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Rosner, Fabian

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Design and Thermo-Economic Analyses of Solid Oxide
Fuel Cell-Gas Turbine Hybrid Systems with Water
Recovery

DISSERTATION

submitted in partial satisfaction of the requirements
for the degree of

DOCTOR OF PHILOSOPHY

In Mechanical and Aerospace Engineering

By

Fabian Rosner

Dissertation Committee:
Professor Scott Samuelsen, Ph.D., Chair
Professor Jack Brouwer, Ph.D.
Professor Jaeho Lee, Ph.D.
Professor Ali Mohraz, Ph.D.
Ashok Rao, Ph.D.

Design and Thermo-Economic
Analyses of Solid Oxide Fuel
Cell-Gas Turbine Hybrid
Systems with Water Recovery

DEDICATION

To
My family.

*“The people who are crazy enough to think they can change the world
are the ones who do.”*

Steve Jobs

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NOMENCLATURE

Roman Symbols

a	Activity, in –
a	Amorphous Carbon Correction Factor, in –
a	GT Map scaling Parameter, in –
A	Area, in m^2
A_s	Surface Area, in m
AER	Annual Escalation Rate, in –
b	GT Map scaling Parameter, in –
c	GT Map scaling Parameter, in –
CCF	Capital charge factor, in –
c_D	Recuperation Efficiency of the Diffuser, in –
CF	Plant Capacity Factor, in –
C_{GT}	GT Cost, in \$
COE	First Year Cost of Electricity, \$/MWh
CSF	Cost Scaling Factor, in \$/kW
D	Diameter, in m
d_{pore}	Average Pore Diameter, in m
D^{eff}	Effective Diffusivity, in m^2/s
E	Cell Potential, in V
F	Faraday Constant, 96,485 C/mol
$\dot{H}^0$	Standard Enthalpy of Combustion based on the Fuel's LHV, in kW
H_i	Van't Hoff Constant, in bar^{-1} and dimensionless for H_2O
i	Current, in A
j	Electric Current Density, in A/m^2
k	Specific Heat Ratio c_p/c_v , in –
k_0	Frequency Factor CH_4 Reforming, in $mol/s/m^2/bar$
$K_{Equi,(k)}$	Equilibrium Constant of Reaction (k), in $bar^{\sum v_i}$
kW	ISO Power Rating of GT, in kW
$\dot{m}$	Mass Flow, in kg/s
M	Molecular Weight, in g/mol
$\bar{M}$	Average Molecular Weight, in g/mol
Ma	Mach Number, in –
MWH	Annual Net-megawatt Hours Generated at 100% CF, in MWh
n	Number of Electrons Participation in the Electrochemical Reaction, in –
N	Operating Spool Speed of the GT, in rpm
n_v	Velocity Profile Exponent, in –
OC_{fix}	Total Fixed Annual Operating Costs, in \$

OC_{var}	Total Variable Annual Operating Costs, in \$
p	Pressure, in bar
$p_{(i)}$	Partial Pressure of Species i , in bar
r	Domain in Radial Direction, in m
R_3	Mixing Chamber Radius, in m
$r_{(k)}$	Reaction Rate of Reaction (k), in kmol/kg _{cat} /h
R	Universal Gas Constant, 8.314 J/mol/K
R	Resistance, in Ω .
RC	Reference Cost, in \$
RFC	Relative Fuel Cell Cost, in –
RP	Reference Parameter Used to Scale the Equipment, dependent on equipment
RY	Reference Year, in –
$\hat{s}_{\text{rxn}}$	Reaction Entropy, in J/mol/K
SC	Scaled Cost, in \$
SP	Scaled Parameter, dependent on equipment
SY	Scaled/Projected Year, in –
T	Temperature, in K/°C/R /°F
TIT	Turbine Inlet Temperature, in K
TOC	Total Overnight Capital, in \$
u	Scaling Exponent, in –
v	Velocity, in m/s
ν_i	Stoichiometric Coefficient of Reaction i , in –
V_{SOFC}	Operating Voltage of the Fuel Cell, in V
x_i	Liquid Mole Fraction of Component i , in –
y_i	Vapor Mole Fraction of Component i , in –

Greek Symbols

α	Electron Transfer Coefficient, in –
β	Equilibrium Factor, in –
δ	Thickness, in m
$\Delta\alpha_{\text{VIGV}}$	Change in VIGV angle, in °degrees
Δp	Pressure Drop, in bar
ΔQ	Heat Loss, in kW
ε	Porosity, in –
γ	Pre-exponential Factor Exchange Current Density, in A/m ²
$\mathcal{A}$	Reforming Reaction Variable, in –
η	Efficiency, in –
η_{act}	Activation Loss, in V
η_{D}	Efficiency of Pressure Conservation in an Ideal Diffuser, in –
η_{ohmic}	Ohmic Loss, in V
η_{conc}	Concentration Loss, in V
ρ	Density, in kg/m ³
π	Pressure Ratio, in –

Ψ	Isentropic Coefficient, in –
Σv	Diffusion Volume, in –
τ	Tortuosity, in –
ξ	Friction Factor, in –
ζ	Friction Coefficient, in –

Subscripts

12	Binary
corr	Corrected
HC	Binary Hydrogen Carbon
i	Running Identifier for Cost Parameter
K	Knudsen
norm	Normalized
OC	Binary Oxygen Carbon
p	Primary Flow
Ref	Reference
s	Secondary Flow
Scale	At Full Scale
Sim	Simulation Value
t	Nozzle Throat
TIT	Turbine Inlet Temperature

Abbreviations

CSF	Cost Scaling Factor
EF	Externally Fired
GT	Gas Turbine
HHV	Higher Heating Value
LHV	Lower Heating Value
LSCF	Lanthanum Strontium Cobalt Ferrite
LSM	Lanthanum Strontium Manganite
PEN	Positive-electrode, Electrolyte, Negative-electrode
RFC	Relative Fuel Cell Cost
SOFC	Solid Oxide Fuel Cell
TET	Turbine Exit Temperature
TIT	Turbine Inlet Temperature
TMC	Transport Membrane Condenser
TPB	Triple Phase Boundary
VIGV	Variable Inlet Guide Vane
YSZ	Yttria-stabilized Zirconia

CURRICULUM VITAE

Fabian Rosner

E-Mail: rosnerf@uci.edu

EDUCATION

Doctor of Philosophy in Mechanical and Aerospace Engineering

University of California, Irvine, USA | 2021

M.S. in Mechanical and Aerospace Engineering

University of California, Irvine, USA | 2018

M.S. in Chemical Engineering

Specialization: Chemical Process Engineering
Technische Universität München, Germany | 2015

B.S. in Chemical Engineering

Technische Universität München, Germany | 2012

Abitur

Carl-Spitzweg-Gymnasium, Germany | 2009

CERTIFICATES

Gas Turbine Combustion Short Course

University of California, Irvine, USA | 2019

Advanced School on Water Energy Food Nexus

Sao Paulo School of Advanced Science, Brazil | 2018

AWARDS

CERC-WET Fellow for Outstanding Contributions to Research, **DAAD Promos** (Scholarship), **Förderung Studienbezogener Auslandsvorhaben** (Scholarship), **SPSAS** (Scholarship)

LANGUAGES

German: Native

English: Proficient

Latin: Final Qualification: Latinum

Chinese: Beginner

EXPERTISE

Sustainable Energy Systems, Energy Conversion, Energy Storage, Renewable Fuels, System Integration, Water Energy Nexus, Model Development, Fuel-Cells, Fuel-Cell Gas Turbine Hybrid Systems, Water Recovery, Gasification, Integrated Gasification Combined Cycles, Reactor Design, Carbon Capture, Nano-Particle Synthesis and Characterization

EXPERIENCE

Process Design, Modeling & Techno-economic Analyses | Ph.D. Research

University of California, Irvine, USA
September 2015 to 2021

Process & Reaction Engineering | M.S. Thesis

University of California, Irvine, USA
September 2015 to June 2018

Surface Chemistry & Catalysis | Employment

University of California, Irvine, USA
April 2015 to June 2015

Surface Chemistry & Catalysis | M.S. Thesis

University of California, Irvine, USA
July 2014 to March 2015

Patent Attorney | Internship

Richter Werdermann Gerbaulet & Hofmann, GER
April 2014 to May 2014

Gasification & Pyrolysis Reactor Design | Research Project

Technische Universität München, GER
October 2013 to June 2014

Technical Mechanics I & II | Teaching

Technische Universität München, GER
May 2013 to March 2014

Reaction & Process Modeling | Internship

CSIRO Process Science and Engineering, AUS
January 2013 to February 2013

Photo-Catalytic Nanoparticle Research | Research Project

The University of Melbourne, AUS
September 2012 to January 2013

Numerical Fluid Mechanics | B.S. Thesis

Technische Universität München, GER
May 2012 to October 2012

Mechanical Design & Engineering | Internship

KADE GmbH, GER
April 2012 to May 2012

Manufacturing | Internship

Schreiner Maiwald, GER
September 2010 to October 2010

Car Design | Internship

BMW Group, GER
March 2005

TEACHING AND MENTORING

Teaching Assistant for Technical Mechanics I & II,
Instructor and TA for the ASPEN Plus Part of
Combustion and Fuel Cell Systems, Supervising
Undergraduate and Graduate Research Projects

COMPUTER SKILLS

Proficient in: MS Word, Excel, Power Point, LaTeX
Worked with CAD: CATIA, SolidWorks, Pro/ENGINEER
Expertise in Programming: C, Fortran, VBA, ACM
Experienced in Simulation: ASPEN Plus, ASPEN CM,
ASPEN EDR, ThermoFlex, AnsysFluent,
OpenFOAM

LABORATORY SKILLS

Auger-Electron Spectroscopy (AES), High Resolution
Electron Energy Loss Spectroscopy (HREELS),
Scanning Electron Microscopy (SEM), UV-Visible
Light Spectrometry (UV-Vis), Fluorescence
Spectrometry, Dynamic Light Scattering (DLS)

PROFESSIONAL SERVICES

Reviewer of: Applied Energy, Energy Conversion and
Management, Applied Thermal Engineering

Member of: The American Society of Mechanical
Engineers, American Institute of Chemical Engineers

PUBLICATIONS

Journal Articles:

- 1) Thermo-economic analyses of isothermal water gas shift reactor integrations into IGCC power plant
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11) Thermo-Economic Analyses of Solid Oxide Fuel Cell-Gas Turbine Hybrids Considering Thermal Cell Gradients

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12) Thermo-Economic Analysis of Integrating an Off-the-shelf Gas Turbine into a Solid Oxide Fuel Cell-Gas Turbine Hybrid

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Manuscript in Preparation

13) Nano-porous Transport Membrane Condenser Modeling and Parametric Design Analysis

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Manuscript in Preparation

14) Thermo-Economic Analyses of Carbon-Neutral IGCC Power Plants via Carbon Capture and Biomass Cofeeding: A Montana Case Study

Rosner, F., Fong, A., Rao, A., Wang, B., Zhao, L., Samuelsen

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Note) Results on water recovery are expected to be published in archival literature in the near future

Other Articles:

1) Economics of Cell Design and Thermal Management in Solid Oxide Fuel Cells under SOFC-GT Hybrid Operating Conditions: Part II, Pressure Dependency

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2) Transforming Water Consuming Power Plants into Water Producing Power Plants

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ABSTRACT OF THE DISSERTATION

Design and Thermo-Economic Analyses of Solid Oxide Fuel Cell-Gas Turbine Hybrid Systems with Water Recovery

By

Fabian Rosner

Doctor of Philosophy in Mechanical and Aerospace Engineering

University of California, Irvine, 2021

Professor Scott Samuelsen, Ph.D., Chair

Highly efficient electric power systems are crucial for the reduction of carbon emissions from fossil fuels, as well as, in a circular economy where renewable fuels like biogas, biomass and hydrogen need to be converted to electricity in an efficient manner to minimize the environmental footprint. Moreover, current thermoelectric electricity generation requires enormous amounts of water, primarily for cooling or heat rejection, which creates tensions and interdependencies between the electricity sector and the water sector. To address these concerns, this dissertation focuses on the design of highly efficient, economically viable solid oxide fuel cell (SOFC)-gas turbine (GT) hybrid systems to minimize the environmental impact of electricity generation including the water use, and further addresses the system integration and the economics of flue gas water recovery technologies.

Two of the greatest challenges to widespread adoption of SOFC technology are thermal cell management and costs. Applying a unique bottom-up SOFC cost approach, cell design parameters are identified that synergistically reduce thermal gradients and specific SOFC stack cost. Using this optimized cell design, the SOFC stack is integrated into a SOFC-GT hybrid system. Through research of this SOFC-GT hybrid system over a wide range of pressure conditions, fuel utilization factors, and operating voltages, cost-competitive operating points and major cost driving factors contributing to the cost of electricity are identified while considering thermal SOFC and GT aerodynamics constraints associated with SOFC-GT systems.

Based upon the results of the hybrid system design study, multiple water recovery technologies, such as an air-cooled condenser, a direct contact condenser, a LiBr absorption system, a monoethylene glycol absorption system and a transport membrane condenser, are integrated into the hybrid and studied in combination with various fuel types at numerous scales. In the studied configurations, the direct contact

condenser and the transport membrane condenser suffer from low water recovery rates, while the air-cooled condenser's large pressure drop leads to substantial system integration losses. The absorption systems achieve the highest water recovery rates and the LiBr absorption system exhibits the most favorable economics. The absorption systems are particularly advantageous in biogas scenarios where the latent heat can be upgraded and made available to the digester.

Chapter 1

Introduction

1.1 The Water Energy Nexus

Flicking a switch and the bathroom lightens up in a clear sheen, a few more steps to turn on the faucet and fresh water pours into the basin; a lifestyle we take for granted, not just at home but in almost every aspect of our lives. However, the question arises: What enables us to freely use electricity and water. At first glance, there does not seem to be any interconnectedness between the two; nevertheless, in reality the resiliency of water and energy systems are strongly coupled. While current thermoelectric electricity generation requires enormous amounts of water, primarily for cooling or heat rejection, water distribution and treatment systems demand electricity for pumping and purification. The growing importance of biomass in electricity generation, due to its low carbon footprint, and the associated water demand and land resources add further complexity to the interconnectedness between water and energy systems. Beyond that, energy resources are needed to ensure reliable feedstock availability and food security. It is important to realize that not just the energy sector has a significant impact on the water and food security but also that water resources influence food production as water is needed to grow food crops, raise livestock, and process food. On the other hand, food production affects water security in the sense of potential water pollution by nutrients and other chemicals. An illustration of the interplay between the different systems affecting the water-energy nexus is shown in **Figure 1-1**.

The reliance of the water and energy systems on one another is critical, particularly in the event of natural disasters where water and electricity are crucial to support the most fundamental human needs. Hurricane Sandy showed how damaged infrastructure and power plants can impair the water distribution and electricity generation. In the future, change of climate and policies addressing greenhouse gas emissions, like the U.S. Department of Energy's (DoE) target of reducing carbon dioxide emissions by 90%, exacerbate the stress on our water and energy systems as many capture technologies increase water consumption and diminish the plant's efficiency. With water, electricity and food needs set to further increase, it is of immense importance to adequately comprehend the linkages between the water and electricity sector as well as the agricultural sector.

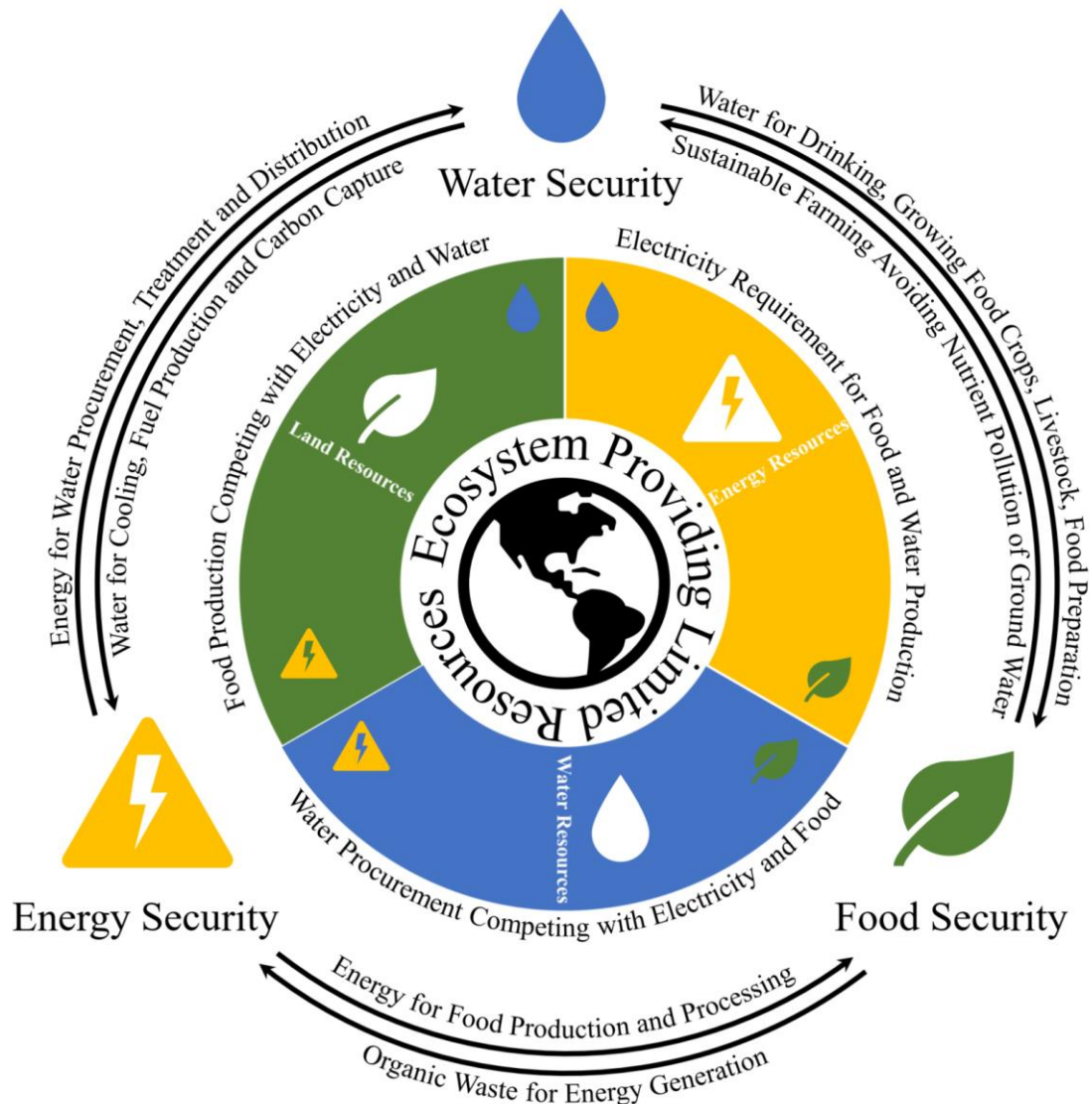


Figure 1-1: The water energy nexus in the context of the ecosystem.

Water is required in every step of energy procurement ranging from natural gas or petroleum production to biomass cultivation, synthesis of biofuels and power plant operation. Today the electricity sector is responsible for the largest water withdrawals in the United States (U.S.). In 2011, 196 billion gallons of water per day were withdrawn from fresh water and saline water sources for heat rejection in thermoelectric power generation. This represents a share of 48% of all water withdrawn, with fresh water accounting for 71% [1]. The development of advanced cooling technologies seeks to reduce these water withdrawals; however, with policy changes shifting away from once-through cooling, increases in water consumption are to be expected [2]. In addition to the water withdrawal and water consumption, power plants and especially coal fired power plants generate highly contaminated wastewater that needs special

treatment on site before it can be discharged creating additional linkage between the water and energy sectors. At the same time the power sector in its current form dissipates enormous quantities of thermal energy due to inefficiencies in electric power generation. The Sankey diagram in **Figure 1-2** illustrates the interplay of various energy carriers and water resources.

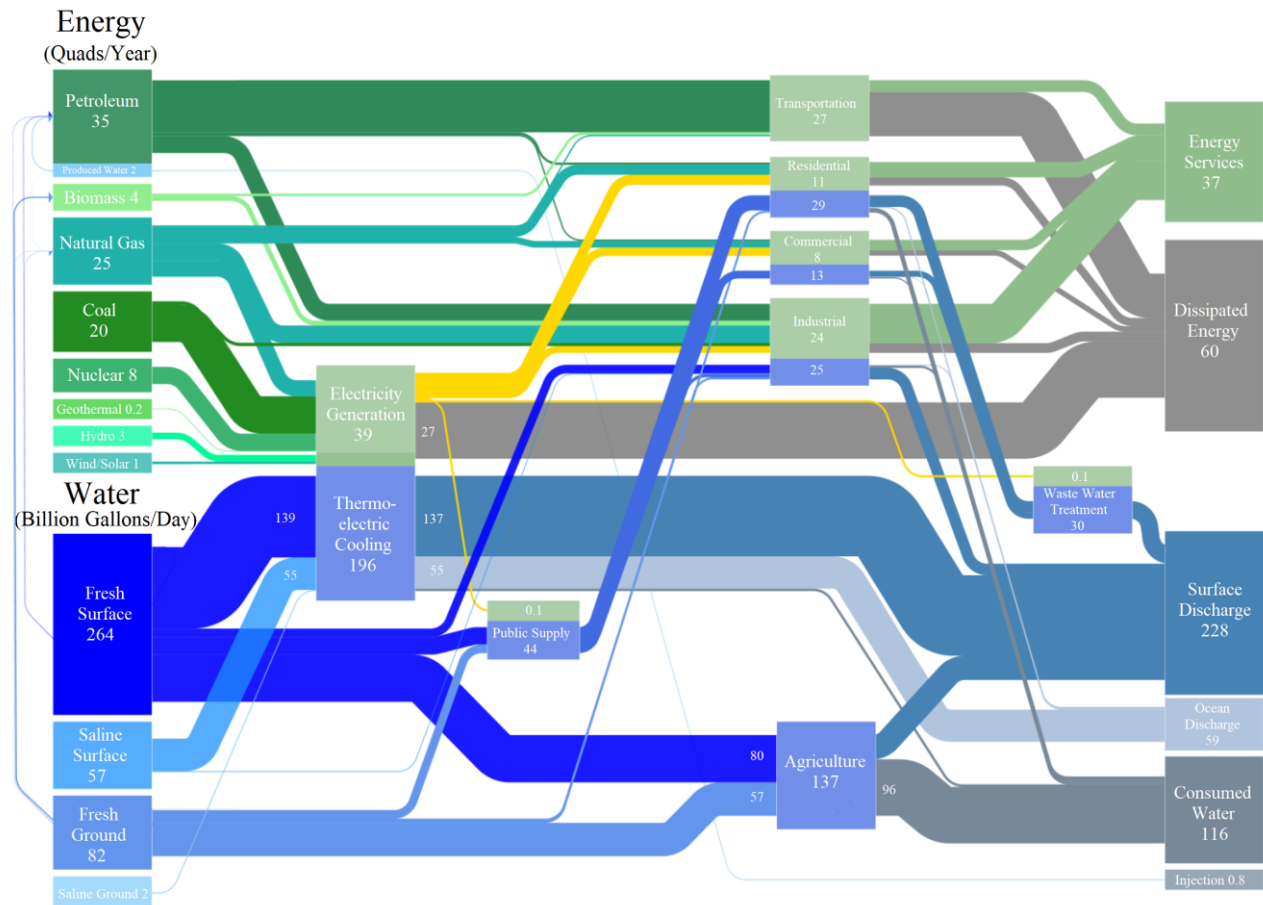


Figure 1-2: Hybrid Sankey diagram of 2011 U.S. interconnected water-energy flows [1].

As it can be seen in **Figure 1-2**, agricultural water use is in direct competition with the power sector. Agricultural resources like water and land are not only used for food production but also for the production of renewable energy resources and renewable fuels. The growing demand for biomass and biogas creates tensions at the interfaces between the different sectors. As a result, food security enters into a rivalry with the water and energy sectors for power, fresh water and land use.

On the other hand, electricity is needed not only for the treatment and purification of wastewater from power plants but also in municipal facilities to treat wastewater from residential, commercial and industrial customers. The treatment of wastewater can account for a large share of a municipality's energy bill. Although the wastewater itself contains significant amounts of energy in the form of organic constituents, the potential remains untapped in most cases. The global energy demand for water treatment reached a scale

The Nexus Approach

comparable to the entire economy of Australia. The largest portion of this energy is supplied in the form of electricity. About 4% of the worlds electricity generated in 2014 was used to extract, distribute and treat water and wastewater, along with 50 million tonnes of oil equivalent of thermal energy (mostly diesel for pumping) [2]. More severe weather events will further challenge our water systems and very costly and energy intensive technologies such as desalination, will be needed to overcome the increasingly severe droughts.

The needs for water, energy and food will increase in the coming years and thus, it is of great importance to identify viable solutions to resolve tensions in the water energy food nexus. Fuel cell technology has proven to operate highly efficient while exhibiting ultra-low pollutant emissions, near-zero acoustic emissions, and zero cooling water demand. In contrast to thermoelectric power generation, fuel cells are scalable and can reach very high efficiencies even in small scale applications, which makes them an ideal candidate for local power generation and co-generation.

By combining fuel cell technology with novel integration schemes and water recovery systems, significant water savings can be achieved. This can help to relieve stress on our energy, water and food systems. In many dry areas today, water is already a limiting factor in electricity generation. Decoupling the interdependency of the water and energy systems will substantially increase the resiliency and reliability of our water and energy infrastructure.

1.2 The Nexus Approach

The word “nexus” originates from Latin and describes the act of binding together or bond. The nexus approach gained increasing awareness in the academic research and policy community over the last decade. The nexus approach can be applied to many areas but attained prominence in the context of the water-energy-food nexus. The essence of the nexus approach is the analysis of interactions of components in the context of an entire network considering linkages and interdependencies, instead of looking at individual components independently. This approach allows to identify synergistic effects between different sectors as well as potential dangers to the functionality and operation of the system. This nexus approach is not just limited to technical aspects but also includes the interactions of technology with socio-political aspects.

As engineers we are mostly concerned with aspects of performance and efficiency. Applying the nexus approach on a purely technical level can help to identify areas where improvements are viable and useful. Translating this into an everyday example with the set goal of cleaning the kitchen counter and floor, one can easily clean the counter by just wiping the breadcrumbs onto the floor. However, this will increase the workload for cleaning the floor. Throwing the breadcrumbs into the trash right away makes the cleaning process of the kitchen counter more resource intense, however, might improve the results of the overall kitchen cleaning process. In a more industrial environment one can think of the following situation:

reducing the heat input at the bottom of a thermal separation column might not be necessary if the heat at the top of the column can be recovered and integrated in such a way that it can benefit the boiler at the bottom of the column. As the systems of interest get larger, the complexity increases significantly, and viable integration strategies are more difficult to identify. When systems reach the size of whole countries like the national electricity grid or water supply network, local optimizations also have to be analyzed on a system level including interactions with interfaces and the environment.

Moving from a purely technical perspective to a level where technology interacts with its environment, these technical solutions can be defined as technical source term as they constitute the source of change to a system. In the case of an energy application this source term could be a more efficient energy conversion device. Dispersion of the source's output variables include electricity transmission and distribution as well as the distribution of unwanted products like emissions. The dispersion determines the range and extent at which source influences can be perceived. The effects on the other hand are concerned with the end use of the source's products but also with the impact of undesirable side effects. Common concerns for electricity generating plants are the water consumption as well as environmental impacts and health risks. These concerns can only be addressed in combination with the consideration of socio-political aspects as shown in the Buford Model [3], **Figure 1-3**.

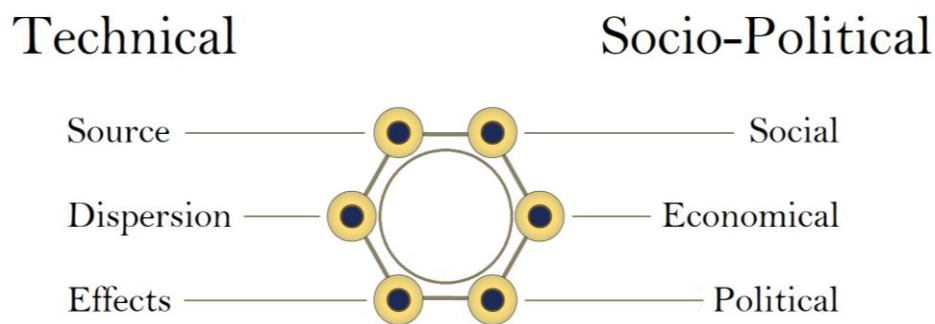


Figure 1-3: Buford Model [3]

Social, economic and political aspects are closely interwoven with the technical side. Even the best technical solution will fail if social, economic and political facets are missing. First and foremost, the technical solutions need to benefit society, and in most cases need to show some economic advantage. With the increasing amount of renewable electricity on our grid and the need for energy storage, this economic aspect becomes a major challenge as many grid supporting systems that provide reliability and resiliency currently do not have a dollar value associated with their service. Politics often acts as a mediator between the different technical and socio-political aspects. Numerous environmental protection and air quality regulations have been signed into law with enormous success. In the 1950s, legislation helped to combat smog plagued cities like Los Angeles. Other successful examples of legislation mitigating harmful

emissions and improving life quality are the regulation of sulfur dioxide, nitrogen oxide and particulate matter emissions. Imposing such regulations is typically causing the price of a product to increase which again impacts social and economic aspects, and thus regulative measures have to be carefully chosen. However, delayed regulatory measures on urgent topics like climate change and carbon dioxide emissions can have fatal consequences for millions of people worldwide. Industrial countries are responsible for a majority of greenhouse gas emissions and social justice demands that technical solutions fighting climate change are affordable and compatible with the often less developed infrastructure in developing countries.

1.3 Sustainable Electricity Generation

Utilizing natural resources to do useful work and satisfy human needs has been achieved since ancient times. Ancient seafarers used sails to capture the wind and farmers once used windmills for grinding grains to flour. Also, the power of the sun has been made use of by humans for centuries for simply drying or sparking fire. During the industrial revolution mankind moved away from these natural resources due to the enormous amounts of energy enclosed in fossil fuels and their convenient and more controllable availability. However, the tremendous success of fossil fuel driven applications led to enormous amounts of combustion products being emitted into the atmosphere. Over time, concerns regarding criteria pollutant emissions grew and later also greenhouse gas emissions and their impact on the climate became an objection.

Today, with the awareness of the urgency to act on climate change and the emissions of carbon dioxide associated with the increasing temperatures, the power and transportation sectors undergo a drastic change away from fossil fuels towards a cleaner, renewable and sustainable future. Soon also other sectors in manufacturing and industry will be confronted with transitioning their production and supply chains to value focused, environmentally neutral processes. Focusing on just one of the aforementioned attributes (clean, renewable, and sustainable) will not suffice to preserve the natural balance. “Clean” production describes a value creation chain without causing harm to humans and environment. A “renewable” production ensures that resources used in the value creation chain can be restored and a “sustainable” production guaranties that the renewable resources will be replenished at the same rate at which they are consumed.

Key in achieving an environmentally neutral production will be the transitioning from cost focused, linear supply chains to a value focused, circular economy. In a circular economy, products or their components are recycled back into the supply chain after usage or consumption in order to utilize the resource as long as possible. At the end of a product’s life, the product is collected, and the rest value is recovered. This reduces raw material extraction and minimizes waste products, which will have to be disposed and often safely stored for a long time. On the technical side of the circular economy raw materials are extracted to manufacture parts and goods, which will eventually be used in society. Different measures

can be taken to maximize the use of resources over their lifetime: 1) maintenance can extend the lifetime of goods, 2) after use, intact goods can be redistributed and reused, 3) faulty equipment can be refurbished and remanufactured, and 4) if completely broken, the raw material can be extracted and used to manufacture new products. An illustration of the resource and product flow patterns in a circular economy are shown in

Figure 1-4.

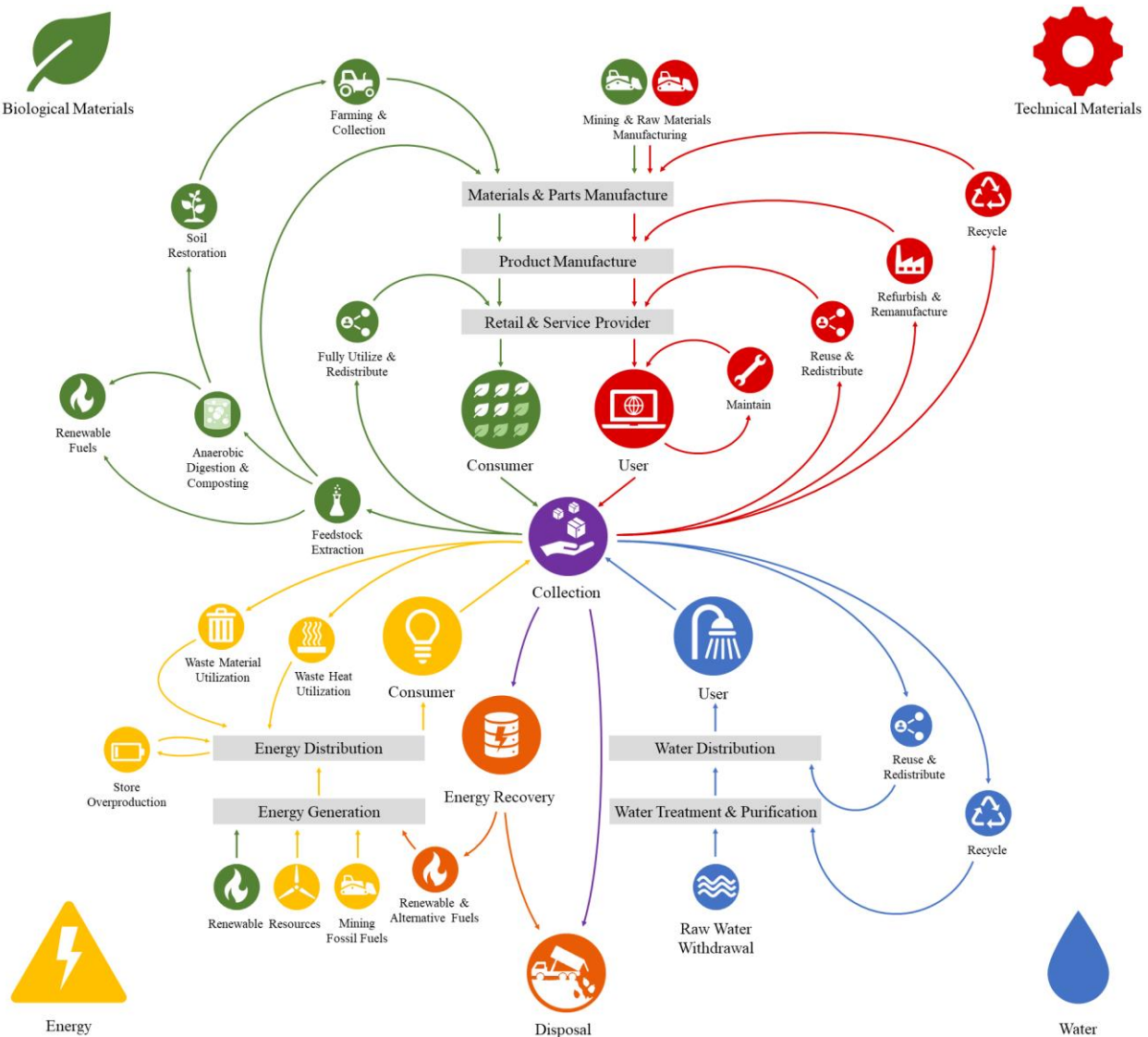


Figure 1-4: The water-energy nexus in the context of the circular economy.

The current economic system is built around cost reduction in material procurement and manufacturing, at the same time, companies are not concerned with environmental impacts (unless laws have been established) and their products afterlife. An example is the manufacturing of lithium-ion batteries. Lithium ore mining leaves behind exploited and destroyed landscapes and at the same time, no infrastructure is in place until today to deal with the battery waste at the end of their lifetime. In order to change the way of

Sustainable Electricity Generation

how companies are operating, political measures have to be taken to have companies focus on product development that enables cost effective extraction of resources from products after reaching their end of life and enable resource recycling.

On the biological product side, products like food are mostly consumed. However, organic waste as well as organic matter passed through the digestive tract contain significant amounts of energy. In order to extract the value left in the “product”, biological processes such as anaerobic digestion can be utilized to produce alternative, renewable fuels but also can help with soil restoration.

The central question left is: how does clean, renewable and sustainable water and energy procurement integrate into the system of a circular economy. Water can be either consumed or used. Consumed water leaves the water infrastructure and may not be available for re-use for a long time, e.g., watering the lawn will lead to large amounts of water seep away and/or evaporate. Used water, on the other hand, can be collected and reused and/or recycled, examples are used shower water or even the water we drink. In order to reduce raw water withdrawal, it is important to increase the water use efficiency, limit water losses and collect used water to enable reuse and or water recycling. As mentioned earlier, the electricity generating sector is responsible for enormous amounts of water withdrawals for cooling. Significant amounts of this water are used for once-through cooling which already has been banned in many places due to the ecological concerns of such systems. However, with the shift to evaporative cooling the water consumption of the electricity sector is set to increase. Since in many applications the used water cannot be reclaimed, the recycling potential in the water sector is very limited. Thus, technologies with a low water footprint are needed which especially applies for the energy sector.

Energy is used in the form of fuels, electricity and heat. For fuels and electricity there are almost no reuse, redistribution and recycling options available. Only waste heat is utilized in some applications. Heat has a low quality of energy and is often a waste product in many energy conversion applications. Therefore, using electricity for heating applications can be seen as wasteful. Combined heat and power systems offer a very effective solution to make complete use of the energy contained in a feedstock. Unique about the electricity sector (part of the energy sector) is that the electricity needs to be generated at the same rate at which it is consumed in order to ensure the stability of the grid. When incorporating renewable energies like wind and solar into the electric grid, storage options are needed to balance natural fluctuations in generation with the demand. For energy storage at the timescale of days batteries offer excellent performance. However, for energy storage over longer periods, e.g., seasonal shifting to balance a demand and generation mismatch of renewable energies between summer and winter, fuel-based energy storage options become more effective. In order to maximize the value of the used resources highly efficient fuel to electricity conversion devices with low emission profile and low water demand are of major importance. A key technology in achieving these goals is fuel cell hybrid technology. High temperature fuel cells like

solid oxide fuel cells (SOFC) can operate on renewable H_2 produced from renewable electricity during times of overcapacities as well as on renewable methane produced via digestion of organic matter which is an important renewable byproduct of the circular economy. However, until sufficient renewable energies are deployed it will be inevitable to operate those systems on conventional fossil fuels where their high efficiency helps to offset carbon emissions until renewable fuels are available.

1.4 Goal and Objectives

The electric power generation sector faces challenges with respect to its environmental sustainability. On the one hand, emissions of criteria pollutants and greenhouse gases originating from the combustion of fossil energy carriers create urban and regional air quality and global climate change challenges. On the other hand, enormous water withdrawal rates deplete water resources and stand in competition with local food production. To address these challenges, this study selects a power generation technology with the potential of virtually zero emission of criteria pollutants (to mitigate if not eliminate degradation of urban and regional air quality), an unusually high fuel-to-electricity conversion efficiency (to mitigate if not eliminate the emission of carbon and the consumption of water), and adaptation to active flue gas water recovery. The power technology selected is the SOFC-GT hybrid with an application range from distributed to central generation and the flexibility to operate from coal to natural gas (for conventional applications) to renewable hydrogen (for future applications). In this study, fuels investigated include: coal and natural gas (used during a transition period while mitigating greenhouse gas emissions due to highly efficient energy conversion), as well as biomass and biogas (future applications that are carbon neutral or even carbon negative).

The goal of this study is to delineate a technology evolution roadmap for SOFC-GT hybrid systems that can resolve tensions in the water energy nexus by transforming power plants from sources of pollutants into zero pollutant and zero carbon sources, and from water consumers into net-producers of water. Turning power plants into sources of water is a substantial step in the direction of higher sustainability, resiliency, and reliability of both: the water infrastructure and the energy infrastructure. Additionally, to further reduce the linkage between the water and energy sector, water treatment strategies and zero liquid discharge systems are investigated in order to eliminate any wastewater discharge from the plant.

Successful accomplishment of this research goal requires a systematic approach. To achieve the research goal, the following seven objectives are prescribed.

OBJECTIVE 1: DEVELOP METHODOLOGY

OBJECTIVE 2: ESTABLISH A DESIGN BASIS AND IDENTIFY STUDY SCENARIOS

OBJECTIVE 3: DEVELOP WATER USE REDUCTION STRATEGIES AND IDENTIFY WATER

RECOVERY TECHNOLOGIES

OBJECTIVE 4: DEVELOP AND CALIBRATE MODELS OF KEY TECHNOLOGIES

OBJECTIVE 5: DEVELOP BASE CASE SCENARIOS

OBJECTIVE 6: DEVELOP CASES WITH WATER RECOVERY

OBJECTIVE 7: THERMO-ECONOMIC ANALYSIS

1.5 Dissertation Outline

The first part of this dissertation provides an introduction to feedstocks, SOFC-GT hybrid systems and water recovery from flue gases. Thereafter, a comprehensive literature review on SOFCs, hybrid technology and water recovery is presented, as well as an analysis of the current technology status.

The second part of the dissertation focuses on the approach of the above-mentioned objectives, which will be restated as tasks with explanations of the work scope. The ensuing section describes the design basis, introduces modeling approaches of key subsystems, and describes the methodology for the economic analysis. Since various feedstocks are investigated in this work, the key components have been categorized in a common process area which includes processes that are common to all plants and various subcategories with case specific process equipment. The results section is divided into two sections: 1) results on SOFC-GT hybrid systems without water recovery that inform the plant design of the reference case without water recovery, which will function as a baseline scenario for the cases with water recovery 2) results of the studies with water recovery. At the end a final summary is presented, important conclusions are summarized in the form of statements and recommendations for further research are given.

Chapter 2

Background

2.1 Fuels and Fuel Processing

Fuels are energy carriers and have been utilized for centuries to generate electricity. Many fuels exhibit exceptional energy densities and are uncomplicated to store. This makes them ideal candidates for on demand energy supplies in stationary as well as mobile power applications. Historically, fossil fuels have dominated the energy sector due to their easy accessibility, however, over the recent years fossil fuels raised concerns with respect to their impact upon climate. As a result, renewable fuels have gained interests to mitigate impacts upon climate while still being able to make use of the above-mentioned positive aspects of fuel-based energy systems. In power systems, fuel or exhaust gas processing is needed regardless of the feedstock type, natural gas, biogas or gasified coal and biomass, in order to comply with the stringent environmental regulations. Moreover, certain contaminants can have negative impacts on the plant operation, plant efficiency and durability of equipment and catalysts. The following section provides a short overview of some of the most common fossil and renewable fuels as well as fuel processing.

2.1.1 Fossil Fuels

Natural Gas

Fuel Characteristics. Natural gas is a fossil fuel mainly consisting of methane. Small amounts of higher alkanes are commonly found in natural gas together with very little quantities of CO₂, N₂, H₂S, and/or He. Natural gas has many advantages compared to other fossil fuels. With its high energy content per carbon atom, methane has a small CO₂ emission profile among the fossil fuels. In many cases only minor clean-up treatments of the raw natural gas are needed to reach pipeline quality and the sulfur content of natural gas is significantly lower than for other common feedstocks. Problems with ash handling, heavy metals and particulate matter emissions are no concern when utilizing natural gas for electricity production. However, for fuel cell systems the feed gas purity requirements are significantly higher than for traditional GTs. Other than combustion-based GTs, fuel cells operate with catalysts which are sensitive to sulfur and oxygen poisoning. Small quantities of H₂S are commonly found in natural gas. Limits for H₂S in pipeline gas are 0.25 – 1.00 gr/100 SCF (7000 gr = 1 lb) [4]. Additionally, sulfur compounds are added to natural gas as odor because of safety reasons. Odors are typically a mixture of different types of mercaptans. Main

component and most widely used are odors based on tert-butyl-mercaptan. Supplier limits on mercaptans concentrations in pipeline gas are 1.0 – 10.0 gr/100 SCF [4].

Oxygen is not inherently found in natural gas. Oxygen is mainly introduced through leakages in the low-pressure gathering system, for example the vapor recovery compressor on an oil production tank battery where the suction pressure can be lower than the atmospheric pressure [5]. Specifications on maximum O₂ content are set by the pipeline operator and can vary from 0.2 mol-% [4] to concentrations as low as 10 ppmv [5]. Oxygen concentrations higher than 50 ppmv might have negative influences on gas processing as well as enhancing pipeline corrosion. Commonly about 0.01 mol-% of O₂ are found in natural gas [6]. In peak times of energy demand, peak shaving with liquified natural gas (LNG) and propane is a common practice to increase the natural gas supply. Liquid natural gas (LNG) is usually generated during times of low demand and does not contain contaminants like N₂. Peak shaving with LNG is rather unproblematic and easy to realize. Due to the higher purity of the LNG the heating value of the pipeline gas increases up to 8.5% [4]. Peak shaving with propane on the other hand is more critical. Instead of injecting LNG into the pipeline, propane is injected. However, propane has a significantly higher heating value per unit of volume and O₂ is used in order to dilute the propane and balance its heating value. This can give rise to an increase in the O₂ concentration in the pipeline gas, reaching O₂ concentrations of up to 4% [4].

The supply pressures of the natural gas at the plant sites may vary by operator, location and demand. While the standard delivery pressure for natural gas is about 8 inch-water, higher delivery pressures can be requested for loads of greater 290 kW (1,000,000 BTU/hour).

Fuel Processing. Although sulfur and O₂ trace amounts are no problem for traditional GT applications, they will have to be removed when using fuel cells to avoid catalyst poisoning. Sulfur components are commonly removed by expendable adsorption processes using activated carbon. Adsorption is driven by concentration differences and favored under high pressure and low temperature conditions. Expendable processes are typically economically more viable than regenerative processes when it comes to natural gas due to the low sulfur concentrations.

Trace amounts of O₂ can be removed with nonregenerative scavengers at lower temperatures. Higher concentrations or high temperature applications require the use of a catalyst in order to react the O₂ with CH₄. If desired the produced water can be removed later in a dehydration step [5]. However, catalytic cleanup processes will affect the heating value of the natural gas negatively as the CO₂ will remain in the gas stream.

Another important processing step, when utilizing natural gas in fuel cells, is pre-reforming. Solid oxide fuel cells are made of ceramics and thus, are sensitive to thermal stresses. Methane reforming is an endothermic reaction and can lead to a steep temperature decrease inside the fuel cell. In order to avoid

conditions of large temperature gradients, pre-reforming is used to mitigate those effects. Other advantages of pre-reforming include a higher fuel flexibility and additional protection of the fuel cell from poisoning.

The fuel supply pressure of a SOFC-GT hybrid is set by the gas turbine pressure ratio, the eductor, and the fuel control valve. For small scale applications, where natural gas is supplied at low pressures, the natural gas might have to be recompressed on site in order to satisfy the system's fuel supply pressure. Especially for systems with eductor and recirculation of the anode off-gas, a sufficiently high pressure is needed to entrain and pressurize the recycle stream. An efficient mean to reduce compression power is via the use of intercoolers.

Coal

Fuel Characteristics. Coal is one of the most abundant fossil energy resources. Worldwide coal reserves surpass oil and gas reserves and will last a very long time into the future. The easy accessibility and uncomplicated handling for shipping makes it an attractive fuel around the globe. Coal is more evenly distributed worldwide than other fossil fuels and it can be found in many countries, which reduces the dependence on oil and gas supplying countries. Particularly in developing countries, like China and India, where large coal reserves exist, coal provides an inexpensive and reliable energy source. However, coal has the highest environmental impact among the fossil energy carriers. Coal with its high carbon content per unit of released energy leads to higher carbon dioxide emissions compared to other fossil fuels. In addition, traditional combustion of coal produces gaseous and solid criteria pollutants (e.g., nitrogen- and sulfur-oxides as well as particulate matter). This raises concerns regarding air pollution and greenhouse gas emissions. Gasification-based energy systems have shown to be an important development on the way to clean power generation from coal combining the advantages of coal usage with low air pollution, low carbon dioxide emission, and high efficiency. During gasification solid coal is converted into a mixture of CH_4 , CO , CO_2 , H_2 , water and many more components in smaller concentrations; also known as synthesis gas or syngas. Syngas can be efficiently cleaned before utilization when trace components are present in much higher concentration before it is diluted with an oxidant stream such as air. A summary of common gasification technologies and their advantages can be found in [7]. More detailed information about gasification can be found in [8], [9].

Coal is formed over millions of years under high pressure and temperature from plant-based organic matter. Based on the development stage or age of the coal, it can be classified in accordance with its carbon content. The older the coal, the higher the carbon content, and higher carbon content ultimately leads to a higher heating value. The evolution of coal starts with biomass, which is first converted to lignite, sub-bituminous and bituminous coal before it eventually becomes anthracite. A summary of the different coals

and their properties on an as-received basis, meaning without any processing at the plant site such as drying, is given in **Table 2-1**.

Table 2-1: Proximate Analysis of Coal Classes [8], [9]

wt-%	Lignite	Sub-Bituminous	Bituminous	Anthracite
Fixed Carbon	25–35	35–45	45–85	85–98
Volatiles	>45	20–45	20–9	<9
Moisture	30–70	10–45	2–16	2–16
Ash	2–30	2–30	2–15	2–10
Lower Heating Value MJ/kg	26–28	28–32	32–36	36–27

This type of analysis is called proximate analysis. Several different bases for coal analysis are common: As-received, air-dry (without free moisture), dry (without any moisture), dry and ash free (daf), and dry and mineral matter free (dmmf). The volatile matter content is a crucial parameter in the proximate analysis with great significance for gasification applications. Volatile matter content is an indicator of the relative reactivity of a coal; the higher the volatile matter content the higher the reactivity, which is beneficial for the gasification process. Structural analyses of coal on an atomistic level are very complex and molecular structures differ widely for different types of coals. However, in most cases the influence of the detailed petrographic composition of the organic compounds has no significant influence on the gasification behavior [9]. Complimentary to the proximate analysis is the ultimate analysis, which provides information about the elemental composition of a coal. Examples of ultimate analyses for various coal types are shown in **Table 2-2**.

Table 2-2: Ultimate Analyses of various Coals [9]

Origin Class	C	H	O	N	S
Germany, Rhein Lignite	67.5	5	26.5	0.5	0.5
U.S., Montana Sub-Bituminous	76.4	5.6	14.9	1.7	1.4
China, Datung Bituminous	84.4	4.4	9.5	0.9	0.8
Germany, Ruhr Anthracite	91.8	3.6	2.5	1.4	0.7

As mentioned before, in order to utilize coal in a downstream Brayton cycle or fuel cell it needs to be gasified; whereby, the resulting syngas still contains many components and trace components originally present in the coal. Typical sulfur contents in coals range from 0.5 – 6 wt.-% and may be present in the form of organic sulfur as well as inorganic sulfur compounds [9]. Sulfur is a poison for many catalysts used in chemical processing (e.g., water-gas-shift reaction promoting catalysts). If combusted in a gas turbine, these sulfur compounds are oxidized to sulfur oxides, which are the source of acid rain if emitted into the atmosphere. In high temperature gasification, 90% of the sulfur components leave the gasifier as H₂S. The remaining sulfur is bound in COS. Trace amounts of sulfur oxides and CS₂ are only present at low temperature gasification processes [8]. In order to remove more than 99% of the sulfur from the syngas,

COS has to be converted to H_2S in a hydrolysis reactor as it is not easily removed by most state-of-the-art acid gas removal systems like SelexolTM.

Next to organic nitrogen bound in the fuel, molecular N_2 can enter the gasifier in the form of air or as transport medium for coal. During the gasification process small amounts of HCN and NH_3 are formed, which originate mostly from the organic nitrogen in the fuel. Nitrogen oxides are essentially not present due to the reducing environment in the gasifier. However, if combusted, both compounds (HCN and NH_3) form nitrogen oxides. Nitrogen oxides are toxic and a contributor to degraded air quality and climate change. Furthermore, the HCN in the raw syngas is known to degrade amines used in some sulfur removal processes. In this case, it is important to remove nitrogen compounds from the syngas before the sulfur is removed. A useful property of HCN and NH_3 with respect to their removal, is their high solubility in water. Syngas scrubbers and/or cooling the syngas below its dew point are efficient and cost-effective ways to remove nitrogen compounds from the syngas.

Beyond the components listed in the ultimate analysis, many other elements present in coal are especially important with respect to environmental and health concerns. Chlorides may be found in coal with chlorine concentrations of up to 2.5 wt-%. Chlorine released from the fuel can form HCl or NH_4Cl , which are not just critical for plant operation but are also concerning for health and environment. HCl can react with metals, such as reactor walls, heat exchangers and piping. The resulting metal salts will start depositing in the syngas cooling section of the plant and pose a severe fouling risk. Similar consequences result from NH_4Cl , which is a salt that solidifies at around 280 °C. Beyond fouling problems, chlorine compounds constitute a poison for catalysts, e.g., catalysts used in low temperature shift reactors, and thus, have to be removed beforehand. Syngas scrubbers are an efficient and cost-effective way to remove these compounds from the syngas.

Furthermore, small amounts of phosphorous are present in many coals [9]. Phosphorous is less critical in coal gasification plants but can become problematic when using biomass feedstocks, which can have significantly higher phosphorus concentrations. Additionally, coal contains traces of heavy metals like mercury, lead and arsenic. Heavy metals have a significant deactivating effect on precious metal catalysts, but more importantly, present a high risk for human health. Furthermore, metals present in coal can form metal carbonyls, such as $\text{Ni}(\text{CO})_4$ and $\text{Fe}(\text{CO})_5$ which are extremely dangerous for any form of living organisms.

Another challenge that comes with the utilization of coal are fine particulates. Particulate matter is not only a problem for human health but also for equipment downstream of the gasifier. Hot ash and carbon particles can stick to heat exchanger surfaces and cause fouling problems. A syngas quench or cooler needs to ensure that the particles are solidified before removal to prevent fouling. However, even the solidified particles can lead to abrasion in piping and other equipment, wherefore, they have to be removed at the

very beginning of the syngas cleaning process. It may also be pointed out that ash properties have a crucial impact on gasification, especially the initial ash deformation temperature. Fluidized bed gasifiers typically operate safely below that temperature to avoid agglomeration. Entrained flow gasifiers on the other hand require a minimum slag viscosity for their slag removal system.

Fuel Processing. Fuel processing can be divided into three sections: 1) feedstock preparation and handling 2) state-of-the-art syngas cold gas cleanup and 3) advanced syngas warm gas cleanup.

Feedstock Preparation. Depending on the gasification technology, a variety of feedstock processing steps are necessary. In general, all gasification technologies require size reduction of the delivered coal. For slurry fed gasifiers, coal is crushed and/or wet milled to particles ranging from 10 – 65 μm in size and mixed with water [10]. The slurry can be pumped to higher pressure and fed into the gasifier. Challenging for slurry feeding systems is the settlement of particles during storage and transport. In dry feed systems the coal is ground and dried using hot gas. The pulverized coal is fed into the gasifier via lock hoppers utilizing an inert transport gas. Generally, dry feeding systems show a higher efficiency as moisture is removed using low quality waste heat instead of high quality chemical energy inside the gasifier. From an economic point of view dry feeding systems are typically viable up to an operating pressure of 40 bar [11]. Depending on the gasification technology the particle size varies between 70 μm for entrained bed gasifiers up to 50 mm for moving bed and fluidized bed gasifiers [11]. Common concerns of dry feeding systems are bridging in sluicing devices and clogging of conveyors.

Cold Gas Cleanup. The general process flow diagram for state-of-the-art syngas cleanup consists of several processing steps. Depending on the gasification technology, the raw syngas leaving the gasifier can have temperatures up to 1600 °C. Critical for such high syngas temperatures are the particulates that are entrained in the gas. In the temperature range between 900 – 1600 °C, ash may be present as a liquid and carries the potential of slag buildup within process equipment. Thus, it is important to cool the hot syngas to temperatures below 900 °C as soon as the gas leaves the gasifier. Options for high temperature cooling are: radiant cooling (heat is transferred to water cooled walls by radiation), water quenching (injection of liquid water into the hot gas, or passing the hot gas through a pool of water), gas quench (injection/recycle of cooled syngas) and chemical quenching (syngas is cooled while the thermal energy is used to produce other valuable products). Further cooling to about 300 °C is achieved by convective heat exchangers while raising steam. At temperatures below 900 °C, the ash is solid and above 300 °C, NH_4Cl is still in vapor form. This temperature range is typically used to remove ash and other particulates. Cyclones and filters, or any combination of those, are typically used in a temperature range between 500 – 300 °C. Alkali compounds are solidified below 500 °C and, as mentioned before, NH_4Cl is still in vapor form at above

300 °C. When cooling syngas below 300 °C, NH_4Cl has to be removed. A common practice is the installation of a wet scrubbing process using water. If the temperature is sufficiently low, this simultaneously removes water soluble components such as NH_3 , HCN , HCl and HF as well as remaining particulates which act as condensation nuclei during the cooling process. Sometimes, bases like NaOH are added to the scrubber to increase the removal of sour gases such as HCl . If no carbon capture is desired, the next cleaning step involves the hydrolysis of COS to H_2S over activated alumina or titanium oxide catalysts (≈ 200 °C, reheating may be needed). This step is only needed when an amine or SelexolTM process is chosen for sulfur removal. The Rectisol[®] process on the other hand also absorbs COS . If carbon capture is desired, a WGS reactor is required to shift CO to CO_2 . This is usually carried out in a high temperature WGS reactor for bulk shifting, which is favorable due to kinetics, in combination with a low temperature shift reactor which thermodynamically favors the formation of CO_2 . High temperature WGS reactors operate between 300-500 °C versus low temperature shift reactors, which are operated around 200 °C. Consequently, substantial reheating and cooling is necessary. Sour Mo-based WGS catalysts also promote COS hydrolysis which makes the installation of a hydrolysis reactor redundant in scenarios with carbon capture (only for sour shift catalyst). Subsequent to the hydrolysis or shifting, the syngas is cooled to around 40 °C and volatile metals such as Hg and As, are removed in an activated carbon bed. After all these treatments, the syngas is ready for acid gas removal. Acid gas removal is often used as a synonym for desulfurization but technically includes other acid gases like CO_2 , too. Many different technologies have been developed to remove acid gases from syngas: chemical absorption or physical absorption by a liquid, absorption or adsorption by a solid and membrane separation. Depending on the concentration and partial pressure, some technologies are preferential to others. Zinc oxide is suitable for low sulfur content syngas (on a volume basis ml/ml) and can be used in low partial pressure regimes as well as high partial pressure regimes. For higher sulfur content, molecular sieves are the first choice. In the range of medium partial pressures (0.1–5 bar) chemical solvents like AmisolTM, SulfinolTM, MEA and MDEA offer best removal efficiencies. In the regime of high partial pressures, physical solvents work best and outperform chemical solvents. Examples of technologies operating on the principle of physical absorption in liquids are SelexolTM, RectisolTM and PurisolTM. The RectisolTM process can achieve sulfur removal down to 0.1 ppmv [8] but requires cooling to at least -30 °C. SelexolTM and PurisolTM can operate at around 0 – 40 °C while reducing the sulfur (H_2S) content down to 1 ppm [8], which is sufficient for IGCC applications. The acid gas removal unit produces an acid gas stream, which can be utilized to produce elemental sulfur in a Claus plant. Some acid gas removal technologies, e.g., dual stage SelexolTM process, are also capable of capturing CO_2 ; however, this has not yet been commercially realized in IGCC plants.

Warm Gas Cleanup. Warm gas cleanup offers several advantages over cold gas cleanup and has attracted research interest since the 1970s [9]. If the gas cleanup is conducted at higher temperatures, efficiency gains can be expected. Losses due to cooling and reheating are reduced and water does not have to be removed from the syngas stream. This water (vapor) can generate electricity when expanded and additional losses due to water treatment of the condensate can be diminished. In general, it is desired to clean and feed the syngas into the gas turbine at the highest possible temperature in order to improve the efficiency. However, limitations are evident independent from the cleanup technology. In IGCC applications, the temperature limit is set by the highest possible operating temperature of the GT fuel control system. Under consideration of alkali corrosion, metallurgy and refractory-lining this temperature is around 500 – 550 °C [8]. Gas turbine vendor guaranties given are typically for much lower temperature ranges. Many different approaches and technologies for warm gas cleanup have been developed and studied. The next paragraph introduces some of these ideas. A complete summary of this research area would go beyond the scope of this study.

Alkali cleaning in cold gas cleanup systems is achieved by solidification in the gas phase when the syngas is cooled below 550 °C and consecutive separation from the syngas by filters. Alternative cleaning methods involve physisorption and chemisorption on activated bauxite or activated alumina which can operate at temperatures exceeding 800 °C [12], [13]. For the removal of HCl, Na and K, compounds such as nahcolite have been identified as effective reagent with operating temperatures of up to 600 °C [8]. When sprayed into the syngas, solid NaCl is formed, which can be captured in the particle removal section together with other solids. The high temperature removal of mercury remains one of the main challenges and is the Achilles tendon of warm gas cleanup. Mercury can also be captured after combustion; however, this introduces penalties compared to pre-combustion cleanup. At elevated temperatures of around 200 °C mostly activated carbon and metal-based materials like ZnO and Pb are currently being investigated. One of the most promising technologies for warm gas cleanup at temperatures of greater than 260 °C, are palladium-based adsorbents, which are described in [14]. Hot gas desulfurization between 350 – 750 °C can be achieved via metal oxides like ZnO or Zn₂TiO₄. RTI and Eastman are commercializing a high temperature desulfurization process based on ZnO and demonstrated the process on a pilot plant scale in 2005 [15], [16]. In this process, the ZnO reacts with H₂S forming ZnS and water. Later the ZnS is regenerated with air under the formation of SO₂. Many different technologies are currently under development for warm gas CO₂ capture. When looking into high temperature CO₂ removal, one needs to distinguish between temperatures above 300 °C and below 300 °C. If carbon is intended to be captured above 300 °C, the capture process has to be integrated into a process that promotes the WGS reaction. The thermodynamic equilibrium favors CO₂ formation at low temperature. As a result of this equilibrium reaction, the CO₂ conversion at high temperatures is insufficient to achieve high carbon capture rates. Thus

the CO₂ needs to be constantly removed from the reactive phase to prevent it from participating in the equilibrium. In gasifiers, CO₂ can be trapped using CaO, which shifts the equilibrium to the product side [17], [18]. This reduces CO concentrations and simultaneously increases hydrogen concentrations. A similar concept is used in WGS reactors with internal carbon capture, however, operating temperatures will be lower than in gasifiers [19]. Also, high temperature membrane reactors with internal WGS have been proposed [20]. Below 250 °C separate carbon capture technologies can be used. These technologies are more developed and closer to commercialization. Stand-alone carbon capture technologies require CH₄ reforming and CO shifting, which is thermodynamically favored at low temperature, before the syngas enters the CO₂ removal unit. The physical principles for warm gas carbon capture are similar to those for cold gas carbon capture; however, it is much more difficult to find materials with the same properties at elevated temperatures, especially for adsorption-based processes.

2.1.2 Renewable Fuels

Biogas

Fuel Characteristics. Biogas is typically produced at small scales via anaerobic digestion of organic matter, in most cases organic waste. Large scale digester plants suffer from feedstock availability, as organic waste streams are not high in energy content, and it is not economically viable to transport those low energy feedstocks over long distances. Even the long-distance transport of biogas itself is not considered practical to justify large central plants. The main component of biogas is methane with concentrations around 50-75 vol-%. Other than natural gas, there are essentially no higher hydrocarbons present in biogas. However, biogas contains about 19-38 vol-% carbon dioxide, which lowers the heating value of the gas significantly. The composition of biogas can vary drastically between different feedstocks. Most common subdivisions of biogas are: household waste, water treatment waste and agricultural wastes. A summary of typical biogas compositions by category is given in **Table 2-3**.

The biogas production from wastewater treatment plants shows some advantages for power applications over the other biogas sources. The benefits of anaerobic digestion of sewage sludge are widely acknowledged and the technology has successfully been proven and matured over the last decade. Biogas from wastewater treatment plants can help to meet the growing energy demand as well as reduce problems concerning sludge storage. Biogas produced from wastewater has one of the highest methane concentrations, which makes it more valuable and anaerobic digestion can provide a steady fuel stream unlike biogas from landfills (landfill biogas production varies over time). Other constituents like nitrogen and oxygen are typically very low and are introduced to the digester through the feedstock. In order to promote the endothermic digestion process heat is added to the digester which operates at slightly elevated temperatures. Consequently, the biogas is saturated with water vapor. Compared to natural gas, biogas has a significantly

Table 2-3: Typical Biogas Composition by Source [21]

Components	Household Waste	Wastewater Treatment Plants Sludge	Agricultural Wastes	Waste of Agrifood Industry
CH ₄ vol-%	50-60	60-75	60-75	68
CO ₂ vol-%	38-34	33-19	33-19	26
N ₂ vol-%	5-0	1-0	1-0	-
O ₂ vol-%	1-0	< 0.5	< 0.5	-
H ₂ O vol-%	6 (at 40 °C)	6 (at 40 °C)	6 (at 40 °C)	6 (at 40 °C)
Total vol-%	100	100	100	100
H ₂ S mg/m ³	100 - 900	1,000 – 4,000	3,000 – 10,000	400
NH ₃ mg/m ³	-	-	50 - 100	-
Aromatic mg/m ³	0 - 200	-	-	-
Organo-Halogens mg/m ³	100-800	-	-	-

higher content of sulfur species, which is introduced to the digester together with the feedstock. Ammonia is a digestion product but not a concern in high temperature fuel cell applications. However, in combustion applications it is a potential source of NO_x emissions. Depending on the feedstock, the biogas can contain substantial amounts of trace contaminants such as halogens, phosphates and siloxanes. Siloxanes are additives in many health and hygiene products like detergents, medical products, shampoos, cosmetics, paper coatings, and textiles. Large amounts of siloxanes end up in wastewater and the sludge that is being used as biogas feedstock. When anaerobically digesting this siloxane contaminated sludge, siloxanes transition from the sludge into the biogas. If the untreated biogas is combusted directly, siloxanes cause problems in combustion turbines as well as fuel cells as they form a white precipitate.

Fuel Processing. In order to utilize biogas for power applications, a thorough gas cleanup procedure is required. Sulfur compounds do not constitute a problem for traditional GTs, however, sulfur species are a poison for catalysts used in SOFCs and pre-reformers. Due to the high sulfur content, regenerative desulfurization processes might show economic advantages over expendable sorbents in biogas scenarios. Nevertheless, a polishing bed will be required to remove trace amounts that are not removed in the regenerative process.

For SOFC applications O₂ removal is necessary. Similarly, to natural gas, O₂ can be removed via nonregenerative scavengers or catalytic oxidation with fuel compounds. Siloxanes and halogens are typically removed using expendable sorbents.

Due to the high CO₂ content in biogas, compression power and energy conversion efficiency of biogas to electricity are lower than for natural gas, which might favor biogas upgrading (CO₂ removal) prior to utilization. Biogas upgrading describes the process of removing CO₂ prior utilization to boost the heating value of the fuel, which makes it suitable to replace natural gas in some applications. Also, in the context of water recovery from flue gas, biogas upgrading can help to increase the water recovery. Various

technologies for biogas upgrading are available ranging from solvent based systems, water wash systems, pressure swing adsorption systems to membrane systems. For small systems, membrane-based CO₂ separation might be most attractive due to the reduced complexity of such a plant. On the other hand, water wash systems have shown to be very effective in removing CO₂ from biogas, however, concerns about water consumption can be problematic in certain areas.

Biogas is supplied at close to ambient pressure and gas compression is needed. For high pressure ratios, intercooling of the biogas is necessary. Since the biogas is supplied at its dew point, intercooling at elevated pressures will lead to vapor condensation. For any sorbent-based gas cleanup process it is critical to ensure that the biogas is introduced with sufficient superheat. Failure to do so will lead to pore condensation and flooding of the sorbent.

Biomass

Fuel Characteristics. Research over the last decade has shown that biomass is an integral part of CO₂ emission reduction. Biomass is considered as a carbon neutral and renewable energy resource, due to the photosynthetic processes, which convert CO₂ from the atmosphere into organic compounds. These organic compounds are the building blocks of plants and thus, biomass. Biomass can be used in many different ways. Depending on the feedstock, biomass can be fermented, aerobically or anaerobically, in order to produce alcohols or gases (suitable for manure, sewage but also plant waste or food waste). Whereas wooden biomass is often pressed into pellets and burned directly. However, gasification of plant-based biomass is gaining more and more attention because of the versatile applications of the produced syngas and its potential to remove CO₂ from the atmosphere in so-called negative emission plants. Negative emission plants will play a crucial part in decarbonizing the transportation, energy and industrial sectors as they are less disruptive and less expensive than cutting all CO₂ emissions to 0% [22]. Thus, negative emission strategies should be viewed as part of the emission reduction portfolio, rather than a technology to remove anthropogenic CO₂ from the atmosphere. Direct air capture technologies are very expensive and carbon mitigation strategies such as afforestation, agricultural practices and biomass energy systems with carbon capture are significantly cheaper and ready for large scale deployment [22].

For biomass, there is a great variety of different feedstocks with substantial deviations their physical and chemical properties, such as heating value, ashing behavior, moisture content, volatile matter, mineral content and many more. Beneficial for the efficiency of the gasification process and overall process are feedstocks with high heating values and little inert material. For gasification, feedstocks with heating values of less than 7-8 MJ/kg cannot be used without a support fuel [23]. Directly related to the heating value is the moisture content. The higher the moisture content the lower the heating value on a mass basis. Additionally, evaporation of feedstock moisture during utilization consumes energy and higher efficiencies

can be reached if the feedstock is dried beforehand. In general, the moisture content of biomass for gasification should be below 15% [24]. Despite that, gasification requires water/steam for H_2 production and in the case of hydrogasification without the use of O_2 as oxidant, water acts as promoter and increases reaction rates [25].

Another aspect that has to be taken into consideration is the ash content of the biomass, which can vary considerably between different feedstocks. Feedstocks exceeding a mineral matter ash content of 12% show strong slagging behavior, if not operated safely below the ash softening temperature [26]. This can lead to agglomeration and in worst case blocking of the reactor. In general, mineral matter contents of less than 6% are expected to be non-slagging [26]. In the range between 6-12% the slagging behavior mostly depends on the mineral matter composition [26] and its base-acid ratio. The low fusion temperatures of herbaceous biomasses are a result of high Si, Ca and K content, whereas, hard coals predominantly contain Si, Al and Fe [8]. Other problems associated with large amounts of ash are ash cooling (and the associated energy loss), ash transportation and recycling/storage of the ash. In fluidized bed gasifiers, feedstocks with high ash melting points are preferred [9].

Further attention needs to be paid on some trace elements like sulfur, chlorine, nitrogen, phosphorous, sodium and potassium. Mercury concentrations in biomass can vary considerably and need to be taken into consideration as necessary. Sulfur contents in biomass are much smaller compared to coal. However, even small amounts of sulfur compounds in the raw syngas can cause catalyst poisoning. If combusted, sulfur forms SO_2 and SO_3 , which are responsible for the formation of acid rain if emitted into the atmosphere. During gasification, large amounts of the sulfur are converted to H_2S , which causes corrosion problems in vessels and equipment. More critical than sulfur is the chlorine content of biomass, which is typically higher than in coal (especially in straws and grasses). Chlorine is mostly taken up in forms of KCl and NaCl as it is found in soil and fertilizers. Biomass grown in coastal areas usually exhibits a higher NaCl content [8]. Chlorine in the feedstock can cause severe corrosion problems. Corrosion damages are responsible for about 50% of all breakdowns of fossil fueled power plants in the U.S. and account for a large portion of maintenance costs [27]. There are four different aspects of how chlorides affect plant equipment [9]; I) deposition of chlorides in the syngas heat exchanger (melting points 350 – 800 °C) and fouling of the exchanger surface, II) chlorides reacting with H_2 forming HCl, which can harm gasifier and piping, III) in combination with nitrogen rich feedstocks NH_4Cl can be formed, depositing in the economizer when temperatures reach 280 °C or less, and IV) poisoning of catalysts. The addition of CaO and MgO to the feedstock can reduce high temperature corrosion [9]. Nitrogen that is bound in the biomass is converted to HCN or NH_3 under gasification conditions and excess H_2 . HCN mainly originates from aromatics, whereas NH_3 has its source in amine compounds [9]. As mentioned before, NH_3 will form NH_4Cl in the presence of chlorides, which can lead to fouling problems in the economizer. Hydrogen cyanide, which is also corrosive,

will be converted to NH_3 when entering the syngas shift unit (only needed in the case of carbon capture). Phosphorus can become a concern when gasifying biomass [9]. Biomass has significantly higher phosphorus concentrations than coal, which condenses at around 589 °C at 3.6 MPa and could lead to fouling of the heat exchangers. Sodium and potassium levels are less critical for the syngas but have a very strong influence on the ash's behavior. Biomass typically has higher sodium and potassium contents as coal. If gasified at high temperatures sodium can leave the gasifier as NaCl or NaCO_3 , which causes problems when the syngas is cooled below 800 °C and 850 °C respectively, as the salts crystallize in the syngas heat exchange section. Furthermore, alkali metals lower the ash melting temperature and enhance slagging and agglomeration in the reactor. This is especially critical for fluidized bed gasifiers. Furthermore, liquid alkali metal slags are very corrosive and dangerous for the gasifier operation. However, sodium and potassium have been identified as good catalysts for catalytic steam gasification and hydrogasification [25], [28], [29].

Next to physical and chemical feedstock properties, also economic, ecological and social factors play an important role when choosing biomass as feedstock. Biomass cultivation stands in competition with agricultural commodities and is seen as risk for the price development of those. This topic gained importance in 2008, as the international food prices rose drastically due to the high biomass production. This increase in food prices has been heavily linked with the production of 1st generation biomass like corn, wheat, sugarcane, soybean, rapeseed and sunflowers. First generation biomass directly competes with food production as these crops can be used for food production as well as for energy production. Thus, there is a strong movement towards 2nd generation biofuels like woody and herbaceous plants which do not stand in direct competition with food production. However, even if using 2nd generation biomass, the crops are still competing for cultivation areas. In order to minimize land usage, crops with high dry matter yields, high energy yields and the ability to grow on marginal land are preferred. Equally important is the water usage and how much biomass can be produced per amount of water. Economic aspects of biomass production include fertilizer, pesticide and herbicide usage as well as harvesting equipment, biomass drying and the bulk density of the obtained biomass. Third generation biomass is still an active area of research and describes biomass derived from algae. Algae-based biomass possesses great potential for future applications as it can be grown offshore, does not compete with food production, and exhibits exceptional growth rates and CO_2 uptake rates.

Fuel Processing. In general, the same fuel processing steps as for coal also apply to biomass. However, biomass is more difficult to dry and feed into a boiler or gasifier. Since biomass has a significantly higher moisture content, energy efficient drying processes are important. Furthermore, the grinding of biomass is much more difficult than grinding brittle hard coal. Biomass has a viscoelastic and plastic structure, which has the capacity to absorb enormous amount of energy before failure. Consequently, the drying as well as

the milling of biomass requires significantly more energy input than the processing of coal. On the other hand, the higher volatiles content increases the reactivity of the biomass particles. While coal particles need to be reduced to an average particle size of approx. 70 μm , wood particles with a particle size of 1 mm are able to burn at the same rate under the same conditions [30]. The syngas cleanup for biomass will be conducted in a similar manner as for the coal derived syngas; however, the individual process units will have to be adjusted and scaled according to the quantity of pollutants present in the syngas.

Hydrogen

Natural gas-based and coal-based SOFC-GT hybrid systems can be seen as advanced technologies utilizing traditional fuels, which in the short term will help to offset critical carbon emissions due to their exceptional efficiency. At the same time, natural gas and coal provide a robust platform for the development of such power systems and a platform for widespread market adoption in order to make use of the economies of scale, especially on the production side. After that, SOFC-GT hybrid systems can transition to biogas and biomass applications, which represent an important pillar of the circular economy where the utilization of alternative organic fuels creates a climate neutral economy with negative emission plants as part of the carbon mitigation portfolio.

Hydrogen forms another important pillar in a sustainable, renewable, climate neutral and clean circular economy, which will be produced from excess renewable electricity (complementary to biogas and biomass, which use organic waste streams in the circular economy). Hydrogen is not part of this study, but as an important fuel of the future it is briefly discussed here. In order to utilize the produced hydrogen, highly efficient fuel cell systems, such as SOFC-GT hybrid systems, are needed to maximize the roundtrip efficiency and minimize resource investment. While batteries are very efficient to store electricity over short periods, H_2 -based energy storage in large quantities will be needed to economically shift the electricity generation and demand difference between seasons. Hydrogen can be stored in pipelines and underground similar to natural gas and it can be assumed that H_2 can be supplied at different pressures depending on the scale and application. Hydrogen produced from electrolysis possesses a very high purity. Potential impurities are small quantities of water vapor that might have to be removed for transportation applications where proton exchange membrane (PEM) fuel cells are used. Once H_2 production reaches larger scales, odorants might be added to the gas due to safety reasons. Ammonia for example will be unproblematic for SOFCs, however, pose a risk for some other fuel cell technologies.

Since water is needed for electrolysis, water concerns may arise in the future since many areas of the world suffer from water scarcity. Thus, in a hydrogen economy, flue gas water recovery systems will play an even more important role in order to supply the water needed for electrolysis. Beyond this, hydrogen pipelines could complement water infrastructure if the water produced from the oxidation of H_2 is captured.

2.2 Solid Oxide Fuel Cell – Gas Turbine Hybrid Systems in Electric Power Generation

Independent of the fuel processing and integration schemes, the heart of each SOFC-system is the fuel cell. The following chapter will provide some fundamentals on SOFC-GT hybrid systems for stationary electric power generation.

2.2.1 Solid Oxide Fuel Cell

Fuel cells convert fuel-bound chemical energy directly to electrical energy. This is a fundamental difference when comparing fuel cells to thermodynamic energy conversion devices like internal combustion engines or gas turbines. Instead of converting the fuel-bound energy to heat, which is then transformed into mechanical energy and electricity in a generator, fuel cells directly harness the electrons participating in the chemical reaction.

In a fuel cell, the chemical reaction is split into two electrochemical half reactions. A reduction reaction and an oxidation reaction:



When these two reactions are spatially separated and connected to an electrically conducting medium, the electrons can be forced to move through an external circuit and do useful work, before they can complete the reaction. The separation of the two reactions can be achieved via an electrolyte that only allows the oxygen ions to migrate through the electrolyte (impermeable to oxygen atoms or molecules as well as electrons). In SOFCs, the electrolyte material is typically a metal oxide ceramic like yttria stabilized zirconia (YSZ), which has shown to have favorable stability properties and oxygen ion conductivity.

However, satisfactory conduction of oxygen ions is only possible at temperature greater than 600 °C. Due to its high operating temperature, SOFCs are more suitable for stationary electricity generation. In stationary electricity generation the high temperature SOFC can be integrated in the balance-of-plant increasing the efficiency. Furthermore, at such high temperatures, the fuel cell can promote internal methane reforming described in Reaction (R 2-3) and the shifting of water gas, Reaction (R 2-4).



Reforming is an endothermic reaction while the water gas shift (WGS) reaction is mildly exothermic. This way internal reforming can internally cool the fuel cell without the supply of a cooling medium and/or heat exchangers. An illustration of the working principles and design of a solid oxide fuel cell is provided in **Figure 2-1**.

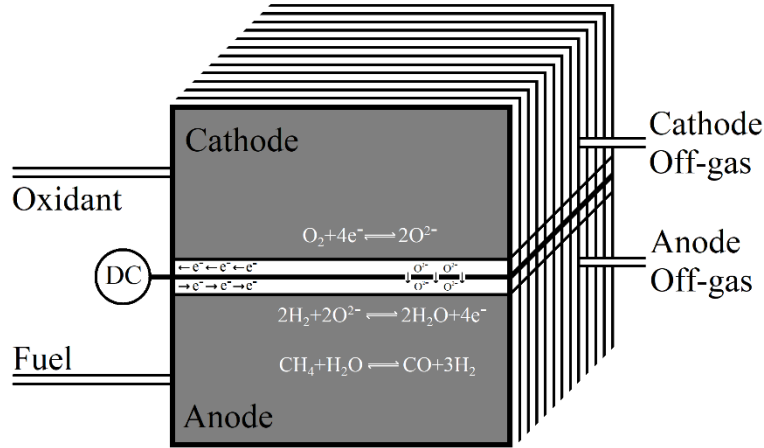


Figure 2-1: Illustration of the operation of a solid oxide fuel cell.

The performance of a fuel cell can be estimated using the Nernst equation, which describes the theoretical cell potential of the two electrochemical half reaction.

$$E_{\text{Nernst}} = \sum E_{\text{Half Reactions}}^0 + \frac{\Delta \hat{s}_{\text{rxn}}}{nF} (T - T_0) + \frac{RT}{nF} \ln \left(\frac{\prod a_{\text{Products}}^{v_i}}{\prod a_{\text{Reactants}}^{v_i}} \right) \quad (\text{E 2-1})$$

E is the cell potential, in V,

$\hat{s}_{\text{rxn}}$ is the reaction entropy, in J/mol/K,

T is the temperature, in K,

R is the universal gas constant, 8.314 J/mol/K,

n is the number of electrons participation in the electrochemical reaction,

F is the Faraday constant, 96,485 C/mol,

a is the activity, and

v_i its corresponding stoichiometric coefficient.

In order to account for irreversibilities, loss terms are subtracted from the theoretical cell voltage. Typically, three different types of losses are accounted for based upon the origin of the loss. These are: activation losses, ohmic losses and concentration losses. Activation losses represent the voltage that must be invested in order to overcome the activation barrier associated with the electrochemical reaction. As all electronics, also fuel cell materials are not perfect conductors and exhibit resistances to the transport of charge. This resistance to charge transport ultimately results in a loss of voltage, known as ohmic losses. In general, the resistance of ionic charge transport through the membrane is by several orders of magnitude larger than the charge transport through the electrodes. The third voltage loss term is concerned with the

gas phase mass transport. As it can be seen from Nernst equation (Equation (E 2-1)), the activity, which is related to the partial pressures in the gas phase, has significant influence on the theoretical voltage. The fuel cell performance is determined by the concentration profile inside the catalyst layer and poor mass transport can lead to a substantial increase in those losses. Optimization of flow pattern and permeability of the catalyst layer can minimize transport losses, however, since the electrochemical reaction is the driver for the concentration gradient, concentration losses cannot be completely eliminated. Details on the estimation of the individual loss terms can be found in [31]. After consideration of the various loss terms the actual operating voltage of the fuel cell is calculated as follows:

$$V_{\text{SOFC}} = E_{\text{Nernst}} - \eta_{\text{act}} - \eta_{\text{ohmic}} - \eta_{\text{conc}} \quad (\text{E 2-2})$$

V_{SOFC} is the operating voltage of the fuel cell, in V,

η_{act} is the activation loss, in V,

η_{ohmic} is the ohmic loss, in V and

η_{conc} is the concentration loss, in V.

Since fuel cells are driven by electrochemical reactions that are promoted by catalysts, side reactions creating criteria pollutants are almost completely suppressed. Harnessing the electrons participating in the reaction directly, enables fuel cells to reach very high efficiencies independent from any thermodynamic cycles, e.g., the Carnot cycle. Driving force for the electrochemical reaction is the chemical potential difference, which allows small scale fuel cells to achieve the same performance characteristics as their large-scale counter parts. Because of this required chemical potential difference, fuel cells cannot utilize 100% of the supplied fuel. In the case of 100% fuel utilization the chemical potential difference drops to zero and so does the electrical voltage and the power output. In order to avoid wasting fuel resources, this fuel can be separated and recycled back to the fuel cell or can be utilized downstream of the fuel cell in another energy conversion device, e.g., a gas turbine.

2.2.2 Gas Turbine Hybrid

Historically, GTs have been developed for natural gas and liquid hydrocarbon fuels but have been advanced to combust syngas and even hydrogen. Gas turbines operate on a Brayton-cycle with air as working fluid. The simple cycle consists of: I) isentropic compression of the gas (air) in the compressor, II) isobaric heat addition in the turbine combustor, III) isentropic expansion in the turbine, IV) isobaric heat rejection of the turbine exhaust. Step IV) is typically disregarded and the Brayton cycle is operated as an open cycle, meaning that constantly fresh air is supplied to the engine. However, sometimes, e.g., in combined cycles, the gas turbine exhaust is cooled for purposes of heat recovery.

In large GTs, compression is conducted in an axial compressor, which is mounted on the same shaft as the expander. Thus, no motor is needed to drive the compressor and losses are minimized. The compressed air is then heated in the combustor by injecting fuel into the combustion chamber, which immediately reacts with O_2 to release energy in the form of heat. The high-pressure, high-temperature gas is then expanded in a turbine expander. For aero-derivative engines this turbine is divided into two parts. The first expander provides the power required for compression. The second expansion step drives an electricity generator. In aero-engines the second expander is not present and the remaining energy in the gas stream is used to generate thrust. In GTs designed for power generation, the compressor, expander and generator are typically mounted on a single shaft. A general flow diagram of a GT is shown in **Figure 2-2**.

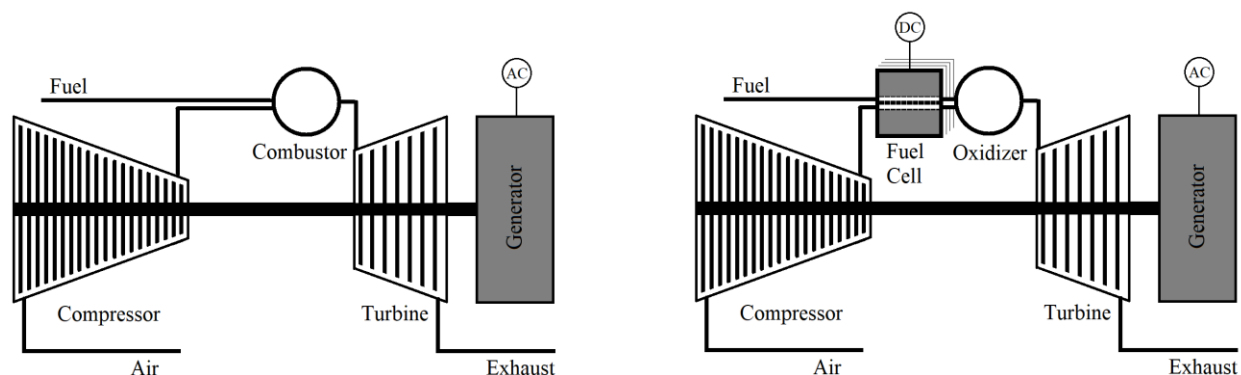


Figure 2-2: Illustration of a gas turbine (left) and a gas turbine-fuel cell hybrid system (right).

The compression work in a Brayton-cycle is significantly larger than in a Rankine-cycle due to the high specific volume of gases. In large machines compression consumes about 50% of the generated power. Thus, it can make sense to use intercoolers to reduce the workload and increase the efficiency. However, intercooling lowers the combustor outlet temperature, which has to be compensated with a higher fuel input.

Another important factor influencing the GT efficiency is the turbine inlet temperature (TIT); the higher the TIT the higher the gas turbine efficiency, limited only by material constraints. If, for example, CH_4 is combusted with air in a stoichiometric mixture, temperatures around $1940^\circ C$ are possible [32]. In order to reduce the TIT to a temperature in the range $1300 - 1400^\circ C$ (state-of-the-art TIT), excess air is used. State-of-the-art TITs are only achievable due to the usage of advanced materials and internal blade cooling via air extracted from the compressor or even steam.

To utilize the unused fuel from a fuel cell, the anode off-gas can be compressed and fed into the gas turbine combustor or the fuel cell can be directly integrated into the gas turbine forming a fuel cell-GT hybrid as shown in **Figure 2-2**. The integration of the fuel cell into the GT has many advantages, especially high temperature fuel cells like SOFCs exhibit synergistic effects in such a configuration.

In a SOFC-GT hybrid, fuel and oxidant are supplied to the SOFC at elevated pressure. From Equation (E 2-1), it becomes apparent that a higher supply pressure of reactants boosts the efficiency of the fuel cell.

The waste heat generated during gas compression inside the gas turbine compressor is used to partly pre-heat the oxidant stream for the SOFC and the waste heat produced during the electrochemical reaction inside the SOFC is used to further upgrade the heat quality of the fuel cell exhaust gasses before they enter the off-gas oxidizer. In the oxidizer the hot fuel cell exhaust gases from anode and cathode are mixed and the unutilized fuel from the SOFC, which then undergoes complete oxidation, further increases the combustor outlet gas temperature to increase the efficiency of the turbine expander. The hot GT exhaust gasses can be utilized for fuel and oxidant preheating and/or running a Rankine cycle.

Challenging for the operation of such a hybrid is the low heating value of the fuel gas entering the GT oxidizer. The heating value is significantly lower than the heating value of natural gas what gas turbines are typically designed for. The higher mass flow through the turbine generates more power, which may be beyond the shaft limitations, and will increase the pressure ratio in the compressor. This will result in a higher turbine inlet or compressor outlet pressure, which can lead to compressor surge [8], [32]. Countermeasures to prevent compressor surge are; increasing turbine cross-sectional area, addition of a compression stage, closing compressor guide vanes or bleeding off air from the compressor. The latter is an attractive solution for integrated gasification combined cycles with air separation unit (ASU) because it does not require expensive modifications of the GT and the compressed air from the GT can be used in the ASU where high pressure air is needed. Closing the inlet guide vanes is the most popular option to reduce the mass flow. Alternative concepts for syngas or even hydrogen combustion are Brayton reheat cycles as described in [32].

These challenges become more pronounced when operating a SOFC-GT hybrid on coal derived syngas. Syngas, which is a mixture of mostly CO, CO₂ and H₂, has very different combustion characteristics than natural gas. Hydrogen is very reactive and increases the flame diffusion speed. A consequence of this property is a high risk of flash back when using dry low-NO_x burners. Hence, conventional combustors have to be used and NO_x control is accomplished by lowering the flame temperature by diluent injection [8]. NO_x control is of special importance in syngas applications since syngas combustion has a lower air requirement which increases the adiabatic flame temperature.

2.3 Flue Gas Water Recovery

The water present in the hot flue gas is in its vapor form. Origin of the water vapor can be traced back to three different sources. Some moisture is introduced into the system together with the fuel feed stream in the form of free moisture or inherent moisture. Secondly, moisture might be present in the oxidant or air stream; however, typically this amount is small compared to other sources. Another portion of the flue gas's water vapor originates from the oxidation of fuel bound hydrogen. In the case of hydrocarbons, CO₂ and H₂O will be formed via Reaction (R 2-5). Molecular H₂ present in the fuel may directly form water via

Flue Gas Water Recovery

Reaction (R 2-6). Similarly, water is formed when oxidizing alternative fuels like alcohols or ammonia. The same overall reactions also hold up for fuel cells, although, the reactions rather occur in multiple partial reactions of chemical and electrochemical nature.



The water content in natural gas and biogas are relatively small when compared to coal or biomass, which can have moisture contents of greater than 50 wt-% at harvest. For slurry fed gasifiers fuel processing can also add a substantial amount of water to the feedstock. Another source of hydrogen in gasification plants is the addition of steam to the gasifier which will produce H_2 and CH_4 during the gasification process.

Flue gases from low-firing-temperature GTs (micro gas turbine) exhibit a moisture content of around 4 vol-%, while higher firing temperature GTs (H-class gas turbine) see values around 10 vol-% (without carbon capture). In IGCC applications without carbon capture, where the GT is fueled with CO_2 -rich syngas, the water content of the flue gas can be as low as 4 vol-% [33], [34]. The moisture content of the flue gas in IGCCs is strongly dependent on the gas cleanup. In traditional state-of-the-art cold gas cleanup scenarios, most of the water is dropped out before it reaches the GT reducing the moisture content of the exhaust gas. However, with advanced warm gas cleanup, which shows enormous efficiency benefits, this will change and significantly increase the moisture content in the flue gas. In IGCC scenarios with carbon capture the moisture content in the flue gas typically varies between 12-21 vol-% [7], [33]–[39]. Thus, carbon capture and water recovery are complimenting each other, which can have synergistic effects on the overall plant performance. Flue gas from pulverized coal power plants typically have a moisture content around 15 vol-% [34]. In SOFC-GT hybrids, the SOFC dictates the fuel to air ratio, thus, the moisture content is a function of the air utilization of the SOFC.

Various strategies for removing moisture from a stream of non-condensable gases exist. In order to separate the vapor from the other gas constituents a driving force that drives the separation needs to be introduced. One option is to cool the gas stream below its dew point temperature, which forces the vapor out of the gas phase. The cold surface or condensed liquid exhibits an equilibrium vapor pressure that is lower than the bulk gas's vapor partial pressure. This difference acts as driving force for mass transfer between the heat exchanger surface area and the bulk gas stream. The lower the vapor pressure, which is a function of temperature, the larger the gradient and the resulting mass flux. The vapor pressure of the liquid water represents the thermodynamic limit of how much water can be removed from the gas stream. To further improve the water recovery, desiccants like salt solutions with a very low vapor pressure are commonly utilized. Desiccants allow water recovery beyond the gas's dew point, which can be an

advantage in areas where low temperature cooling is impractical due to high ambient temperatures and/or insufficient cooling water supply. However, desiccant need to be regenerated under the input of energy in order to release the captured water and being able to reuse it. Another approach to separate the water vapor from the flue gas is via the use of membranes. The working principle for membranes is similar and a driving force is needed to transport the water molecules across the membrane. Membranes can work in the condensing regime where the liquid's vapor pressure represents the driving force or the gas permeation regime where the partial pressure of the water vapor on the permeate represents the driving force. Thus, in the case of gas permeation, a sufficiently high vacuum needs to be applied on the permeate side if the flue gas is close to ambient pressure.

Critical for the operation of a flue gas water recovery unit downstream of a GT is the pressure drop. High GT backpressures can be problematic for the operation of the GT. Additionally, a higher GT back pressure reduces the pressure ratio across the turbine expander and consequently the power output. Pressure drop is less important if the flue gas water recovery unit is installed at a boiler plant. Furthermore, pollutants present in the flue gas and water need to be considered. Condensation of water in the flue gas might lead to gas absorption and formation of undesirable byproducts such as sulfuric acid and carbonic acid. While this might have positive aspects with respect to the stack emissions, these corrosive and potentially toxic liquids can be harmful for equipment, environment and human health. Desiccant-based and membrane-based water recovery systems might be able to mitigate some of these negative effects.

2.4 Solid Oxide Fuel Cell Technology Status and Literature Review

2.4.1 History of Solid Oxide Fuel Cells

The idea of generating power from inverse water electrolysis was introduced by Sir William Grove in 1838 [40] who discovered power generation from a cell of iron and porcelain in a solution of sulphate of copper and dilute acid with a device he called “gas battery”. Even more than half a century later, the working principle of this device was still not understood, as the science community was struggling to understand basic principles of matter, energy and electricity. Wilhelm Ostwald, the founder of modern physical

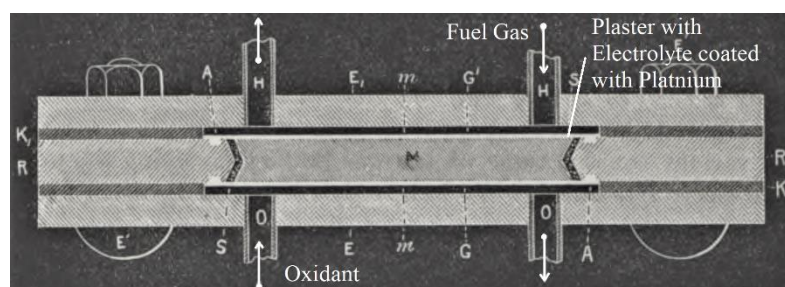


Figure 2-3: Illustration of the fuel cell developed by Langer and Mond [42].

chemistry, called it “a puzzle”. Although it caught his theoretical interest, he described its practical use as irrelevant [41]. Nevertheless, in 1896 Charles Langer and Ludwig Mond – who later introduced the term “fuel cell” – planned to commercialize

a fuel cell device and developed the first practical fuel cell using coal gas in 1889. Their fuel cell was able to generate up to 35 A/m^2 at 0.73 V [42]. However, due to severe problems with the stability of their fuel cell and the invention of the gasoline engine, they were not able to introduce their fuel cell into the market. Research around Charles Alder Wright [43] and Louis Paul Cailletet [44] resulted in the development of similar cells as the one of Mond and Langer and encounter comparable problems with cell stability and leakage of gases. Especially the substantial amount of noble metals needed to operate the fuel cell at a considerable rate seemed to be an inhibiting factor on the way to commercialization.

In the first half of the twentieth century the German electro-chemist Emil Baur, who started researching fuel cells at the Technische Universität München and his later work at the Eidgenössische Technische Hochschule Zürich, set the ground work for high temperature fuel cells like molten carbonate and solid oxide fuel cells [45]–[47]. It was not until the 1950's that the fuel cell technology experienced a renaissance: Bacon [48] demonstrated a practical 6 kW system using a less corrosive alkaline electrolyte and less expensive nickel electrodes, Ihrig [49] at Allis-Chalmers Manufacturing Company, presented a 20 hp tractor using a 1,008 cell alkaline fuel cell operating at low-pressure and low-temperature, and NASA employed fuel cells in their Apollo missions to the moon. Onboard the spaceship fuel cells quietly supplied not only the needed electricity but also drinking water for



Figure 2-5: Allis-Chalmers' 15 kW fuel cell tractor [49].

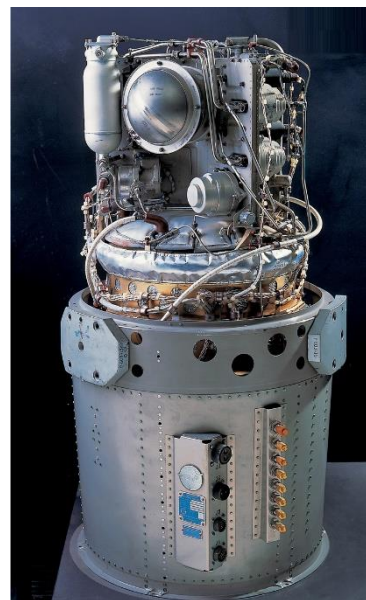


Figure 2-4: Fuel cell system from the Apollo missions to the moon [50].

the crew without creating any pollutants [50]. However, on the first flights these modules did not contain diaphragms to filter the water before use and small bubbles of hydrogen were dissolved in the produced water leading to gastronomic discomfort of the astronauts.

In the last couple of decades, fuel cells attracted more and more attention from car makers and energy utilities as they constitute a highly efficient and environmentally friendly energy

conversion device for the future. Before large scale central power plants can utilize fuel cells, more research will be needed in order to improve availability, efficiency and price-performance ratio. Since the first mentioning of hydrogen fuel cells, many different types of fuel cells have been developed: proton exchange membrane fuel cells (PEMFC), phosphoric acid fuel cells (PAFC), solid acid fuel cells, alkaline fuel cells (AFC), molten carbonate fuel cells (MCFC) and solid oxide fuel cells (SOFC). SOFCs are of interest when considering fuel cells for central power generation due to their high efficiency, relatively low cost and reliability. However, MCFCs gained more significance in the last couple of years as they can operate highly efficient and achieve higher power densities at lower temperatures than SOFCs. SOFCs typically operate between 650-1000 °C, while MCFCs operate around 500-650 °C. Lower operating temperatures are desired as this shortens the ramping time and allows the usage of less expensive materials. However, higher temperatures are desirable in SOFC-GT hybrid applications. When considering fuel cells for the direct system integration into a GT hybrid under pressurized conditions, PEMs, PAFCs, AFCs and MCFCs face challenges due to their liquid electrolyte and/or low operating temperature. Temperatures below 500 °C become unattractive for stationary power applications as elevated temperatures increase fuel processing flexibility and enable utilization of carbon-based fuels, e.g., syngas from coal and biomass, natural gas, biogas or even pure carbon monoxide (carbon monoxide is a poison in low temperature fuel cells but not for SOFCs and MCFCs). Furthermore, elevated temperatures allow a better system integration, enable internal reforming inside the fuel cell and support co-production of heat, cooling or fuels. A summary of the various fuel cell technologies is provided in **Table 2-4**. With the steady goal of improving efficiencies, SOFC-GT hybrids first entered the stage in the year 2000. With the successful concept demonstration in the early 2000s, the interest in SOFC-GT hybrids in the research community continues to grow due to their exceptional performance characteristics.

Table 2-4: Comparison of Major Fuel Cell Types [31]

	PEMFC	PAFC	AFC	MCFC	SOFC
Electrolyte	Polymer Membrane	Liquid H_3PO_4	Liquid KOH	Molten Carbonate	Ceramic Membrane
Charge Carrier	H^+	H^+	OH^-	CO_3^{2-}	O^{2-}
Operating Temperature	80 °C	200 °C	60-220 °C	500-650 °C	650-1000 °C
Catalyst	Platinum	Platinum	Platinum	Nickel	Perovskites
Cell Components	Carbon-based	Carbon-based	Carbon-based	Stainless-based	Ceramic-based
Fuel Compatibility	H_2 , MeOH	H_2	H_2	H_2 , CH_4	H_2 , CH_4 , CO

Solid Oxide Fuel Cell Technology Status and Literature Review

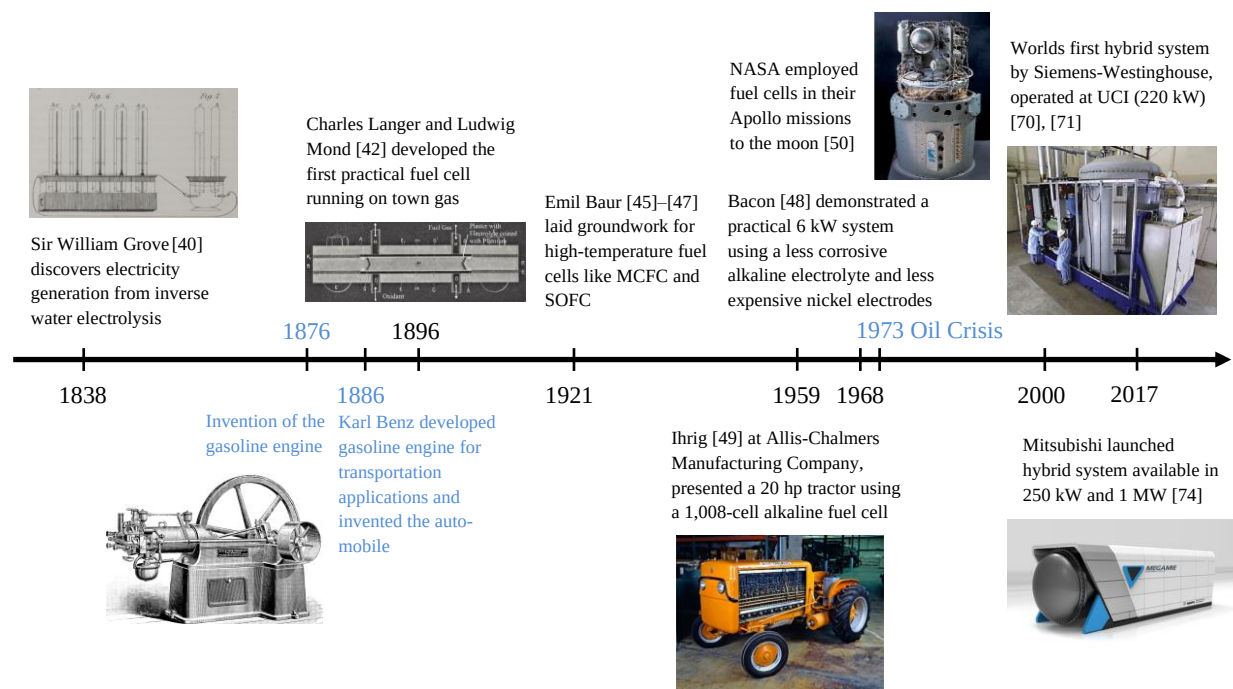


Figure 2-6: Fuel Cell History

2.4.2 Technology Status of Solid Oxide Fuel Cells and Hybrid Systems

Since the great advancement in fuel cell technology over the last 50 years, fuel cells have been considered for integration into large-scale power systems. Today numerous power plants based on fuel cell technology have been installed. At the end of 2016, the U.S. had 56 large-scale fuel cell power plants (larger than 1 MW) generating a total of 137 MW, whereby, California had a share of 36% of the total U.S. fuel cell capacity. Fuel cells collectively provided 810,000 MWh of electricity in 2016, representing 0.02% of total U.S. electricity market [51]. The largest plant is currently operating in South Korea. The Gyeonggi Green Energy facility in Hwasung City has a capacity of 59 MW and consists of 21 2.8-MW hydrogen fuel cells supplied by FuelCell Energy operating at an efficiency of greater than 50%-LHV [52]. This record will not last for very long as the next mega project is already on the way. Korea Western Power Co., Hanp Inc., KB Securities Co., Chungcheongbuk-do provincial government, and the Jincheon-gun office signed an agreement for the deployment of a 80 MW fuel cell power plant to be built in Jincheon in South Korea's North Chungcheong Province on 6th July 2018 [53].

Most of the above-mentioned plants operate on MCFCs, however, the highest efficiencies can be reached in hybrid systems integrating the fuel cell into a Brayton cycle. While fuel cells in combination with Rankine cycle can provide better performance than standalone fuel cell systems, it has been shown that in general fuel cell-GT hybrid systems outperform combinations of fuel cells with Rankine cycles. The advantage of a fuel cell-GT hybrid cycle is that the fuel cell efficiency increases as the operating pressure

risers and the waste heat from the fuel cell together with its unused fuel from the anode off-gas can be utilized in the Brayton cycle. The pressurization of fuel cells is challenging. Although SOFCs currently have some disadvantages versus MCFCs when operating at ambient pressures, SOFCs are believed to be more suitable for elevated pressure operation than MCFCs. At elevated pressures MCFCs experience intensified cathode corrosion leading to cathode dissolution as well as sealant and interconnection problems. In the context of the water energy nexus; a combination of fuel cells with Brayton cycle is advantageous due to the elimination of the substantial cooling requirement of the Rankine cycle and the higher efficiency of the SOFC-hybridized Brayton cycle.

Since the 1950s companies like GE are conducting research in SOFC technology with the goal of commercialization. However, problems with melting of the cell material and short circuits, among many other challenges, prohibited a quick market introduction. Today about three dozen SOFC developers exist worldwide. In North America Atrex Energy, Bloom Energy, FuelCell Energy (mostly known for its MCFC technology) and LG Fuel Cell Systems belong to the biggest players in SOFC technology. In Europe there are: Bosch Thermotechnology, Convion, Haldor Topsøe AS, SOLIDpower to only list a few, and Aisin Seiki, h2e Power Systems Inc., Mitsubishi-Hitachi Heavy Industries and POSCO Energy are some of the major SOFC companies in Asia. Current research is seeking to reduce cost and develop small scale units for residential and utility applications. Challenges faced during the commercialization process are mostly related to reliability, interconnects, seals and lowering the operating temperature to improve startup time. DoE's Solid-State Energy Conversion Alliance (SECA) program targets a SOFC stack cost of less than 175 \$/kW to pursue large scale power generation in the 100 MW range and life time performance degradation of less than 0.2% per 1000 hours of operation while achieving a 40,000 hour lifespan [54].

Today there are several SOFC systems available on the market. Atrex uses a tubular fuel cell design and offers its ARP remote power generator with 500 W, 1000 W and 1500 W running on propane or natural gas with a fuel efficiency of up to 35% [55]. Bloom Energy offers its Bloom Energy Server 5 in three SOFC system configurations: 200 kW, 250 kW and 300 kW. Each system reaches an efficiency of greater 60%-LHV and is designed to cover baseload [56]. Delphi commercialized a 5 kW auxiliary power unit for stationary and transportation application that can operate on pump gas, diesel, bio-diesel, military fuel, propane, natural gas and syngas [57]. Protonex developed a robust 200W SOFC system for powering remote sensors, signaling, and communication systems [58]. In 2018 Protonex' SOFC business has spun off and is now part of Upstart Power. Ultra Electronics has SOFC systems for military, civilian and industrial sector in its portfolio. The Performer P250i is available in two variants: 250 W and 500 W and is designed for backup power applications. The portable Defender D350 is a light weight portable 350 W power supply for expeditionary missions using microtubular technology and propane as fuel. It stands out with its power density of 1 kWh per pound which is ten times the energy density of advanced batteries.

The Defender D350XR is an upgraded version of the D350 and is built for long endurance missions, e.g., unmanned aerial vehicles and unmanned ground vehicles [59]. Adelan offers solutions for combined heat and power where up to 250 W of electricity and up to 1 kW_t of heat are needed. Adelan's SOFC system can run on liquid propane, liquid butane, liquid natural gas, methane and natural gas. The system is able to start-up in a time of 20 minutes [60]. Convion's C50 SOFC products are designed for electrical power outputs in range of 50 kW to 300 kW, while reaching efficiencies of greater than 53%-LHV. In combined heat and power operation efficiencies of greater than 80% are possible [61]. Hexis' combined heat and power unit Galileo (1 kW_e, 1.8 kW_t, efficiency 35%-LHV) was introduced into the market in 2013 but is not part of the portfolio anymore. Meanwhile Viessmann took over Hexis, working on the next generation of the Galileo combined heat and power (CHP) unit [62]. New Enerday offers three auxiliary power supply solution with SOFC technology ranging from 250 W to 1000 W operating on liquid hydrocarbons with efficiencies of 30 – 35%-LHV [63]. BlueGen is a compact CHP unit produced by SOLIDpower. The unit is designed for residential and industrial applications providing 1.5 kW and 0.6 kW heat at an electrical efficiency of up to 60%-LHV [64]. Sunfire's CHP for residential and commercial buildings is rated for 20 kW and achieves efficiencies of greater 50%-LHV [65]. ZEGpower is targeting the industrial scale SOFC utility market utilizing hydrocarbon and renewable fuels. ZEGpower is focused on the upscaling of SOFC technology with carbon capture and hydrogen coproduction. Largest project so far was a 50 kW (30 kW_{H₂}, 20 kW_e) biogas plant [66]. Aisin Seiko also offers a 700 W heat and power cogeneration system for residential use with an electrical efficiency of 52%-LHV and 35% thermal efficiency [67]. h2e offers 6 variants of its HCP-battery hybrid system ranging from 250 W to 9.9 kW with electrical efficiencies ranging from 43%-LHV to 49%-LHV [68]. Huatsing Jingkun New Energy Technology Company offers a SOFC CHP, HS-201, with optional carbon capture. The electrical efficiency of the 1 kW system is 50%-LHV and 85%-LHV combined [69]. The variety of SOFC system keeps increasing as many companies are currently in pre-commercial testing phase, e.g., Bosch and Ceres Power.

All the above-mentioned SOFC systems operate close to ambient pressure. Sealant challenges currently limit the operation of SOFCs under pressurized conditions. Nonetheless, Siemens-Westinghouse commercialized the first SOFC-GT hybrid system based on tubular cells, which was delivered in 2000 and operated at the National Fuel Cell Research Center at the University of California, Irvine. A simplified flowsheet of the SOFC-GT hybrid cycle is depicted in **Figure 2-7**.

Air is compressed in a micro-gas turbine and pre-heated to 550 °C against the hot turbine exhaust stream before it is fed into the solid oxide fuel cell. The tubular design of the SOFC allows the air to further heat up before it gets in contact with the cathode of the SOFC. The fuel, natural gas, is compressed and desulfurized before it enters the SOFC-pressure vessel. Inside the pressure vessel the fuel mixes with a hot

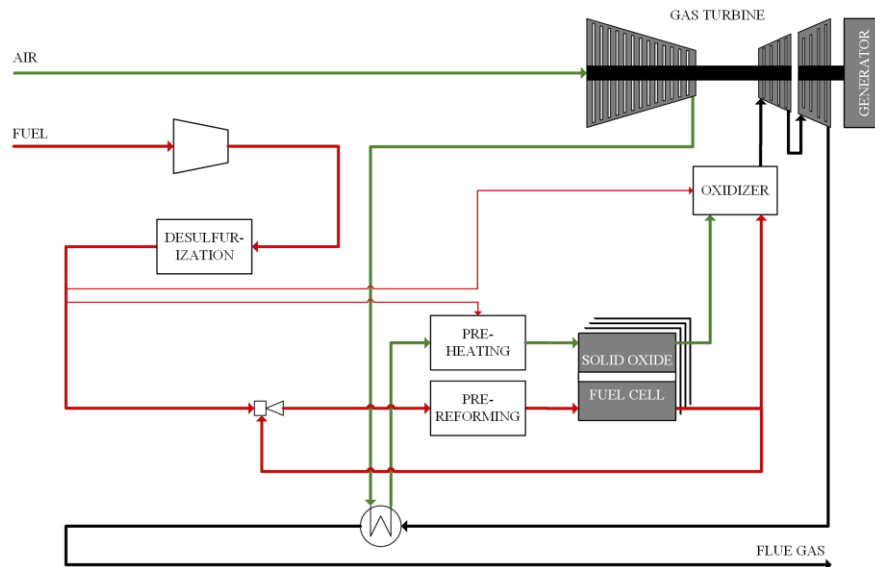


Figure 2-7: Design of the commercial Siemens Westinghouse SOFC-GT hybrid system. Adapted from [70], [71].

recycle stream and is pre-reformed at 550 °C before it contacts the internal reformer and the fuel cell tubes. Heat for the internal reformer is provided by the exothermic electrochemical reaction. At the outlet of the fuel cell, where anode off-gas and cathode off-gas meet, is an oxidation zone converting the unutilized fuel to heat. Additional fuel is used to for air preheating and in the oxidizer. It is not clear if this fuel diversion is only used during startup and part load operation or whether this is also used during full load operating at steady state conditions. With its operating pressure of 3.5 bar the hybrid reaches an efficiencies of 53%-LHV. Siemens Westinghouse offered the system in 220 kW and 320 kW versions [70], [71].

The second company to offer a commercial SOFC-GT hybrid system is Mitsubishi-Hitachi Heavy Industries (MHI) and is the only company that is currently offering a SOFC-GT hybrid system. The system, which was introduced into the market in September 2017, has now received the first order (February 2018) and is expected to start operation by February 2019. The system will be installed at the Marunouchi Building, located near Tokyo Station, and owned and operated by Mitsubishi Estate Co Ltd. This project is being subsidized by the Ministry of Economy, Trade and Industry (METI), which is supporting the introduction of Ene-Farm (residential) and other systems to promote the widespread use of fuel cells for commercial and industrial use [72]. A second order was placed by Hazama Ando Corporation in December 2018 and will be delivered by the middle of 2019 [73]. Mitsubishi's SOFC-GT hybrid system, Megamie, is available as 250 kW system and 1 MW system. The design of the MHI SOFC-GT hybrid is shown in **Figure 2-8**.

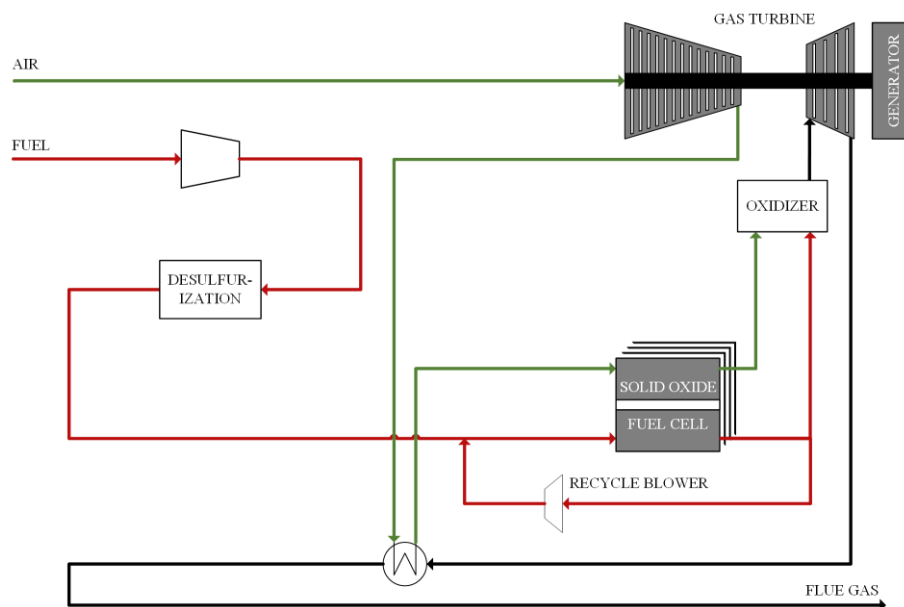


Figure 2-8: Design of the commercial Mitsubishi-Hitachi Heavy Industries SOFC-GT hybrid system. Adapted from [74].

The design is similar to the Siemens Westinghouse system, however, the MHI design uses a recycle blower instead of an eductor to recycle the anode off-gas. Information regarding pre-reforming and internal reforming were not available and are not shown in the flowsheet. Also, the MHI SOFC consists of tubular fuel cells consisting of a tube with several compartments like a stack. The tubes are bundled to a cartridge and arranged in a pressure vessel. During demonstration the system has shown to reach efficiencies of 55%-LHV [74]. Beyond the SOFC-GT hybrid cycle, MHI is demonstrating the principle of triple cycles on a 220 kW system consisting of a SOFC-GT hybrid system with a bottoming Rankine cycle [75].

LG, former Rolls Royce, commenced the development of a SOFC-GT hybrid in 2008 for a 1 MW unit scalable in the range from 250 kW to 20 MW with the ultimate goal to obtain know-how for the integration of SOFCs into large scale integrated gasification plants. Similar to the design of Siemens-Westinghouse and Mitsubishi-Hitachi Heavy Industries, LG uses tubular fuel cells. The design of the SOFC-GT hybrid that is currently being developed is illustrated in **Figure 2-9**.

The system comprises an anode off-gas recycle loop similar to the Siemens Westinghouse design in order to enable pre-reforming. Additionally, a cathode-recycle to minimize component size and cost, and combustor exhaust gas recycle for heat integration are employed. By June 2017 about 1500 hours of operation on a 200 kW scale system were completed [76].

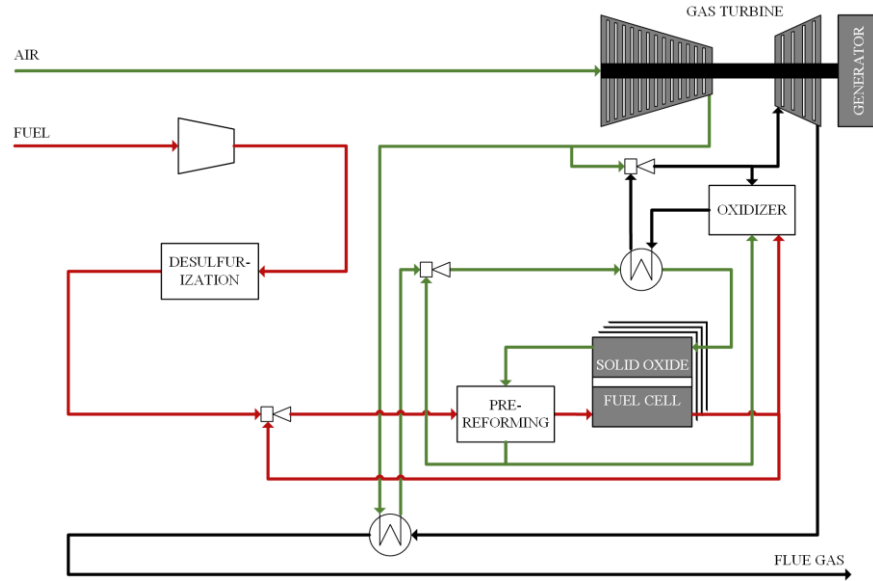


Figure 2-9: Design of the for commercialization developed LG SOFC-GT hybrid system. Adapted from [76].

GE is researching SOFC-GT hybrid systems using planar SOFC cells. Expected benefits from using planar SOFCs are higher power densities and cost reduction. GE designs their SOFC-GT hybrid without any recycle loops. In order to avoid carbon deposition steam is mixed with the desulfurized natural gas before it enters the pre-reformer. Heat integration without recycle loops becomes more challenging and the combustor outlet has to be used to heat the pre-reformer and inlet air [77]. A flowsheet of the GE SOFC-GT hybrid system is shown in **Figure 2-10**.

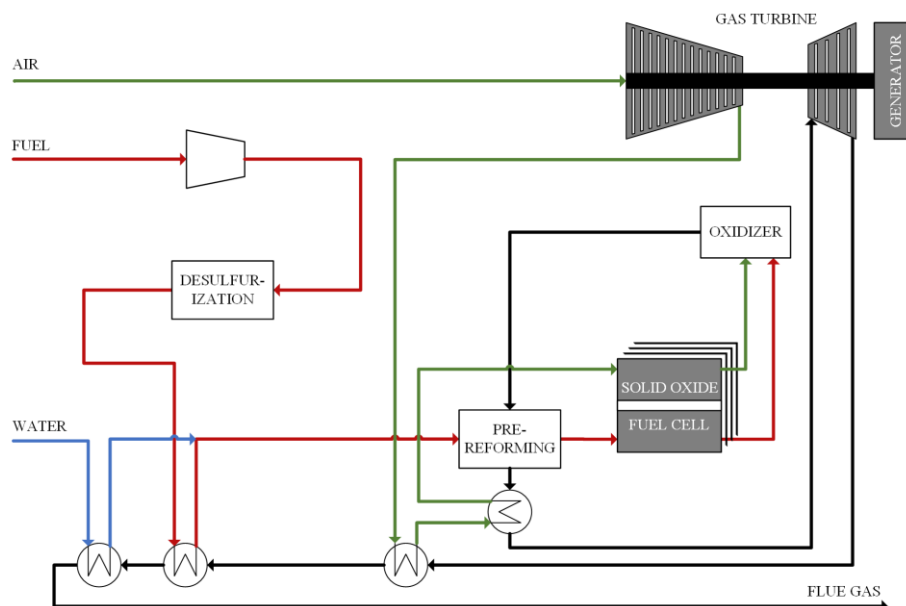


Figure 2-10: Design of the for commercialization developed GE SOFC-GT hybrid system. Adapted from [77].

2.4.3 Literature Review on Solid Oxide Fuel Cell – Gas Turbine Hybrid Systems

Current efficiencies of commercial SOFC-GT hybrid systems are comparatively low as technical and operational know-how of real systems are limited. Over time with improved materials and more experience, more advanced systems, e.g., with higher pressure ratio or higher operating voltage, as proposed in literature will become achievable resulting in higher system efficiencies.

Natural Gas without Carbon Capture. Over the last decades the interest in natural gas-based SOFC-GT hybrid power systems rose due to the high theoretical efficiencies and the availability of this relatively clean fuel. One of the first investigators to analyze such a SOFC-GT hybrid system was Ide et al. [78]. Ide et al. investigated three fuel cell-GT hybrid systems with natural gas as feedstock and pre-reforming. He compared the performances of a PAFC, MCFC and SOFC system at a fuel cell pressure of 8.1 bar. His study concluded that the MCFC system showed the highest efficiency with 51%-LHV but with advancements in fuel cell technology, SOFCs will become the first choice for hybrid systems, also because of their higher operating temperature, which is favorable for the integration into a GT. Ide's SOFC-GT hybrid recovered water from the anode off-gas when the gas is cooled in order to compress it and recycle it back to the reformer. However, quantity and economic value of the recovered water is not considered. Harvey and Richter [79], [80] studied a SOFC-GT hybrid system with flue gas recycling as oxygen carrier to reduce irreversibilities associated with the combustion process in the GT combustor. They were one of the first ones to recover water from flue gas, however, it was mostly due to their specific plant design. In order to re-compress the flue gas, it is necessary to decrease the exhaust stream's temperature, whereby the cooling leads to water condensation. Their natural gas-based SOFC-GT hybrid systems; I) with external reforming and II) with pre-reforming and internal reforming, used a water-cooled condenser to recover water from the flue gas. The condenser uses cooling water at 30 °C and a temperature approach of 10 °C. The recovered water is used to satisfy the internal demand and for water export. The water recovery efficiency of their design is 62.5%. However, the authors remark that under most circumstances the economic meaningfulness of the condenser is not given.

A simple SOFC-GT hybrid system with steam addition to the natural gas and without anode off-gas recycling was studied by Chan et al. [81], [82] focusing on the effects of pressure variation and fuel flow rate. In [83] Calise et al. performed an exergy analysis of a simple SOFC hybrid system without anode gas recycling and identified the SOFC as the largest source of exergy destruction despite its high efficiency. Similarly, Haseli et al. [84] concluded that exergy destruction mostly originates from the SOFC and combustor, whereby, it is desirable to convert most of the fuel in the SOFC even if that leads to a lower firing temperature and turbine efficiency. Also, in the work of Eisavi et al. [85] exergy destruction was found to occur predominantly in the after burner and SOFC. Additionally, different stack configurations

were investigated and a serial air flow in combination with parallel fuel injection was found to be advantageous despite the higher pressure drop. Calise et al. [86] investigated a similar plant design with anode off-gas recirculation, replacing steam addition, for part load behavior under constant speed gas turbine operation and suggest that for part load operation the fuel to air ratio should be kept constant to obtain higher efficiencies. Uechi et al. [87] conducted research on an equivalent system with external reformer coupled with the SOFC looking at turbine inlet temperature, pressure ratio, ambient temperature and steam to carbon ratio. The same system has been studied by Kimijina and Kagsagi [88] for part load conditions which revealed advantages of variable speed control versus constant spool speed operation. Also Yang et al. [89] studied hybrid cycles, similar to the ones aforementioned, looking at fuel cell operating temperatures for systems with internal or external reformer. An optimization study conducted by Shirazi et al. [90] optimized a SOFC-GT hybrid system with GT exhaust recuperation and anode off-gas recycle. The exergy minimization optimization revealed that efficiencies of 63.4%-LHV are possible while the lowest cost was achieved at a pressure ratio of 7.1, a fuel utilization factor of 0.79 and an operating voltage of 0.628 V. Lundberg et al. [91] investigated the integration of a SOFC into a Mercury 50 GT without bottoming cycle and with bottoming steam and ammonia Rankine cycles. The GT was simulated considering design point operation with constant geometry. However, minor modifications were necessary to accommodate the fuel cell. The inlet stators have been adjusted, the air inlets to the combustion chamber were resized due to the higher inlet temperature and the combustor main case has been redesigned using high temperature material. Park et al. [92] integrated a SOFC into a fixed geometry micro GT and found a significant reduction of the GT power generation compared to the design point. The GT power outlet can be increased by increasing the operating temperature of the SOFC. However, this requires the GT to operate close to the compressor surge line. Costamagna et al. [93] conducted simulations on the integration of a SOFC into a micro GT using compressor performance parameters and found that even for very small systems of 300 kW efficiencies of greater than 61%-LHV are possible. Furthermore, variable speed operation has shown to be superior to constant speed operation of the GT under part load conditions. Barelli et al. [94] showed that under part load the efficiency can be maintained at a very high level when using variable speed control for the GT. A similar system was studied by Campanari [95] under part load conditions who came to the same conclusion and additionally showed that variable speed operation of the GT is advantageous (with respect to efficiency) under varying ambient conditions.

Likewise, more complicated designs have been studied. Massardo and Lubelli [96] studied four SOFC-GT configurations with a bottoming Rankine cycle. Two atmospheric SOFCs with internal reforming that indirectly transfer heat to the working fluid of the GT and two cases where the SOFC is directly integrated into the Brayton cycle. Efficiencies of greater than 75%-LHV are possible. In [97] Massardo and Magistri analyzed the plant configurations from [96] with respect to economics and found that plant configurations

with the highest efficiency do not represent the most cost-efficient solution, and that the lowest cost of electricity is obtained when the ratio of SOFC cost to total plant cost is around 0.7. Direct integration of the SOFC into the Brayton cycle has shown to be more economical. In [98] Campanari and Macchi compared SOFC-GT hybrid cycles in combination with various bottoming Rankine cycle configurations reaching efficiencies close to 75%-LHV. Four enhanced GT-cycles in combination with SOFCs have been analyzed by Kuchonthara et al. [99]. The proposed triple cycle, consisting of a SOFC-GT hybrid and a downstream Rankine cycle, as well as the two variants of steam injection cycles haven shown lower efficiencies than the HAT-cycle hybrid. While cycles with SOFC off-gas recuperation exhibited higher efficiencies at higher pressure ratios, exhaust gas recuperated cycles like the HAT-cycle showed the highest efficiency at a pressure ratio of 5. A GT-steam injection cycle has also been investigated by Motahar and Alemrajabi [100] and showed that steam injection can substantially increases power output and efficiency. Bevc et al. [101] compared two variants of intercooled, recuperated reheat SOFC-GT hybrid cycles with a standard hybrid cycle and saw efficiency gains of up to 7% points. Yi et al. [102] studied a hybrid with GT intercooling and fuel humidification. Araki et al. [103] compared high temperature SOFC hybrid system to a serial low-high temperature SOFC configuration. Granovskii et al. [104] compared a SOFC-GT hybrid plant with steam generation for reforming with an anode off-gas recycle and bottoming Rankine cycle scheme and found that the anode off-gas recycle scheme is more efficient under various fuel to oxidant ratios and oxidant utilization rates. Rao and Samuelsen [105], [106] studied SOFC-HAT cycle combinations, which achieved efficiencies of up to 76%-LHV. Cost of electricity were based on a SOFC cost of 1000 \$/kW and ranged between 3.54 ct/kWh and 4.02 ct/kWh. Sghaier et al. [107] studied a SOFC hybrid with ammonia refrigeration cycle to partly cool the gas turbine inlet air in order to reduce compression power. Campanari et al. [108] compared a highly efficient large scale atmospheric SOFC-Rankine cycle to a pressurized SOFC-Brayton cycle plant fed by natural gas and showed that efficiencies of 75.2%-LHV and 78.7%-LHV are theoretically possible.

Natural Gas with Carbon Capture. Furthermore, systems with carbon capture have been reported in literature. Araki et al. [109] studied SOFC-GT hybrid cycles with carbon capture comparing a amine process to a high fuel utilization case reaching overall efficiencies of 68.5%-LHV with 100% carbon capture. As a side product of carbon capture, water is condensed from the anode off-gas in the high fuel utilization case. Möller et al. [110] compared a base case without carbon capture to a MEA process downstream of the gas turbine reaching efficiencies of 63.0%-LHV while capturing 67% of the carbon limited by the amount of heat available for absorbent regeneration. A decrease of 1.6% points versus the base case is observed. In order to increase the CO₂ concentration in the expanded flue gas, water is condensed via direct contact condenser prior carbon removal. Various oxy-combustion cases with 100%

carbon capture have been studied by Duan et al. [111] reaching efficiencies of up to 63.3%-LHV. Simultaneously, 100% of the water were removed from the flue gas via condensation most likely through intercooling of the compressed sequestration stream. Similar oxy-combustion configurations were investigated by Mahisanana et al. [112] reaching electrical efficiencies of up to 69.0%-LHV while capturing 100% of CO₂ and recovering 100% of the water in the anode off-gas. Isfahani and Sedaghat [113] studied a chemical looping process for carbon capture. Their zero-water plant reaches an efficiency of 51.4%-LHV, while capturing 100% of the produced CO₂. A further study by Campanari et al. [114] investigates the effects of carbon capture for the SOFC systems described in [108] using a cryogenic CO₂ separation. In the carbon capture scenarios 66% of the water is being recovered as a side product from carbon removal. However, the produced water from hydrogen combustion in the burner is released into the atmosphere. The efficiency of the Brayton cycle setup was reduced by 7.9% points (LHV) due to the reduced mass flow through the turbine and the auxiliary load of CO₂ capture and compression.

Dynamics. Since the first commercialization of a SOFC-GT hybrid system, dynamic system operation has become a major interest of the research community. Roberts et al. [115] investigated the dynamic operation of the commercial Siemens Westinghouse system and compared it to simulation results. Similar systems have been studied by Barelli et al. [116], McLarty et al. [117] and Mueller et al. [118], [119]. Also, degradation phenomena have been studied with respect to the optimization of the GT design and plant revenue [120]. For more details on dynamic models, it is referred to above-mentioned literature or this comprehensive review paper [121] as dynamic simulations are beyond the scope of this work.

Biogas. Biogas fueled SOFC-GT hybrid systems have been studied to a lesser extent. Krummrein et al. [122] identified challenges for biogas utilization in the combustor and a decrease in efficiency of 2.1% points (LHV) when utilizing biogas compared to methane. Wongchanapai et al. [123] studied effects of pressure ratio, TIT and internal temperature gradients in a direct biogas SOFC hybrid system. Kim and Kim [124] studied a simple hybrid cycle configuration for a biogas site in Korea, which reaches efficiencies between 50 and 55%-LHV. The payback period was estimated to be 5.8 years with a plant cost of \$7000/kW. Zabihian et al. [125] investigated SOFC-GT hybrids with and without anode off-gas recirculation for methane, natural gas and various biogas sources. Anode off-gas recirculation has shown to be 4.8-7.9% points (LHV) more efficient than steam injection and efficiency penalties for using biogas versus natural gas range between 4.2 and 4.7% points (LHV) for the anode recirculation configurations. Kaneko et al. [126] investigated control strategies for fluctuations in the biogas composition in a simple SOFC-GT hybrid cycle configuration and showed that with a bypass valve around the recuperator, the temperature and power output of the hybrid system can be stabilized. Curletti et al. [127] studied the techno-economics of large-scale biogas SOFC-GT hybrids with and without carbon capture to achieve negative carbon emissions. The

economics revealed that it is most profitable to operate the fuel cell at fuel utilization factors of less than 0.7. Also, Siefert and Litster [128] investigated the economic performance of a hybrid plant. The plant was fed by biogas from a wastewater treatment facility and their analysis showed that the economic performance of the hybrid configuration is superior to micro gas turbines or internal combustion engines, mainly due to the high efficiency of the system.

Syngas. Similar studies have been conducted for integrated gasification fuel cell (IGFC) cycles; whereby, the SOFC is integrated into a GT forming a hybrid. Lobachyov and Richter [129] were one of the first ones to study such a system. Their system consists of a CO₂ acceptor gasification process, which is comprised of a gasifier and a regenerator. Thus, a nitrogen free syngas is obtained without needing an air separation unit. Their system configurations reached efficiencies of up to 62.0%-HHV. Kuchonthara et al. [130] studied an IGFC system with thermochemical recuperation comprised of an oxygen-blown fluidized bed gasifier and a SOFC. Lanzini et al. [131], [132] studied IGFCs with shell gasifier and carbon capture. Methanation of the low methane content syngas has shown efficiency advantages, however, due to the higher investment cost, cost-of-electricity remained almost constant. Also, Huberg [133] investigated CO₂ capture in IGFC plants comparing oxy-combustion, H₂-membranes and O₂-membranes. Rao et al. [134] designed a hybrid system with an air-blown advanced transport gasifier in combination with a SOFC for power production and hydrogen production with and without carbon capture. A similar system has been studied by Verma et al. [135] for reaching ultra-low emission targets. A very comprehensive study conducted by the U.S. DoE [136] investigates combinations of E-GasTM gasifier or catalytic gasifier with atmospheric or pressurized SOFCs with internal reforming as well as natural gas injection. Water is being recovered during CO₂ cooling and CO₂ dehydration of the oxy-combustor exhaust, however, no details are available about the water recovery and the employed cooling towers consume water at an enormous rate, which exceeds the amount of recovered water by far. A further study conducted by the U.S. DoE [137] focuses solely on catalytic gasification. Catalytic gasification has shown to be superior to high temperature gasification technologies as lower temperatures promote methane formation leading to a higher cold gas efficiency, which is desirable in power applications. High methane content is of special interest when integrating gasifiers with high temperature fuel cells, as internal reforming inside the SOFC increases the efficiency. Hydrogasification is a technology currently under development which is able to produce syngas with very high methane content and has shown to be superior to other gasification technologies when combined with high temperature fuel cells. Li et al. [138] studied catalytic hydrogasification with oxygen as agent for the regenerator in combination with 90% carbon capture and showed that efficiencies of 61.5%-HHV are possible. Another study by Li et al. [139] employed hydrogasification and focused on the stack configuration of the SOFC using a quasi 2-dimensional SOFC model. The results show that it is important

to use a SOFC model that can resolve internal temperature gradients. Li observed efficiency losses of at least 3.2% points HHV when comparing the quasi 2-dimensional model to a 0-dimensional model due to thermal limitations that are not visible in 0-D models.

Hydrogasification. SOFC-GT hybrid efficiencies reached by IGFCs with hydrogasification as gasification technology show great synergies between the hydrogasifier and the SOFC. Hydrogasification has been used since the 1930's to produce syngas from solid fuels. Due to the slow kinetics, especially at lower reactor temperatures, a catalyst is used to improve the throughput and the efficiency. Catalytic coal gasification has not been tested beyond early development stages and over the last decades there has been no significant research interest in this technology. One of the most extensive studies was conducted by Exxon in the 1970's which identified K_2CO_3 and Na_2CO_3 as effective catalysts [28]. For plant simulations, mostly thermodynamic models based on Gibbs free energy are used [140]–[142]. A chemical equilibrium model has been developed by Yan et al. [143], which showed excellent agreement with experimental data. Furthermore, Yan et al. [144] developed a kinetic model for coal hydrogasification. CFD studies on entrained flow hydrogasifier were conducted by Xuan et al. [145] and Yan et al. [146]. In recent years, more experimental work has been conducted in the field of hydrogasification and especially steam hydrogasification because of the improved kinetics. The University of California, Riverside became a center of expertise and developed their own hydrogasification technology under the CE-CERT program [142], [147]. The program focuses on steam hydrogasification investigating effects of gasification atmospheres, catalysts and feedstock as well as the formation of different sulfur components during hydrogasification [148]–[150]. Also in China, the interest in hydrogasification is increasing and several experimental studies on hydrogasification have been conducted [151], [152]. Hydrogasification simulation work is still in the very early stages due to the complexity of the gasification reaction but will be an important part for the development of commercial hydrogasifiers as the development of gasification technology is still mostly empirical and thus capital intensive.

Industrial engagement in hydrogasification currently is low. Hydrogasification has been known for a long time but suffers from low carbon conversion. The first intensive research program on hydrogasification was launched in the 1970's by Exxon as a reaction to the oil crisis. Until today hydrogasification did not make it beyond the research scale. In the U.S. Viresco Energy sponsored some of the initial research of the CE-CERT program and planned to construct a 5 ton per day pilot plant in Kanab, Utah in 2010 [153]. However, the funding of the project was curtailed in 2013. Today, only ENN in China is pursuing commercialization of hydrogasification. ENN's hydrogasification process aims to produce a large amount of oils as salable byproduct. The gasifier is an entrained flow reactor without catalyst. The current reactor is designed for capacities of up to 50 tons per day. A large-scale reactor (400 tons per day) is expected to

be constructed by 2018. The gasifier produces a syngas with 10-15% oils and more than 80% methane. Operating temperatures are in the range from 800 °C to 1000 °C and operating pressure is around 7.0 MPa (operating range 5-10 MPa with 7 MPa resulting in the highest oil production). The process utilizes low rank coals, such as sub-bituminous coal and/or lignite. Hydrogen is recovered from the methane rich syngas via membranes and oxygen injection into the hydrogen feed gas is used to preheat the feed stream. The carbon conversion of the hydrogasification process is around 60% and the unreacted char is fed into a steam gasifier to produce additional hydrogen for the hydrogasification process. A process flow diagram of ENN's hydrogasification process can be seen in **Figure 2-11**.

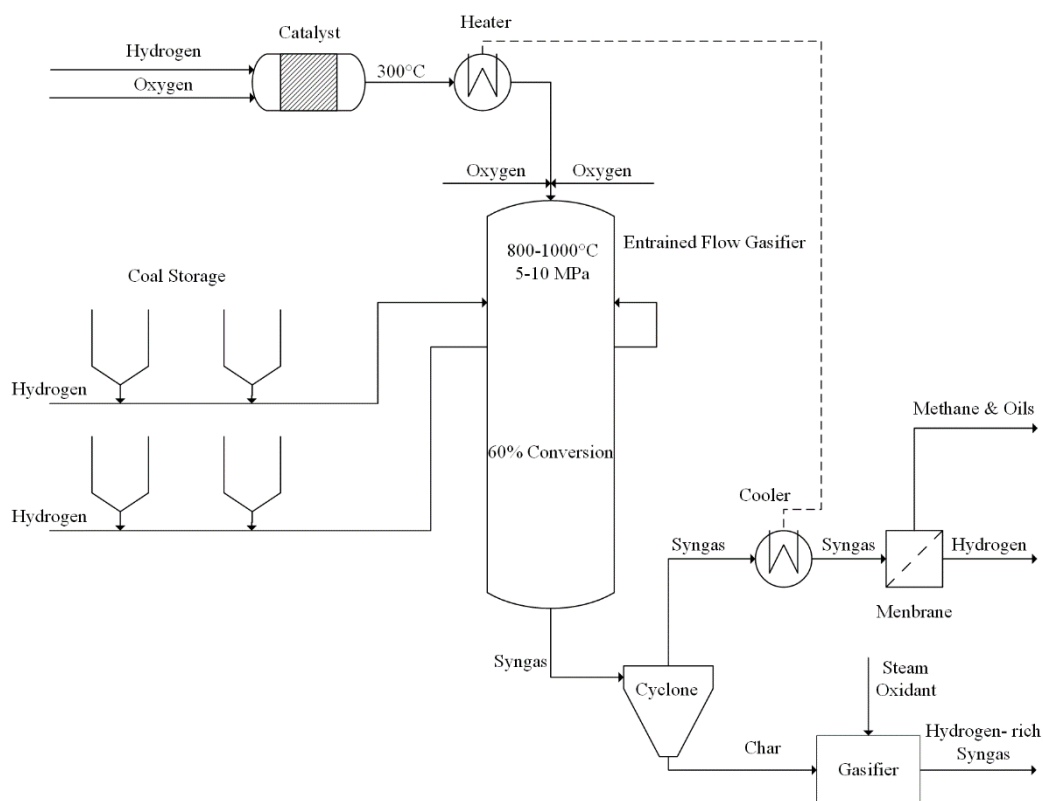


Figure 2-11: Flow diagram of ENN's hydrogasification process.

Economics. Another major obstacle on the way to commercializing SOFCs is their cost (due to the relatively low level of research and development and production scale compared to internal combustion engines or batteries) and projections for SOFC costs considering large-scale production are still difficult to make. Most system studies rely on rough estimates or cost targets set by institutions like the U.S. DoE.

Rao and Samuelsen [106] studied the cost of SOFC-GT hybrid configurations assuming a base cost of 1000 \$/kW based upon projections of the Siemens-Westinghouse hybrid. A study by Kim and Kim [124] uses a much larger value of 7000 \$/kW and assumes a SOFC cost of 3500 \$/kW for large scale production. Curletti et al. [127] based their cost estimates on U.S. DoE targets [154]. Siefert and Litster [128] estimated

the SOFC cost with 1,700 \$/m² of active cell area and Arsalis [155] employed a cost correlation dependent on active cell area and operating temperature, which was then also adopted in [85], [90]. Massardo and Magistri [97] proposed a cost correlation based upon the specific power (kW/m² cell area), resulting in specific SOFC costs (cost on a per kW basis) ranging from 700 – 2,200 \$/kW. However, all these cost estimates or methodologies are missing a more fundamental basis. In many instances detailed kinetic models for the SOFC are not available to construct accurate costs and instead specific SOFC costs are simply employed. For accurate thermo-economic system studies, such specific cost estimates are imprecise due to the wide range of operating conditions encountered in SOFC systems, especially when considering pressurized operating conditions. Even if the active cell area is being determined, the above-mentioned cell area specific cost correlations cannot capture costs correctly; while some cell components might scale with the active cell area, others such as framing or assembly scale on a different basis. One of the most sophisticated and recent studies on SOFC costs has been produced by Battelle Memorial Institute [156], which developed a detailed SOFC cost estimation method and established cost projections for atmospheric pressure SOFC combined heat and power systems. The developed cost model contains various individual cell components and manufacturing processes, but no parametric study of the cell design nor an optimization with respect to costs was conducted as it would require detailed modeling of the entire system to establish the required operational data, which appears to be beyond the scope of their study.

2.4.4 Literature Review on Flue Gas Water Recovery

Over the course of the last decade, water recovery technologies from flue gas have attracted academic and industrial research due to more frequent and severe draught events and the resulting water scarcity in many areas including the south west of the U.S. and northern parts of China. Currently, commercially available systems for flue gas water recovery include surface condensers as well as direct contact condensers. Most studies found in literature employ conventional surface condensers for water recovery, which are problematic with respect to fouling, pressure drop, heat sink availability and only a portion of the water in the flue gas can be recovered. Water recovery from boiler flue gases has been studied by Levy et al. [157]. The study purely focuses on flue gases from coal boiler plants as well as emission reduction through water condensation. Levy et al. showed that with surface condensers water recovery efficiencies of 70% may be achievable. Important factors on the condensation rate are the cooling water temperature as well as the water vapor content in the flue gas. Furthermore, finned tube heat exchangers showed superior performance compared to regular tube heat exchangers. Emissions from coal fired boiler plant power generation were effectively reduced by flue gas water condensation. Mercury emissions were reduced by 60% and substantial amounts of sulfur oxides reacted with water during the condensation process. Utilizing the sensible heat for boiler feed water pre-heating increased the heat rate by 1.8 – 3.9% but limits the water

recovery. Using ambient air, to reject the heat showed water recovery efficiencies of greater 70%. For a pulverized coal-fired boiler plant the condensed vapor could replace between 6.4 – 24.8% of the cooling tower's makeup water, depending on the moisture content of the coal/flue gas. The simulations are focused on the stack section of the plant and do not include the system integration into the power plant as a whole. Treatment strategies for the recovered water, in order to comply with water quality guidelines of the cooling tower equipment, has been addressed in [158]. The suggested treatment consists of a filtration system and an ion exchange bed. Due to severe corrosion of the captured water Alloy 22 and 690 along with FEP and PTFE showed acceptable corrosion resistivity, whereby, PTFE showed essentially no degradation. However, due to the low thermal conductivity a PTFE-based heat exchangers, three times the size of an Alloy 22 heat exchanger would be required. As a consequence, a PTFE heat exchanger would be more expensive than an Alloy 22 heat exchanger despite its lower material cost. Below the water dew point, 304 SS would be the preferred exchanger material. The economic assessment showed advantages of using boiler feed water-cooled flue gas condensers to displace water imports. Xiong et al. [159] employed plastic heat exchangers for water recovery using a slip stream of a lignite PC plant. Flue gas waste heat recovery of up to 80% and water recovery efficiencies of up to 70% were observed, while achieving heat transfer rates of 275 W/m²/K. Machackova et al. [160] studied heat exchangers models for better prediction of heat transfer during condensation of water vapor from coal flue gas. Terhan and Comakli [161] investigated the potential of water recovery from a 60 MW natural gas fired district heating system. The economic analysis estimates the payback period for the investment in the heat exchanger to be less than 1 year.

Direct contact water columns, which can generate significantly larger surface areas at lower cost compared to surface condensers, might be more cost attractive than conventional condensers. Direct contact columns have been considered for water condensation from flue gas for a long time and many column types – trays and packed beds – have been studied for water condensation [162]–[166]; yet, the design of direct contact heat exchangers is still mostly empirical. Direct contact heat exchangers have been mostly employed in PC plants with post combustion CO₂ capture to cool the flue gas before the CO₂ absorption unit [166]–[169] or in SOFC-GT hybrid systems with post-combustion CO₂ capture [110]. Also, the potential of direct contact condensers as a flue cleaning technology has been recognized (e.g., reduction of SO_x and NO_x) [166], [169], however, the treatment of the recovered water is not addressed. Beyond post-combustion CO₂ capture, direct contact condensers have been considered in humidified air turbine (HAT) cycles [170], though, most HAT systems are still in developmental stage and only studied in non-commercial applications.

Water recovery technologies have not just been developed for industrial flue gas applications but also attracted attention for drinking water recovery from humid air. The Fraunhofer Institute developed a technology for drinking water generation from moist air based on a regenerative desiccant process. This

technology is able to recover water from air with a relative humidity of less than 100% without cooling via a saline solution [171]. Similar technologies could be applied to water recovery from flue gases, too. Siemens developed a technology for water recovery in industrial power plant applications using a liquid desiccant that can be sprayed into the stack and absorbs moisture with a minimal pressure loss [172], [173]. Siemens considered various desiccants and determined the following order of desiccant strength: LiBr, LiCl, LiNO₃, CaCl₂, triethylene glycol (TEG). The corrosivity of the respective compound follows the reverse order. The cost factors associated with the compounds were estimated to be 10 for LiBr, 9 for LiCl, 9 for LiNO₃, 1 for CaCl₂ and 7 for TEG. Siemens estimated the environmental impact factors of the desiccants as 10 except for TEG which is 8. Based on those criteria Siemens proceeded with the development of a CaCl₂ absorber system. Riffat et al. [174] studied a water absorption and heat recovery system in the context of boilers for central heating and concluded that absorption systems are more economical than conventional surface condenser in heating applications. Westerlund et al. [175] studied a similar heating system utilizing biomass and showed that flue gas water absorption systems can significantly reduce particulate matter emissions, while increasing the efficiency of the heating system. However, this might be only the case for heating systems, which can utilize low temperature heat sources. Similar systems have been studied by Ye et al. [176], Wei et al. [177], Wang et al. [178] and Wang et al. [179].

A technology that is still under development, but has been used for filtration in the past, consists of transport membranes, which can be utilized for condensation purposes (TMC). TMCs consist of nano-engineered ceramics that work on the principle of capillary condensation. Wang [180] investigated TMCs installed in boiler plants, which have been operated for more than 3 years without degradation. A 12% reduction in fuel consumption and a 20% reduction in water consumption could be observed after employing this technology in a plant [180]. A short economic analysis for boiler plants has been conducted for various coals but the integration of the system into IGFC and its economic analysis are yet to be done. In further research, Wang et al. [181] showed that the high-quality water recovered by the TMC can be used internally and improve the overall plant efficiency. A 40% water recovery led to an 5% efficiency increase. More experimental work on TMCs was conducted by Zhao et al. [182]. Increased water flow rates showed to be an effective tool to maximize water recovery and heat recovery. Higher water vapor content can significantly improve heat and water transfer rates; however, has little impact on the overall recovery. Chen et al. [183] investigated the impact of feed gas flow rate, cooling water and vacuum on the water side on water and heat recovery. The study shows water recovery efficiencies of 80% and a heat recovery of 40%. Bao et al. [184] studied TMCs and compared the heat transfer rate to conventional metal tube condensers, showing that the Nusselt number in case of the TMC is 50-80% higher due to simultaneous heat and mass transfer. Lin et al. [185] derived a Nusselt number correlation based on CFD simulations using macro-kinetic rate expressions for evaporation and condensation. Soleimanikutanaei et al. [186] studied TMCs

numerically based on a first principle hybrid condensation model and showed that in tube bundles the condensation rate and heat transfer rate increases with a decrease in tube spacing. In the U.S., the TMC technology is being developed by the Gas Technology Institute (GTI) and the U.S. DoE, and research activities are currently expanding to investigate TMCs for latent heat recovery in pressurized oxy-combustion applications. GTI licensed its technology to Cannon Boiler Works, Inc. since 2009, which is currently offering TMC technology together with their Ultramizer® boiler technology [187].

2.4.5 Summary Literature Review

Various SOFC-GT hybrid cycles have been studied in literature, ranging from simple systems without carbon capture to complex integrated gasification systems with carbon capture and co-generation. Since system studies consider a large network of reaction vessels, simplifications are necessary. While some studies utilize more detailed models to study certain aspects of a plant configuration, widespread application of detailed models for multiple plant components remains challenging at this time. Key areas in a SOFC-GT hybrid system that require detailed analysis are the SOFC, GT, eductor and prediction of carbon deposition in pre-reformer and SOFC.

Thermal SOFC-cell gradients constitute one of the greatest challenges for the operation of SOFCs and require careful evaluation. Although this problem is widely acknowledged, most system studies still rely on 0-D SOFC models due to the simplicity and reduced computational time. Overall, there is only a limited number of full system studies that use higher dimensional fuel cell models [103], [109], [120], [123], [188]–[190], yet they illustrate how important it is to consider the reaction kinetics inside the fuel cell to prevent critical thermal stresses.

Equally important in a SOFC-GT hybrid is the GT. Gas turbines are highly complex machines engineered for the use of natural gas. The operation of GTs in a hybrid shifts the operating point away from its design point. System studies in literature heavily rely on hypothetical GTs based on the current development status of natural gas fired engines. Since GT development costs are extremely high it is unlikely that GT vendors will provide GTs specifically for SOFC-GT hybrid application in the foreseeable future. Thus, since we are aiming for a timely introduction of this hybrid technologies current GT offering need to be considered. Current literature made some efforts to address this issue. Lundberg et al. [91] studied a Mercury 50 gas turbine for integration into a hybrid. While they matched the SOFC operating pressure to the GT's compression ratio and mass flows, they did not account for the flow mismatch and essentially operated the gas turbine at on-design condition. Some studies [82], [86], [88], [126], [191], [192] include compressor and/or turbine maps, e.g., for studying part load behavior, but the maps are normalized to the design point condition of the hybrid instead of the natural gas design point. In [92], [93] fixed geometry micro GTs have been paired/integrated with SOFCs and investigated for part load operation, which showed

that under part load variable speed control is outperforming constant speed operation. Detailed design point optimization studies using fixed geometry GTs under consideration of thermal SOFC gradients, fuel utilization factors, operating mode and surge margin are still unavailable in current literature.

Crucial for the operation of the SOFC and pre-reformer is the avoidance of carbon deposition on the catalyst. The application of an actual gas stability analysis to determine the minimum amount of recirculation to prevent carbon deposition in the pre-reformer and/or fuel cell is rarely seen [123]. The vast majority of studies found in literature base anode gas recirculation on the steam/carbon ratio, which can be misleading since hydrogen and carbon monoxide also participate in the equilibrium.

Another essential component in SOFC-GT hybrid plants with anode gas recirculation is the eductor, which allows compression of the recycle stream via a high-pressure stream without the need of cooling beforehand, like necessary when using compressors. Simple 1-D thermodynamic eductor models tend to overpredict the performance as boundary effects are neglected [193]. Advanced eductor models have been utilized in [194]–[197]; however, this model was only used in combination with 0-D fuel cell models, which cannot resolve critical temperature gradients within the fuel cell and without precise prediction of the carbon deposition equilibrium.

Economic considerations of SOFC-GT hybrids are insufficiently addressed in literature. While most system studies focus entirely on thermodynamic performance using price targets or simple cost correlations [54], [106], [124], it was not until recently that more sophisticated SOFC cost projections became available (even though only for atmospheric CHP systems) [156]. Nevertheless, detailed cell design cost analyses have not been employed in system studies, yet.

A major effort of this research is to investigate SOFC-GT hybrid systems in the context of water use reduction and flue gas water recovery. No work has been reported in literature investigating water use reduction and flue gas water recovery in SOFC-GT hybrid systems nor a study that compares different flue gas water recovery technologies. Furthermore, plant internal water treatment for net water export and in combination with SOFC-GT hybrid systems remains unaddressed in literature. Flue gas recovery technologies have been investigated in isolated scenarios [157], [158], [184]–[186], [198], [159], [161]–[163], [172], [173], [180], [181], however, the full plant integration of most of the technologies still has to be shown.

Net-water has been produced by SOFC-GT hybrid configurations as a side product of employing early-stage SOFC technology without internal reforming [78], [79], as a side product of carbon capture using oxy-combustion [111], [112], [127], or as a side product of other carbon capture technologies that require water removal prior to carbon capture [108], [110], [113], [114]. In all these cases, the recovered water is not treated and rather handled as waste stream, which adds further stress on the water and electricity infrastructure. In the above-mentioned studies, water recovery is only modeled on a thermodynamic basis

without addressing cooling loads and the plant's water management. First-principle water recovery models tailored to the individual operating condition have not been studied in literature. Furthermore, effects of plant internal water treatment for water export and zero liquid discharge have not been discussed in SOFC-GT hybrid literature. Beyond that, synergetic effects between water recovery and carbon capture have yet to be explored. This work seeks to fill this gap in literature and study various water recovery technologies in the context of SOFC-GT hybrid cycles (using a combination of advanced modeling strategies as described above) with net-water production and zero liquid discharge to mitigate stress on the water and energy infrastructure and improve resiliency. In this context various feedstocks — natural gas, biogas, coal and biomass — and plant scales —10 MW, 50 MW and 100 MW — are investigated. Furthermore, water recovery and carbon capture share certain characteristics and synergistic effects can be expected, despite the fact that many carbon capture technologies increase water consumption.

Chapter 3

Approach

Process simulation software is a powerful and cost-effective tool to evaluate emerging technologies, while substantially accelerating the technology's development. In this work, system studies using thermodynamic solution strategies in combination with detailed first-principle models will serve as foundation for performance and economic analyses. Detailed first-principle models will be developed for important key components and allow to better understand the physical phenomena involved in the process. The approach used in this study seeks to evaluate the performance and economics of SOFC-GT hybrid cycles using process simulation software in combination with other computational tools.

In the approach section, tasks are developed for each of the stated objectives and the work scope is delineated to achieve the aforementioned goal of this study.

TASK 1: DEVELOP METHODOLOGY

The current technology status of SOFC-GT hybrid cycles will be assessed, and a literature study will be conducted to evaluate the progress on SOFC-GT hybrid research. After this first evaluation a strategy will be put forward that will support the development of SOFC-GT hybrid systems. Future needs with respect to feedstock, greenhouse gas emissions and other factors have to be considered. Centered around technologies available today, scales for next generation SOFC-GT hybrids and more futuristic plants will be proposed to systematically support the introduction of fuel cell hybrid technology into the market.

Furthermore, the development of a methodology requires evaluating modeling strategies and critical parameters will have to be identified that can serve as characteristic input variables for simulations and economics. Moreover, a strategy for cost estimation will be developed to predict the cost of electricity and cost of recovered water. The simulations will be carried out in a well-established commercially available chemical processing software.

TASK 2: ESTABLISH A DESIGN BASIS AND IDENTIFY STUDY SCENARIOS

Secondly, a design basis needs to be established, which identifies and defines the used feedstocks and site characteristics. This basis will be the same for all simulations to ensure comparability between the different scenarios. Furthermore, the design basis defines critical operating parameters that influence the plant performance, e.g., pinch temperatures or purity of product streams. Feedstocks will have to be chosen

based on availability, reliability and steadiness over the entire operating period of the plant. Additionally, feedstocks have to be suitable for the respective technologies chosen in the individual cases. Cases are established based upon the current technology status to offer a market introduction in near future. Furthermore, global and local trends for feedstock selection, fossil fuels and biofuels, as well as regulations with respect to greenhouse gas emissions will have to be considered.

TASK 3: DEVELOP WATER USE REDUCTION STRATEGIES AND IDENTIFY WATER RECOVERY TECHNOLOGIES

A central part of this work is to investigate water use reduction strategies and active flue gas water recovery in order to address tensions faced in the water energy nexus. To this extended, water use reduction strategies will have to be identified. Areas of interest for water use reduction are: I) efficiency improvement, II) reduction of cooling load, III) alternative cooling strategies, IV) plant integration, and V) internal water management. Additionally, flue gas water recovery will be assessed, and water recovery strategies will have to be identified. Various physical principles will be considered for vapor separation from flue gases. Furthermore, this task deals with the model development of promising flue gas water recovery technologies and the elaboration of their system integration strategies. Another important aspect of the water energy nexus is wastewater management as part of the overall plant design.

TASK 4: DEVELOP AND CALIBRATE MODELS OF KEY TECHNOLOGIES

To accurately predict the performance and more importantly, the feasibility of the plant design, thermodynamic models have to be combined with more detailed first-principle models. First-principle models allow spatial resolution of characteristic parameters. The fuel cell operating range for example is limited by temperature gradients and the temperature dip in the pre-reformer determines the amount of water needed to prevent carbon deposition. This is particularly important for the water recovery technologies, where the pressure drop is of major importance besides the amount of water recovered. Equipment downstream of the GT has a significant impact on the overall plant efficiency as it changes the GT back pressure. Next to the water recovery technologies other critical key models need to be established and validated. This process is crucial for establishing the performance of the plant, which serves as basis for the thermo-economic analysis.

TASK 5: DEVELOP BASE CASE SCENARIOS

Before the water technologies can be compared, a base case scenario will be developed, which acts as reference case for determining important performance characteristics and the value of the recovered water. The base case scenario will be established after a series of design and optimization studies in order to select the most economical base case for the implementation of the water recovery technologies. Furthermore, a

preliminary study on the fuel cell design will be conducted considering cost of manufacturing while minimizing temperature gradients.

TASK 6: DEVELOP CASES WITH WATER RECOVERY

The identified water recovery technologies will be implemented in the most promising base case scenario and evaluated. In this context, different integration schemes will be evaluated and compared. Integration strategies for effective heat management and reduction in cooling loads will be investigated in combination with dry-cooling technologies. Different water recovery technologies will be evaluated for the small-scale plants and the most promising technologies will be considered for future large-scale plants. Larger scale plants include carbon capture scenarios, and in this context synergies between water recovery and carbon capture will be investigated and explored.

TASK 7: THERMO-ECONOMIC ANALYSIS

Performance characteristics based upon thermodynamics like efficiency are very important. However, the thermo-economic analysis is of even greater importance. The thermo-economic analysis will use the steady state simulation results for the different plants as input to determine the economic performance of the power plant. Cost of electricity and costs associated with carbon capture and water recovery are very important to evaluate the profitability of a plant. For the cost analysis scaling parameters and scaling exponents will be derived from literature data. A financing structure of the plant will be elaborated to estimate the cost of electricity. A cost analysis for various water recovery technologies will be conducted, and key values will be compared.

Chapter 4

Methodology

4.1 Design Basis

4.1.1 Site Characteristics

The plant site is considered to be in the U.S. Due to the variety of geographical climate zones within the U.S., ISO conditions are chosen as basis for ambient conditions. ISO conditions are commonly chosen in thermodynamic power plant analyses and will allow to draw direct comparisons to many other results found in literature. However, it needs to be mentioned that biogas, biomass as well as low rank coals are not valuable enough to be shipped over large distances and plants will be located close to the resource. Biogas and biomass tend to be more evenly distributed than coal mines and can be considered available throughout the country. Coal mines, on the other hand, are only found at specific locations with coal resources. Powder River Basin (PRB) coal for example is found in the north western U.S. and a potential plant using this resource will have to be built close to the mine mouth with cooler ambient conditions than ISO conditions. However, since this study seeks to investigate the impact of low water consumption and water recovery technologies, e.g., dry cooling and flue gas condensers, using different ambient conditions can significantly skew the results. Penalties due to dry cooling in colder regions can be mitigated by the cooler ambient air. In order to avoid effects of ambient air on the system's performance, all plants will use the same ambient conditions to facilitate this comparison of these new technologies. Site characteristics are summarized in **Table 4-1**.

Table 4-1: Site Characteristics

Parameter	Value	Unit
Elevation	0	m
Barometric Pressure	1.013	bar
Design Ambient Temperature, Dry bulb	15.0	°C
Design Ambient Temperature, Wet bulb	10.8	°C
Design Ambient Relative Humidity	60	%
Design Cooling Water Temperature	15.8	°C

Design Basis

Considering an ambient wet bulb temperature of 10.8 °C and using a 5 °C temperature approach, a cooling water temperature of 15.8 °C is obtained. The cooling tower range is assumed to be 10 °C. Conventional heat exchangers are simulated with a pinch temperature of 11 °C and air-cooled exchangers are assumed to operate with a pinch temperature of 25 °C. The composition of ambient air used in the simulations is provided in **Table 4-2**.

The site is assumed to be of level topology and accessible by rail and highway. Special site conditions such as flood plain considerations, soil conditions, rainfall and snowfall criteria, seismic design, noise regulations are not considered in the site condition.

Carbon dioxide is stored in a 171 m thick saline formation at a depth of 1239 m. The pressure of the formation is 84 bar, which is representative for an average storage site. The injection rate per well is determined to be 9360 t/d. For the transport of about 80 km, the CO₂ needs to be compressed to a pressure of 153 bar before it leaves the plant to reach the storage facility at a high enough pressure [199]. The CO₂ is injected into the pipeline as super-critical fluid and maintained at supercritical pressure at all times to prevent transients similar to water hammer [200]. A high purity of CO₂ is required to limit compression power, storage capacities and avoid corrosion. Limits of trace components are listed in **Table 4-3** [201]. Nitrogen, Ar, CH₄ and H₂ impurities are mostly limited to reduce compression power and storage capacity concerns but might also cause miscibility issues under high pressure. The presence of N₂, CH₄ and H₂ may further lead to ductility issues and require an increased pipe strength. Oxygen limits are in place to reduce corrosion potential as well as concerns related to the growth of aerobic bacteria. Similarly, H₂O is limited to reduce the formation of acids in the pipeline. Carbon monoxide and NH₃ are limited due to safety concerns.

Table 4-2: Air Composition

Component	Vol.-%
N ₂	77.315
O ₂	20.741
H ₂ O	0.987
Ar	0.924
CO ₂	0.033

Table 4-3: CO₂ Pipeline Specifications

Component	Unit	Value
Inlet Pressure	bar	153
Inlet Temperature	°C	35
N ₂	vol-%	<4
Ar	vol-%	<4
CH ₄	vol-%	<4
H ₂	vol-%	<4
O ₂	ppmv	<10
H ₂ O	ppmv	<500
CO	ppmv	<1,200
NH ₃	ppmv	<300

4.1.2 Fuel Specifications

Natural Gas

The average natural gas composition in the U.S. has been determined by the U.S. DoE. The values for the natural gas composition are average values from over 6,800 samples of pipeline quality measurements within the U.S. taken in 26 of its major metropolitan areas [202], [203]. The composition then has been modified with trace components that are critical for the operation of fuel cells; O₂ and sulfur compounds. According to [6], 0.01 mol-% of O₂ is commonly found in natural gas, along with 1.7 ppmv of sulfur-based odorants. Odorants are a mixture of different types of mercaptans. Main component and most widely used odorant is tert-butyl-mercaptan, which will be used in this study to represent the odorant's sulfur components. After taking trace components into account, the composition presented in **Table 4-4** is obtained. The lower heating value on a mass basis is 47.2 MJ/kg and the higher heating value on a mass basis is 52.3 MJ/kg. It is important to understand that the systems designed for natural gas must also be able to withstand peak shaving events where the O₂ concentration might rise to 4% [4] and other dynamics in the gas grid.

Table 4-4: Natural Gas Composition¹

Component	Mole-%
Methane	93.08746
Ethane	3.19957
Propane	0.69991
n-Butane	0.39995
Carbon Dioxide	0.99987
Nitrogen	1.59978
Water	0.00329
Oxygen	0.01000
Hydrogen Sulfide	0.00004
Tert-Butyl Mercaptan	0.00013

The supply pressures of the natural gas at the plant sites may vary by operator, location and demand. Based upon the Southern California Gas transmission and distribution lines the following supply pressures for the three different plant sizes of 10 MW, 50 MW and 100 MW are assumed. Many pipelines are exposed to the environment, thus, a supply temperature of 15 °C is used in this work.

¹ “Notes: 1. The reference data reported the mean volume percentage of higher hydrocarbons (C₄+) to be 0.4%. For simplicity, the above composition represents all the higher hydrocarbons as n-butane (C₄H₁₀). 2. The reference data reported the mean volume percentage of CO₂ and N₂ (combined) to be 2.6%. The above composition assumes that the mean volume percentage of CO₂ is 1.0%, with the balance (1.6%) being N₂.” [168]

Table 4-5: Natural Gas Supply Pressure

Plant Size	10 MW	50 MW	100 MW
Pressure	0.41MPa	0.41 MPa	1.38 MPa

Coal

Since hydrogasification faces challenges with respect to its carbon conversion it is important to choose a reactive coal. Subbituminous coal is very suitable for hydrogasification due to its high reactivity, relatively high energy content and high amount of volatile matter. The coal used in this study is a mine mouth Powder River Basin (PRB) coal from Montana Rosebud, area D [204]. A detailed analysis of the coal is given in **Table 4-6**. The higher heating value (HHV) on a dry basis is 26.8 MJ/kg and 19.9 MJ/kg on an as-received basis. The lower heating value (LHV) is given as 25.8 MJ/kg on a dry basis and 19.2 MJ/kg on an as-received basis [204]. As coal is stored outside, exposed to the ambient conditions, a feedstock temperature of 15 °C is assumed when entering the feeding system. The ash composition is not further specified, however, details on the ash composition can be found in [204]. The ash composition is not of major importance when analyzing gasification systems and can be treated as a pseudo component that leaves the gasifier unreacted. Nevertheless, it is important to note that the ash composition has great influence on ash properties like the initial deformation temperature and will have to be considered when choosing a coal or

Table 4-6: Montana Rosebud PRB, Area D, Western Energy Co. Mine, Subbituminous Coal [204]

Proximate Analysis	Dry Basis, wt-%	As Received, wt-%
Moisture	0.00	25.77
Ash	11.04	8.19
Volatile Matter	40.87	30.34
Fixed Carbon	48.09	35.70
Total	100.00	100.00
Ultimate Analysis	Dry Basis, wt-%	As Received, wt-%
Carbon	67.45	50.07
Hydrogen	4.56	3.38
Nitrogen	0.96	0.71
Sulfur	0.98	0.73
Chlorine	0.01	0.01
Ash	11.03	8.19
Moisture	0.00	25.77
Oxygen	15.01	11.14
Total	100.00	100.00
Heating Value	Dry Basis	As Received
HHV, kJ/kg	26,787	19,920
LHV, kJ/kg	25,810	19,195
Hardgrove Grindability Index	57	

gasification technology respectively. For fluidized bed gasifiers it is important to stay below the initial ash deformation temperature to avoid agglomeration of the coal particles. Entrained flow gasifiers, on the other hand, require a certain minimum viscosity of the slag in order to operate properly. The initial ash deformation temperature of the PRB coal under reducing conditions is typically in the range of 1180 to 1230°C [204]. The mercury content of this coal is 0.081 ppmw on an as-received basis [204].

Biogas

Biogas from digesters in sewage treatment facilities offers great potential for steady fuel supply while reducing waste. The content of higher hydrocarbons in biogas is significantly lower than in natural gas; however, the content of trace components is higher in biogas. Especially sulfides and siloxanes pose risks for continuous operation of fuel cell and GT. A detailed analysis of a wastewater treatment plant's biogas composition is given in **Table 4-7**. The biogas composition is taken from the King County Power Plant [205], which offers a very well documented composition of its sewage-derived fuel gas. Along with hydrocarbon compounds, CO₂, N₂, O₂ and water, trace components like siloxanes, sulfur components and chlorine components are reported, which are crucial for the application of fuel cell cycles. The physical properties of this biogas are: supply temperature 26.7 °C, LHV 17.8 MJ/kg, HHV 19.8 MJ/kg, pressure 5.5 inch water above atmosphere and the biogas is saturated with water [205].

Table 4-7: Biogas Composition [205]

Component	Mole-%	Component	Mole-%
Methane	60.5	H ₂ S	150 ppmv
Ethane	40 ppmv	Sulfur	CS ₂ +SO ₂
Propane	1 ppmv		Mercaptans
Butanes	0.7 ppmv		Organic di-/sulfides
Pentanes	7 ppmv		Thiophenes
Hexanes+	50 ppmv	Chlorine	Methyl Chloride
Carbon Dioxide	37.5		Cis 1,2-Dichloroethane
Nitrogen	1.0		Chlorobenzene
Oxygen	1.0		1,4 Dichlorobenzene
Water	saturated	Siloxane	Tetramethylsiloxane
Benzene	30 ppbv		Hexamethyldisiloxane
Toluene	400 ppbv		Octamethyltrisiloxane
Xylenes	850 ppbv		Octamethylcyclotetrasiloxane
			Decamethylcyclopentasiloxane
LHV 17.8 MJ/kg	HHV 19.8 MJ/kg		

Biomass

As mentioned before, next to physical and chemical feedstock properties, economic, ecological and social factors also play an important role when choosing biomass as feedstock. In order to reduce competition of biomass production with food production, only 2nd generation biomass crops are considered in this work. Important considerations for biomass production are: land usage, dry matter yields, energy yields, growing conditions, water use efficiency, fertilizer and pesticide usage as well as harvesting equipment, biomass drying and bulk density. In order to find the best suited biomass crop for this project some of the most prominent crops are compared on the basis of the above-mentioned criteria. The five selected crops for this comparison are 3 species from the hard wood category and 2 species from the grasses.

The first hard wood is hybrid poplar; hybrid poplar is a combination of various poplar species and is one of the fastest growing trees and can be grown in all parts of the U.S. Willow, the second hardwood selected, has fast growth rates and the wood has almost the same energy content as other hard woods like oak or maple. Most parts of the U.S. are suitable for willow cultivation including the northern areas. Third, eucalyptus trees have shown some incredible annual dry matter yields but are limited to the southern regions of the US. Grasses have the advantage of annual and simple harvesting while producing high dry matter yields. Two intensively studied grasses are switchgrass and miscanthus. All selected biomass crops belong to the 2nd generation crops and do not stand in direct competition with food production.

For gasification, feedstocks with high amounts of volatiles are desired. The higher the amount of volatiles, the more reactive is the feedstock, which leads to a shorter residence time and thus, to a smaller reactor size or higher through put. Eucalyptus has the highest volatile matter content among the hard woods (eucalyptus at 77.08%, hybrid poplar at 75.48%, willow at 71.11%) as well as the grasses. Miscanthus with its 69.78% volatile matter content has an about 4.59%-point higher value than switchgrass. When looking at the dry feedstock data, one can see that miscanthus has a higher volatile matter content than willow, which can be traced back to the higher moisture content of the as-received miscanthus. For woody biomass, the moisture content after pre-drying is typically 11%. Considering storage, transport and milling of the biomass a moisture content of 11% is very desirable [206]. Herbaceous biomass like switchgrass and miscanthus are often field dried and baled after harvest, which requires a moisture content of 15% [207]. Woody biomass in general has a very low ash content, which is beneficial for gasification and gasifier operation. Depending on which part of the tree is being gasified, the ash content may vary. Bark and leaves have higher ash contents than wood. Switchgrass has a comparably high ash content, which is due to its summer harvest. In summer switchgrass has the maximum biomass yield, after anthesis switchgrass loses about 10 - 20% of its biomass [208], which is mainly in the form of leaves [209]. Leaves in general have a greater ashing behavior and contribute to the higher ash content of switchgrass. Miscanthus, on the other hand, is harvested very late in the season at very low moisture contents after having passed its maximum

biomass yield, which results in a very low ash content. A summary of the proximate analyses of the selected biomass feedstocks can be found in **Table 4-8**.

Table 4-8: Fuel Composition of Selected Biomass Feedstocks

Parameter	Unit	Hybrid Poplar [210]		Willow [211]		Eucalyptus [212]		Switchgrass [213]		Miscanthus [214]		
		ar ¹	dry	ar ¹	dry	ar ¹	dry	ar ¹	dry	ar ¹	dry	
Proximate	Fixed Carbon	wt-%	11.12	12.49	16.84	18.92	11.45	12.86	12.19	14.34	13.94	16.40
	Volatiles	wt-%	75.48	84.81	71.11	79.9	77.08	86.61	65.19	76.69	69.78	82.10
	Moisture	wt-%	11.00	0.00	11.00	0.00	11.00	0.00	15.00	0.00	15.00	0.00
	Ash	wt-%	2.40	2.70	1.05	1.18	0.47	0.53	7.62	8.97	1.27	1.50
Ultimate	C	wt-%	44.66	50.18	44.23	49.7	44.07	49.51	39.68	46.68	41.14	48.40
	H	wt-%	5.39	6.06	5.43	6.10	5.11	5.75	4.95	5.82	5.35	6.30
	O	wt-%	35.98	40.43	37.91	42.59	39.14	43.98	31.77	37.38	36.78	43.27
	N	wt-%	0.53	0.60	0.36	0.40	0.13	0.14	0.65	0.77	0.26	0.30
	S	wt-%	0.02	0.02	0.03	0.03	0.03	0.03	0.16	0.19	0.09	0.10
Trace	Chlorine	mg/kg	89.0	100.0	32.9	37.0	490.8	551.5	1615	1900	1105	1300
	Potassium	mg/kg		3700	[215]	4058		1791	[216]	9500	[217]	4100
	Sodium	mg/kg		40	[215]	37		359	[216]	33	[218]	290
	Phosphorous	mg/kg		320	[219]	1215	[220]	348	[221]	3750	[217]	760

¹ as-received

Table 4-8 shows the ultimate analyses of the various feedstocks along with chlorine, potassium, sodium and phosphorus levels. Carbon content across the feedstocks varies from 49.68-44.66% with switchgrass having the lowest and hybrid poplar having the highest carbon content. The spectrum of hydrogen content fluctuates from 4.95% (switchgrass) to 5.43% (willow). High amounts of carbon and hydrogen are desired for a higher methane output per unit of coal. The oxygen content of the samples listed in the table are not measured but calculated by difference. Oxygen in biomass is commonly found in the form of hydroxyl groups, which will be lost in the process of coalification. The high oxygen content in biomass and low rank coals imparts a high non-specific reactivity but lowers the energy content at the same time [223]. However, coal reactivity does not solely depend on the oxygen content, other factors like degree of order of the carbon structure and hydrogen content are important aspects, as well. The effects of biomass bound nitrogen have been discussed earlier and suggest a rather low nitrogen content as desirable. All listed feedstocks have relatively low nitrogen concentrations compared to coal with switchgrass (0.65%) on the upper end and miscanthus (0.26%) on the lower end. Same is true for the sulfur content of biomass feedstock, which is significantly lower than the sulfur content of coal. Generally speaking, it can be concluded from **Table 4-8** that the grasses, switchgrass and miscanthus, have a slightly higher sulfur content. This should not be a problem regarding their suitability for gasification but has to be considered in downstream syngas processing. The chlorine content in the different feedstocks is typically more critical.

Switchgrass and miscanthus contain a significantly higher chlorine content when compared to the hard woods. Nevertheless, these numbers are still in an acceptable range (1000 - 2000 mg/kg). Most coals have a chlorine content well below 1 wt-% but some coals have chlorine contents up to 2.5 wt-% [8]. Potassium and sodium levels in biomass are significantly higher in biomass feedstocks compared to coal. For gasification no limits for these elements are reported except their concentrations should be taken into consideration when selecting construction materials for the gasifier and downstream of the gasifier. In the case of hydrogasification, potassium and sodium have shown to have beneficial effects upon the gasification behavior and have been investigated as promising catalysts to enhance reaction rates [28]. As discussed before, phosphorus is critical with respect to syngas cooling. Thus, switchgrass with its extremely high phosphorous content of 3750 mg/kg is by far the least favorable feedstock in this category. However, it needs to be mentioned that during plant growth, nutrition concentration and soil have a significant influence on the mineral concentrations in the feedstock and can vary drastically from one location to another. In this regard, values of chlorine, potassium, sodium and phosphorus should be interpreted with caution. Another trace component of importance is mercury. Mercury concentrations in biomass can vary over a wide range, 0.03-0.40 ppmw on an as-received basis with 15% moisture [224]. In this work an average value of 0.215 ppmw on an as-received basis has been assumed.

Differences in the mineral content of the feedstock has a direct influence on the ash composition and the ash melting behavior. Hard woods have an ash fusion temperature similar to hard coals, which is in the range of 1200-1400 °C [8]. Herbaceous biomass' fusion temperatures are significantly lower and can vary a lot. The ash fusion temperature is important for the operation of the gasifier. Advanced low temperature gasification technologies usually operate in a temperature range of 700 – 900 °C in a non-slagging mode. In order to prevent agglomeration, the ash fusion temperature has to be above the gasification temperature. In **Table 4-9** ash composition and fusion temperatures of the selected feedstocks are summarized.

Values for the initial deformation temperatures and melting temperatures are taken from different sources in literature under varying conditions and do not represent the exact values for the ash compositions tabulated. However, they provide estimates and represent typical values for the respective feedstock. The very high amounts of sodium oxide, potassium oxide, phosphorous pentoxide as well as silicon oxide lead to a lower ash melting temperature of biomass, in particular of herbaceous biomass, compared to coal. Because of the extreme aggressiveness of all these salts towards refractory materials, biomass has to be gasified at low temperatures in a non-slagging mode. As it can be seen in **Table 4-9** the ash fusion temperatures for the woods are higher than for the grasses, with switchgrass having the lowest ash melting temperature at 1016 °C. The heating values, higher heating value (HHV) and lower heating value (LHV), of the selected feedstocks are listed in **Table 4-10**.

Table 4-9: Ash Analysis of Selected Biomass Feedstocks

Parameter	Unit	Hybrid Poplar [210]	Willow [211]	Eucalyptus [212]	Switchgrass [213]	Miscanthus [214]
Initial Ash Deformation Temperature	°C	-	1570 ¹	1160 ^{2 [225]}	-	920 ^{1 [226]}
Melting Temperature	°C	1350 ^{2 [227]}	-	1236 ^{2 [225]}	1016 ^{2 [227]}	1240 ^{1 [226]}
Silica, SiO ₂	wt-%	5.90	0.43	17.83	65.18	39.00
Aluminum Oxide, Al ₂ O ₃	wt-%	0.84	0.30	7.87	4.51	1.60
Iron Oxide, Fe ₂ O ₃	wt-%	1.40	0.19	-	2.03	1.10
Titanium Oxide, TiO ₂	wt-%	0.30	0.02	-	0.24	-
Calcium Oxide, CaO	wt-%	49.92	30.78	26.52	5.60	8.60
Magnesium Oxide, MgO	wt-%	18.40	5.14	7.25	3.00	5.90
Sodium Oxide, Na ₂ O	wt-%	0.13	0.24	4.98	0.58	2.20
Potassium Oxide, K ₂ O	wt-%	9.64	26.50	7.20	11.60	27.00
Sulfur Trioxide, SO ₃	wt-%	2.04	3.00	-	0.44	4.90
Carbon Dioxide, CO ₂	wt-%	8.18	-	-	-	0.54
Phosphorous Pentoxide, P ₂ O ₅	wt-%	1.34	11.46	29.11	4.50	6.30

¹ reducing atm. ² oxidizing atm.

Table 4-10: Heating Values of selected Biomass Feedstocks

Parameter	Unit	Hybrid Poplar [210]		Willow [211]		Eucalyptus [212]		Switchgrass [213]		Miscanthus [214]	
		ar ¹	dry	ar ¹	dry	ar ¹	dry	ar ¹	dry	ar ¹	dry
HHV	MJ/kg	16.93	19.02	17.58	19.75	17.1	19.22	15.35	18.06	16.64	19.58
LHV	MJ/kg	15.48	17.7	16.13	18.42	15.72	17.96	13.91	16.79	15.11	18.21

¹as received

All feedstocks have relatively high heating values. The woody biomass has heating values higher than the herbaceous biomass. Willow has the highest heating value with 17.58 MJ/kg. Miscanthus has a heating value almost as high as the woody biomass with 16.64 MJ/kg. Switchgrass, on the other hand, possesses a somewhat lower heating value of 15.35 MJ/kg. The heating values are not only important for gasification, they are also used to determine the energy yields of the crops in order to evaluate which crop can produce the most energy per hectare and year. Thus, the cultivation area can be minimized and the competition between energy crops and food crops can be reduced.

Dry matter yields of hybrid poplar, willow and eucalyptus have been reported as 20 t_{DM}/ha/a [228], 11 t_{DM}/ha/a [228] (unfertilized, with irrigation and fertilizer up to 27 t_{DM}/ha/a [228] in the US) and 35 t_{DM}/ha/a [228]. Maximum yields of up to 32 t/ha/a of air dry biomass for hybrid poplar have been observed in India [229]. For willow, annual dry matter biomass production rates of up to 27 t_{DM}/ha/a have been reported under irrigated and fertilized conditions in the northeastern and midwestern area of the U.S. [228]. In Brazil, dry matter yields of 50 t_{DM}/ha/a have been reached for eucalyptus [227]. Heaton et al. [230] studied the growth of switchgrass and miscanthus under equal conditions in the US. The study reports an

average yield for switchgrass of 10 t_{DM}/ha/a over three years and three locations, northern Illinois, central Illinois and southern Illinois. Miscanthus under the same conditions produced an average biomass yield of 30 t_{DM}/ha/a. In 1990 outstanding yields of up to 35 t_{DM}/ha/a have been reached for switchgrass varieties Alamo and Kanlow under fertilized conditions. The mean across all eight switchgrass varieties in that year was 14 t_{DM}/ha/a [227]. For miscanthus annual dry matter yields as high as 41 t_{DM}/ha/a have been observed in Portugal [227]. An advantage for the cultivation of switchgrass [231] and miscanthus are that they can be cultivated on marginal land with very little application of fertilizer, weed control and pesticides [227]. Furthermore, the grasses can be harvested with commercial harvester and mowers, and baled for storage. The woods, even if harvested after only 3-5 years, require specially designed harvesting machinery, which is not commercially available in large scale yet. After the harvest, switchgrass can be field dried, which reduces the moisture content from originally 65-70% to 15% in about 72 h [232]. Miscanthus is usually harvested after the winter. This reduces the harvestable amount of biomass but at the same time it makes drying unnecessary [233]. Woody biomass is more dense than herbaceous biomass and requires more time for drying. Over months, the wood has to be stored and later actively dried, which demands additional resources [206]. Additionally, switchgrass and miscanthus are harvested annually which reduces economic and supply risks of losing the cultivation due to insect plaques and/or diseases. A summary of these points is provided in **Table 4-11**.

Eucalyptus has the highest energy yield among the selected crops; however, it can only be grown in the southern regions of the US. Although woody biomass, such as eucalyptus, has a high bulk density, which positively affects transportation and storage costs, this local limitation is inhibiting widespread adoption of biomass in the U.S. (note the bulk densities presented are bulk densities of chopped biomass and is a function of the particle size. Thus, the numbers presented should be treated with caution. Nevertheless, a significant difference can be observed between woody biomass and grass). The long growing cycles of woody biomass (which makes it more susceptible to droughts and pests) and the drying requirements make it less attractive compared to the grasses. Furthermore, the grasses can be grown on marginal land with very little to no fertilizer, pesticide and herbicide use, while at the same time reaching exceptional energy yields. Thus, it can be concluded that due to reasons above, switchgrass and miscanthus are strongly favored as feedstock. Comparing switchgrass with miscanthus; miscanthus is the preferred choice for biomass cultivation for energy applications like gasification. Miscanthus has a higher content of volatile matter and a lower ash content, which also has a higher ash melting point. The high heating value together with a high annual dry matter yield results in a 325% higher energy yield of miscanthus. At the same time, the water use efficiency of miscanthus is more than twice as high as the water use efficiency of switchgrass. Another important environmental advantage of miscanthus is that it is grown from rhizomes. Switchgrass, on the other hand, is an invasive grass, which can have serious impacts on ecosystems.

Table 4-11: Additional Information on Selected Biomass Feedstocks

		Hybrid Poplar	Willow	Eucalyptus	Switchgrass	Miscanthus
DM Yield in U.S.	t _{DM} /ha/a	20 (32) ¹	11 (27) ¹	35 (50) ¹	10 (35) ¹	30 (41) ¹
Energy Yield	GJ/ha/a	380.4	217.3	672.7	180.6	587.4
Bulk Density	kg/m ³	301 [234]	227 [211]	375 [235]	105 [236]	139 [226]
Water Use Efficiency	kg _{DM} /ha/ mmH ₂ O	-	-	-	10 [237]	20-28 [237]
Soil /Nutrition	-	Sandy soil and loamy clay, good supply water, no fertilizer [227]	Loams, sandy loams moderately drained [238]	Low fertility clay, salt tolerant and well drained, sensitive to low temperature [227]	Marginal land/ little suppl. fertilizer [227]	Marginal land/ little suppl. fertilizer [227]
Pesticide/Herbicide	-	Requires herbicide, pesticide [227]	Typically not needed [238]	Typically not needed	Not needed [227]	Not needed [227]
Harvest	-	3-5 years [228] retrofitted forestry harvesting equipment [228]	3-4 years [238] modified forage harvester [238]	3-8 years [228] retrofitted forestry harvesting equipment [228]	6/12 months [232] traditional hay swathers and balers [232]	12 months [233] Silage-harvester or mower-conditioner [233]
Drying	-	From 47-55% to 11% [239] Store over months self-drying to 30-35% active drying to 11% [206]	From 45-50% to 11% [206] Store over months self-drying to 30-35% active drying to 11% [206]	From 50% to 11% [240] Store over months self-drying to 30-35% active drying to 11% [206]	From 70%/48% to 15% Field drying approx. 72 h [232]	Harvested at 15% [233]
Other	-	-	-	-	invasive	rhizomes

¹ reported maximum value for annual yield globally

4.1.3 Environmental Targets

The environmental targets for this study were considered on a fuel-specific and plant size basis. These include current emission regulations and emission reduction trends. Particulate matter (PM), sulfur dioxide (SO₂) and nitric oxides (NO_x) emission limits are regulated by the Environmental Protection Agency in 40 CFR CH. I, §40 subpart Da.

Based on the CFR 2015 title 40 volume 7 part 60 subpart Da, emission limits for new energy generating units combusting a fossil fuel at a rate greater than 73 MW_t cannot exceed the following values; for PM 0.090 lb/MWh gross energy output or 0.097 MWh lb net energy output [241], for SO₂ 1.0 lb/MWh gross energy output or 1.2 lb/MWh net energy output [241], for NO_x 0.70 lb/MWh gross energy output or 0.76 lb/MWh net energy output, as well as 0.5 lb/MWh heat input for gaseous coal derived fuels and 0.2 lb/MWh for all other gaseous fuels [241]. According to the EPA Federal Register volume 71, No. 111 (Friday, June 9, 2006) mercury emission limits are constraint to 20 x 10⁻⁶ lb/MWh gross energy output for IGCC power plants, producing more than 25 MW electricity for sale [242]. Since there are no regulations for IGFC power plant emission values, IGCC emission limits will be used as reference values for this study.

Regulatory implementations for CO₂ emissions have not been realized, yet. However, it is foreseeable that in near future such regulations will be imposed due to the contribution of CO₂ to global warming. Because of their small size, it will be unlikely that 10 MW plants will be subject to such a regulation. Currently it is not clear if there will be specifications for biomass derived fuels such as biogas and synthesis gas from biomass (negative emission plants). Thus, for biomass derived gaseous fuels the goal is to find an acceptable compromise between CO₂ capture and efficiency that allows economic operation of the negative emission plants. In several IGCC and NGCC studies conducted by the U.S. DoE, the carbon capture target was set to 90%. However, that would penalize more efficient power plants and a more fundamental approach based upon CO₂ emissions per MWh should be used instead. Based on the work conducted by the U.S. DoE on IGCCs utilizing bituminous coals, 90% capture typically translates to values smaller than 100 kg/MWh (225 lb/MWh) [34] and for subbituminous coals to values smaller than 113 kg/MWh (250 lb/MWh) [33]. For NGCCs, the CO₂ emission footprint with 90% carbon capture is 39 kg/MWh (85 lb/MWh) [34].

One major goal of this study is to investigate efficient methods of reducing water usage of power plants. No guidelines with respect to reduction of water usage have been established or proposed as of today. Goal of this study will be to identify setups that minimize water usage and maximize economic performance.

Table 4-12: Environmental Targets and Emission Limits

Plant Size	Carbon Capture	PM lb/MWh	SO ₂ lb/MWh	NO _x lb/MWh	Hg lb/MWh	CO ₂ lb/MWh
10 MW Natural Gas	No	-	-	-	-	-
10 MW Biogas	No	-	-	-	-	-
50 MW Natural Gas	No	0.090 ¹ /0.097 ²	1.0 ¹ /1.2 ²	0.070 ¹ /0.076 ²	-	-
50 MW Natural Gas	Yes	0.090 ¹ /0.097 ²	1.0 ¹ /1.2 ²	0.070 ¹ /0.076 ²	-	85
100 MW Natural Gas	Yes	0.090 ¹ /0.097 ²	1.0 ¹ /1.2 ²	0.070 ¹ /0.076 ²	-	85
100 MW Coal	Yes	0.090 ¹ /0.097 ²	1.0 ¹ /1.2 ²	0.070 ¹ /0.076 ²	20·10 ⁻⁶ ¹	250
100 MW Biomass	Yes	0.090 ¹ /0.097 ²	1.0 ¹ /1.2 ²	0.070 ¹ /0.076 ²	20·10 ⁻⁶ ¹	< 0

¹ gross power output, ² net power output

4.2 Process Description and Modeling

4.2.1 SOFC-GT Hybrid Plant Common Process Areas

Solid Oxide Fuel Cell

The SOFC model is based upon the Nernst Equation for the reaction between hydrogen and oxygen. Although CO reacts electrochemically in high temperature fuel cells, its reaction rate is slow compared to the water gas shift reaction and the subsequent electrochemical oxidation of the produced hydrogen.

$$E_{\text{Nernst}} = \sum E_{\text{Half Reactions}}^0 + \frac{\Delta \hat{s}_{\text{rxn}}}{nF} (T - T_0) + \frac{RT}{nF} \ln \left(\frac{\prod a_{\text{Products}}^{v_i}}{\prod a_{\text{Reactants}}^{v_i}} \right) \quad (\text{E 4-1})$$

E is the cell potential, in V,

$\hat{s}_{\text{rxn}}$ is the reaction entropy, in J/mol/K,

T is the temperature, in K,

R is the universal gas constant, 8.314 J/mol/K,

n is the number of electrons participation in the electrochemical reaction,

F is the Faraday constant, 96,485 C/mol,

a is the activity, and

v_i its corresponding stoichiometric coefficient.

The cell potential E^0 is determined at ambient pressure and temperature and then corrected for pressure and temperature effects. After the correction of the ideal cell potential to the operating conditions the three major losses present during fuel cell operation are subtracted from the ideal potential in order to obtain the real operating voltage of the cell.

$$V_{\text{SOFC}} = E_{\text{Nernst}} - \eta_{\text{act}} - \eta_{\text{ohmic}} - \eta_{\text{conc}} \quad (\text{E 4-2})$$

V_{SOFC} is the operating voltage of the fuel cell, in V,

η_{act} is the activation loss, in V,

η_{ohmic} is the ohmic loss, in V and

η_{conc} is the concentration loss, in V.

Activation losses are due to energy that has to be invested to overcome the activation barrier of the electrochemical reaction at the catalyst interface. These effects are dominant at low current densities and become less pronounced at higher temperatures. Activation losses are present at anode as well as cathode.

$$\eta_{\text{act}} = \eta_{\text{act}}^{\text{an}}(j) + \eta_{\text{act}}^{\text{cath}}(j) \quad (\text{E 4-3})$$

Where,

$$\eta_{\text{act}}^i(j) = \frac{RT}{\alpha nF} \sinh^{-1} \left(\frac{j}{2j_{\text{act}}^i} \right) \quad (\text{E 4-4})$$

α is the electron transfer coefficient, dimensionless, and

j is the electric current density, in A/m².

The electric current density is calculated using the Butler-Volmer equation.

$$j = j_0 \left[\exp \left(\frac{\alpha nF \eta_{\text{act}}^i}{RT} \right) - \exp \left(- \frac{(1 - \alpha) nF \eta_{\text{act}}^i}{RT} \right) \right] \quad (\text{E 4-5})$$

The exchange current density j_0 can be expressed as Arrhenius-type equation.

$$j_0^{\text{an}} = \gamma_{\text{an}} \left(\frac{p_{\text{H}_2}}{p_{\text{amb}}} \right) \left(\frac{p_{\text{H}_2\text{O}}}{p_{\text{amb}}} \right) \exp \left(-\frac{E_{\text{A}}^{\text{an}}}{RT} \right) \quad (\text{E 4-6})$$

$$j_0^{\text{cat}} = \gamma_{\text{cat}} \left(\frac{p_{\text{O}_2}}{p_{\text{amb}}} \right)^{0.25} \exp \left(-\frac{E_{\text{A}}^{\text{cat}}}{RT} \right) \quad (\text{E 4-7})$$

γ is the pre-exponential factor, in A/m².

Ohmic losses are due to charge transport inside a medium. In SOFCs, the loss in charge transfer tends to be from oxygen ions moving through the electrolyte; however, also electrons contribute to ohmic losses.

$$\eta_{\text{ohmic}} = iR_{\text{ohmic}} = i(R_{\text{ionic}} + R_{\text{elec}}) = i(R_{\text{PEN}} + R_{\text{IC}}^{\text{an}} + R_{\text{IC}}^{\text{cat}}) \quad (\text{E 4-8})$$

i is the current, in A, and

R is the resistance, in Ω .

Concentration or diffusion polarization is due to mass transfer from the bulk of the flow channel to the triple phase boundary (TPB) and can be separated in anode side and cathode side losses. At low current densities diffusion losses are small but can become significant at high species conversion rates.

$$\eta_{\text{conc}} = \eta_{\text{conc}}^{\text{an}} + \eta_{\text{conc}}^{\text{cath}} = \frac{RT}{2F} \ln \left(\frac{x_{\text{H}_2}^{\text{bulk}} x_{\text{H}_2\text{O}}^{\text{TPB}}}{x_{\text{H}_2\text{O}}^{\text{bulk}} x_{\text{H}_2}^{\text{TPB}}} \right) + \frac{RT}{4F} \ln \left(\frac{x_{\text{O}_2}^{\text{bulk}}}{x_{\text{O}_2}^{\text{TPB}}} \right) \quad (\text{E 4-9})$$

x is the mole fraction of component i , dimensionless.

By relating the diffusive flow of H₂, H₂O, and O₂ to the electric current density j via the Faraday's law and assimilating multicomponent diffusion to binary diffusion where required, the mole fractions of H₂, H₂O, and O₂ at the TPBs are calculated using the following equations:

$$x_{\text{H}_2}^{\text{TPB}} = x_{\text{H}_2}^{\text{bulk}} - \frac{jRT\delta_{\text{an}}}{2Fp_{\text{an}}D_{\text{an}}^{\text{eff}}} \quad (\text{E 4-10})$$

$$x_{\text{H}_2\text{O}}^{\text{TPB}} = x_{\text{H}_2\text{O}}^{\text{bulk}} + \frac{jRT\delta_{\text{an}}}{2Fp_{\text{an}}D_{\text{an}}^{\text{eff}}} \quad (\text{E 4-11})$$

$$x_{\text{O}_2}^{\text{TPB}} = 1 + (x_{\text{O}_2}^{\text{bulk}} - 1) - \frac{jRT\delta_{\text{cat}}}{4Fp_{\text{cat}}D_{\text{cat}}^{\text{eff}}} \quad (\text{E 4-12})$$

δ is the thickness, in m, and

D^{eff} is the effective Diffusivity, in m²/s.

The diffusivities can be calculated via:

$$D_{\text{an}}^{\text{eff}} = \left(\frac{p_{\text{H}_2\text{O}}}{p_{\text{an}}} \right) D_{\text{H}_2}^{\text{eff}} + \left(\frac{p_{\text{H}_2}}{p_{\text{an}}} \right) D_{\text{H}_2\text{O}}^{\text{eff}} \quad (\text{E 4-13})$$

$$D_{\text{cat}}^{\text{eff}} = D_{\text{O}_2}^{\text{eff}} \quad (\text{E 4-14})$$

Since the anode and cathode material is porous and the pore size is on the length scale of the free mean path, bulk diffusion and Knudsen diffusion have to be considered.

$$D^{\text{eff}} = \frac{\varepsilon}{\tau} \left(\frac{1}{D_{12}} + \frac{1}{D_{\text{K},i}} \right)^{-1} \quad (\text{E 4-15})$$

ε is the porosity of the electrode, dimensionless,

τ is the tortuosity of the electrode material, dimensionless,

D_{12} is the binary diffusivity of components 1 and 2, in m^2/s , and

D_{K} is the Knudsen diffusivity, in m^2/s .

The binary diffusivity can be determined using the Fuller equation.

$$D_{12} = \frac{0.00143 T^{1.75}}{p \left(\frac{2}{\frac{1}{M_1} + \frac{1}{M_2}} \right)^{1/2} ((\Sigma v)_1^{1/3} + (\Sigma v)_2^{1/3})^2} \quad (\text{E 4-16})$$

M is the molecular weight, in g/mol , and

Σv is the diffusion volume of the species, dimensionless.

Knudsen diffusivity is determined via the following equation.

$$D_{\text{K},i} = \frac{d_{\text{pore}}}{3} \sqrt{\frac{8RT}{\pi M_i}} \quad (\text{E 4-17})$$

d_{pore} is the average diameter of the material's pores, in meter.

Next to the fundamental fuel cell processes, SOFCs support internal reforming. Thus, simultaneously to the above-described processes, methane steam reforming and water gas shifting are executed. While the water gas shift reaction is sufficiently fast to always be in equilibrium, the methane reforming reaction is kinetically controlled and modeled using power law kinetics as described by Achenbach and Riensche [243], which considers the forward as well as the backward reactions. The reforming reactions of higher hydrocarbons (in the SOFC) are considered to be irreversible and are described in the pre-reformer section.



The model is setup as a quasi 2D finite volume model solving mass, energy and species balances for every individual control volume and using the output of the control volume as input for the downstream control volumes, as shown in **Figure 4-1**.

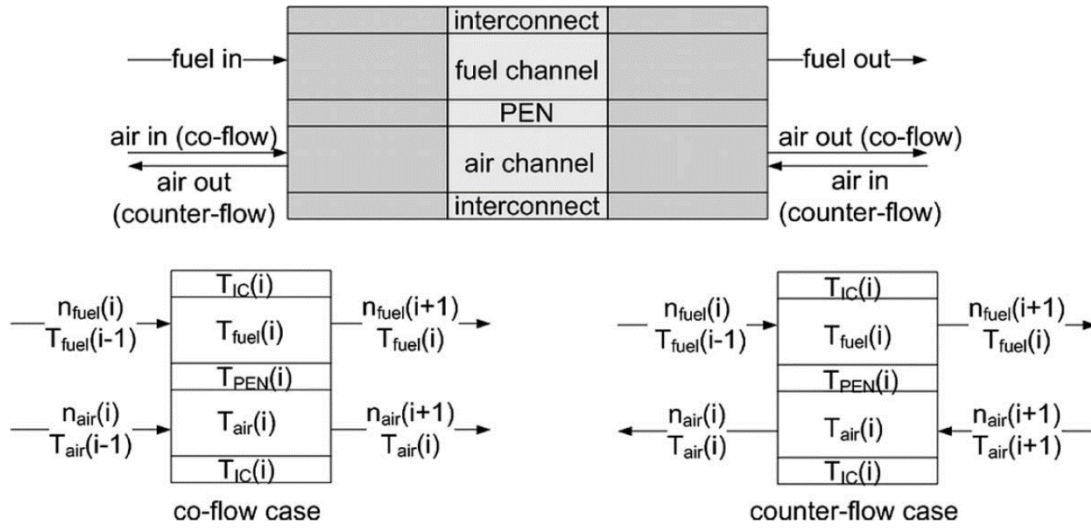


Figure 4-1: Discretization and calculation domain of the SOFC model for co-flow and counter-flow operation.

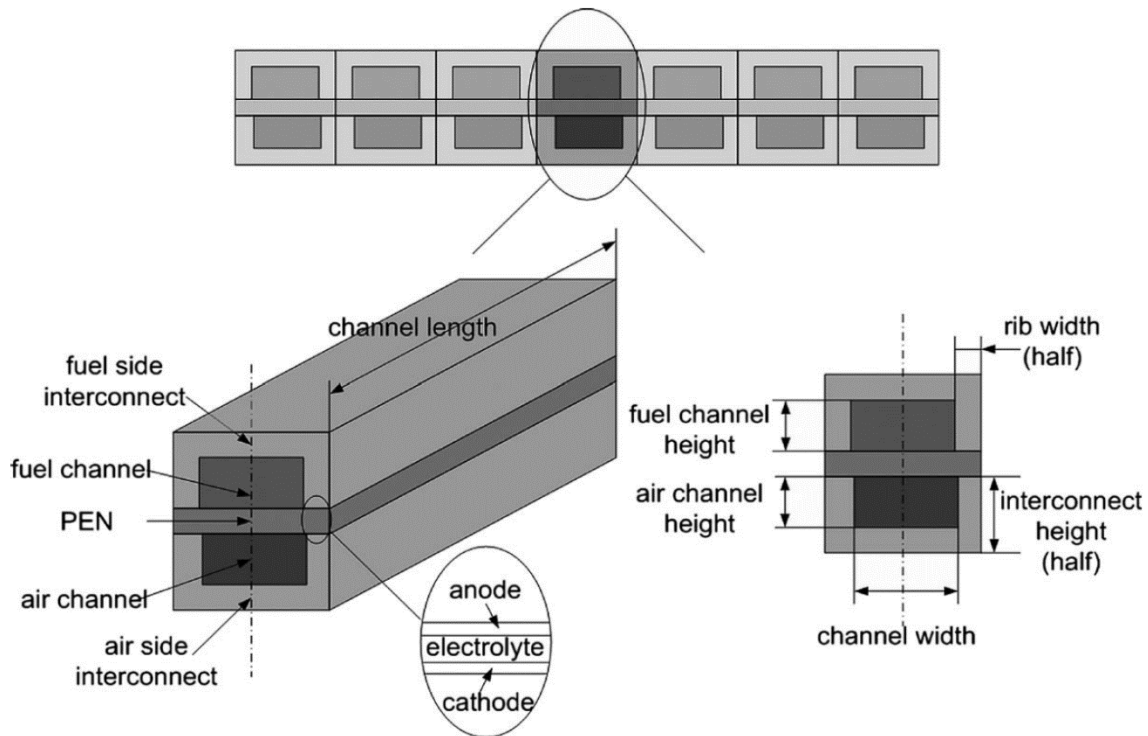


Figure 4-2: Geometrical configuration of the SOFC model domain.

An overview of the geometrical arrangement and input variables can be seen in **Figure 4-2**. A summary of the SOFC model key parameters is provided in **Table 4-13**. The electrolyte material is yttria stabilized zirconia which is typically operated in a temperature range from 600-1000 °C and cell materials are not changed through this study. The interconnect of this SOFC is made of steel. Important assumptions regarding the SOFC model are summarized below. Details, applicability and additional assumptions can be found in [244]:

- Homogeneous compound distribution in finite control volume.
- Electrochemical oxidation of H₂ occurs at the anode-electrolyte interface, with the reaction kinetics controlled by the local PEN temperature.
- 100% of the surface area under the interconnect rib is active for H₂ oxidation but inactive for CH₄ reformation.
- The Peclet number is large: thus, it is reasonable to neglect axial diffusion effects (thermal and mass diffusion) in the gas phases.

Table 4-13: Summary of Internal Reforming Solid Oxide Fuel Cell Parameters

Geometry	Value	Unit
Fuel/Air Channel Length	0.20	m
Fuel/Air Channel Width	0.002	m
Rib Width	0.003	m
Number of Channels	50	-
Fuel Channel Height	0.001	m
Air Channel Height	0.002	m
Interconnect Height	1.5	mm
Anode Thickness	400	μm
Electrolyte Thickness	8	μm
Cathode Thickness	20	μm
Max. Edge Length (Manufacturing Limit)	0.5	m
Max. Cell Area (Manufacturing Limit)	0.25	m ²
Operation Parameters		
Max. Operating Pressure	20	bar
Pressure Drop Anode (at 7.5 bar)	0.26	bar
Pressure Drop Cathode (at 7.5 bar)	0.35	bar
Inlet Temperature Anode	973.15	K
Inlet Temperature Cathode	973.15	K
Max. Overall Temperature Difference	150	K
Max. Local Temperature Gradient	15	K/cm
Heat Loss to Ambient	0.5	%
Material Properties	Value	Unit
Thermal Conductivity PEN Structure	2	W/m/K
Thermal Conductivity Interconnect	25	W/m/K
Emissivity Metal	0.3	-
Emissivity Cermet	0.85	-

Internal Reforming Reactions		Value	Unit
Pre-exponential Factor Methane Reforming		4274	mol/s/m ² /bar
Reaction Order Methane		1	-
Reaction Order Vapor		0	-
Activation Energy		82	kJ/mol
Steam Methane Reforming Kinetics	$k_0 p_{\text{CH}_4}^m p_{\text{H}_2\text{O}}^n \exp\left(\frac{-E_A}{R T_{\text{PEN}}}\right) A_S (1 - \beta)$		mol/s
Water Gas Shift Reaction is in Equilibrium			
Water Gas Shift Equilibrium Constant	$\exp\left(\frac{4276}{T_{\text{Fuel}}} - 3.961\right)$		-
Activation Polarization		Value	Unit
Pre-exponential Factor Anode		$5.5 \cdot 10^8$	A/m ²
Activation Energy Anode		55	kJ/mol
Pre-exponential Factor Cathode		$7.0 \cdot 10^8$	A/m ²
Activation Energy Cathode		110	kJ/mol
Transfer Coefficient		0.5	-
Ohmic Polarization		Value	Unit
Specific Resistivity Anode	$\left[\frac{95 \cdot 10^6}{T_{\text{PEN}}} \exp\left(-\frac{1150}{T_{\text{PEN}}}\right)\right]^{-1}$		Ωm
Specific Resistivity Electrolyte	$\left[\frac{42 \cdot 10^6}{T_{\text{PEN}}} \exp\left(-\frac{1200}{T_{\text{PEN}}}\right)\right]^{-1}$		Ωm
Specific Resistivity Cathode	$\left[3.34 \cdot 10^4 \exp\left(-\frac{10,300}{T_{\text{PEN}}}\right)\right]^{-1}$		Ωm
Area Specific Resistivity Interconnect Oxide Scale	$5 \cdot 10^{-7}$		Ωm^2
Diffusion Polarization		Value	Unit
Porosity Anode		0.5	-
Tortuosity Anode		3.0	-
Pore Diameter Anode		1	μm
Porosity Cathode		0.5	-
Tortuosity Cathode		3.0	-
Pore Diameter Cathode		1	μm

Gas Turbine

The performance of the GTs is estimated using vendor specifications for current gas turbine offerings. Gas turbines are extremely complex equipment and are historically engineered for the use of natural gas. The market for other fuels with low heating value, e.g., syngas or biogas, is small and it is not believed that gas turbine vendors will re-design their engines for such applications in the short. Thus, current GTs will have to be operated in off-design mode, which typically leads to lower efficiencies. In off-design operation the GT can either operate at grid frequency (constant speed) or at variable speed. If operated at constant speed the generated electricity can be fed into the grid without frequency adjustment. For variable speed the generated electricity will have to be synchronized with the grid before it can be exported. This can be achieved via AC/DC converters and DC/AC inverters; however, those conversions are associated with losses. That is why gas turbines are usually operated at constant speed. Although these conversion losses reduce the power output of a system operated at variable speed, it might be justifiable if the performance gains of the GT outweigh the AC/DC and DC/AC conversion losses. Part-load operation is another form of

off-design operation and it has been shown in literature [93]–[95] that variable speed operation is substantially more efficient than constant speed operation for the part-load operation of SOFC-GT hybrid systems.

The GT is modeled as a non-intercooled axial or radial compressor (as specified by the user), a combustor and an expander. The model equations are described in accordance with the description in **Figure 4-3**. For the model it is assumed that the gas turbine has a fixed geometry and that the modeled GT has the same geometry as in the calibration case. The performance of the gas turbine compressor, combustor and expander are calibrated with open access vendor data.

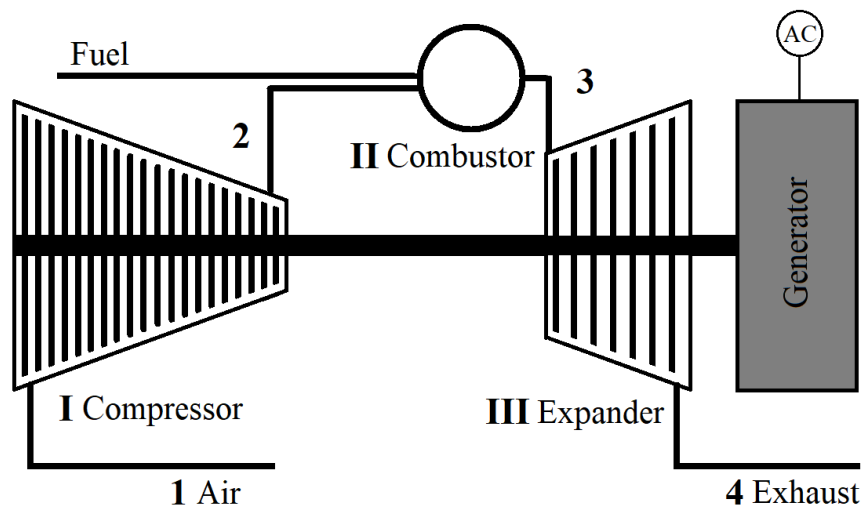


Figure 4-3: Stream description of the gas turbine model.

Three different operating modes of the GT are considered. The first mode considers a reference gas turbine's technology status, e.g., compressor efficiency, expander efficiency, combustor pressure drop, maximum turbine inlet temperature, and applies those parameters to the operating condition in the simulation. It represents a transfer of technology characteristics from the on-design performance of a reference GT using natural gas to a customized GT for an application using the fuel composition from the simulation. The reference conditions refer to the GT manufacturer's on-design conditions. The simulation refers to the operating conditions encountered in the investigated design. The second mode uses a generalized compressor map and expander map for the performance calculations when operating the GT at constant spool speed. The third model also considers compressor and expander maps, however, operates at variable spool speed allowing for adjustments of the compressor discharge pressure. Additionally, mode two and three have variable inlet guide vane control to optimize GT operation.

Process Description and Modeling

The calculations involving the combustor are identical for all three operating modes and are expressed by Equation (E 4-18) to Equation (E 4-20). The first equation determines the pressure drop of the combustor.

$$\Delta p_{II} = -p_2 \cdot \Delta p_{II, \text{Ref}} \quad (\text{E 4-18})$$

Δp_{II} is the pressure drop inside the combustor, in bar,

p_2 is the inlet pressure of the combustor, in bar, and

$\Delta p_{II, \text{Ref}}$ is the pressure drop of the reference combustor, in fraction of inlet pressure.

It may be noted that if the combustor geometry is held constant, the pressure drop will increase with increasing velocity (higher flow rate, e.g., due to different fuel). However, it is common practice today to redesign the combustor according to the fuel used and it is assumed that a similar fractional pressure drop can be obtained by this combustor modification approach.

In a similar manner, the heat loss of the combustor is estimated by Equation (E 4-19).

$$\Delta Q_{II} = -\dot{H}_2^0 \cdot \Delta Q_{II, \text{Ref}} \quad (\text{E 4-19})$$

ΔQ_{II} is the heat loss of the combustor, in kW,

$\dot{H}_2^0$ is the flow of standard enthalpy of combustion based on the fuel's LHV, in kW, and

$\Delta Q_{II, \text{Ref}}$ is the heat loss of the reference combustor, in fraction of LHV.

In order to correct for differences in the gas composition, the firing temperature is adjusted to maintain the same blade metal temperature as in the reference case [245]. This equation corrects the firing temperature based on the H₂O and CO₂ content in the syngas. Water and CO₂ have higher heat capacities than N₂ and O₂, which impacts the heat transfer to the expander blades [246].

$$TIT = TIT_{\text{Ref}} + \frac{644.23}{1.8} \left(y_{\text{H}_2\text{O}+\text{CO}_2, 3, \text{Ref}} - y_{\text{H}_2\text{O}+\text{CO}_2, 3} \right) \quad (\text{E 4-20})$$

TIT is the expander inlet temperature (T_3), in K,

TIT_{Ref} is the expander inlet temperature (T_3) in the reference case, in K,

y_i is the mole fraction of component i entering the expander, dimensionless, and

$y_{i, \text{Ref}}$ is the mole fraction of component i entering the expander in the reference case, dimensionless.

Mode One. This mode models the GT as a custom GT provided by a manufacturer for this specific application. Compared to current GTs built for natural gas applications, which can act as a reference GT for the technology level of this GT, this mode will require redesign of the compressor to adjust for the desired pressure ratio, redesign of the combustor to extract air from the compressor and utilize the low heating value anode off-gas and redesign of the expander to adjust for mass flow and pressure ratio changes. For the first operational mode the efficiencies of compressor and expander (mechanical and thermodynamic)

are set equal to the reference GT that is chosen for representing the technology level. Furthermore, the compressor discharge pressure is set to the desired pressure value. This ultimately means that the compressor flow, compressor discharge pressure, fuel flow and heating value as well as expander flow encountered in this application are identical to the GT manufacturer's custom design conditions.

Thermo-economic evaluation of a system requires specifying the size of the GT, which is set by the compressor size of the reference GT; although, it is a custom build GT. Typically, the smaller an engine, the lower the efficiency. This is true for the compressor as well as for the expander. Thus, the technology level is also a function of GT size. Using the compressor as reference for the scale is a more conservative approach for the utilization of low energy content fuels (lower than the reference fuel, in most cases natural gas), such as the anode off-gas of a SOFC-GT hybrid. Utilization of a low energy content fuel will lead to an increased fuel demand to reach the same firing temperature and ultimately lead to a higher mass flow through the expander. Furthermore, designing a compressor from a "clean sheet" is a major undertaking and it is more reasonable to expect that the manufacturer is taking an existing design and adding or removing back-end stages for the required pressure ratio as long as the Mach number does not constraint this design approach. The mass flow through the compressor can be calculated via ideal gas law assuming constant suction volume of the compressor.

$$\dot{m}_{1,Scale} = \frac{\dot{m}_{1,Ref} T_{1,Ref}}{p_{1,Ref} \bar{M}_{1,Ref}} \cdot \frac{p_{1,Sim} \bar{M}_{1,Sim}}{T_{1,Sim}} \quad (E\ 4-21)$$

$\dot{m}_1$ is the mass flow at compressor inlet, in kg/s,

T_1 is the temperature at compressor inlet, in K,

p_1 is the pressure at compressor inlet, in bar, and

$\bar{M}_1$ is the average molecular weight of the gas at compressor inlet, in g/mol.

The scaling of the simulation values (subscript Sim) to the actual engine size is realized by relating the simulation values to the calibration values (subscript Ref). The scaling factor for maximum GT utilization is needed for the thermo-economic evaluation and can be calculated via Equation (E 4-22).

$$F_{Scale} = \frac{\dot{m}_{1,Scale}}{\dot{m}_{1,Sim}} \quad (E\ 4-22)$$

$\dot{m}_{1,Scale}$ is the mass flow at compressor inlet calculated in the model, in kg/s,

$\dot{m}_{1,Sim}$ is the mass flow at compressor inlet in the simulation, in kg/s,

Typically, GT manufacturers provide a generator set together with the gas turbine, which are matched in a manner that allows the gas turbine to operate over a wide range of ambient conditions. Thereby, ambient temperatures between -10 °C and 0 °C are considered to represent the operating condition constituting the shaft limit and generator sizing as they result in the highest power output. Consequently, lower ambient

conditions are desired in power applications to maximize the economic potential of the gas turbine and maximize the scale factor.

Map Interpolation. Modes two and three utilize compressor and expander maps. The gas turbine selected in this work is a Dresser Rand KG2-3G EF, which is rated for 1.83 MW under ISO conditions using CH₄ as fuel [247]. With its design point compression ratio of 7:1 and mass flow rate of 9.48 kg/s [247] it is suitable for the integration into SOFC-GT hybrid systems of approximately 10 MW. The EF (externally fired) version of this GT will allow for an easier GT integration as compared to GTs where the combustor is integrated into the GT casing. Furthermore, Dresser Rand offers a Power Oxidizer with this model for utilization of low heating value gases while achieving ultra-low emissions. Calibration data for the Dresser Rand KG2-3G EF are provided in **Table 4-14**.

Table 4-14: Summary of Gas Turbine Parameters

GT Operational Parameters (Design Point)	Value	Unit
Compressor Polytropic Efficiency	84.35	%
Expander Polytropic Efficiency	84.36	%
Compressor Mechanical Efficiency	98.03	%
Expander Mechanical Efficiency	98.03	%
Turbine Inlet Temperature	943.3	°C
Combustor Pressure Drop	3	%
Combustor Heat Loss	1	%-LHV

This GT uses radial compressor and expander technology in order to accommodate for this fuel flexible operation. Single stage radial compressor [248] and expander [249] maps have been used to develop the generalized maps shown in **Figure 4-4** and **Figure 4-5**.

Characteristic for radial compressors is the curvature of the on-design point operating speed line and relatively wide range of mass flows on this speed line. Large engines with axial compressors typically exhibit more vertical speed lines.

For map interpolation a linear approach is used. First, the interpolation algorithm locates the closest two speed lines relative to the operating speed line. For every datapoint on the lower speed line, the surface normal is determined. Then the intercept of the surface normal with the higher speed line is calculated and the two closest data points, left and right, are identified. Thereafter, values for mass flow, pressure ratio and efficiency are linearly interpolated between these two points to identify the properties at the intercept. These properties at this new point on the higher speed line are now used to determine mass flow, pressure ratio and efficiency via linear interpolation at the actual operating speed. The first point on the speed line (surge limit) and the last point are directly interpolated. Since this is done for every point on the speed line, a new matrix of values with the actual operating speed is created. The same procedure is used for compressor and turbine maps.

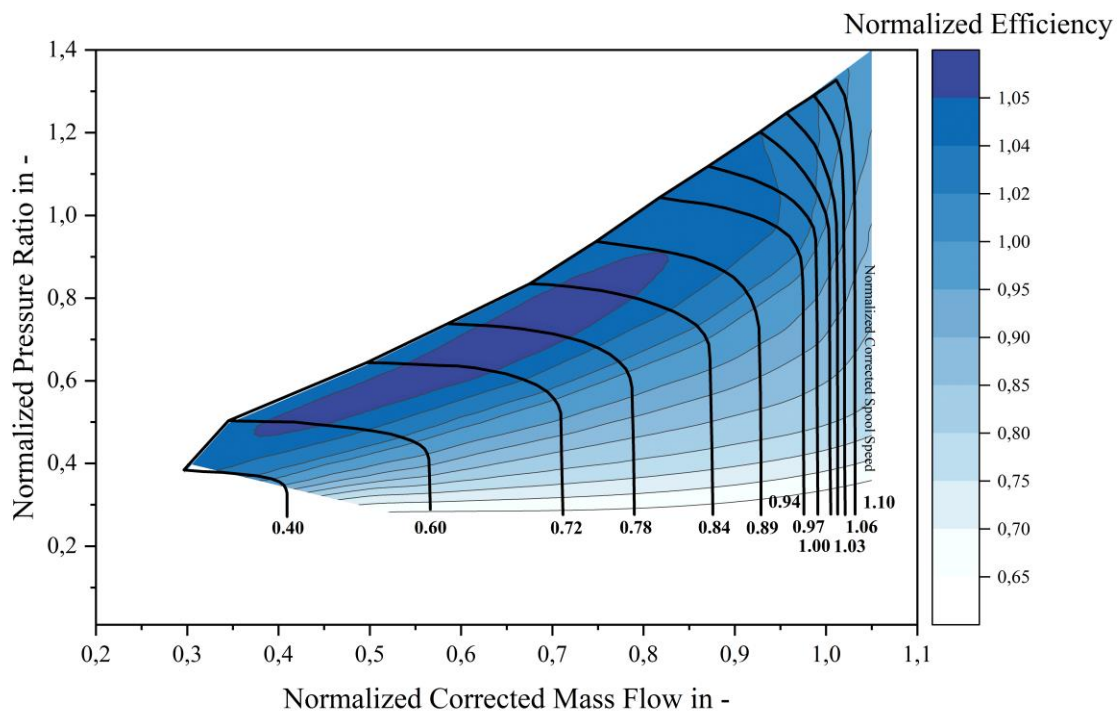


Figure 4-4: Compressor map of a gas turbine employing a single stage radial compressor.

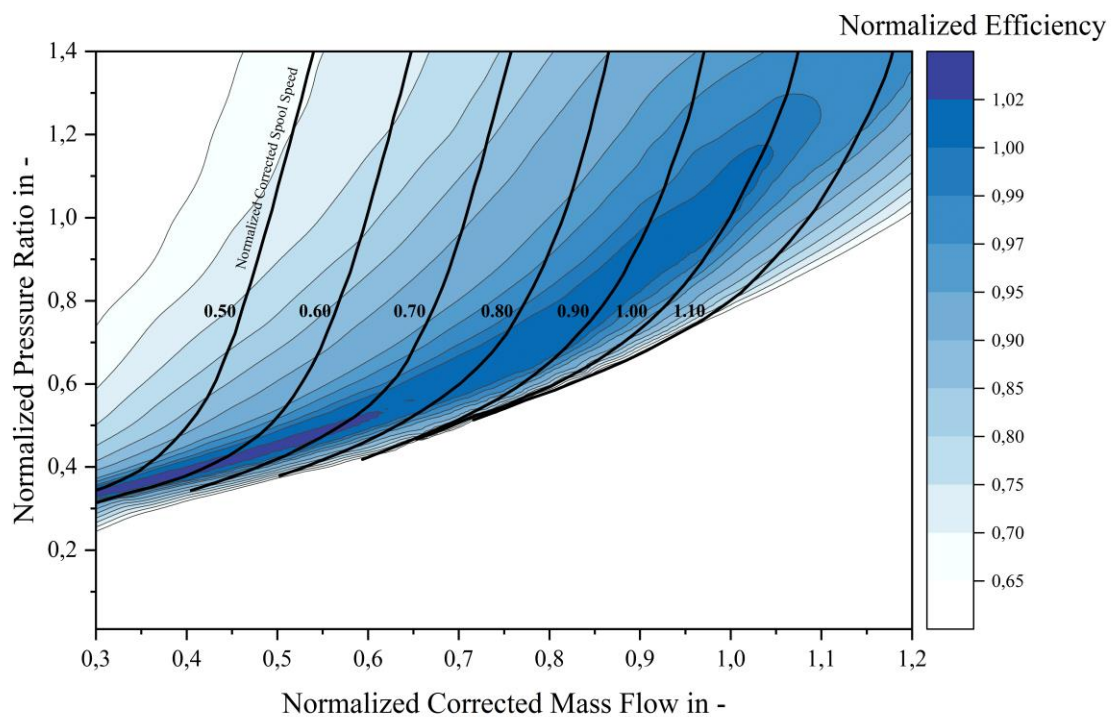


Figure 4-5: Turbine expander map of a gas turbine employing single stage radial turbine expander.

Process Description and Modeling

Variable Inlet Guide Vanes. Variable inlet guide vanes (VIGV) regulate the air flow to the first compressor rotor and can be used to optimize the GT performance in off-design operation. Changing the inlet guide vane angle has not just implication for the mass flow but also the pressure ratio and compressor efficiency. According to Kim and Hwang [250], changes to the compressor map due to VIGV can be expressed as follows:

$$\frac{\dot{m}_{1,\text{VIGV}} \cdot \sqrt{T_{1,\text{VIGV}}}}{p_{1,\text{VIGV}}} = a \cdot \frac{\dot{m}_1 \cdot \sqrt{T_1}}{p_1}, \quad (\text{E 4-23})$$

$$\frac{p_{2,\text{VIGV}}}{p_{1,\text{VIGV}}} = b \cdot \frac{p_2}{p_1}, \quad (\text{E 4-24})$$

$$\eta_{\text{I,VIGV}} = c \cdot \eta_{\text{I}}, \quad (\text{E 4-25})$$

$\dot{m}_1$ is the mass flow at compressor inlet, in kg/s,
 T_1 is the temperature at compressor inlet, in K,
 p_1 is the pressure at compressor inlet, in bar,
 p_2 is the pressure at compressor outlet, in bar, and
 a, b, c are scaling parameters, dimensionless.

The parameters a , b and c are a functions of the VIGV angle and have been derived from [251] for radial compressors. The used correlations are:

$$a_{\text{rad}} = -9.14 \cdot 10^{-5} \cdot \Delta\alpha_{\text{VIGV}}^2 + 1, \quad (\text{E 4-26})$$

$$b_{\text{rad}} = -8.1781 \cdot 10^{-4} \cdot \Delta\alpha_{\text{VIGV}} + 1, \quad (\text{E 4-27})$$

$$c_{\text{rad}} = -3.804 \cdot 10^{-5} \cdot (\Delta\alpha_{\text{VIGV}} - 24.08)^2 + 1.022057, \quad (\text{E 4-28})$$

$\Delta\alpha_{\text{VIGV}}$ is the change in VIGV angle, in °degrees.

A positive VIGV angle indicates closing VIGV and a negative number indicates more open VIGVs. The radial compressor data have been derived for an angle range from 0° to +60°.

Mode Two. The second mode is the off-design operation of a GT at constant speed utilizing generalized compressor and expander maps for determining the performance. The maps depict correlations between the corrected normalized mass flow ($\dot{m}_{\text{corr,norm}}$), normalized pressure ratio ($\pi_{\text{corr,norm}}$), corrected normalized speed ($N_{\text{corr,norm}}$) and normalized efficiency (η_{norm}). The corrected mass flow, pressure ratio and corrected speed can be calculated via:

$$\dot{m}_{j,\text{corr}} = \frac{\dot{m}_j \cdot \sqrt{T_j}}{p_j} \quad (\text{E 4-29})$$

$$\pi_{\text{corr}} = \frac{p_k}{p_j} \quad (\text{E 4-30})$$

$$N_{\text{corr}} = \frac{N}{\sqrt{T_j/288.15 \text{ K}}} \quad (\text{E 4-31})$$

$\dot{m}_j$ is the mass flow at position $j=1$ for compressor, $j=3$ for turbine (see **Figure 4-3**), in kg/s,
 T_j is the temperature at position $j=1$ for compressor, $j=3$ for turbine, in K,
 p_j is the pressure at position $j=1$ for compressor, $j=3$ for turbine, in bar,
 p_k is the pressure at position $k=2$ for compressor, $k=4$ for turbine, in bar, and
 N is the operating spool speed of the GT, in rpm.

The normalized values of these quantities are obtained by dividing the value by the corresponding reference quantity from the calibration GT. In order to obtain the GT operational point, an iterative procedure between the compressor map and the expander map is needed. The normalized corrected spool speed of the compressor under constant spool speed operation ($N_{\text{Sim}} = N_{\text{Ref}}$) is calculated via:

$$N_{\text{I,corr, norm}} = \frac{\frac{N_{\text{Ref}}}{\sqrt{T_{1,\text{Sim}}/288.15 \text{ K}}}}{\frac{N_{\text{Ref}}}{\sqrt{T_{1,\text{Ref}}/288.15 \text{ K}}}} \quad (\text{E 4-32})$$

Based on the normalized corrected speed, data from the map are interpolated between the two closest speed lines tabulated and values for normalized pressure ratio, normalized corrected mass flow and normalized efficiency are calculated and saved. The iteration is initialized with an initial guess for the normalized compressor pressure ratio $\pi_{\text{I,norm}}$. A suitable initial guess could be the pressure ratio of the reference GT, $\pi_{\text{I,norm,Ref}}$. Then the normalized corrected mass flow and normalized efficiency are determined from the map based on the normalized pressure ratio and normalized corrected spool speed. Thereafter, the mass flow is scaled to maximize the GT capacity utilization and to determine the expander mass flow.

$$F_{\text{Scale}} = \frac{\dot{m}_{1,\text{corr,norm}} \dot{m}_{1,\text{Ref}} \sqrt{T_{1,\text{Ref}} p_{1,\text{Sim}}}}{p_{1,\text{Ref}} \sqrt{T_{1,\text{Sim}}} \dot{m}_{1,\text{Sim}}} \quad (\text{E 4-33})$$

The absolute compressor mass flow scaled to max. GT capacity utilization is then calculated by:

$$\dot{m}_{1,\text{Scale}} = F_{\text{Scale}} \cdot \dot{m}_{1,\text{Sim}} \quad (\text{E 4-34})$$

To maintain the same operating condition as in the simulation, the ratio of compressor mass flow to expander mass flow, Φ_{CER} , is held constant (equivalent to constant air to fuel ratio). Then the absolute expander inlet mass flow is calculated as:

Process Description and Modeling

$$\dot{m}_{3,\text{Scale}} = \frac{\dot{m}_{1,\text{Scale}}}{\Phi_{\text{CER}}} \quad (\text{E 4-35})$$

Similar to the normalized corrected compressor mass flow, the normalized corrected expander mass flow can be expressed by the following equation.

$$\dot{m}_{3,\text{Scale,corr,norm}} = \frac{\dot{m}_{3,\text{Scale}} \sqrt{T_{3,\text{Sim}}} p_{3,\text{Ref}}}{p_{3,\text{Sim}} \dot{m}_{3,\text{Ref}} \sqrt{T_{3,\text{Ref}}}} \quad (\text{E 4-36})$$

The normalized corrected turbine expander speed is calculated using Equation (E 4-37),

$$N_{\text{III,corr,norm}} = \frac{\frac{N_{\text{Ref}}}{\sqrt{T_{3,\text{Sim}}/288.15 \text{ K}}}}{\frac{N_{\text{Ref}}}{\sqrt{T_{3,\text{Ref}}/288.15 \text{ K}}}} \quad (\text{E 4-37})$$

Using the information about the operating condition of the turbine expander, $\dot{m}_{3,\text{Scale,corr,norm}}$ and $N_{\text{III,corr,norm}}$, the normalized expander pressure ratio can be determined with the turbine expander map. The obtained normalized pressure ratio from the turbine expander follows the expression in Equation (E 4-38).

$$\pi_{\text{III,corr,norm}} = \frac{p_{3,\text{Sim}}^* p_{4,\text{Ref}}}{p_{4,\text{Sim}} p_{3,\text{Ref}}} \quad (\text{E 4-38})$$

This equation can be rearranged to find the $p_{3,\text{Sim}}^*$. This value can now be used to calculate a new compressor pressure ratio by considering the pressure drop occurring between the compressor and the turbine expander.

$$\pi_{\text{I,corr,norm}}^* = \frac{p_{3,\text{Sim}}^* - \Delta p_{\text{II}}}{p_{1,\text{Sim}}} \quad (\text{E 4-39})$$

If $\pi_{\text{I,corr,norm}}^*$ is different from the initial guess the model is not converged and a new initial guess, e.g. $\pi_{\text{I,corr,norm}}^*$ or a relaxed value between $\pi_{\text{I,corr,norm}}^*$ and the initial guess, will be used until the error is sufficiently small.

Mode Three. Mode three operates very similarly to the second mode, however, instead of solving for the pressure, the operating speed represents the free variable and the pressure, $p_{2,\text{Sim}}$, is set by the user. Thus, in this mode the pressure ratio is fixed to:

$$\pi_{\text{I,corr,norm}} = \frac{p_{2,\text{Sim}} p_{1,\text{Ref}}}{p_{1,\text{Sim}} p_{2,\text{Ref}}} \quad (\text{E 4-40})$$

and the actual operating speed $N_{I,Sim}$ is initially guessed, e.g., the design operating speed, $N_{I,Ref}$, which is then used to determine $N_{I,corr,norm,Ref}$. Thus, the input variables to the map interpolation algorithm are the same as in the second model mode and the solution strategy is identical. Thereafter, Equation (E 4-33), Equation (E 4-34), Equation (E 4-35) and Equation (E 4-36) are solved and the normalized turbine expander mass flow, $\dot{m}_{3, Scale,corr,norm}$, is obtained. Furthermore, the very same variable, $\dot{m}_{3, Scale,corr,norm}$, is determined using the expander map and will be called $\dot{m}_{3, Scale,corr,norm}^*$. A comparison of the two variables will function as convergence criterion.

For the expander map in the third mode, normalized corrected speed and normalized pressure ratio will function as inputs. The normalized turbine operating speed is calculated via Equation (E 4-41).

$$N_{III,corr,norm} = \frac{\frac{N_{I,Sim}}{\sqrt{T_{3,Sim}/288.15 \text{ K}}}}{\frac{N_{Ref}}{\sqrt{T_{3,Ref}/288.15 \text{ K}}}} \quad (\text{E 4-41})$$

Next the turbine expander pressure ratio is calculated by

$$\pi_{III,corr,norm} = \frac{p_{2,Sim} + \Delta p_{II}}{p_{4,Sim}} \quad (\text{E 4-42})$$

Based on these two input variables the normalized corrected mass flow through the turbine expander is determined. If the normalized corrected expander mass flow does not match with the result from Equation (E 4-36) a new guess for the actual operating speed is needed, which can be approached by the following equation.

$$N_{I,Sim}^* = \frac{\dot{m}_{3, Scale,corr,norm}^*}{\dot{m}_{3, Scale,corr,norm}} \cdot N_{I,Sim} \quad (\text{E 4-43})$$

The new guess can be used directly or relaxed using the initial guess, $N_{I,Sim}$. The procedure is repeated until the error in the mass flows is sufficiently small.

Carbon Deposition Equilibrium

Crucial to the operation of the SOFC as well as the pre-reformer is the gas stability of the stream entering the reforming section. Unstable gas mixtures lead to solid carbon deposition onto the catalyst reducing its activity. The required amount of steam or anode off-gas recirculation is determined by an equilibrium code. Predicting whether a gas mixture will form solid carbon under certain conditions of temperature, pressure and chemical composition is not an exact science and requires experimental validation. However, taking thermodynamic equilibrium into consideration, it is possible to identify safe regions where the formation

of solid carbon is impossible. From a thermodynamic equilibrium standpoint, CO decomposes to solid carbon if the temperature is too low. Methane, on the other hand, cracks if the temperature is too high. Based on this information, it is impossible to predict whether the gas mixture is stable, unless it is in equilibrium with other components of the gas phase. Using anode off-gas recirculation makes the interpretation of a steam to carbon ratio, as often used for hydrocarbon reforming, meaningless. A steam to carbon ratio is only meaningful if the carbon to hydrogen ratio of the feedstock is fixed, in the case of CH₄ 1:4, as hydrogen, carbon and oxygen take part in the equilibrium reactions. Thus, it is important in the case of anode off-gas recirculation to consider the entire equilibrium of the gas phase components. For solid carbon formation the following heterogenous reactions are being considered.



The Reaction (R 4-3) is known as reverse hydrogasification. Reaction (R 4-4) is the Boudouard reaction and Reaction (R 4-5) is the water gas reaction. The respective equilibrium constants can be formulated as follows:

$$K_{\text{Equi},(\text{R 4-3})} = \frac{p_{\text{H}_2}^2}{p_{\text{CH}_4}} \quad (\text{E 4-44})$$

$$K_{\text{Equi},(\text{R 4-4})} = \frac{p_{\text{CO}}^2}{p_{\text{CO}_2}} \quad (\text{E 4-45})$$

$$K_{\text{Equi},(\text{R 4-5})} = \frac{p_{\text{H}_2} p_{\text{CO}}}{p_{\text{H}_2\text{O}}} \quad (\text{E 4-46})$$

$K_{\text{Equi},r}$ is the equilibrium constant of reaction r , in atm,
 p_i is the partial pressure of component i , in atm.

The partial pressures are expressed in atmospheres and the activity of solids is unity. Gas phase reactions and equilibria can be obtained by various combinations of the above stated reaction equations and equilibrium constants. The water-gas-shift reaction for example can be expressed as a combination of Reaction (R 4-4) and Reaction (R 4-5). Expressions for the equilibrium constants for graphite are given in [252],

$$K_{\text{Equi},(\text{R 4-3})} = \exp\left(15.6525 - \frac{19,069.9}{T}\right) \quad (\text{E 4-47})$$

$$K_{\text{Equi},(\text{R 4-4})} = \exp\left(23.8898 - \frac{37,243.6}{T}\right) \quad (\text{E 4-48})$$

$$K_{\text{Equi},(\text{R } 4-5)} = \exp\left(19.8794 - \frac{29,286.6}{T}\right) \quad (\text{E } 4-49)$$

$K_{\text{Equi},r}$ is the equilibrium constant of reaction r, in psia,
 T is the temperature of the gas mixture, in Rankine.

In order to account for the possibility of formation of amorphous carbon, the following modifications are used [252]:

$$K'_{\text{Equi},(\text{R } 4-3)} = \frac{K_{\text{Equi},(\text{R } 4-3)}}{a_{(\text{R } 4-3)}} \quad (\text{E } 4-50)$$

$$K'_{\text{Equi},(\text{R } 4-4)} = K_{\text{Equi},(\text{R } 4-4)} \cdot a_{(\text{R } 4-4)} \quad (\text{E } 4-51)$$

$$K'_{\text{Equi},(\text{R } 4-5)} = K_{\text{Equi},(\text{R } 4-5)} \cdot a_{(\text{R } 4-5)} \quad (\text{E } 4-52)$$

where,

$$a_{(\text{R } 4-3)} = 1.0 + \exp\left(\frac{400.3 - T}{78.2}\right) \quad (\text{E } 4-53)$$

$$a_{(\text{R } 4-4)} = 1.0 + \exp\left(\frac{500.3 - T}{117}\right) \quad (\text{E } 4-54)$$

$$a_{(\text{R } 4-5)} = \sqrt{a_{(\text{R } 4-3)} \cdot a_{(\text{R } 4-4)}} \quad (\text{E } 4-55)$$

T is the temperature of the gas mixture, in Rankine.
 a is the amorphous carbon correction factor, dimensionless.

For determining the equilibrium curve for a certain temperature considering a binary sub-system, the ratio involving the partial pressures according to the equilibrium constant (Equation (E 4-44)-Equation (E 4-46)) can be determined by the temperature dependent equilibrium constant (Equation (E 4-47)-Equation (E 4-49)). Since the equilibrium curve is the point where carbon deposition is just about to occur, the following equation can be formulated:

$$p_{\text{H}_2} + p_{\text{CH}_4} = p_{\text{HC}} \quad (\text{E } 4-56)$$

p_{HC} is the binary hydrogen and carbon, in bar.

In order to start the calculation, the binary p_{HC} requires an initial guess. Combining Equation (E 4-56) and Equation (E 4-47) results in a quadratic equation. The solution of the positive root (negative root has no physical meaning) is of the form:

$$p_{\text{H}_2} = -\frac{K_{\text{Equi},(\text{R } 4-3)}}{2} + \sqrt{\frac{K_{\text{Equi},(\text{R } 4-3)}^2}{4} + K_{\text{Equi},(\text{R } 4-3)} \cdot p_{\text{HC}}} \quad (\text{E } 4-57)$$

p_{CH_4} is then calculated from the difference using the result from Equation (E 4-57) and the initial guess from p_{HC} . The same approach is taken for the binary oxygen-carbon. The results of both binaries are used to determine the partial pressure of water using the equilibrium constant $K_{\text{Equi},(\text{R 4-5})}$. After all partial pressures are obtained from the initial guess, the system pressure is normalized to the real pressure and iterated until the error is acceptable. This method is used for several points until enough points are obtained to draw an equilibrium line. Details on the algorithm and iteration scheme can be found in [252]. The results can be displayed in a ternary diagram as seen in **Figure 4-6**.

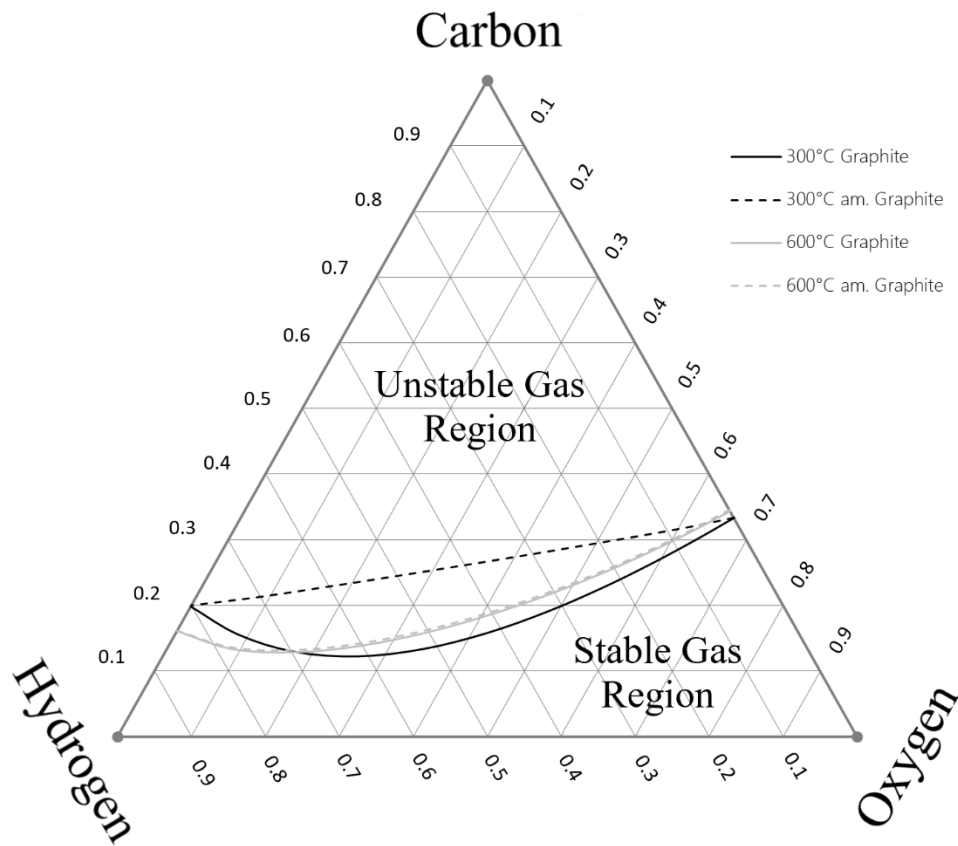


Figure 4-6: Ternary equilibrium chart for solid carbon formation at 300 °C and 600 °C.

The area above the equilibrium is the unstable gas region where carbon deposition can occur. Below the equilibrium line, the gas mixture is thermodynamically stable and carbon deposition is impossible. For the operation of SOFCs, as well as reformers and water gas shift reactors, this area constitutes safe reaction conditions. For the SOFCs, reformers and water gas shift reactors in this work, the solid carbon equilibrium criterion will be used for determining the steam requirement. Thereby, it is important to maintain a small safety margin and not operate the reactors exactly on the equilibrium curve.

4.2.2 Natural Gas and Biogas Plant Common Process Areas

Eductor

The eductor is a crucial part of the hybrid plant. It compresses the anode recirculation gas by converting pressure from the primary motive into kinetic energy and entraining the lower pressure stream before reconverting the kinetic energy back into static pressure. An illustration of an eductor is shown in **Figure 4-7**.

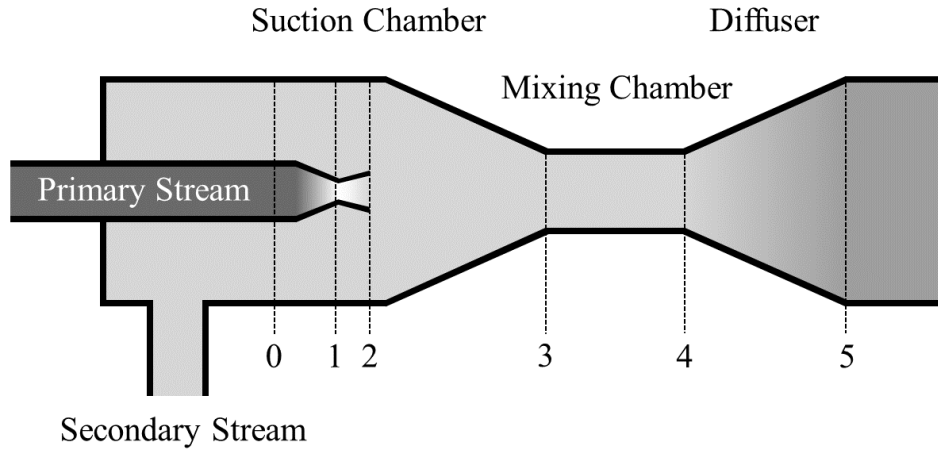


Figure 4-7: Eductor design and stream description.

Most models employed in system simulations are 1-dimensional thermodynamic models; however, these models tend to over predict the performance of eductor pumps as they do not account for the velocity distribution due to boundary effects. Zhu et al. [193] proposed a model for the entrainment section of eductors that can size the eductor based on the operating condition and accounts for boundary effects.

The nozzle throat area is calculated according to:

$$A_t = \frac{\dot{m}_{p,0}}{\sqrt{\Psi_p k R T_{p,0}} \left(\frac{2}{k+1} \right)^{\frac{k+1}{2k-2}} \rho_{p,0}} \quad (\text{E 4-58})$$

A_t is the nozzle throat area, in m^2 ,

$\dot{m}_p$ is the mass flow of the primary flow, in kg/s ,

Ψ_p is the isentropic coefficient of the primary flow, assumed to be 0.98,

k is the specific heat ratio, c_p/c_v , of the gas, dimension less,

R is the gas constant, in J/kg/K ,

T is the gas temperature, in K , and

ρ_p is the density of the primary flow, in kg/m^3 .

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Assuming that the temperature of the primary flow at location 3 is identical to the inlet of the secondary flow (primary mass flow is small compared to secondary mass flow) and assuming that the secondary flow temperature remains constant until it mixes with the primary flow, the Mach number of the primary flow can be determined as follows based on the further assumption that the primary flow fully expands to the secondary flow's pressure at location 3. Although the primary flow expands into the suction chamber, mixing will not occur until entering the mixing chamber (primary flow will expand to $R_{p,3}$, secondary flow occupies space from $R_{p,3}$ to the wall R_3).

$$Ma_{p,3} = \sqrt{\frac{2 \left(\frac{p_{p,0}}{p_{s,0}} \right)^{\frac{k-1}{k}} - 2}{k-1}} \quad (E 4-59)$$

Ma is the Mach number, dimensionless,

$p_{p,i}$ is the primary flow pressure at location i ,

$p_{s,i}$ is the secondary flow pressure at location i ,

From the Mach number, the velocity can be calculated.

$$v_{p,3} = Ma_{p,3} \sqrt{kRT_{s,0}} \quad (E 4-60)$$

$v_{p,i}$ is the velocity of the primary flow at location i , in m/s.

To determine the flow area of the primary flow at location 3, mass and energy balances for the primary flow from location 0 to location 3 are equated. The actual expansion diameter of the primary flow at location 3 can be calculated via

$$D_{p,3} = \frac{D_t}{\zeta} \sqrt{\frac{1}{Ma_{p,3}^2} \left(\frac{2 + (k-1)Ma_{p,3}^2}{k+1} \right)^{\frac{k+1}{4(k-1)}}} \quad (E 4-61)$$

D_p is the diameter of the primary flow, in m,

D_t is the nozzle throat diameter, in m, and

ζ is the friction coefficient, assumed to be 0.98.

To approximate the boundary effects of the secondary flow, Zhu et al. [193] propose the following behavior.

$$v_r = \begin{cases} v_{p,3} & 0 \leq r \leq R_{p,3} \\ v_{p,3} \left(\frac{R_3 - r}{R_3} \right)^{\frac{1}{n_v}} & R_{p,3} < r \leq R_3 \end{cases} \quad (E 4-62)$$

r is the domain in radial direction, in m,

R_3 is the mixing chamber radius, in m,

n_v is the velocity profile exponent, dimensionless, and determined by

$$n_v = \frac{\ln\left(1 - \frac{R_{p,3}}{R_3}\right)}{-\ln(\text{Ma}_{p,3})} \quad (\text{E 4-63})$$

After defining the mean mass flow of the secondary profile and evaluating the integral within the domain, the following expression for the mass flow is obtained. To obtain numerical values for all variables, Equation (E 4-63) and Equation (E 4-64) have to be solved iteratively.

$$\dot{m}_{s,3} = 2\pi v_{p,3} \rho \left[\frac{n_v R_3^2}{n_v + 1} \left(1 - \frac{R_{p,3}}{R_3}\right)^{\frac{n_v+1}{n_v}} - \frac{n_v R_3^2}{2n_v + 1} \left(1 - \frac{R_{p,3}}{R_3}\right)^{\frac{2n_v+1}{n_v}} \right] \quad (\text{E 4-64})$$

$\dot{m}_s$ is the mass flow of the secondary flow, in kg/s,

By assuming that the initial inlet velocities are very small compared to the thermal energy in the fluid the mass balance, energy balance and momentum balance can be written as follows [195].

$$\dot{m}_p + \dot{m}_s = \frac{p_4 v_4 A_3}{RT_4} \quad (\text{E 4-65})$$

$$(\dot{m}_p + \dot{m}_s) \left(c_p T_4 + \frac{1}{2} v_4^2 \right) = \dot{m}_p c_{p,p} T_{s,0} + \dot{m}_s c_{p,s} T_{s,0} \quad (\text{E 4-66})$$

$$p_4 A_4 + \frac{v_4^2}{\xi} (\dot{m}_p + \dot{m}_s) = (\dot{m}_p v_{p,3} + p_{p,3} A_{p,3}) + (\dot{m}_s v_{s,3} + p_{s,3} A_{s,3}) \quad (\text{E 4-67})$$

A is the cross-sectional area, in m^2 ,

ξ is a friction factor, assumed to be 0.8.

The diffuser is handled by an approach described by Vincenzo et al. [196] where the outlet pressure is determined by the following equation.

$$p_5 = p_4 + \frac{1}{2} \eta_D c_D \rho_4 v_4^2 \quad (\text{E 4-68})$$

c_D is the recuperation efficiency of the diffuser, assumed to be 0.75, and

η_D is the efficiency of total pressure conservation in an ideal diffuser, determined by Equation (E 4-69).

$$\eta_D = 1 - \left(\frac{A_4}{A_5} \right)^2 \quad (\text{E 4-69})$$

Whereby the area at the diffuser outlet is designed for a typical gas velocity found in plants. Here 10 m/s are assumed [32]. The energy balance for location 5 is formulated similarly to Equation (E 4-66).

Pre-Reformer

Pre-reforming is needed to reduce the temperature gradients inside the fuel cell as well as to protect the fuel cell from carbon deposition. Carbon deposition using natural gas is mostly due to the presence of higher hydrocarbons, yet, experimental work using biogas showed that also in biogas systems carbon deposition in the anode of the SOFC is problematic [253] (note: no thermodynamic gas stability analysis was conducted by the referenced authors and no steam added before the biogas entered the SOFC). However, the effect is not as dominant as in natural gas. A further advantage of a pre-reformer is the increase in flexibility of handling different fuel types. This is important, especially for biogas systems, as the composition of biogas can fluctuate over time. The pre-reformer is integrated in the manifold-pipes inside the SOFC furnace, which can be thermally integrated as described in [254]. By integrating the reformer into the pipes, pressure drop can be minimized. Conversion is limited due to the small surface area available inside the pipe compared to a fixed bed reactor, however, it is desired to mostly pre-reform higher hydrocarbons, which are present in low concentrations and only small amounts of methane in order to take advantage of direct internal reforming. In this design, the conversion can be either limited by the amount of catalyst present or the heat transfer from the SOFC to the reformer tube.

For the pre-reformer, the following overall reactions are considered:



The heat transfer from the SOFC stack to the reformer tube is modeled by a specific heat transfer coefficient, the geometry of the reformer tubes and the average SOFC temperature. Based upon the heat transfer rate and the amount of catalyst present in the pre-reformer, the conversion is determined based on the following kinetic expressions for steam reforming [255] over a nickel catalyst:

$$r_{(R\ 4-6)} = 4.225 \cdot 10^{15} \frac{\text{kmol} \cdot \text{bar}^{0.5}}{\text{kg} \cdot \text{h}} \exp\left(\frac{-257.0 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \left(p_{\text{CH}_4} p_{\text{H}_2\text{O}} - \frac{p_{\text{CO}} p_{\text{H}_2}^3}{K_{\text{Equi},(R\ 4-6)}} \right) \cdot \frac{1}{p_{\text{H}_2}^{2.5} \cdot A^2} \quad (\text{E 4-70})$$

$$r_{(R\ 4-7)} = 1.955 \cdot 10^6 \frac{\text{kmol}}{\text{kg} \cdot \text{h} \cdot \text{bar}} \exp\left(\frac{-89.2 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \left(p_{\text{CO}} p_{\text{H}_2\text{O}} - \frac{p_{\text{CO}_2} p_{\text{H}_2}}{K_{\text{Equi},(R\ 4-7)}} \right) \cdot \frac{1}{p_{\text{H}_2} \cdot A^2} \quad (\text{E 4-71})$$

$$r_{(R\ 4-8)} = 1.020 \cdot 10^{15} \frac{\text{kmol} \cdot \text{bar}^{0.5}}{\text{kg} \cdot \text{h}} \exp\left(\frac{-236.7 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \left(p_{\text{CH}_4} p_{\text{H}_2\text{O}}^2 - \frac{p_{\text{CO}_2} p_{\text{H}_2}^4}{K_{\text{Equi},(R\ 4-8)}} \right) \cdot \frac{1}{p_{\text{H}_2}^{3.5} \cdot A^2} \quad (\text{E 4-72})$$

$r_{(k)}$ is the reaction rate of reaction (k), in kmol/kg_{cat}/h,

R is the universal gas constant, 8.314 J/mol/K,

T is the temperature, in K

$p_{(i)}$ is the partial pressure of species i , in bar,

$K_{\text{Equi},(k)}$ is the equilibrium constant of reaction (k), in bar ^{$\sum v_i$} ,

H_i the Van't Hoff constant, in bar⁻¹ and dimensionless for H₂O

A and is calculated via Equation (E 4-73).

$$A = 1 + H_{\text{CH}_4} p_{\text{CH}_4} + H_{\text{CO}} p_{\text{CO}} + H_{\text{H}_2} p_{\text{H}_2} + \frac{H_{\text{H}_2\text{O}} p_{\text{H}_2\text{O}}}{p_{\text{H}_2}} \quad (\text{E 4-73})$$

Methane conversion and the water gas shift reaction are considered as equilibrium reactions with the following equilibrium constants:

$$K_{\text{Equi},(R\ 4-6)} = \exp\left(\frac{-26830}{T} + 29.898\right) \quad (\text{E 4-74})$$

$$K_{\text{Equi},(R\ 4-7)} = \exp\left(\frac{4276}{T} - 3.961\right) \quad (\text{E 4-75})$$

$$K_{\text{Equi},(R\ 4-8)} = K_{\text{Equi},(R\ 4-6)} \cdot K_{\text{Equi},(R\ 4-7)} \quad (\text{E 4-76})$$

The Van't Hoff parameters can be calculated using Equation (E 4-77) to Equation (E 4-80) [256], [257].

$$H_{\text{CH}_4} = 6.65 \cdot 10^{-4} \frac{1}{\text{bar}} \exp\left(\frac{38.28 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \quad (\text{E 4-77})$$

$$H_{\text{CO}} = 8.23 \cdot 10^{-5} \frac{1}{\text{bar}} \exp\left(\frac{70.61 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \quad (\text{E 4-78})$$

$$H_{H_2} = 6.12 \cdot 10^{-9} \frac{1}{\text{bar}} \exp\left(\frac{82.90 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \quad (\text{E 4-79})$$

$$H_{H_2O} = 1.77 \cdot 10^5 \frac{1}{\text{bar}^0} \exp\left(\frac{88.68 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \quad (\text{E 4-80})$$

The conversion of higher hydrocarbons ($C > 1$) can be considered as irreversible [258] and is modeled by the following kinetic expressions [259]–[261]:

$$r_{\text{Ethane}} = 0.48 \frac{\text{mol}}{\text{g} \cdot \text{s} \cdot \text{bar}^{1.8}} \exp\left(\frac{-43.8 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \cdot p_{C_2H_6}^{1.3} \cdot p_{H_2O}^{0.5} \cdot p_{H_2}^0 \quad (\text{E 4-81})$$

$$r_{\text{Propane}} = 43 \frac{\text{mol}}{\text{g} \cdot \text{s}} \exp\left(\frac{-42.5 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \cdot p_{C_3H_8}^1 \cdot p_{H_2O}^1 \cdot p_{H_2}^0 \cdot \frac{83.562}{(1 + 73.3 \cdot p_{C_3H_8} + 1.14 \cdot p_{H_2O})^2} \quad (\text{E 4-82})$$

$$r_{\text{Butane}+} = 0.79 \cdot 10^{-3} \frac{\text{mol}}{\text{g} \cdot \text{s} \cdot \text{bar}^{0.92}} \exp\left(\frac{-1.8 \frac{\text{kJ}}{\text{mol}}}{R T}\right) \cdot p_{C_4H_{10}}^{1.01} \cdot p_{H_2O}^{-0.09} \cdot p_{H_2}^0 \quad (\text{E 4-83})$$

Hydrocarbons higher than butane are approximated with butane reaction kinetics. Reaction rates provided in Equation (E 4-81), Equation (E 4-82) and Equation (E 4-83) have been measured under intrinsic conditions. Under real operating conditions the catalyst pellet underlies transport phenomena limiting the transport of reactants and products. Assuming a catalyst layer on the inside wall of the manifold/piping with a thickness of 1 mm, a catalyst effectiveness of $\eta_{\text{Cat}} = 0.3$ can be expected based upon the data provided in [255]. However, experimental validation will be needed for this concept in order to ensure sufficient mass transport from the bulk gas to the catalyst coated wall.

De-oxidizer

The de-oxidizer removes trace amounts of oxygen from the fuel stream. Even small quantities of O_2 can be problematic for reforming catalysts inside the pre-reformer and SOFC. Even for natural gas applications, the O_2 content can vary and reach values exceeding acceptable limits especially during times of peak shaving. Biogas generally contains larger amounts of O_2 and depending on the biogas cleanup, de-oxidation is almost always needed. Various technologies are available to remove O_2 from a fuel stream: 1) Catalytic oxidation with H_2 , 2) catalytic oxidation with CH_4 and 3) and oxidation of a metal such as Cu. Using a metal based de-oxidation process requires regeneration with H_2 which complicates the process design and is unfavorable for small scale application. While the catalytic oxidation with CH_4 is certainly possible in natural gas and biogas applications, the elevated operating temperatures of around 350 °C [262] make the process integration more challenging than using H_2 . Oxygen can be very effectively removed using H_2 at

ambient temperatures, which make it favorable for the use in SOFC-GT hybrid systems where minimizing pressure drop in the high temperature section between the eductor and SOFC is crucial due to the recycle. In order to completely remove O_2 from the fuel, 1.25 times the stoichiometric amount of H_2 is required [262]. The hydrogen is provided by the anode off-gas via a small slip stream. The de-oxidizer uses a Pt catalyst and operating temperatures of min. 50 °C are needed to guarantee complete O_2 reduction [262].

Natural Gas Plant Common Process Areas

Desulfurization. Desulfurization of natural gas is accomplished by an expendable sulfur adsorption process offered by SulfaTrap™, LLC. When low concentrations of sulfur species are present in the fuel gas stream expendable processes offer economic advantages over regenerative processes. The desulfurization process design includes two vessel and sorbent beds, whereby, the vessels are operated in a single column serial downflow configuration. The system is configured as a lead/lag system in which the first bed provides a high enough sulfur removal efficiency to operate the plant and the second bed acts as a polishing bed. Once the sulfur sorbent of the lead bed reaches its end-of-life, the lead bed is bypassed while its sorbent is replaced, and the lag unit becomes the new lead unit. This way the plant is not shut down and the fresh sorbent is always used in the polishing unit. The removal of H_2S , thiophenes and mercaptans from the natural gas is achieved simultaneously using SulfaTrap™ R2 sorbents. The operating temperature for sorbent-based desulfurization using SulfaTrap™ is typically below 50 °C and sulfur species in the clean gas stream are reduced down to < 4 ppbv [263]. The sorbent capacity ranges between 1-4 wt-% and the unit is designed for an operational period of 365 days before the sorbent needs to be replaced [264].

Biogas Plant Common Process Areas

Anaerobic Digester. The anaerobic digester is considered to be operated at a wastewater treatment facility and fed by sewage. The digester is operated under mesotrophic conditions at a temperature of 36 °C, which is typical for industrial applications. The heat demand for the digester is based upon an average value of 37.47 kWh/tonne of sewage [265], which is partly supplied by the GT exhaust and partly by biogas combustion. The heat is supplied by a closed loop water circulation system, which enters the digester with a temperature difference of 20 °C [266] and a pinch temperature of 11 °C is assumed. The power demand is based upon an average value of 13.23 kWh/tonne of sewage [265]. The sewage itself is modeled with a volatile solid content of 60.75% (in reference to the total solid content VS_{kg}/TS_{kg}) [265]. Moisture represents 89% of the feed stream [265] and the CH_4 yield is $0.24 \text{ m}^3_{CH_4}/\text{kg}_{VS}$ [265].

Desulfurization. In the case of biogas, bulk desulfurization is achieved in a regenerative process due to the substantially higher concentration of sulfur compounds. The regenerative process is also offered by SulfaTrap™, LLC and uses a two-column cyclic operation. While one column operates in adsorption mode

removing sulfur from the fuel gas stream, the second column is operating in regeneration mode. Adsorption is carried out at low temperatures up to 50 °C and elevated pressure. Regeneration is conducted at temperatures around to 400 °C and close to ambient pressure [267] using a hot CO₂-rich, O₂-free purge gas. The presence of O₂ will lead to oxidation of the adsorbed sulfur species and poses the risk of bed overheating and runaway temperatures [267]. The typical sorbent capacity is about 2.4 wt-% [267]. The cycle time is designed for 48 h (24 h adsorption, 24 h desorption) and the sorbent life is assumed to be 3 years. The sulfur removal efficiency of the regenerative desulfurization process is around 96.5 mol-%[267]. Thus, for SOFC applications a subsequent polishing bed is required. In the case of biogas this sulfur polisher can be combined with siloxane removal.

Siloxane Removal. The siloxane removal unit is another SulfaTrapTM, LLC process, which simultaneously removes siloxanes as well as sulfur and is located downstream of the bulk sulfur removal unit. The process is an expendable, adsorption-based process that is composed of a single column, downflow vessel. The absorption process operates at elevated pressures with temperatures up to 50 °C. Sorbents for siloxane removal reach capacities of up to 9 wt-% before breakthrough and reduce sulfur and siloxane levels down to ppbv levels [267]. The polisher is designed for an operating period of 365 days.

Chlorides. For chloride removal another SulfaTrapTM, LLC product is used. The sorbent R8C can operate at temperatures around 50 °C. With its capacity of approximately 2 wt.-% the sorbent is able to essentially remove all chlorides present in the biogas [263].

Fuel Upgrading. As mentioned before, fuel upgrading is beneficial for the SOFC and GT operation as it increases the methane concentration and heating value of the fuel stream. Additionally, it leads to a higher water vapor concentration in the GT exhaust stream, which is beneficial for flue gas water recovery. Various technologies are available for removing CO₂ from biogas. Some of the most prominent technologies are: water wash, amine scrubbing and membranes. While water wash and amine scrubbing show economic advantages in large scale applications, small scale applications typically favor membrane technology due to their lower capital cost and simpler plant design [268]. It may be pointed out that solvent-based systems require makeup chemicals and/or water, which can be problematic in certain regions. Furthermore, a preliminary study showed that membrane-based biogas upgrading has a lower energy requirement compared to using a water wash, which required circulation of large quantities of water due to the low CO₂ solubility in water. Thus, a membrane-based biogas upgrading solution has been chosen for this study.

Biogas upgrading is modeled using a hollow fiber carbon molecular sieve membrane. The model is a finite volume element model that is set up for counter flow operation. Typical dimensions are a fiber length of 40 cm, a fiber inner diameter of 150 µm and a wall thickness of 25 µm [269] and the following

permeabilities have been assumed for an operating temperature of 90 °C: CO₂ $1.10 \cdot 10^{-8}$ mol/m/s/bar, CH₄ $2.36 \cdot 10^{-10}$ mol/m/s/bar, O₂ $3.78 \cdot 10^{-9}$ mol/m/s/bar, H₂ $5.99 \cdot 10^{-8}$ mol/m/s/bar [270].

Carbon molecular sieves are commercially used in air separation PSA systems. While carbon molecular sieve membranes have demonstrated their potential in separating CO₂ from CH₄ on a small scale, this technology is still under development but is expected to reach commercialization in the near future. A methodology for predicting the cost of such a membrane unit is described in [269], [271].

4.2.3 Gasification Plant Process Areas

Fuel Feeding System

Coal. The coal is delivered to the plant site by rail cars. The unloading is accomplished via trestle bottom dumpers, which unload the arriving coal into two hoppers. The coal from each hopper is fed directly into a vibratory feeder and conveyors with an intermediate transfer tower are assumed to convey the coal to the coal stacker. The about 8 centimeters large coal lumps are either stored at a long-term storage pile or a reclaim area. The reclaimer feeds two vibratory conveyors located in the reclaim hopper under the pile. A belt conveyor transfers the coal to the coal surge bin located in the crusher tower. Then, the coal is crushed to a size of approximately 3 cm. Via a transfer tower the coal is routed to a tripper, which loads the coal into silos [33].

Before the coal is fed into the gasifier, it is dried from an initial moisture content of 25.8 wt-% to a moisture content of 18.0 wt-% in a roll mill incorporated flash dryer. Low temperature gasification technologies such as the studied steam hydrogasification process can efficiently handle high moisture content feedstock. The main motivation to remove moisture is to enable feeding into the gasifier. The dryer itself operates at a superheat of 13.9 °C. The drying air (nitrogen enriched air to keep the oxygen level below 11.3 vol-% due to safety concerns) is heated to 169 °C (which is safely below the devolatilization and autoignition temperature of PRB coal, 432 ± 4 °C [272] and 357 °C [273]) and dries the coal particles in a drying column. The incorporated mill reduces the coal particle size down to 700 µm. This particle size is larger than for other gasification technologies, e.g., entrain flow gasifier. Nevertheless, larger particles are acceptable in fluidized bed gasifiers due to improved heat and mass transfer as well as longer residence times. Subsequently, the coal is separated from the gas stream by cyclones and a bag house. The moisture rich gas is moved by an induced draft fan to the cooling section where the moisture is condensed. The gas is then recycled to the heater, which is supplied with saturated steam [33], [274]–[276]. The coal is fed into the gasifier via lock hoppers using high purity CO₂ from the carbon capture unit as pressurization medium.

Biomass. Biomass is considered to be produced in a more localized fashion. However, to support a large central plant of 100 MW or more, a large catchment area has to be taken into account. While short distance

biomass transport can be achieved by trucking, longer distances will have to rely on rail transport. The handling system of the biomass is similar to the coal handling system described above. However, the biomass is assumed to arrive at the plant site with a size of about 2.5-5.0 cm, which is typical for dry chopping with forage harvesters and further milling before storing the biomass in the silos is not needed [277].

After field drying the biomass possesses a moisture content of 15% when it arrives at the unloading dock. Biomass comminution is challenging and much more energy intensive than coal milling. In order to increase friability, the biomass is dried to a moisture content of 3 wt-% in a vapor compression dryer. Vapor compression drying has been developed to efficiently remove moisture from high moisture content biomass. In the case of low moisture removal, a substantial amount of compression power is required to supply enough energy for drying. Thus, additional heat is supplied in the form of steam to reduce the compression power and total power consumption of the biomass dryer. The biomass is preheated and fed into the fluidized bed dryer, which uses a portion of the removed moisture for fluidization after it has been compressed. The other portion of the removed moisture is condensed, and the heat is recovered and utilized in the dryer. The dryer operates at a temperature of 102 °C, which is safely below the devolatilization temperature of miscanthus (250 °C, onset 180 °C [278]) and ensures efficient moisture removal [278]. Since the dryer operates in an inert atmosphere containing water vapor, autoignition is not a concern [33]. Before the dry biomass is leaving the dryer more heat is recuperated.

The dry biomass is then transported to a hammer mill, which reduces the biomass chips to an average size of 1.5 mm. Larger particle sizes are acceptable in biomass applications due to its higher reactivity [30]. The comminution power is estimated based upon experimental data on miscanthus [279]. The biomass is then fed into the gasifier via lock hoppers using high purity CO₂ from the carbon capture unit as pressurization medium.

Gasifier

Hydrogasification appears to be favorable in combination with SOFCs due to the high CH₄ content of the produced syngas. However, currently no commercial hydrogasification technology is offered on the market. In this work, a steam hydrogasification process, similar to the one that is being development under the CE-CERT project, is adopted [142], [147]. The gasifier consists of two reaction vessels: I) the steam hydrogasifier and II) the char converter. The CE-CERT steam hydrogasification reactor is typically operated within a temperature range of 750-850 °C and an operating pressure around 30 bar [280]. Other hydrogasification technologies have been studied within a temperature range of 750-1200 °C and operating pressures ranging from 20 bar all the way up to 137 bar [281]. In this work a relatively low operating pressure of 20 bar has been selected due to the low pressure requirement of the SOFC. Low operating

temperatures are desirable from a thermodynamic and economic perspective. However, hydrogasification suffers from slow reaction kinetics and carbon conversion, which requires careful evaluation of the operating temperature. In literature, carbon conversion ranges from 33.7% to 89.0% [281]. An effective tool to improve the reaction kinetics of hydrogasification is the addition of steam like in the CE-CERT steam hydrogasification process [282]. In this work a H_2/C_{Coal} mole-ratio of 2.38 and a H_2O/C_{Coal} mole-ratio of 1.31 is used [280]. In order to drive the overall endothermic reactions in the steam hydrogasification reactor, hot sand is introduced into the gasifier. Hence, the gasifier is a fluidized bed reactor that allows for sufficient residence time to achieve high carbon conversion. In this study a carbon conversion of 80% has been assumed [280]. The unconverted carbon, together with the sand, are separated from the syngas via cyclones and introduced into the char converter. In the char converter the unconverted carbon is combusted with 20% excess oxygen [280]. Some of the heat released by the combustion process is absorbed by the sand, which is separated from the flue gas and ash via cyclones and filters, and returned to the steam hydrogasification reactor. The composition of the produced syngas is based upon chemical equilibrium considerations, which have shown to provide accurate results when compared to experimental data [143].

In power applications the formation of tars and oils during gasification is a concern. When surveying hydrogasification literature this may be counter intuitive, as many studies were carried out with the goal of producing low-oxygen containing bio-oils. Research on tar formation suggests that tar formation is only mildly influenced by pressure; however, tar decomposition is favored in the gas phase at higher pressures [281]. Above 600 °C, the formation of tars during hydrogasification decreases with increasing temperature. Simultaneously, increasing temperatures lead to an accelerated breakdown of tar components once they have been formed. Nevertheless, tars were still found in the syngas from hydrogasification at temperatures as high as 1200 °C [281]. Effective ways to break down tars are an increase in residence time, which increases the CH_4 yield and reduces the amount of produced tars and oils [281]. Furthermore, the addition of steam substantially increases the kinetics of tar cracking [281]. By adding small amounts of oxygen tars can be reduced to even smaller amounts. Based upon these observations it can be assumed that steam hydrogasification will lead to small amounts of tars and oils; though, experimental evaluation will be necessary to evaluate the exact amount and composition of tars and oils.

Various methods for tar and oil removal from syngas are available or under development. Generally, they can be divided into two categories I) separation or removal processes and II) tar conversion processes. Tar and oil separation can be achieved via water or oil scrubbers and sorbents. Water scrubbing of tars and oils, however, is only possible to a limited extend due to the mostly hydrophobic character of tars and oils. Water can only remove hydrophilic tar compounds. Hydrophobic tar removal efficiencies reach only around 22.1-38.8% and the remaining tar might pose a fouling risk to the scrubber when the oils and tars start condensing [283]. Due to these limitations, oil scrubbing processes have gained interest, which have shown

to reach tar removal efficiencies of up to 99% [284]. Oil based absorbents have to be regenerated and are typically used to recover tars and oils as a byproduct. Sorbents like activated carbon can reduce tars by about 81.5% but are rarely used and mostly have the purpose of a polishing bed [285]. Furthermore, sorbents can be costly and might require frequent replacement depending on the capacity and tar concentration (note: if tar concentrations are high a substantial amount of tar-bound energy can be captured in the sorbent and lost).

Catalytic candle filters are another option to reduce tars and oils. They are integrated into the gasifier head space and are made of α -alumina impregnated with a Ni-catalyst. They rather reform the tars and oils via catalytic cracking than removing them from the syngas. Promoters are added to improve the sulfur poisoning resistance. However, larger concentrations of sulfur species can cause problems related to their activity. Concentration of up to 100 ppm H_2S seem acceptable [286]. Additionally, CaO, MgO [287] and Fe-oxides [288] have found to show catalytic activity towards tar breakdown and could be directly integrated into the fluidized bed of the steam hydrogasifier. Instead of reforming tars and oils inside the gasifier at high temperatures, tars and oils can be reformed further downstream after potential catalyst poisons such as H_2S have been removed from the syngas. Reforming essentially converts all the tars present in the syngas and can be done using a Ni-based catalyst or noble metal catalysts [289]. Ni-based catalysts are typically more susceptible to carbon deposition and require sufficient steam supply [289]. Ni-based catalysts have been extensively used for tar reformation due to their high activity towards tar destruction. Also, noble metals have shown to have an exceptional activity for tar destruction and long term stability; however, their high cost makes them commercially less attractive than Ni-based catalysts [289]. Also, partial oxidation can be used to reduce tar content. However, temperatures around 800 °C have shown to primarily reduce tars to polycyclic aromatic hydrocarbons such as naphthalene [284]. A study by Praxair suggests that peak temperatures of at least 1250 °C (at 0.5 s residence time) are needed to achieve complete tar reforming.

After consideration of the above-mentioned advantages and disadvantages, metal oxide catalysts added to the sand will be used to breakdown tars and oils; however, this approach will have to be further studied and validated during gasifier development in order to ensure sufficient tar breakdown. A schematic of the dual-bed steam hydrogasification process is shown in **Figure 4-8**.

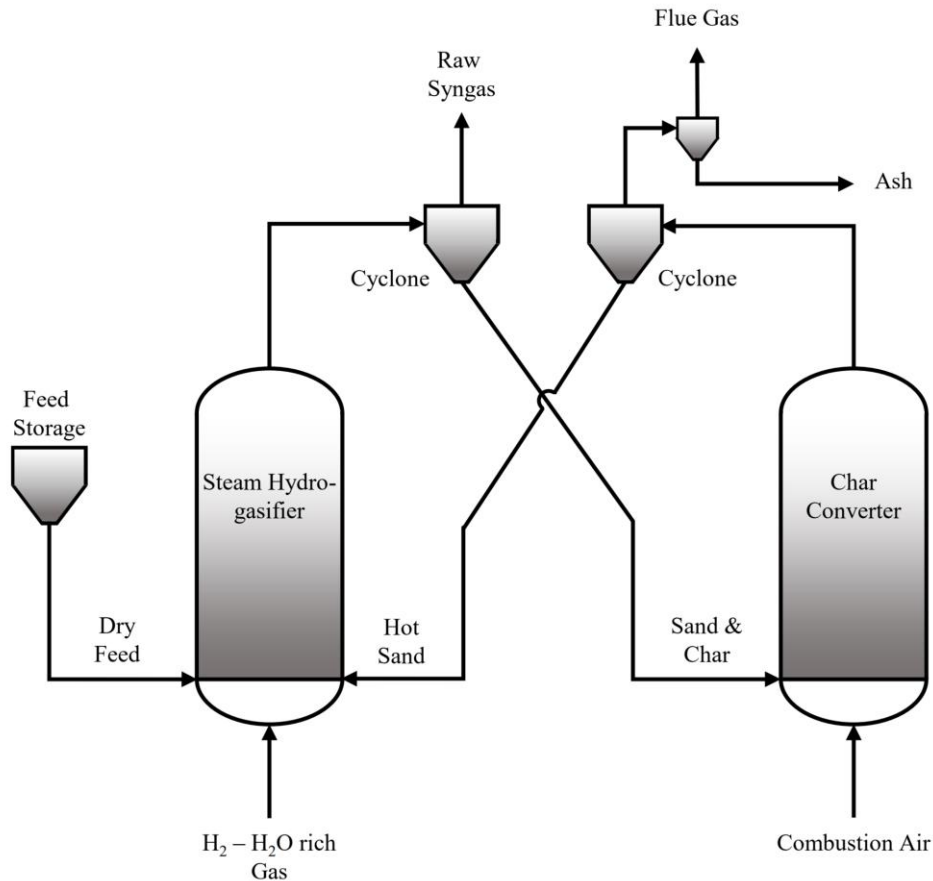


Figure 4-8: Process flow diagram of the dual-bed steam hydrogasification process.

Syngas Particulate Removal

The gasifier fluidization medium together with coal particulates and ash particulates are separated via cyclones at the outlet of the gasification reactor. However, some particulates will remain entrained in the syngas and need processing further downstream. It is important to remove particulates as early as possible to reduce the wear and tear of pipes and equipment. Larger particulates and during the cooling process solidified matter is removed using high pressure rigid, barrier-type filters. The filter will effectively capture all of the high-density dust and will virtually eliminate all fines larger than 5 microns. Typical particulate matter removal efficiencies are exceeding 99.5% [290]. In IGCC applications, these type of filters are typically operated in a temperature range 350 – 500 °C [33], [290], [291]. The filter elements are pulsed with recycle syngas to remove the trapped ash. The collected particulates are cooled and depressurized using a continuous fine ash removal system. For disposal, the particulates are combined and treated together with the ash from the gasifier. In order to remove the remaining particulates from the syngas a venturi scrubber is employed.

Syngas Alkali Removal

Gas turbine manufacturers have strict limits for alkali metals contained in the fuel. Many alkali metals form salts, such as NaCl, Na₂CO₃, KCl and K₂CO₃, which solidify in the temperature range 770 – 890 °C. In the gas turbine combustor these temperatures are often exceeded, and the highly corrosive molten salts can attack turbine blades. Also, in gasifier applications where the syngas is hotter than the melting temperature of the respective salt, they can cause fouling problems on high temperature heat exchangers. While most alkali salts will be present in their solid form at moderate temperatures and can be removed by filters, a proportion of each species will exist as a vapor, depending on the vapor pressure. As alkali salts are highly soluble in water, they can be efficiently removed by the venturi scrubber and no specific additional equipment in addition to the particulate control devices is needed for their removal.

Syngas Chloride Removal

Chlorides are mostly present in the form of alkali salts that can be removed by filters and scrubbing. Also, HCl is very water soluble and is effectively removed by water scrubbing at low temperatures. In order to improve the HCl removal efficiency at elevated temperatures the scrubber water is mixed with NaOH until a pH value around 10 is obtained [34]. NaOH chemically reacts with HCl forming NaCl and water, which enhances HCl mass transfer from the gas phase into the liquid. To ensure the removal of trace amounts of HCl, that could potentially damage downstream equipment, a sorbent guard bed is installed. Guard beds made of alkali metals are commonly used for the adsorption of HCl and can operate at temperatures up to 400 °C [292].

Syngas Volatile Metal Removal

For warm gas volatile trace metal cleanup, another TDA process is used, which can operate in the temperature range of 230 – 260 °C and utilizes a low-cost, high capacity, expendable sorbent. It removes trace metals such as Hg, As, Se, as well as sulfur, halides and nitrogen compounds. Trace metal are harmful for environment, human health and also lead to degradation of the SOFC [293]. If installed upstream of the sulfur removal unit, sulfur saturation of the sorbent is quickly reached, and sulfur components will break through. Similar behavior can be seen for NH₃. The saturation of the sorbent with sulfur does not prevent the removal of trace metals [294], however, the capacity will be reduced and requires more frequent replacement. Volatile trace metals are typically removed at efficiencies of greater than 90% with contaminant concentrations in the ppb levels leaving the unit [295].

Syngas Desulfurization

For warm gas desulfurization, a circulating bed adsorption process developed by RTI is used. RTI, who is industrializing this process, has completed large scale pre-commercial testing with over 3500 hours of total syngas operation at the Polk County IGCC [296]. RTI's test facility uses a 20% split stream of the raw gas produced at the Polk County IGCC, which is equivalent to a 50 MW plant scale. The syngas, cleaned in the test facility, is shifted in sweet-shift reactors and 90% of carbon is removed in a downstream aMDEA[®] process. Operational experience has shown that, while NH₃ passes through the desulfurization unit, other trace components such as Se, As, HCl and HCN interact with the sorbent and should be removed to improve the sorbent life [297]. The SO₂ produced in the regeneration loop of RTI's desulfurization unit is used to produce sulfuric acid. The technology is now ready for demonstration and deployment at full commercial scale. RTI's desulfurization process consists of an adsorber and regenerator.

In the adsorber, the raw syngas is mixed with regenerated zinc oxide sorbent, which is entrained in the gas flow and carried upwards in the adsorber. This enables sulfur components, such as H₂S and COS, to react with the sorbent to form ZnS. The adsorber typically operates in a temperature range from 315 – 535 °C. A great advantage of this process is the simultaneous removal of H₂S and COS. Most commercial processes can only remove one or the other, and a conversion reactor is needed before the sulfur can be removed. The chemical reactions of H₂S and COS with zinc oxide are described in Reaction (R 4-13) and Reaction (R 4-14).



As indicated in Reaction (R 4-13) and Reaction (R 4-14) the reaction of sulfur species with ZnO is an equilibrium reaction and their reaction extent is estimated based upon equilibrium data of the process generated by RTI [298], seen in **Figure 4-9**. In order to increase the heat capacity of the sorbent and limit the adiabatic temperature rise, alumina is recirculated together with the sorbent [299]. The used sorbent is then separated from the clean syngas via a cyclone and accumulated in a standpipe. Some of the used sorbent is recycled into the adsorber. The other portion is regenerated in a regeneration loop. In the regeneration loop, air is used to oxidize the ZnS at 650 – 815 °C.



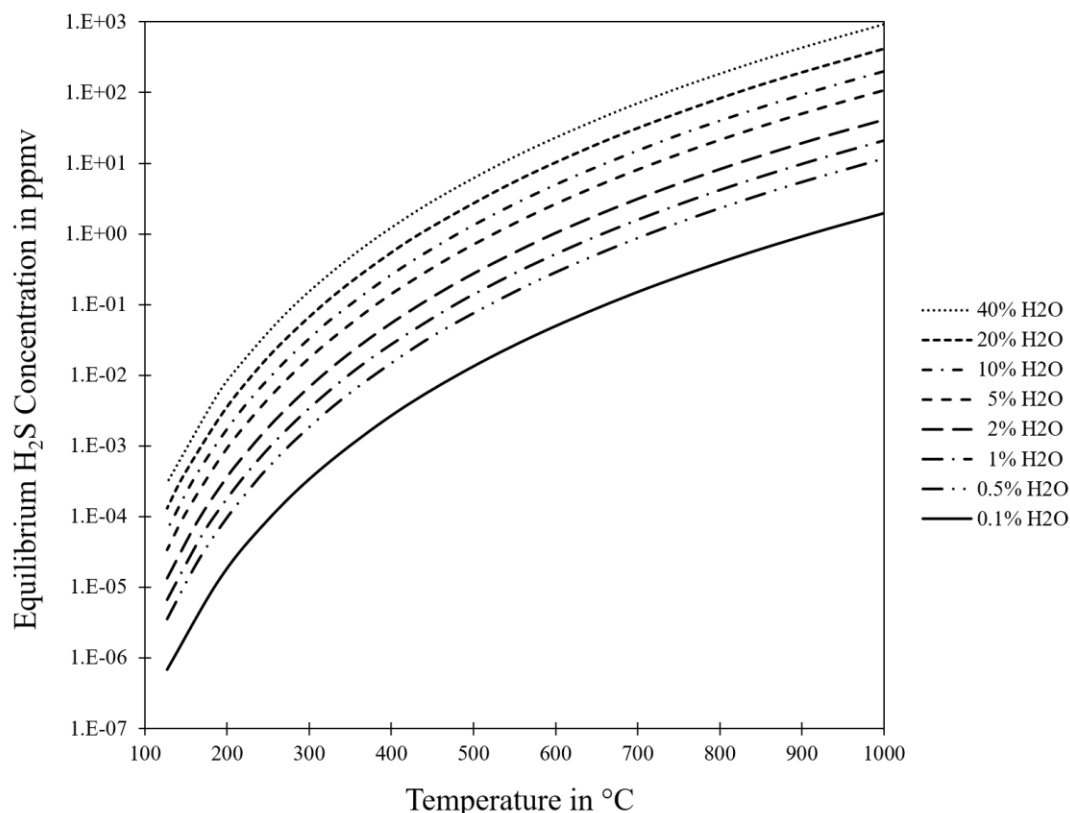


Figure 4-9: Warm gas sulfur removal equilibrium concentrations of H₂S as function of temperature and syngas H₂O content.

After separation of the regenerated sorbent, the sorbent is ready for reuse. The SO₂ rich off-gas can be used in a direct sulfur recovery process, sulfuric acid unit or modified Claus unit. The overall process can operate at pressures between 7 – 82 bar and achieves sulfur removal rates of greater than 99.9% [16]. SOFCs typically require a sulfur content of less than 0.1 ppm, which according to the minimum operating temperature of 315 °C is only achievable for syngas with very low water vapor content. To protect the fuel cell an additional expendable guard bed is needed to reduce the trace sulfur to levels below 0.1 ppm. Various metal oxides such as ferric oxides and manganese oxide have been developed that can operate at a temperature range of 280-500 °C [300].

Syngas Decarbonization

The warm gas carbon dioxide removal technology simulated in this study is based upon a combined WGS-pressure swing adsorption (PSA) technology developed by TDA Research, Inc. [301]–[304]. Performance data used in this study were provided from TDA and founded on realistic operating conditions at their demonstration sites at the National Carbon Capture Center and at the Wabash River IGCC. TDA’s system uses a commercial low temperature WGS catalyst in combination with a selective CO₂ sorbent that can

operate at elevated temperatures around 220 °C. TDA's sorbent selectively removes carbon dioxide from syngas via physical adsorption without forming covalent bonds. The sorbent consists of mesoporous carbon modified with surface functional groups. The heat of adsorption, 4.9 kcal/mol, is similar to the heat of absorption in the Selexol™ process, ≈ 4.0 kcal/mol [305]. This reduces the energy requirement for regeneration substantially compared to amine solvents and other chemical sorbents. In order to maximize regeneration efficiency, a combination of temperature swing, pressure swing and concentration swing is used. Especially, the regeneration at low pressure has a significant impact on the regeneration behavior and the required amount of purge steam. Since the adsorption is carried out in a fixed bed (batch process), several adsorption vessels are needed to enable continuous operation. Each adsorber will have to go through several operation and regeneration stages. If the number of adsorbers is equal to the number of process stages, a continuous operation is possible with one adsorber operating at each process stage. Also, the time interval for each step plays an important role. If a step/process is short, two or more processes can be executed during the same operational stage. For continuous operation of the combined WGS-PSA process a similar process design as for the individual CO₂-PSA process [306] (also developed by TDA) is anticipated. The system design consists of 8 beds, which operate in 8 stages/cycles, which in total involves 10 steps. An illustration of the sequencing of the single steps is given in **Figure 4-10**.

	Stage 1		Stage 2		Stage 3		Stage 4		Stage 5		Stage 6		Stage 7		Stage 8	
Bed 1	ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE		EQ3R EQ2R		EQ1R PRESS					
Bed 2	EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE		EQ3R EQ2R		EQ1R PRESS			
Bed 3	EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE					
Bed 4	PURGE		EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE			
Bed 5	PURGE		EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE			
Bed 6	EQ3D CoDEP		PURGE		EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE	
Bed 7	EQ1D EQ2D		EQ3D CoDEP		PURGE		EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D		EQ3D CoDEP	
Bed 8	ADS		EQ1D EQ2D		EQ3D CoDEP		PURGE		EQ3R EQ2R		EQ1R PRESS		ADS		EQ1D EQ2D	

Figure 4-10: Process stages of TDA's pre-combustion carbon dioxide capture system using solid sorbents [343].

The adsorption pressure is determined by the operating pressure of the SOFC or respectively the GT compressor discharge pressure. For the integration into a SOFC-GT hybrid cycle, the following pressure profile is assumed for the natural gas base case and adjusted according to the operating conditions for the coal and biomass cases. In this design, the syngas released during depressurization is utilized to pressurize a regenerated vessel:

1. Carbon Dioxide Adsorption at approx. 6.2 bar (ADS)
2. Pressure Equalization to approx. 5.3 bar (EQ1D)
3. Pressure Equalization to approx. 4.1 bar (EQ2D)
4. Pressure Equalization to approx. 2.9 bar (EQ3D)

5. Blowdown to approx. 1.3 bar (CoDEP)
6. Regeneration at approx. 1.2 bar (PURGE)
7. Pressure Equalization to approx. 2.8 bar (EQ3R)
8. Pressure Equalization to approx. 4.0 bar (EQ2R)
9. Pressure Equalization to approx. 5.2 bar (EQ1R)
10. Pressure Equalization to approx. 6.2 bar (PRESS)

An illustration of the bed cycling, and the respective pressure profiles is provided in **Figure 4-11**. It shows the interplay between the 8 different adsorbers during various stage of operation. All steps at each stage are engineered to be completed at the same time for a smooth transition into the next stage of operation. The data provided in **Figure 4-10** and **Figure 4-11** are as of August 2017 and might be subject to modifications as TDA is constantly working to improve its process. The advantage of integrating the WGS catalyst into the sorbent makes the unit more compact and drives the equilibrium limited WGS reaction towards the product side in a single step. This increases yield and reduces the amount of shift steam needed. Average carbon capture during test campaigns exceeded 96% despite the presence of substantial amounts of CO in the syngas [301]. Integrating the WGS catalyst into the sorbent can lead to undesirable heat effects,

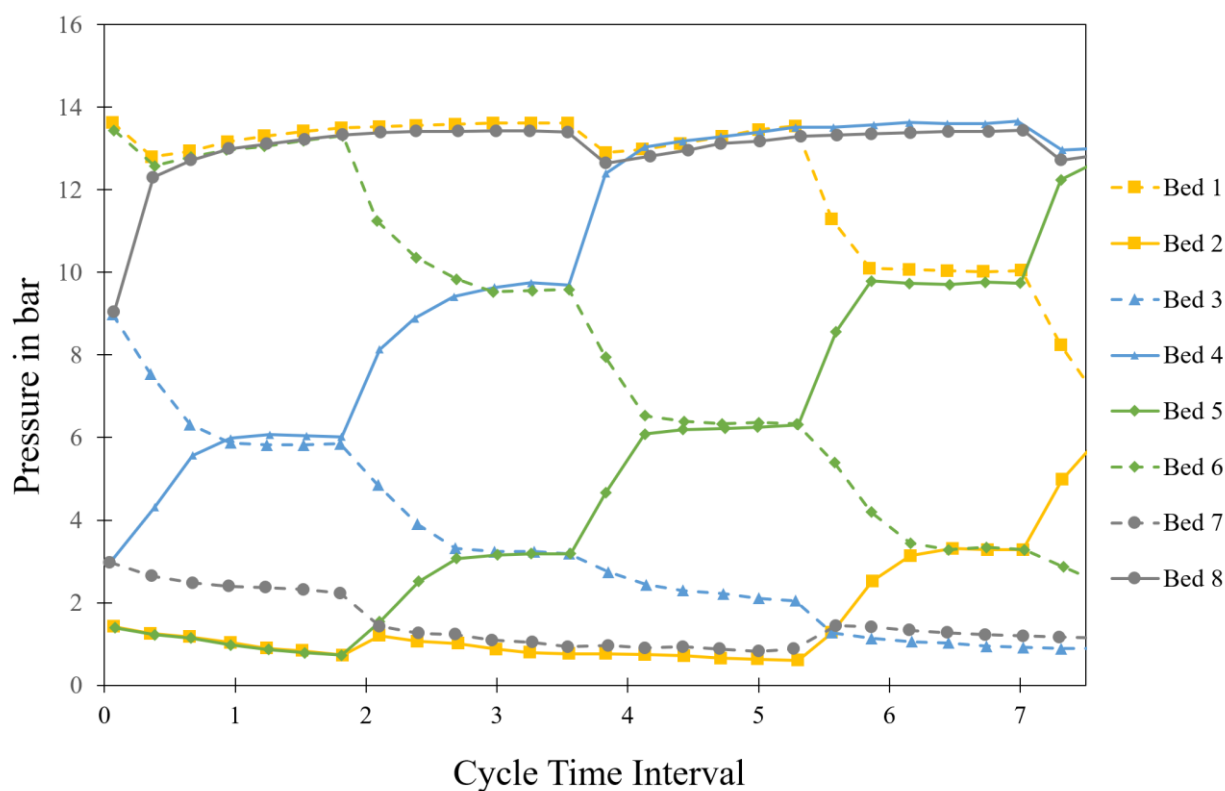


Figure 4-11: Process cycle diagram of TDA's pre-combustion carbon dioxide capture system using solid sorbents [343].

which can be mitigated using liquid water injection. However, due to the low CO content in the anode off-gas such measures are not considered in this study. Several test campaigns for proof-of-concept demonstrations have been completed at facilities at the Wabash River IGCC and the National Carbon Capture Center in Wilsonville, AL. Throughout the demonstration, the sorbent reliably removed CO₂ from the syngas, while the sorbent was able to maintain its activity and capacity. Furthermore, large scale sorbent production has been accomplished as well as long-term testing with over 20,000 cycles. Pilot scale testing and commercial demonstration readiness are believed to be achieved in the coming years which makes this technology a promising candidate for carbon removal in near future.

Flue Gas Alkali Removal

As described above, alkali metals can lead to corrosion problems. This is not only true for the syngas but also, the flue gas leaving the char converter. Thus, alkali metals must be removed from the flue gas before the high temperature, high pressure gas is expanded in the gasifier expander. While a low temperature scrubber can be employed for the syngas, this option is not available if substantial cooling and reheating of the flue gas is to be avoided. In this study high temperature aluminosilicate sorbents are employed to reduce alkali vapors in the flue gas down to ppbv concentrations. Alkali metal removal at high temperatures has been studied in the context of coal combustion and gasification for various temperature ranges. Punjak et al. [12] studied the adsorption of NaCl on kaolinite, bauxite and emathlite. While all sorbents are suitable for removal of alkali vapors, kaolinite and bauxite are more suitable for temperature exceeding 1000 °C. The adsorption behavior of alkali chloride on kaolinite and emathlite has shown to be irreversible, however chloride is released back into the gas phase as HCl vapor. Bauxite retains a portion of the chlorine and the adsorption is partially reversible. A further study by Punjak et al. [13] confirmed the findings and showed that also KCl is effectively removed by kaolinite. Concentrations as low as 50 ppbv have found to be possible when using bentonite and activated bauxite [307].

Flue Gas Particulate Removal

The flue gas leaving the char converter is initially sent to a series of cyclones disengaging sand and ash particles. However, before the high temperature flue gas can be sent to the expander, fine particles are removed in ceramic candle filters. Downstream of the expander another particle separation step is employed as part of the SNOXTM process using electrostatic particle precipitator. Thereafter, the particulate free flue gas can be desulfurized and reduced in NO_x.

Flue Gas Desulfurization and NO_x Control

After particulate removal, the flue gas is cleaned using the SNOXTM process developed by Haldor Topsoe [308]. In a first step NO_x is removed by addition of NH₃ upstream of the DeNO_x reactor at a temperature of around 400 °C [309]. Inside the DeNO_x reactor the NO_x (mostly NO) is catalytically reduced via Reaction (R 4-16).



The degree of NO_x reduction depends on the NH₃/NO_x ratio and the degree of mixing. Using a NH₃/NO_x ratio of 1.0 will typically result in NO_x levels of 10 ppm and NH₃ levels of 20 ppm at the DeNO_x reactor outlet [309]. Any NH₃ slip is destroyed later in the SO₂ reactor.

Subsequently to the NO_x removal, the flue gas enters the SO₂ oxidation reactor, which catalytically oxidizes about 98% of the flue gas's SO₂ to SO₃.



The NH₃ slip entering the SO₂ reactor oxidizes to N₂ and NO_x leading to an overall NO_x removal efficiency of 95% after the SO₂ oxidation reactor. Furthermore, any hydrocarbons left in the flue gas will be oxidized in the SO₂ reactor. Both catalysts, DeNO_x and SO₂ oxidation, have an expected life time of 10 years [309].

After the SO₂ oxidation the flue gas is cooled below its dew point in a vertical glass tube heat exchanger. During condensation, the SO₃ in the flue gas reacts with condensed water vapor forming sulfuric acid.



The acid with a concentration of approximately 95-96% is collected at the bottom of the condenser and further cooled to ambient temperature. Any acid droplets entrained in the flue gas are removed in a guard demister installed in the duct section upstream of the stack. Pressure drop across the SNOXTM unit is typically 45 mbar [309].

4.2.4 Water Recovery

Next to thermodynamics and thermal kinetics of water recovery it is important to consider the chemistry between sour gases in the flue gas and the condensate. For this purpose, the equilibrium water chemistry of the following gases has been implemented throughout all water recovery scenarios: CO₂, NH₃, SO₂, SO₃, NO, NO₂ and HCl.

Air-Cooled Surface Condenser

An air-cooled surface condenser uses ambient air as cooling medium to condense water contained in the process stream. Typical designs rely on fans on the air side to improve heat transfer through forced convection as seen in **Figure 4-12**. The air-cooled surface condenser is modeled on a thermodynamic level with a pinch temperature approach of 25 °C. Later the exchanger is sized based upon the thermodynamic performance considering heat



Figure 4-12: Air-cooled surface heat exchanger with fans [344].

transfer kinetics. Sizing is done in a commercially available heat exchanger design and rating software, which will also provide the pressure drop and fan-power consumption of the air-cooled heat exchanger. As water is directly condensed from the flue gas, substantial amounts of combustion products and acidic byproducts can be expected in the recovered water stream. In order to use the water in society, e.g., for irrigation purposes, the water needs to be purified. According to the expected water quality, a water treatment strategy consisting of an ultra-filtration unit, reverse osmosis unit and pH-adjustment step has been selected.

Direct Contact Condenser

The direct contact condenser uses a cooling medium, in this case recirculated water, that is in direct contact with the flue gas in a counter-flow packed bed column. The advantage of the direct contact condenser is its large surface area, which is created by the packed bed and does not have to be designed in the same way as the heat exchange surface area in the air-cooled condenser. Furthermore, the cooling medium in the direct contact condenser is not separated from the flue gas by metal tubes, which allows a lower pinch temperature. In this design a pinch temperature of 6 °C has been used. The recovery of water from the flue gas in the direct contact condenser is based upon the same physical principle as in the air-cooled condenser. Once the flue gas reaches its dew point, condensate is formed. The condensate instantly mixes with the cooling medium and leaves the condenser at the bottom together with the cooling water. An illustration of the direct contact condenser is shown in **Figure 4-13**.

The direct contact condenser is modeled with a Koch-Glitsch 2Y plastic packing with an HETP of 48.5 cm [310]. The mass transfer is solved based upon a correlation developed for packed beds by Bravo

at al. [311] and the heat transfer is solved following the Chilton-Colburn analogy. The column hydrodynamics are designed to approach 70% flooding.

Downstream of the direct contact condenser, the recovered water is split from the cooling water. The cooling water is recirculated and cooled via an air-cooled condenser before reuse. The recovered water stream is expected to contain substantial amounts of trace components from the flue gas. According to the expected water quality, a water treatment strategy consisting of an ultra-filtration unit, reverse osmosis unit and pH-adjustment step has been selected.

To reduce cost of the direct contact condenser, the column is made of a corrosion resistant plastic material (polypropylene). Polypropylene possesses excellent chemical resistance even at elevated temperatures. However, plastics can only be used at relatively low operating temperatures and the inlet temperature into the direct contact condenser is limited to 110 °C. Thus, before the hot flue gas enters the direct contact condenser, the gas is cooled via water spray cooling.

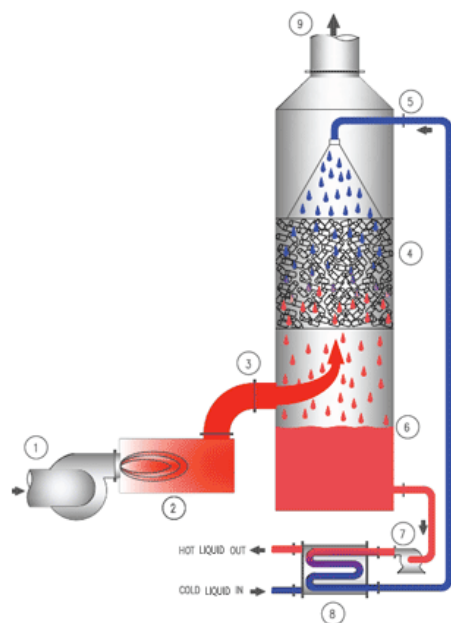


Figure 4-13: Illustration of direct contact condenser [345].

Liquid Desiccant Absorption Systems

The absorbers of the water absorption systems investigated in this study are designed in an identical manner to the direct contact condenser and are also constructed of polypropylene, which limits the inlet temperature to 110 °C. The absorber is a packed bed (Koch-Glitsch 2Y plastic packing with an HETP of 48.5 cm [310]), rate-based model that utilizes the Bravo et al. [311] mass transfer correlation in conjunction with the Chilton-Colburn analogy. The flooding approach of the absorber is 70%.

Using a desiccant instead of water in the column substantially increases the water recovery potential. When using a desiccant, the water recovery no longer depends on the gas's dew point temperature. It is rather the desiccant's vapor pressure that determines the amount of water that can remain in the gas (from a thermodynamic equilibrium perspective). In general, there are two types of liquid desiccants: aqueous salt solutions and pure component liquid desiccants. While most salts are solid at ambient temperatures, pure liquid desiccant can potentially evaporate depending on their vapor pressure. This is especially important for desiccant regeneration via thermal separation. In this study two different desiccants have been studied: an aqueous LiBr solution (aqueous salt solution) and monoethylene glycol (MEG) (pure component liquid desiccant).

Both system configurations consist of an absorber where the desiccant contacts the humid flue gas in a counter current packed bed. Before the gas leaves the absorber, it passes through a mist eliminator, which is designed to remove 99.99% of entrained droplets with a size greater than $20\text{ }\mu\text{m}$. The demister pad is a York Style 421 wire mesh, which is constantly flushed with water to avoid salt buildup. The pressure drop is estimated based on the dry pressure drop and the load dependent pressure drop as presented in [312]. The loaded desiccant leaves the column at the bottom and all or a portion of the desiccant is heated against the hot flue gas, which simultaneously cools the flue gas to the desired column inlet temperature of $110\text{ }^{\circ}\text{C}$. Dependent on the selected desiccant two different regeneration strategies are investigated. A general illustration of a desiccant based gas dehydration system with thermal regeneration is provided in **Figure 4-14**.

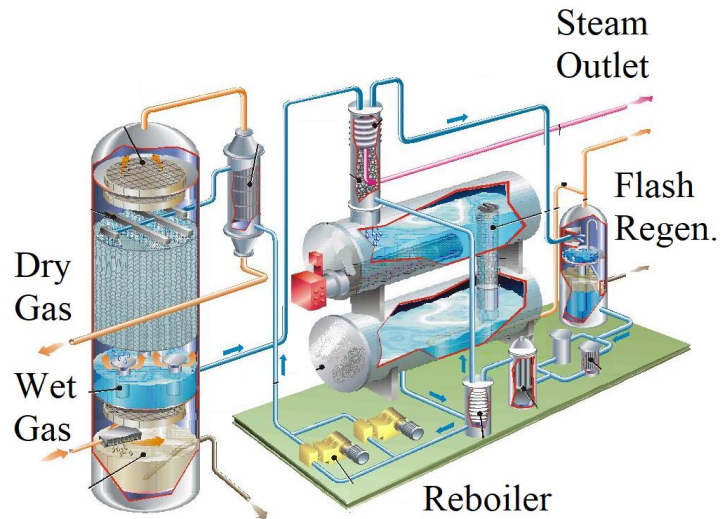


Figure 4-14: Illustration of gas dehydration system using liquid desiccants [346].

Lithium Bromide Absorption System. In the LiBr absorption system, the water-loaded, heated LiBr solution is regenerated in a vacuum flash drum. The recovered water vapor is condensed in an air-cooled condenser, whereby, the condensation temperature sets the vacuum pressure in the flash drum similarly to steam condensers in a Rankine cycle. To remove non-condensable flue gas components that have been absorbed by the desiccant from the regenerator, a compressor is used. After regeneration, the lean desiccant is cooled, makeup added and recycled to the desiccant storage tank. The recovered water is of relatively high quality [173] and it has been assumed that a pH adjustment step will be sufficient to export the water for irrigation purposes.

Monoethylene Glycol Absorption System. After leaving the absorber, the MEG solution is heated against the hot flue gas. However, the decomposition temperature of MEG needs to be respected and the hot flue gas is cooled to $240\text{ }^{\circ}\text{C}$ via spray cooling before it enters the MEG heat exchanger. Thereafter, the water is separated via a novel pervaporation membrane, which favors the transport of water across the membrane and limits the amount of MEG evaporation. The membrane is a poly(N,N-dimethylaminoethyl methacrylate)/polysulfone composite membrane and a kinetic pervaporation model has been developed based on the performance data provided in [313] to predict the water recovery rate, water quality and MEG loss in this application. The backpressure on the permeate site is determined by the cooling temperature of

the air-cooled condenser. Additionally, a vacuum compressor is installed to ensure the removal of non-condensables. In order to reuse the regenerated desiccant, the retentate needs to be cooled before recycling it back into the MEG feed tank. The expected water quality from this process is moderate due to some MEG pervaporation. The suggested water treatment strategy for this water quality consists of an ultra-filtration unit, reverse osmosis unit and pH adjustment step.

Transport Membrane Condenser

The transport membrane condenser (TMC) uses the capillary condensation effect, which lowers the vapor pressure of a liquid above the surface of a concave meniscus. This increases the driving force for mass transfer from the bulk gas to the surface and water can be condensed above the bulk gas's dew point. This effect becomes noticeable for pore diameters of 40 nm and smaller as seen in **Figure 4-15**. TMCs have been manufactured with pore diameters as small as 4 nm [198], which substantially increases the condensation temperature difference (allowable bulk gas superheat that will still lead to condensation in the

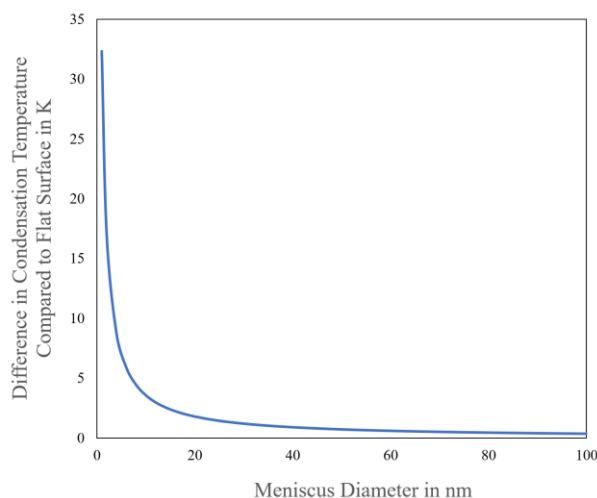


Figure 4-15: Condensation temperature difference of water at 25 °C for a concave meniscus compared to a flat surface.

pore) due to its exponential behavior. For large scale production, GTI suggests 7 nm alumina tubes [198] with an anticipated condensation temperature difference of approximately 4 °C, which also has been adopted in this study. To ensure that the TMC operates in the condensation regime, the inlet gas has to be sufficiently cooled prior to entering the TMC unit; otherwise, water evaporation will occur. The inlet temperature in this study has been set to 110 °C, which is achieved via water spray cooling upstream of the TMC unit.

To enable condensation in a TMC, cooling water has to be supplied on the tube side. The cooling water cools the flue gas to drive the condensation and simultaneously captures the recovered water. An illustration of the TMC design and design configuration is shown in **Figure 4-16**. The pinch temperature approach for the TMC is 11 °C. A first-principle model of the TMC has been developed to provide guidance with the design, sizing and economics. The heat transfer is modeled based upon an empirical Nusselt correlation specifically developed for the convective heat transfer observed in TMCs [184]. For mass transfer, the heat and mass transfer analogy is applied; however, the condensation on the TMC's surface leads to a suction

effect resulting in a mass transfer enhancement. Thus, the bird suction parameter is employed to adjust for suction effects [314].

After the heated cooling water leaves the TMC the water is cooled and water for export is split from the recycle. The water quality from the TMC is expected to be high and free of minerals [198]. Contaminants such as CO_2 , O_2 , NO_x and SO_x are inhibited from passing through the nano-porous membrane [198], which reduces the necessary water treatment steps. In this study only a pH-adjustment step is used before the water is exported from the plant site.

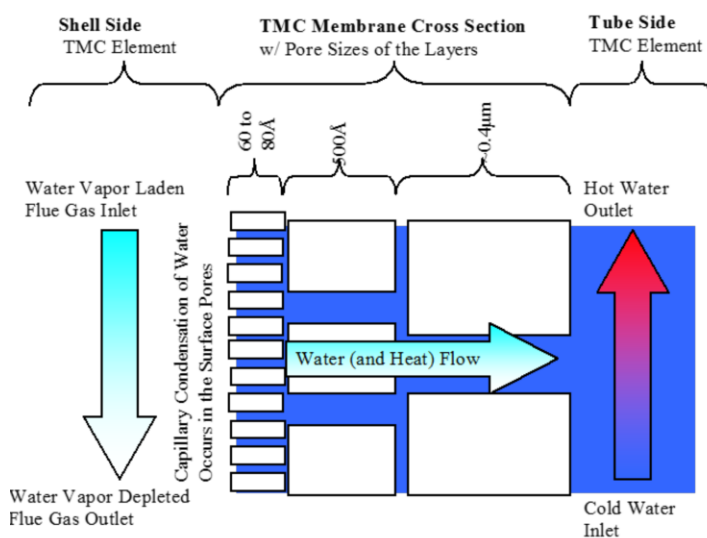


Figure 4-16: Schematic of the cross section of a transport membrane condenser tube [198].

Water Purification

This section describes the various water treatment technologies used in this study. The water purification strategies are chosen based upon the expected water quality of the recovered water. The respective treatment strategies are discussed together with the water recovery technologies and will be further clarified in the result section of the individual cases.

Ultrafiltration. The ultrafiltration unit has been designed with Dow Chemical's Water Application Value Engine (WAVE) and consists of a strainer, an ultrafiltration module, backwash system, clean in place system and air scour. The software provides removal efficiencies of suspended solids and organics, pressure drop as well as power and chemical requirements. The overall ultrafiltration system recovery is 97.8% and the wastewater is either disposed or treated in a zero liquid discharge system.

Reverse Osmosis. The reverse osmosis unit is a single pass two stage process with an overall water recovery of 95%. The performance of the reverse osmosis unit is based upon Dow Chemical's Water Application Value Engine (WAVE). The software provides removal efficiencies of the individual compounds as well as pressure drops on permeate and retentate side.

pH Adjustment. The pH adjustment system constantly measures the pH value of the recovered and treated water. Before export from the plant site the pH of the water is adjusted to 7.0 using a 50 wt.-% KOH solution.

Zero Liquid Discharge System

This section describes the zero liquid discharge system used in this study. Zero liquid discharge systems are only used for some plant configurations, which is further clarified in the result section of the individual cases.

Brine Concentrator. The brine concentrator is a vapor compression cycle, consisting of a pre-heater to recover heat from the vapor condensate, a deaerator, a falling film brine concentrator, a brine circulation pump and a vapor compressor that boosts the pressure of the recovered water vapor so that its latent heat can be reused in the brine concentrator. The brine concentrator recovers about 95% of the water from the brine before the brine is crystallized.

Spray Dryer/Crystallizer. For small applications spray dryers/crystallizers are a simple and cost-effective solution to crystallize brines. The brine is finely dispersed in a hot gas stream, in this case hot flue gas with a temperature of over 300 °C, and any water in the brine is instantly evaporated. The crystals formed during the process are separated by a cyclone and a baghouse. The baghouse operates under a weak vacuum and a downstream fan is employed to overcome this pressure drop of about 25” water.

4.3 Economic Analysis

4.3.1 Capital Cost Estimation

In order to evaluate the studied plants on a 2020-dollar cost basis, reference equipment costs are escalated to a 2020-dollar basis using Equation (E 4-84) and an annual escalation rate of 3%.

$$SC = RC \cdot (1 + AER)^{SY-RY} \quad (E\ 4-84)$$

SC is the scaled cost, RC the reference cost, AER the annual escalation rate, SY the scaled year and RY the respective reference year. Total plant cost (TPC) includes cost of process equipment, on-site facilities and infrastructure in support of the plant, both direct and indirect labor, and engineering services, procurement and construction (EPC). Detailed equipment design, contractor permitting and project/construction management costs are included in the EPC. Process contingencies as well as project contingencies are also included in the TPC. Process contingencies and project contingencies typically vary between 0-30% depending on the type of equipment and application. Data found in [33] have been used to establish the basis for contingencies.

SOFC Equipment Cost

The SOFC stack cost is based upon the manufacturing costs developed by the Battelle Memorial Institute [156] for a factory production scale of 1000 modules per year (module size > 300 kW). The cost of the stack is developed by considering 13 individual stack components/process steps: ceramic cell, end plates, interconnect, anode frame, picture frame, cathode frame, laser welding, ceramic-glass sealing, anode mesh, cathode mesh, stack assembly, stack brazing, and testing and conditioning. Each of these categories is divided into 5 subcategories: material cost, labor cost, machine cost, scrap cost and tooling cost. The cost estimation is based upon vendor quotes for materials, production equipment and outsourced items. Custom manufacturing processes have been evaluated based upon machine, energy and labor requirements. Furthermore, cost learning curves have been applied for material and manufacturing costs as function of the production scale. An illustration of the SOFC cell design is provided in **Figure 4-17**.

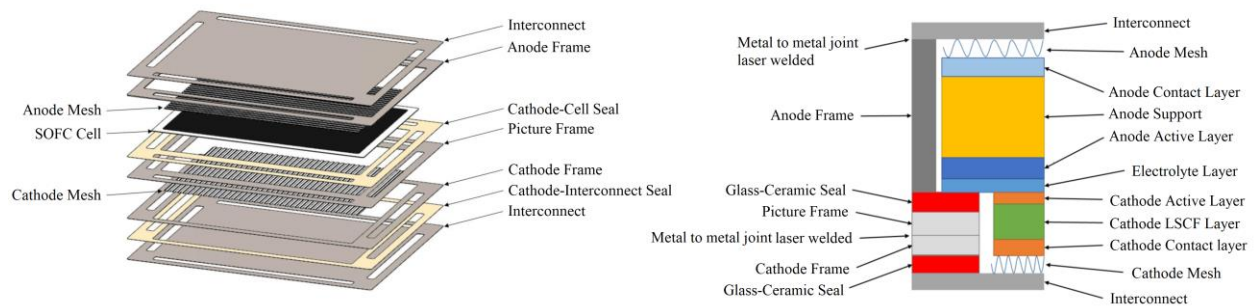


Figure 4-17: SOFC cell design. SOFC repeat unit on the left, detailed assembly on the right (not drawn to scale) [156].

Every stack consists of two end plates made of Hastelloy X. The end plates are sand casted, and face milled before machining. The interconnects are manufactured from a ferritic steel sheet. The sheets are stamped and laser cut to size. Thereafter, a ceramic coating is applied, and the interconnect is heat treated at 1000 °C for 4 hours. The anode frame, which supports the interconnect on the anode side, the picture frame, which supports the periphery of the electrolyte, and the cathode frame, which supports the interconnect on the cathode side, are also manufactured from ferritic steel via stamping and openings for gas paths are punched. For anode and cathode frames laser welding and glass-ceramic seals are used to connect the frames. The meshes on anode and cathode side are produced from stainless steel sheets of appropriate dimensions and laser-cut to final dimensions. The ceramic cell itself is built in layers using aqueous ceramic slurries. The anode support is created by tape casting of a YSZ (8%-mol)/NiO slurry, which is then oven-dried, hot-pressed and cut. The subsequent layers are applied via screen printing and dried in a fired kiln dryer. The anode active layer and contact layer are made of YSZ/NiO. The cathode

material is LSCF and the cathode active layer and contact layer are made of LSM/YSZ. The stacks are assembled in an assembly line and brazed, tested and conditioned before shipping.

In this work the cost of the stack as a whole is calculated based upon the geometric parameters provided in **Table 4-13** (unless otherwise specified) and the above-mentioned cost items. The stack cost represents the stack only and does not include power conditioning systems, housing, final assembly and installation, nor other balance of plant equipment. In order to arrive at the stack sales price a markup of 50% is used [156].

The hybrid DC/AC inverter cost after sales markup is assumed to be 0.37 \$/W [156]. Other balance of plant equipment directly related to the SOFC, such as eductors, valves, pre-reformers, etc. as well as assembly, testing and conditioning are determined in accordance with [156]. The SOFC pressure vessel and insulation have been sized according to the stack size and prices are established based upon cost data provided in [315]. All these costs are summarized in the total SOFC cost. Degradation of the SOFC is accounted for in the form of spare stacks and a 1st order degradation of 0.2% per 1000 h has been assumed. End of life is assumed to be reached after 10 years.

GT Equipment Cost

Numerous new technologies in the energy sector are currently being developed in order to reduce emissions and utilize unconventional energy resources. A significant number of such technologies rely on GT technology in combination with low heating value fuels such as ultra-low emission catalytic oxidizers [316], integrated gasification combined cycles [35] or SOFC-GT hybrids [317]. The economic evaluation of such energy systems is crucial to gauge their competitiveness in the free market. Various methodologies have been developed to estimate the cost of GTs ranging from machine cost rate modeling [318] to production and developing cost correlations based on cost improvement functions, turbine inlet temperature (TIT) and dry engine weight [319]. Brennan and Finizie [320] developed a correlation to estimate the Maurer factor for aero engines based on TIT, air flow and the Kronecker delta function, which then can be used to estimate the GT cost. However, some of these approaches require detailed knowledge of the engine design. Thus, it is more common for early stage system assessments to use simpler cost correlations based on a \$/kW basis [247]. While these correlations are fairly accurate for natural gas [247], they will lead to skewed results when considering customized GTs for low heating value fuels. Custom engineered GTs are an important part in system analysis studies to fully evaluate the technical potential of a technology. However, in order to conduct a fair economic comparison n-th plant cost of custom gas turbines are needed. With the current tools available, it is difficult to estimate the n-th plant cost of custom gas turbines and a more fundamental cost basis is needed. Thus, GT equipment costs have to be distinguished between commercially available off-the-shelf cost and n-th plant custom designed GTs.

Commercial Off-the-Shelf GT Equipment Cost. In this work the equipment cost for off-the-shelf GTs is estimated based upon the following correlation (expressed in 2018-dollar) [247].

$$C_{GT} = 9650 \cdot kW^{0.7} \quad (\text{E 4-85})$$

C_{GT} is the price of the GT in 2018-dollar and kW is the ISO power rating of the GT in kW as provided by the vendor. Next to the GT equipment cost, the total GT cost includes costs for foundation, piping and insulation as well as instrumentation and electrical equipment. These costs are established in accordance with [33]. For the Dresser Rand KG2-3G EF no surcharge for GT modifications in order to accommodate the SOFC has been assumed. The EF version of this GT has been designed for external heat sources, which is assumed to allow for the integration of an SOFC without major changes.

N-th Plant Custom Engineered GT Equipment Cost. In order to determine the n-th plant cost of a GT, cost driving factors have to be analyzed and correlated. For this purpose, a total of 29 GT data sets have been compiled based on data provided in [247], [321]. The data consist of thermal efficiency, turbine exhaust temperature (TET), exhaust mass flow, compression ratio, efficiency, TIT, and genset price based on 2019-dollar for GT models offered by various vendors. The compressor flow rate is determined assuming that 98% of the exhaust flow is air, which is a typical value for GTs firing natural gas. The GT sizes for the derivation of a n-th plant GT cost correlation range from 1.8 MW – 117.0 MW. The full data set is provided in **Table 4-15**.

In order to identify important key cost driving factors of the GT, parameters shown in **Table 4-15** have been classified into “input variables” and “output variables”. “Input variables” are design variables which are chosen by the engine design engineer. “Output variables” are defined as variables whose value is a result of the input variables or engine design. The commercially available software STATA® 14 is utilized as a statistical tool to identify input variables that are correlated to the budget price of the GT. In this study, the following definitions are used: $0.3 < r < 0.7$ for moderately correlated and $0.7 < r < 1.0$ for highly correlated. Considering the sample size, an alpha level (significance level) of 0.1 is used to determine statistical significance. The statistically significant input variables are then selected for regression over the budget price to derive the n-th plant cost of a custom GT.

The variables shown in **Table 4-15** have been classified as follows: power (output variable, depends on engine size/mass flow, TIT and pressure ratio), efficiency (output variable, depends on TIT and pressure ratio²), TET (output variable, depends on TIT and pressure ratio), exhaust flow (input variable, almost

² Note: efficiency also depends on GT size as compressor and turbine efficiency typically increase with size.

Table 4-15: Gas Turbine Data

GT Model	Power ^[245] kW	Efficiency ^[245] %-LHV	TET ^[245] K	Exhaust Flow ^[245] kg/s	Compressor Flow ^[245] kg/s	Pressure Ratio ^[245] -	TIT ^[313] K	Budget Price ^[245] \$M
M1A-17D	1810	28.1	795	8.1	7.9	10.5	1255	1.44
OP16-3C	1850	25.1	846	9.0	8.8	6.7	1275	1.60
501-KB5S	3980	29.7	833	15.4	15.1	10.3	1353	3.25
Centaur 50	4600	29.3	783	19.0	18.6	10.6	1328	3.50
501-KB7S	5380	32.3	767	21.3	20.9	13.9	1330	4.30
SGT-100	5400	31	814	20.9	20.4	15.6	1583	4.00
SGT-300	7901	30.6	815	30.2	29.6	13.7	1394	5.00
Taurus 70	7965	34.3	789	26.7	26.1	17.6	1422	4.90
Mars 100	11350	32.9	758	42.2	41.4	17.7	1380	6.40
SGT-400	14326	35.4	813	44.3	43.4	18.9	1561	7.25
Titan 130	16450	35.5	764	55.8	54.7	17.1	1422	8.50
Titan 250	21745	39.3	736	69.8	68.4	24.1	1478	11.00
LM2500DLE 50Hz	22417	35.4	820	68.4	67.1	18.1	1505	12.40
LM2500DLE 60Hz	23200	36.6	812	68.2	66.8	18.0	1505	12.45
SGT-600	24480	33.6	816	81.3	79.7	14.0	1455	11.15
LM2500+ DLE 60 Hz	31900	38.8	799	87.0	85.2	23.1	1505	13.50
RB211-GT DLE	32130	39.3	783	93.8	92.0	21.5	1533	13.75
LM2500+ G4 DLE 60Hz	34500	39.2	808	91.5	89.7	23.6	1478	14.75
SGT-750	37031	40.4	741	115.0	112.7	24.3	1533	14.00
H-25	41030	36.2	842	114.8	112.5	17.9	1533	13.60
6B.03	44000	33.5	824	145.0	142.1	12.7	1408	16.75
LM6000PF DLE	45000	42.1	734	124.6	122.2	31.0	1516	20.00
LM6000PF Sprint	50000	42.1	734	131.9	129.3	31.0	1516	21.00
LM6000 SAC	54000	40	744	143.3	140.4	33.5	1516	22.00
SGT-800	57000	40.1	838	136.6	133.9	21.8	1575	17.50
AE64.3A	80000	36.4	853	215.0	210.7	18.3	1700	26.00
6F.03	88000	36.8	895	218.8	214.4	16.4	1600	29.00
7E.03	91000	33.9	825	295.0	289.1	13.0	1386	24.00
SGT6-2000E	117000	35.4	805	367.9	360.5	12.0	1422	31.90

same as compressor flow), compressor flow rate (input variable), pressure ratio (input variable), and TIT (input variable). The exhaust mass flow and compressor mass flow are highly correlated with the GT price ($r = 0.93$, $p = 0.00$), however, the exhaust mass flow and compressor mass flow are also highly correlated with each other ($r = 1.00$, $p = 0.00$). As a result, only the compressor mass flow is used for regression as the compressor constitutes a major component of GT development and production costs. The pressure ratio ($r = 0.56$, $p = 0.00$) and TIT ($r = 0.40$, $p = 0.03$) are moderately correlated with the GT price. The correlation of the compressor mass flow with the pressure ratio is weak and does not pass the significance threshold ($r = 0.14$, $p = 0.46$). Thus, the null hypothesis cannot be rejected. The correlation between the TIT and the pressure ratio ($r = 0.57$, $p = 0.00$) and the correlation between the TIT and the compressor mass flow ($r = 0.36$, $p = 0.06$) is moderate, which indicates that higher TITs tend to be found in GTs with higher compressor mass flow and higher pressure ratios. In conclusion, the compressor mass flow, pressure ratio and TIT are significant drivers of the GT price and can be used to derive the n-th plant cost correlation. The derived cost correlation based upon compressor mass flow, pressure ratio and TIT is shown in Equation (E 4-86).

$$C_{GT} = 0.1391 \cdot m_{Air}^{0.73} + 0.0474 \cdot m_{Air}^{0.73} p_c^{0.55} + 8.8221 \cdot 10^{-13} \cdot m_{Air}^{0.73} T_{TIT}^{3.50} + 0.0356 \quad (E\ 4-86)$$

The R^2 -value of the correlation is 0.975. The air flow $m_{Air}^{0.73}$ is significantly positively associated with the GT price with a p-value of 0.01. Also, the pressure ratio factor $m_{Air}^{0.73} p_c^{0.55}$ is significantly positively associated with the GT price with a p-value of 0.00 and the TIT factor $m_{Air}^{0.73} T_{TIT}^{3.50}$ is significantly positively associated with the GT price with a p-value of 0.01. The validation of the correlation against the data in **Table 4-15** shows great accuracy (see **Figure 4-18**). To arrive at the total GT cost, costs for foundation, piping and insulation as well as instrumentation and electrical equipment are added according to [33].

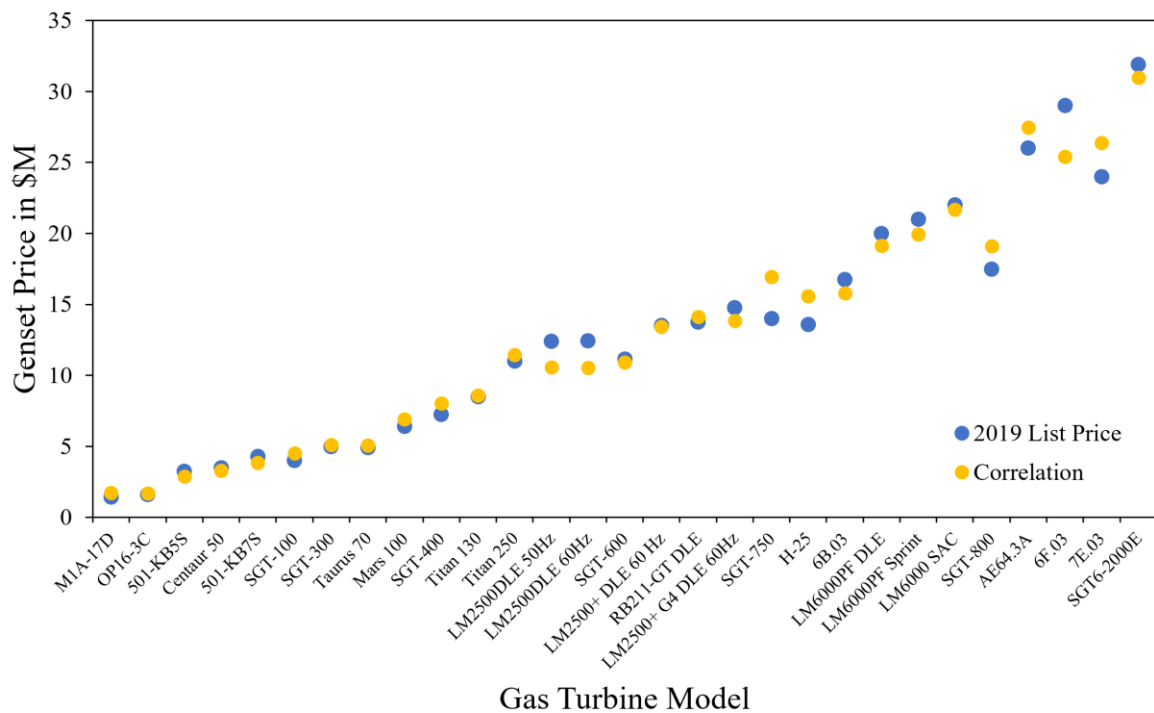


Figure 4-18: Comparison of GT list price data with GT price correlation results.

Other Plant Equipment

Other plant equipment costs are obtained from individual cost correlations [315] or vendor quotes. Heat exchangers have been sized and priced by a commercially available exchanger design software. In order to obtain cost estimates for the investigated hybrid plant, comparable process equipment is scaled to the required capacity using the following equation.

$$SC = RC \left(\frac{SP}{RP} \right)^u \left(\frac{TS}{TR} \right)^{0.9} \quad (E\ 4-87)$$

SC represents the scaled cost, RC represents the reference cost, SP is the scaled parameter and RP is the reference parameter used to scale the equipment. u is the scaling exponent. The scaling exponent is equipment specific and can be found in [315] for a large variety of plant equipment. TS is the number of trains/purchased quantity of each equipment of the scaled plant and TR is the number of trains/purchased quantity of each equipment in the reference case. The exponent 0.9 accounts for the cost reduction when more than one of the same pieces of equipment is purchased and installed. The expected accuracy of this methodology for cost estimation is expected to be in the range of -15% to +30%.

4.3.2 System Analysis

Although SOFCs are arranged in modules, the majority of the plants are designed as single trains. Designs of the SOFC-GT hybrid power stations investigated in this study (all sizes and feedstocks) are comprised of one fuel processing train (1 x 100%) (in case of coal of biomass this includes gasifier), multiple GTs and off-gas oxidizers with a unit size of approximately 10 MW (MW/10 x 1000/MW%), single train heat recuperation equipment (1 x 100%) and multiple pressurized SOFC modules (N x 100/N%). Each SOFC module consists of 22 stacks, each stack with its own eductor and pre-reformer and one pressure vessel with thermal insulation per module. Cases with carbon capture and sequestration have one train of CO₂-PSA and CO₂-compression (1 x 100%). Likewise, water recovery is designed as single train (1 x 100%).

Using a modular design for the SOFC-GT hybrid unit enables a faster market introduction and cost savings in engineering design and mass production. The integration of a SOFC into a GT is challenging and requires detailed engineering and system design studies, which at the current time is resource intensive as GTs are primarily offered for natural gas applications. This modular approach limits the economies of scale of these SOFC-GT hybrid power plants but also has advantages with respect to scheduled and unscheduled maintenance. In case of a dysfunctional module, the hybrid module can be isolated from the rest of the plant and repaired while the other modules can remain online, minimizing plant downtime.

Transport cost, storage cost and monitoring cost (TS&M) for CO₂ sequestration are not included in the TPC but will be accounted for separately. Operating and maintenance costs are expenses associated with the daily operation of the power plant and include: operating labor, maintenance material and labor, administrative and support labor, consumables, fuel, waste disposal and byproduct sales. Operating and maintenance cost can be separated into fix and variable costs. Fix costs are comprised of annual operating labor, maintenance labor, administrative and support labor as well as property tax and insurance.

The 10 MW NG plants are assumed to be controlled by two skilled operators per shift (3 shifts per day) who are paid an hourly salary of 40.85 \$/h (escalated from [34]). Additional operators are needed for additional operating units such as anaerobic digester, water recovery unit or carbon capture unit. A summary of the number of plant operators needed in the individual cases is provided in **Table 4-16**. The operating

labor burden is estimated at 30% and the overhead charge rate is assumed to be 25% at a plant capacity factor of 1.0 [33]. Labor related to maintenance activities is accounted for separately as 35% of the maintenance cost. The maintenance cost itself is based on individual cost relationships [33]. Administrative and support labor are projected to be 25% of operating and maintenance labor. Other fixed costs include property tax and insurance costs, which are typically around 2% of the TPC [33].

Table 4-16: Summary of Plant Operators

Plant Size	10 MW				50 MW				100 MW					
Fuel	Natural Gas		Biogas		Natural Gas		Natural Gas		Natural Gas		Subbituminous Coal		Biomass	
Carbon Capture	No	No	No	No	No	No	Yes	Yes	Yes	Yes	Yes	Yes	Yes	Yes
Water Recovery	No	Yes	No	Yes	No	Yes	No	Yes	No	Yes	No	Yes	No	Yes
Number of Operators	2	2.3	3	3.3	4	4.3	5	5.3	6	6.3	10	10.3	10	10.3

Variable operating costs such as maintenance costs are dependent on the availability of the plant. Other variable costs that must be considered are consumables such as fuel, sorbents and catalysts. Fuel costs for the various fuels used in this study are summarized in **Table 4-17**.

Table 4-17: Summary of Fuel Costs

Fuel	Value	Unit	Cost Basis	Reference
Natural Gas	3.96	\$/MMBtu	2018	[322]
Sewage	0	\$/kg _{Solid}	2018	assumed ¹
PRB Coal	21.90	\$/tonne _{ar}	2018	[323]
Miscanthus Biomass	120.36	\$/tonne _{ar}	2019	[324]

¹ sewage is a waste stream which is typically associated with a disposal cost

Costs for replacing the catalysts and sorbents at the end of their individual lifetime are also accounted for. Furthermore, catalysts and sorbents require an initial fill before startup, which contributes to the upfront costs. Consumables are evaluated on their consumption rate, which varies for different types of catalysts and sorbents. After reaching their end-of-life, these materials need to be disposed of and any associated disposal fees are included in the cost analysis. Costs for CO₂ transport, storage and monitoring are treated separately. For CO₂ storage in an average saline formation, the cost of CO₂ transport, storage and monitoring are 22.00 \$/t [325].

New technologies are often considered as high-risk investments, which requires an equity of 45%. Considering a current dollar cost of 2.94% (debt) and 7.84% (equity), an average weighted capital cost of 4.73% after tax is obtained. In this study an income tax rate of 25.74% is assumed based upon an effective federal tax rate of 21% and state tax rate of 6% [326]. Capital depreciation has a declining balance of 150% over the course of 20 years. One hundred percent of the total overnight capital is considered to be depreciable [326]. Investment tax credits are not considered in the financial analysis as well as tax holidays are assumed to be 0 years. Expenditure periods for the 10 MW hybrid plants have been assumed to be 1.5

years with a total overnight capital distribution of: 10.0%, 60.0% and 30.0% per half-year. Expenditure periods for the 50 MW hybrid plants have been assumed to be 2.0 years with a total overnight capital distribution of: 10.0%, 37.5%, 30.0% and 22.5% per half-year. Expenditure periods for the 100 MW natural gas hybrid plants have been assumed to be 2.5 years with a total overnight capital distribution of: 10.0%, 27.0%, 24.0%, 21.0% and 18.0% per half-year. Expenditure periods for the 100 MW gasification hybrid plants have been assumed to be 4.0 years with a total overnight capital distribution of: 10.0%, 18.2%, 16.4%, 14.6%, 12.9%, 11.1%, 9.3% and 7.5% per half-year. Capital cost escalation during expenditure is assumed to be 3.0%. For all plants, a plant life of 30 years is considered. These financing structures can be approximated with a capital charge factors (CCF) of 0.0740 (10 MW), 0.0750 (50 MW), 0.0760 (100 MW NG) and 0.0796 (100 MW gasification). Using the above-described capital charge factors the following equation can be used to calculate the first-year levelized cost of electricity (COE).

$$COE = \frac{(CCF)(TOC) + OC_{fix} + (CF)(OC_{var})}{(CF)(MWH)} \quad (E\ 4-88)$$

where COE is the first year's levelized cost of electricity, CCF the capital charge factor, TOC the total overnight capital, OC_{fix} the total fixed annual operating costs, OC_{var} the total variable annual operating costs, CF the plant capacity factor and MWH the annual net-megawatt hours generated at 100% CF.

The cost of water (COW) is determined based on a reference case without water recovery, which provides a reference COE. Based on this reference COE, the COW is determined by adjusting the \$/L price to match the COE from the reference case.

Chapter 5

Results and Discussion

5.1 SOFC-GT Hybrid Systems

5.1.1 Economic Analysis of SOFC-Cell Design and Thermal Management in SOFCs under SOFC-GT Hybrid Operating Conditions

Introduction

Critical for the operation of SOFCs are thermal gradients within the cell and/or the stack. Over the recent years research efforts have tried to address challenges experienced with respect to thermal cell management. Robinson et al. [327] experimentally studied thermal stresses under SOFC operating conditions using combined synchrotron X-ray diffraction and thermal imaging. Also, Jasinski et al. [328] studied temperature distributions of a SOFC cell experimentally but using infrared thermography.

Detailed experimental data on the processes occurring inside a SOFC-cell under real operating conditions are difficult to obtain. Thus, a strong focus of the research community lies in utilization of numerical tools. Janardhanan et al. [329] investigated temperature profiles inside a planar SOFC under various operating conditions. Similarly, Xu et al. [330] conducted simulations of a planar SOFC specifically looking at the temperature distribution inside the cell. Numerical simulations were also conducted by Yahya et al. [331] who studied the impact of multiple SOFC design parameters upon the temperature distribution. Thermal stresses resulting from the non-uniform temperature distribution have been studied in [332]. Chaudhary et al. [333] used CFD to gain insights into temperature gradients and thermal stresses under different operating conditions.

The performance of a single cell can vary when integrated into a stack. Amiri et al. [334] modeled multiple layers of planar SOFCs in a stack aiming to account for temperature gradients in both; flow direction and stacking direction. The study discusses three temperature reduction approaches with the goal to minimize thermal gradients inside the cell. Other challenges arise from the dynamic operation. Zhang et al. [335] studied control strategies for SOFC systems focusing on maximizing electrical efficiency under thermal cell management constraints. Control strategies to minimize thermal gradients during operation have also been studied by Wu et al. [336].

Another major obstacle on the way to commercialization of SOFCs is their cost (due to the relatively low level of research and development and production scale compared to internal combustion engines or batteries) and projections for full scale production costs of SOFCs are still difficult to make. Most system studies rely on rough estimates or cost targets set by institutions like the U.S. Department of Energy.

Rao and Samuelsen [106] studied the cost of SOFC-GT hybrid configurations assuming a base cost of 1000 \$/kW based upon projections of the Siemens-Westinghouse hybrid. A study by Kim and Kim [124] uses a much larger value of 7000 \$/kW and assumes a cost of 3500 \$/kW for large scale production. Curletti et al. [127] based their cost estimates on U.S. Department of Energy targets [154]. Siefert and Litster [128] estimated the SOFC cost with 1,700 \$/m² of active cell area and Arsalis [155] employed a cost correlation dependent on active cell area and operating temperature which was then also adopted in [85], [90]. Massardo and Magistri [97] proposed a cost correlation based upon the specific power (kW/m² cell area) resulting in specific SOFC costs (cost on a per kW basis) ranging from 700–2,200 \$/kW. However, all these cost estimates or methodologies are missing a more fundamental basis. In many instances detailed models for the SOFC are not available to construct accurate costs and instead specific SOFC costs are simply employed. For accurate thermo-economic system studies, such specific costs are imprecise due to the wide range of operating conditions encountered in SOFC systems especially when considering pressurized operating conditions. Even if the active cell area is being determined, the above-mentioned cell area specific cost correlations cannot capture costs correctly; while some cell components might scale with the active cell area, others such as framing or assembly scale on a different basis. One of the most sophisticated and recent studies on SOFC costs has been produced by Battelle Memorial Institute [156] which developed a detailed SOFC cost estimation method and established cost projections for atmospheric pressure SOFC combined heat and power systems. The developed cost model contains various individual cell components and manufacturing processes, but no parametric study of the cell design nor an optimization with respect to costs was conducted as it would require detailed modeling of the entire system to establish the required operational data which appears beyond the scope of their study.

In summary, current literature is missing detailed insights into the cost of SOFC stacks under SOFC-GT hybrid operating conditions as well as how cell design parameters impact thermal stresses. To the best of the authors' knowledge this is the very first study to investigate synergistic effects between cell design economics and thermal management. This study will provide a set of favorable cell design parameters and detailed cost estimates for researchers conducting thermo-economic SOFC-GT hybrid system studies. Based upon the cell design and production scale, specific cost in \$/kW can then be derived.

Methodology

Typical operating parameters of a SOFC-GT hybrid system have been established at steady state conditions. State point data of the streams entering the fuel cell were utilized to study the influence of various SOFC-cell design variables, e.g., interconnect thickness, and operating variables, e.g., degree of pre-reforming, upon thermal cell management and economics. Most critical for SOFCs are thermal gradients in axial direction and the temperature profile across the membrane can be considered as essentially constant [337]. In this study, fuel utilization, air utilization and operating voltage were held constant while the number of cells was adjusted to meet the utilization specifications. Based upon the particular cell design and number of cells, the cost of the stacks has been established, which is then converted to specific cost in $\$/kW_{DC}$ on a 2020 cost basis.

Design Basis

In a SOFC-GT hybrid, a gas turbine is used to pressurize the SOFC while serving as the “bottoming cycle”. Higher SOFC operating pressures allow for higher fuel utilization, higher operating voltage and higher current density. Additionally, thermal energy generated in the SOFC and unutilized fuel can be further utilized in the gas turbine. These synergistic effects allow the SOFC-GT hybrid to reach very high efficiencies. Various hybrid configurations have been studied in literature with pressure ratios ranging from 3 bar to 50 bar [91], [102], [106], [107], [110]. For this study the configuration shown in **Figure 5-1** has been used with a SOFC inlet pressure of 10 bar and a SOFC inlet temperature of 700 °C. The anode exhaust recycle has been determined based upon thermodynamic equilibrium considerations [252] to avoid carbon deposition. The fuel composition prior to reforming is: 18.18 vol.-% CH_4 , 0.62 vol.-% C_2H_6 , 0.14 vol.-% C_3H_8 , 0.08 vol.-% C_4H_{10} , 6.52 vol.-% H_2 , 4.04 vol.-% CO , 23.53 vol.-% CO_2 , 46.16 vol.-% H_2O , 0.73 vol.-% N_2 . The corresponding air composition entering the cathode side is: 20.74 vol.-% O_2 , 77.32 vol.-% N_2 , 0.92 vol.-% Ar , 0.03 vol.-% CO_2 and 0.99 vol.-% H_2O . Fuel and air are in co-flow and the cell is operated at a cell voltage of 0.83 V, a single pass fuel utilization of 80% and an air utilization of 35%. For all cases, except the sensitivity study on reforming, use a pre-reformer that converts 25% of the incoming CH_4 .

A summary of the SOFC model key geometrical and operating parameters used in for the cell design study is provided in **Table 5-1**. The electrolyte is yttria stabilized zirconia, which is typically operated in a temperature range from 600 – 1000 °C and cell materials are not changed throughout this investigation.

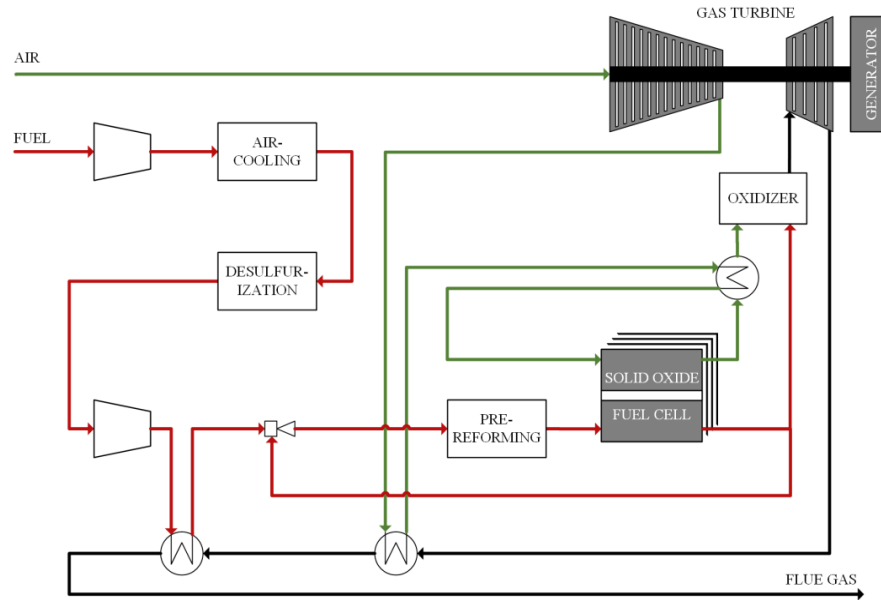


Figure 5-1: Solid oxide fuel cell - gas turbine hybrid configuration.

Table 5-1: Summary of Internal Reforming Solid Oxide Fuel Cell Design Parameters

Geometry	Value	Unit
Fuel/Air Channel Length	0.2	m
Fuel/Air Channel Width	0.003	m
Rib Width	0.002	m
Number of Channels	50	-
Fuel Channel Height	0.001	m
Air Channel Height	0.002	m
Interconnect Height	250	μm
Anode Thickness	400	μm
Electrolyte Thickness	8	μm
Cathode Thickness	20	μm
Operation Parameters	Value	Unit
Operating Pressure	10	bar
Inlet Temperature Anode	973.15	K
Inlet Temperature Cathode	973.15	K

Results

Economics of Production Scale. Battelle Memorial Institute developed cost scenarios for four different annual production scales: 100 modules per year, 1,000 modules per year, 10,000 modules per year and 50,000 modules per year. Furthermore, distinction between module sizes of greater than 300 kW and smaller than 300 kW are made. Characteristic trends for the cost scaling factor (CSF) under SOFC-GT hybrid operating conditions for these production scenarios have been derived and presented in **Figure 5-2**.

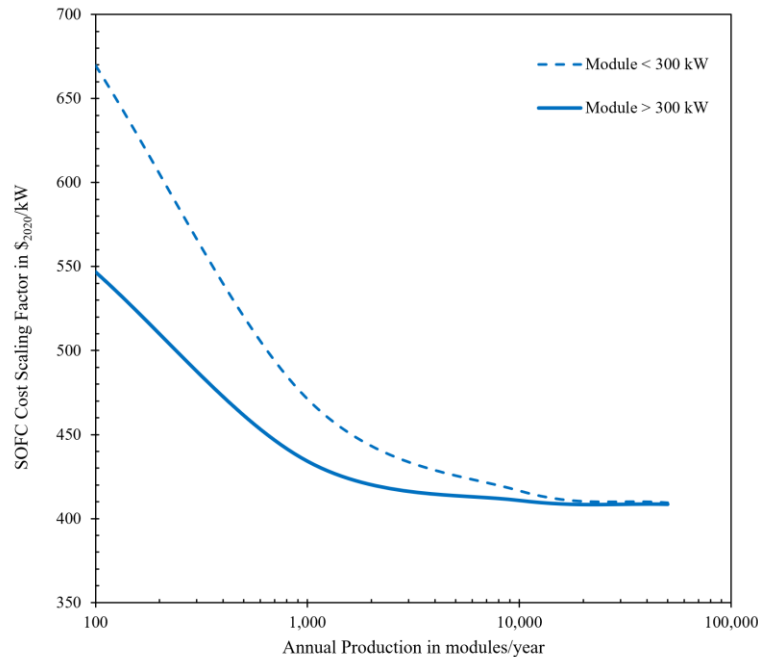


Figure 5-2: Solid oxide fuel cell cost scaling factor expressed in 2020 dollar as function of annual production scale and module size.

Small production scales lead to a substantial difference in the specific SOFC cost between module sizes of > 300 kW and < 300 kW. Once reaching a production scale of 10,000 modules per year, this cost difference between the two module sizes almost vanishes. For production scales between 100 and 1,000 modules per year, stack costs are strongly dependent on the production scale while the impact of the production scale for annual production scales between 10,000 and 50,000 is very small. For the standard geometric parameters listed in **Table 5-1**, SOFC stack costs vary between 405 \$/kW and 670 \$/kW.

Economics of Cell Design and Thermal Management. In the following section 13 SOFC design and operating parameters have been investigated with respect to their impact on economics and thermal cell management.

Cell Length. The cell length has been varied from 0.05 m to 0.50 m while maintaining a single pass fuel utilization of 80% and an air utilization of 35%. The temperature of the positive-electrode, electrolyte,

negative-electrode (PEN) structure along with the local temperature gradients for the different cell lengths is shown in **Figure 5-3**.

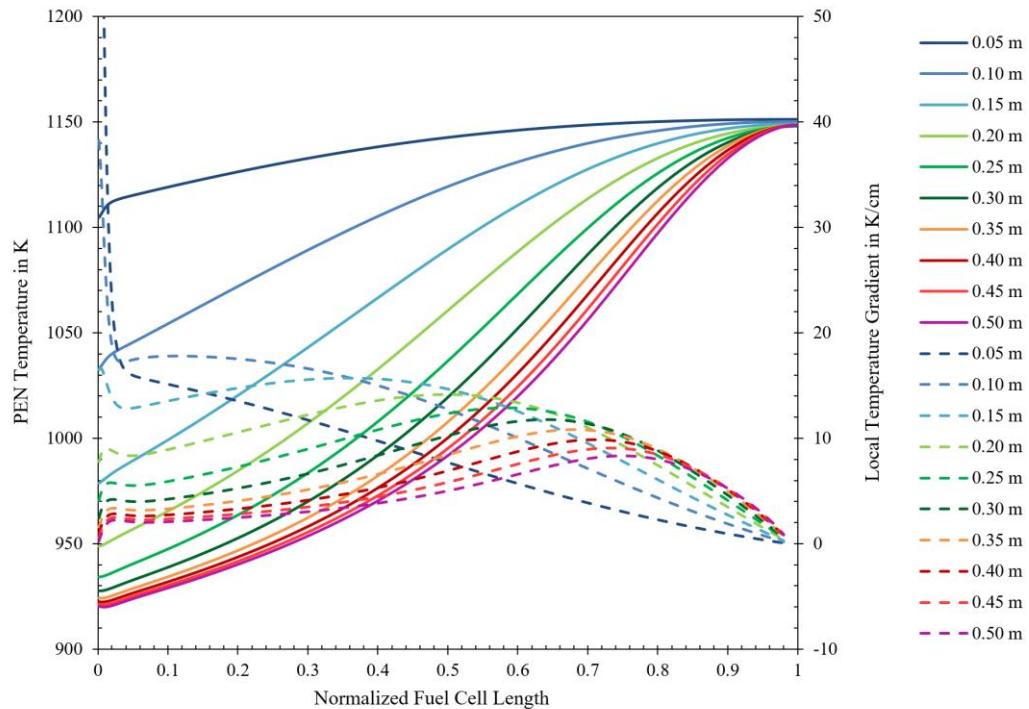


Figure 5-3: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various channel lengths.

Due to constant utilization factors, the PEN temperatures at the cell outlet are almost identical and independent of the cell length. However, the PEN temperature at the inlet varies substantially. Total temperature difference between inlet and outlet are reduced for smaller cell sizes. This is due to the shorter distance between inlet and outlet which has similar effects as increased thermal conductivity at constant cell length. An increased PEN temperature at the inlet ultimately accelerates reaction kinetics and leads to higher current densities. Important for the operation of SOFCs are temperature gradients. The temperature gradient curves show three major trends. At the inlet of the channel, impacts of the incoming gas streams can be observed. Fuel and air enter the fuel cell at 973.15 K. If the PEN temperature is higher than 973.15 K, the incoming streams heat up by absorbing heat from the PEN structure. Thus, a rapid temperature increase is observed at the inlet. Conversely, if the PEN temperature is less than the gas inlet temperature, the PEN structure absorbs heat from the incoming gas streams which leads to the opposite effect on the thermal gradient. In general impacts from inlet steam temperatures on the PEN structure are limited to the first 3% of the channel length. The second major trend is a dip in the temperature gradient curve which is observed in the first half of the channel. This dip is caused by the endothermic internal reforming reactions. Longer

cells have a higher gas velocity (less stacks needed) and internal reforming is more spread out while in shorter cells reforming is predominantly found in vicinity of the inlet. Thirdly, the irreversibilities in the electrochemical reaction of hydrogen lead to a heat release. Reaction kinetics are improved at higher temperatures and higher hydrogen partial pressures; thus, this effect is more pronounced towards the second half of the channel leading to a temperature increase (also, because of the absence of the reforming reaction). All the way to the end of the channel, the temperature increase levels off due to the reduced driving force and the temperature gradient approaches a value of zero.

As mentioned before, shorter cells exhibit a higher current density. This is beneficial as it reduces the cell area needed to achieve the desired fuel utilization. However, the increased total number of cells needed outweigh the benefits of the increased current density. Especially assembly, testing and conditioning costs are significantly increased when considering smaller cells. A cost minimum is reached between 0.15 m and 0.20 m. After that, the cost remains almost constant and savings in cell manufacturing are countered by the reduced current density. The results of the cost analysis are shown in **Figure 5-4**.

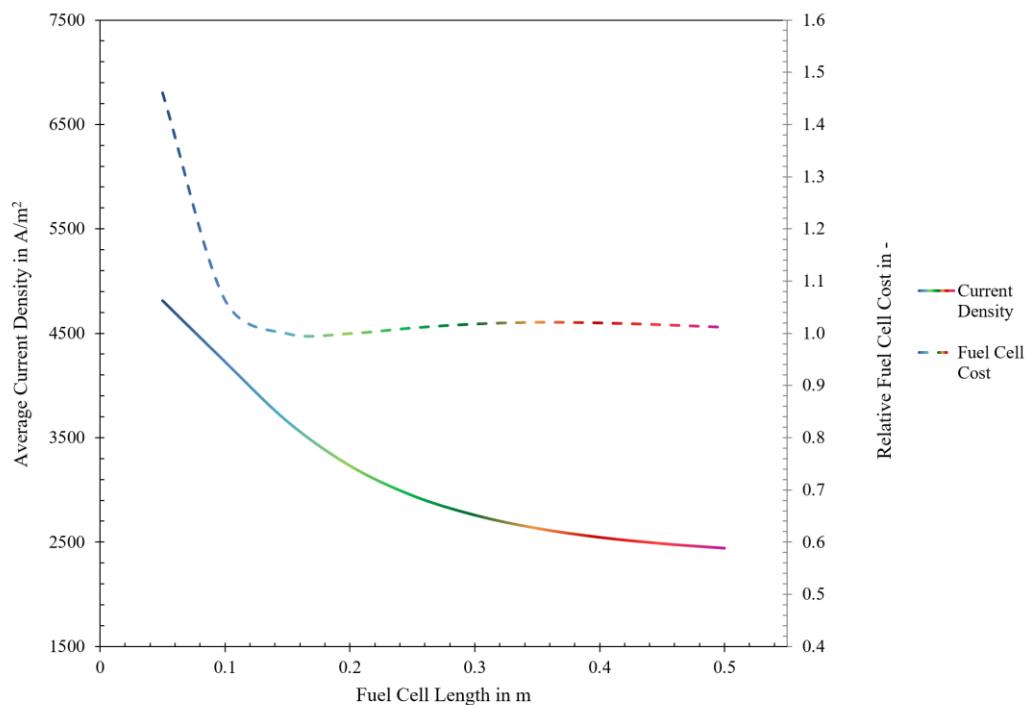


Figure 5-4: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of cell length.

Cell Width. The cell width is changed by changing the number of channels that make up the cell. The number of channels has been varied from 15 to 85. Since the channels operate in parallel and the fuel utilization is held constant the inlet condition for an individual channel does not change (more channels per cell but less cells overall). Thus, changes in temperature profile and current density are not expected. However, manufacturing cost will be impacted by this change (similar to cell length). If the cell is not wide

enough a significant increase in costs associated with single cell and or stack handling, e.g., assembly, testing and conditioning, lead to an overall cost increase. Since increasing the number of channels per cell does not impact the current density, a significant decrease in cost is obtained. Note that the fuel and air utilizations are being held constant; thus, addition of more channels to the cell reduces the number of cells while the inlet conditions to each channel do not change. The relationship between cost and number of channels is illustrated in **Figure 5-5**.

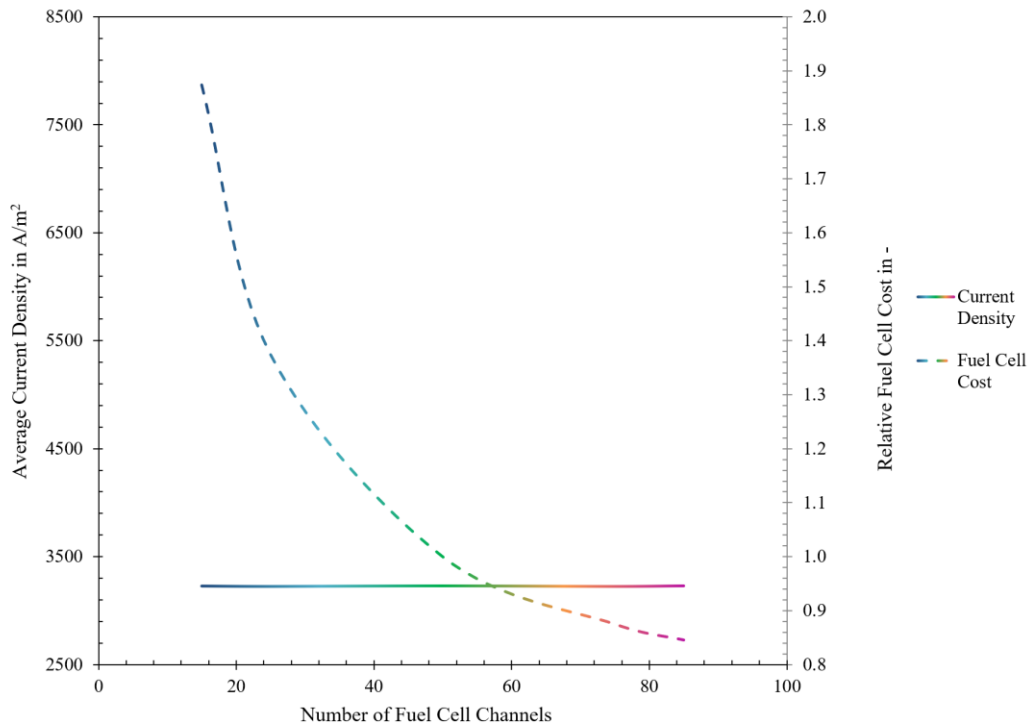


Figure 5-5: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of cell width.

Rib Width. The rib width was varied from 0.5 mm to 7.0 mm. A wider rib promotes heat transfer from the end of the channel to the front but reduces reforming reactions while still allowing H_2 to react electrochemically. Consequently, the heat absorption by the endothermic reforming reaction is reduced and the average cell temperature increases with wider ribs. Temperature gradients on the other hand are reduced due to the increased heat transport from the end of the cell to the front. The effects of rib width on cell thermal management are shown in **Figure 5-6**.

Higher cell temperatures enhance ion transport through the electrolyte and increase the current density. Although wider ribs increase the cost of an individual cell due to their larger size and increased rib material used, the higher current density is able to decrease the number of stacks needed and ultimately results in a cost reduction when wider ribs are used. The impact of rib width on current density along with its impact on specific fuel cell cost are shown in **Figure 5-7**.

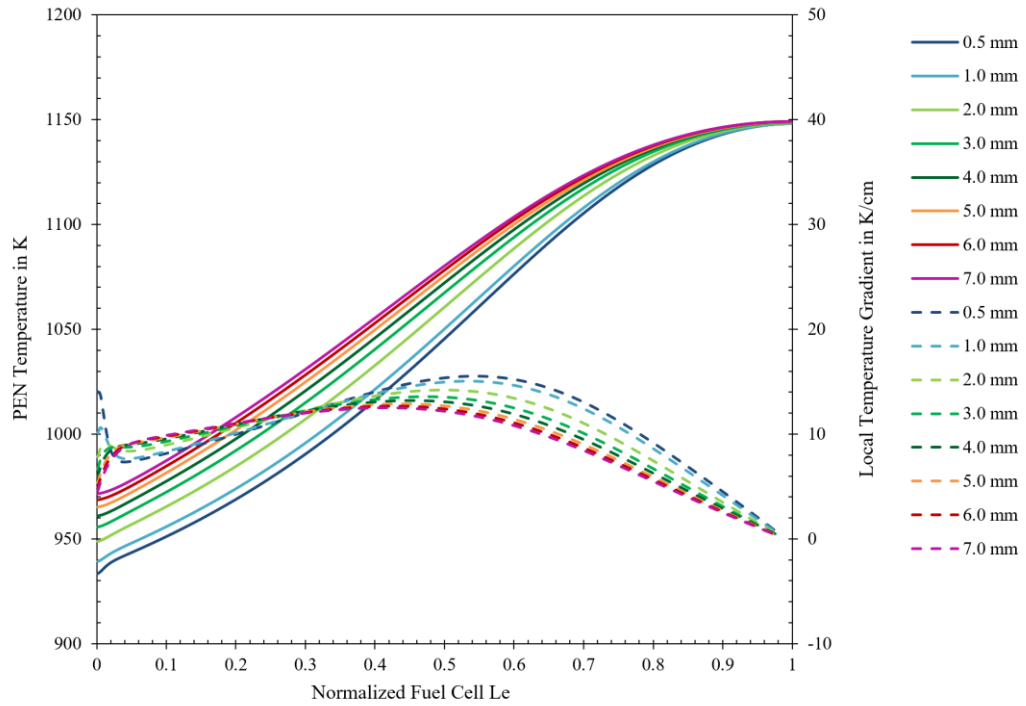


Figure 5-6: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various rib widths.

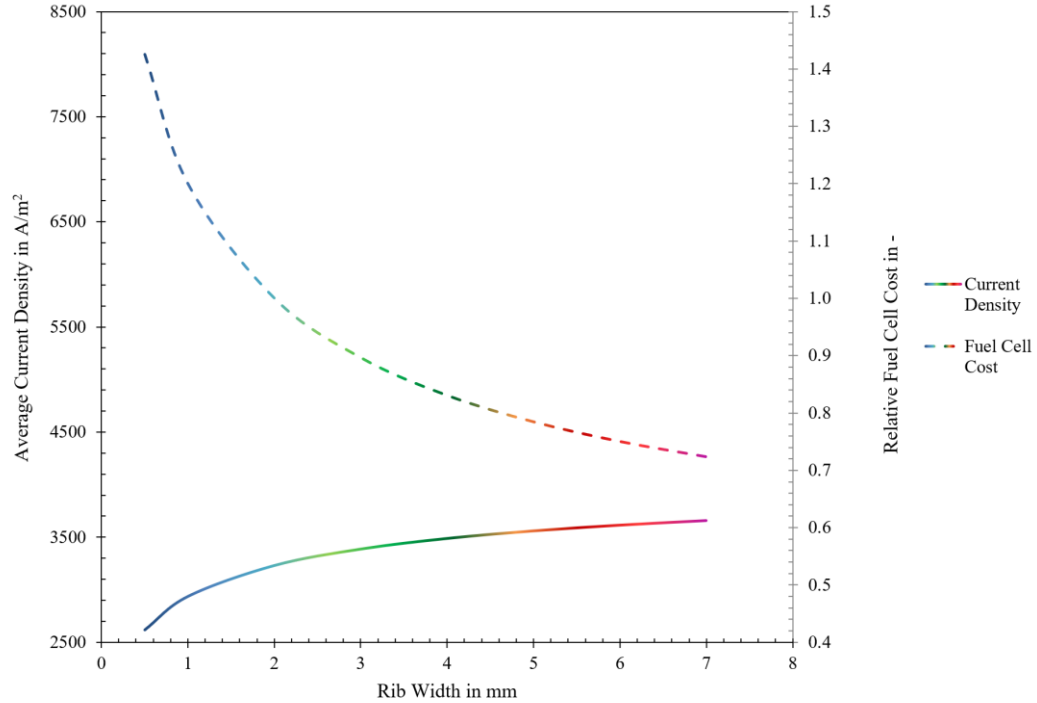


Figure 5-7: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of rib width.

Interconnect Thickness. The interconnect thickness has a similar effect as the rib width; however, only impacts the heat conduction from the end of the cell to the front and has no influence on the reforming reaction. Increasing the thickness of the interconnect increases the average cell temperature due to better heat transport and lowers the absolute temperature difference of the PEN structure along the channel. As a result, thermal gradients in the middle of the cell are reduced. The increased PEN temperature at cell inlet, however, leads to a more pronounced interaction with the cooler incoming air increasing temperature gradients at the inlet (see **Figure 5-8**).

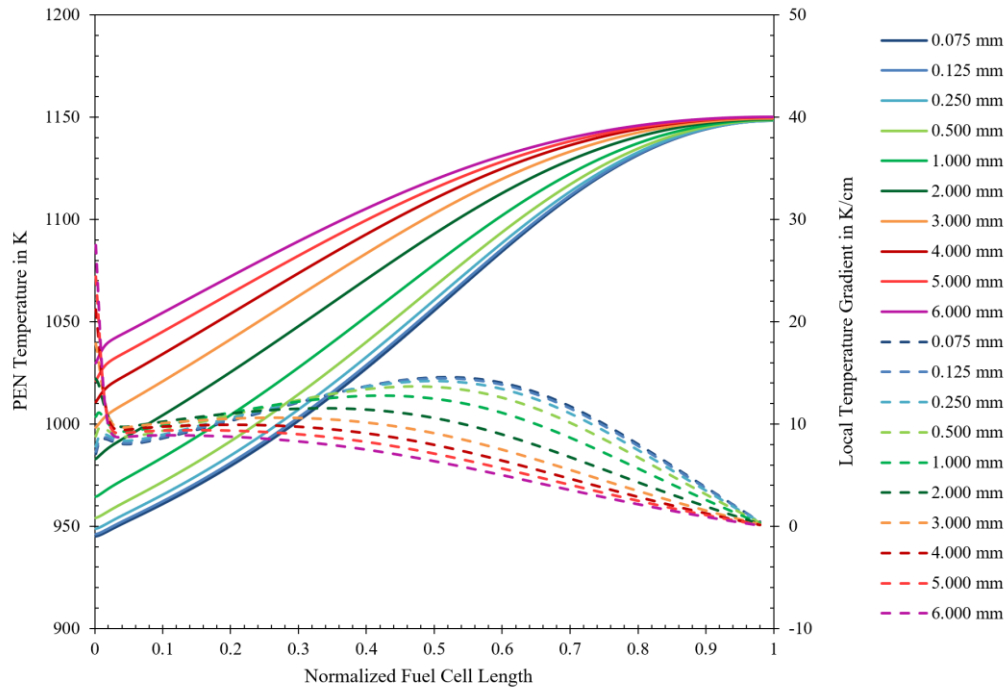


Figure 5-8: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various interconnect thicknesses.

As previously mentioned, higher average cell temperatures result in faster kinetics and increase the current density. This effect is also observed when increasing the interconnect thickness. The specific fuel cell cost initially decreases with increasing interconnect thickness up to a thickness of approximately 1 mm. Beyond that, the increase in current density and the resulting reduction in required stacks to obtain the desired fuel utilization cannot compensate for the increased cost of the thicker interconnect plate. Thus, higher costs are obtained for interconnect thicknesses of greater than 2 mm. The results are summarized in **Figure 5-9**.

Cathode Channel Height. Increasing the cathode channel height increases the residence time of the air but more importantly leads to a higher rib height and more material for heat conduction in channel direction. Thus, the effects of increasing the channel height are similar to increasing the interconnect thickness and the same explanations are valid. General trends are that increasing the cathode channel height results in

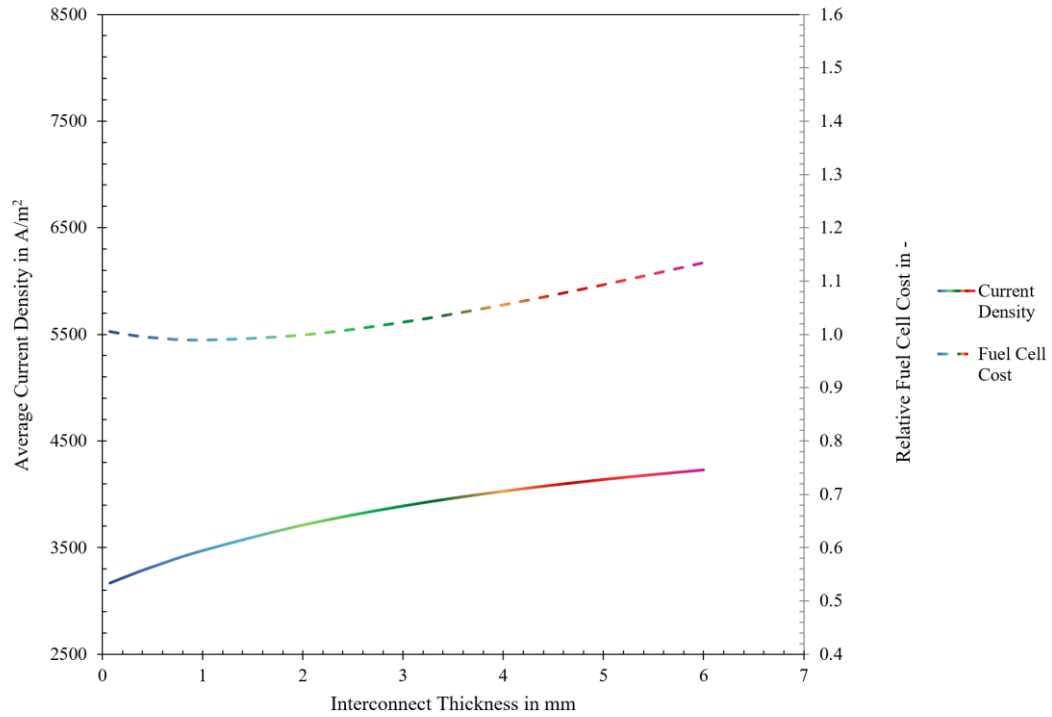


Figure 5-9: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of interconnect thickness.

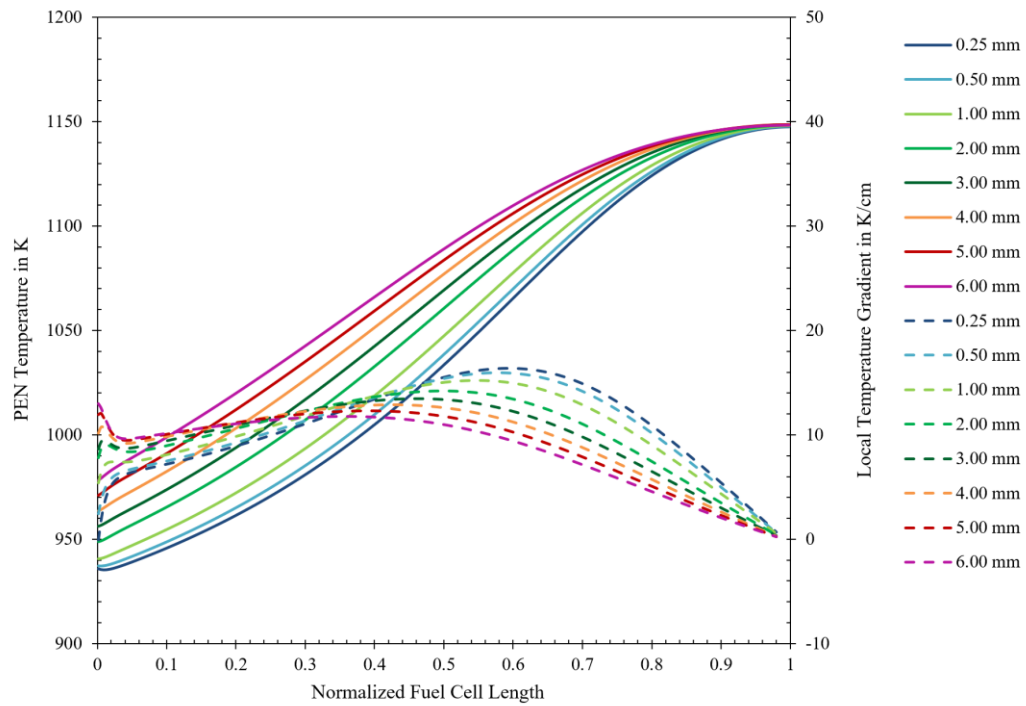


Figure 5-10: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various cathode channel heights.

higher average cell temperatures and lower thermal gradients in the PEN structure. These trends are illustrated in **Figure 5-10**.

Due to the enhanced heat transport, the average cell temperature is increased and thus, the current density improves with higher channel height. The cost analysis exhibits a minimum around 2 mm. Cathode channels higher than 2 mm cannot sufficiently raise the current density to offset the increased cost due to the increased cell height. The current density and cost results are shown in **Figure 5-11**.

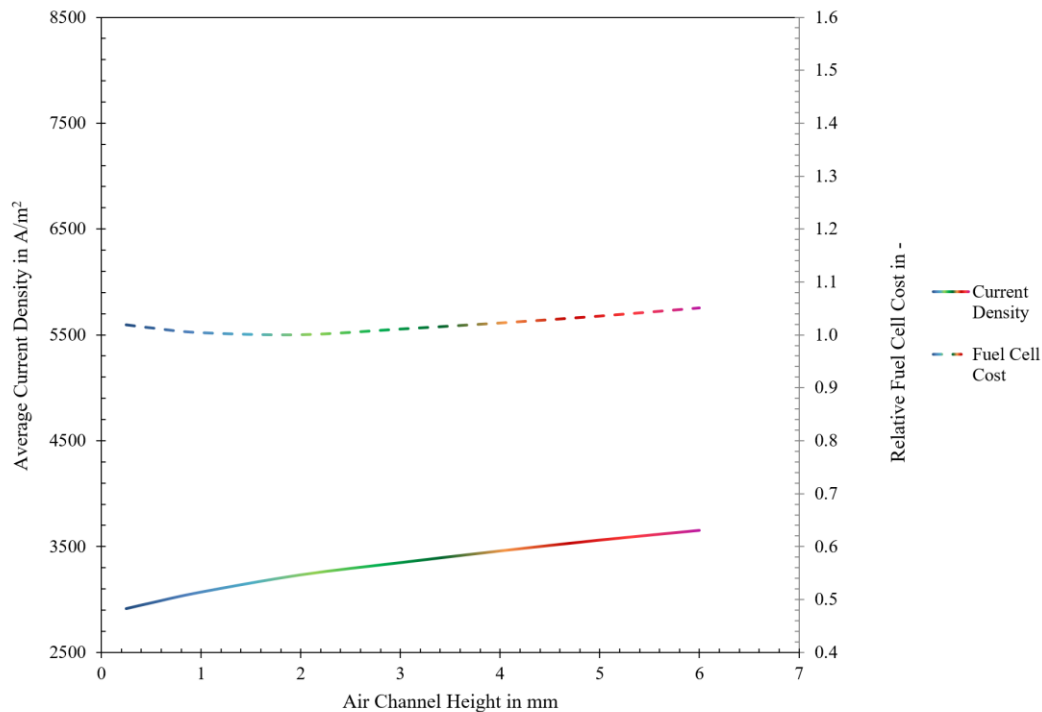


Figure 5-11: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of cathode channel height.

Anode Channel Height Analogous to the cathode channel height increasing the anode channel height has similar impacts on the temperature profiles. Higher anode channel height increases the fuel's residence time, but most importantly increase the rib material which allows better heat transport between the outlet and inlet of the cell. Thus, the overall temperature difference as well as the thermal gradients in the center of the cell are reduced with increasing anode channel height. However, entrance effects might increase thermal stresses at the channel inlet. The results of the thermal cell management are shown in **Figure 5-12**.

The cost analysis results of the anode channel height are different from previous results on the cathode channel height. Due to the more expensive cell framing, the specific fuel cell cost increases monotonically with the channel height despite the fact that the better heat conduction increases the current density. The cost analysis results of the anode channel height are presented in **Figure 5-13**.

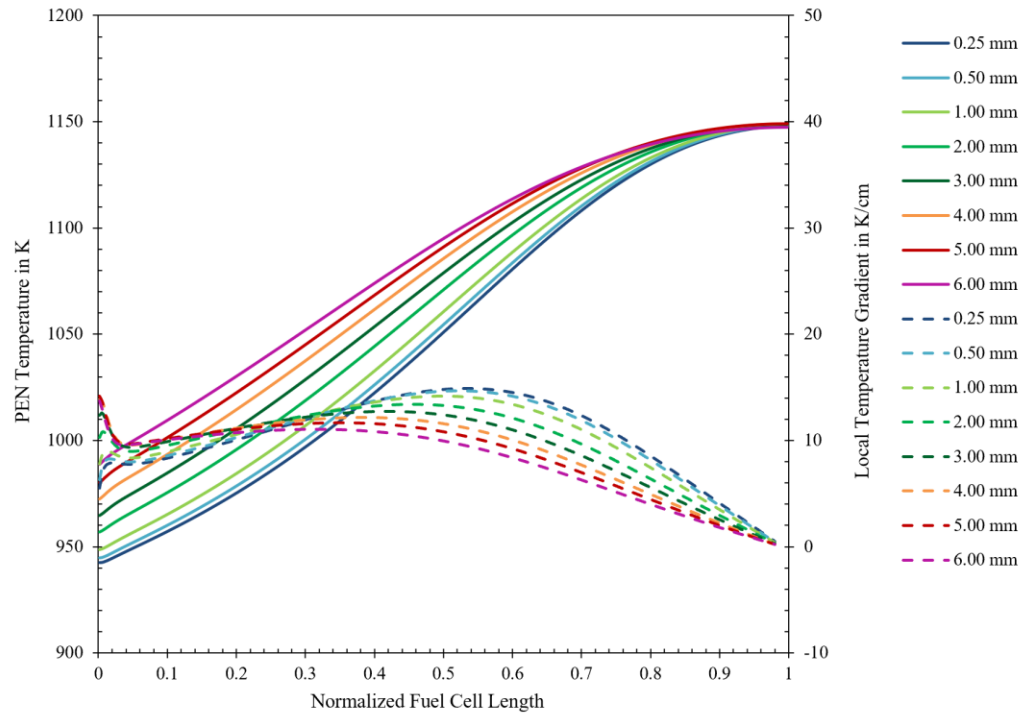


Figure 5-12: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various anode channel heights.

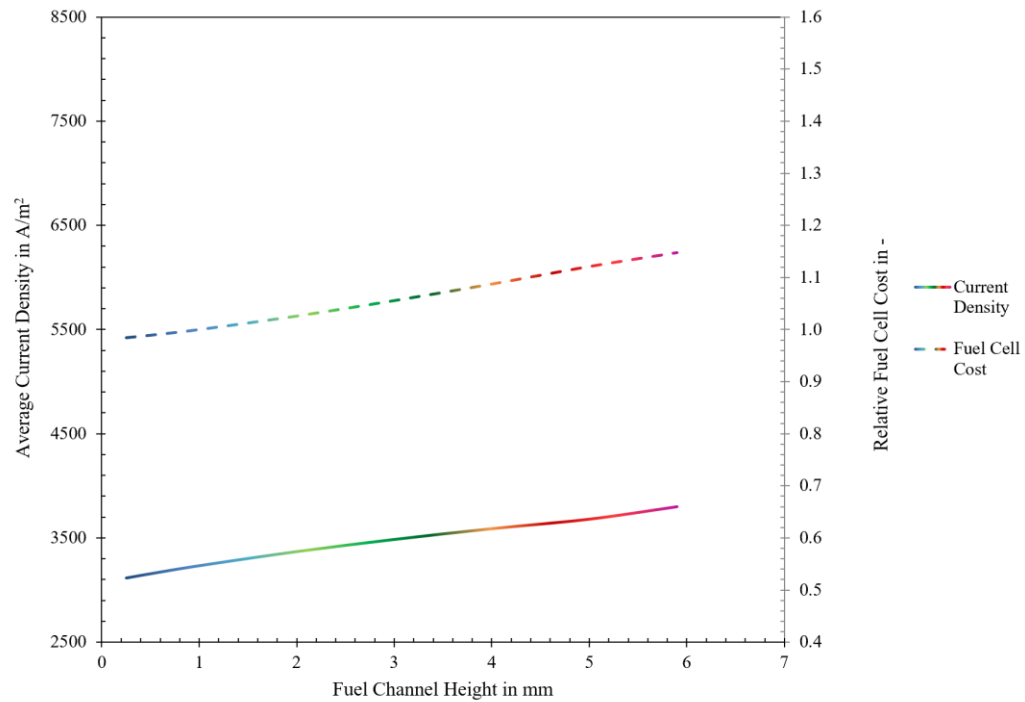


Figure 5-13: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of anode channel height.

Anode and Cathode Thickness. The impacts of anode thickness and cathode thickness on performance and thermal cell management are minor. The anode thickness mostly influences diffusion losses related to the fuel. However, in SOFCs ohmic losses and activation losses on the cathode dominate the loss terms which results in negligible performance reduction due to the diffusion terms. A thicker anode layer allows for a minor increase in heat conduction. The reduction of thermal gradients is small. Thus, the current density remains almost constant and so does the specific fuel cell cost. For completeness, temperature profiles and fuel cell cost are shown in Appendix A, **Figure A-1** and **Figure A-2**. The results rely on the assumption that the reforming reaction only occurs at the top layer of the anode as proposed by Achenbach [243]. Other authors [329], however, suggest that the thickness of the anode layer impacts the reforming reaction and thus, the H_2 concentration and current density.

The cathode thickness on the other hand increases diffusion losses on the cathode side which have negligible impact on the SOFC performance. Although a thicker cathode layer reduces the activation losses on the cathode, the relative scale of this improvement is small. Furthermore, a thicker cathode layer has almost no impact on the temperature profiles which results in an almost constant current density. Cost associated with the thicker cathode layer are minor leaving the specific fuel cell cost essentially constant. The graphs of the temperature profiles and fuel cell cost are shown in Appendix A, **Figure A-3** and **Figure A-4**.

Electrolyte Thickness. Oxygen ion transport through the electrolyte constitutes the largest loss term (ohmic loss) in SOFCs. Thus, electrolyte thickness plays a crucial role in the design of SOFCs. The thicker the electrolyte, the larger the resistance for the O^{2-} ions traveling through the electrolyte. As a result, the driving force is reduced, and the current density drops. This lower current density combined with the increased cost of the thicker electrolyte leads to a steep increase in specific fuel cell cost, as shown in **Figure 5-14**.

A lower current density results in a higher number of cells needed to reach the desired fuel utilization. Consequently, the flow velocity is reduced and the residence time increased. As a result, the endothermic reforming reaction moves closer the channel entrance which leads to an overall higher average cell temperature. Higher temperatures in general increase current density but the temperature increase cannot compensate for the increased resistance caused by the thicker electrolyte. The higher resistance of O^{2-} ions slows down the oxygen transport and leads to slower kinetics of the electrochemical H_2 reaction. As a result, the temperature gradient in the middle section of the cell is decreased. The results are depicted in **Figure 5-15**.

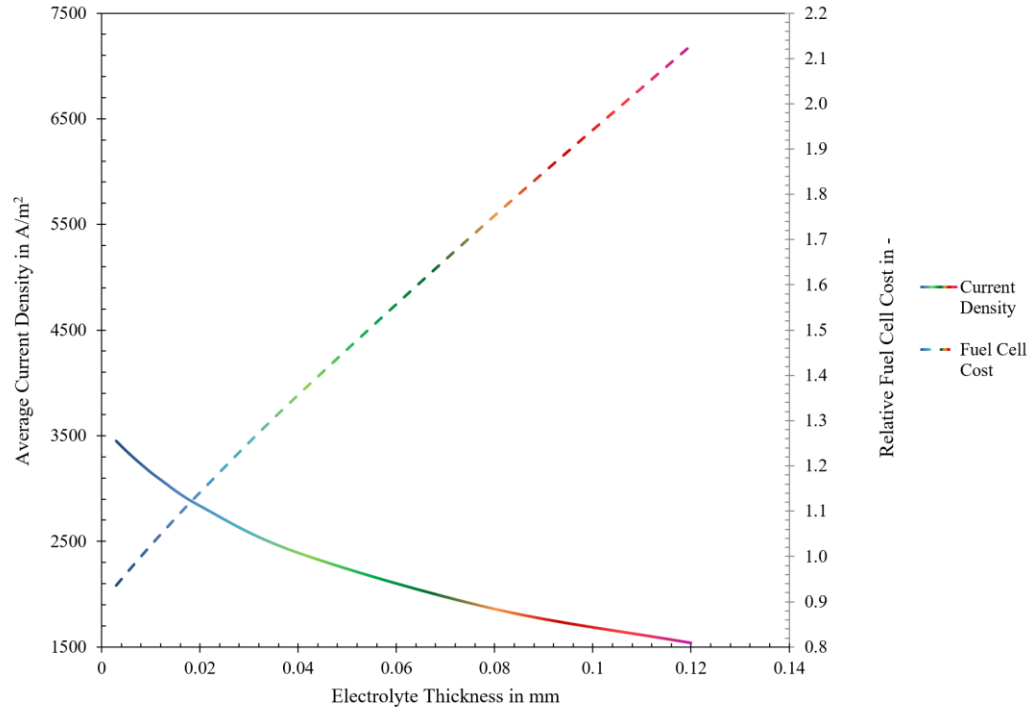


Figure 5-14: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of electrolyte thickness.

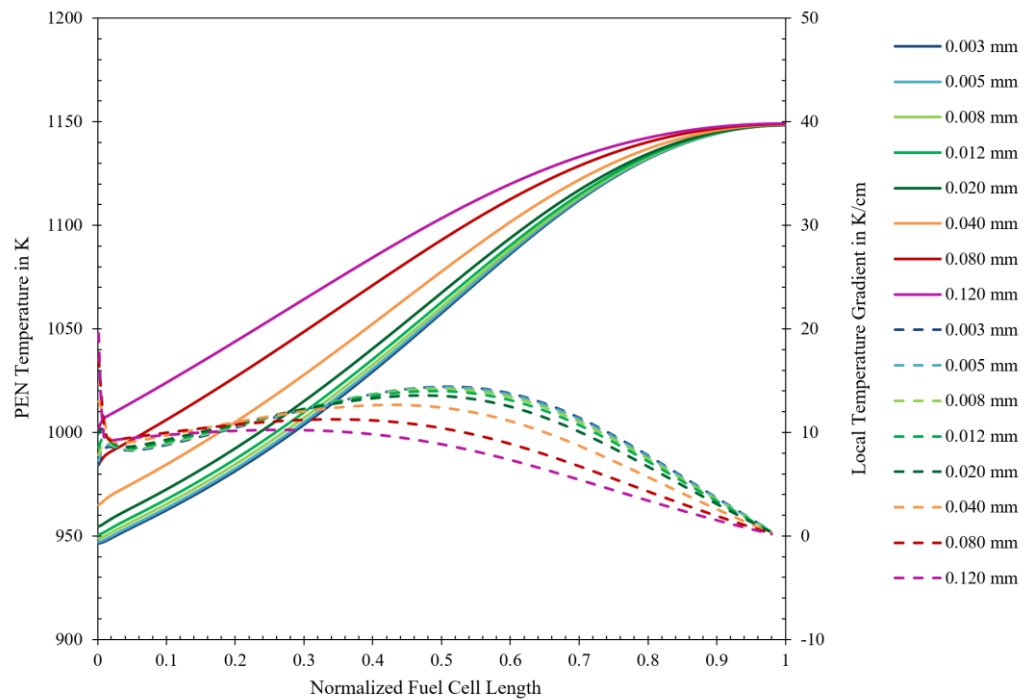


Figure 5-15: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various electrolyte thicknesses.

Pre-Reforming. Pre-reforming is an important part of SOFC operation. From an efficiency standpoint it is desirable to completely reform the CH_4 inside the fuel cell. Reforming upgrades CH_4 to H_2 and simultaneously cools the fuel cell. However, reforming tends to occur near the cell inlet and can lead to significant thermal gradients. Advantage of the configuration used in this study in which the anode off-gas is recycled back to the SOFC is that it increases the H_2 , H_2O , CO and CO_2 content of the fuel gas which is similar to the impact of pre-reforming. Thus, the impact of pre-reforming on the thermal gradients is not as significant. Also, the overall temperature difference between outlet and inlet remains relatively constant. At the same time, the PEN structure's temperature profile shifts upwards when more pre-reforming is done as shown in **Figure 5-16**. Less pre-reforming lowers the PEN temperature at the channel inlet since more CH_4 is being reformed internally which mostly occurs close to the inlet. Additionally, more cooling inside the fuel cell by the reforming reaction leads to a lower outlet temperature compared to a scenario with a higher degree of pre-reforming.

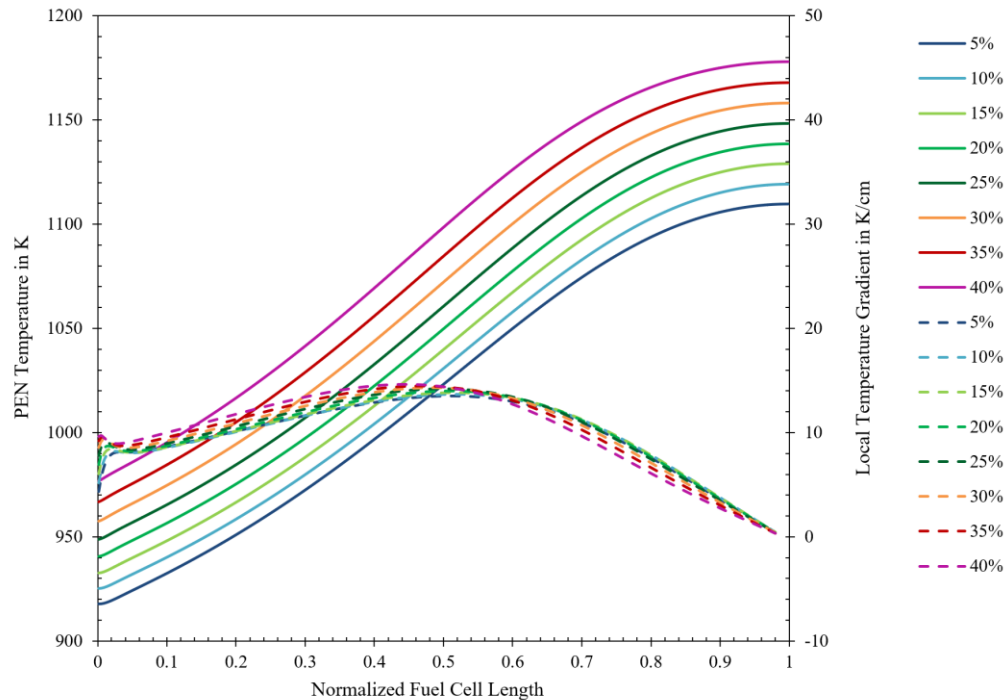


Figure 5-16: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various degrees of pre-reforming.

The higher average cell temperature achieved with higher degrees of pre-reforming ultimately helps to improve the current density. Since the SOFC cell design is not impacted by the degree of pre-reforming in this analysis, a cost reduction directly related to the current density can be observed. The results are illustrated in **Figure 5-17**.

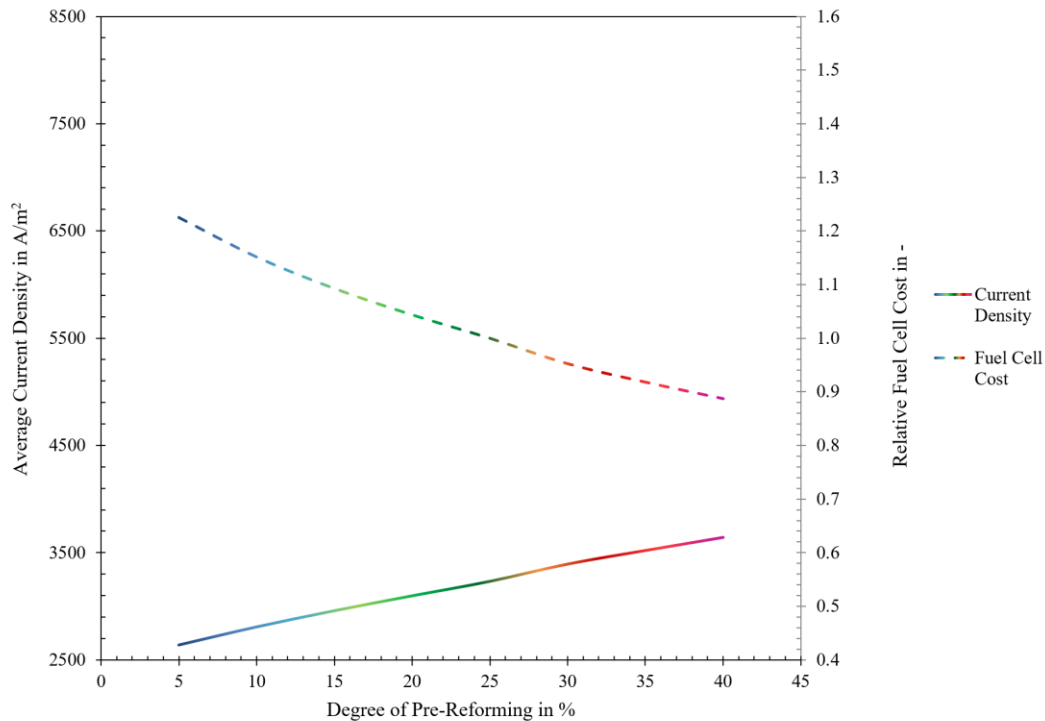


Figure 5-17: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of degree of pre-reforming.

Air Utilization. Air utilization is another important operating variable for SOFCs. Air utilization has an impact on the chemical potential difference as well as the thermal management. Typically, the amount of air needed is determined by the temperature gradients experienced in the cell. Higher air flow rates and thus, lower air utilization (with constant fuel utilization) help in cooling the cell and lead to lower cell temperatures, lower gas outlet temperatures and reduced thermal gradients as can be seen in **Figure 5-18**.

Lower air flow rates increase cell temperature which is beneficial for improving the current density as it increases the conductivity of the electrolyte. This increase in current density relates directly to a reduction in specific fuel cell cost as the cell design remains unchanged. However, for very high air utilization factors the O_2 concentration near the SOFC outlet is drastically reduced which reduces the chemical potential difference and ultimately leads to a lower current density. The maximum current density and lowest fuel cell cost is achieved at an air utilization factor of about 45%. Higher air utilization factors lead to lower current densities and higher cost. The current density and cost results as functions of the air utilization factor are shown in **Figure 5-19**.

Operating Pressure. The operating pressure in a SOFC-GT hybrid can vary as it depends upon the cycle design and the GT. If the SOFC operating pressure is increased, the chemical potential between anode and cathode increases according to Nernst-Equation, which in turn increases the driving force of the electro-

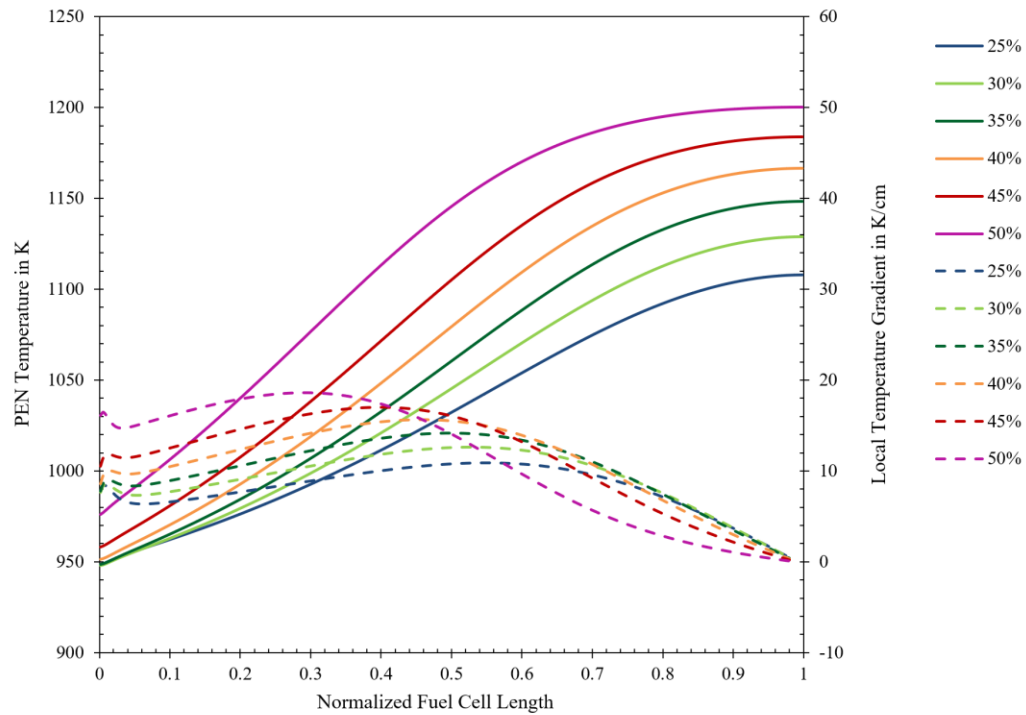


Figure 5-18: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various air utilization factors.

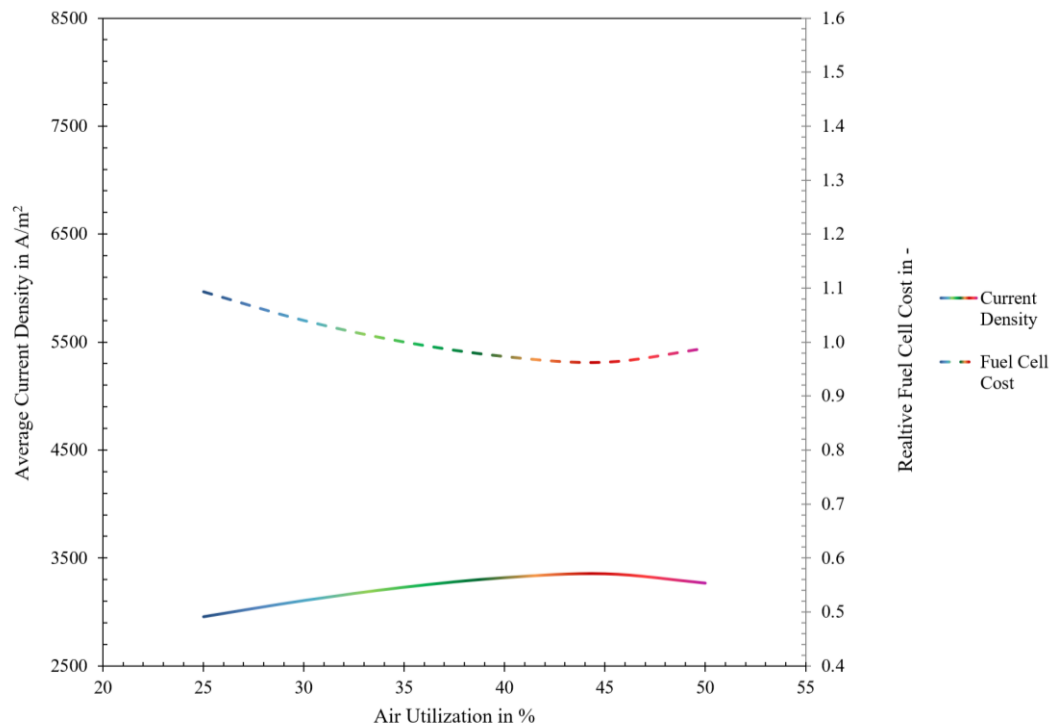


Figure 5-19: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of air utilization factor.

chemical reaction leading to an increase in current density. This increase in current density can be seen in **Figure 5-20**. However, as the pressure keeps increasing the current density levels off. At the low-pressure side, the current density will eventually reach a value of 0 and the fuel utilization (FU) of 80% at an air utilization of 35% is not attainable for pressures left of the intersection with the x-axis. The relative specific fuel cell cost shows the inverse behavior (compared to the current density). As the current density decreases more active cell area is needed in order to maintain the same FU. Thus, the specific SOFC cost increases. At low pressures, the relative fuel cell cost almost doubles compared to the reference operating pressure of 10 bar. However, this cost does not include the costs associated with the SOFC pressure vessel. Pressure vessel costs tend to increase with pressure and a cost optimum can be expected.

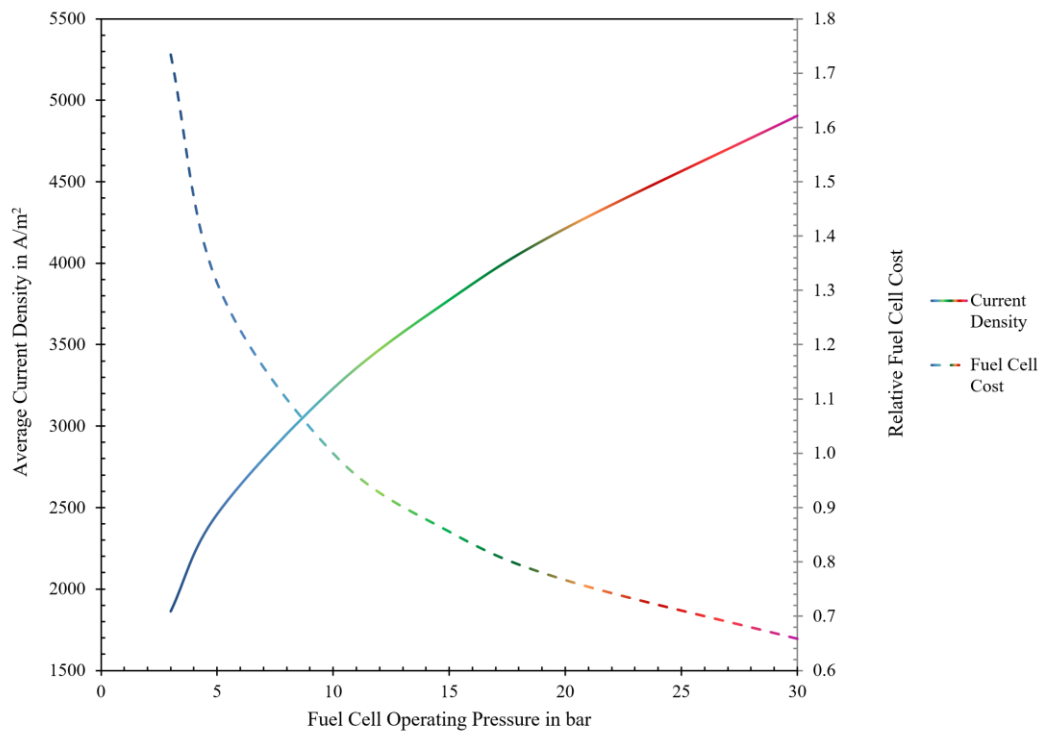


Figure 5-20: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of SOFC operating pressure.

Local temperature gradients inside the cell increase with lower operating pressure and shift towards the cell inlet (note: this is for constant FU and constant air utilization). Lower pressure operation requires more cells and reduces the flow velocity, which leads to more endothermic reforming at the cell inlet. Simultaneously, thermodynamic equilibrium favors reforming at lower pressure. The trends for the temperature profiles and thermal gradients are presented in **Figure 5-21**.

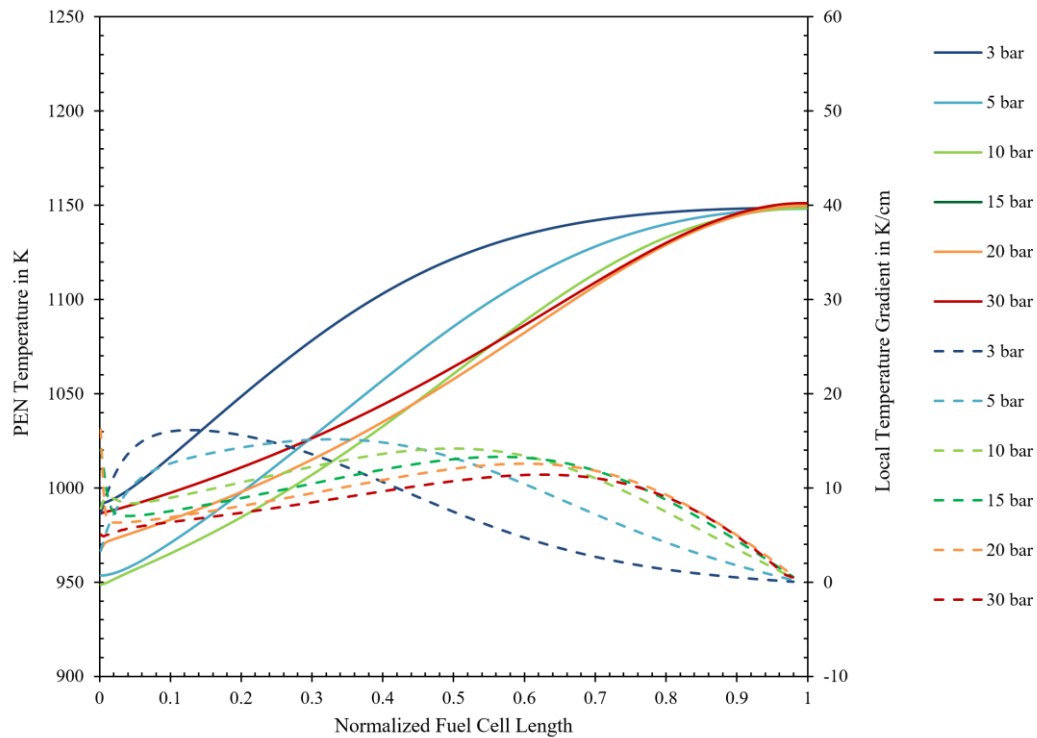


Figure 5-21: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various SOFC operating pressures.

Voltage. Lastly, the SOFC operating voltage is studied and its impact on thermal gradients and cost. The voltage in SOFCs defines the SOFC thermal efficiency and impacts the specific power output. Since the fuel utilization in this study is held constant, the current produced remains constant when varying the voltage. If the current remains constant and the voltage increases the power output goes up. On the other hand, a higher voltage reduces the chemical driving force of the electrochemical reaction and the kinetics. Thus, if the voltage is increased the current density is reduced and vice versa. The average current density in the cell decreases almost linearly from over 8000 A/m² at 0.70 V to approximately 2600 A/m² at 0.85 V. The cost associated with this lower current density increases rapidly from initially 0.48 RFC to 1.21 RFC (see **Figure 5-22**).

Temperature gradients in the cell increase with lower operating voltage. Local gradients significantly increase with the decline of internal reforming reaction towards the end of the channel as a lower voltage improves O²⁻ ion diffusion. Furthermore, the overall temperature difference between inlet and outlet increases due to lower operating voltages as the energy conversion of the fuel to electricity is less efficient. Dominant loss factors are the ohmic losses, which drastically increase due to the higher current density and

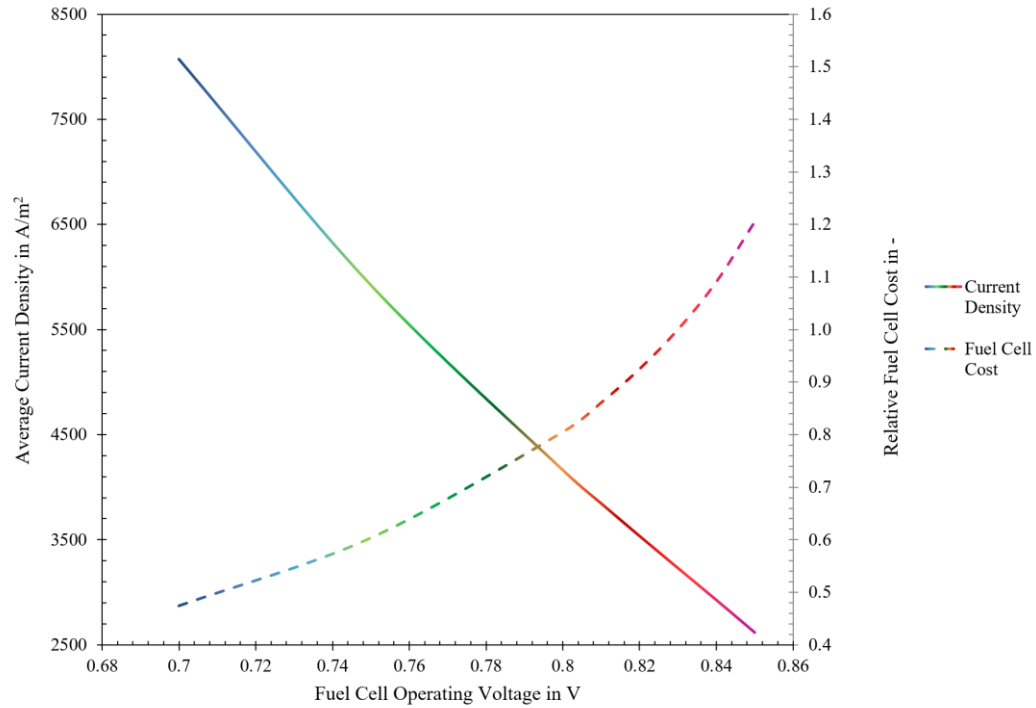


Figure 5-22: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of operating voltage.

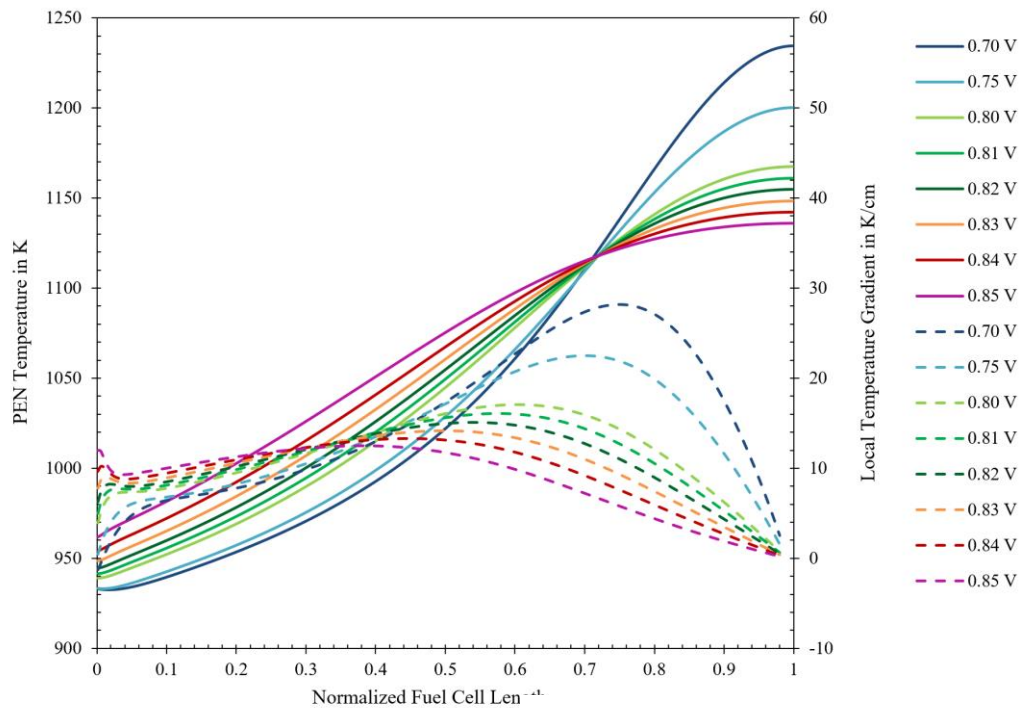


Figure 5-23: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various operating voltages.

also, diffusion losses start becoming significant. The results of the thermal analysis are provided in **Figure 5-23**.

Cost Estimation. The SOFC design impacts are analyzed in dimensionless manner and independent of the production scale. In order to estimate the stack cost, the following equation is proposed which predicts the specific SOFC cost as function of production scale and relative cost factors (determined according to section 3.2). The accuracy of this equation which is based upon the super-positioning of the multiple parameters is expected to be within $\pm 15\%$. Inaccuracies arise from the interaction of different parameters which are not captured in this cost function.

$$\frac{\$_{2020}}{\text{kW}_{\text{DC}}} = \text{CSF} \cdot \prod_{i=1}^{13} \text{RFC}_i(\text{parameter } i) \quad (\text{E } 5-1)$$

CSF is the cost scaling factor and is determined by the annual production scale as shown in **Figure 5-2**. *RFCs* are the relative cost factors and dependent upon fuel cell parameters *i*, where *i* represents the 9 geometrical parameters such as cell length, cell width, rib width, etc., and the 4 operating parameters such as pre-reforming, air utilization, operating pressure and operating voltage. The geometric parameters shown in **Table 5-1** represent typical values for a SOFC cell and have an RFC factor of 1. Equally, the operating parameters discussed in earlier possess an RFC factor of 1.

Summary

SOFC costs have been analyzed with respect to production scale, various cell design parameters, operating parameters and their impact upon the thermal cell management. Findings show that in co-flow SOFCs specific costs (\$/kW) can be effectively reduced by maximizing the cell width/number of parallel channels without impacting current density and thermal cell management. The cost optimum for the cell length was found to be around 15–20 cm. While overall temperature differences were reduced in shorter cells, local gradients were reduced in longer cells. The optimum cost of the SOFC is dependent on the maximum allowable temperature difference across the fuel cell and thermal gradient within the cells but some qualitative generalizations may be made. Increasing the rib width offers a cost-effective solution for reducing thermal gradients in the cell. Interconnect thickness and cathode channel height can be cost-effective to reduce thermal stresses under the right conditions. Under the given circumstances an interconnect thickness of 1mm has shown to offer the most cost-effective solution. On the cathode side a channel height of 2 mm has shown to have the highest cost advantage. The electrolyte thickness has shown to be a crucial part to reduce cost in SOFCs. From a cost perspective it is desirable to reduce the electrolyte thickness to a minimum. However, thermal gradients in the middle of the cell as well as overall temperature differences are larger for cells with thinner electrolyte.

The degree of pre-reforming in a SOFC-GT hybrid configuration with anode off-gas recycling has shown to be insignificant from a thermal gradient perspective; however, specific SOFC costs are reduced for higher degrees of reforming. Air utilization reaches a cost optimum around 45% and the operating voltage shows that lower cell voltages increase overall temperature differences and local thermal gradients while reducing specific SOFC cost. As a result of this analysis, the cell parameters presented in **Table 4-13** have been identified as favorable cell design parameters and are used for the design of the SOFC-GT hybrid systems studied in this work. Results have been published in [317], [338]

5.1.2 Thermo-Economic Analyses of SOFC-GT Hybrids Considering Thermal Cell Gradients

Introduction

Over the recent years significant research efforts have been undertaken to study SOFC-GT hybrid systems in more depth. Eisavi et al. [85] studied exergy destruction in SOFC-GT hybrids and Krummrein et al. [122] compared the operating range of a hybrid system running on methane to biogas. Advanced systems with GT air inlet cooling or oxy-combustion and CO₂ capture have been investigated by Sghaier et al. [107] and Mahisanana et al. [112]. An important consideration when designing SOFC systems are thermal gradients within the cell and/or the stack that occur during the operation of the system; however, only a limited number of hybrid system studies have employed dimensional SOFC models that can provide insights into thermal cell gradients and can identify critical operating conditions [120], [123], [335], [336]. Wongchanapai et al. [123] identified strong thermal gradients at the cell inlet due to the endothermic reforming reaction. Cuneo et al. [120] investigated the impact of cell degradation on the temperature profile and Zhang et al. [335] studied thermal cell gradients during load transients. A study conducted by Wu et al. [336] focuses on strategies to better control thermal gradients during SOFC operation. While dimensional SOFC models have been used in some cases, a systematic investigation on how thermal gradients in SOFC-GT hybrid systems influence the hybrid's operating condition is yet to be conducted.

Another major hurdle to widespread adoption of SOFC systems is their current cost due to relatively small production volumes. Moreover, SOFC cost projections for large scale production are still difficult to make. Consequently, and as described below, most system studies rely on conceptual estimates or cost targets set by institutions like the U.S. Department of Energy.

Such a cost target [154] was used in a SOFC-GT hybrid study by Curletti et al. [127]. Rao and Samuelsen in their investigation [106] used a SOFC-GT hybrid cost of 1000 \$/kW as projected by Siemens-Westinghouse for their hybrid. Another hybrid study conducted by Kim and Kim [124] uses a substantially higher cost of 7000 \$/kW which is assumed to be reduced to 3500 \$/kW for large scale production. An area specific SOFC price of 1,700 \$ per m² of active cell area was employed by Siefert and Litster [128]. Arsalis

[155] estimated the SOFC cost based on the active cell area and operating temperature which was also used in [85], [90]. A cost correlation proposed by Massardo and Magistri [97] is based upon the specific power (kW/m² cell area) which results in specific SOFC costs (cost on a per kW basis) stretching from 700 – 2,200 \$/kW. In cases where 0D SOFC models are employed these simplified cost estimates are useful since 0D models do not provide the data necessary to construct more accurate cost estimates. However, due to the wide operating range of SOFC systems, especially when considering pressurized operating conditions, these simplified cost estimates can be imprecise as these cost estimates or methodologies are missing a more fundamental basis.

Even in cases where the active cell area is being considered, the above-mentioned cost correlations based upon specific cell area cannot scale the various cost components of a SOFC stack correctly; while some stack parts might scale with the active cell area, others such as enclosures or assembly might scale on a different basis.

More recently, more sophisticated SOFC cost predictions have become available considering various individual cell components and manufacturing processes. The Battelle Memorial Institute [156] has developed a detailed SOFC cost estimation method based on individual manufacturing processes and established cost projections for small and large scale SOFC production of the studied atmospheric pressure SOFC combined heat and power systems. In [317] this methodology has been adopted to identify cost effective cell designs while reducing thermal gradients; however, this detailed economic modeling has not been used to study SOFC-GT hybrid systems until now.

In summary, various SOFC-GT hybrid systems have been investigated in literature, but current literature mostly employs 0D SOFC models ignoring thermal gradients inside the cell. Although, some studies utilize 1D SOFC models, no systematic investigation of the impact of operating pressure, fuel utilization (FU) and operating voltage upon thermal efficiency and cost of electricity (COE) of a SOFC-GT hybrid system has been conducted while considering thermal gradients. This study is filling this gap in literature. Additionally, a detailed SOFC cost model is employed to predict SOFC cost more accurately under given operating conditions as opposed to the simplified cost estimation currently applied in literature.

Methodology

Steady state operating conditions are established, which serve as input for the thermo-economic SOFC-GT hybrid evaluation. Three different recuperation strategies for heating the air discharged from the GT compressor are considered and various operating conditions such as operating pressure, FU and SOFC operating voltage are studied under the constrain of local thermal cell gradients and overall temperature difference in order to find the optimal operating condition with respect to efficiency and COE. The following section provides a summary of the modeling approach.

Design Basis

SOFC-GT hybrid systems are relatively fuel flexible as long as the fuel can be delivered to the SOFC in a gaseous state and is free of contaminants which may harm the catalyst or anode. Natural gas, after preconditioning, is not only a viable fuel but can be used advantageously in an SOFC due to its high operating temperature which promotes internal reforming. The endothermic reforming reaction converts thermal energy to chemical energy by the formation of hydrogen inside the SOFC. This increases the efficiency and simultaneously provides cooling to the cell. The natural gas composition used in this work is provided in **Table 4-4**.

Solid Oxide Fuel Cell Hybrid Configuration. In a SOFC-GT hybrid, a GT is used to pressurize the SOFC while serving as the “bottoming cycle”. Higher operating pressures allow the SOFC to reach higher FU, higher operating voltage, and higher current density. Moreover, high temperature heat generated in the fuel cell itself together with heat generated from unconverted fuel can be utilized in the expander section of the GT. This synergy between the SOFC and GT allows the SOFC-GT hybrid to reach extremely high efficiencies. Consistent with previous work, a configuration with anode off-gas recycle has been selected as this can provide the needed O-atoms to avoid carbon deposition. The recycle amount is determined based upon thermodynamic gas phase equilibrium considerations [252] to avoid the formation of whisker carbon. The system’s flowsheet is provided in **Figure 5-24**.

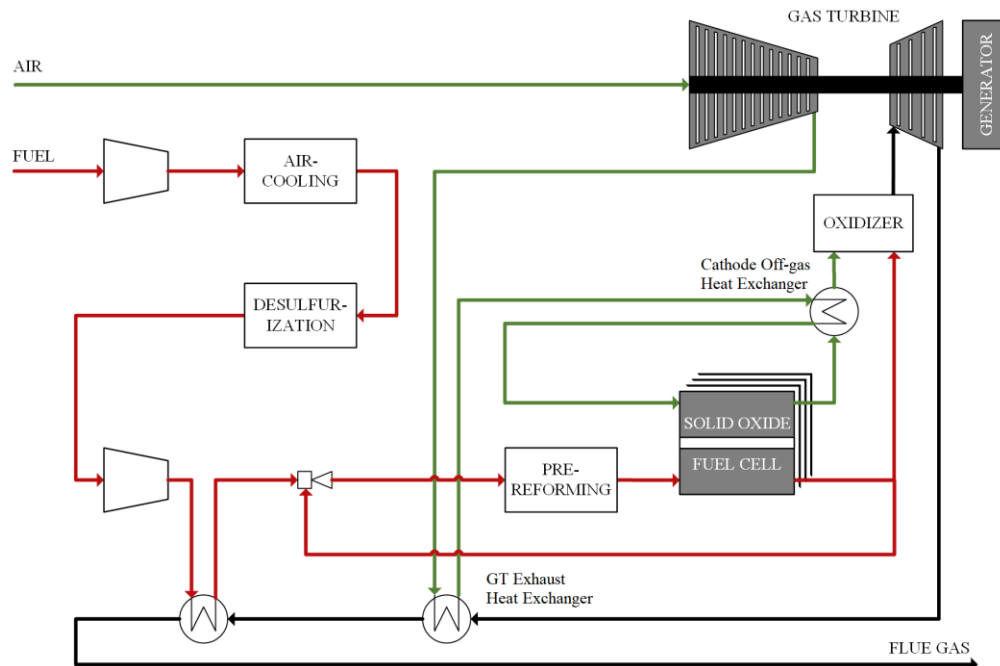


Figure 5-24: Solid oxide fuel cell - gas turbine hybrid configuration.

The ambient air is considered at ISO conditions (1.013 bar, 15 °C) and its composition can be found in **Table 4-2**. Three different heat recuperation strategies for air heating (prior to the SOFC) are employed in this study: (I) using only the GT exhaust to heat the air (no cathode off-gas heat exchanger), (II) using the GT exhaust as much as possible and the cathode off-gas heat exchanger as needed, and (III) using only the cathode off-gas heat exchanger (no GT exhaust gas recuperation).

Power Block Modeling. The anode recirculation is achieved via an eductor, which uses a high pressure primary motive fluid to compress the anode recycle by converting pressure from the primary motive fluid into kinetic energy and entraining the lower pressure stream before reconvertng the kinetic energy back into static pressure. Most models employed in system simulations are 1D thermodynamic models; however, these models tend to over predict the performance of eductor pumps as they do not account for the velocity distribution due to boundary effects. In this study, a model that accounts for the velocity profile in the suction chamber and simultaneously sizes the eductor has been adopted.

The pre-reformer for all cases is designed to convert 20% of the incoming CH₄ to H₂ and CO. Additionally to the reforming reaction, the reforming catalyst promotes water gas shifting. Pre-reforming is needed in SOFC applications to reduce the temperature gradients inside the fuel cell in addition to protecting the fuel cell from carbon deposition. The pre-reformer is integrated in the manifold-pipes and is thermally coupled with the SOFC furnace via radiation similar to the design described in [254]. In this design, the catalyst is finely dispersed on the pipe wall. The conversion in this setup is limited due to the small surface area available inside the pipe compared to a fixed bed reactor, however, it is desired to mostly pre-reform higher hydrocarbons which are present in low concentrations and only small amounts of methane in order to take advantage of direct internal reforming inside the SOFC. In this design the conversion can be either limited by the amount of catalyst present or the heat transfer from the SOFC to the reformer tube.

The SOFC is a planar, anode supported cell operating in co-flow and has been extensively described in the model section. The cell design is based upon previous work on the cell's thermal management and cell cost. Cell geometrical parameters are summarized in **Table 4-13**. The SOFC operating conditions for this study are summarized in **Table 5-2**.

The GT is modeled as compressor, combustor, and expander with compressor and turbine maps as previously described. The GT is assumed to be a customized GT for this specific application and is considered to be operated at its design point for every case in order to evaluate the full potential of every configuration. Costs of this GT vary depending upon the operating conditions, such as pressure ratio and TIT. Details regarding the GT cost and other economic aspects can be found in the section Economic Analysis. The efficiencies and compressor air flow are obtained from calibration of a commercial 1.8 MW

engine which is sufficient to supply high pressure air for a SOFC-GT hybrid plant of about 10 MW. The bearing losses and generator losses are lumped together into the “mechanical” efficiency. A summary of key GT operating parameters is also provided in **Table 5-2**.

Table 5-2: Summary of Internal Reforming Solid Oxide Fuel Cell and Gas Turbine Operating Parameters

SOFC Operational Parameters	Value	Unit
Operating Pressure	1–20	bar
Operating Voltage	0.82 & 0.85	V
Inlet Temperature Anode	973.15	K
Inlet Temperature Cathode	973.15	K
Max. local PEN Temp. Gradient	15	K/cm
Overall PEN Temp. Difference	150	K
GT Operational Parameters	Value	Unit
Compressor Polytropic Efficiency	84.85	%
Expander Polytropic Efficiency	83.31	%
Compressor Mechanical Efficiency	98.04	%
Expander Mechanical Efficiency	98.04	%
Combustor Pressure Drop	4	%
Combustor Heat Loss	1	%-LHV

Results

Hybrid Operating Range. Before evaluating the performance, the operating window is evaluated for two operating voltages within a pressure range of 3-20 bar and a FU of 80-100%. For the entire analysis, the operating voltage is held constant at 0.82 V or 0.85 V, while the number of stacks has been adjusted to obtain the desired fuel utilization. The magnitude of air flow rate or air utilization is a result of the thermal cell management and is adjusted to operate the SOFC closely to the cell’s thermal constraints as specified in **Table 5-2**.

Three different air pre-heating strategies are employed: (I) using only the GT exhaust to heat the air (no cathode off-gas heat exchanger), (II) maximizing use of the GT exhaust and the cathode off-gas heat exchanger as needed, and (III) using only the cathode off-gas heat exchanger (no GT exhaust gas recuperation). It is important to note that the air pre-heating strategy III can also operate in the same region as the air pre-heating strategies I and II. Similarly, the combined air pre-heating strategy II can also operate in the operating window of the air pre-heating strategy I. However, from an efficiency standpoint it is desirable to recover as much heat from the exhaust gas as possible to pre-heat the air to the desired SOFC inlet temperature as this leads to an increase in the GT’s TIT (less cooling of the cathode off-gas before it enters the oxidizer of the GT). Thus, the air-preheating strategies I and II are used when possible and air pre-heating strategy III is employed where no GT exhaust gas recuperation is feasible.

As it can be seen from **Figure 5-25**, GT exhaust gas heat recuperation is only possible at low pressure ratios. Low pressure ratios result in lower GT compressor outlet temperatures, which allows recuperation of more heat from the GT exhaust. Additionally, higher pressure ratios reduce the exhaust temperature (considering a constant TIT) as the working fluid is expanded over a larger pressure range and more energy is extracted from the working fluid and converted to electricity in the GT generator.

Secondly, GT exhaust recuperation is more likely to be feasible at low FUs. Lower FUs lead to more unutilized fuel entering the GT oxidizer which increases the TIT and thus, the GT exhaust temperature. This effect is further enhanced by the cooling air requirement. Lower FUs require less cooling air in order to maintain the thermal gradients within the SOFC. This ultimately leads to less dilution in the GT oxidizer resulting in higher TITs and exhaust temperatures.

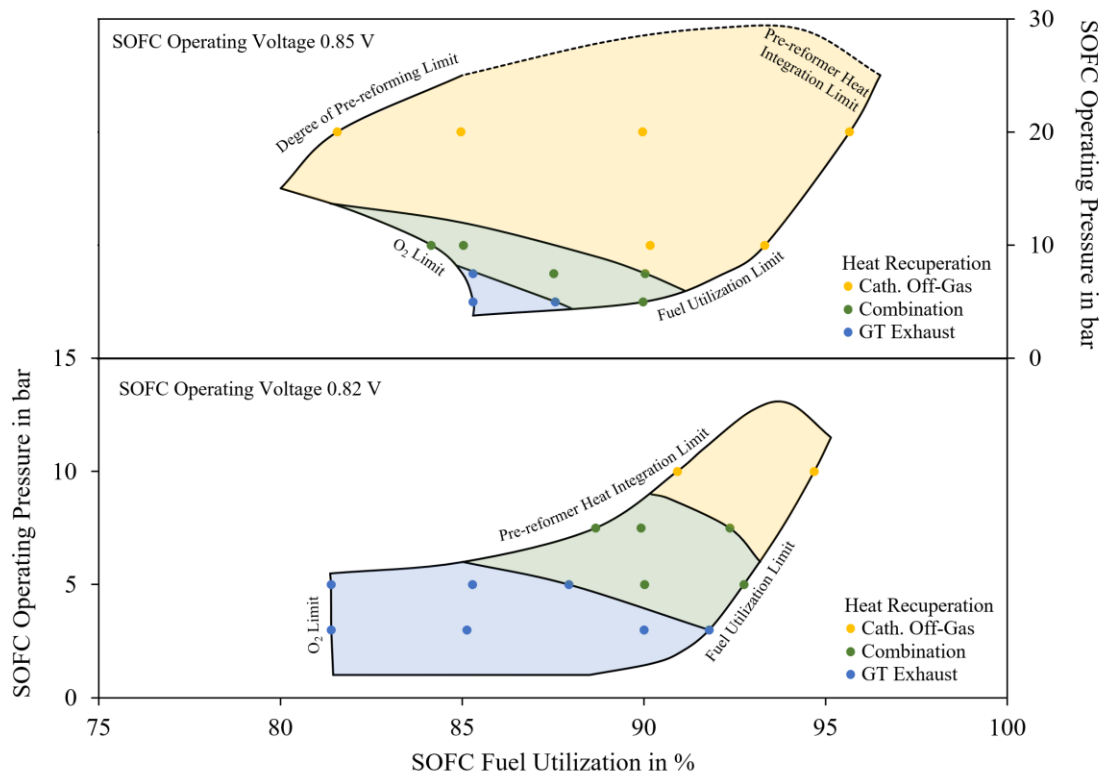


Figure 5-25: SOFC-GT Hybrid operating ranges for SOFC operating voltages of 0.82 V and 0.85 V. Thermal cell gradients are held constant at 15 K/cm and/or an overall PEN temperature difference of 150 K.

On the right side of the operating window shown in **Figure 5-25**, the plant operation is limited by the maximum achievable FU. This is the case for the 0.82 V case as well as for the 0.85 V case. In accordance with Equation (E 4-1) the FU limit increases with the pressure ratio as the pressure increases the chemical potential between anode and cathode. For 0.85 V the FU limit at a certain pressure is reached sooner as for the 0.82 V since a higher chemical potential difference is needed to obtain the higher voltage.

Two other factors have an impact on the FU limit. Anode off-gas is recirculated to avoid carbon deposition and adjusted based upon the gas composition. This typically leads to lower H_2 concentrations at SOFC inlet and outlet since the anode off-gas also contains CO , CO_2 and H_2O . Thus, when running the SOFC at high FUs, the increased recirculation leads to a further decrease of the chemical potential difference.

Secondly, the air flow is determined based upon the thermal cell management, with lower pressures leading to lower power densities, which reduces the amount of cooling air flow. This ultimately leads to a higher air utilization leading to lower O_2 concentrations and a lower chemical potential difference between anode and cathode. As a result of these effects, high SOFC operating voltages can be challenging for configurations operating at low pressure ratios and no feasible operating condition for the 0.85 V scenario at 3 bar could be identified for the configuration studied in this work. However, it may be pointed out that there are operating points with smaller thermal gradients as specified in **Table 5-2**, which is typically acceptable for plant operation.

While this region of lower thermal gradients can be entered anywhere by just increasing the air flow, the O_2 limit line represents the last operating points where the above-mentioned thermal cell limits can be reached. It can also be viewed as the line of maximum air utilization or minimum air flow which is set by a minimum O_2 concentration of 3% in the GT oxidizer exhaust gas in order to ensure complete fuel oxidation. In general, the O_2 limit moves to lower FUs as the operating pressure increases which is a result of the higher power density at higher pressure ratios but also the reforming kinetics play a role (CH_4 reforming is favored at low pressures). As mentioned earlier, left of the O_2 limit line possible operating points exist but since these operating points exhibit smaller thermal gradients they are excluded from this analysis as they are typically economically less attractive.

Critical for this type of plant configuration is the heat integration of the pre-reformer which is thermally coupled with the SOFC via radiation. If the SOFC operating pressure is high the power density increases and less stacks are needed. However, a certain number of stacks are needed to provide enough external surface area for heat transport from the stack to the pre-reformer via radiation. This limit is illustrated by the pre-reformer heat integration limit which is first reached for low FUs as the lower FUs are obtained via reduction of SOFC stacks and not by voltage variations. The downwards slope close to the FU limit is a result of the increased air flow needed at high FUs which at some point substantially decreases the average stack temperature and thus the radiation heat transfer.

For the higher operating voltage, another limiting effect, similar to the O_2 limit, is observed. This is not a hard operating limit, however, increasing the pressure shifts the equilibrium of the reforming reaction to the reactant side making it impossible to reach the 20% CH_4 reforming for the higher operating pressures (as shown in **Figure 5-25**).

Hybrid Thermodynamic Performance. The thermodynamic performance of the hybrid has been evaluated at steady state for multiple operating conditions based on the operating range. For the selected GT, the net power output of the plant varies between 7,474 – 13,992 kW for the 0.82 V cases and 4,922 – 15,027 kW for the 0.85 V cases, as seen in **Figure 5-26**. The SOFC power output is closely tied to its cooling air requirement which is supplied by the GT compressor. In general, the higher the power density (due to higher FU, higher pressure, lower operating voltage), the higher is the cooling air required and the lower is the plant's power output of the SOFC scaled to match the GT since the compressor size remains constant for all cases. Another significant factor impacting the thermal management is the reforming kinetics inside the SOFC. Internal reforming tends to occur at the inlet of the SOFC, which causes severe temperature gradients

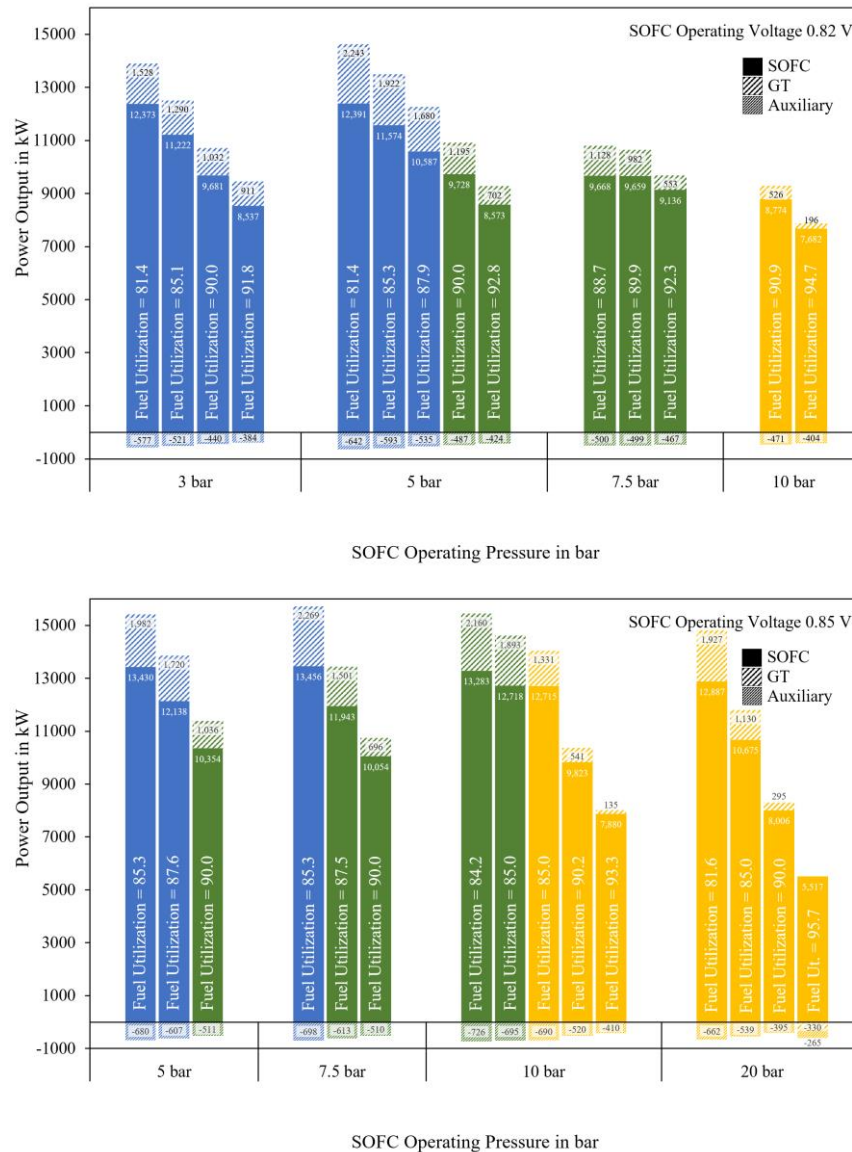


Figure 5-26: SOFC-GT hybrid plant gross power generation and consumption for an operating voltage of 0.82 V (top) and 0.85 V (bottom).

due to its endothermicity. At higher pressures, the reforming kinetics are slowed down as the gas composition entering the SOFC is closer to its equilibrium state. This can substantially reduce the cooling requirement and increase the overall plant power output as seen for the low FU cases at 20 bar of the 0.85 V scenarios.

Generally, higher FUs and higher SOFC cooling demands result in lower TITs reducing the GT efficiency and GT power output. Furthermore, higher cooling demand results in a lower fuel to air ratio decreasing the mass flow through the turbine section. Another factor to consider is the pressure ratio. If the TIT is very high, higher pressure ratios can be beneficial as more energy from the hot working fluid can be converted to electricity. However, if the TIT is not high enough the increased compressor power demand can be detrimental to the system efficiency. In the extreme case of the 95.7% FU at 0.85 V and 20 bar, the power generated by the GT expander is not enough to offset the power consumed by the GT compressor and the GT requires a motor instead of a generator. Auxiliaries are primarily related to the SOFC power conditioning equipment and fuel compression.

The plant efficiency has a moderate impact on the overall power output and in general higher operating voltages improve the system efficiency. The efficiencies for the various operating conditions are shown in **Figure 5-27**. As seen in **Figure 5-26**, the majority of the electricity is generated by the SOFC. Thus, increasing the operating voltage of the SOFC has a significant impact on the plant performance leading to overall higher plant efficiencies for the 0.85 V cases compared to the 0.82 V cases.

First the 0.82 V scenarios are discussed. For system configurations operating at 3 bar, the efficiency increases with increasing FU as the fuel is more efficiently used in the SOFC compared to the GT expander. All cases operating at 3 bar use only exhaust gas recuperation to pre-heat the air to SOFC inlet temperature (blue) since there is enough high quality exhaust heat available. The same trend is observed for the exhaust gas recuperation cases (blue) of 5 bar scenarios. Comparing the efficiencies of the 5 bar scenarios with the 3 bar scenarios at the same FU a significant efficiency increase is observed. This increase is associated with the GT (SOFC operates at constant voltage). If the GT exhaust is hot enough to completely pre-heat the air, higher pressure ratios are favorable from an efficiency perspective as more heat is converted to electricity. For the higher FU cases of the 5 bar scenarios (green), the temperature of the GT exhaust is insufficient to pre-heat the air. Here, a combination of high pressure ratios across the GT expander and high FUs lead to lower GT exhaust temperatures. Thus, the cathode off-gas needs to be utilized in order to increase the air temperature to the desired SOFC inlet temperature. Unfortunately, this further decreases the TIT. The power output decrease of the GT is so substantial that the higher power generation in the SOFC cannot compensate for this loss. This trend is also observed in the 7.5 bar scenarios. These observations are similar to efficiency trends observed in recuperated gas turbines without SOFC integration. The 10 bar scenarios utilize only the cathode off-gas to pre-heat the air (yellow) and are more comparable to simple cycle unrecuperated GTs.

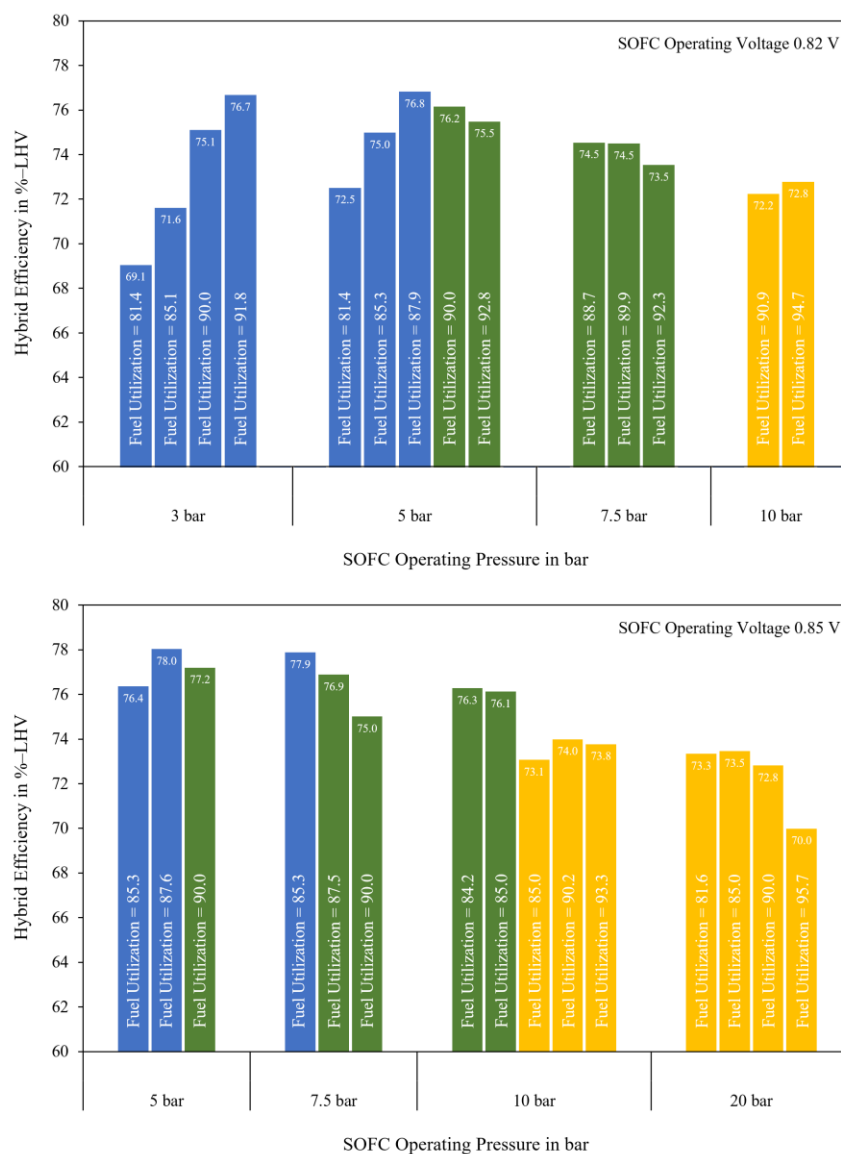


Figure 5-27: SOFC-GT hybrid plant efficiency for an operating voltage of 0.82 V (top) and 0.85 V (bottom).

Both 10 bar cases require the same amount of heat per kg of air from the cathode off-gas to preheat the air since this air preheating method does not utilize the GT exhaust. Under these conditions an increase in FU generates more power in the SOFC than the power reduction in the GT due to the increased SOFC cooling air needs leading to an overall higher efficiency.

Initially, the same trend is observed for the 0.85 V scenarios at 10 bar and 20 bar using only the cathode off-gas for air pre-heating (yellow); however, for the highest FU cases of these scenarios a change in this trend is observed and the efficiency starts decreasing. The 0.85 V scenarios in general require less cooling air compared to the 0.82 V cases, which helps to increase the TIT and boosts efficiency. However, the cooling demand increases more significantly with FU causing an inversion point of this trend where the

efficiency starts to fall. This efficiency decrease is associated with a drastic increase in cooling requirement, which substantially decreases the TIT. A special case is the 20 bar, 81.6% FU case which exhibits a surprisingly high efficiency compared to the other cases considering the general trends. This case additionally benefits from the slower reforming kinetics resulting in a lower cooling air demand and higher efficiency. All other scenarios at 0.85 V follow the trends of the lower voltage scenarios and the same explanations are valid.

Hybrid Economics. While key performance characteristics such as efficiency are important, an economic analysis is needed to gauge a technology's competitiveness on the free market. In the economic analysis tradeoffs between high efficiency, which reduces some operating costs, and capital cost are analyzed over the plant's lifetime.

First the specific plant costs and the specific cost for the two power-generating plant units (SOFC and GT) have been analyzed on a \$/kW basis, which can be seen in **Figure 5-28**. The specific plant cost is plotted on the top and the specific GT cost and SOFC cost are mirrored at the center line. In general, the specific GT cost increases with increasing FU throughout the different pressure scenarios and operating voltages. This is directly related to the low heating value of the anode off-gas and the increased cooling air demand, which reduce the firing temperature and power output of the GT. This trend is observed despite the fact that lower TITs lead to a lower GT equipment cost. The lowest specific GT costs are achieved for the cases with the highest power output (see **Figure 5-26**) and pressure effects have only a moderate impact on the specific GT cost.

Also, the specific SOFC cost increases with higher FUs as higher FUs required more cell area at a constant voltage. An exception are the two 85% FU cases at a pressure of 10 bar and a voltage of 0.85 V, which is due to the different heat recuperation strategies impacting the fuel pre-heating. Comparing different pressure scenarios at constant FU shows that for the 0.82 V scenarios, the specific SOFC cost decreases with increasing pressure. Increasing the pressure from initially 3 bar to 5 bar reduces the specific SOFC cost by over 21%. At higher pressures this effect is less pronounced. Higher SOFC operating pressures increase the current density; however, this effect gets smaller on a per pressure increment basis as the pressure resistant equipment, such as the SOFC housing, keeps getting more expensive as the pressure requirements increase. This will eventually lead to a cost optimum somewhere above 10 bar. For the higher operating voltage of 0.85 V, the same trend is observed; however, the cost optimum is expected at a higher operating pressure due to the generally higher specific SOFC cost when operating at a higher voltage.

SOFC-GT Hybrid Systems

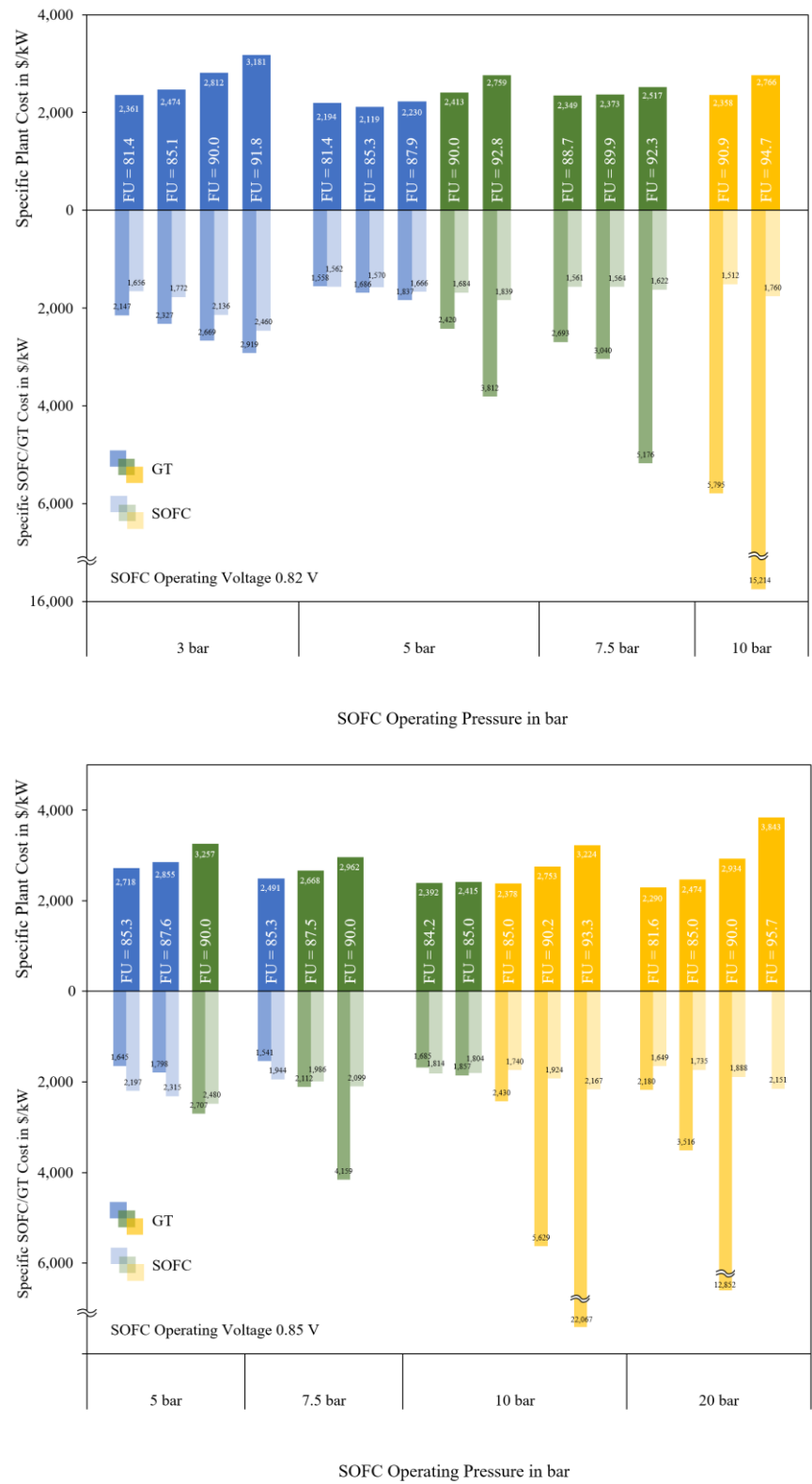


Figure 5-28: Specific plant cost, specific SOFC cost and specific GT cost for the 0.82 V scenarios (top) and 0.85 V scenarios (bottom).

The lowest specific plant costs are achieved for scenarios with low specific SOFC cost and low specific GT cost. However, other plant equipment also comes into the picture when considering the entire power plant. While most other plant equipment remains fairly unchanged, the cost of heat recuperation can change considerably between different cases. In general, the combined air pre-heat strategy using the GT exhaust and cathode off-gas to pre-heat the air is more expensive than the other two strategies of pre-heating the air in a single heat exchanger. Furthermore, heat recuperation from the exhaust gas of low FU cases results in a higher exhaust temperature increasing the cost of heat exchange equipment. As a result, the lowest specific plant cost is obtained for the 85.3% FU case at a pressure of 5 bar and 0.82 V with a value of 2,119 \$/kW. A breakdown of how the various plant components contribute to the overall plant cost for this case can be found in **Figure 5-29**.

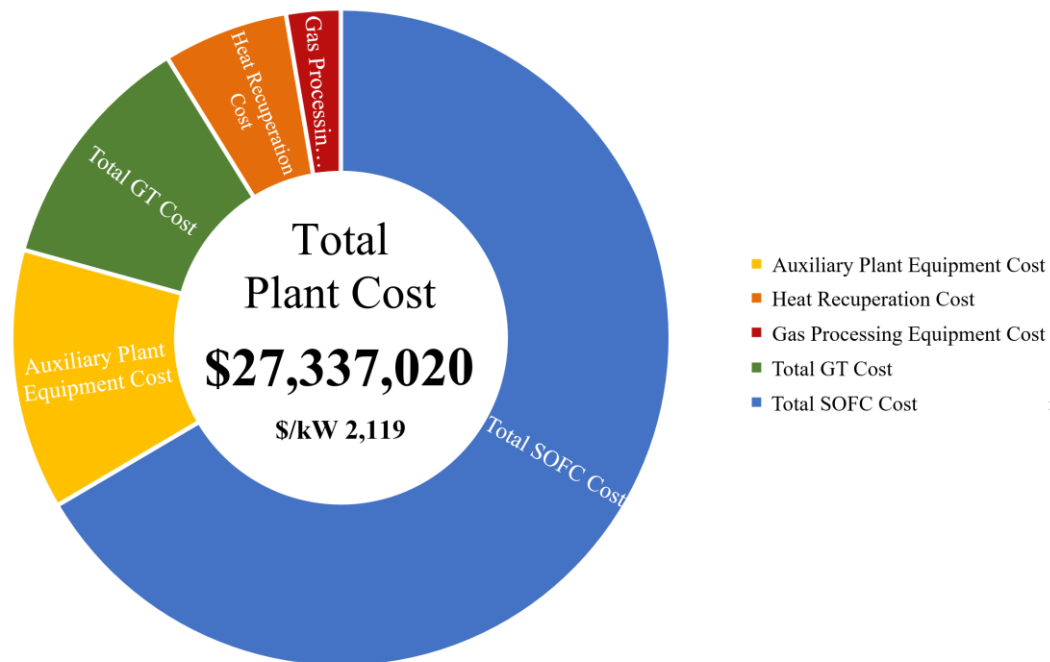


Figure 5-29: SOFC-GT Hybrid plant cost breakdown of the total plant cost (0.82 V, 85.3% FU, 5 bar).

The vast majority of the TPC, 66.5%, is associated with the SOFC island and includes stack cost, power conditioning equipment, housing and final assembly as well as installation cost. Auxiliary plant equipment such as electric plant accessories, instrumentation and controls, site improvements, and buildings and structures account for 12.8% of the TPC. The GT generates about 14.2% of the plant's gross power while

accounting for 11.9% of the TPC. The heat recuperation cost represents 6.2% of the TPC and gas processing contributes 2.7%.

The specific plant cost is a strong indicator for the COE and the same trends are observed. The lowest electricity cost is obtained for the 85.3% FU case at a pressure of 5 bar and 0.82 V with a COE of 72.60 \$/MWh. A complete summary of COEs for all study cases is provided in **Figure 5-30**.

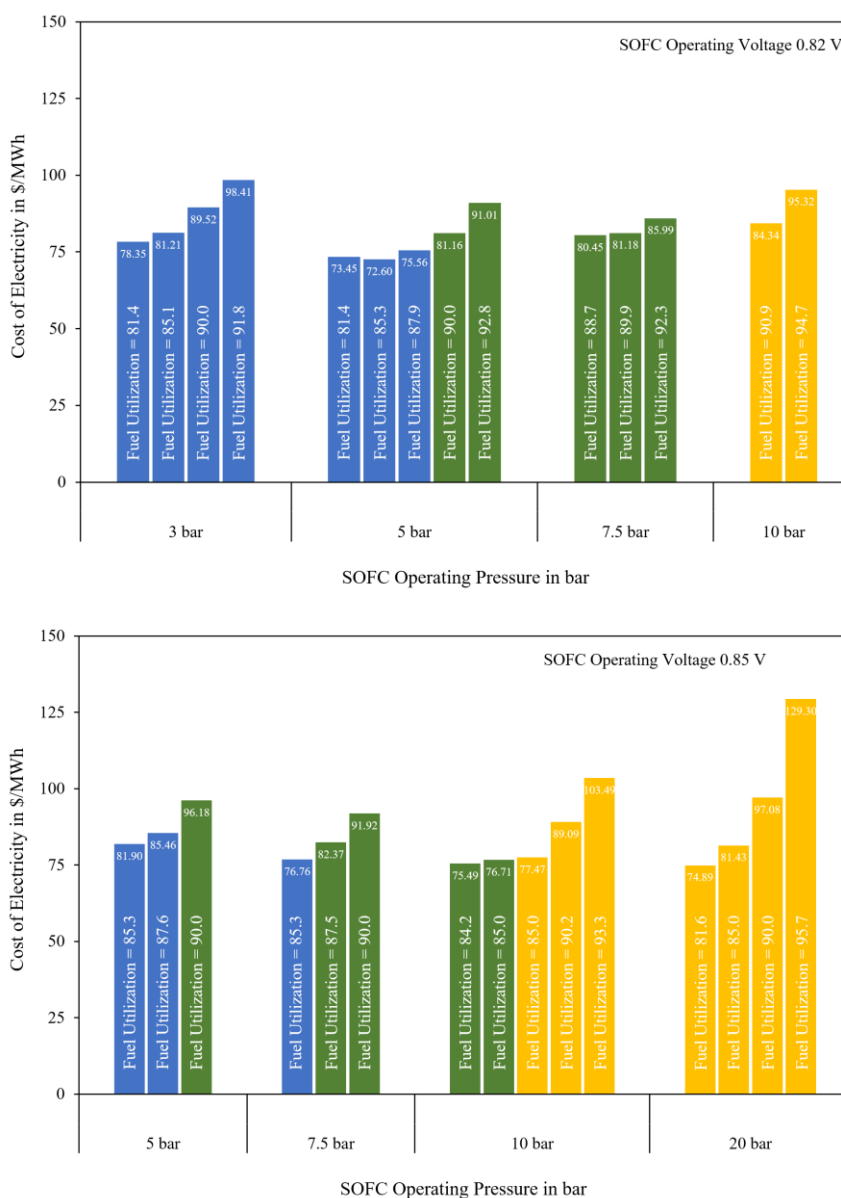


Figure 5-30: SOFC-GT hybrid plant power cost-of-electricity for an operating voltage of 0.82 V (left) and 0.85 V (right).

A detailed breakdown of the various COE driving factors is shown in **Figure 5-31**. Variable operating costs represent the largest contribution to the COE with a share of 35.3%. Most of the variable operating costs and single largest COE cost component with 26.4%, is the feed cost despite the high efficiency of the

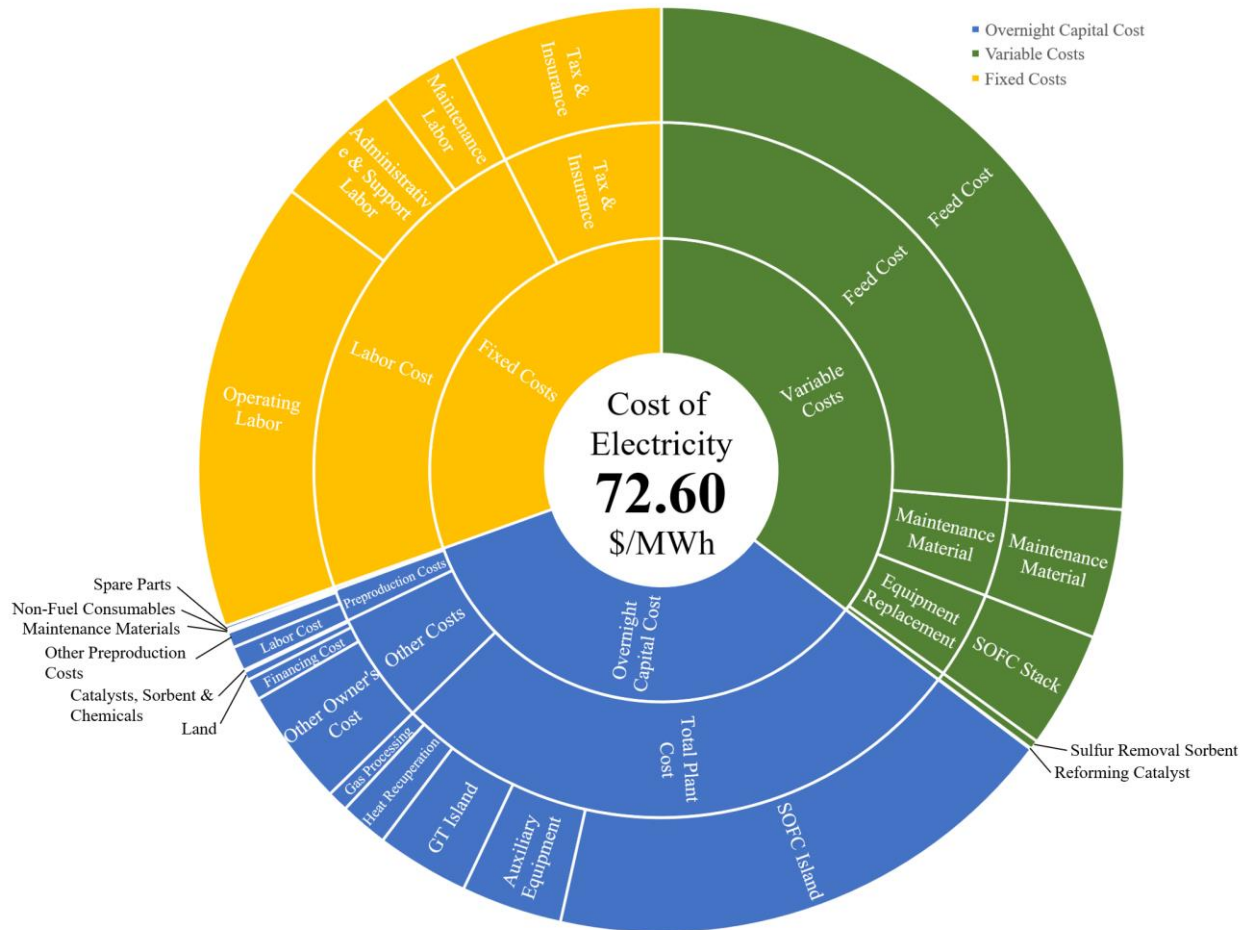


Figure 5-31: SOFC-GT Hybrid plant cost breakdown of the cost-of-electricity (0.82 V, 85.3% FU, 5 bar).

hybrid. Maintenance material costs account for 4.5% and the annual levelized SOFC stack replacement for 4.1% of the overall COE.

The overnight capital cost is the second largest category among the COE cost driving factors at 34.2%. Most of the overnight capital cost is the TPC, whereby, the SOFC island cost contributes 18.2% to the overall COE. Other overnight capital costs include preliminary feasibility studies, economic development, legal fees, permitting, financing cost, land cost, etc. as well as pre-production costs and initial fills. Fixed operating costs associated with tax, insurance and labor account for 30.5% of the COE.

Summary

The thermodynamic performance of a SOFC-GT hybrid system as shown in **Figure 5-24** has been investigated at steady state operating conditions considering overall cell temperature differences and local thermal gradients inside the SOFC. The results show that operating the system under higher fuel utilization and high pressure ratios results in a substantial increase in air flow to the SOFC in order to maintain the

same temperature differences and thermal gradients. In general, higher SOFC operating voltages reduce the cooling requirement and increase the efficiency. High air utilization factors are desirable to increase the net power output of the plant.

The simulation results served as input for the thermo-economic evaluation of the SOFC-GT hybrid. The specific cost of the SOFC as well as the GT are reduced when operating the hybrid under low FU. While the GT specific cost is predominately affected by its power output (pressure ratio having only a moderate influence on cost), the pressure dependency of the specific cost of the SOFC exhibits an optimum. On the one hand, the physical size of the SOFC is reduced at higher operating pressures as the current density increases (while maintaining the same FU and voltage). On the other hand, the SOFC housing and pressure vessel costs increase with pressure which become dominant at higher pressures. As the operating voltage is increased, this optimum is shifted to higher pressures due to the generally higher specific SOFC cost.

Besides the fuel utilization and SOFC operating pressure, the operating voltage has a significant impact on the SOFC specific cost, which increases with operating voltage. As a result, specific plant costs are reduced by operating the plant at relatively low FUs, relatively low voltages and moderate compression ratios. The specific plant cost is a strong predictor of the COE and the lowest COE is obtained at an operating voltage of 0.82 V, an operating pressure of 5 bar and a FU of 85.3%.

5.1.3 Thermo-Economic Analyses of Integrating an Off-the-shelf Gas Turbine into a Solid Oxide Fuel Cell-Gas Turbine Hybrid

Introduction

With current operating practices used in the power sector, it is not possible to maintain the sensitive ecological balance of the earth's ecosystems. The steady increase of greenhouse gasses in the atmosphere related to the use of fossil fuels together with the emission of criteria pollutants lead to environmental degradation and changes in rainfall and weather patterns. To diminish the environmental footprint of the power sector, highly efficient and environmentally sensitive energy conversion systems are of major importance. When utilizing carbon fuels, the thermal efficiency directly translates to lower resource depletion and lower carbon emissions. Solid oxide fuel cell (SOFC)-gas turbine (GT) hybrids have demonstrated reaching efficiencies of greater than 75%-LHV [96], [102], [114]. While SOFC-GT hybrid systems can operate on natural gas during a transition period, they can also be operated on carbon free H_2 generated from renewable resources [99]. The ability of reaching very high efficiencies and being able to operate on natural gas as well as H_2 uniquely qualifies SOFC-GT hybrids as an important key technology for the energy infrastructure of the future.

In a SOFC-GT hybrid, most of the fuel is converted to electricity via highly efficient electrochemical reactions. This chemistry precludes the formation of pollutants, and consequently, SOFC-GT hybrids emit virtually no criteria pollutants.

Over the years substantial efforts have been undertaken to study SOFC-GT hybrid systems in detail. Most work focuses on operational aspects of the SOFC such as carbon deposition or thermal gradients. Carbon deposition inside the SOFC leads to catalyst deactivation and has been analyzed in the context of varying steam to carbon ratios and its impact on the plant performance [155] as well as in the context of gas stability [123]. In order to study thermal gradients within the ceramic electrolyte membrane, 1D models have been utilized which take heat and mass transfer phenomena inside the cell into account [120], [123], [335], [336].

Aspects of GT integration into a SOFC-GT hybrid have been studied to a lesser extent. Historically, commercially offered GTs have been engineered for energy dense fuels such as natural gas. The utilization of lower heating value fuels (e.g., the anode off-gas from a SOFC) can result in a flow mismatch between the GT compressor and GT expander. This issue is eliminated with the use of custom-engineered GTs, which is useful for the evaluation of the full thermodynamic potential of a technology. However, GT vendors are not expected to offer GTs specifically designed for SOFC-GT hybrid applications. Thus, engineers will need to rely on current GT offerings for the design of SOFC-GT hybrid systems.

In this context Lundberg et al. [91] investigated the integration of a SOFC into a Mercury 50 gas turbine without a bottoming cycle and with bottoming steam and ammonia Rankine cycles. The gas turbine was simulated at its design point condition with constant geometry. Only minor design modifications were necessary to accommodate the fuel cell. The inlet stators were adjusted, the air inlets to the combustion chamber were resized due to the higher inlet temperature and the combustor main case has been redesigned using high temperature material. Park et al. [92] integrated a SOFC into a fixed geometry micro gas turbine and found a significant reduction of the gas turbine power generation compared to the design point. The gas turbine power outlet can be increased by increasing the operating temperature of the SOFC. However, this requires the gas turbine to operate close to the compressor surge line. Costamagna et al. [93] conducted simulations on the integration of a SOFC into a micro turbine using compressor performance parameters and found that even for very small systems of 300 kW, efficiencies of greater than 61%-LHV are possible. Furthermore, variable speed operation was shown to be superior to constant speed operation under part load conditions. Barelli et al. [94] showed that, under part load, the efficiency can be maintained at a very high level when using variable speed control for the GT. A similar system was studied by Campanari [95] under part load conditions who came to the same conclusion and additionally showed that variable speed operation of the GT is advantageous (with respect to efficiency) under varying ambient conditions. The economic impact of these varying operating conditions however remains unexplored.

In summary, various SOFC-GT hybrid systems have been studied in literature; however, key aspects of the integration of commercial GTs into a SOFC-GT hybrid are not fully addressed. While variable speed operation has been found to be favorable with respect to efficiency under part load conditions, it is not clear how variable speed operation impacts full load operation, surge margin, and economics. This study aims to close this gap in the literature by comparing three GT operating modes, I) constant speed, II) variable speed and III) variable inlet guide vanes (VIGV) to a reference case with a custom-designed GT under various fuel utilization factors considering surge margin, efficiency and cost of electricity (COE).

Methodology

Three GT operating modes I) constant speed, II) variable speed and III) VIGV are evaluated under full load operating conditions for various fuel utilization factors at steady state. The SOFC underlies thermal constraints and the GT operates on a compressor map and turbine map. The thermodynamic and kinetic results obtained from the simulation serve as basis for equipment sizing and the economic evaluation which includes first-year levelized COE.

Design Basis

The SOFC-GT hybrid systems studied in this work use natural gas (see **Table 4-4**) supplied at a pressure of 4.1 bar. Natural gas can be internally reformed inside the SOFC producing H_2 and CO. Carbon monoxide typically undergoes further conversion via the water gas shift reaction producing additional hydrogen. While CO can be oxidized electrochemically inside the SOFC, it is almost exclusively H_2 that reacts inside the SOFC due to kinetic limitations. The reforming reaction is of endothermic nature and can provide cooling to the SOFC.

Solid Oxide Fuel Cell Hybrid Configuration. In a SOFC-GT hybrid, a gas turbine is used to pressurize the SOFC while serving as the “bottoming cycle.” Higher operating pressures allow the SOFC to reach higher fuel utilizations (FU), higher operating voltage and higher current density. Moreover, high quality heat generated in the fuel cell itself can be exploited together with heat generated from unutilized fuel in the expander section of the GT. This synergy between the SOFC and GT allow the SOFC-GT hybrid to reach extremely high efficiencies. In this work, the same configuration as above has been chosen to allow direct comparisons between the optimized case using a custom-designed GT and results from this work using an off-the-shelf GT. In this design, an eductor recycles a portion of the anode off-gas back to the pre-reformer to avoid carbon deposition. The amount of recycle is determined based upon gas phase stability considerations [252]. Compared to the lowest COE case from above where the air is completely preheated by the GT exhaust, the plant configuration in this study is slightly modified because of differences between

the custom engineered GT and the off-the-shelf-GT. The higher pressure ratio of the off-the-shelf GT requires additional air heating using the cathode off-gas. A flowsheet of the plant configuration is provided in **Figure 5-32**. Ambient conditions correspond to ISO conditions of 1.013 bar and 15 °C. Air composition is provided in **Table 4-2**.

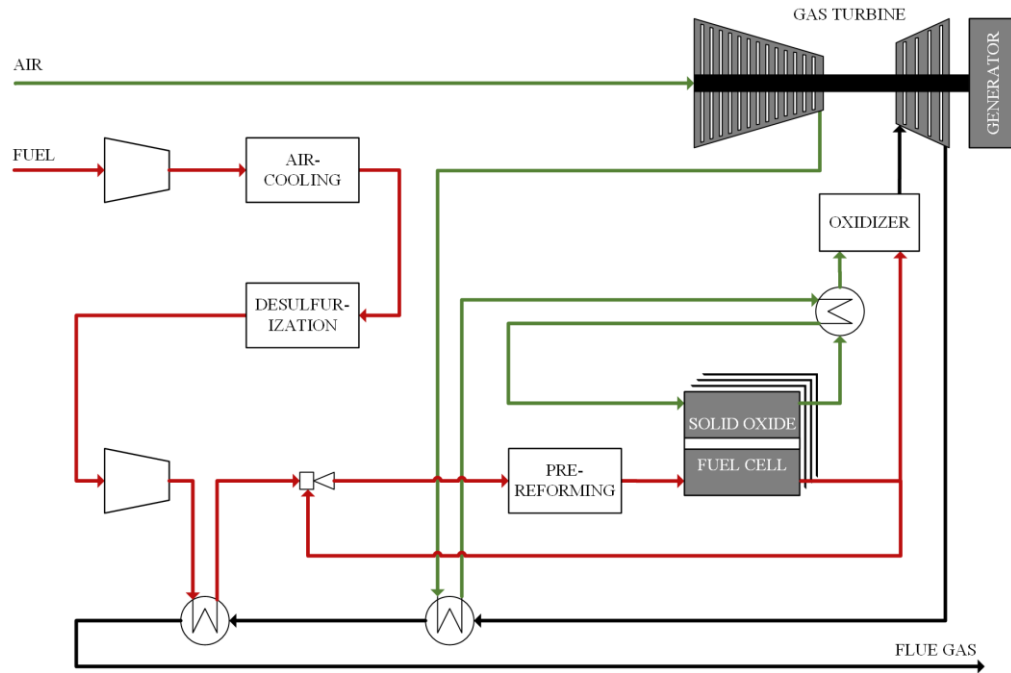


Figure 5-32: Solid oxide fuel cell - gas turbine hybrid configuration with off-the-shelf gas turbine.

Power Block Modeling. An eductor is used to recycle a portion of the anode off-gas back to the pre-reformer. Eductors are static devices that use a high-pressure stream to compress a lower pressure stream. The use of an eductor is advantageous in SOFC applications as the hot recycle stream does not have to be cooled prior to the eductor opposed to when using a compressor. Furthermore, equipment cost and maintenance cost for eductors are very low. In this work a 2-dimensional eductor model developed by Zhu et al. [193] has been employed which accounts for the velocity profile in the suction chamber. Traditional 1-dimensional thermodynamic models tend to overpredict the eductor performance as they do not consider frictional losses [193].

Before the methane rich fuel gas is introduced into the SOFC, 20% of the methane is pre-reformed in an external pre-reformer. In this work, the pre-reformer is integrated into the SOFC furnace in a fashion that allows radiation heat transfer between the SOFC stack and the pre-reformer. The thermal integration in this design is similar to the design described in [254]. Under these conditions, the reforming reactions can be either limited by the quantity of catalyst present or the heat transfer from the SOFC to the pre-

reformer. The purpose of the pre-reformer is to pre-reform higher hydrocarbons which are present in low concentrations and pose a risk of carbon deposition. Only limited amounts of CH_4 conversion are desirable in the pre-reformer to lower thermal gradients inside the SOFC in order to take advantage of internal reforming.

The SOFC is modeled as a planar, anode supported cell with anode and cathode streams flowing in co-flow. The model utilizes the Nernst equation for the reaction between hydrogen and oxygen. Although carbon monoxide can react electrochemically in SOFCs, the kinetics are sufficiently slow compared to the dominant hydrogen reaction that it can be neglected. Therefore, CO is rather shifted to CO_2 producing hydrogen than electrochemically oxidized. The reforming reaction rate is based upon kinetic expressions, whereas the water gas shift reaction is considered to quickly reach equilibrium. Three major losses are accounted for, in order to obtain the real cell operating voltage: activation losses, ohmic losses and concentration losses. The cell geometrical parameters used in this study are summarized in **Table 4-13**. The SOFC operating conditions for this study are summarized in **Table 5-3**

Table 5-3: Summary of Internal Reforming Solid Oxide Fuel Cell and Gas Turbine Parameters for off-the-shelf GT

SOFC Operational Parameters	Value	Unit
Operating Pressure	~7	bar
Operating Voltage	0.82	V
Inlet Temperature Anode	973.15	K
Inlet Temperature Cathode	973.15	K
Max. local PEN Temp. Gradient	15	K/cm
Overall PEN Temp. Difference	150	K
GT Operational Parameters (Design Point)	Value	Unit
Compressor Polytropic Efficiency	84.35	%
Expander Polytropic Efficiency	84.36	%
Compressor Mechanical Efficiency	98.03	%
Expander Mechanical Efficiency	98.03	%
Turbine Inlet Temperature	943.3	°C
Combustor Pressure Drop	3	%
Combustor Heat Loss	1	%-LHV

The gas turbine selected in this work is a Dresser Rand KG2-3G EF which is rated for 1.83 MW under ISO conditions using CH_4 as fuel [247]. With its design point compression ratio of 7:1 and mass flow rate of 9.48 kg/s [247], it is suitable for the integration into a SOFC-GT hybrid systems of approximately 10MW, similar to the previous study. The EF (externally fired) version of this GT will allow for an easier GT integration as compared to GTs where the combustor is integrated into the GT casing. Furthermore, Dresser Rand offers a Power Oxidizer with this model for the utilization of low heating value gases while achieving ultra-low emissions. This GT uses radial compressor and expander technology in order to accommodate for

this fuel flexible operation. Single stage radial compressor [248] and expander [249] maps have been used to develop generalized maps as shown in **Figure 4-4** and **Figure 4-5**, which are used to determine the off-design operation parameters. Thermodynamically, the GT is modeled as compressor, combustor and expander whereby the design point performance input parameters such as polytropic efficiency, mechanical losses (incl. generator losses), heat loss and pressure drop are obtained from the on design point calibration case using vendor data [247]. In order to not exceed the maximum permissible blade metal temperature in the expander section of the GT, Equation (E 4-20) for the turbine inlet temperature (TIT) is employed.

Results

Operating Conditions. In the previous study, a SOFC operating pressure of 5 bar and a relatively low FU of 85% result in the lowest COE (0.82V); although, higher efficiencies were obtained at higher fuel utilization factors and higher operating voltages. This finding is based upon the use of a custom-engineered GT specifically design for the operating condition encountered in the respective SOFC-GT hybrid. While this is an appropriate assumption when evaluating the ultimate potential of SOFC-GT hybrid systems, GT vendors are not expected to offer such GTs in the near future and currently available GTs will, as a result, need to be utilized in near-future SOFC-GT hybrid applications.

A Dresser Rand KG2-3G EF GT has been selected in this work due to its fuel flexibility and rating at ISO conditions, which closely match the desired GT size and operating conditions. Nevertheless, the differences of the KG2-3G EF to the ideal GT from the study above require small modifications in the heat recuperation strategy. While it has been shown that using only GT heat recuperation to pre-heat the SOFC air to 700°C is favorable from an efficiency standpoint, the slightly higher pressure ratio of the KG2-3G EF makes it necessary to use additional heat from the cathode off-gas for air pre-heating as shown in the flowsheet in **Figure 5-32**.

When integrating a commercial GT designed for natural gas into a SOFC-GT hybrid, the operating point moves away from its natural gas design point. In a SOFC-GT hybrid, the fuel for the Brayton cycle is the unutilized fuel from the SOFC anode outlet which has a characteristically low heating value ranging from 1.09 MJ/kg to 1.74 MJ/kg in this work, depending on the fuel utilization (high fuel utilization leads to a low heating value gas stream leaving the SOFC and vice versa). In this work, the FU is varied between a lower limit set by the heat integration of the pre-reformer, around 88%, and an upper end of around 92% where the electrochemical reaction starts to severely slow down due to the limited chemical potential difference. The anode off-gas is rich in H₂O and CO₂ leading to an increase in the H₂O and CO₂ concentrations in the gas stream entering the GT expander section. Due to the higher heat capacity of CO₂ and H₂O compared to diatomic gases, an adjustment of the TIT is required to maintain the same blade metal temperature as in the natural gas calibration case.

For constant speed operation, the maximum permissible firing temperatures (TIT_{MAX}) are: 909 °C (FU 88.2%), 908 °C (FU 90.0%) and 915 °C (FU 92.0%), a substantial reduction when compared to the NG design point with a TIT of 943 °C. The changes in TIT_{MAX} are observed due to variations in the fuel to air ratio which is dictated by the SOFC. The higher the FU, the higher the cooling air requirement which helps to decrease the concentration of H_2O and CO_2 in the hot combustor outlet gas. A second effect impacting the air flow requirement is the SOFC operating pressure. Higher operating pressures increase the current density of the SOFC and thermal gradients which results in a higher cathode air flow requirement in order to maintain the same thermal cell gradients. At constant speed operation, the 88.2% FU scenario exhibits the highest operating pressure with a GT compressor discharge pressure of 7.77 bar. At a FU of 88.2% sufficient quantities of fuel are left in the anode off-gas such that the maximum permissible TIT of 909 °C can be reached. In the 90.0% scenario the TIT decreases 893 °C which is 15 °C below the maximum permissible TIT. A further increase in FU exacerbates this effect and the 92.0% FU scenario reaches a TIT of 809 °C, 106 °C lower than the maximum permissible TIT. A lower TIT and fuel to air ratio reduce the compressor discharge pressure resulting in a compressor discharge pressures of 7.72 bar in the 90.0% FU scenario and 7.44 bar in the 92.0% FU scenario which is still above the design point pressure ratio of 7.0 bar. The operating points are shown in **Figure 5-33**.

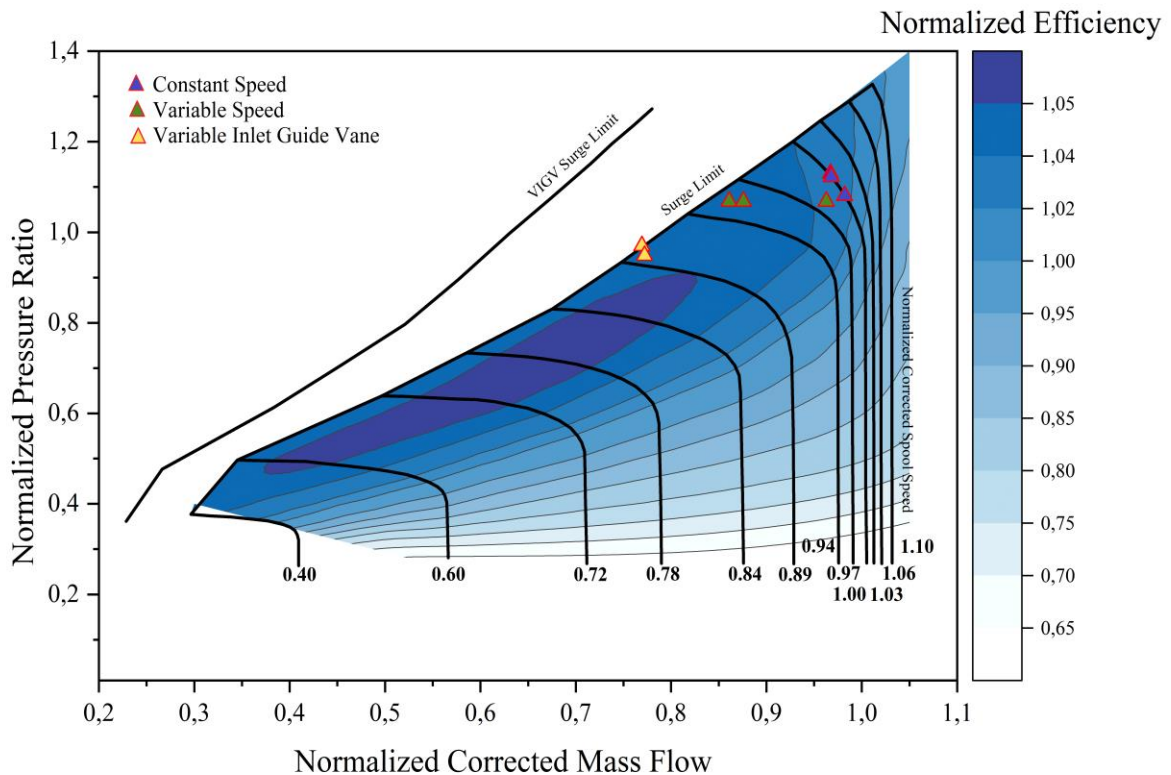


Figure 5-33: SOFC-GT Hybrid operating points at off-design conditions.

The surge margin (definition used in this paper refers to the surge margin at constant corrected mass flow) at the design point is 31.0%, which reduces to 18.7% for the 92.0% FU scenario and even 11.7% for the 88.2% FU scenario as shown in **Figure 5-34**.

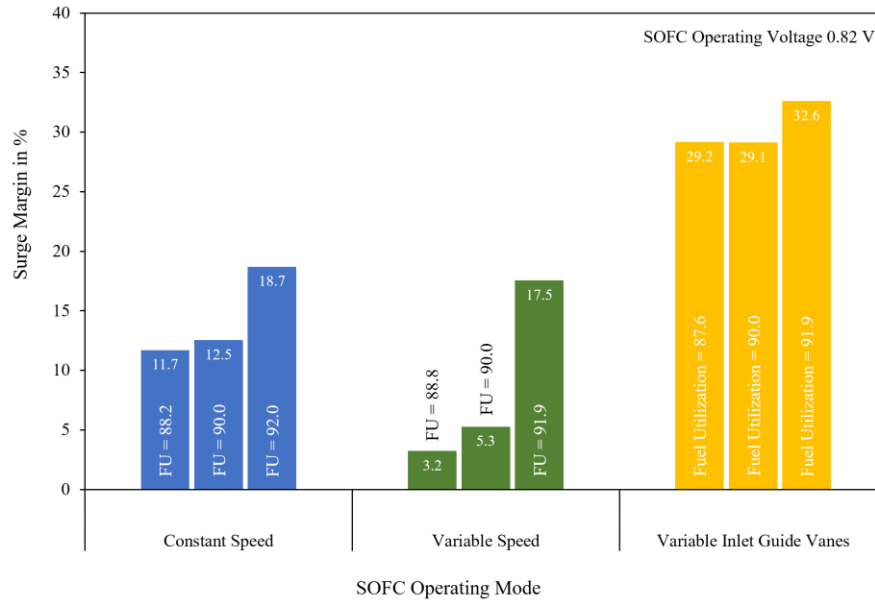


Figure 5-34: Surge margins of SOFC-GT hybrid plant at respective off-design operating point.

For variable speed control the degree of freedom of the system increases as the operating speed is not fixed. Thus, it is desirable to choose an operating condition avoiding concerns related to the surge margin. However, the operating range of the GT under variable speed is limited by the compressor and expander mismatch as well as the TIT. A compressor discharge pressure of 7.35 bar has been chosen as it is the lowest feasible discharge pressure to accommodate the three FU scenarios. At constant pressure, pressure effects on the cooling requirement of the SOFC are eliminated. Thus, the maximum permissible TIT_{MAX} is only related to the SOFC's FU. The lowest FU scenario of 88.8% FU needs the least cooling air flow in the SOFC and thus results in the lowest maximum permissible TIT_{MAX} of 908 °C. The 90.0% FU scenario arrives at a TIT_{MAX} of 909 °C and the 91.9% scenario allows a TIT_{MAX} of up to 915 °C. While the TIT_{MAX} is achievable in the 88.8% FU scenarios, the higher FU scenarios cannot reach their respective TIT_{MAX} and operate at 889 °C (90.0% FU) and 803 °C (91.9% FU). As the FU decreases the operating point moves away from the design point horizontally towards to surge line. From **Figure 5-34** it becomes apparent how drastically the surge margin is reduced. For the 88.8% FU scenario surge margin reduces to 3.2%. The 91.9% FU scenario has a substantially higher surge margin of 17.5%, however, this is still lower than the constant speed operation at similar FU.

Under VIGV operation the compressor map changes which leads to a left shift of the surge limit. The GT still operates at constant speed but also the constant speed lines have shifted. This ultimately leads to a

reduction in air mass flow and a lower compressor discharge pressure. Under a VIGV closing angle of 50°, the 87.6%, 90.0% and 91.9% FU scenarios reach a compressor discharge pressure of 6.67 bar, 6.67 bar and 6.53 bar. The lower operating pressure allows the 87.6% and 90.0% FU scenarios to operate at their respective TIT_{MAX} of 907 °C and 910 °C. Just the higher fuel utilization scenario of 91.9% FU operates at a TIT of 858 °C despite the fact that temperatures up to 917 °C are permissible. Closing the inlet guide vanes has a positive impact on the surge margin and the design point surge margin of 31.0% can be almost maintained in the lower FU scenarios. The 91.9% FU scenario even exceeds the design point surge margin, as seen in **Figure 5-34**.

Hybrid Thermodynamic Performance. The GT power output ranges from 1.41 MW to 0.49 MW which is substantially lower than the GT's ISO rating of 1.83 MW. The GT power output is influenced by the GT efficiency and the mass flow rates. The GT efficiency then again is affected by the compressor and expander efficiencies (off-design operation) as well as the TIT. As previously discussed, the TIT in this SOFC-GT hybrid application decreases compared to the ISO rating, due to the higher concentration of H₂O and CO₂ in the combustor outlet gas stream as well as challenges with the heat integration. The mass flow rates on the other hand are reduced at higher pressure ratios, lower spool speeds and partially closed inlet guide vanes. Lower mass flows ultimately lead to less working fluid and less power generation.

The constant speed scenarios have the highest GT power output, as shown in **Figure 5-35**. In **Figure 5-33**, it can be seen that the constant speed operation results in the largest compressor air flow. When comparing the different FU scenarios within the constant speed scenario, it may be noticed that the GT power output decreases with increasing FU. In general, lower FUs lead to higher TITs resulting in higher GT efficiencies and higher power outputs. The effects of changes in pressure ratio and fuel to air ratio, which increases the mass flow through the GT expander, are rather small and do not significantly impact the overall GT power output in the constant speed scenarios. Changes in compressor and turbine efficiencies for the constant speed and VIGV scenarios are relatively small, too.

In the variable speed scenarios, pressure effects have been completely eliminated. The GT efficiency, impacted by the TIT and component efficiencies, and mass flow rate are the dominating factors determining the GT power output. Although a substantially higher mass flow in the variable speed, 91.9% FU scenario is obtained and the compressor operates in a region of higher efficiency, it cannot fully compensate for the power loss associated with the decrease in TIT. At variable speed operation, lower FUs lead to lower mass flow rates as less air is needed to maintain a constant pressure level.

Closing the VIGVs has a similar effect on the compressor as reducing the speed. By controlling the flow cross section, air flow into the compressor is restricted causing lower air mass flow rates. The degree

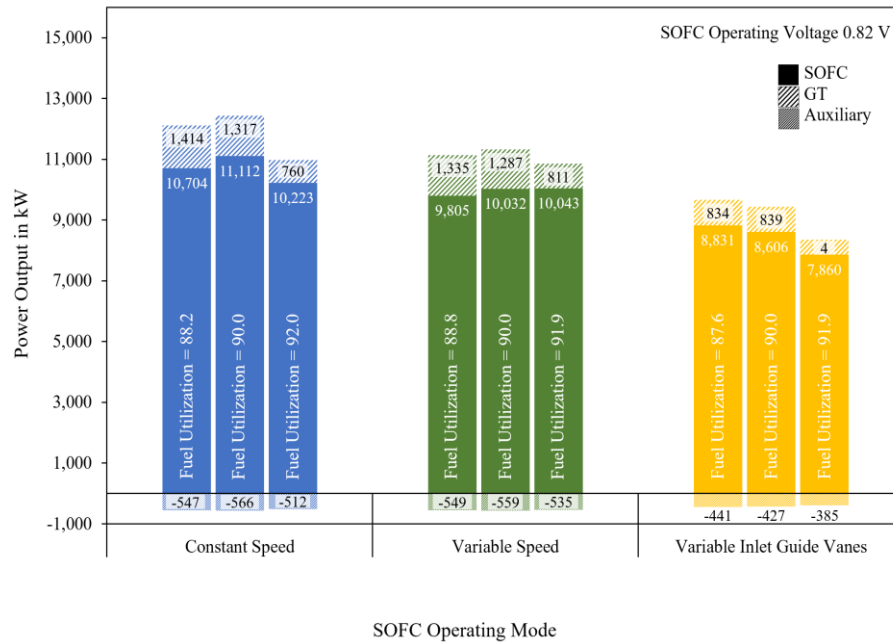


Figure 5-35: Power generation and auxiliary load of SOFC-GT hybrid plants.

to which variable speed control or VIGVs are adapted can be freely specified in real world applications, in this work rather extreme cases are studied to identify the effectiveness of these measures. Under these circumstances, closing the VIGVs by 50° results in a flow reduction of more than 23%, leading to the lowest GT power outputs among the studied scenarios. The GT power output trends within the VIGV operating mode follows the FU as discussed before.

The SOFC power output is strongly coupled to the FU and fuel to air ratio, since the operating voltage is held constant in all cases. The higher the FU, the more fuel is electrochemically converted in the SOFC increasing the power output. However, the GT air flow rate and the fuel to air ratio determine how many SOFCs can be used when paired with this specific GT. In a SOFC, the cooling air needed to maintain the same thermal gradients is dependent on the operating condition. In general, the cooling air requirement is increased when higher operating pressures and/or a higher FU is used. In fact, the cooling requirement increases drastically when operating the SOFC beyond 90% FU. Since the air flow rate is set by the GT, the fuel flow and respectively the size of the SOFC are adjusted so that the air provided by the GT is sufficient for the amount of SOFCs installed. Thus, while a higher FU leads to more fuel being converted to electricity, it also leads to a down scaling of the SOFC due to the increased cooling requirement. Additionally, the GT air flow rate determines how many SOFC can be accommodated. The balancing of these effects leads to the results shown in **Figure 5-35**.

The efficiency of the hybrid plant is mostly influenced by the FU and GT operating point. In a SOFC the voltage can be interpreted as a measure of efficiency; however, since the voltage in this study is held

constant, only the SOFC's FU has an impact on the plant efficiency. An increase in the FU shifts power generation from the GT to the SOFC. SOFCs typically operate at a higher efficiency compared to GTs, leading to an overall increase in plant efficiency. Yet, if at the same time the GT operating point moves into a region of extremely low efficiency, the gains of shifting electricity generation from the GT to the SOFC might be diminished. This behavior can be seen at high FUs for all operating modes and is illustrated in **Figure 5-36**.

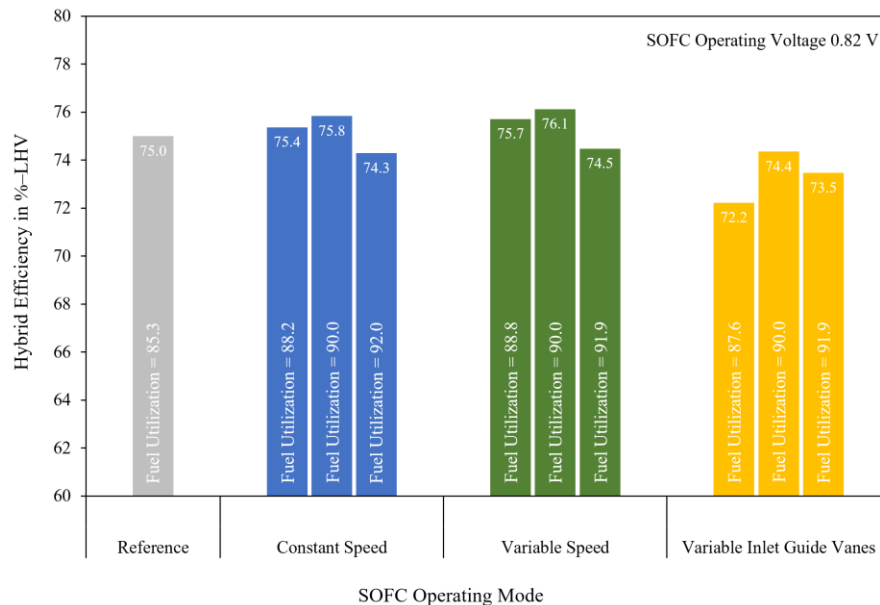


Figure 5-36: Efficiency of SOFC-GT hybrid plants with off-the-shelf GT and reference case with custom-engineered GT.

When comparing the constant speed, variable speed and VIGV scenarios at a constant FU, differences between the different operating modes are mostly related to the GT. The impact of pressure ratio is small in that case and the TITs at constant FUs are similar across the different operating modes. Also, the effect of higher fuel to air ratio, which increases the mass flow through the GT expander but at the same time increases upstream fuel compression work, has only modest impact on the plant efficiency when comparing constant FUs.

Most of the efficiency gains in this constant FU comparison originate from the GT compressor and GT expander efficiency. While the GT operating points are relatively close within the constant speed, variable speed and VIGV scenarios, the operating points shift considerably between the aforementioned scenarios, as seen in **Figure 5-33**. Such changes in GT component efficiencies can have a significant impact as GT compressor power consumption and GT expander power output are multiple times higher than the net power output of the GT. The highest efficiency is reached in the variable speed, 90.0% FU scenario where the operating point of the GT is located in a region of high efficiency, as seen in **Figure 5-33**. This high efficiency is reached despite the fact that variable speed control requires frequency adjustment which is

associated with additional conversion losses in order to export the electricity to the grid. When comparing the efficiencies to the reference case with custom-designed GT, the reference case does not have a higher efficiency than the variable speed, 90.0% FU scenario. The reason for this is that the reference case has been selected based upon the lowest COE and not the highest efficiency. In the previous study higher efficiency cases have shown to be feasible, however, these ultra-high efficiency cases were less economical than the selected reference case.

Hybrid Economics. In order to gauge a technology's competitiveness on the free market economic performance characteristics are needed to analyze tradeoffs between high efficiency, which reduces some operating costs, and capital cost over the plant's lifetime.

The total plant cost (TPC) is established based on the individual plant components and categorized into: SOFC cost, GT cost, auxiliary plant equipment cost, heat recuperation cost and gas processing cost. A characteristic breakdown of the TPC is shown in **Figure 5-37** using the constant speed, 88.2% FU scenario as an example.

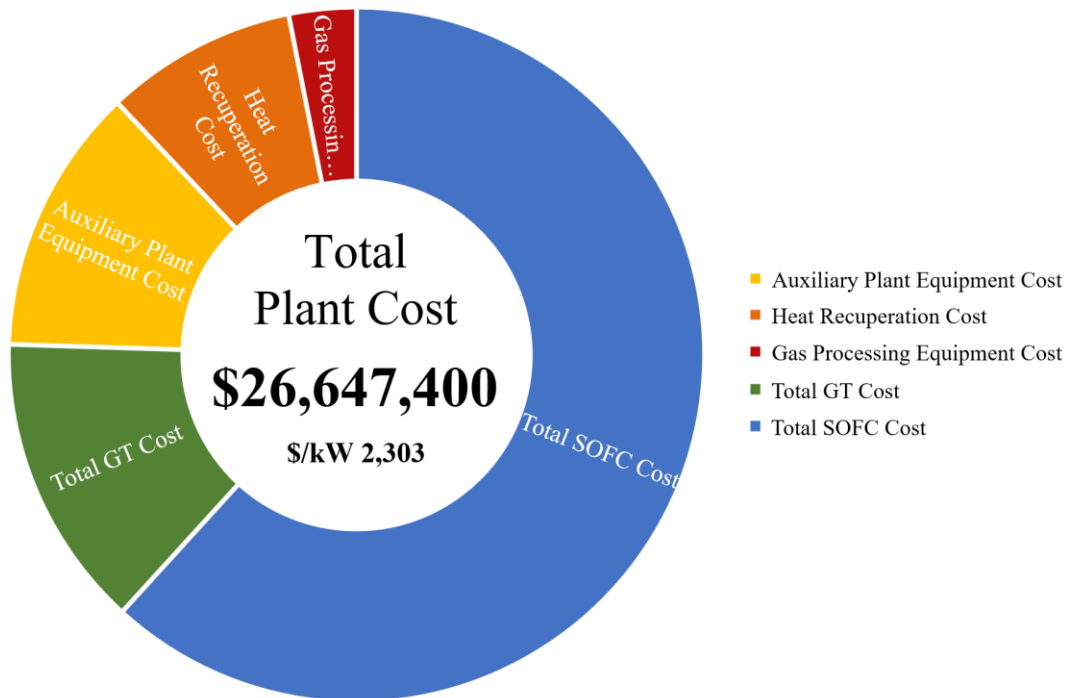


Figure 5-37: SOFC-GT hybrid plant cost breakdown of the total plant cost (Const. Speed, 88.2% FU).

The SOFC cost constitutes the largest cost of the TPC with 61.8%. This phenomenon is even more pronounced when moving to higher FU scenarios where the SOFC cost can be as high as 62.7%. The GT represents the second most expensive TPC category. The GT is an off-the-shelf GT which is purchased at a fixed cost and the rest of the plant is scaled to match the GT. While the GT generates about 11.7% of the

electricity, it is responsible for 13.7% of the TPC. Auxiliary plant equipment represents 12.4%, heat recuperation 8.9% and gas processing 3.1% of the TPC.

Next, the specific plant costs as well as the specific cost for the two power-generating plant units are analyzed on a \$/kW basis as shown in **Figure 5-38**. On the top of the zero-line, the specific plant cost is plotted. The specific GT costs and specific SOFC costs are mirrored at the zero-line. The specific SOFC cost remains fairly constant throughout the various scenarios. Low specific SOFC costs are achieved at low FUs. High FUs require over-proportionally more cell area at constant cell voltage as the chemical potential difference decreases. Other factors impacting the specific cell cost are the SOFC operating pressure. At the encountered pressures, benefits associated with the increase in current density and reduction in cell area outweigh cost increases associated with the SOFC pressure vessel. Thus, for constant FUs, the lowest specific SOFC cost is obtained at the highest pressure ratio.

The specific GT cost exhibits a much wider range of fluctuations. As mentioned before, the GT is an off-the-shelf item that comes at a fixed cost. However, the variable speed operation requires frequency adjustment equipment since the electricity is not generated at grid frequency. Costs associated with frequency adjustment equipment are added to the GT cost in the variable speed scenarios. When correcting

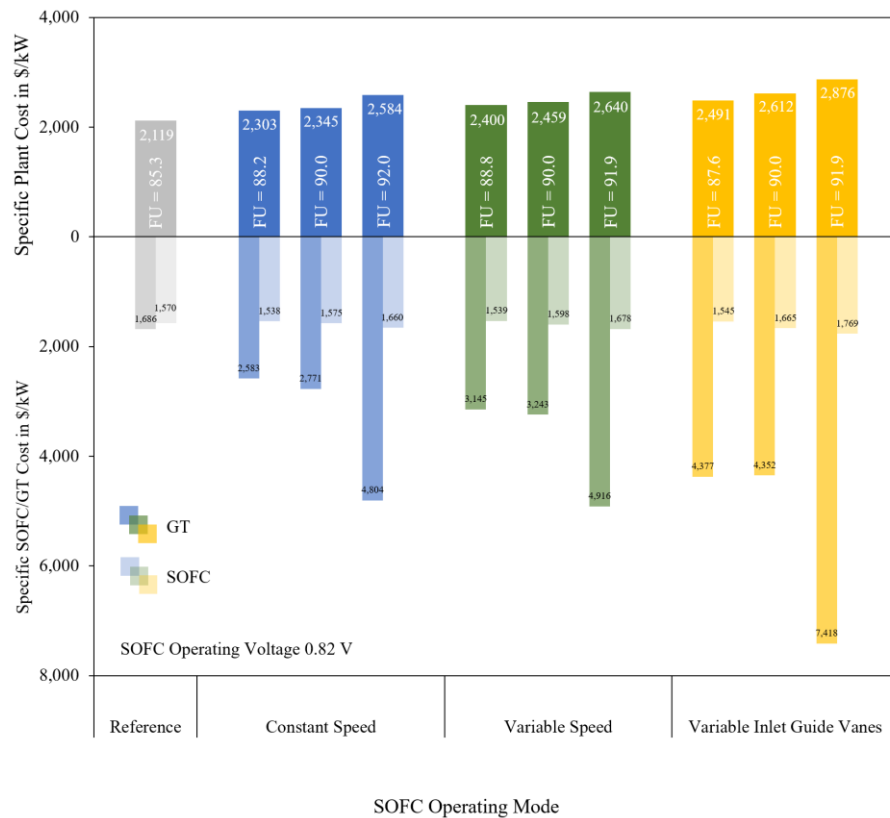


Figure 5-38: Specific plant cost, specific SOFC cost and specific GT cost for SOFC-GT hybrid plants with off-the-shelf GT and reference case with custom-engineered GT.

for this higher GT cost in the variable speed scenarios, the specific GT cost is directly related to the GT power output and follows the inverse trend of the GT power output as shown in **Figure 5-35**.

The lowest specific plant costs are achieved for scenarios where low specific SOFC costs and low specific GT costs are obtained. The TPC and specific plant cost are also dependent on other equipment costs, however, gas processing equipment cost, heat recuperation equipment cost and auxiliary plant equipment cost remain relatively constant throughout the various scenarios and have no significant impact on the trend. Compared to the reference case with custom-designed GT it becomes clear that operating the hybrid not at its ideal conditions adds to the specific SOFC cost, specific GT cost and specific plant cost.

The specific plant cost is a strong indicator for the COE. Nevertheless, plant power output also plays an important role. Fixed costs like operating labor are tied to the number of operators present at the plant and the number of operators is based upon the plant's operating units. Since all plants in this study have the same number of operating units, the same number of operators are needed. Thus, variations in the net power output of the plant have an impact on the operating labor cost on a \$/MWh basis. Increasing the net plant power output decreases the relative labor cost.

All these effects are reflected in the COE and the lowest COEs are achieved for plants with high net power output and low specific plant cost. The lowest COE with 77.47 \$/MWh is reached in the constant speed, 88.2% FU scenario which has the lowest specific plant cost and the second highest net power output. A slightly higher COE is obtained in the constant speed, 90.0% FU scenario which has shown the second lowest specific plant cost and the highest net power output. Also from a purely technical point of view, the lower FU scenario appears more attractive due to slightly lower thermal stress in the SOFC (considering

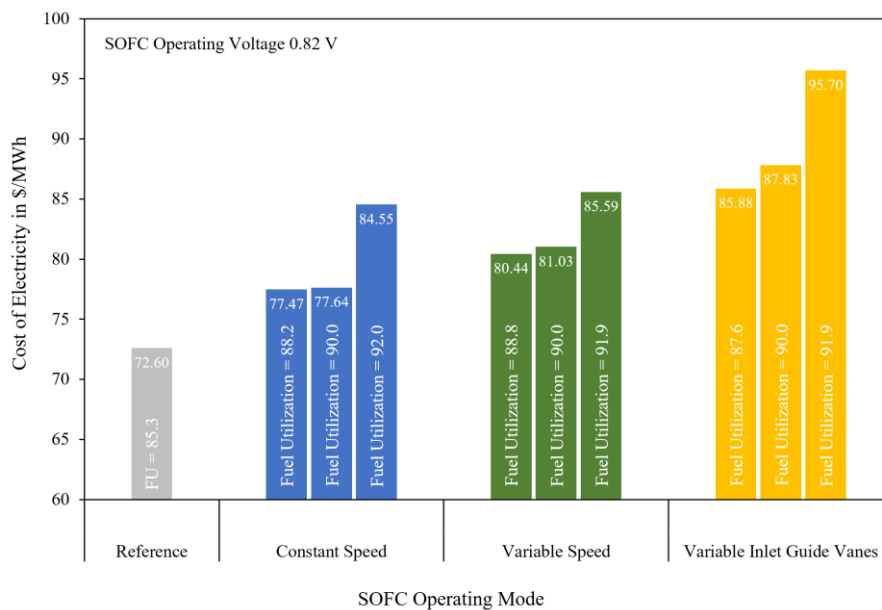


Figure 5-39: Cost-of-electricity of SOFC-GT hybrid plants with off-the-shelf GT and reference case with custom-engineered GT.

the non-limiting thermal constrain parameter) and a higher heating value of the anode off-gas which is beneficial in maintaining stabile oxidation. Compared to the reference case with custom-designed GT, the COE in the best-case scenario with off-the-shelf GT increased by 6.7%. All COE results are summarized in **Figure 5-39**.

In **Figure 5-40**, a detailed breakdown of the different COE driving factors is shown for the constant speed, 88.2% FU scenario. Overnight capital costs have the largest impact on the COE with a share of 34.8% whereby the SOFC island alone is contributing 17.2% to the COE. Variable costs have a share of 33.5%. The by far largest variable cost is the fuel cost with 24.6% despite the high fuel efficiency of the SOFC-GT hybrid plant. Maintenance material costs account for 4.8% of the overall COE and the annual levelized SOFC stack replacement for 3.8%. Fixed costs constitute 31.7% of the COE. Operating labor is the dominant factor in the fixed costs with 16.5%. Administrative & support labor are 4.8% and maintenance labor is 2.9%. Taxes & insurance account for 7.5% of the COE.

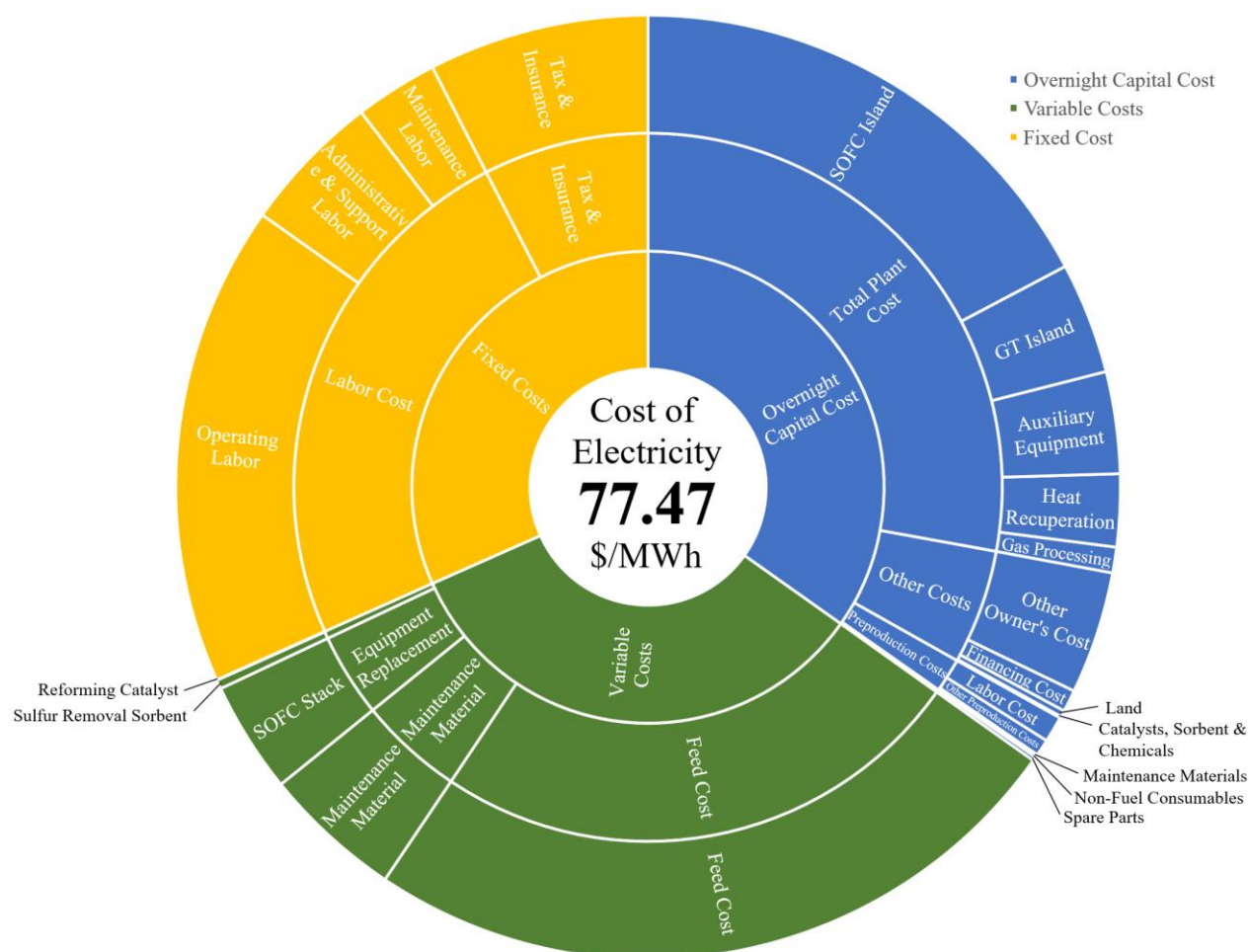


Figure 5-40: SOFC-GT hybrid plant cost breakdown of the cost-of-electricity (Const. Speed, 88.2% FU).

Summary

A SOFC-GT hybrid system with off-the-shelf GT, shown in **Figure 5-32**, has been studied with respect to its thermodynamic and economic performance under various operating conditions considering overall cell temperature differences and local thermal gradients inside the SOFC. Three different GT operating modes have been studied: I) constant speed, II) variable speed and III) VIGV control. The results show that operating the system at constant spool speed leads to the highest net power output. Variable speed control showed to have the highest efficiency and VIGV control was able to provide the largest surge margin. Furthermore, for the three different operating modes, FUs of around 90% are shown to have the highest efficiency, and higher FUs are shown to increase the surge margin.

The economic analysis revealed that the specific cost of the SOFC as well as the GT are reduced when operating the hybrid under low FU factors. As a result, specific plant costs and COE are reduced by operating the plant at relatively low FUs. When comparing a hybrid system with an off-the-shelf-GT to a hybrid system with a custom-engineered GT, the off-the-shelf system results in an increase in specific GT cost, an increase in specific SOFC cost, and an increase in COE of 6.7%.

5.2 SOFC-GT Hybrid Systems with Water Recovery

5.2.1 10 MW Natural Gas SOFC-GT Hybrids

Reference Case

The design of the reference case for the water recovery cases is based upon the previous SOFC-GT hybrid studies. One difference of this design compared to the previous work is the addition of a deoxidizer to remove trace amounts of O_2 that can poison the reforming catalyst in the pre-reformer and/or fuel cell. While O_2 is commonly found in very low concentrations, local and temporal fluctuations, for example during peak shaving, can exceed tolerable O_2 concentrations. To facilitate the removal of O_2 at relatively low concentrations and temperatures, a noble metal catalyst is used to promote the reaction of O_2 with recycled H_2 from the anode off-gas (the reaction of O_2 with CH_4 in the fuel would require substantially higher temperatures and make the plant system integration more complicated). All water recovery studies are conducted using the same thermal SOFC constraints as the previous SOFC-GT hybrid cases, which can be found in **Table 4-13**, together with the other SOFC design parameters. An illustration of the plant design used in the reference case is provided in **Figure 5-41**. The corresponding state-point steam data are shown in **Table 5-4**.

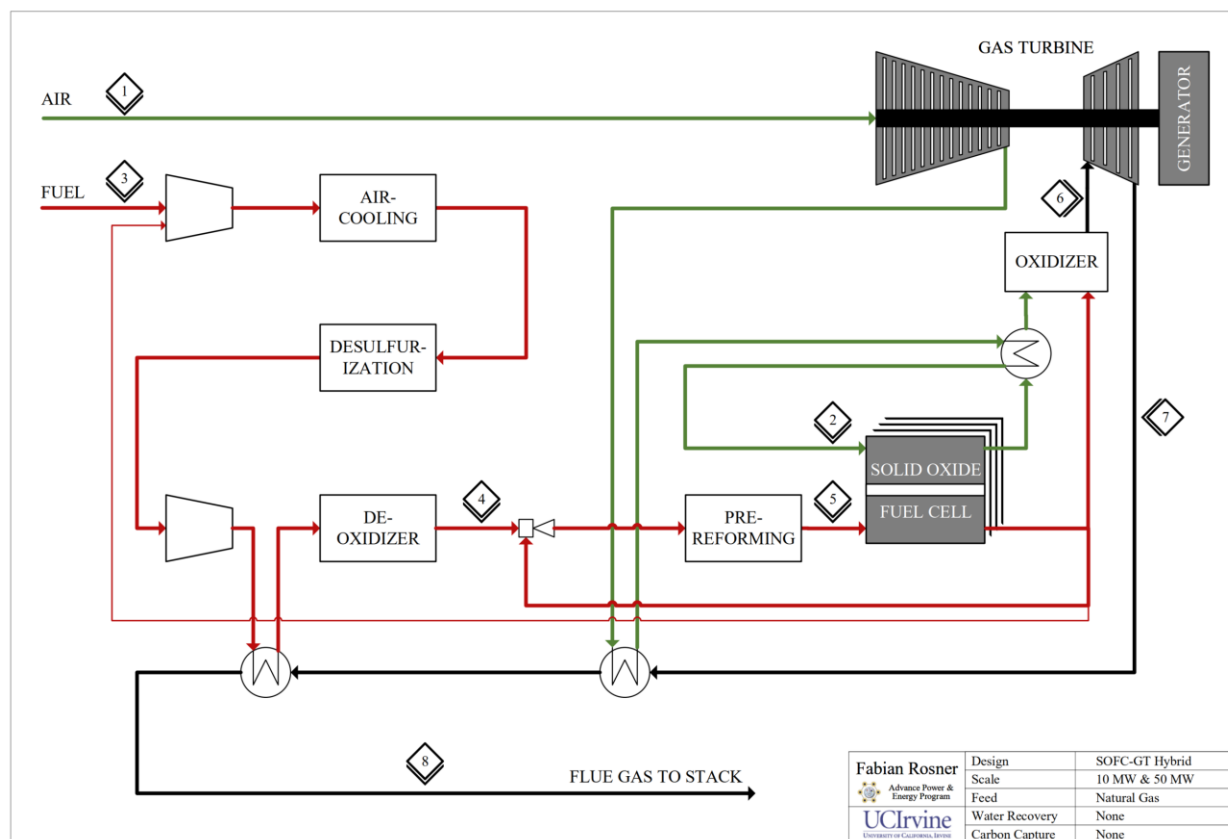


Figure 5-41: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system without water recovery.

Table 5-4: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	909	556	327
Pressure	bar	1.01	7.17	4.10	21.28	7.17	6.57	1.05	1.01
Molar Flow Rate	kmol/h	1,076.4	1,076.4	67.6	67.7	351.2	1,145.9	1,145.9	1,145.9
Mass Flow Rate	kg/h	31,061	31,061	1,171	1,175	7,146	32,232	32,232	32,232
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07486	0.07486	0.07486
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72718	0.72718	0.72718
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00868
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22862	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09023	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01069	0.20398	0.06174	0.06174	0.06174
H ₂ O	-	0.00987	0.00987	0.00003	0.00165	0.32707	0.12754	0.12754	0.12754
CH ₄	-	0.00000	0.00000	0.93087	0.92860	0.14337	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1

Thermodynamic Performance. The net power output of the nominal 10 MW natural gas reference case is 11,554 kW, whereby the SOFC produces 10,683 kW of electricity and the gas turbine produces another 1,414 kW of electricity. The auxiliary load accounts for 543 kW, which includes the DC-AC inverter losses and transformer losses. The DC-AC inverter losses are associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 59% of the total auxiliary load. Second largest auxiliary load is the fuel gas compression, which accounts for 31% of the total auxiliary load.

The 10MW reference case reaches an electrical net efficiency of 75.27%-LHV. The efficiency of the SOFC stack on a standalone basis is 82.0%-LHV, which is similar to the commercially offered BlueGen system with approximately 80%-LHV [339], and is possible due to the high degree of internal reforming. The GT operates at an efficiency of 60.9%-LHV utilizing the low heating value anode off-gas from the SOFC. While the off-design operation reduces the component efficiency of the expander, the efficiency of the GT increases substantially from initially 24.5% at ISO conditions using NG [247]. This is a result of the very high fuel and air inlet temperatures as well as the low heating value of the anode off-gas. Most of the energy is already contained in the air and fuel streams in the form of high temperature heat when they reach the oxidizer, and the oxidation of the unutilized anode off-gas is only able to increase the air temperature by 248 °C, from initially 661 °C to 909 °C. Note that this increase in efficiency is observed despite a decrease in TIT compared to NG operation where a TIT of 943 °C is reached. However, due to the change in exhaust gas composition 909 °C is the maximum permissible TIT in order to maintain the same blade metal temperature.

The plant design requires no water input as it is equipped with dry cooling technology. Furthermore, no continuous liquid or solid waste streams (does not refer to catalysts and sorbents, which are disposed of after their end-of-life) arise from the operation of this plant. Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. The CO₂ emissions of the 10MW natural gas hybrid are 269 kg/MWh, which is substantially lower than PC plants with CO₂ emission values of greater than 775 kg/MWh [34], IGCCs (without carbon capture) with values of greater than 750 kg/MWh [34], and state-of-the-art F-class NGCCs (net-efficiency of 59.4%-LHV) with a CO₂ emission value of 336 kg/MWh [34]. The thermodynamic performance results are summarized in **Table 5-5**.

Table 5-5: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	10,683.43
Gas Turbine	kW	1,413.54
Total Generation	kW	12,096.96
Auxiliary Load		
DC-AC Inverter Losses	kW	320.50
Fuel Gas Compression	kW	169.11
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.85
Transformer Losses	kW	45.49
Total Auxiliary Load	kW	542.97
Net Power Generation	kW	11,553.99
Plant Performance		
Net Efficiency	%-HHV	67.88
Net Efficiency	%-LHV	75.27
Consumables		
Fuel Gas	kg/h	1,171
Thermal Input	MJ _{LHV} /h	55,245
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3114
Carbon Emission	kg/MWh	269
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Economics. The economic analysis follows the methodology outlined earlier. The TPC of the 10MW NG hybrid without water recovery is \$26,736,870 (2,314 \$/kW), whereby, the SOFC island accounts for \$16,445,830 (61.5%). The TPC for the SOFC island includes: stack cost, power conditioning, pressure vessel, housing and assembly, and system installation. The SOFC stack with \$6,330,870 is the single most expensive category of the SOFC island followed by the power conditioning system, which includes the inverters, with \$6,286,450. The GT island contributes \$3,651,010 to the plant's TPC, a share of 13.7%, while generating 11.7% of the gross power. Fuel gas processing represents a relatively small cost. Due to the NG supply pressure of 4.1 bar, on-site compression is reduced, lowering compressor cost. The \$316,170 for gas processing includes an expendable carbon bed process for sulfur removal and the deoxidizer. Heat recuperation accounts for 8.9% of the plant's TPC. In particular, the high temperature heat exchangers for air heating prior to the SOFC are a major cost driving factor due to the expensive materials used for construction. Auxiliary plant equipment costs are estimated with \$3,352,310, which includes electric plant accessory, instrumentation and controls, improvement to site, and buildings and structures. An

illustration of the TPC breakdown into the major categories is shown in **Figure 5-42**. A more detailed overview of the capital cost is provided in **Table 5-6**.

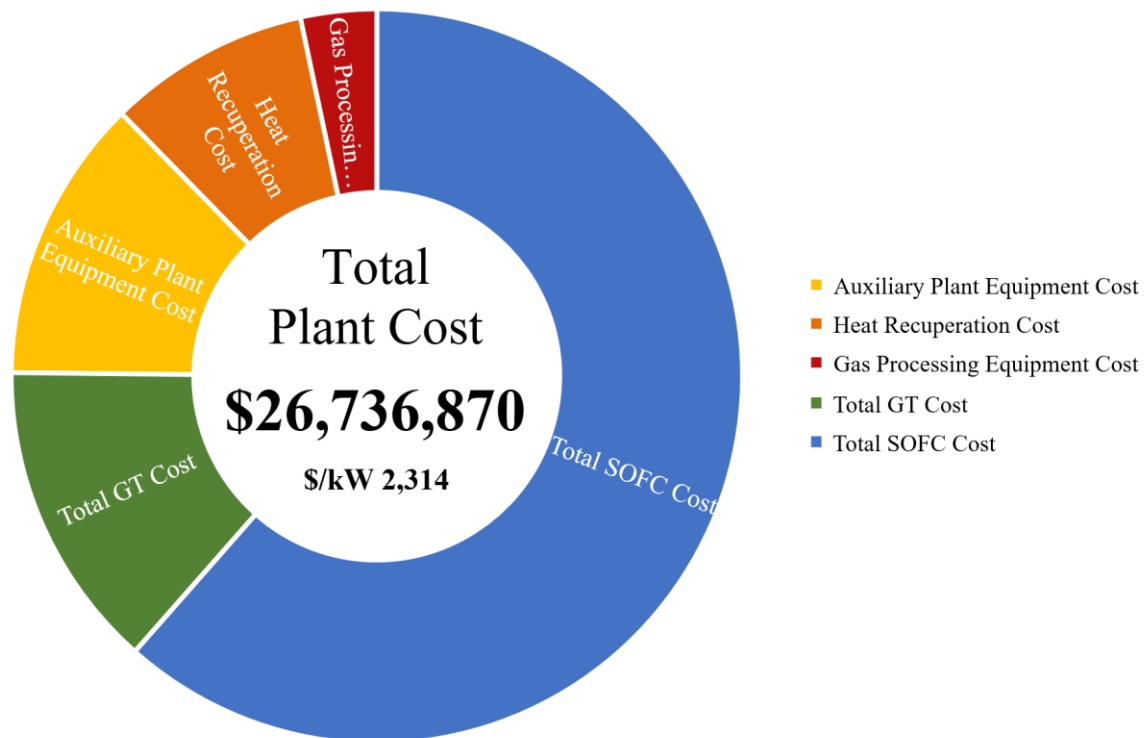


Figure 5-42: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid without water recovery.

The total overnight capital cost is \$33,372,660 and is comprised of the TPC, the preproduction costs with \$1,487,130, the inventory capital costs with \$137,970, and other costs including financing and land costs with \$5,010,690. Variable annual costs are mostly fuel costs with \$1,735,940. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, and waste disposal costs, which amount to \$632,930 per year. Fixed costs are comprised of annual operating labor with \$1,707,210 and costs associated with tax and insurance with \$534,740 per year. The resulting COE for the 10 MW SOFC-GT hybrid plant without water recovery is 77.73 \$/MWh. A detailed breakdown of the overnight capital cost, variable costs and fixed costs is provided in **Table 5-7**. **Figure 5-43** shows how the various costs impact the cost COE.

Table 5-6: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	16,445,830
SOFC Stacks	\$	6,330,870
Power Conditioning	\$	6,286,450
Pressure Vessel, Housing & Assembly	\$	3,252,090
System Installation	\$	576,420
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	898,620
Fuel Compressors and Intercooler	\$	452,110
Fuel Processing	\$	316,170
Flue Gas Stack System	\$	130,340
Heat Recuperation	\$	2,389,090
Cathode Off-gas Heat Exchanger	\$	1,014,810
Air Preheater	\$	1,109,250
Fuel Preheater	\$	265,030
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	N/A
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,352,310
Accessory Electric Plant	\$	1,161,230
Instrumentation & Controls	\$	981,830
Improvement to Site	\$	255,280
Buildings & Structures	\$	953,970
Total Plant Cost	\$	26,736,870
Specific Plant Cost	\$/kW	2,314

Table 5-7: Economics of the 10 MW Natural Gas SOFC-GT Hybrid without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	26,736,870
Preproduction Costs		
6 Months Labor Cost	\$	853,600
1 Month Maintenance Materials	\$	31,380
1 Month Non-Fuel Consumables	\$	27,210
1 Month Waste Disposal	\$	20
25% of 1 month fuel cost at 100% CF	\$	40,180
2% of TPC	\$	534,740
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,290
0.5% of TPC (Spare Parts)	\$	133,680
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	61,030
Land	\$	217,230
Other Owner's Costs (15% of TPC)	\$	4,010,530
Financing Costs	\$	721,900
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,735,940
Other Annual Variable Operating Cost at above CF for Initial Year	\$	632,930
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	0
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,707,210
Property Tax & Insurance	\$	534,740
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	N/A

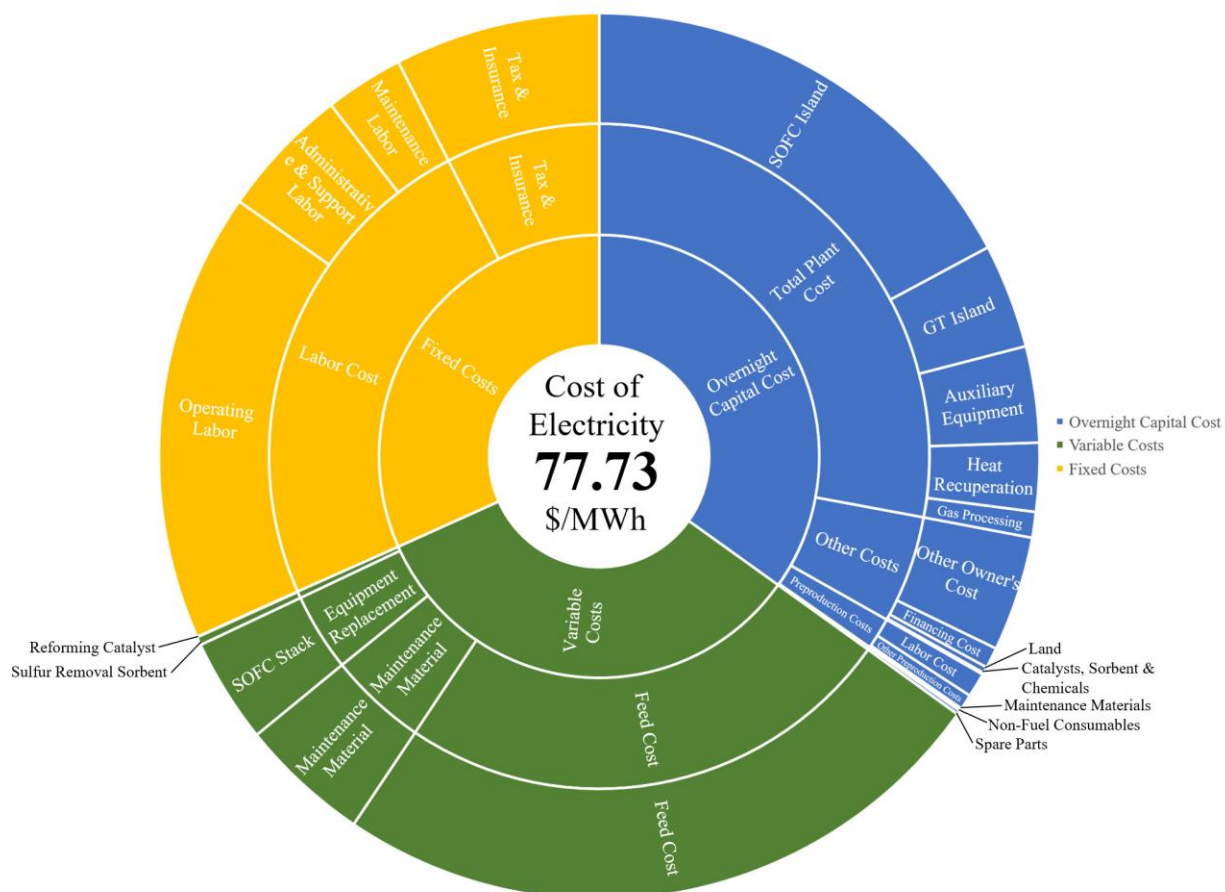


Figure 5-43: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid without water recovery.

Air-Cooled Condenser

The basic configuration of the air-cooled condenser case is identical to the 10 MW NG reference case. In the air-cooled condenser case, an air-cooled condenser is used downstream of the heat recuperation heat exchangers to cool the flue gas below its dew point temperature. After being cooled to about 40 °C, the flue gas leaves the plant through the stack. Critical for this water recovery technology is the flue gas stack design. With a pinch temperature of 25 °C in the air-cooled condenser, buoyancy forces of the flue gas might not suffice to achieve the necessary plume rise and the associated dispersion effects [340]. Thus, additional pressure drop needs to be accounted for to allow for a high stack exit velocity. With a temperature difference of 25 °C between the flue gas and the ambient air, and an gas exit velocity of 20 m/s, a plume rise of 33.5 m can be expected [340], which might be acceptable in some areas. Another important factor to consider is the potential for immediate fog formation as the flue gas leaves the stack at its dew point and will continue cooling down after leaving the stack. Next to the stack's pressure drop, there is a substantial amount of pressure drop associated with the air-cooled condenser, due to the large surface area required to facilitate

the heat exchange between the two gases because of the low heat transfer coefficients. As a result, the gas turbine back pressure increases by 0.09 bar, reducing its power output. The recovered water from the air-cooled condenser is in direct contact with the flue gas, which results in considerable amounts of flue gas constituents ending up in the water. To convert the recovered water to useable water, it is treated in an ultrafiltration unit, reverse osmosis unit and pH adjustment unit. A general flow diagram of the 10MW NG case with air-cooled condenser water recovery is shown in **Figure 5-44**. The respective state-point stream data are provided in **Table 5-8**.

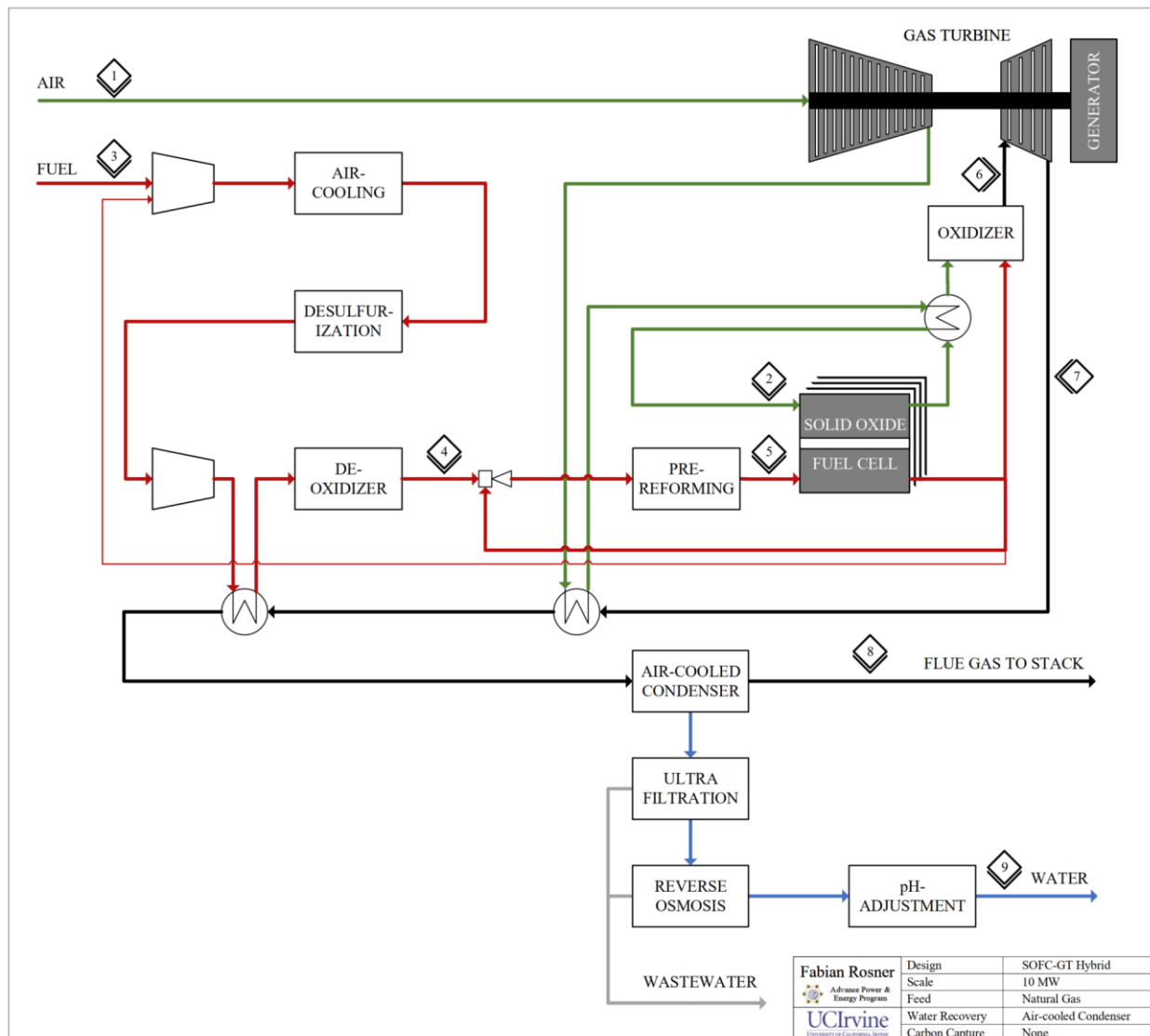


Figure 5-44: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with air-cooled condenser for water recovery.

Table 5-8: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	574	40	41
Pressure	bar	1.01	7.27	4.10	21.74	7.27	6.66	1.14	1.01	2.00
Molar Flow Rate	kmol/h	1,069.2	1,069.2	65.9	66.1	340.6	1,137.0	1,137.0	1,072.6	59.6
Mass Flow Rate	kg/h	30,854	30,854	1,142	1,147	6,946	31,997	31,997	30,836	1,074
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07708	0.07708	0.08170	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00673	0.72795	0.72795	0.77161	0.00090
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00869	0.00869	0.00921	0.00003
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22636	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.08917	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01073	0.20477	0.06071	0.06071	0.06436	Trace
H ₂ O	-	0.00987	0.00987	0.00003	0.00172	0.32874	0.12557	0.12557	0.07311	0.99902
CH ₄	-	0.00000	0.00000	0.93087	0.92850	0.14422	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03191	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	0.99995

Thermodynamic Performance. The net power output of the nominal 10MW NG plant with air-cooled condenser for water recovery is 11,123 kW, 431 kW less than the reference plant. The gas turbine power output decreased from 1,414 kW in the reference case to 1,203 kW in the air-cooled condenser water recovery case, a decrease of 14.9%. This reduction in GT power output is related to the increased backpressure, which reduces the pressure ratio across the GT expander, despite the fact that the compressor pressure ratio increases. As a result of the higher compressor discharge pressure the GT operates closer to its surge limit with a surge margin of 9.4%, a reduction of 2.1% points compared to the reference case (11.5% surge margin). The higher pressure ratio also has implications for the SOFC, which now operates at a slightly higher pressure. This leads to a small increase in cooling demand in order to not exceed the thermal constraints of the cell, which in turn leads to a downscaling of the SOFC (10,485 kW, 198 kW less than in the reference case). The change in compressor air flow due to the change in pressure ratio is small compared to the downscaling effect of the SOFC and responsible for approximately 30% of the change in scale. The operating voltage remains constant throughout all study cases; similarly, the overall fuel utilization is essentially the same as in the reference case, 88.6% versus 88.2% (a change in the FU by 0.1% points at the respective plant scale results in an approximately 12 kW change in SOFC power output). Thus, SOFC efficiency and FU effects account for about 25% of the change in SOFC power output. In a SOFC-GT hybrid the SOFC and GT are strongly coupled, and the respective operating points impact one another. First, the GT compressor discharge pressure causes a change in the SOFC's cooling demand, which then impacts the maximum permissible TIT of the GT, which again impacts the pressure ratio. Due to the higher cooling demand, the fuel to air ratio changes and the maximum permissible TIT increases from 909.3 °C

to 910.5 °C. However, the increased TIT cannot compensate for the power output decrease resulting from the increased GT backpressure. The auxiliary load is largely similar to the reference case; however, the water recovery and water treatment systems slightly increase the auxiliary power consumption to 565 kW and represent about 5.7% of the auxiliary load.

The electrical efficiency of the air-cooled water recovery case decreased by 1.00% points to 74.27%-LHV and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GT due to the backpressure. The GT efficiency decreased to 55.3%-LHV. The stack efficiency remains almost unchanged at 81.9%-LHV. The minimal decrease is a result of the higher FU but also internal reforming impacts the stack efficiency.

Due to the plant size reduction, carbon emissions are slightly decreased; however, the decline of the efficiency increases the carbon emissions on a per MWh basis to 273 kg/MWh. The water recovery from the flue gas in the air-cooled condenser is 45.1% and limited by the ambient temperature. At the scale of this nominal 10 MW plant, a total of 25,735 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW

Table 5-9: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Gross Power Generation	Unit	Air-cooled Condenser
Fuel Cell	kW _{DC}	10,485.28
Gas Turbine	kW	1,202.71
Total Generation	kW	11,687.99
Auxiliary Load		
DC-AC Inverter Losses	kW	314.56
Fuel Gas Compression	kW	166.60
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	29.77
Water Treatment	kW	2.18
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.58
Transformer Losses	kW	43.95
Total Auxiliary Load	kW	564.67
Net Power Generation	kW	11,123.32
Plant Performance		
Net Efficiency	%-HHV	66.98
Net Efficiency	%-LHV	74.27
Consumables		
Fuel Gas	kg/h	1,142
Thermal Input	MJ _{LHV} /h	53,898
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3038
Carbon Emission	kg/MWh	273
Liquid Waste	kg/h	87
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	45.1
Flue Gas Water Recovery	kg/d	25,735

SOFC-GT Hybrid Systems with Water Recovery

SOFC-GT hybrid plant with air-cooled condenser water recovery system can produce enough water for the irrigation of over 13,300 m² (including 90% capacity factor). While there have been no liquid or solid waste streams in the reference case, the need for water treatment of the recovered water results in the generation of liquid waste from the ultrafiltration unit and the reverse osmosis unit. This lowers the amount of export water and is also associated with disposal costs. The plant performance data are summarized in **Table 5-9**.

Economics. The TPC of the 10 MW NG case with air-cooled condenser for water recovery is \$27,477,110 and the resulting specific plant cost is 2470 \$/kW. The cost contribution of the SOFC island remains almost constant with a value of \$16,346,920. The stack cost slightly increased due to pressure effects and the fact that for both plants, the reference case and the air-cooled condenser cases, the same amount of SOFC modules are needed, despite the lower power output in the air-cooled condenser case. However, unused spare stacks are counted towards the stack replacement after 10 years. Power condition costs are reduced due to the reduced power output and costs associated with the pressure vessel, housing, and assembly slightly increased due to the higher SOFC operating pressure in the air-cooled condenser case. GT island costs are identical to the reference case since the same Dresser Rand KG2-3G/EF GT is used in all the water recovery cases. Fuel gas processing costs and flue gas stack costs are slightly reduced, due to reduced flow rates, as well as the heat recuperation costs. Air heating of the GT compressor discharge is slightly shifted

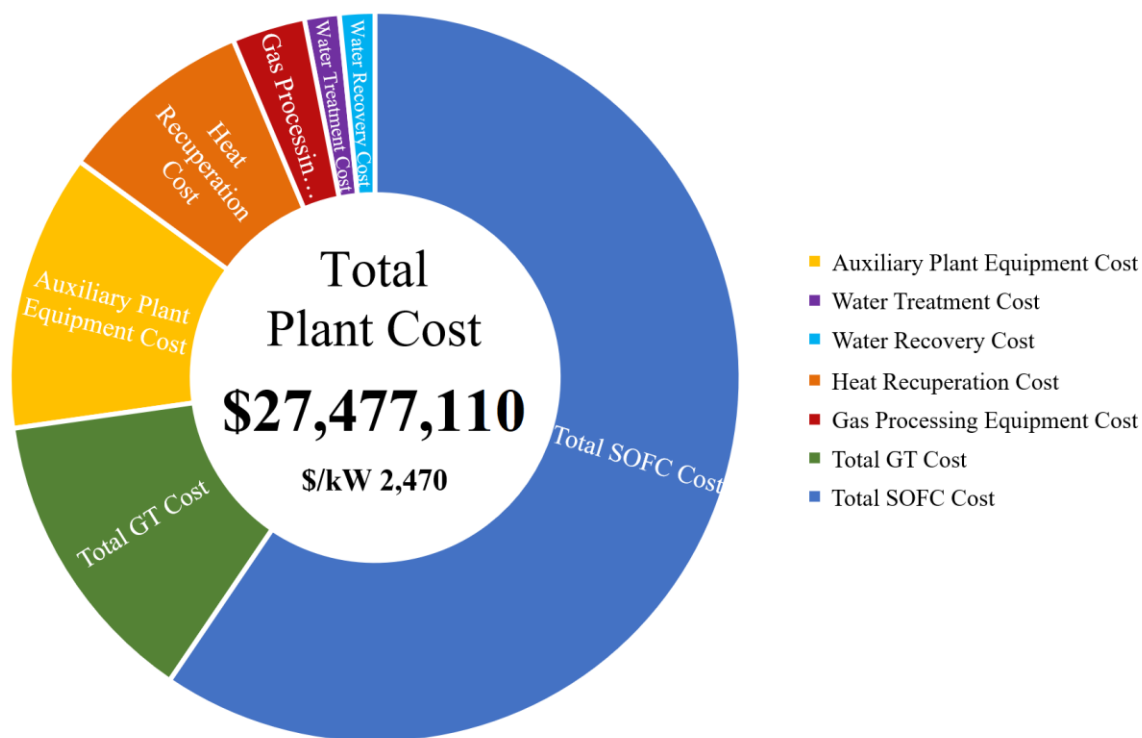


Figure 5-45: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser for water recovery.

towards the GT exhaust gas, which is also reflected in the heat recuperation cost. The water recovery and water treatment add a cost of \$858,970 to the plant's TPC (3.1%), whereby the water recovery accounts for \$426,660 and the water treatment accounts for \$432,310. The auxiliary plant equipment cost is responsible for \$3,373,710. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-45**. A more detailed cost summary of all plant equipment is provided in **Table 5-10**.

Table 5-10: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Equipment Cost	Unit	Air-cooled Condenser
Fuel Cell Island	\$	16,346,920
SOFC Stacks	\$	6,335,270
Power Conditioning	\$	6,165,200
Pressure Vessel, Housing & Assembly	\$	3,272,630
System Installation	\$	573,820
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	882,290
Fuel Compressors and Intercooler	\$	440,370
Fuel Processing	\$	312,150
Flue Gas Stack System	\$	129,760
Heat Recuperation	\$	2,364,200
Cathode Off-gas Heat Exchanger	\$	959,820
Air Preheater	\$	1,143,160
Fuel Preheater	\$	261,230
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	858,970
Water Recovery System	\$	426,660
Water Treatment	\$	432,310
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,373,710
Accessory Electric Plant	\$	1,121,980
Instrumentation & Controls	\$	1,009,020
Improvement to Site	\$	262,340
Buildings & Structures	\$	980,380
Total Plant Cost	\$	27,477,110
Specific Plant Cost	\$/kW	2,470

The total overnight capital cost is \$34,396,030, which represents an increase of over \$1,023,370 compared to the reference case, mostly due to the TPC increase in the water recovery case. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. In the water recovery cases, water byproduct sales reduce the variable operating costs. In the air-cooled condenser case, the water byproduct revenue is \$600,660 for the initial year. The fixed operating cost in the air-cooled condenser case increases

due to labor costs associated with the water recovery as well as tax and insurance costs of the additional equipment. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COEs for all 10 MW NG cases remain constant at 77.73 \$/MWh. The COW in the air-cooled condenser case is 0.0709 \$/L. A summary of the economic key parameters can be found in **Table 5-11**. A detailed breakdown of the COE is shown in **Figure 5-46**.

Table 5-11: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Item	Unit	Air-cooled Condenser
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	27,477,110
Preproduction Costs		
6 Months Labor Cost	\$	979,770
1 Month Maintenance Materials	\$	32,620
1 Month Non-Fuel Consumables	\$	27,300
1 Month Waste Disposal	\$	2,050
25% of 1 month fuel cost at 100% CF	\$	39,210
2% of TPC	\$	549,540
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,210
0.5% of TPC (Spare Parts)	\$	137,390
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	59,500
Land	\$	223,880
Other Owner's Costs (15% of TPC)	\$	4,121,570
Financing Costs	\$	741,880
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,693,710
Other Annual Variable Operating Cost at above CF for Initial Year	\$	669,260
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	600,660
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,959,550
Property Tax & Insurance	\$	549,540
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0709

Figure 5-46 shows the COE value of 77.73 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-6.85 \$/MWh) and the whole circle represents the COE without byproduct credit (84.57 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 84.57 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0709 \$/L in order to reach the COE of 77.73 \$/MWh.

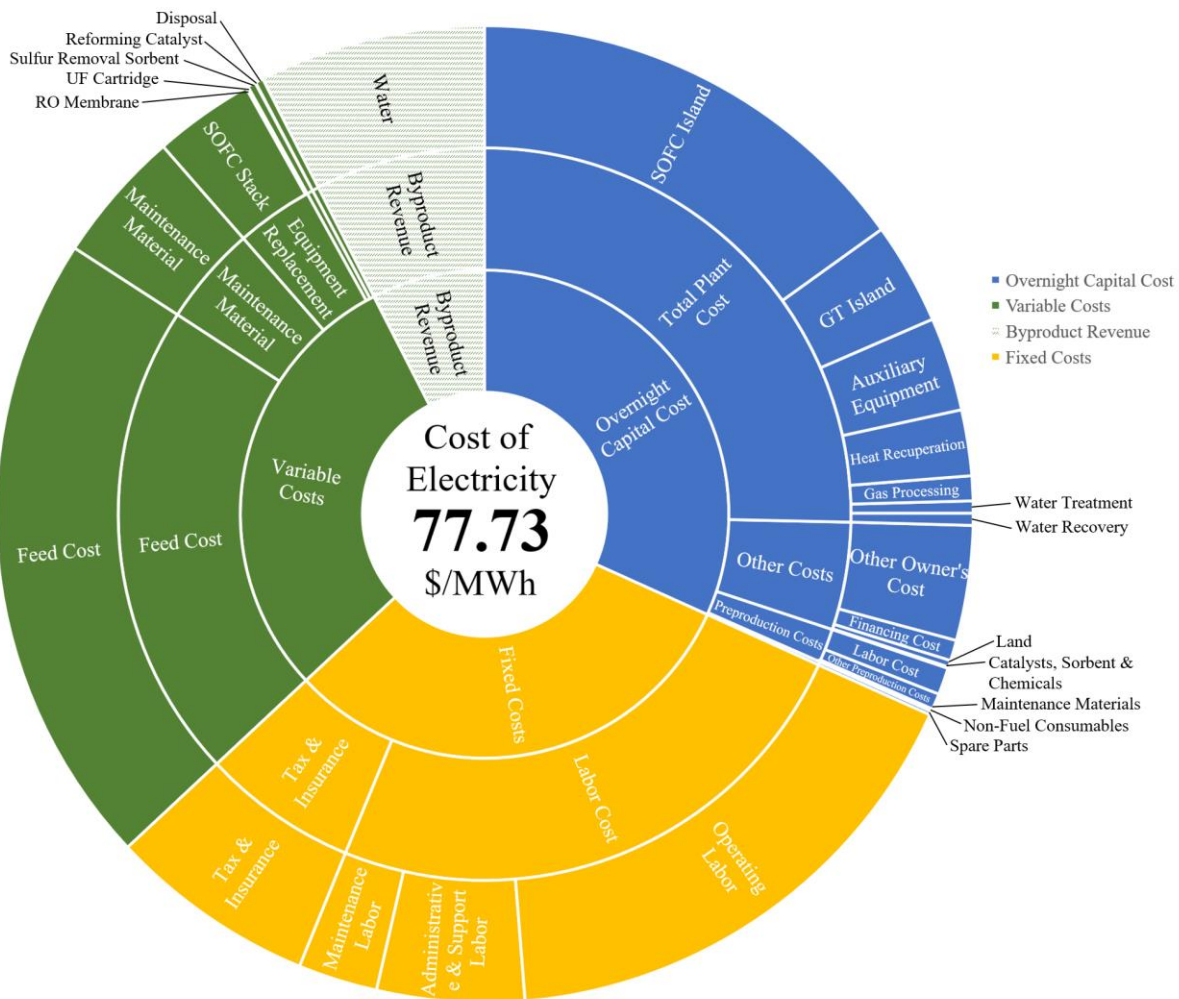


Figure 5-46: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser for water recovery.

The COW is composed of costs associated with water recovery, water treatment and performance penalty costs. The water treatment accounts for 35.7% of the COW, whereby the fixed operating costs, mostly labor, represent the largest share (in this analysis, the labor costs associated in with the water recovery and treatment are equally split between recovery and treatment). Variable water treatment costs account for about 5.9% of the COW, which includes annual maintenance, equipment replacement and disposal costs. The water treatment TPC accounts for 5.8% of the COW. Water recovery contributes 30.8% to the COW with fixed operating costs being the largest contributor. The TPC of the water recovery unit has a 5.7%-share of the COW. Performance penalties come from the parasitic power consumption of the water recovery (3.0% of COW) and water treatment (0.2% of COW) as well as system integration losses, such as higher GT backpressure, which reduces the GT power output. As mentioned before, the GT power output decreased by 14.9% in the air-cooled condenser case compared to the reference case, which

substantially contributes the system integration losses and its share in the COW is 30.3%. A detailed breakdown of the COW can be found in **Figure 5-47**.

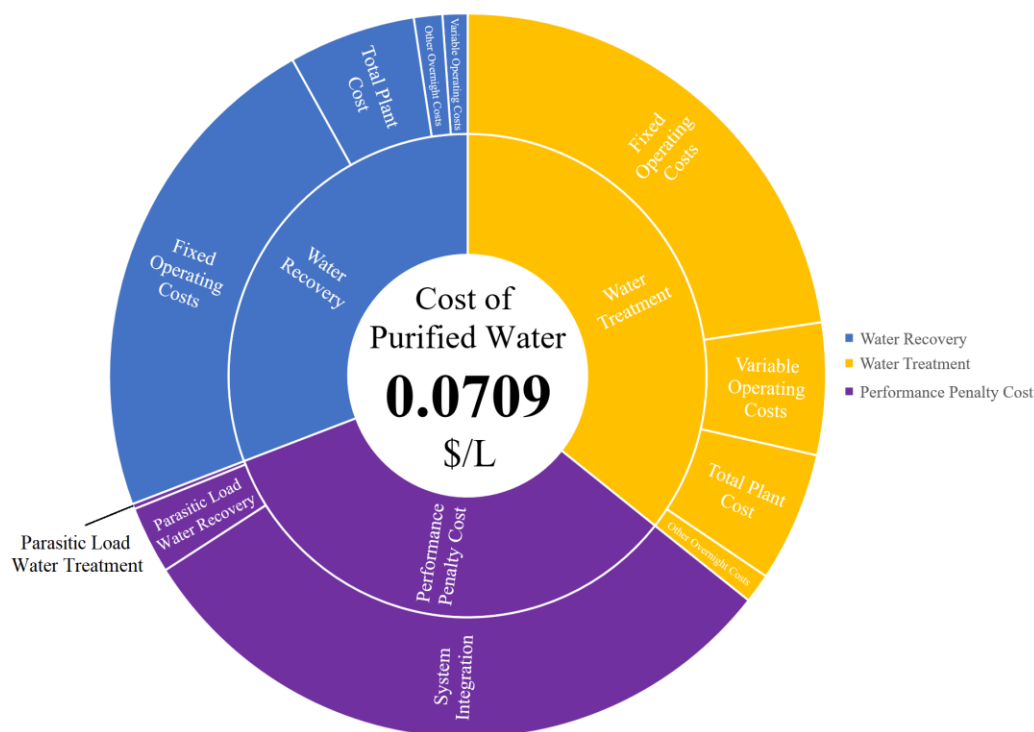


Figure 5-47: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser for water recovery.

Air-Cooled Condenser with Zero-Liquid-Discharge System

The design of the air-cooled condenser case with zero-liquid-discharge system is largely identical to the air-cooled condenser case without zero-liquid-discharge system. Differences in the flowsheet arise from the integration of a brine concentrator and a spray crystallizer. The brine concentrator is a falling film evaporator utilizing a vapor compression cycle as energy source to drive the evaporation. The brine concentrator feed is the rejected water from the UF (backwash and clean-in-place) and the RO unit (retentate). The brine concentrator increases the content of dissolved solids in the rejected water and simultaneously recovers water from this waste stream, which increases the net water production of the power plant. The concentrated wastewater stream is then injected into a spray crystallizer, which uses a small portion of the hot flue gas to provide the heat for the evaporation of the remaining water contained in the waste stream. A general flow diagram of the 10MW NG case with air-cooled condenser water recovery is shown in **Figure 5-48**. The respective state-point stream data are provided in **Table 5-12**.

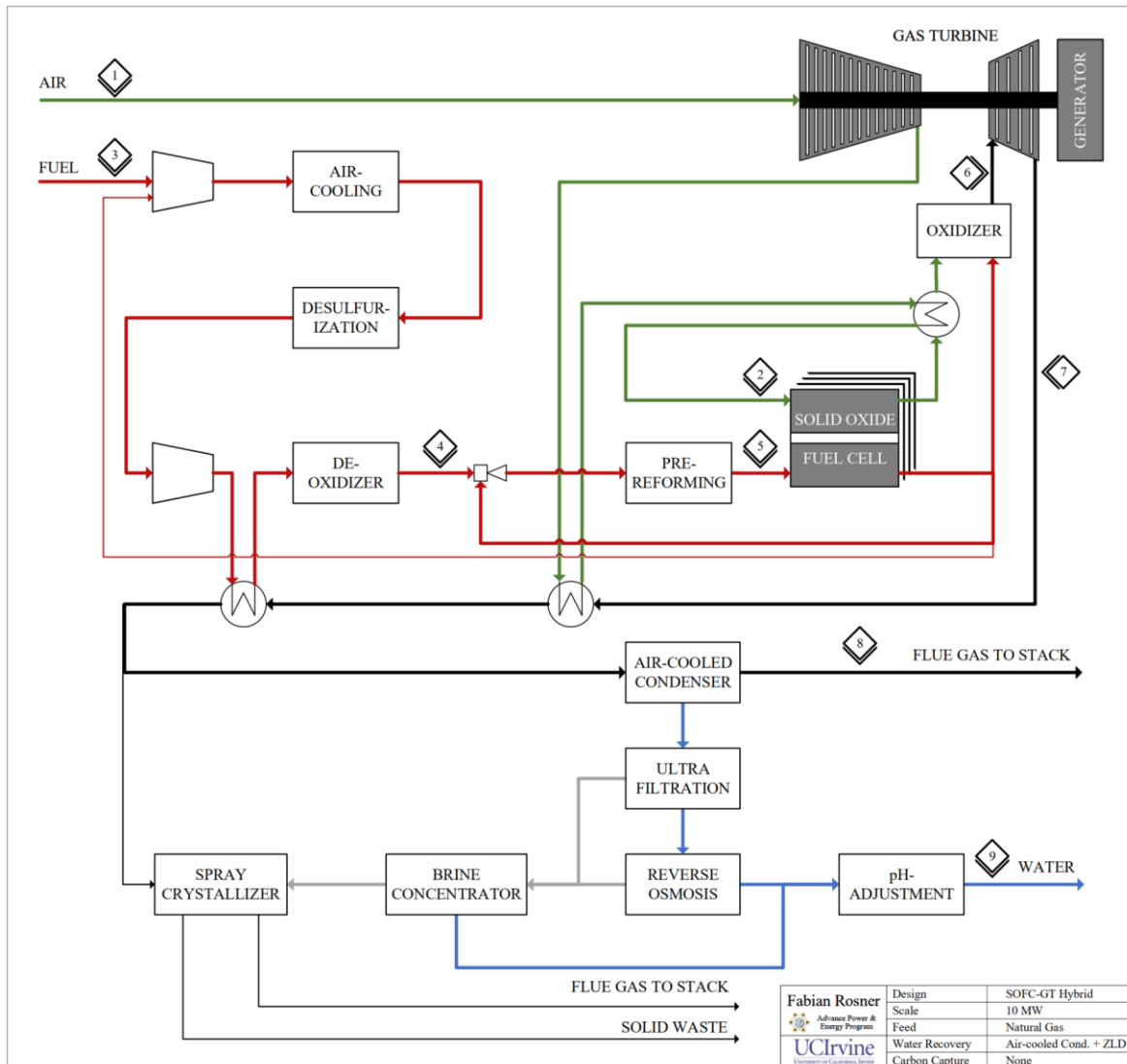


Figure 5-48: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with air-cooled condenser and zero-liquid-discharge system for water recovery.

Table 5-12: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser with Zero-Liquid-Discharge System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	574	40	41
Pressure	bar	1.01	7.27	4.10	21.74	7.27	6.66	1.14	1.01	2.00
Molar Flow Rate	kmol/h	1,069.2	1,069.2	65.9	66.1	340.6	1,137.0	1,137.0	1,070.2	63.9
Mass Flow Rate	kg/h	30,854	30,854	1,142	1,147	6,946	31,997	31,997	30,767	1,151
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07708	0.07708	0.08170	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00673	0.72795	0.72795	0.77161	0.00084
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00869	0.00869	0.00921	0.00003
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22636	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.08917	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01073	0.20477	0.06071	0.06071	0.06436	Trace
H ₂ O	-	0.00987	0.00987	0.00003	0.00172	0.32874	0.12557	0.12557	0.07311	0.99909
CH ₄	-	0.00000	0.00000	0.93087	0.92850	0.14422	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03191	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	0.99995

Thermodynamic Performance. The thermodynamic performance of the plant is summarized in **Table 5-13**.

Plant modifications due to the zero-liquid-discharge system are limited to the water side downstream of the air-cooled condenser (and the flue gas slip stream for the spray crystallizer). Thus, there are no changes in the power generating units as well as the majority of the auxiliary loads. The auxiliary load of the water recovery unit is slightly reduced due to the slightly lower flue gas flow to the condenser and the load associated with the water treatment is slightly increased due to the overall higher amount of water exported from the plant. Additionally, the zero-liquid-discharge system requires electricity for moving gas and liquid. As a result, the efficiency of the air-cooled condenser case with zero-liquid-discharge system is with 74.25%-LHV minimally lower than the air-cooled condenser case without zero-liquid-discharge system.

Emission values of the plant remain virtually constant; however, the liquid waste was reduced to 0 kg/h via the zero-liquid-discharge system by turning it into a solid waste stream of 85 g/h. While the flue gas water recovery remains constant at 45.1%, the water recovery, from the rejected water from the UF and RO, enhances the plant's water output to 27,590 kg/d.

Table 5-13: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser and Zero-Liquid-Discharge System for Water Recovery

Gross Power Generation	Unit	Air-cooled Condenser + ZLD
Fuel Cell	kW _{DC}	10,485.28
Gas Turbine	kW	1,202.71
Total Generation	kW	11,687.99
Auxiliary Load		
DC-AC Inverter Losses	kW	314.56
Fuel Gas Compression	kW	166.60
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	29.70
Water Treatment	kW	2.19
Zero Liquid Discharge System	kW	3.32
Miscellaneous Balance of Plant	kW	7.58
Transformer Losses	kW	43.95
Total Auxiliary Load	kW	567.93
Net Power Generation	kW	11,120.06
Plant Performance		
Net Efficiency	%-HHV	66.96
Net Efficiency	%-LHV	74.25
Consumables		
Fuel Gas	kg/h	1,142
Thermal Input	MJ _{LHV} /h	53,898
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,038
Carbon Emission	kg/MWh	273
Liquid Waste	kg/h	0
Solid Waste	g/h	85
Water Recovery		
Flue Gas Water Recovery	%	45.1
Flue Gas Water Recovery	kg/d	27,590

Economics. Due to the zero-liquid-discharge system, the TPC increases from \$27,477,110 in the air-cooled condenser case without zero-liquid-discharge system to \$28,564,790 in the air-cooled condenser case with zero-liquid-discharge system. The TPC for the GT island, gas processing and heat recuperation remain constant. The fuel cell island TPC remains essential constant too, except a minor cost reduction in the power conditioning equipment at the grid interface. The zero-liquid-discharge system alone adds \$998,330 to the TPC, which is shown in **Figure 5-49** under the rubric water treatment. The auxiliary plant cost also increases as the need for additional instrumentation, controls, land, buildings and structures increases. A detailed summary of the TPC is provided in **Table 5-14**.

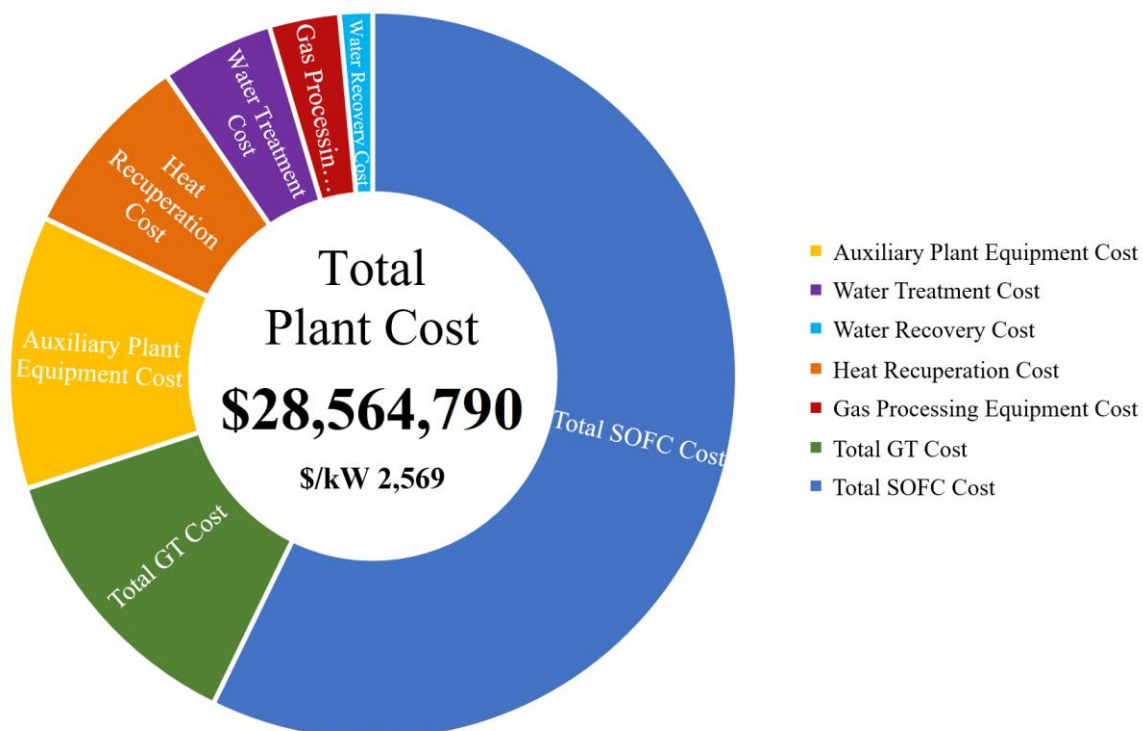


Figure 5-49: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser and zero-liquid-discharge system for water recovery.

The overnight capital cost in the air-cooled condenser case with zero-liquid-discharge system is \$35,711,090. The zero-liquid-discharge system leads to an increase in pre-production costs due to the increase in labor, maintenance materials, non-fuel consumables and other TPC related preproduction costs. However, the costs associated with waste disposal could be drastically reduced as the waste stream was reduced from 87 kg/h of wastewater to 85 g/h of solid waste. In order to obtain a COE of 77.73 \$/MWh (same as in reference case) the exported water needs to be sold at a price of 0.0811 \$/L, which leads to a total revenue of \$736,480 per year. Thus, the increased water export from the plant site due to the zero-liquid-discharge system cannot offset the increased costs associated with it. If the water would be provided to the local community at no cost, the COE would increase to 86.12 \$/MWh. A summary of the plant economics is provided in **Table 5-15** and a detailed breakdown of the cost of electricity is shown in **Figure 5-50**.

Table 5-14: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser and Zero-Liquid-Discharge System for Water Recovery

Equipment Cost	Unit	Air-cooled Condenser + ZLD
Fuel Cell Island	\$	16,346,850
SOFC Stacks	\$	6,335,270
Power Conditioning	\$	6,165,130
Pressure Vessel, Housing & Assembly	\$	3,272,630
System Installation	\$	573,820
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	882,290
Fuel Compressors and Intercooler	\$	440,380
Fuel Processing	\$	312,150
Flue Gas Stack System	\$	129,760
Heat Recuperation	\$	2,364,200
Cathode Off-gas Heat Exchanger	\$	959,820
Air Preheater	\$	1,143,160
Fuel Preheater	\$	261,230
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	1,857,580
Water Recovery System	\$	426,250
Water Treatment	\$	433,000
Zero Liquid Discharge System	\$	998,330
Auxiliary Plant Equipment Cost	\$	3,462,850
Accessory Electric Plant	\$	1,121,980
Instrumentation & Controls	\$	1,048,960
Improvement to Site	\$	272,730
Buildings & Structures	\$	1,019,190
Total Plant Cost	\$	28,564,790
Specific Plant Cost	\$/kW	2,569

Figure 5-51 shows the various driving factors of the COW. The zero-liquid-discharge system is responsible for 31.4% of the COW. However, it needs to be mentioned that the operating labor for the water recovery systems are now equally split between the water recovery, water treatment and zero-liquid-discharge system (same number of operators needed as in the air-cooled condenser case without zero-liquid-discharge system). The TPC of the zero-liquid-discharge system accounts for 10.9% of the COW and the variable operating costs of the zero-liquid-discharge system account for 2.4% of the COW. The second largest category is the performance penalty cost with 28.4%, whereby the system integration losses are responsible for 25.5% points. The share of the parasitic load of the zero-liquid-discharge system is small with a value of 0.3%. Water treatment accounts for 20.6% of the COW and the actual water recovery is the smallest category with a share of 19.6%.

Table 5-15: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with Air-cooled Condenser and Zero-Liquid-Discharge System for Water Recovery

Item	Unit	Air-cooled Condenser + ZLD
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	28,564,790
Preproduction Costs		
6 Months Labor Cost	\$	987,600
1 Month Maintenance Materials	\$	34,560
1 Month Non-Fuel Consumables	\$	27,290
1 Month Waste Disposal	\$	20
25% of 1 month fuel cost at 100% CF	\$	39,210
2% of TPC	\$	571,300
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,210
0.5% of TPC (Spare Parts)	\$	142,820
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	59,500
Land	\$	223,820
Other Owner's Costs (15% of TPC)	\$	4,284,720
Financing Costs	\$	771,250
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,693,710
Other Annual Variable Operating Cost at above CF for Initial Year	\$	668,300
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	736,480
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,975,190
Property Tax & Insurance	\$	571,300
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0811

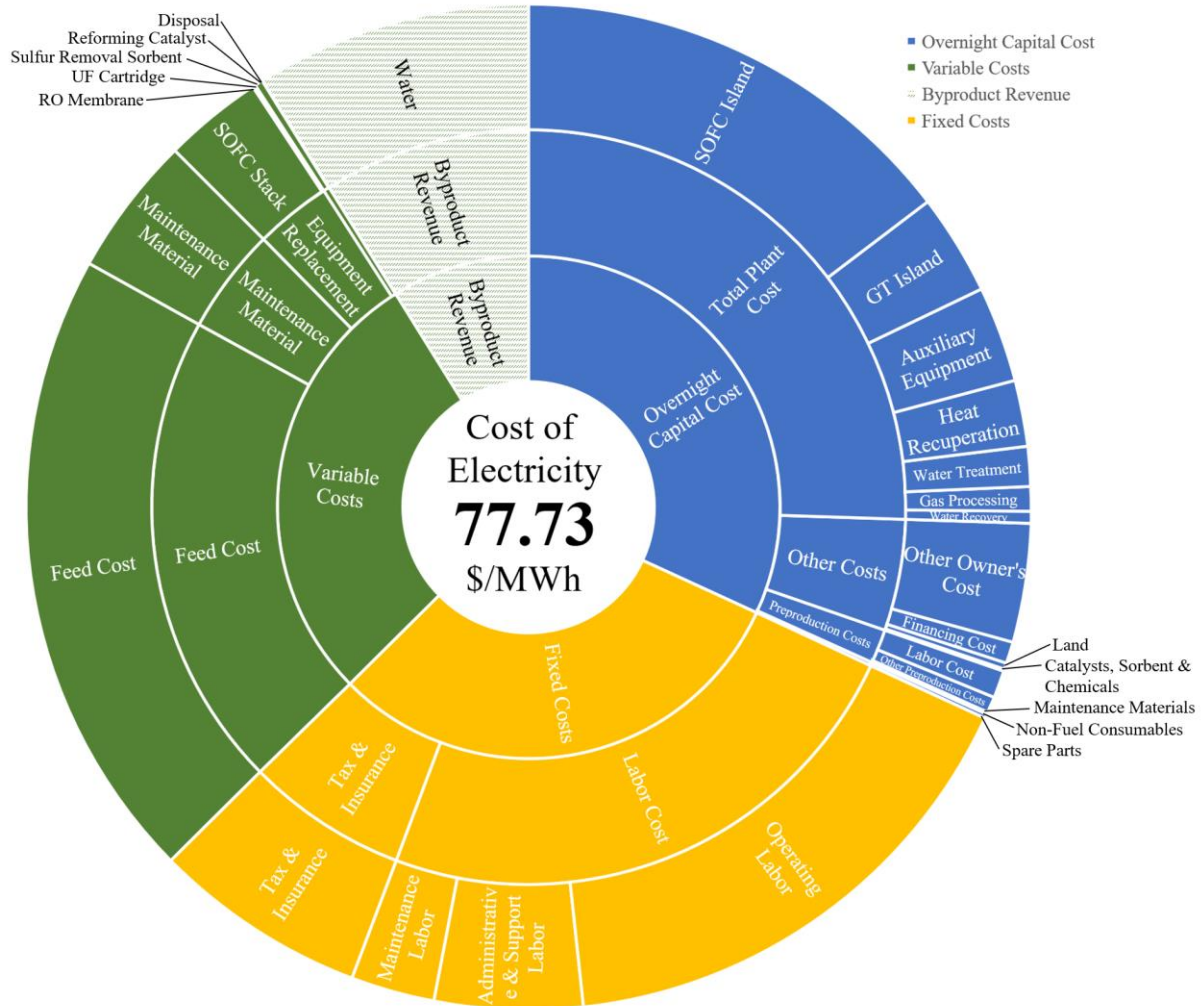


Figure 5-50: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser and zero-liquid-discharge system for water recovery.

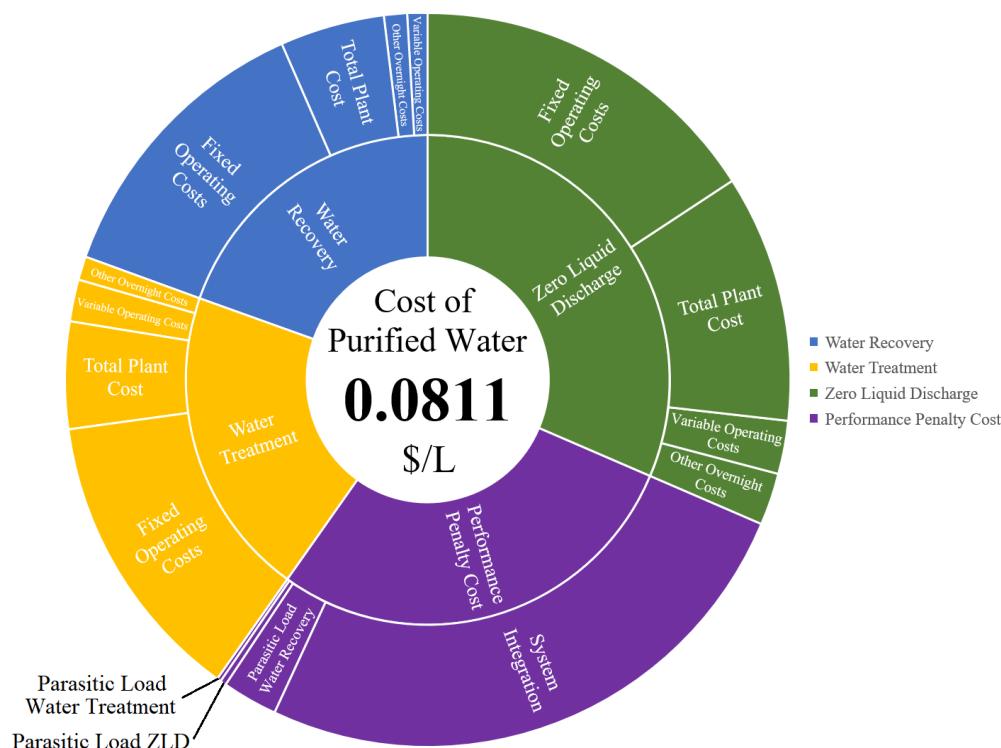


Figure 5-51: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with air-cooled condenser and zero-liquid-discharge system for water recovery.

Direct Contact Condenser

The plant configuration of the SOFC-GT hybrid with direct contact condenser is very similar to the air-cooled condenser case without zero liquid discharge system. The direct contact condenser is a counterflow column made of polypropylene that is installed downstream of the heat recuperation section. Prior to the column a water spray is employed to reduce the flue gas temperature to acceptable inlet temperatures. The column uses a structured bed to create a large surface area inside the column to allow for effective heat and mass transfer. The advantage of using a structured packing is its low pressure drop. In a direct contact condenser, there is no barrier (e.g. metal tube) between the hot fluid and the liquid coolant allowing for lower pinch temperatures compared to the air-cooled condenser design. Nonetheless, the sensitive heat and latent heat from the flue gas needs to be rejected and the recycled water/coolant temperature is limited by the ambient air temperature. Consequently, the flue gas outlet temperature in the direct contact condenser is higher than in the air-cooled condenser case. This is beneficial with respect to plume rise since it increases buoyancy forces; however, the flue gas still leaves the plant at a saturated state, meaning that fog formation is possible. Since the direct contact condenser works on the principle of gas dew point condensation, the higher outlet temperature results in a reduction of water recovery. Furthermore, in the column, the water is in direct contact with the flue gas, which allows trace components to dissolve in the water, making it

necessary to treat the water adequately before it can be used in society. For the water treatment, the same strategy as in the air-cooled condenser case is used, consisting of an ultra-filtration, reverse osmosis, and pH-adjustment step. The flowsheet of the direct condenser case can be found in **Figure 5-52**. The accompanying state-point stream data as indicated in **Figure 5-52** can be found in **Table 5-16**.

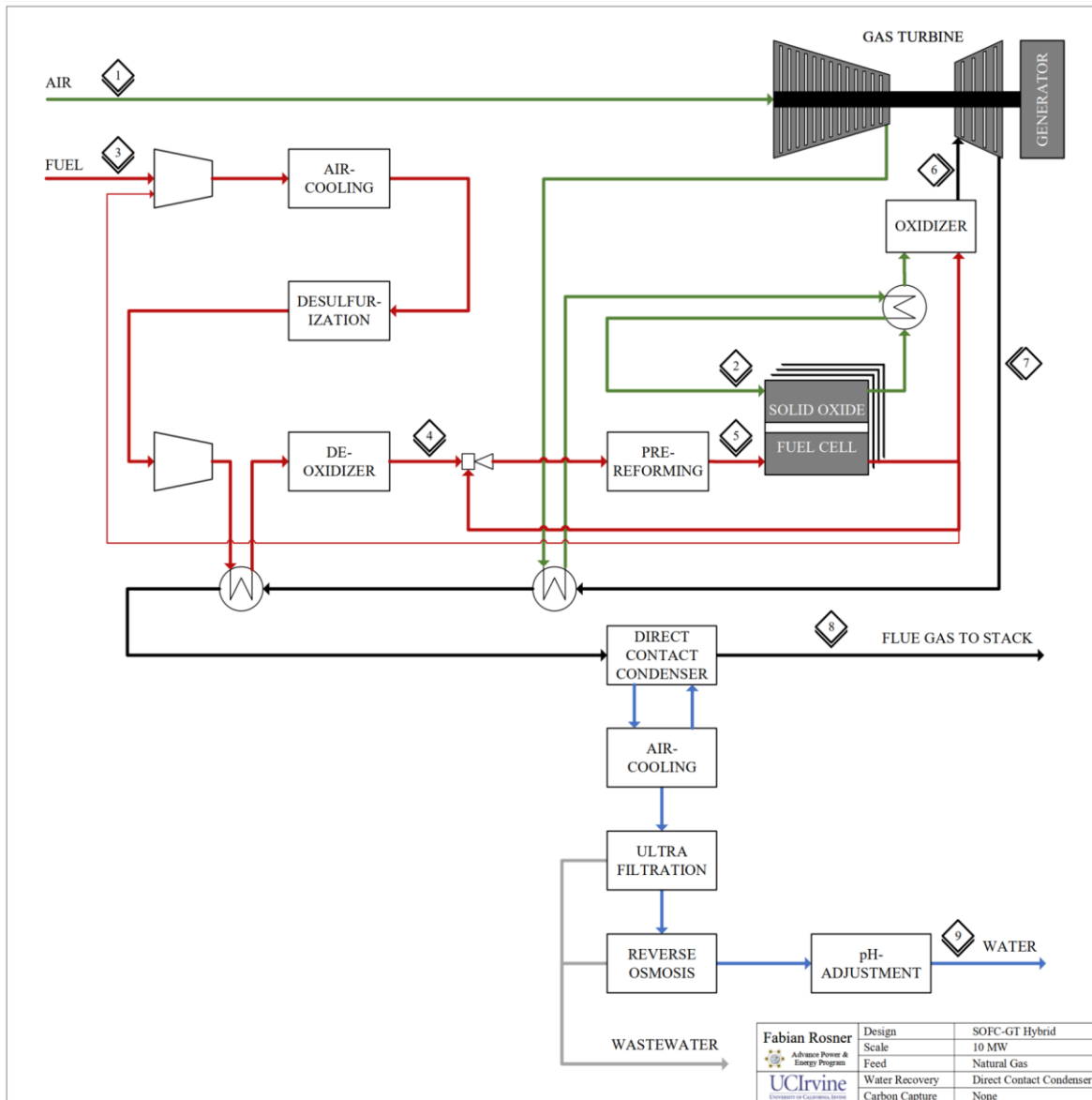


Figure 5-52: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with direct contact condenser for water recovery.

Table 5-16: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	559	46	41
Pressure	bar	1.01	7.19	4.10	21.34	7.19	6.59	1.07	1.01	2.00
Molar Flow Rate	kmol/h	1,075.1	1,075.1	67.3	67.5	349.7	1,144.4	1,144.4	1,104.0	39.1
Mass Flow Rate	kg/h	31,024	31,024	1,167	1,171	7,117	32,191	32,191	31,428	706
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07513	0.07513	0.07788	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72728	0.72728	0.75114	0.00234
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00891	0.00008
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22850	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09017	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01069	0.20401	0.06161	0.06161	0.06378	0.00002
H ₂ O	-	0.00987	0.00987	0.00003	0.00166	0.32711	0.12729	0.12729	0.09829	0.99745
CH ₄	-	0.00000	0.00000	0.93087	0.92859	0.14348	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	0.99989

Thermodynamic Performance. The power output of the nominal 10 MW hybrid plant with direct contact condenser for water recovery is 11,436 kW, whereby, the SOFC produces a net power of 10,658 kW_{DC} and the GT 1,377 kW. Due to the lower backpressure of the GT, associated with the lower pressure drop of the direct contact condenser, the GT produces 174 kW more than in the air-cooled condenser case. The change in back pressure results in a SOFC operating pressure of 7.19 bar and a GT surge margin of 11.1%. The general implications of changes in SOFC operating pressure have been discussed for the air-cooled condenser case and also apply here. The FU in this scenario is 88.2% and essentially constant when compared to the other cases (0.032% points higher than in the reference case). The auxiliary load accounts for 599 kW and the common plant equipment (non-water recovery and water treatment) scales with the plant size. The auxiliary load for the water recovery system is significantly higher than in the air-cooled condenser case due to large amounts of cooling water recirculation needed and the pumps' power consumption associated with that. The water treatment load on the other hand, is much smaller than in the air-cooled condenser case due to the lower water recovery; however, this auxiliary load reduction is very small when compared to the total auxiliary load.

The efficiency of the direct contact condenser case is 74.76%-LHV. This is 0.51% points lower than the reference case. While the GT power output in this case is higher than in the air-cooled condenser case and closer to the GT power output from the reference case, the higher power demand of the direct contact condenser water recovery system offsets some of these benefits. The GT efficiency in this scenario is 59.8%-LHV and the SOFC stack efficiency is 82.0%-LHV.

On an absolute scale the carbon emissions are 3,099 kg/h, which translates to 271 kg/MWh. The water recovery from the flue gas using the direct contact condenser is 25.5% and limited by the ambient air temperature. At the scale of this nominal 10 MW plant, a total of 16,868 kg of water can be recovered per day. The hybrid plant with direct contact condenser water recovery system can produce enough water for the irrigation of over 8,700 m² (including 90% capacity factor, 1.74 l/m² of irrigation water per day [341]). The need for water treatment results in a liquid waste stream of 57 kg/h, which needs to be disposed and lowers the water export. The plant performance data are summarized in **Table 5-17**.

Table 5-17: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Gross Power Generation	Unit	Direct Contact Condenser
Fuel Cell	kW _{DC}	10,657.53
Gas Turbine	kW	1,377.30
Total Generation	kW	12,034.83
Auxiliary Load		
DC-AC Inverter Losses	kW	319.73
Fuel Gas Compression	kW	168.58
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	56.33
Water Treatment	kW	1.04
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.81
Transformer Losses	kW	45.25
Total Auxiliary Load	kW	598.77
Net Power Generation	kW	11,436.07
Plant Performance		
Net Efficiency	%-HHV	67.41
Net Efficiency	%-LHV	74.76
Consumables		
Fuel Gas	kg/h	1,167
Thermal Input	MJ _{LHV} /h	55,053
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,099
Carbon Emission	kg/MWh	271
Liquid Waste	kg/h	57
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	25.5
Flue Gas Water Recovery	kg/d	16,868

Economics. The TPC of the 10 MW hybrid with direct contact condenser for water recovery is \$28,395,790, which corresponds to a specific plant cost of 2,483 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The gas turbine island cost remains constant at \$3,651,010 since the same GT is used in all study cases. Other equipment costs, such as gas processing and flue gas stack, scale with the plant size but

remain relatively constant. Heat recuperation in the direct contact condenser case is shifted towards the air preheater due to the higher GT backpressure, which allows for more heat recovery from the GT exhaust (compared to the reference case). At the same time, the cathode off-gas exchanger cost is reduced. The direct contact condenser water recovery system costs \$1,185,640 and is more than 2.7 times more expensive than the air-cooled condenser. The water treatment equipment cost is \$361,960 and is lower than in the air-cooled condenser case due to the smaller scale (lower water recovery). An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-53**. A more detailed cost summary of all plant equipment is provided in **Table 5-18**.

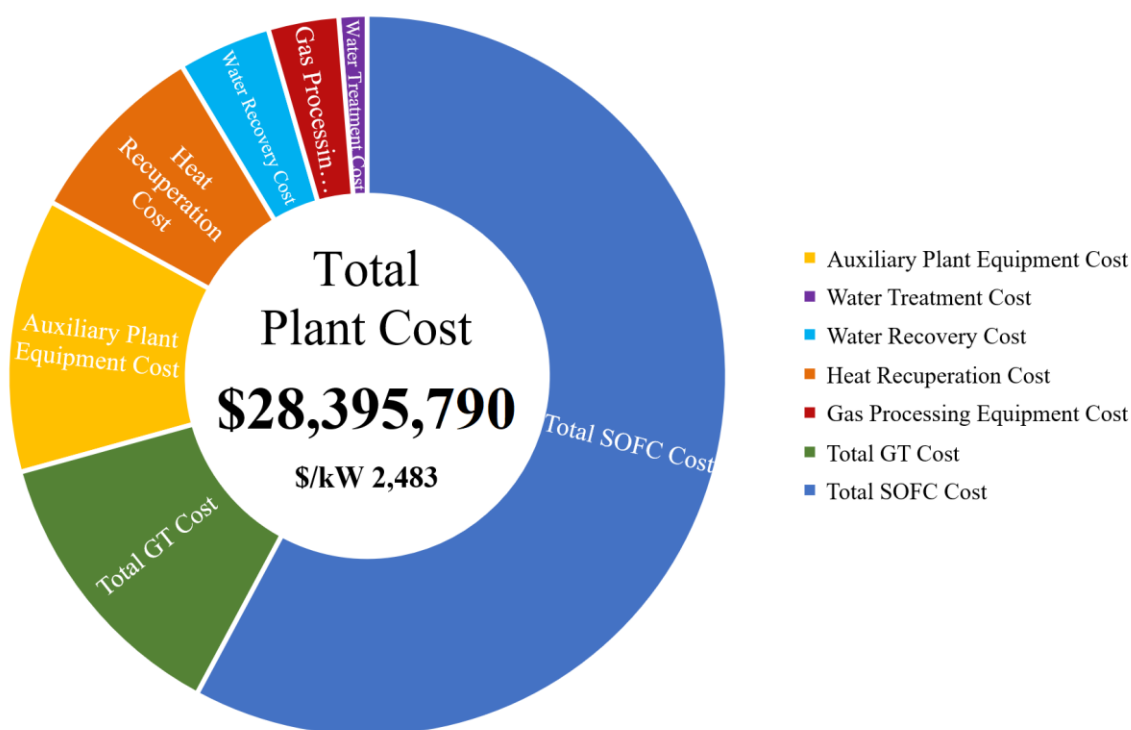


Figure 5-53: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser for water recovery.

The overnight capital cost in the direct contact condenser case is \$35,514,820, an increase of over \$2,142,160 compared to the reference case, and \$1,118,790 compared to the air-cooled condenser case. This increase is mostly due to the TPC increase of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. In the direct contact condenser case, water byproduct sales reduce the variable operating costs by \$563,200 per year.

Table 5-18: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Equipment Cost	Unit	Direct Contact Condenser
Fuel Cell Island	\$	16,428,970
SOFC Stacks	\$	6,331,680
Power Conditioning	\$	6,269,280
Pressure Vessel, Housing & Assembly	\$	3,252,020
System Installation	\$	575,990
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	896,000
Fuel Compressors and Intercooler	\$	450,150
Fuel Processing	\$	315,610
Flue Gas Stack System	\$	130,240
Heat Recuperation	\$	2,389,910
Cathode Off-gas Heat Exchanger	\$	1,010,220
Air Preheater	\$	1,113,450
Fuel Preheater	\$	266,240
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	1,547,610
Water Recovery System	\$	1,185,640
Water Treatment	\$	361,960
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,482,290
Accessory Electric Plant	\$	1,155,270
Instrumentation & Controls	\$	1,042,750
Improvement to Site	\$	271,120
Buildings & Structures	\$	1,013,160
Total Plant Cost	\$	28,395,790
Specific Plant Cost	\$/kW	2,483

The fixed operating cost in the direct contact condenser case increases due to tax and insurance costs of the additional equipment. The operating labor cost in the air-cooled condenser and direct contact condenser are identical; however, when comparing the labor cost on a relative basis, the labor cost in the direct contact condenser case is significantly higher due to the low water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 77.73 \$/MWh. The COW in the direct contact condenser case is 0.1012 \$/L. A summary of the economic key parameters can be found in **Table 5-19**. A detailed breakdown of the COE is shown in **Figure 5-54**.

Table 5-19: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Item	Unit	Direct Contact Condenser
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	28,395,790
Preproduction Costs		
6 Months Labor Cost	\$	984,440
1 Month Maintenance Materials	\$	33,780
1 Month Non-Fuel Consumables	\$	27,490
1 Month Waste Disposal	\$	1,370
25% of 1 month fuel cost at 100% CF	\$	40,050
2% of TPC	\$	567,920
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,330
0.5% of TPC (Spare Parts)	\$	141,980
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	60,840
Land	\$	230,780
Other Owner's Costs (15% of TPC)	\$	4,259,370
Financing Costs	\$	766,690
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,730,020
Other Annual Variable Operating Cost at above CF for Initial Year	\$	676,580
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	563,200
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,968,870
Property Tax & Insurance	\$	567,920
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.1012

Figure 5-54 shows the COE value of 77.73 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-6.25 \$/MWh) and the whole circle represents the COE without byproduct credit (83.97 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 83.97 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.1012 \$/L in order to reach the COE of 77.73 \$/MWh. This COW is higher than in the air-cooled condenser case despite the fact that if the water would be provided at no cost the COE would be lower than in the air-cooled condenser case. It is important to understand that the zero-cost-of-water COE does not account for how much water is recovered, while the COW is taking the amount of water recovered into consideration. Thus, the zero-cost-of-water COE is more representing system integration losses (as well as the capital cost of the water recovery equipment) than the economic value of the recovered water.

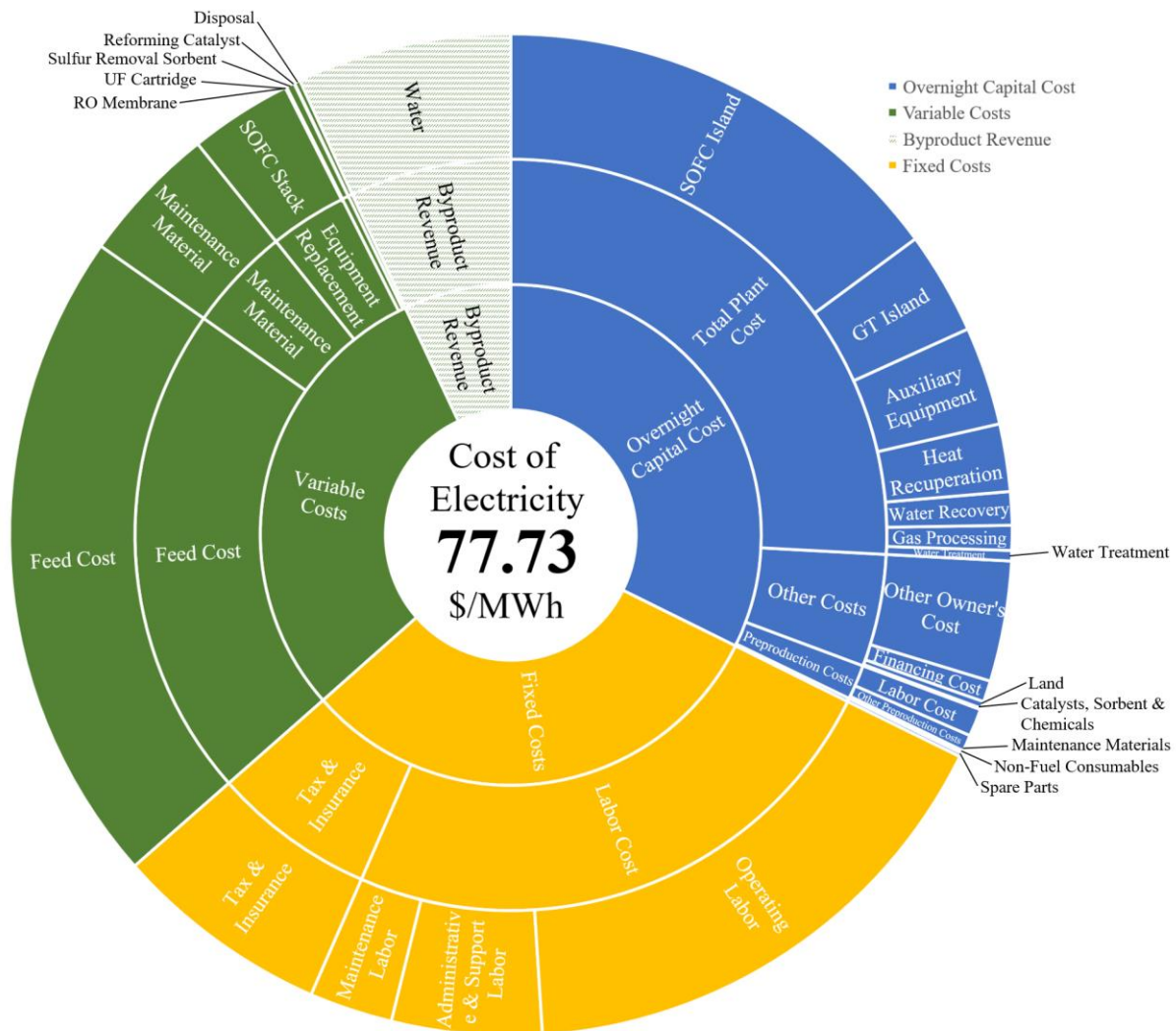


Figure 5-54: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser for water recovery.

This can be also seen in **Figure 5-55**, where the performance penalty cost of the direct contact condenser is significantly lower than in the air-cooled condenser case. While the performance penalty cost in the air-cooled condenser case represents 33.6% of the COW, it is only 13.9% in the direct contact condenser case. The fixed operating costs are still the largest contributor to the COW with over 52.1%, whereby, the operational labor for this analysis is equally split between the water recovery system and the water treatment system. The water recovery alone accounts for 51.7% of the COW (16.9% TPC water recovery). The water treatment costs are again dominated by the fixed operating costs, and the TPC of the water treatment is responsible for 5.2% of the COW. A detailed breakdown of the COW can be found in **Figure 5-55**.

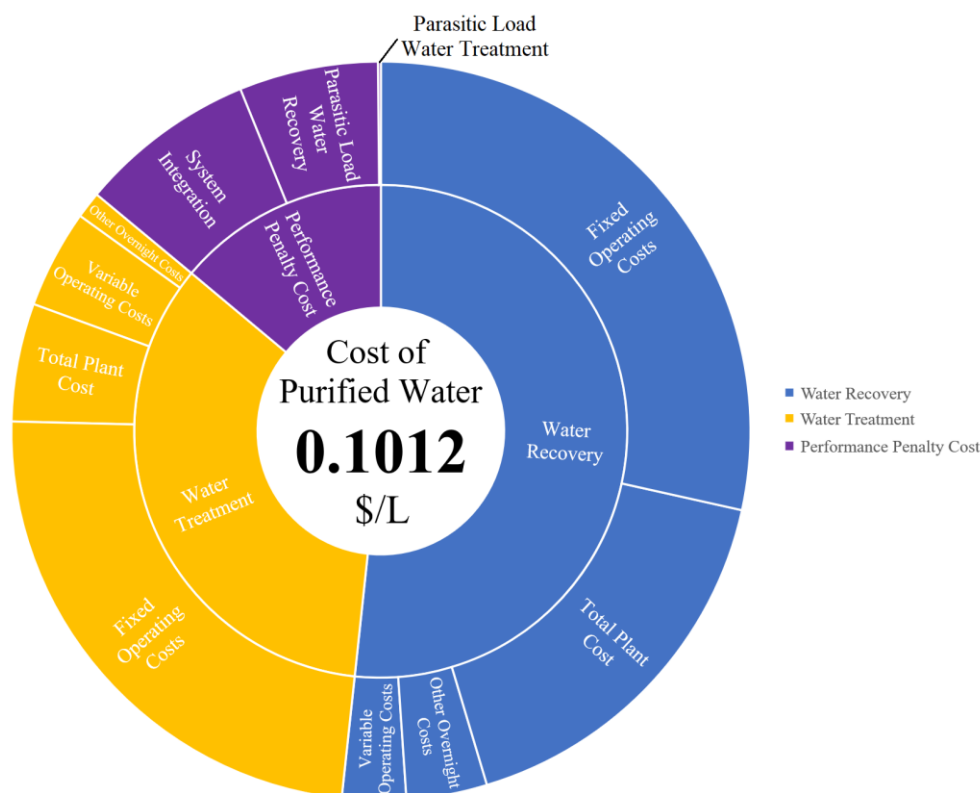


Figure 5-55: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser for water recovery.

Direct Contact Condenser with Zero-Liquid-Discharge System

The design of the direct contact condenser case with zero-liquid-discharge system is almost identical to the direct contact condenser case without zero-liquid-discharge system. Like in the air-cooled condenser case with zero-liquid-discharge system, a brine concentrator and a spray crystallizer are used to eliminate any liquid waste streams that originate from the water treatment. The brine concentrator is a falling film evaporator utilizing a vapor compression cycle as energy source to drive the evaporation. The brine concentrator feed is the rejected water from the UF (backwash and clean-in-place) and the RO unit (retentate). The brine concentrator increases the content of dissolved solids in the rejected water and simultaneously recovers water from this waste stream, which increases the net water production of the power plant. The concentrated wastewater stream is then injected into a spray crystallizer, which uses a small portion of the hot flue gas to provide the heat for the evaporation of the remaining water contained in the waste stream. A general flow diagram of the 10MW NG case with direct contact condenser water recovery and zero-liquid-discharge system is shown in **Figure 5-56**. The respective state-point stream data are provided in **Table 5-20**.

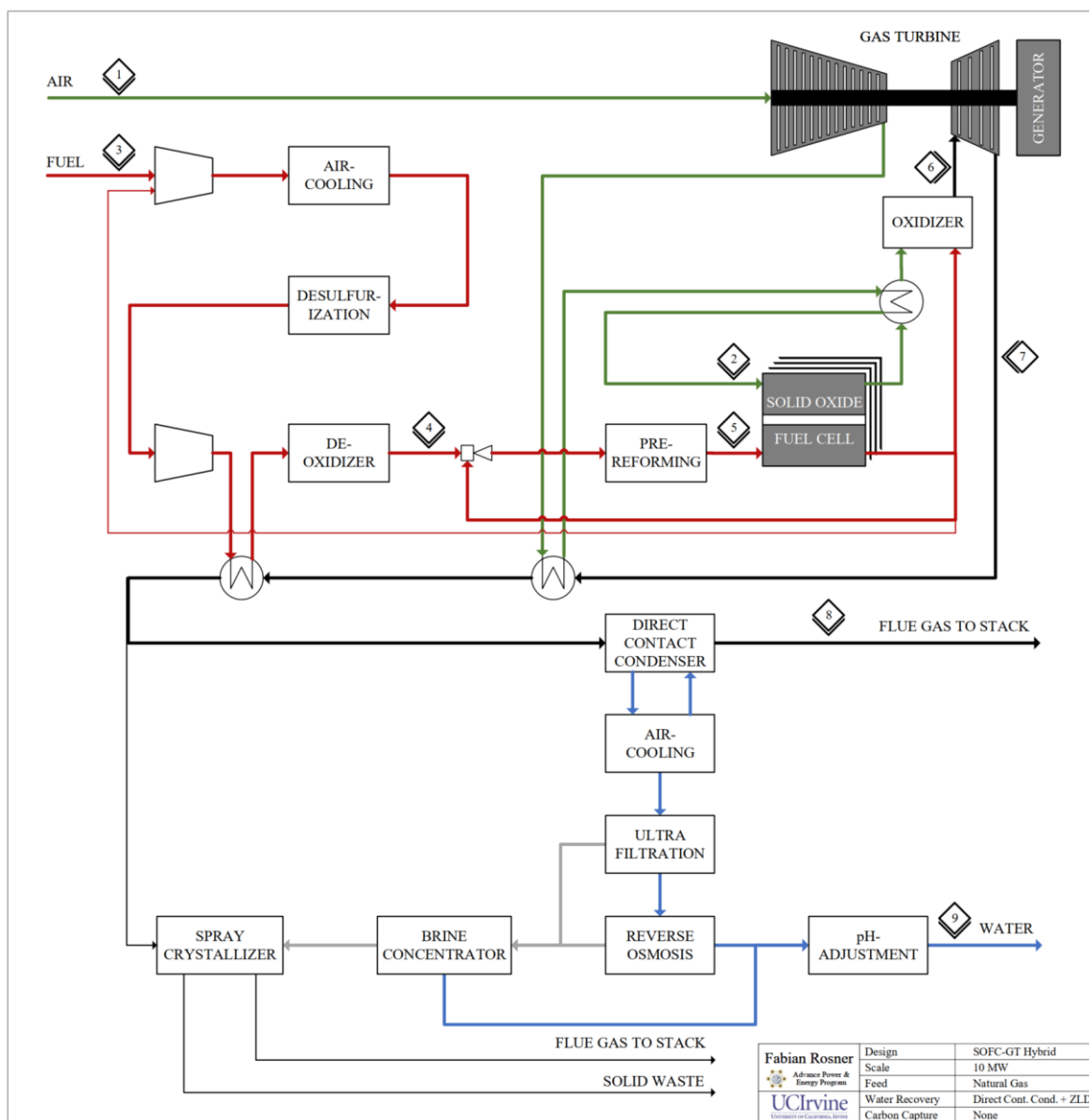


Figure 5-56: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with direct contact condenser and zero-liquid-discharge system for water recovery.

Table 5-20: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser with Zero-Liquid-Discharge System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	559	45	41
Pressure	bar	1.01	7.19	4.10	21.23	7.19	6.59	1.07	1.01	2.00
Molar Flow Rate	kmol/h	1,075.1	1,075.1	67.3	67.5	349.7	1,144.4	1,144.4	1,102.0	42.2
Mass Flow Rate	kg/h	31,024	31,024	1,167	1,171	7,117	32,191	32,191	31,375	761
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07513	0.07513	0.07789	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72728	0.72728	0.75130	0.00218
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00891	0.00007
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22850	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09017	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01069	0.20401	0.06161	0.06161	0.06380	0.00001
H ₂ O	-	0.00987	0.00987	0.00003	0.00166	0.32711	0.12729	0.12729	0.09810	0.99763
CH ₄	-	0.00000	0.00000	0.93087	0.92859	0.14348	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	0.99989

Thermodynamic Performance. The thermodynamic performance of the plant is summarized in **Table 5-21**.

Changes to the plant configuration due to the zero-liquid-discharge system are limited to the water side downstream of the direct contact condenser (and the flue gas slip stream for the spray crystallizer). Thus, there are no changes in the operating conditions, power generating units, as well as the balance-of-plant equipment.

Compared to the direct contact condenser case without zero-liquid-discharge system, the auxiliary load of the water recovery unit is minimally reduced due to the slightly lower flue gas flow to the condenser. On the other hand, the load associated with the water treatment is slightly increased due to the overall higher amount of water exported from the plant. Additionally, the zero-liquid-discharge system requires electricity for moving gas and liquid. As a result, the efficiency of the direct contact condenser case with zero-liquid-discharge system is with 74.75%-LHV minimally lower than the direct contact condenser case without zero-liquid-discharge system.

Emission values of the plant remain virtually constant; however, the liquid waste was reduced to 0 kg/h via the zero-liquid-discharge system by turning it into a solid waste stream of 55 g/h. With a flue gas water recovery of 25.7% and the water recovery from the rejected water from the UF and RO, the plant's water output increased to 18,193 kg/d, an increase of 7.9% versus the direct contact condenser case without zero-liquid-discharge system.

Table 5-21: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser and Zero-Liquid-Discharge System for Water Recovery

Gross Power Generation	Unit	Direct Contact Condenser + ZLD
Fuel Cell	kW _{DC}	10,657.53
Gas Turbine	kW	1,377.30
Total Generation	kW	12,034.83
Auxiliary Load		
DC-AC Inverter Losses	kW	319.73
Fuel Gas Compression	kW	168.15
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	56.32
Water Treatment	kW	1.05
Zero Liquid Discharge System	kW	2.29
Miscellaneous Balance of Plant	kW	7.81
Transformer Losses	kW	45.25
Total Auxiliary Load	kW	600.62
Net Power Generation	kW	11,434.21
Plant Performance		
Net Efficiency	%-HHV	67.40
Net Efficiency	%-LHV	74.75
Consumables		
Fuel Gas	kg/h	1,167
Thermal Input	MJ _{LHV} /h	55,053
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,099
Carbon Emission	kg/MWh	271
Liquid Waste	kg/h	0
Solid Waste	g/h	55
Water Recovery		
Flue Gas Water Recovery	%	25.7
Flue Gas Water Recovery	kg/d	18,193

Economics. Due to the zero-liquid-discharge system, the TPC increases from \$28,395,790 in the direct contact condenser case without zero-liquid-discharge system to \$29,234,240 in the direct contact condenser case with zero-liquid-discharge system. The TPC for the GT island, gas processing and heat recuperation remain constant. The fuel cell island TPC remains essentially constant too, except a minor cost reduction in the power conditioning equipment at the grid interface. The zero-liquid-discharge system alone adds \$770,500 to the TPC, which is included in **Figure 5-57** in the rubric water treatment. The auxiliary plant cost also increases as the need for additional instrumentation, controls, land, buildings and structures increases. A detailed summary of the TPC is provided in **Table 5-22**.

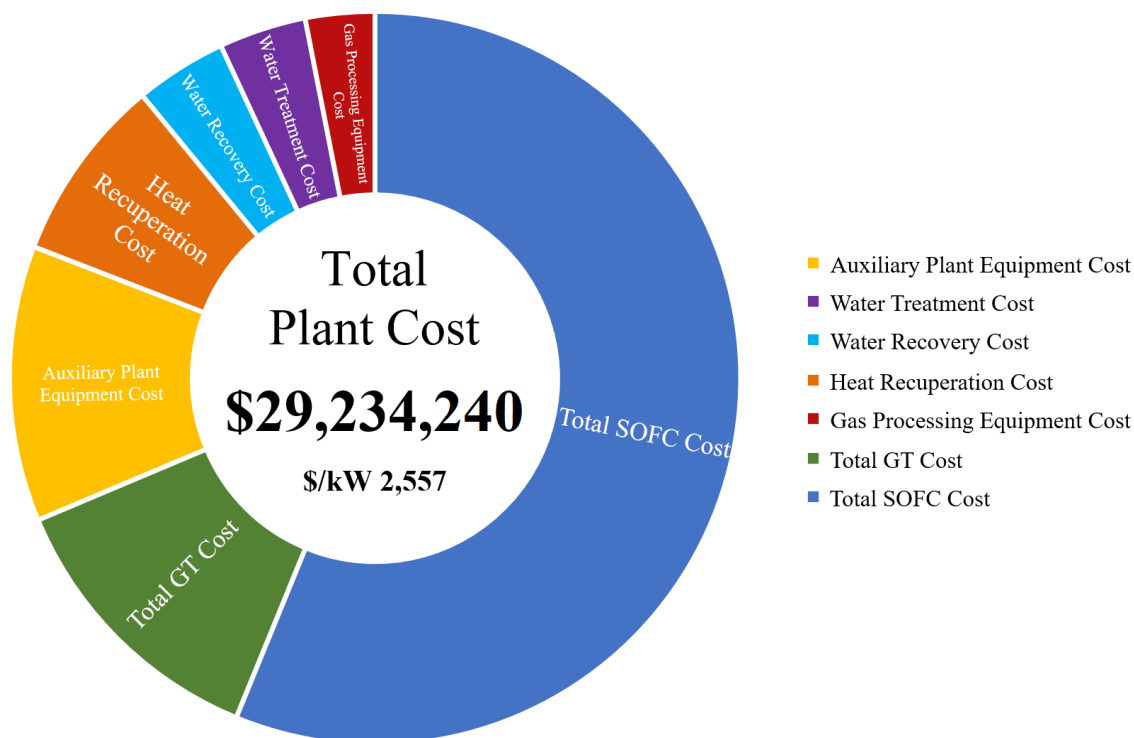


Figure 5-57: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser and zero-liquid-discharge system for water recovery.

The overnight capital cost in the direct contact condenser case with zero-liquid-discharge system is \$36,528,870. The zero-liquid-discharge system leads to an increase in pre-production costs due to the increase in labor, maintenance materials, non-fuel consumables and other TPC related preproduction costs. However, the costs associated with waste disposal could be drastically reduced as the waste stream was reduced from 57 kg/h of wastewater to 55 g/h of solid waste. In order to obtain a COE of 77.73 \$/MWh (same as in reference case) the exported water needs to be sold at a price of 0.1116 \$/L, which leads to a total revenue of \$669,760 per year. Thus, the increased water export from the plant site due to the zero-liquid-discharge system cannot offset the increased costs associated with it. If the water would be provided to the local community at no cost, the COE would increase to 85.15 \$/MWh. A summary of the plant economics is provided in **Table 5-23** and a detailed breakdown of the COE is shown in **Figure 5-58**.

Table 5-22: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser and Zero-Liquid-Discharge System for Water Recovery

Equipment Cost	Unit	Direct Contact Condenser + ZLD
Fuel Cell Island	\$	16,428,930
SOFC Stacks	\$	6,331,680
Power Conditioning	\$	6,269,240
Pressure Vessel, Housing & Assembly	\$	3,252,020
System Installation	\$	575,990
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	895,330
Fuel Compressors and Intercooler	\$	449,480
Fuel Processing	\$	315,610
Flue Gas Stack System	\$	130,240
Heat Recuperation	\$	2,390,290
Cathode Off-gas Heat Exchanger	\$	1,010,220
Air Preheater	\$	1,113,450
Fuel Preheater	\$	266,620
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	2,317,680
Water Recovery System	\$	1,183,820
Water Treatment	\$	363,350
Zero Liquid Discharge System	\$	770,500
Auxiliary Plant Equipment Cost	\$	3,551,000
Accessory Electric Plant	\$	1,155,270
Instrumentation & Controls	\$	1,073,540
Improvement to Site	\$	279,120
Buildings & Structures	\$	1,043,070
Total Plant Cost	\$	29,234,240
Specific Plant Cost	\$/kW	2,557

Figure 5-59 shows the various driving factors of the COW. The zero-liquid-discharge system is responsible for 29.4% of the COW. Like previously mentioned, the operating labor for the water recovery system, water treatment and zero-liquid-discharge systems are now equally split between the three units (same number of operators needed as in the direct contact condenser case without zero-liquid-discharge system). The TPC of the zero-liquid-discharge system accounts for 9.3% of the COW and the variable operating costs of the zero-liquid-discharge system account for 2.0% of the COW. Performance penalty costs remain relatively small with 12.5%. The share of the parasitic load of the zero-liquid-discharge system is small with a value of 0.2%. Water treatment accounts for 20.7% of the COW and the actual water recovery has a share of 37.4% on the COW.

Table 5-23: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with Direct Contact Condenser and Zero-Liquid-Discharge System for Water Recovery

Item	Unit	Direct Contact Condenser + ZLD
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	29,234,240
Preproduction Costs		
6 Months Labor Cost	\$	990,460
1 Month Maintenance Materials	\$	35,270
1 Month Non-Fuel Consumables	\$	27,500
1 Month Waste Disposal	\$	20
25% of 1 month fuel cost at 100% CF	\$	40,050
2% of TPC	\$	584,680
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,330
0.5% of TPC (Spare Parts)	\$	146,170
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	60,840
Land	\$	230,840
Other Owner's Costs (15% of TPC)	\$	4,385,140
Financing Costs	\$	789,320
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,730,020
Other Annual Variable Operating Cost at above CF for Initial Year	\$	678,140
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	669,760
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,980,930
Property Tax & Insurance	\$	584,680
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.1116

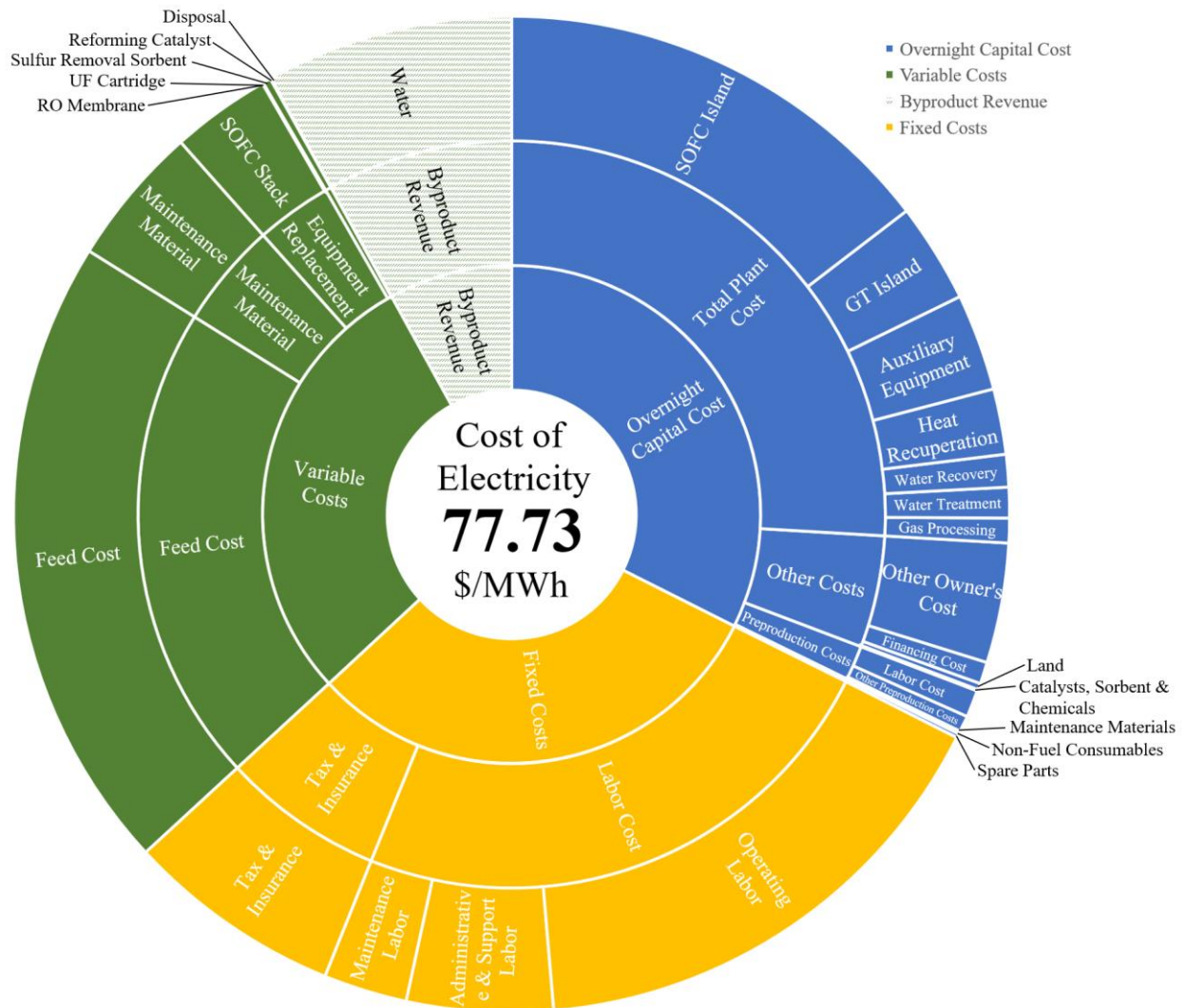


Figure 5-58: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser and zero-liquid-discharge system for water recovery.

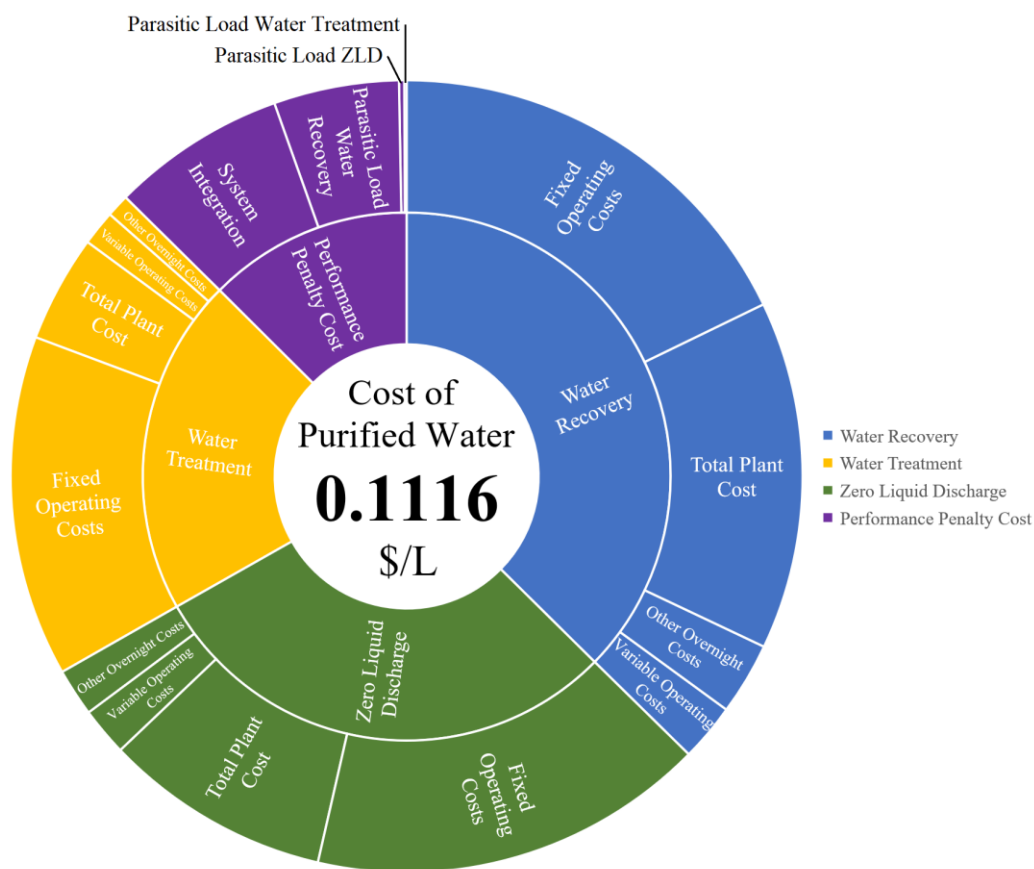


Figure 5-59: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with direct contact condenser and zero-liquid-discharge system for water recovery.

LiBr Absorption System

The configuration of hybrid plant with LiBr absorption system follows the basic configuration of the reference case and fuel processing, fuel cell island, GT island and heat integration are identical to the reference case. In the LiBr absorber case, a flue gas heat exchanger and a packed-bed column are installed downstream of the fuel pre-heater. The packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream heat exchanger is used to provide the heat for sorbent regeneration, while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, an air-cooled condenser, supported by a vacuum compressor to remove any non-condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high

quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is similar to the direct contact condenser case and a similar plume rise can be expected. However, the difference between the flue gas from the LiBr absorber case and the air-cooled condenser, direct contact condenser cases is that the flue gas in the LiBr condenser case is leaving the stack above its dew point temperature minimizing the risk of plume formation (flue gas dew point temperature around 14 °C). A general flow diagram of the 10MW NG case with LiBr absorption water recovery is shown in **Figure 5-60**. The respective state-point stream data are provided in **Table 5-24**.

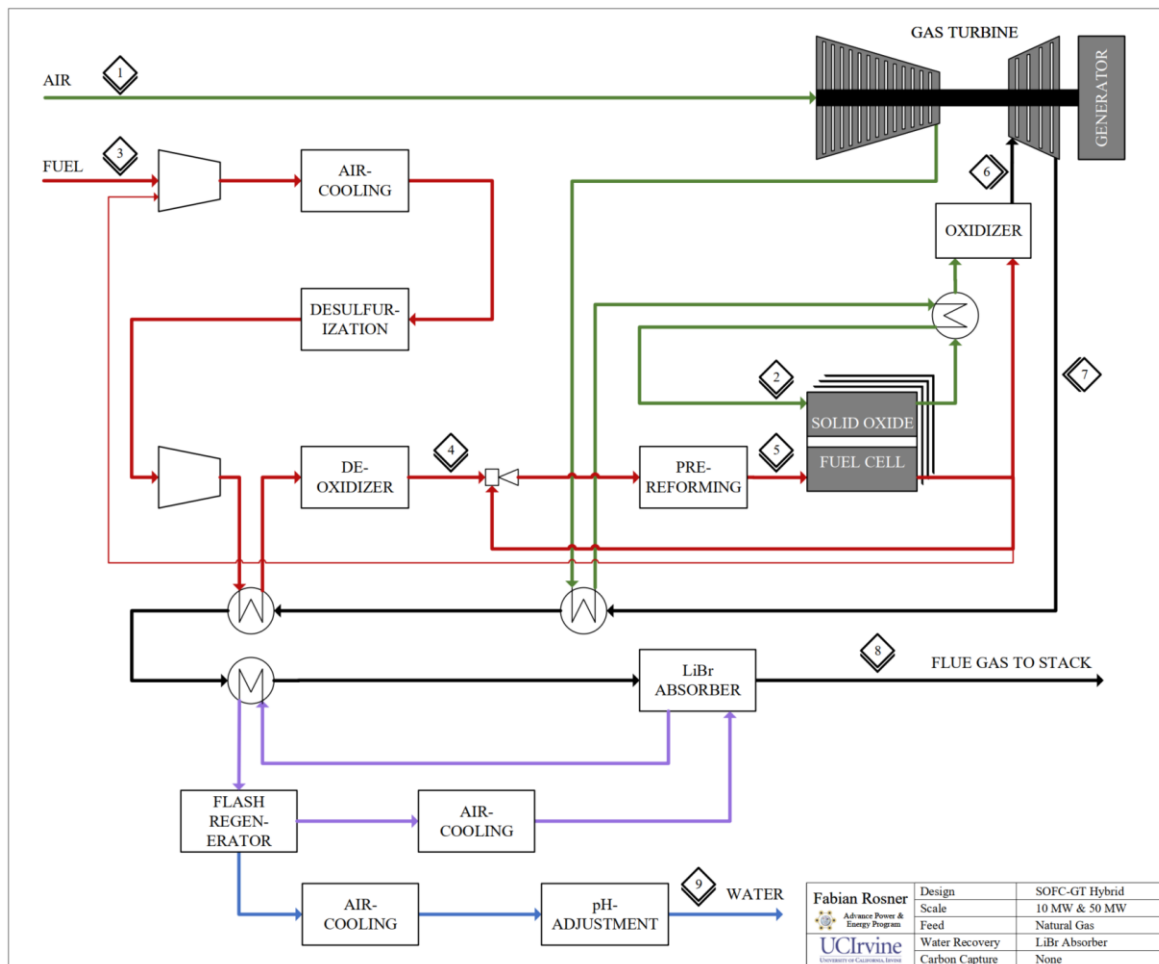


Figure 5-60: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with LiBr absorption system for water recovery.

Table 5-24: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	562	49	40
Pressure	bar	1.01	7.21	4.10	21.33	7.21	6.60	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,073.9	1,073.9	67.0	67.2	348.9	1,142.9	1,142.9	1,013.7	129.3
Mass Flow Rate	kg/h	30,991	30,991	1,161	1,165	7,096	32,152	32,152	29,824	2,329
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07559	0.07559	0.08523	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72744	0.72744	0.82018	Trace
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00979	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22898	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09042	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01068	0.20386	0.06140	0.06140	0.06923	Trace
H ₂ O	-	0.00987	0.00987	0.00003	0.00164	0.32685	0.12689	0.12689	0.01557	1.00000
CH ₄	-	0.00000	0.00000	0.93087	0.92862	0.14317	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	1

Thermodynamic Performance. The net power output of the nominal 10MW NG plant with LiBr absorber for water recovery is 11,312 kW, which is between the air-cooled condenser case and the direct contact condenser case. The fuel cell island produces 10,593 kW_{DC} and the GT generates an additional 1,344 kW. Despite the LiBr absorber being a packed bed like the direct contact condenser, the GT backpressure in the LiBr absorber case is slightly higher than in the direct contact condenser case due to the LiBr heater. This explains why the GT power output, as well as the total plant scale, ranks between the air-cooled condenser case and the direct contact condenser case. The GT surge margin in the LiBr absorber case is 10.8% and the SOFC operates at an operating pressure of 7.21 bar at a FU of 88.1%.

In a SOFC-GT hybrid the SOFC and GT are strongly coupled, and the respective operating points impact one another. First, the GT compressor discharge pressure causes a change in the SOFC's cooling demand, which then impacts the maximum permissible TIT of the GT, which again impacts the pressure ratio. Due to the higher cooling demand, the fuel to air ratio changes and the maximum permissible TIT increases from 909.3 °C (reference case) to 909.8 °C. However, the TIT increase is too little to offset the losses associated with the increased back pressure. The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 625 kW and represent about 13.9% of the auxiliary load.

The electrical efficiency of the LiBr absorption case decreased by 0.97% points to 74.30%-LHV compared to the reference case and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GT due to the

backpressure. The GT efficiency decreased to 58.0%-LHV compared to the reference case and the stack efficiency remains almost unchanged at 82.1%-LHV.

The plant with LiBr absorber emits 3,090 kg/h of CO₂, which translates to 273 kg/MWh. The water recovery from the flue gas in the LiBr absorber case is 89.1%. Sorbent-based water recovery systems are independent of the flue gas's dew point temperature and can remove substantially higher quantities of water. Simultaneously, they are less dependent upon ambient temperatures. While the ambient temperature still plays a part in the water recovery by setting the condenser backpressure (pressure in the regenerator), this effect can be moderated by the heat input required for sorbent regeneration. At the scale of this nominal 10 MW plant, a total of 55,909 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 28,900 m² of farmland (including 90% capacity factor). The high water quality of the recovered water minimizes water treatment and eliminates any wastewater streams from the plant, which reduces waste disposal costs but also maximizes water export. The plant performance data are summarized in **Table 5-25**.

Table 5-25: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	10,593.34
Gas Turbine	kW	1,343.72
Total Generation	kW	11,937.06
Auxiliary Load		
DC-AC Inverter Losses	kW	317.80
Fuel Gas Compression	kW	167.70
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	86.12
Water Treatment	kW	0.71
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.75
Transformer Losses	kW	44.89
Total Auxiliary Load	kW	624.98
Net Power Generation	kW	11,312.08
Plant Performance		
Net Efficiency	%-HHV	67.00
Net Efficiency	%-LHV	74.30
Consumables		
Fuel Gas	kg/h	1,161
Thermal Input	MJ _{LHV} /h	54,812
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,090
Carbon Emission	kg/MWh	273
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	89.1
Flue Gas Water Recovery	kg/d	55,909

Economics. The TPC of the 10 MW hybrid with LiBr absorber for water recovery is \$29,070,800, resulting in a specific plant cost of 2,570 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The gas turbine island cost remains constant at \$3,651,010, since the same GT is used in all study cases. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. Heat recuperation in the LiBr absorber case is shifted towards the air preheater due to the higher GT backpressure, which allows for more heat recovery from the GT exhaust (compared to the reference case). At the same time, the cathode off-gas exchanger cost is reduced. The LiBr absorption water recovery system costs \$2,093,160. The water treatment equipment cost is \$122,830 and substantially reduced compared to other water recovery cases despite the fact that more water is recovered in this case. The high water quality obtained in the LiBr absorption case reduces the water processing equipment and thus, the water treatment cost. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-61**. A more detailed cost summary of all plant equipment is provided in **Table 5-26**.

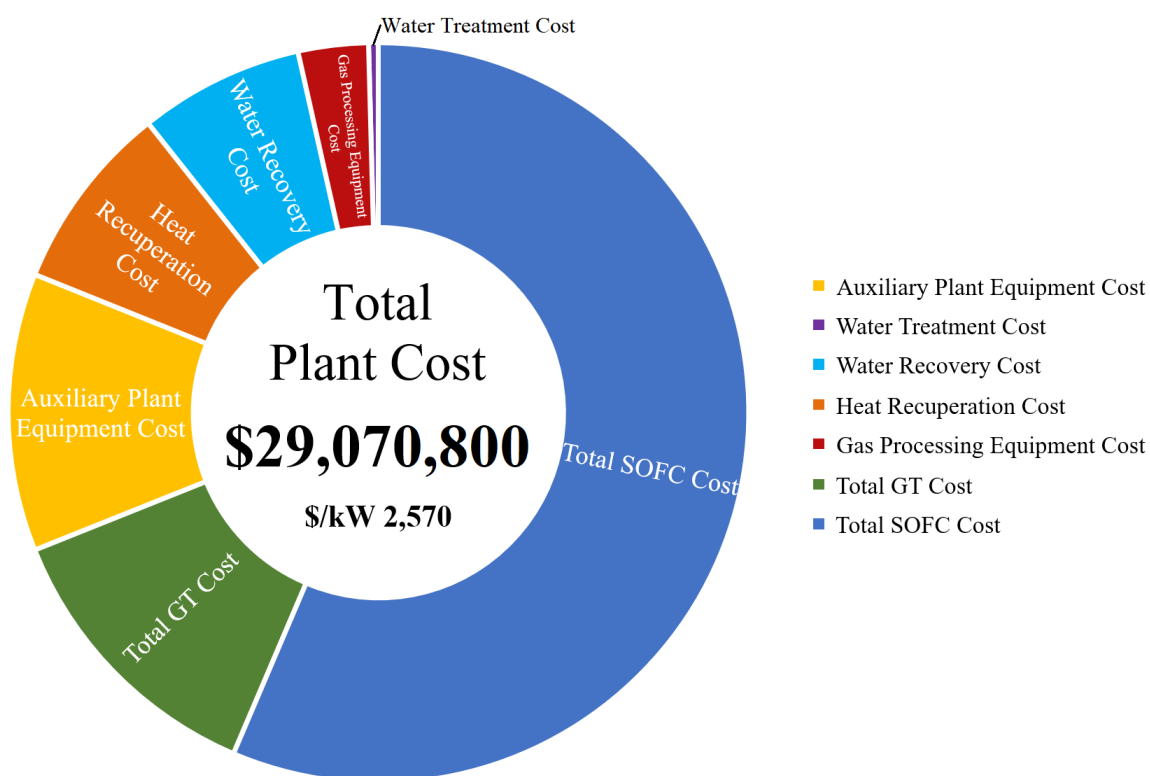


Figure 5-61: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with LiBr absorption system for water recovery.

Table 5-26: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	16,396,640
SOFC Stacks	\$	6,332,400
Power Conditioning	\$	6,230,330
Pressure Vessel, Housing & Assembly	\$	3,258,720
System Installation	\$	575,190
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	892,190
Fuel Compressors and Intercooler	\$	447,220
Fuel Processing	\$	314,830
Flue Gas Stack System	\$	130,140
Heat Recuperation	\$	2,386,740
Cathode Off-gas Heat Exchanger	\$	1,008,770
Air Preheater	\$	1,112,500
Fuel Preheater	\$	265,470
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	2,215,990
Water Recovery System	\$	2,093,160
Water Treatment	\$	122,830
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,528,220
Accessory Electric Plant	\$	1,145,880
Instrumentation & Controls	\$	1,067,540
Improvement to Site	\$	277,560
Buildings & Structures	\$	1,037,240
Total Plant Cost	\$	29,070,800
Specific Plant Cost	\$/kW	2,570

The overnight capital cost in the LiBr absorber case is \$36,766,490, an increase of over 10.2% compared to the reference case. This increase is mostly related to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. While the air-cooled condenser case and the direct contact condenser case do not require any chemicals for water recovery, the LiBr absorber system requires an initial fill and continuous LiBr makeup. The 55 wt-% LiBr solution is assumed to cost 2.04 \$₂₀₁₅/kg and the initial fill costs about \$422,040. The LiBr loss is very small and entirely due to entrainment since LiBr is a salt with extremely low vapor pressure. With a mist eliminator, the LiBr losses are expected to be around 10 kg per year. In the LiBr absorber case, water byproduct sales reduce the variable operating costs by \$742,320 per year.

The fixed operating cost in the LiBr absorber case increases due to tax and insurance costs of the additional equipment. The operating labor costs in the LiBr absorber case are identical to the air-cooled

condenser and direct contact condenser cases; however, when comparing the labor cost on a relative basis, the labor cost in the LiBr absorber case is significantly lower due to the increased water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 77.73 \$/MWh. The COW in the LiBr absorber case is 0.0404 \$/L. A summary of the economic key parameters can be found in **Table 5-27**. A detailed breakdown of the COE is shown in **Figure 5-62**.

Table 5-27: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	29,070,800
Preproduction Costs		
6 Months Labor Cost	\$	989,990
1 Month Maintenance Materials	\$	35,150
1 Month Non-Fuel Consumables	\$	26,940
1 Month Waste Disposal	\$	20
25% of 1 month fuel cost at 100% CF	\$	39,860
2% of TPC	\$	581,420
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,260
0.5% of TPC (Spare Parts)	\$	145,350
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	482,690
Land	\$	244,500
Other Owner's Costs (15% of TPC)	\$	4,360,620
Financing Costs	\$	784,910
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,721,830
Other Annual Variable Operating Cost at above CF for Initial Year	\$	670,760
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	742,320
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,979,980
Property Tax & Insurance	\$	581,420
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0404

Figure 5-62 shows the COE value of 77.73 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-8.32 \$/MWh) and the whole circle represents the COE without byproduct credit (86.05 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 86.05 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0404 \$/L in order to reach the COE of 77.73 \$/MWh. This COW is substantially lower than in the air-cooled condenser case and direct contact condenser case despite the fact that the capital cost in the LiBr absorber case is higher.

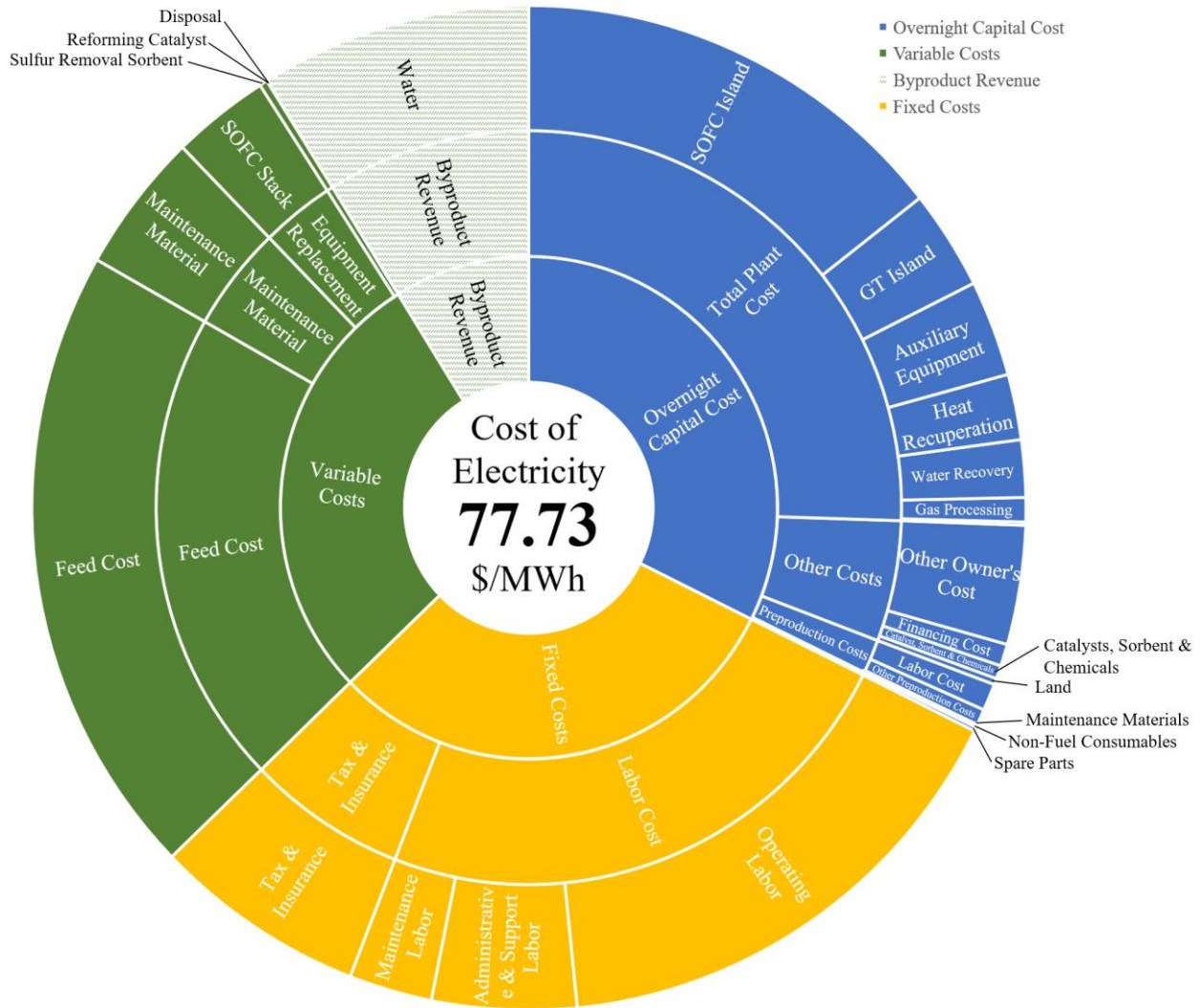


Figure 5-62: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with LiBr absorber for water recovery.

Figure 5-63 shows a detailed breakdown of the COW. For the LiBr absorber case the total operating labor costs are shown under the water recovery rubric, which is different from the air-cooled condenser case and the direct contact condenser case where the operating labor is split between the water recovery and the water treatment. However, the water treatment in the LiBr absorber case is minor and does not justify the addition of operating labor to its costs. The performance penalty cost in the LiBr absorber case is 19.0% and a bit higher than in the direct contact condenser case. System integration losses account for 11.9% and the parasitic load of the water recovery (mostly desiccant and water pumping) accounts for 7.1% of the COW. The fixed operating costs are still the largest contributor to the COW with 43.0%. Overall, fixed operating costs are reduced from 45.3% in the air-cooled condenser case and 52.1% in the direct

contact condenser case, mainly due to the large amount of water recovered. The water treatment is responsible for only 2.6% of the COW.

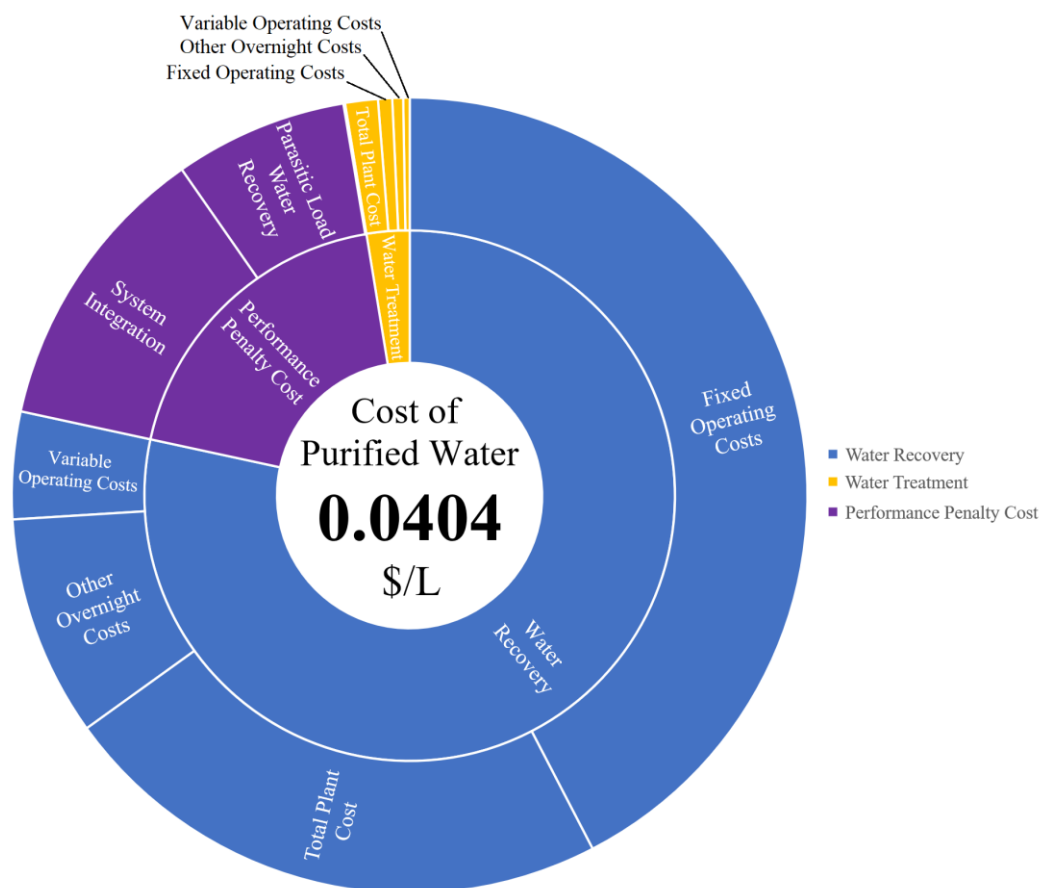


Figure 5-63: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with LiBr absorber for water recovery.

Glycol Absorption System

The plant design of the SOFC-GT hybrid plant with MEG absorption system is very similar to the LiBr absorption case. MEG has a decomposition temperature around 240 °C [342] and the hot flue gas is cooled to this temperature prior to the MEG heater via water spray cooling to avoid MEG degradation. Also in the MEG absorber case, a packed-bed column is used to bring the moist flue gas in contact with the liquid MEG solution. Despite the fact that a membrane is used for desiccant regeneration, the separation principle of this membrane is pervaporation and requires heat input. Thus, the MEG heater is required to drive the water-MEG separation and is not only installed to cool the flue gas to acceptable column inlet temperatures. As MEG is a volatile liquid itself, the membrane increases the selectivity and reduces MEG loss during regeneration. The pressure on the permeate side is set by the vapor condensation temperature, again an air-cooled condenser is used for cooling/condensing the permeate vapor. The air-cooled condenser is further

supported by a vacuum compressor to remove any non-condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires extensive treatment via ultra-filtration, reverse osmosis and pH adjustment due to MEG contamination of the recovered water. A portion of the recovered water is used in the mist eliminator at the column outlet to reduce fouling. With ambient air as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 45 °C. This flue gas stack temperature is similar to the LiBr absorption case and a similar plume rise can be expected. Also in the MEG absorption case, the flue gas is leaving the stack above its dew point temperature, minimizing the risk of plume formation; however, the dew point temperature of the flue gas is above ambient temperature (flue gas dew

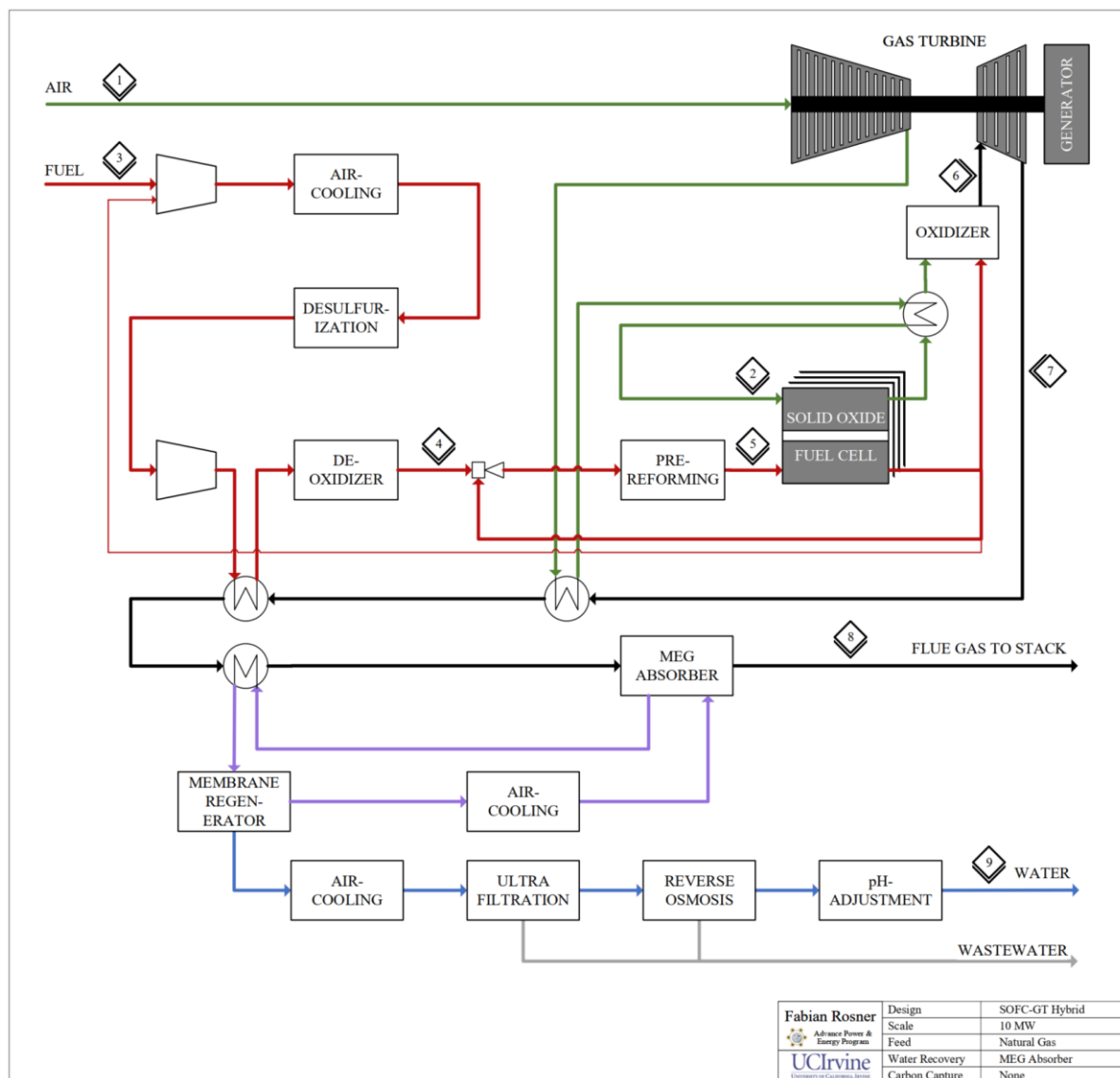


Figure 5-64: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with MEG absorption system for water recovery.

point temperature around 24 °C). A general flow diagram of the 10MW NG case with MEG absorption water recovery system is shown in **Figure 5-64**. The respective state-point stream data are provided in **Table 5-28**.

Table 5-28: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with MEG Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	561	45	41
Pressure	bar	1.01	7.20	4.10	20.55	7.20	6.60	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,074.3	1,074.3	67.2	67.4	348.9	1,143.4	1,143.4	1,026.0	110.4
Mass Flow Rate	kg/h	31,000	31,000	1,165	1,169	7,101	32,164	32,164	30,024	1,996
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07527	0.07527	0.08369	0.00000
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72733	0.72733	0.80979	0.00000
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00965	0.00000
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22828	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09009	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01070	0.20407	0.06155	0.06155	0.06739	0.00000
H ₂ O	-	0.00987	0.00987	0.00003	0.00166	0.32727	0.12717	0.12717	0.02908	0.99834
CH ₄	-	0.00000	0.00000	0.93087	0.92858	0.14356	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	0.99960	0.99834

Thermodynamic Performance. The net power output of the nominal 10MW NG plant with MEG absorber for water recovery is 11,389 kW. The fuel cell island produces 10,643 kW_{DC} and the GT generates an additional 1,352 kW. While the absorber column height is kept constant between the MEG and LiBr cases, the lower density of MEG slightly reduces the pressure drop in the absorber and mist eliminator, lowering the GT backpressure. This explains why the GT power output as well as the total plant scale is slightly higher than in the LiBr case. The GT surge margin in the MEG absorber case is 10.9% and the SOFC operates at an operating pressure of 7.20 bar at a FU of 88.3%.

In a SOFC-GT hybrid the SOFC and GT are strongly coupled, and the respective operating points impact one another. In the MEG absorber case, the maximum permissible TIT of 909.7 °C is reached. The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 606 kW, slightly less than in the LiBr absorber case.

The electrical efficiency of the MEG absorption case is 74.59%-LHV and improved versus the LiBr absorption system, mostly due to the reduced back pressure and a lower pumping load. The GT efficiency is 59.0%-LHV and the stack efficiency remains almost unchanged at 82.0%-LHV.

The plant with MEG absorber emits 3,098 kg/h of CO₂ corresponding to 272 kg/MWh. The water recovery from the flue gas in the MEG absorber case is 78.8%. At the scale of this nominal 10 MW plant,

a total of 47,652 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. uses 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW SOFC-GT hybrid plant with MEG water recovery system can produce enough water for the irrigation of over 24,600 m² of farmland (including 90% capacity factor). The low water quality requires more extensive water treatment than in the LiBr absorber case and generates about 277 kg/h of wastewater. Even after treatment some MEG remains in the product water; however, MEG is biodegradable, and the water quality should be sufficient for irrigation purposes. The plant performance data are summarized in **Table 5-29**.

Table 5-29: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with MEG Absorption System for Water Recovery

Gross Power Generation	Unit	MEG Absorber
Fuel Cell	kW _{DC}	10,642.70
Gas Turbine	kW	1,352.30
Total Generation	kW	11,995.00
Auxiliary Load		
DC-AC Inverter Losses	kW	319.28
Fuel Gas Compression	kW	165.17
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	65.65
Water Treatment	kW	3.08
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.78
Transformer Losses	kW	45.11
Total Auxiliary Load	kW	606.09
Net Power Generation	kW	11,388.91
Plant Performance		
Net Efficiency	%-HHV	67.26
Net Efficiency	%-LHV	74.59
Consumables		
Fuel Gas	kg/h	1,165
Thermal Input	MJ _{LHV} /h	54,956
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,098
Carbon Emission	kg/MWh	272
Liquid Waste	kg/h	277
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	78.8
Flue Gas Water Recovery	kg/d	47,652

Economics. The TPC of the 10 MW hybrid with MEG absorber for water recovery is \$31,003,090, resulting in a specific plant cost of 2,722 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The gas turbine island cost remains constant at \$3,651,010 since the same GT, a Dresser Rand KG2-3G EF, is used in this case. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. Heat recuperation in the MEG absorber case is \$2,390,020. The MEG absorption

water recovery system costs \$3,324,920 and is substantially more expensive than the LiBr water recovery system with \$2,093,160. While the absorber column, heat exchangers and pumping equipment is very similar, the MEG membrane regenerator alone accounts for \$1,109,720. Also, the water treatment equipment cost is more expensive compared to the LiBr case due to the ultra-filtration and reverse osmosis units, and accounts for \$631,860 in this case. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-65**. A more detailed cost summary of all plant equipment is provided in **Table 5-30**.

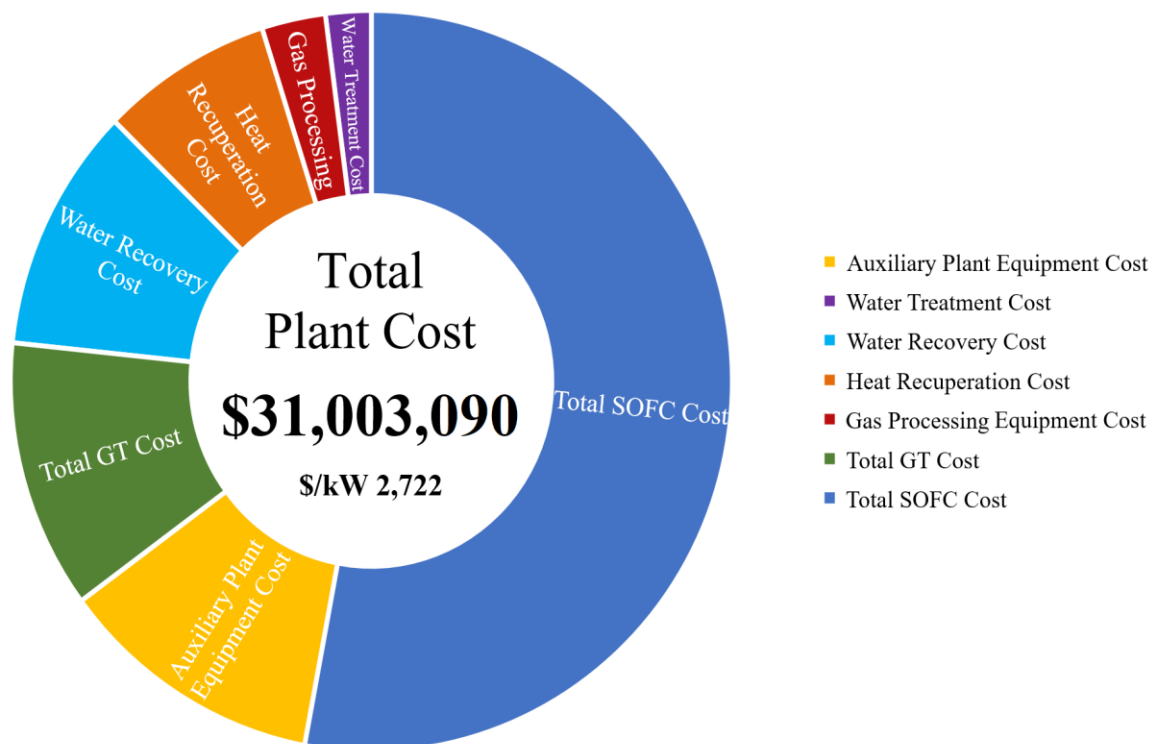


Figure 5-65: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with MEG absorption system for water recovery.

The overnight capital cost in the MEG absorber case is \$39,052,780, an increase of over 17.0% compared to the reference case. This increase is mostly related to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery (compared to reference case). Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. While the air-cooled condenser case and the direct contact condenser case do not require any chemicals for water recovery, the MEG absorber system requires an initial fill and continuous MEG makeup. The 100 wt-%

Table 5-30: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with MEG Absorption System for Water Recovery

Equipment Cost	Unit	MEG Absorber
Fuel Cell Island	\$	16,423,500
SOFC Stacks	\$	6,332,170
Power Conditioning	\$	6,259,880
Pressure Vessel, Housing & Assembly	\$	3,255,590
System Installation	\$	575,850
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	889,640
Fuel Compressors and Intercooler	\$	444,170
Fuel Processing	\$	315,300
Flue Gas Stack System	\$	130,170
Heat Recuperation	\$	2,390,020
Cathode Off-gas Heat Exchanger	\$	1,005,940
Air Preheater	\$	1,115,390
Fuel Preheater	\$	268,690
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	3,956,780
Water Recovery System	\$	3,324,920
Water Treatment	\$	631,860
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,692,140
Accessory Electric Plant	\$	1,151,450
Instrumentation & Controls	\$	1,138,500
Improvement to Site	\$	296,010
Buildings & Structures	\$	1,106,190
Total Plant Cost	\$	31,003,090
Specific Plant Cost	\$/kW	2,722

MEG solution is assumed to cost 0.65 \$₂₀₁₆/kg and the initial fill costs about \$115,270. The MEG loss by entrainment in the absorber is small; however, the losses in the regeneration process, associated with MEG's vapor pressure, are significant. The annual MEG makeup costs amount to \$877,640, which need to be offset by the byproduct's revenue, \$1,944,990 per year.

The fixed operating cost in the MEG absorber case increases due to tax and insurance costs of the additional equipment. The operating labor costs in the MEG absorber case are identical to the air-cooled condenser and direct contact condenser cases and split equally between the water recovery unit and the water treatment unit. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 77.73 \$/MWh. The COW in the MEG absorber case is 0.1235 \$/L. A summary of the economic key parameters can be found in **Table 5-31**. A detailed breakdown of the COE is shown in **Figure 5-66**.

Table 5-31: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with MEG Absorption System for Water Recovery

Item	Unit	MEG Absorber
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	31,003,090
Preproduction Costs		
6 Months Labor Cost	\$	1,005,210
1 Month Maintenance Materials	\$	38,920
1 Month Non-Fuel Consumables	\$	110,100
1 Month Waste Disposal	\$	6,280
25% of 1 month fuel cost at 100% CF	\$	39,970
2% of TPC	\$	620,060
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	164,590
0.5% of TPC (Spare Parts)	\$	155,020
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	175,990
Land	\$	246,000
Other Owner's Costs (15% of TPC)	\$	4,650,460
Financing Costs	\$	837,080
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,726,760
Other Annual Variable Operating Cost at above CF for Initial Year	\$	1,677,310
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	1,944,990
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	2,010,410
Property Tax & Insurance	\$	620,060
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.1235

Figure 5-66 shows the COE value of 77.73 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-21.66 \$/MWh) and the whole circle represents the COE without byproduct credit (99.39 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 99.39 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.1235 \$/L in order to reach the COE of 77.73 \$/MWh. This COW is substantially higher than in the air-cooled condenser case, direct contact condenser case and LiBr absorber case due to the higher capital cost and substantial cost of the MEG makeup.

Figure 5-67 shows a detailed breakdown of the COW. In the MEG absorption case, the water recovery is the dominating cost driving factor accounting for over 78.5% of the COW. The variable operating costs of the water recovery unit, which includes the MEG makeup, are the largest category of the COW with 48.1%. Water treatment has a share of 15.5% and the performance penalty cost is around 5.9%. The fixed

operating costs, which have dominated the COW for the previously investigated technologies, represent a share of only 19.8% due to the fact that the variable operating costs of the water recovery unit are so dominant.

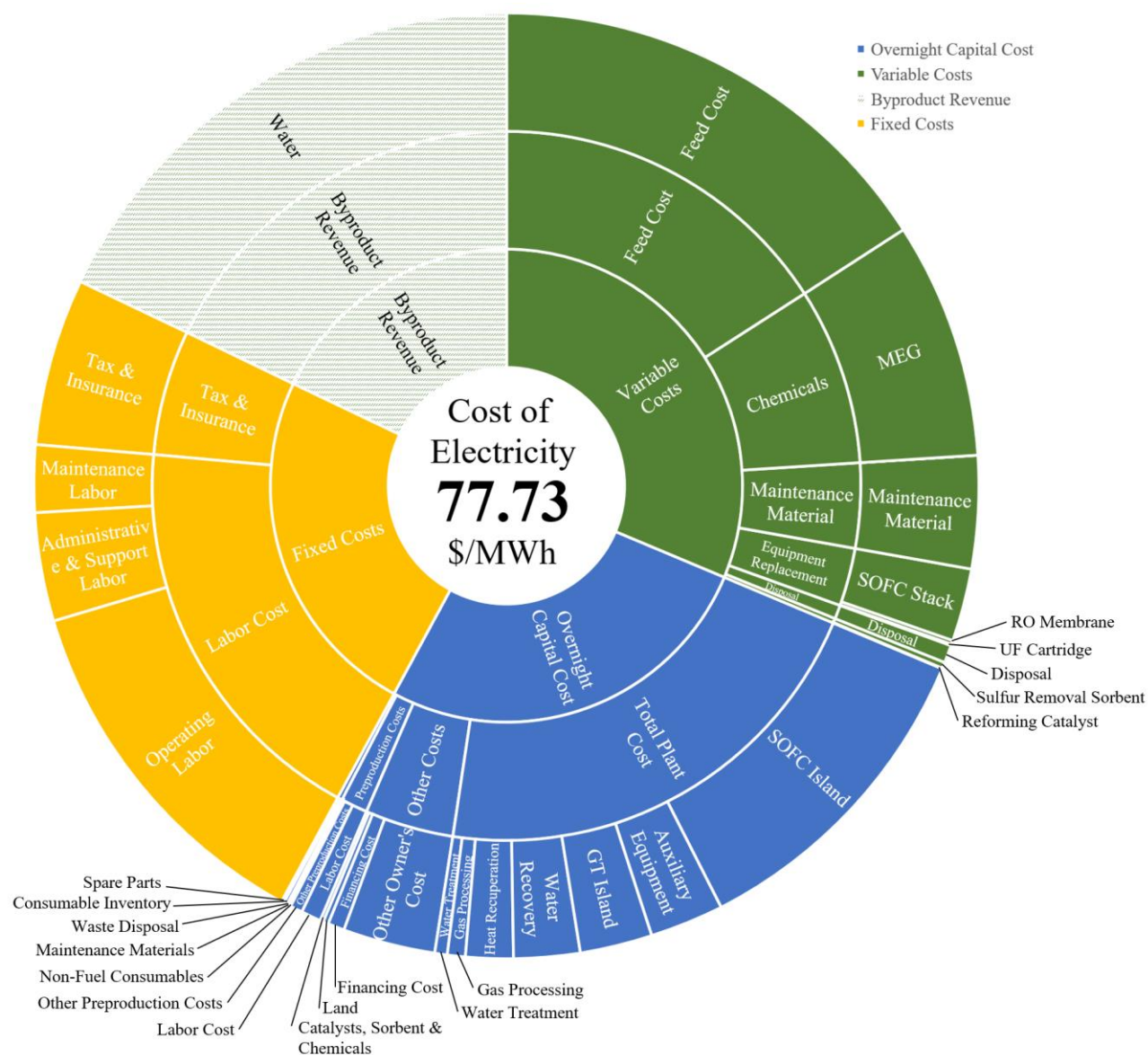


Figure 5-66: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with MEG absorber for water recovery.

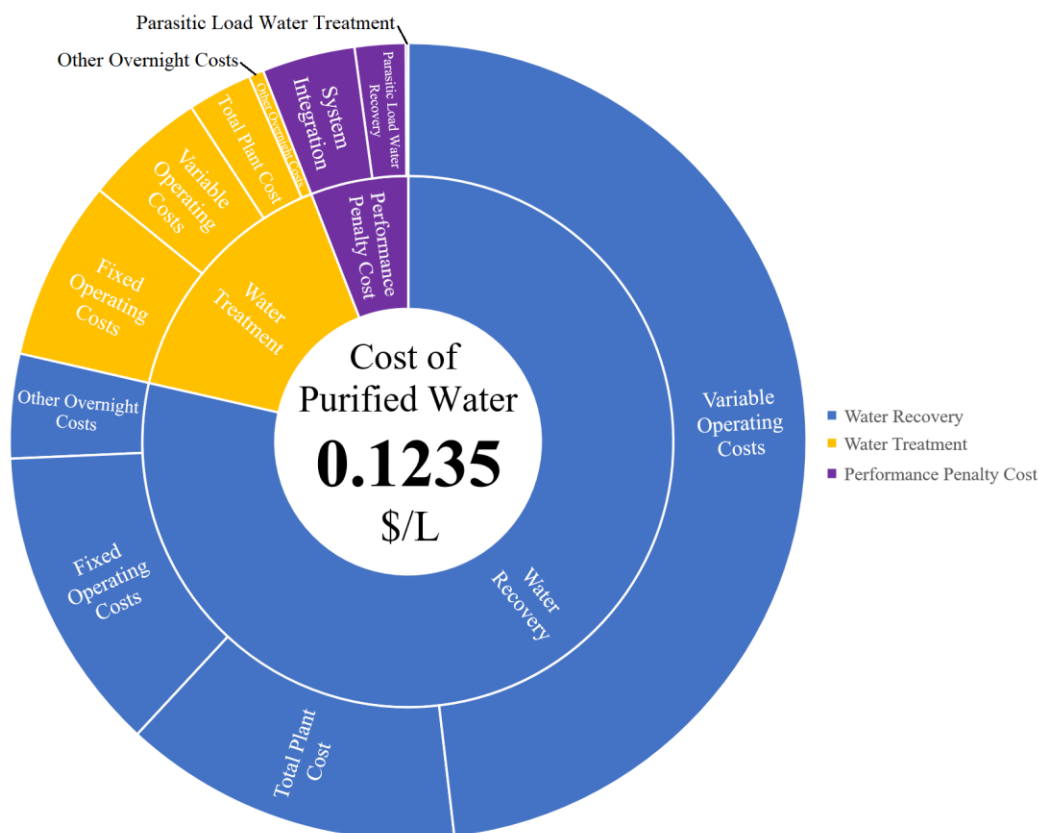


Figure 5-67: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with MEG absorber for water recovery.

Transport Membrane Condenser

The TMC has a similar function as a water-cooled heat exchanger; however, the membrane allows the water to condense in the pores above its dew point temperature and the condensed water instantly mixes with the cooling water inside the cooling tube. In order to avoid cooling water evaporation from the TMC tubes at the inlet section of the TMC, the hot flue gas is cooled to more acceptable inlet temperatures (110 °C) via spray cooling. The water stream leaving the TMC (cooling water + recovered water) is cooled in an air-cooled exchanger and the majority of the water is recirculated back to the TMC as cooling water, while the other portion (net water recovered) is prepared for export. Due to the water transport through the membrane, the recovered water is of high quality [198] and only needs minor treatment (pH-adjustment). The water recovery from the flue gas is limited by the ambient air as the ambient air determines the cooling water temperature. With a 25 °C pinch temperature in the air-cooled exchanger and an 11 °C pinch temperature in the TMC, the flue gas leaves the plant at 51 °C. The pores in the membrane reduce the effective

condensation temperature by around 3.5 °C (superheat); however, this effective condensation temperature is still higher than the flue gas exit temperature in the direct contact condenser resulting in an even lower water recovery than in the direct contact condenser case. The flue gas exit temperature is the highest among the studied scenarios and is beneficial with respect to plume rise. The moderate super heat of the flue gas helps to prevent immediate fog formation but with an ambient temperature of 15 °C fog formation might still occur depending upon gas mixing/dispersion and gas cooling rates. The simplified flow diagram of the 10MW NG case with TMC is shown in **Figure 5-68**. The respective state-point stream data are provided in **Table 5-32**.

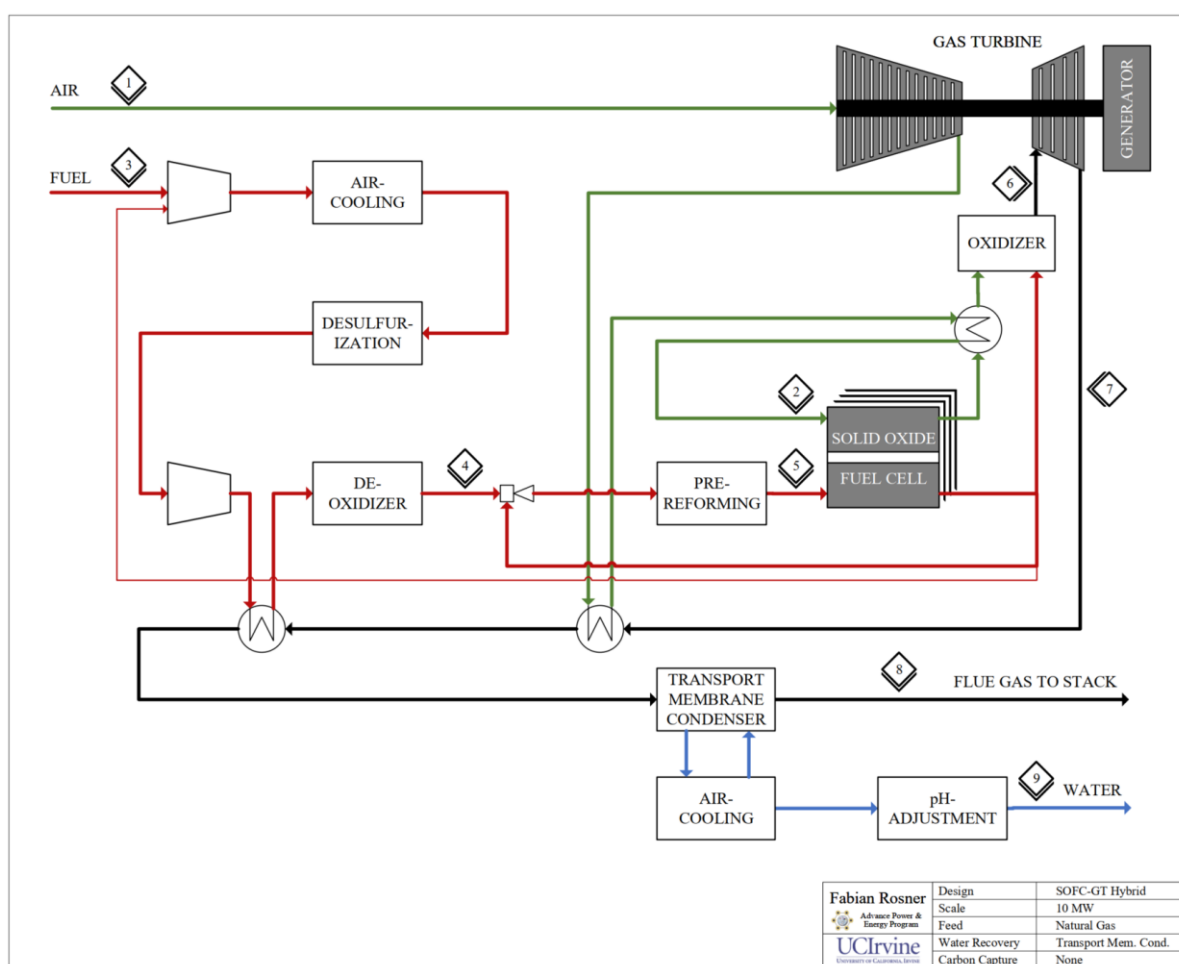


Figure 5-68: Flowsheet of the 10 MW natural gas SOFC-GT hybrid system with transport membrane condenser for water recovery.

Table 5-32: Stream Summary of the 10 MW Natural Gas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	558	51	40
Pressure	bar	1.01	7.18	4.10	21.31	7.18	6.58	1.06	1.02	2.00
Molar Flow Rate	kmol/h	1,075.5	1,075.5	67.4	67.5	349.8	1,144.8	1,144.8	1,118.7	26.1
Mass Flow Rate	kg/h	31,035	31,035	1,168	1,172	7,118	32,203	32,203	31,732	471
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07509	0.07509	0.07684	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72726	0.72726	0.74412	0.00008
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00888	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22836	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09012	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01070	0.20403	0.06163	0.06163	0.06307	Trace
H ₂ O	-	0.00987	0.00987	0.00003	0.00166	0.32719	0.12733	0.12733	0.10709	0.99991
CH ₄	-	0.00000	0.00000	0.93087	0.92858	0.14357	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	0.99999

Thermodynamic Performance. The net power output of the nominal 10MW NG plant with TMC for water recovery is 11,480 kW. The fuel cell island produces 10,669 kW_{DC} and the GT generates an additional 1,389 kW. The TMC case has the lowest pressure drop among the studied water recovery technologies and results in the largest power generation as well as the largest net power generation. The lower pressure drop downstream of the GT (compared to the other water recovery technologies) increases the GT power output, increases the surge margin (11.3%), reduces the SOFC operating pressure (7.18 bar) and leads to a scale up of the SOFC, which is associated with its cooling demand. The FU in this scenario is 88.2%.

The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 578 kW, slightly less than in the LiBr absorber and MEG absorber cases as well as the direct contact condenser case. The electrical efficiency of the TMC case is 74.99%-LHV, which is the highest among the studied water recovery cases. The GT efficiency is 60.4%-LHV and the stack efficiency remains almost unchanged at 82.0%-LHV.

The plant with TMC emits 3,106 kg/h of CO₂ equivalent to 270 kg/MWh. As a result of the high flue gas outlet temperature and a moderate dew point temperature increase of 3.5 °C due to pore condensation, the effective water recovery reaches just 17.8%. At the scale of this nominal 10 MW plant, a total of 14,318 kg of water can be recovered per day. This is equivalent to the water supply needed for the irrigation of about 7,400 m² of farmland in the U.S. [341]. The advantage of the TMC is that the water quality is high, and no liquid or solid waste streams are created. The plant performance data are summarized in **Table 5-33**.

Table 5-33: Thermodynamic Performance of the 10 MW Natural Gas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Gross Power Generation	Unit	TMC
Fuel Cell	kW _{DC}	10,669.13
Gas Turbine	kW	1,389.08
Total Generation	kW	12,058.20
Auxiliary Load		
DC-AC Inverter Losses	kW	320.07
Fuel Gas Compression	kW	168.57
Fuel Gas Intercooler	kW	0.02
Water Recovery System	kW	36.00
Water Treatment	kW	0.07
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.82
Transformer Losses	kW	45.34
Total Auxiliary Load	kW	577.91
Net Power Generation	kW	11,480.30
Plant Performance		
Net Efficiency	%-HHV	67.63
Net Efficiency	%-LHV	74.99
Consumables		
Fuel Gas	kg/h	1,168
Thermal Input	MJ _{LHV} /h	55,101
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	3,106
Carbon Emission	kg/MWh	270
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	17.8
Flue Gas Water Recovery	kg/d	14,318

Economics. The TPC of the 10 MW hybrid with TMC for water recovery is \$29,049,020 resulting in a specific plant cost of 2,530 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The gas turbine island cost remains constant at \$3,651,010 since the same GT, a Dresser Rand KG2-3G EF, is used in this case. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. Heat recuperation in the TMC case is \$2,390,770. The TMC water recovery system costs \$2,038,590 and its cost is comparable to the LiBr water recovery system with \$2,093,160. Also, the water treatment equipment cost with \$95,660 is similar to the LiBr case due to the high water quality. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-69**. A more detailed cost summary of all plant equipment is provided in **Table 5-34**.

SOFC-GT Hybrid Systems with Water Recovery

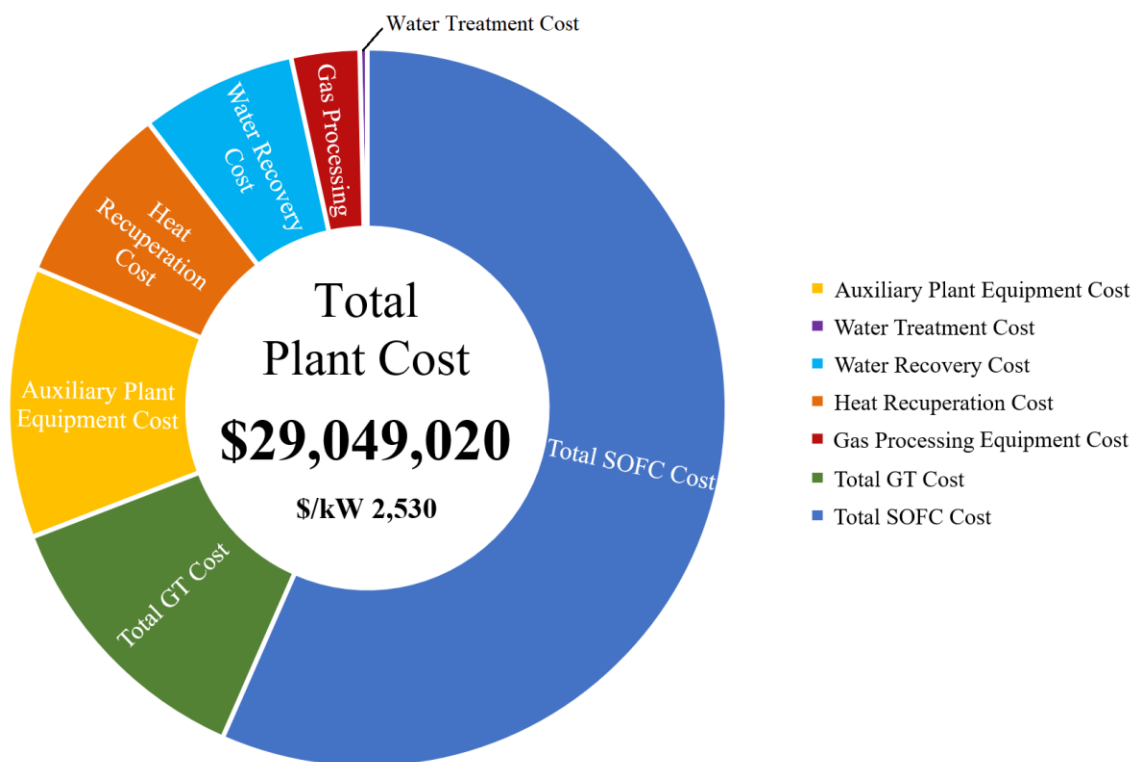


Figure 5-69: Total plant cost and specific plant cost of the 10 MW natural gas SOFC-GT hybrid with transport membrane condenser for water recovery.

The overnight capital cost in the TMC case is \$36,299,540, which is comprised of the TPC, preproduction costs with \$1,672,420, inventory capital cost with \$149,630 and other costs with \$5,429,470, such as initial fill and chemical costs, land, and financing. Annual feed costs remain virtually constant at \$1,731,280. Small changes in variable operating costs, compared to the reference case, are observed due to the slightly lower efficiency and the plant scale. Other operating costs increase due to the water recovery and water treatment equipment.

The fixed operating cost in the TMC case increases due to tax and insurance costs of the additional equipment. The operating labor costs in the TMC case are identical to the previous cases but entirely allocated to the water recovery unit like in the LiBr absorber case due to the minimal water treatment needed in the TMC case. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 77.73 \$/MWh. The COW in the TMC case is 0.1644 \$/L. A summary of the economic key parameters can be found in **Table 5-35**. A detailed breakdown of the COE is shown in **Figure 5-70**.

Table 5-34: Equipment Cost of the 10 MW Natural Gas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Equipment Cost	Unit	TMC
Fuel Cell Island	\$	16,438,540
SOFC Stacks	\$	6,331,450
Power Conditioning	\$	6,276,780
Pressure Vessel, Housing & Assembly	\$	3,254,060
System Installation	\$	576,240
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Gas Processing and Flue Gas Stack	\$	896,380
Fuel Compressors and Intercooler	\$	450,380
Fuel Processing	\$	315,730
Flue Gas Stack System	\$	130,260
Heat Recuperation	\$	2,390,770
Cathode Off-gas Heat Exchanger	\$	1,009,700
Air Preheater	\$	1,114,590
Fuel Preheater	\$	266,480
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	2,134,250
Water Recovery System	\$	2,038,590
Water Treatment	\$	95,660
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	3,538,070
Accessory Electric Plant	\$	1,157,510
Instrumentation & Controls	\$	1,066,740
Improvement to Site	\$	277,350
Buildings & Structures	\$	1,036,460
Total Plant Cost	\$	29,049,020
Specific Plant Cost	\$/kW	2,530

Figure 5-70 shows the COE value of 77.73 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-6.75 \$/MWh) and the whole circle represents the COE without byproduct credit (84.47 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 84.47 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.1644 \$/L in order to reach the COE of 77.73 \$/MWh. This COW is substantially higher than in the air-cooled condenser case, direct contact condenser case, LiBr absorber case and even the MEG absorber case due to the small amount of water recovered or the relatively high capital costs of the water recovery equipment on a per-liter-of-recovered-water basis. Also, the operating labor cost on a per-liter-of-recovered-water basis becomes more expensive if the water recovery is low.

Figure 5-71 shows a detailed breakdown of the COW. In the TMC case, the water recovery is the dominating cost driving factor accounting for over 88.1% of the COW. The fixed operating cost of the water recovery unit is the single largest category with 50.7%, whereby, the operating labor accounts for

SOFC-GT Hybrid Systems with Water Recovery

31.7% of the COW. Water treatment has a share of 2.5% and the performance penalty cost is around 9.4% of the COW.

Table 5-35: Economics of the 10 MW Natural Gas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Item	Unit	TMC
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	29,049,020
Preproduction Costs		
6 Months Labor Cost	\$	988,200
1 Month Maintenance Materials	\$	34,710
1 Month Non-Fuel Consumables	\$	27,440
1 Month Waste Disposal	\$	20
25% of 1 month fuel cost at 100% CF	\$	40,080
2% of TPC	\$	580,980
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	4,390
0.5% of TPC (Spare Parts)	\$	145,250
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	60,860
Land	\$	226,930
Other Owner's Costs (15% of TPC)	\$	4,357,350
Financing Costs	\$	784,320
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	1,731,280
Other Annual Variable Operating Cost at above CF for Initial Year	\$	671,390
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	610,820
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	1,976,400
Property Tax & Insurance	\$	580,980
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.73
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.1644

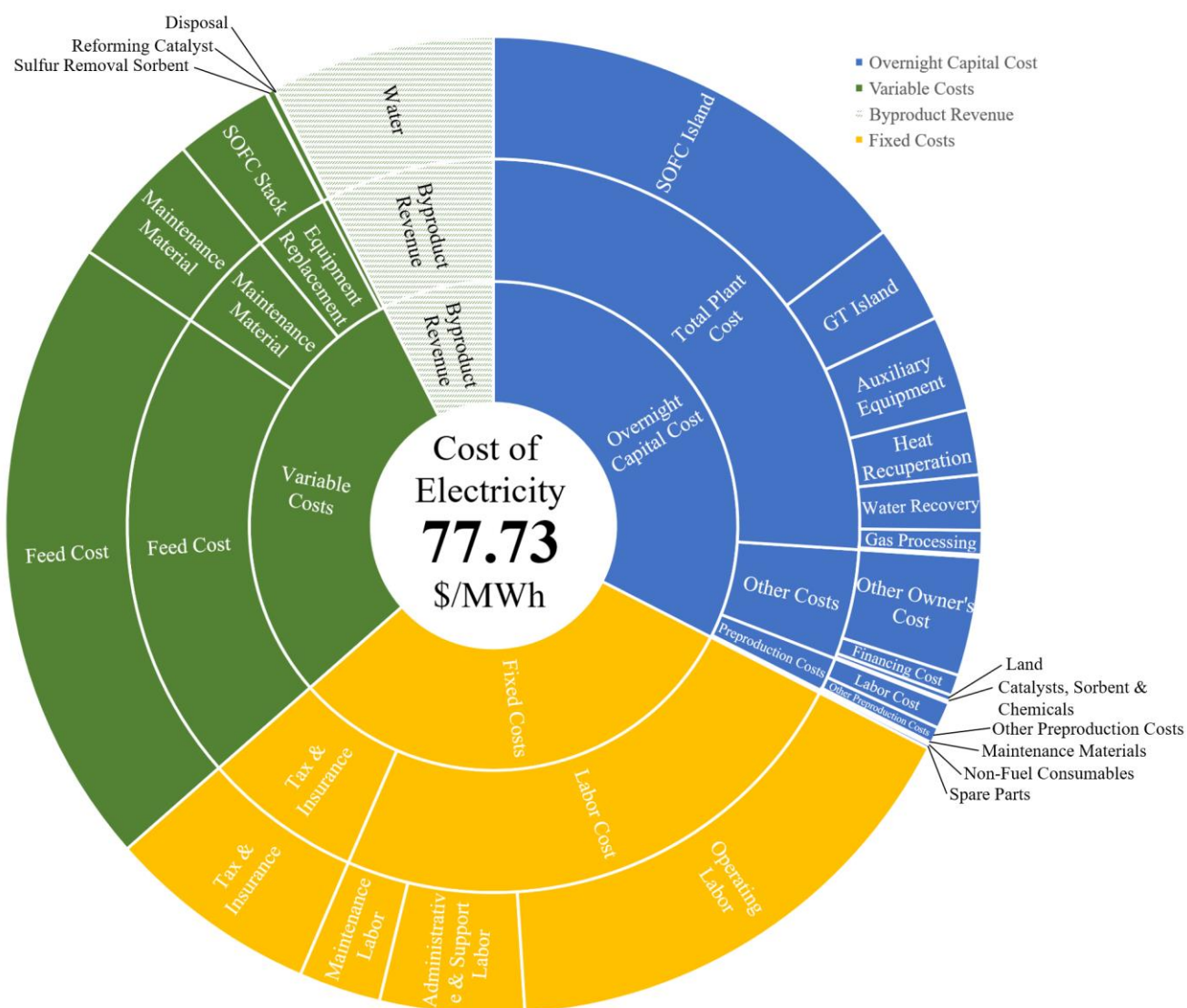


Figure 5-70: Breakdown of the cost-of-electricity of the 10 MW natural gas SOFC-GT hybrid with transport membrane condenser for water recovery.

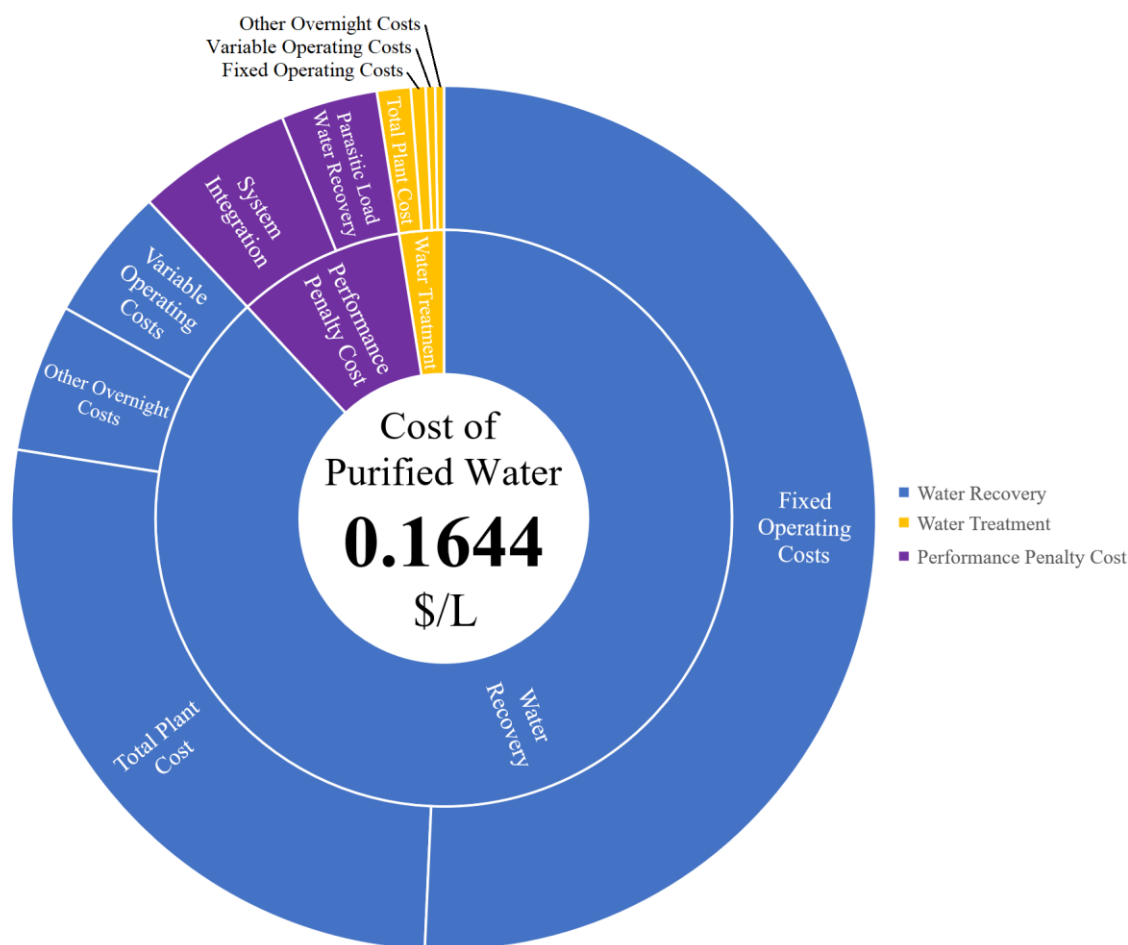


Figure 5-71: Breakdown of the cost-of-water of the 10 MW natural gas SOFC-GT hybrid with transport membrane condenser for water recovery.

Summary of 10 MW Natural Gas SOFC-GT Hybrids

In total five different water recovery technologies, an air-cooled condenser, a direct contact condenser, a LiBr absorption system, a MEG absorption system and a transport membrane condenser, have been integrated into an SOFC-GT hybrid power plant and compared to a reference case without water recovery. The installation of the water recovery technologies is associated with a performance decrease of the SOFC-GT hybrid when compared to the reference case. This performance decrease ranges between 0.28-1.02% points-LHV. The performance decrease can be traced back to two major effects: the increase of GT backpressure and the parasitic load of the water recovery and treatment. The lowest efficiency is obtained for the air-cooled condenser cases (ZLD system further decreases efficiency due to increased parasitic load). While the parasitic load in the air-cooled condenser cases is relatively low, the substantial pressure drop in the air cooled condenser case increases the GT backpressure and thus, reduces the GT power output. The lowest pressure drop is obtained in the TMC scenario, which results in the highest efficiency of the water

recovery cases. The back pressure is not just important with respect to efficiency but also has an influence upon the surge margin (highest surge margin is maintained at low backpressures). For the direct contact condenser, LiBr absorption system, MEG absorption system and the TMC, cooling water or desiccant recirculation is the major parasitic load. A comparison of the electrical efficiencies is provided in **Figure 5-72**.

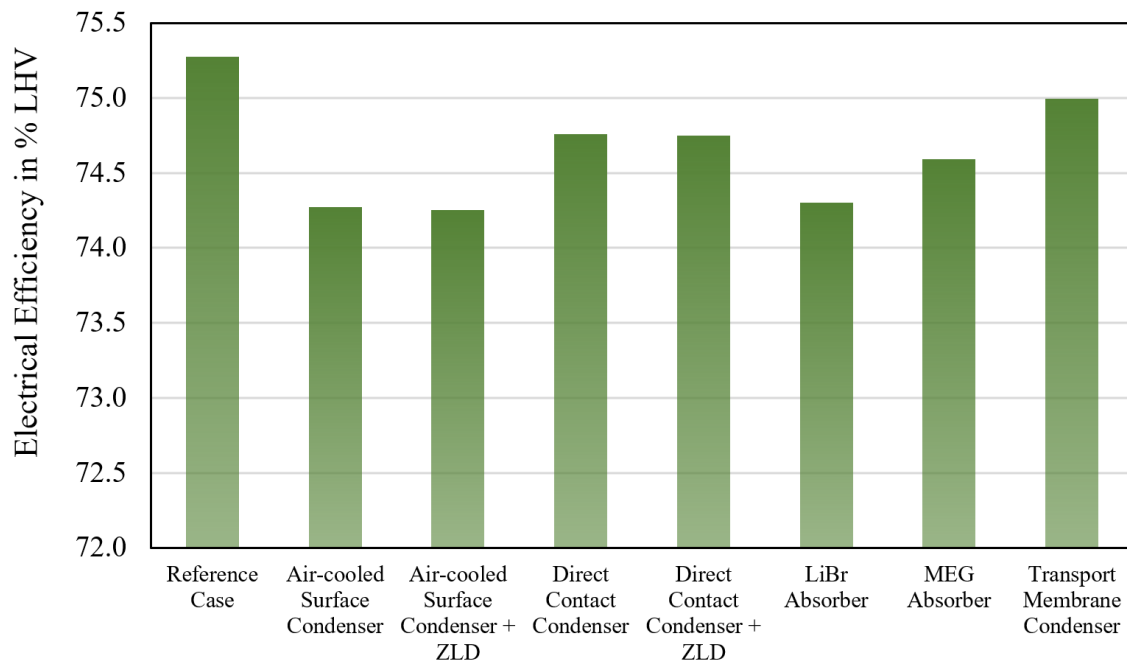


Figure 5-72: Comparison of the efficiencies of the 10 MW natural gas SOFC-GT hybrids with water recovery.

The COE is established in the reference case without water recovery and its value is 77.73 \$/MWh. The COE is held constant throughout the water recovery scenarios and the water sales price is adjusted in order to obtain the same COE as in the reference case. Three main factors contribute to the COW: the capital cost, the operating cost and the amount of water recovered. The air-cooled condenser has low capital and operating costs, and the amount of water recovery is the third highest among the studied technologies, which results in the second lowest COW. The direct contact condenser and TMC mostly suffer from low water recovery and the MEG absorption system exhibits enormous operating costs due to MEG loss. As a result, the LiBr absorption system with its high water recovery and low operating cost is the most economical water recovery technology despite its high investment costs. Although the ZLD system has shown to increase water recovery, the ZLD system has shown to not be of any economic value. A summary of the COE and COW for the various water recovery technologies is presented in **Figure 5-73**

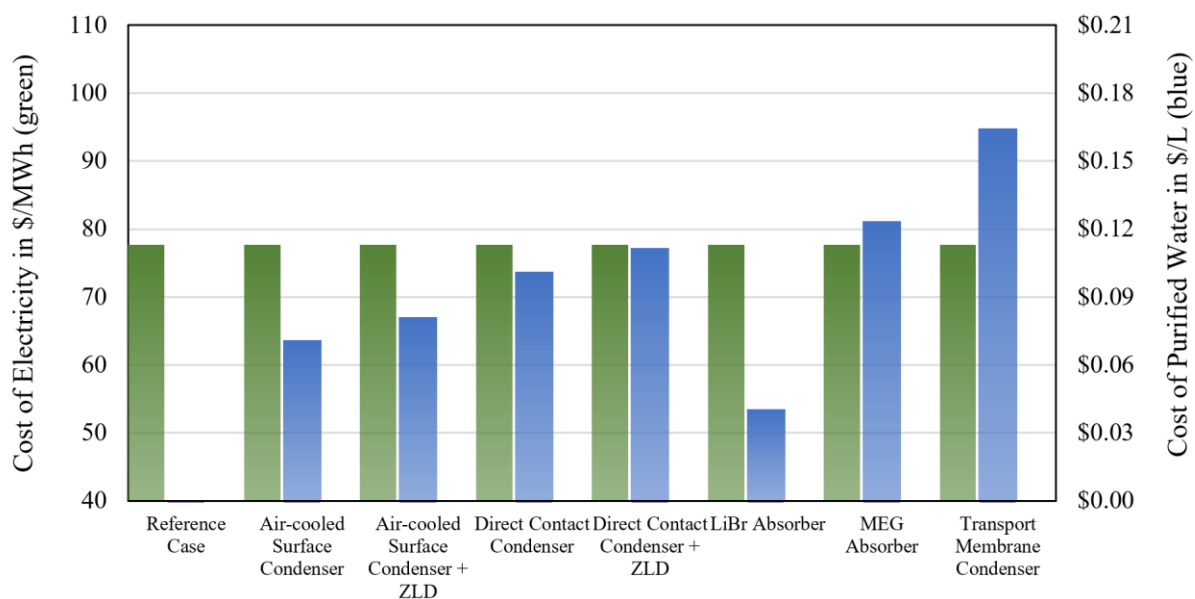


Figure 5-73: Comparison of the cost-of-electricity and cost-of-water of the 10 MW natural gas SOFC-GT hybrids with water recovery.

The water recovery technologies use different physical principles to remove the water vapor from the flue gas. The air-cooled condenser as well as the direct contact condenser rely on cooling the flue gas below the gas's dew point temperature. Once the dew point temperature is reached, the water vapor starts condensing. For both, the air-cooled condenser and the direct contact condenser, the ambient air is the cooling medium. In the case of the air-cooled condenser, ambient air directly cools the flue gas; however, a large surface area is needed to facilitate the heat exchange. In a direct contact condenser, this surface area is created by a packed bed with significantly less pressure drop; however, a secondary cooling water loop is needed, which ultimately leads to a higher flue gas outlet temperature (addition of pinch temperatures ambient air-cooling water, cooling water-flue gas). Thus, the air-cooled condenser is able to recover more water than the direct contact condenser. The absorption cases use a liquid desiccant and no longer rely on the gas's dew point temperature. The desiccant's vapor pressure determines the maximum amount of water that can be recovered from the flue gas. Concentrated aqueous LiBr is an exceptional desiccant and is able to recover more than double the amount of water compared to the air-cooled condenser case. MEG has similar properties as LiBr and the MEG absorption process also reaches very high levels of water recovery. On the downside, the MEG absorption process produces water at a lower quality and the recovered water needs to be treated, which leads to some loss of water. The TMC works upon the principle of nano-capillary condensation. The curvature of a meniscus in a pore changes the vapor pressure of a liquid above its surface. Using this principle, the TMC is able to condense water above the flue gas's dew point temperature. However, current TMC technology uses pore diameters that only lead to dew point temperature increases

of around 3.5 °C, which is not enough to compensate for the increased pinch temperature of the TMC. Thus, the TMC case reaches the lowest water recovery. The results of the water recovery rates are plotted in **Figure 5-74**.

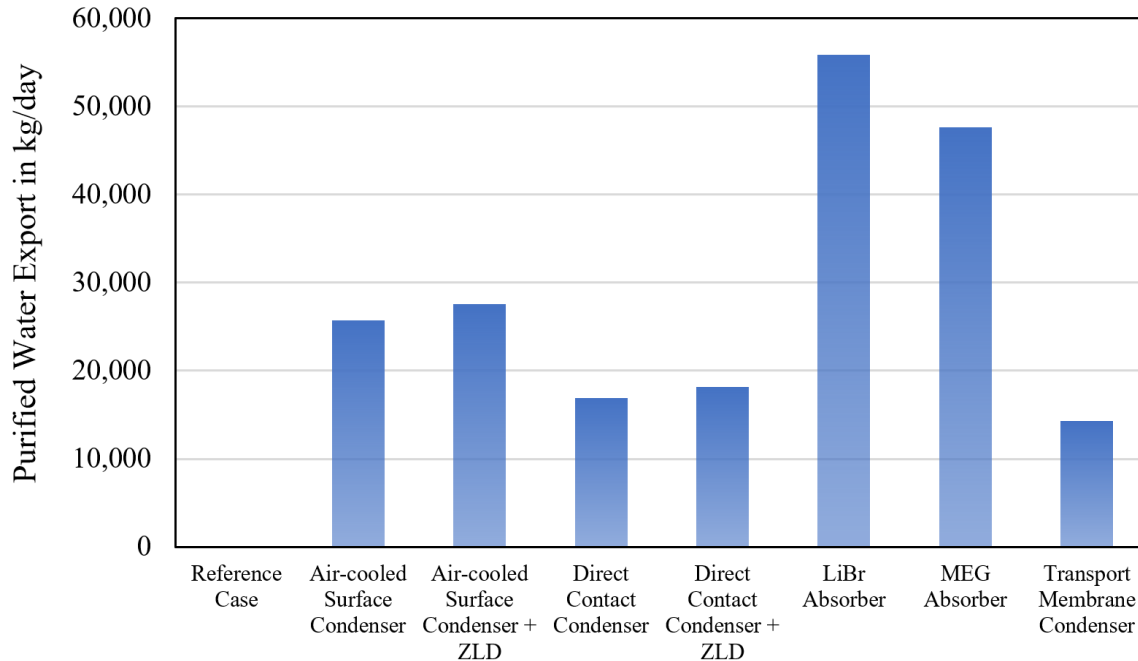


Figure 5-74: Comparison of the water recovery rate of the 10 MW natural gas SOFC-GT hybrids with water recovery.

5.2.2 10 MW Biogas SOFC-GT Hybrids

Reference Case

The design of the SOFC-GT hybrid using biogas requires the integration of an anaerobic digester into the system design, and also requires a thorough biogas cleanup strategy to remove any trace components found in the biogas that can harm the SOFC or GT. To produce biogas from sewage, a heat source is needed to drive the biochemical conversion. The exhaust gas of the SOFC-GT hybrid after heat recuperation (cathode and anode gas pre-heating) still contains considerable amounts of heat, which can be provided to the anaerobic digester (in the NG reference case, the flue gas leaving the plant has a temperature of 327 °C). The amount of heat recovery from the flue gas is limited by the digester's operating temperature as well as the flue gas's dew point temperature. Due to cost benefits, the heat exchanger for the anaerobic digester is made of carbon steel, which requires the "cold" fluid to enter the heat exchanger at or above the flue gas's dew point temperature in order to prevent any condensation on the heat exchanger tubes. With this lower

temperature boundary, heat recovery is limited, and a duct burner is installed upstream of the anaerobic digester's heat exchanger to provide the necessary heat for digestion.

The biogas leaving the digester consist of large amounts of CH_4 and CO_2 but also trace amounts of sulfur compounds, chlorine compounds, oxygen and siloxanes. Since the biogas is supplied at a pressure close to ambient, a three-stage compression with intercooling is chosen to reduce compression power. Sulfur, chlorine and siloxane compounds are removed via sorbents after the first compression stage where the trace components are present at higher partial pressures. Intercooling after the first compression stage reduces the biogas temperature to acceptable inlet values. Subsequent to intercooling, the biogas leaves the air-cooled exchanger at its dew point temperature and a small portion of the biogas is bypassed around the intercooler to provided sufficient superheat to avoid pore condensation in the downstream sorbent beds. While siloxane and chlorine removal are accomplished via expendable sorbents, sulfur removal is a regenerative process, which is more economical in situations with high sulfur concentrations. Deoxidization requires an inlet temperature of around 90°C and is accomplished after the second compression stage (air-cooled exchanger is designed to provide this temperature) where the biogas is mixed with some H_2 -containing anode off-gas in the deoxidizer. Upstream of the deoxidizer, a two-stage membrane separation step is employed to remove CO_2 and upgrade the biogas. Biogas upgrading has four advantages in this design: I) the CO_2 -rich 2nd stage permeate can be used as purge stream for the desulfurization unit and the small quantities of CH_4 in the permeate provide the heat needed for regeneration II) without upgrading the

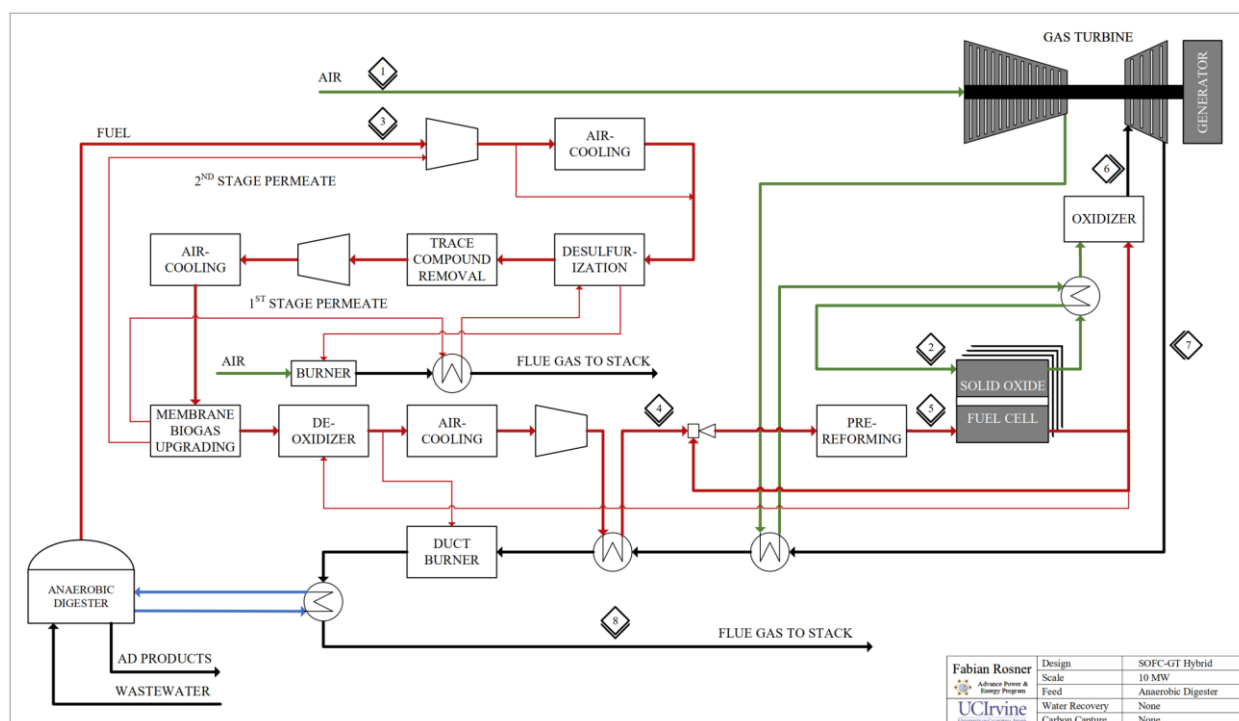


Figure 5-75: Flowsheet of the 10 MW biogas SOFC-GT hybrid system without water recovery.

biogas, the CO₂ would end up in the flue gas and reduce the partial pressure of water, which reduces water recovery III) biogas upgrading increases the chemical potential difference in the SOFC and reduces its size IV) higher heating value fuels tend to increase surge margin in the GT (surge limit is already a concern even with NG). After the biogas cleanup, the biogas contains almost 90% of CH₄, which is similar to natural gas. Thus, the downstream SOFC-GT hybrid configuration is identical to the previous NG cases.

The biogas cases are using the same thermal SOFC constraints as the previous SOFC-GT hybrid cases, which can be found in **Table 4-13**, together with the other SOFC design parameters. An illustration of the plant design used in the reference case is provided in **Figure 5-75**. The corresponding state-point steam data are shown in **Table 5-36**.

Table 5-36: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	27	322	700	902	556	63
Pressure	bar	1.01	7.16	1.01	24.96	7.16	6.56	1.07	1.01
Molar Flow Rate	kmol/h	1,077.0	1,077.0	135.8	78.0	334.6	1,143.3	1,143.3	1,150.7
Mass Flow Rate	kg/h	31,080	31,080	3,600	1,422	7,064	32,201	32,201	32,337
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07402	0.07402	0.06197
N ₂	-	0.77315	0.77315	0.00966	0.01617	0.00774	0.72935	0.72935	0.72477
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00865
H ₂	-	0.00000	0.00000	0.00000	0.00271	0.19509	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00829	0.07883	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06824	0.21854	0.06273	0.06273	0.06859
H ₂ O	-	0.00987	0.00987	0.03381	0.01164	0.33324	0.12519	0.12519	0.13601
CH ₄	-	0.00000	0.00000	0.58439	0.89285	0.16654	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1

Thermodynamic Performance. The net power output of the nominal 10 MW biogas reference case is 9,396 kW, whereby, the SOFC produces 10,845 kW of electricity and the GT produces another 1,320 kW of electricity. The auxiliary load accounts for 2,768 kW and includes the balance-of-plant equipment as well as the anaerobic digester load. In the NG reference case, the inverter losses are the dominating auxiliary load followed by fuel compression. In the biogas reference case, fuel compression is significantly higher due to the lower supply pressure, low heating value of the biogas, and recycling associated with biogas upgrading. However, the largest power consumer is the anaerobic digester accounting for over 54.3% of the auxiliary load.

The efficiency of the 10MW biogas reference case is 53.07%-LHV. The efficiency of the SOFC stack on a standalone basis is 82.1%-LHV and is possible due to the high degree of internal reforming. The GT operates at an efficiency of 58.5%-LHV utilizing the low heating value anode off-gas from the SOFC. While the off-design operation reduces the component efficiency of the expander, the efficiency of the GT

increases substantially from initially 24.5% at ISO conditions using NG [247]. This is a result of the very high fuel and air inlet temperatures as well as the low heating value of the anode off-gas. Most of the energy is already contained in the air and fuel streams in the form of high temperature heat when they reach the oxidizer, and the oxidation of the unutilized anode off-gas is only able to increase the air temperature by 243 °C, from initially 659 °C to 902 °C. This increase in efficiency is observed despite a decrease in TIT compared to NG operation where a TIT of 943 °C is reached. The TIT in the biogas reference case is limited by the heat integration and 902 °C is the maximum achievable TIT despite the fact that a TIT of up to 909 °C would be permissible (maintaining blade metal temperature). The FU of the biogas reference case is 88.7% and the surge margin is 11.8% (lower TIT helps to increase the surge margin). Overall, this shows that the hybrid performance on biogas is very similar to NG after upgrading, only the auxiliary load for the digester and gas compression lead to a reduced efficiency.

Table 5-37: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrid without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	10,844.62
Gas Turbine	kW	1,319.98
Total Generation	kW	12,164.59
Auxiliary Load		
DC-AC Inverter Losses	kW	325.34
Anaerobic Digester	kW	1,504.50
Fuel Gas Compression	kW	801.50
Fuel Gas Intercooler	kW	0.26
Regen. Sulfur Removal	kW	83.17
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.89
Transformer Losses	kW	45.74
Total Auxiliary Load	kW	2,768.40
Net Power Generation	kW	9,396.20
Plant Performance		
Net Efficiency	%-HHV	47.65
Net Efficiency	%-LHV	53.07
Consumables		
Fuel Gas	kg/h	3,600
Thermal Input	MJ _{LHV} /h	63,738
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Net Carbon Emission (Absolute)	kg/h	0 (5,670)
Net Carbon Emission (Absolute)	kg/MWh	0 (603)
Liquid Waste	kg/h	201
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Also, the biogas hybrid plant requires no water input as it is equipped with dry cooling technology. In contrast to the NG reference case, the biogas hybrid plant produces liquid waste streams in the gas processing part of the plant, which amounts to 201 kg/h. Since the plant is build at a wastewater treatment facility, these waste streams are directly treated within the facility and no disposal fee has been assumed for liquid waste streams in the biogas scenarios. The CO₂ emissions of the 10MW biogas hybrid are considered to be virtually zero as the feedstock, sewage, is made of atmospheric carbon (note that biogas production might result in some indirect CO₂ emissions). On an absolute basis, CO₂ emissions are 5,670 kg/h, which could be captured to make this plant carbon negative. The thermodynamic performance results are summarized in **Table 5-37**.

Economics. The TPC of the 10MW biogas hybrid without water recovery is \$46,181,660 (4,915 \$/kW), whereby, the SOFC island accounts for \$17,291,670 (37.4%). The TPC for the SOFC island includes: stack cost, power conditioning, pressure vessel, housing and assembly, and system installation. The SOFC stack with \$6,817,330 is the single most expensive category of the SOFC island followed by the power conditioning system, which includes the inverters with \$6,331,090. Overall, the SOFC island is slightly more expensive than in the NG reference case, which is mostly related to the scale of the plant and the higher number of spare stacks installed due to the fixed number of stacks per module. The GT island contributes \$3,651,010 to the plant's TPC, a share of 7.9%, while generating 10.9% of the gross power. The anaerobic digester adds another \$8,810,340 to the TPC. The cost of fuel gas processing drastically increased compared to the 10 MW NG reference case. Due to the low biogas supply pressure, compression cost is increased, resulting in a cost of \$2,905,450 for compressors and intercoolers. Other gas processing equipment includes the regenerative sulfur removal process, the expendable sorbents for sulfur removal, chlorine removal and siloxane removal as well as the CO₂ membrane process and the deoxidizer. In particular, the CO₂ membrane process with \$3,085,570 and the regenerative sulfur removal with \$2,126,140 make the gas processing very expensive. Heat recuperation accounts for 5.0% of the plant's TPC. Auxiliary plant equipment costs are estimated with \$4,952,290, which includes accessory electric plant, instrumentation and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-76**. A more detailed overview of the capital cost is provided in **Table 5-38**.

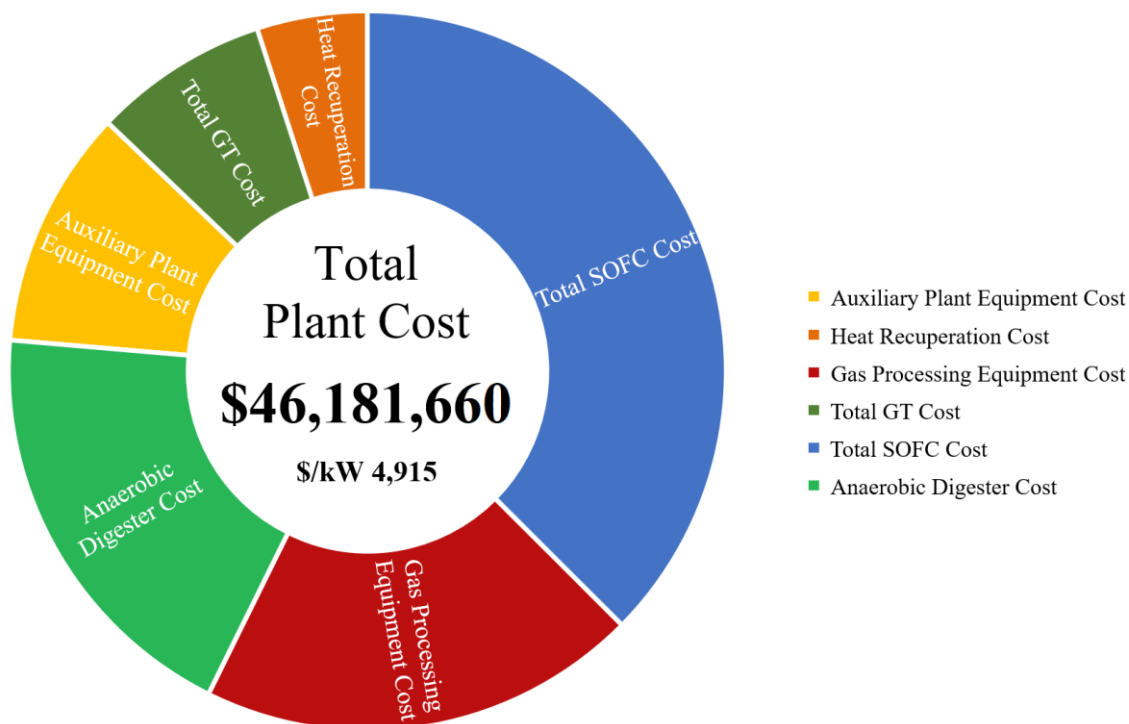


Figure 5-76: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid without water recovery.

The total overnight capital cost in the 10 MW biogas reference case is \$61,137,420 and is comprised of the TPC, the preproduction costs with \$2,464,680, the inventory capital costs with \$331,290 and other costs with \$12,159,780 including financing and land costs. Land cost (\$3,555,070) is substantially increased over the 10 MW NG case due to the enormous amount of land required for the digesters. Variable annual costs are \$2,006,640 and mostly maintenance material cost, equipment replacement costs for the SOFCs and CO₂ membranes as well as chemical costs associated with the digester operation. Fixed costs are comprised of annual operating labor with \$2,710,870 and costs associated with tax and insurance (\$923,630 per year). A detailed breakdown of the overnight capital cost, variable costs and fixed costs is provided in **Table 5-39**.

Table 5-38: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrid without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	17,291,670
SOFC Stacks	\$	6,817,330
Power Conditioning	\$	6,331,090
Pressure Vessel, Housing & Assembly	\$	3,535,710
System Installation	\$	607,540
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Anaerobic Digester	\$	8,810,340
Anaerobic Digester	\$	8,073,240
AD Heat Supply	\$	737,090
Gas Processing and Flue Gas Stack	\$	9,176,550
Fuel Compressors and Intercooler	\$	2,905,450
Fuel Processing	\$	6,140,840
Flue Gas Stack System	\$	130,260
Heat Recuperation	\$	2,299,790
Cathode Off-gas Heat Exchanger	\$	1,009,900
Air Preheater	\$	1,117,370
Fuel Preheater	\$	172,530
Water Recovery	\$	0
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	4,952,290
Accessory Electric Plant	\$	1,167,730
Instrumentation & Controls	\$	1,695,880
Improvement to Site	\$	440,930
Buildings & Structures	\$	1,647,750
Total Plant Cost	\$	46,181,660
Specific Plant Cost	\$/kW	4,915

The resulting COE for the biogas fueled 10 MW SOFC-GT hybrid plant without water recovery is 137.19 \$/MWh. The detailed COE breakdown in **Figure 5-77** shows that the increased capital cost for gas processing and anaerobic digestion substantially increase the COE. Overnight capital cost in this biogas scenario contributes over 44.5% to the COE. On the other hand, variable operating costs remain roughly constant on an absolute basis (reduced to 19.7% on a relative basis) despite the fact that the anaerobic digester adds 5.82 \$/MWh to the COE. The reason is that in the biogas case fuel expenditures are saved since the fuel is produced on site from sewage. Within the fixed costs, additional operating labor associated with the digestion process and increased tax and insurance expenses lead to a further increase in the COE.

Table 5-39: Economics of the 10 MW Biogas SOFC-GT Hybrid without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	46,181,660
Preproduction Costs		
6 Months Labor Cost	\$	1,355,430
1 Month Maintenance Materials	\$	65,650
1 Month Non-Fuel Consumables	\$	119,920
1 Month Waste Disposal	\$	50
25% of 1 month fuel cost at 100% CF	\$	0
2% of TPC	\$	923,630
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	100,380
0.5% of TPC (Spare Parts)	\$	230,910
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	430,550
Land	\$	3,555,070
Other Owner's Costs (15% of TPC)	\$	6,927,250
Financing Costs	\$	1,246,900
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	2,004,640
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	0
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	2,710,870
Property Tax & Insurance	\$	923,630
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	N/A

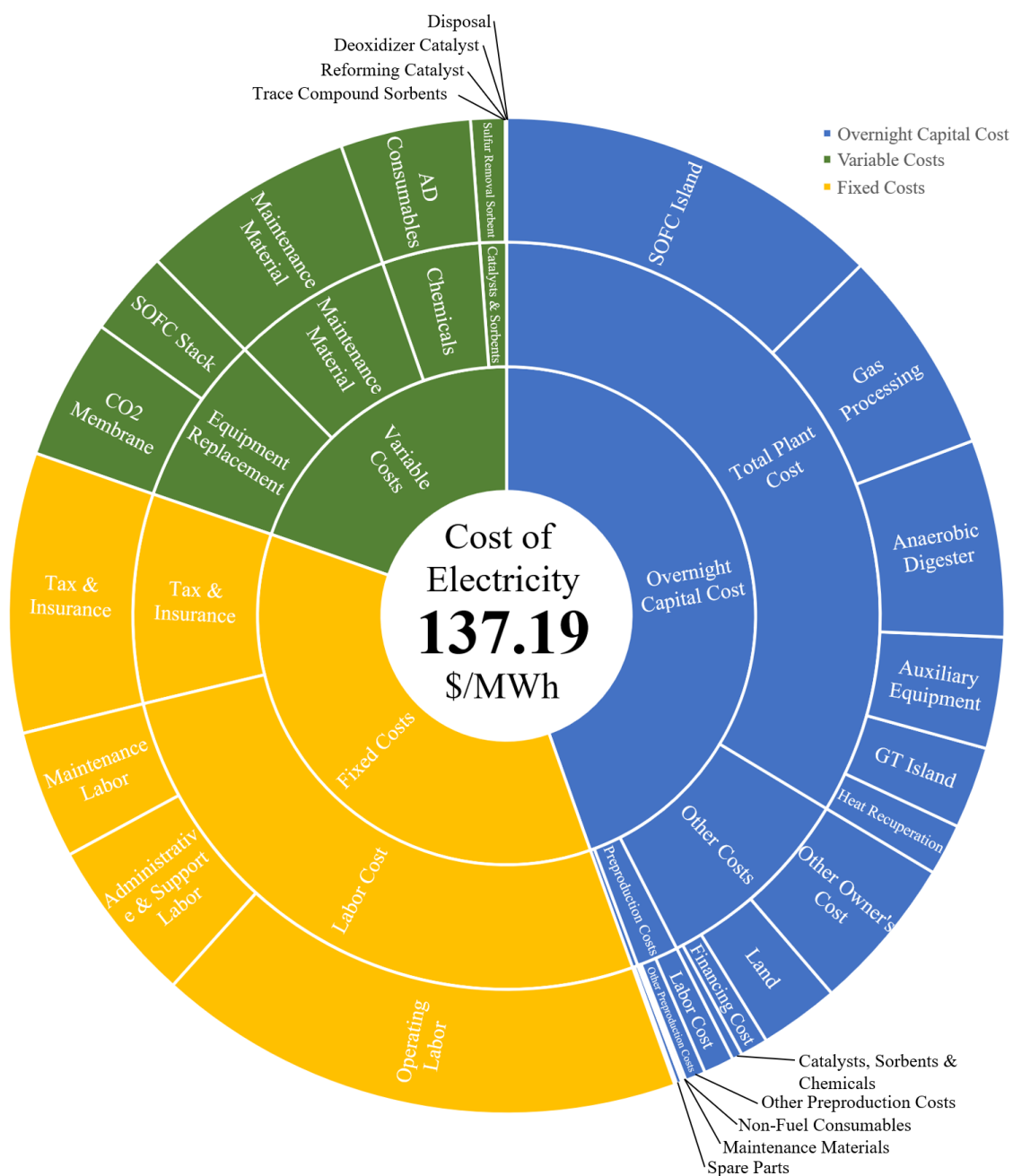


Figure 5-77: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid without water recovery.

Air-Cooled Condenser

The design of the nominal 10 MW biogas SOFC-GT hybrid plant with air-cooled condenser for water recovery is largely identical to the 10 MW biogas reference case with the exception of the water recovery and treatment equipment downstream of the anaerobic digester heat exchanger. Due to the additional heat recovery (anaerobic digester) the inlet temperature to the air-cooled condenser (65 °C) for water recovery is significantly lower than in the 10 MW NG case, which helps to reduce the size of the air-cooled condenser. The air-cooled condenser operates with a pinch temperature of 25 °C (identical to NG case) and the flue gas leaves the plant through the stack at a saturated state and a temperature of 40 °C. This raises the same concerns about plume rise and fog formation as previously discussed in the 10 MW NG air-cooled condenser case.

Although the size of the air-cooled condenser can be reduced in this biogas scenario, the pressure drop associate with the condenser remains relatively high at a value of 0.08 bar. The recovered water from the air-cooled condenser is in direct contact with the flue gas, which results in considerable amounts of flue gas constituents ending up in the water. To convert the recovered water to useable water, it is treated in an ultrafiltration unit, reverse osmosis unit and pH adjustment unit. A general flow diagram of the 10MW

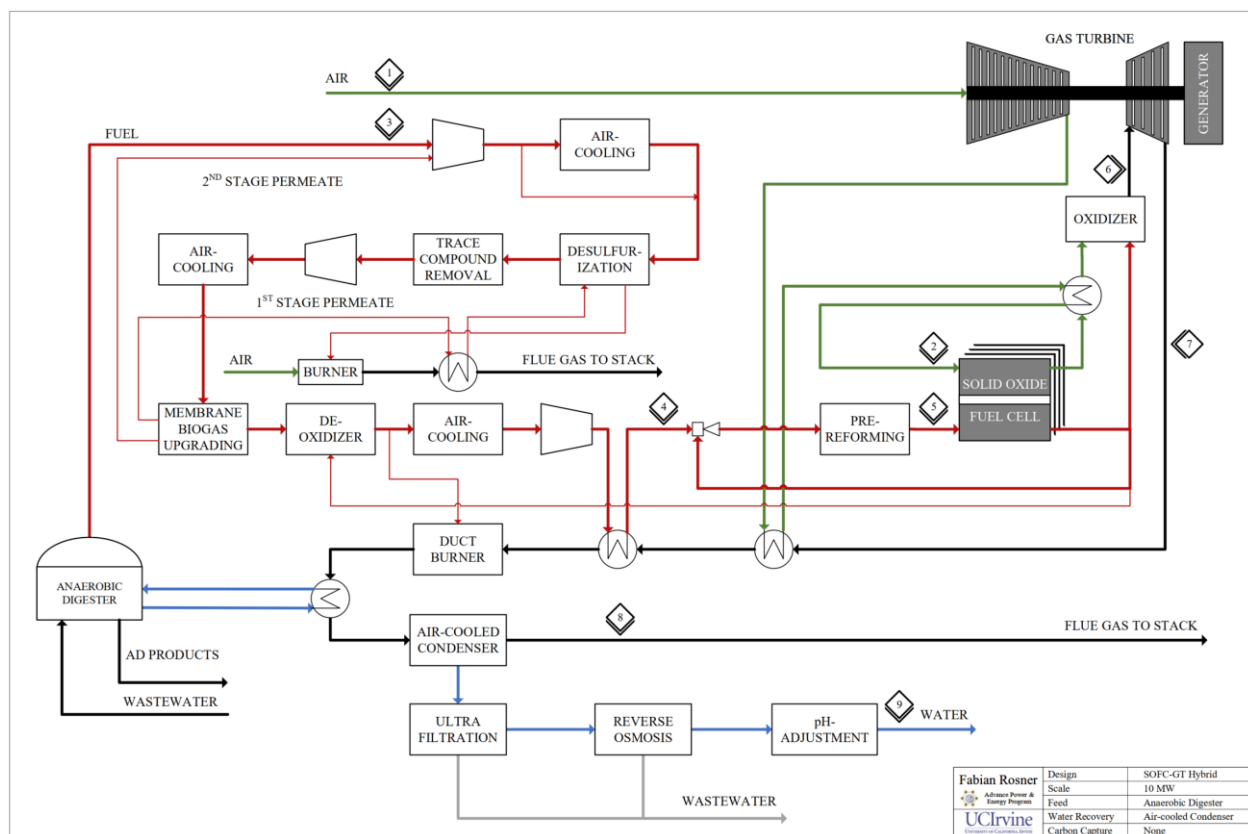


Figure 5-78: Flowsheet of the 10 MW biogas SOFC-GT hybrid system with air-cooled condenser for water recovery.

biogas case with air-cooled condenser water recovery is shown in **Figure 5-78**. The respective state-point stream data are shown in **Table 5-40**.

Table 5-40: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	910	575	40	41
Pressure	bar	1.01	7.27	1.01	25.18	7.27	6.67	1.15	1.01	2.00
Molar Flow Rate	kmol/h	1,069.1	1,069.1	133.5	77.5	333.6	1,135.2	1,135.2	1,065.2	70.6
Mass Flow Rate	kg/h	30,850	30,850	3,540	1,411	7,037	31,970	31,970	30,710	1,273
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07373	0.07373	0.06779	Trace
N ₂	-	0.77315	0.77315	0.00966	0.01618	0.00773	0.72912	0.72912	0.77707	0.00090
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00927	0.00003
H ₂	-	0.00000	0.00000	0.00000	0.00271	0.19589	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00827	0.07923	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06757	0.21806	0.06289	0.06289	0.07285	Trace
H ₂ O	-	0.00987	0.00987	0.03381	0.01144	0.33293	0.12556	0.12556	0.07302	0.99901
CH ₄	-	0.00000	0.00000	0.58439	0.89374	0.16614	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1	0.99994

Thermodynamic Performance. The net power output of the nominal 10 MW biogas case with air-cooled condenser for water recovery is 9,184 kW. The GT power output decreased from 1,320 kW in the reference case to 1,178 kW in the air-cooled condenser water recovery case, a decrease of 10.8%. Again, this reduction in GT power output is related to the increased backpressure, which reduces the pressure ratio across the GT expander, despite the fact that the compressor pressure ratio increases. As a result of the higher compressor discharge pressure, the GT operates closer to its surge limit with a surge margin of 9.3%, a reduction of 2.5% points compared to the reference case (11.8% surge margin). Due to the increased pressure ratio, the SOFC operates now at a pressure of 7.27 bar. Consequently, the cooling air flow needs to be adjusted to stay within the operating limits set by the thermal constraints. The FU in the air-cooled condenser case is 88.5%, slightly lower than in the biogas reference case, which is associated a reduced SOFC power output of 25 kW.

In a SOFC-GT hybrid the SOFC and GT are strongly coupled, and the respective operating points impact one another. First, the GT compressor discharge pressure causes a change in the SOFC's cooling demand, which then impacts the maximum permissible TIT of the GT, which again impacts the pressure ratio. In the biogas reference case, the maximum permissible TIT was not reached due to limitations in the heat integration. In the air-cooled condenser case, the higher GT backpressure increases the GT's exhaust temperature and allows the TIT to reach its maximum permissible value of 909.8 °C. This higher GT

exhaust temperature further improves heat recovery for the digester and less fuel needs to be burned in the duct burner. The auxiliary load is largely similar to the reference case. Fuel gas compression is reduced due to less fuel demand by the duct burner and water recovery and water treatment add new loads to the plant.

The electrical efficiency of the air-cooled water recovery case decreased by 0.31% points to 52.76%-LHV and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GT due to the backpressure. The GT efficiency decreased to 51.8%-LHV. The stack efficiency remains almost unchanged at 82.2%-LHV. The minimal increase is a result of the lower FU but also internal reforming impacts the stack efficiency.

Carbon emissions are not particularly meaningful when utilizing carbon neutral biogas but show how much carbon could be captured if desired to make the plant carbon negative. The 10 MW biogas plant with air-cooled condenser emits 5,572 kg of CO₂ per hour. The water recovery from the flue gas in the air-cooled condenser is 49.5%, slightly more than in the NG case due to a higher water partial pressure in the exhaust

Table 5-41: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Gross Power Generation		Unit	Air-cooled Condenser
	Fuel Cell	kW _{DC}	10,766.81
	Gas Turbine	kW	1,178.12
Total Generation		kW	11,944.93
Auxiliary Load			
	DC-AC Inverter Losses	kW	323.00
	Anaerobic Digester	kW	1502.12
	Fuel Gas Compression	kW	788.53
	Fuel Gas Intercooler	kW	0.26
	Regen. Sulfur Removal	kW	81.87
	Water Recovery System	kW	9.46
	Water Treatment	kW	2.58
	Zero Liquid Discharge System	kW	N/A
	Miscellaneous Balance of Plant	kW	7.75
	Transformer Losses	kW	44.92
Total Auxiliary Load		kW	2760.49
Net Power Generation		kW	9,184.45
Plant Performance			
	Net Efficiency	%-HHV	47.36
	Net Efficiency	%-LHV	52.76
Consumables			
	Fuel Gas	kg/h	3,540
	Thermal Input	MJ _{LHV} /h	62,672
	Water	kg/h	0
Emissions and Waste Streams			
	Carbon Capture	%	N/A
	Net Carbon Emission (Absolute)	kg/h	0 (5,572)
	Net Carbon Emission (Absolute)	kg/MWh	0 (607)
	Liquid Waste	kg/h	300
	Solid Waste	g/h	0
Water Recovery			
	Flue Gas Water Recovery	%	49.5
	Flue Gas Water Recovery	kg/d	30,501

after the duct burner compared to the NG case. At the scale of this nominal 10 MW plant, a total of 30,501 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW biogas SOFC-GT hybrid plant with air-cooled condenser water recovery system can produce enough water for the irrigation of over 15,700 m² of farmland (including 90% capacity factor). Liquid waste stream in this case amount to 300 kg/h. The plant performance data are summarized in **Table 5-41**.

Economics. The TPC of the 10 MW biogas SOFC-GT hybrid with air-cooled condenser for water recovery is \$46,788,980, which results in a specific plant cost of 5,094 \$/MWh. The SOFC island cost remains relatively stable at \$17,262,600. Small differences arise from the modular SOFC design and the plant scale, which has an impact on the number of spare stacks installed. GT island costs remain constant at \$3,651,010 as the GT is an off-the-shelf item. The anaerobic digester cost is reduced due to the lower fuel demand due to the reduced plant scale as well as increased heat recovery from the flue gas. This increased GT exhaust heat recovery from the flue gas also shifts the heat recuperation costs from the cathode off-gas exchanger towards the air pre-heater. Cost for the water recovery and treatment equipment is \$812,650, whereby, the water recovery equipment accounts for 42.8% of the cost. Compared to the NG case with air-cooled condenser, the condenser equipment cost reduced by 18.5% due to the lower gas inlet temperature.

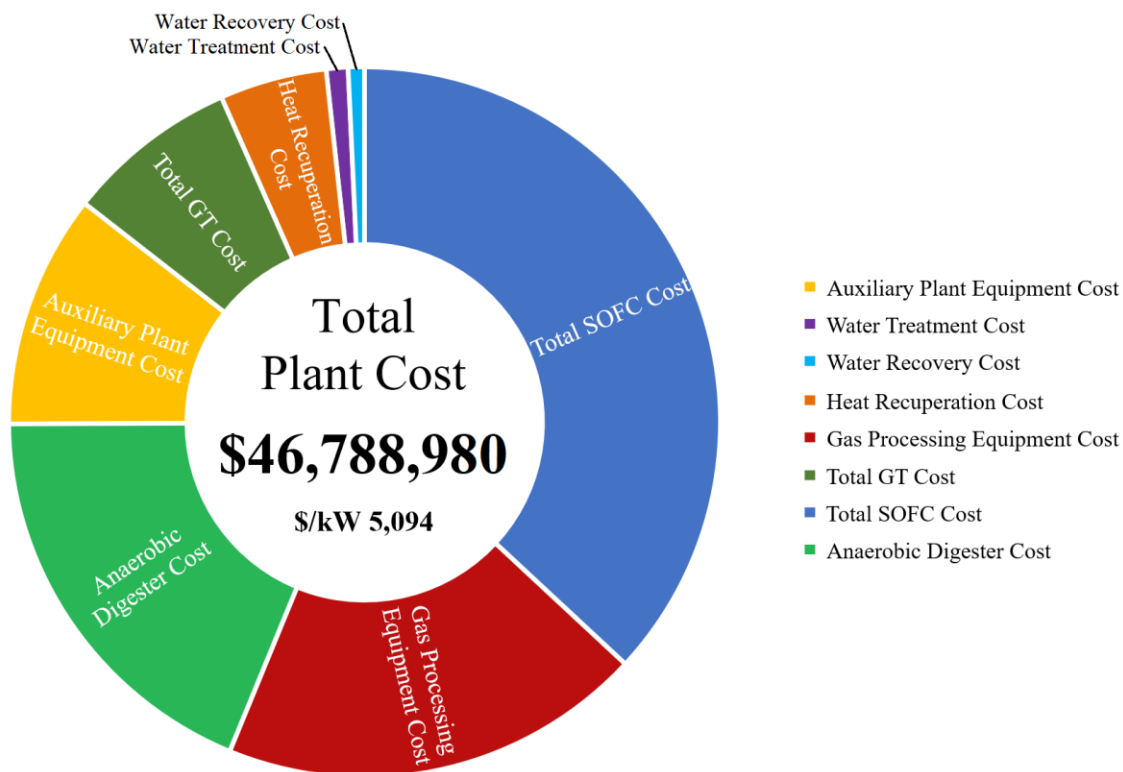


Figure 5-79: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid without water recovery.

Auxiliary plant equipment accounts for \$4,980,980. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-79**. A more detailed cost summary of all plant equipment is provided in **Table 5-42**.

Table 5-42: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Equipment Cost	Unit	Air-cooled Condenser
Fuel Cell Island	\$	17,262,600
SOFC Stacks	\$	6,822,700
Power Conditioning	\$	6,282,560
Pressure Vessel, Housing & Assembly	\$	3,550,810
System Installation	\$	606,520
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Anaerobic Digester	\$	8,765,610
Anaerobic Digester	\$	7,991,460
AD Heat Supply	\$	774,150
Gas Processing and Flue Gas Stack	\$	9,035,510
Fuel Compressors and Intercooler	\$	2,841,380
Fuel Processing	\$	6,064,430
Flue Gas Stack System	\$	129,700
Heat Recuperation	\$	2,280,630
Cathode Off-gas Heat Exchanger	\$	981,200
Air Preheater	\$	1,125,850
Fuel Preheater	\$	173,580
Water Recovery	\$	812,650
Water Recovery System	\$	347,900
Water Treatment	\$	464,750
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	4,980,980
Accessory Electric Plant	\$	1,146,640
Instrumentation & Controls	\$	1,718,190
Improvement to Site	\$	446,730
Buildings & Structures	\$	1,669,420
Total Plant Cost	\$	46,788,980
Specific Plant Cost	\$/kW	5,094

The total overnight capital cost is \$61,946,550, which represents an increase of \$809,130 compared to the biogas reference case, mostly due to the TPC increase in the water recovery case. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain at zero. Other variable operating costs increase due to the water recovery and water treatment equipment. In the water recovery cases, water byproduct sales reduce the variable operating costs. In the air-cooled condenser case, the water byproduct revenue is \$551,780 for the initial year. The fixed operating cost in the air-cooled condenser case increases due to labor costs associated with the water recovery as well as tax and insurance costs of the additional equipment. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COEs for all 10 MW biogas cases remain constant at 137.19 \$/MWh. The COW in the air-

cooled condenser case is 0.0549 \$/L and 0.016 \$/L cheaper than in the equivalent NG case. A summary of the economic key parameters can be found in **Table 5-43**. A detailed breakdown of the COE is shown in **Figure 5-80**.

Table 5-43: Economics of the 10 MW Biogas SOFC-GT Hybrid with Air-cooled Condenser for Water Recovery

Item	Unit	Air-cooled Condenser
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	46,788,980
Preproduction Costs		
6 Months Labor Cost	\$	1,480,390
1 Month Maintenance Materials	\$	66,590
1 Month Non-Fuel Consumables	\$	119,020
1 Month Waste Disposal	\$	50
25% of 1 month fuel cost at 100% CF	\$	0
2% of TPC	\$	935,780
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	98,760
0.5% of TPC (Spare Parts)	\$	233,940
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	423,750
Land	\$	3,517,620
Other Owner's Costs (15% of TPC)	\$	7,018,350
Financing Costs	\$	1,263,300
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	2,005,140
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	551,780
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	2,960,780
Property Tax & Insurance	\$	935,780
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0549

Figure 5-80 shows the COE value of 137.19 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-7.62 \$/MWh) and the whole circle represents the COE without byproduct credit (144.81 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 144.81 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0549 \$/L in order to reach a COE of 137.19 \$/MWh.

The COW is composed of costs associated with water recovery, water treatment and performance penalty costs. The water treatment accounts for 36.1% of the COW, whereby, the fixed operating costs, mostly labor, represent the largest share (in this analysis, the labor costs associated with the water recovery and treatment are equally split between recovery and treatment). Variable water treatment costs account for about 5.9% of the COW, which includes annual maintenance, equipment replacement and disposal costs.

SOFC-GT Hybrid Systems with Water Recovery

The water treatment TPC accounts for 2.9% of the COW. Water recovery contributes 31.4% to the COW with fixed operating costs being the largest contributor. The TPC of the water recovery unit has a 5.1%-share in the COW. Performance penalties come from the parasitic power consumption of the water recovery (1.8% of COW) and water treatment (0.5% of COW) as well as system integration losses, such as higher GT backpressure, which reduces the GT power output. As mentioned before, the GT power output decreased by 10.8% in the air-cooled condenser case compared to the biogas reference case, which substantially contributes the system integration losses, and system integration losses represent 30.2% of the COW. A detailed breakdown of the COW can be found in **Figure 5-81**.

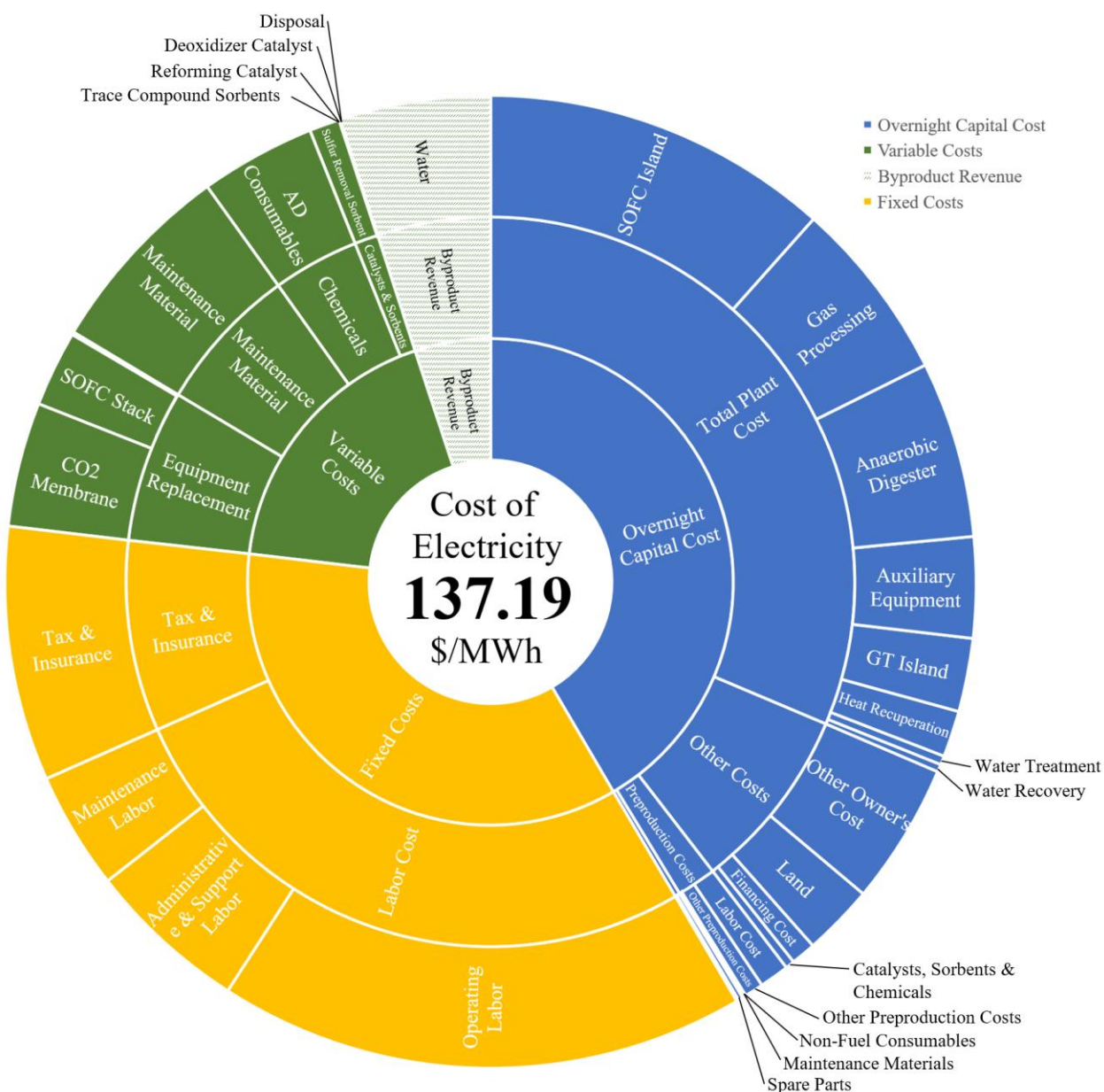


Figure 5-80: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with air-cooled condenser for water recovery.

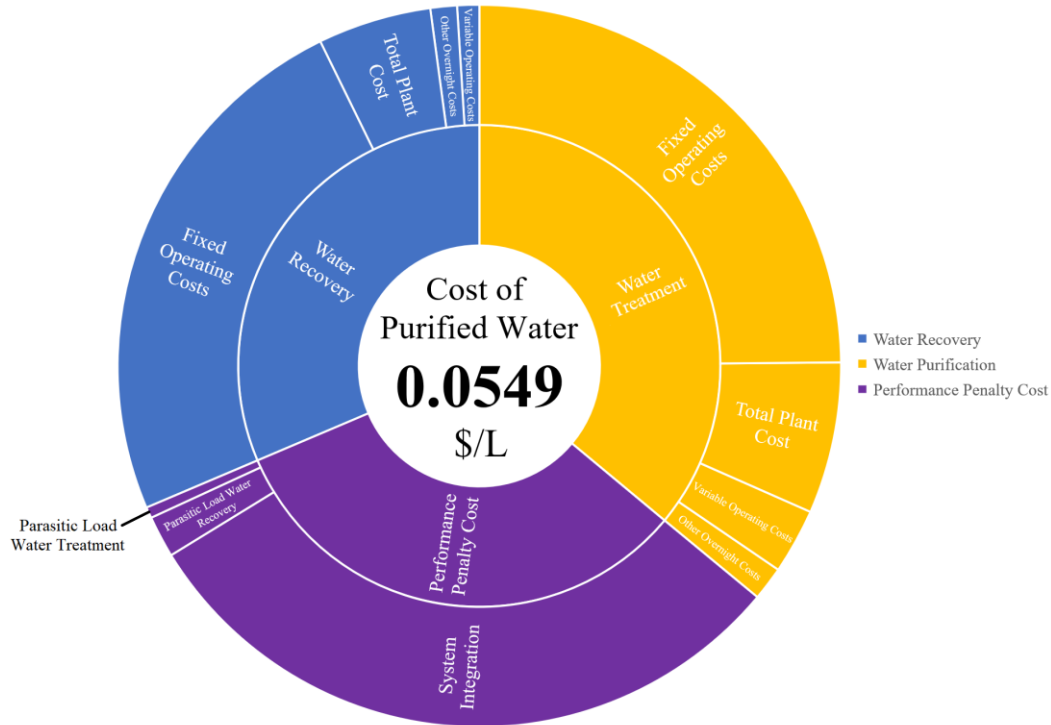


Figure 5-81: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with air-cooled condenser for water recovery.

Direct Contact Condenser

The plant configuration of the 10 MW biogas SOFC-GT hybrid with direct contact condenser is very similar to the air-cooled condenser case. The direct contact condenser is a counterflow column made of polypropylene that is installed downstream of the digester heat recuperation exchanger. Due to the anaerobic digester integration, the flue gas arrives at the direct contact condenser at sufficiently low temperatures so that no water spray is needed to cool the flue gas before it enters the condenser column. The column uses a structured bed to create a large surface area inside the column to allow for effective heat and mass transfer. The advantage of using a structured packing is its low pressure drop. In a direct contact condenser, there is no barrier (e.g., metal tube) between the hot fluid and the liquid coolant allowing for lower pinch temperatures compared to the air-cooled condenser design. Nonetheless, the sensitive heat and latent heat from the flue gas needs to be rejected and the recycled water/coolant temperature is limited by the ambient air temperature. Consequently, the flue gas outlet temperature in the direct contact condenser is higher than in the air-cooled condenser case. This is beneficial with respect to plume rise since it increases buoyancy forces; however, the flue gas still leaves the plant at a saturated state, meaning that fog formation is possible. Since the direct contact condenser works on the principle of gas dew point condensation, the higher outlet temperature results in a reduction of water recovery. Furthermore, in the column, the water is

in direct contact with the flue gas, which allows trace components to dissolve in the water, making it necessary to treat the water adequately before it can be used in society. For the water treatment, the same strategy as in the air-cooled condenser case is used, consisting of an ultra-filtration, reverse osmosis and pH-adjustment step. The flowsheet of the direct condenser case can be found in **Figure 5-82**. The accompanying state-point stream data as indicated in **Figure 5-82** can be found in **Table 5-44**.

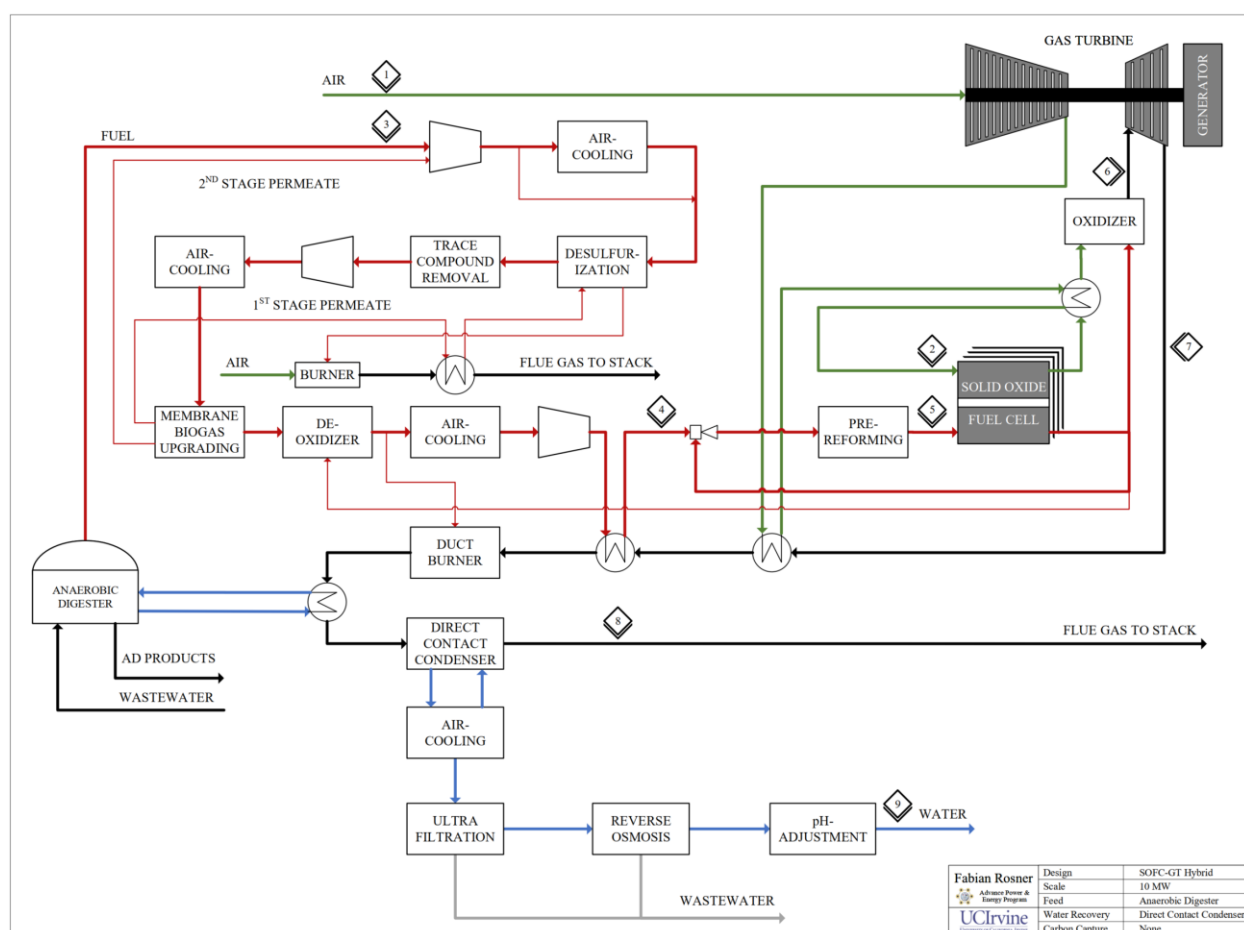


Figure 5-82: Flowsheet of the 10 MW biogas SOFC-GT hybrid system with direct contact condenser for water recovery.

Table 5-44: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	903	559	45	41
Pressure	bar	1.01	7.18	1.01	24.97	7.18	6.58	1.09	1.01	2.00
Molar Flow Rate	kmol/h	1,075.7	1,075.7	135.0	77.7	333.3	1,141.8	1,141.8	1,100.6	44.8
Mass Flow Rate	kg/h	31,042	31,042	3,581	1,417	7,039	32,159	32,159	31,418	807
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07435	0.07435	0.06532	Trace
N ₂	-	0.77315	0.77315	0.00966	0.01618	0.00774	0.72948	0.72948	0.75686	0.00066
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00871	0.00871	0.00903	0.00002
H ₂	-	0.00000	0.00000	0.00000	0.00271	0.19476	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00829	0.07871	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06843	0.21873	0.06258	0.06258	0.07132	Trace
H ₂ O	-	0.00987	0.00987	0.03381	0.01161	0.33349	0.12488	0.12488	0.09747	0.99928
CH ₄	-	0.00000	0.00000	0.58439	0.89269	0.16656	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1	0.99996

Thermodynamic Performance. The power output of the nominal 10 MW biogas hybrid plant with direct contact condenser for water recovery is 9,319 kW, whereby, the SOFC produces a net power of 10,809 kW_{DC} (88.8% FU) and the GT 1,290 kW. Due to the lower backpressure of the GT associated with the lower pressure drop of the direct contact condenser, the GT produces 112 kW more than in the air-cooled condenser case. The change in back pressure results in a SOFC operating pressure of 7.18 bar and a GT surge margin of 11.3%. The general implications of changes in SOFC operating pressure have been discussed for the 10MW NG and biogas air-cooled condenser cases and also apply here. The auxiliary load accounts for 2,779 kW and the common plant equipment (non-water recovery and water treatment) scales with the plant size. The auxiliary load for the water recovery system is significantly higher than in the air-cooled condenser case and is related to the water pumping power of the cooling water recirculation. On the other hand, water treatment load is reduced by 54.3% due to the lower water recovery; however, this auxiliary load reduction is very small compared to the total auxiliary load on an absolute basis.

The efficiency of the direct contact condenser case is 52.93%-LHV. This is 0.14% points lower than the reference case. While the GT power output in this case is higher than in the air-cooled condenser case and closer to the GT power output from the reference case, the higher power demand of the direct contact condenser water recovery system offsets some of these benefits. The GT efficiency in this scenario is 57.7%-LHV and the SOFC stack efficiency is 82.1%-LHV.

The negative CO₂ emission potential for this hybrid plant is 5,638 kg/h or 605 kg/MWh. The water recovery from the flue gas using the direct contact condenser is 31.1%, which is slightly higher than in the NG case due to a higher water partial pressure in the exhaust gas in the biogas case. At the scale of this

nominal 10 MW plant, a total of 19,352 kg of water can be recovered per day. Thus, the hybrid plant with direct contact condenser water recovery system can produce enough water for the irrigation of over 10,000 m² of farmland (including 90% capacity factor, 1.74 l/m² of irrigation water per day [341]). Water condensation after biogas compression and intercooling, as well as wastewater created during water treatment results in a liquid waste stream of 267 kg/h, which is disposed of at the on-site wastewater treatment plant. The plant performance data are summarized in **Table 5-45**.

Table 5-45: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Gross Power Generation	Unit	Direct Contact Condenser
Fuel Cell	kW _{DC}	10,808.71
Gas Turbine	kW	1,289.84
Total Generation	kW	12,098.56
Auxiliary Load		
DC-AC Inverter Losses	kW	324.26
Anaerobic Digester	kW	1499.02
Fuel Gas Compression	kW	797.01
Fuel Gas Intercooler	kW	0.26
Regen. Sulfur Removal	kW	83.29
Water Recovery System	kW	20.91
Water Treatment	kW	1.18
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	7.85
Transformer Losses	kW	45.49
Total Auxiliary Load	kW	2779.28
Net Power Generation	kW	9,319.28
Plant Performance		
Net Efficiency	%-HHV	47.51
Net Efficiency	%-LHV	52.93
Consumables		
Fuel Gas	kg/h	3,581
Thermal Input	MJ _{LHV} /h	63,390
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Net Carbon Emission (Absolute)	kg/h	0 (5,638)
Net Carbon Emission (Absolute)	kg/MWh	0 (605)
Liquid Waste	kg/h	267
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	31.1
Flue Gas Water Recovery	kg/d	19,352

Economics. The TPC of the 10 MW biogas hybrid with direct contact condenser for water recovery is \$47,525,420, which corresponds to a specific plant cost of 5,100 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and SOFC operating pressure effects (see previous explanations, for example air-cooled condenser case). The GT island cost remains constant at \$3,651,010 since the same GT is used in all study cases. Other equipment costs such as anaerobic digester, gas processing and flue gas stack scale with the plant size but remain relatively constant. Heat recuperation in the direct contact condenser case is shifted towards the air preheater due to the higher GT backpressure, which allows for more heat recovery from the GT exhaust (compared to the reference case). Thus, the cathode off-gas exchanger cost is reduced. The direct contact condenser water recovery system costs \$943,780 and is cheaper than in the 10 MW NG case with direct contact condenser. The cost of the direct contact condenser column is significantly impacted by the gas volume flow. In the biogas case the inlet temperature is lower than in the NG case reducing its volumetric flow rate. Furthermore, the water spray in the biogas case is eliminated, reducing the gas flow rate by another 13.0%. The water treatment equipment cost is \$384,420 and is lower than in the air-cooled condenser case due to less water recovered. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-83**. A more detailed cost summary of all plant equipment is provided in **Table 5-46**.

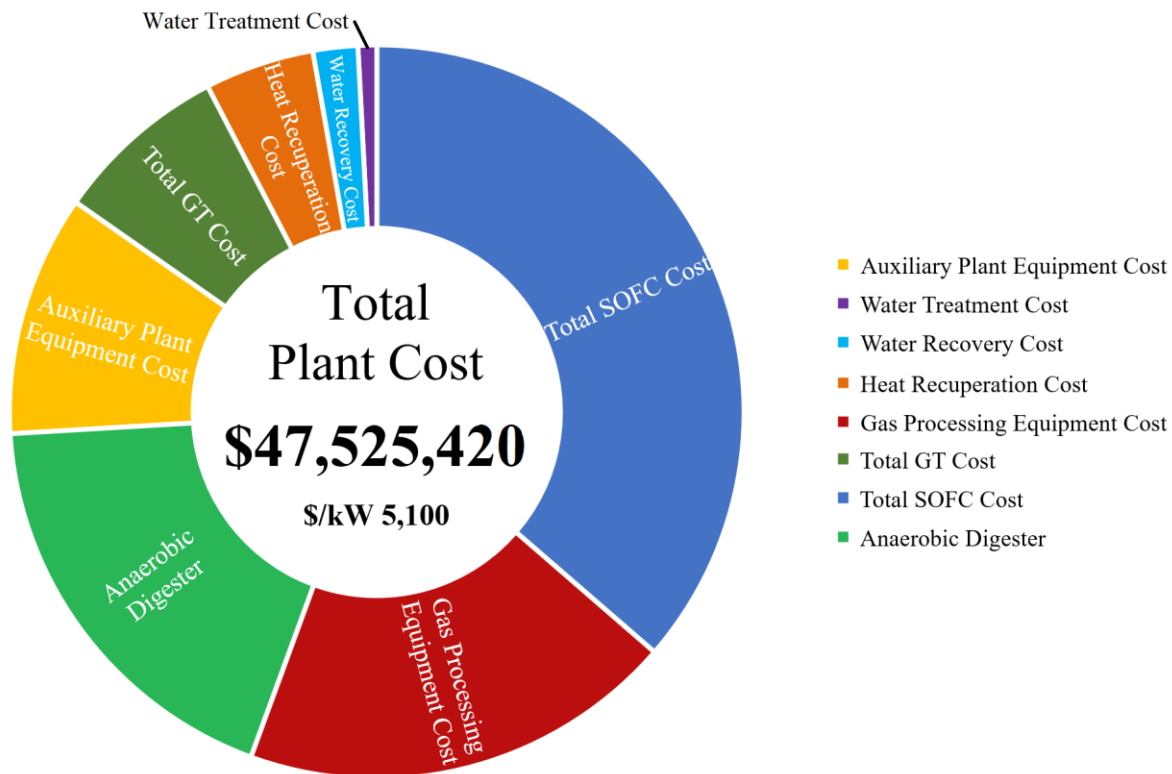


Figure 5-83: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with direct contact condenser for water recovery.

Table 5-46: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Equipment Cost	Unit	Direct Contact Condenser
Fuel Cell Island	\$	17,273,790
SOFC Stacks	\$	6,818,290
Power Conditioning	\$	6,309,150
Pressure Vessel, Housing & Assembly	\$	3,539,300
System Installation	\$	607,040
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Anaerobic Digester	\$	8,787,520
Anaerobic Digester	\$	8,046,720
AD Heat Supply	\$	740,800
Gas Processing and Flue Gas Stack	\$	9,133,650
Fuel Compressors and Intercooler	\$	2,883,810
Fuel Processing	\$	6,119,690
Flue Gas Stack System	\$	130,160
Heat Recuperation	\$	2,295,170
Cathode Off-gas Heat Exchanger	\$	997,750
Air Preheater	\$	1,124,820
Fuel Preheater	\$	172,600
Water Recovery	\$	1,328,200
Water Recovery System	\$	943,780
Water Treatment	\$	384,420
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	5,056,080
Accessory Electric Plant	\$	1,161,390
Instrumentation & Controls	\$	1,745,230
Improvement to Site	\$	453,760
Buildings & Structures	\$	1,695,700
Total Plant Cost	\$	47,525,420
Specific Plant Cost	\$/kW	5,100

The overnight capital cost in the direct contact condenser case is \$62,877,400. The increase of the overnight capital cost compared to the biogas reference case is mostly due to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Other variable operating costs increase due to the water recovery and water treatment equipment. In the direct contact condenser case, water byproduct sales reduce the variable operating costs by \$516,440 per year.

The fixed operating cost in the direct contact condenser case increases due to tax and insurance costs of the additional equipment. The operating labor cost in the air-cooled condenser and direct contact condenser are identical; however, when comparing the labor cost on a relative basis, the labor cost in the direct contact condenser case is significantly higher due to the low water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 137.19 \$/MWh. The COW in the direct contact condenser case is 0.0811 \$/L. A summary of the

economic key parameters can be found in **Table 5-47**. A detailed breakdown of the COE is shown in **Figure 5-84**.

Table 5-47: Economics of the 10 MW Biogas SOFC-GT Hybrid with Direct Contact Condenser for Water Recovery

Item	Unit	Direct Contact Condenser
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	47,525,420
Preproduction Costs		
6 Months Labor Cost	\$	1,484,410
1 Month Maintenance Materials	\$	67,590
1 Month Non-Fuel Consumables	\$	119,780
1 Month Waste Disposal	\$	50
25% of 1 month fuel cost at 100% CF	\$	0
2% of TPC	\$	950,510
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	99,860
0.5% of TPC (Spare Parts)	\$	237,630
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	428,450
Land	\$	3,551,700
Other Owner's Costs (15% of TPC)	\$	7,128,810
Financing Costs	\$	1,283,190
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	2,024,070
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	516,440
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	2,968,830
Property Tax & Insurance	\$	950,510
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0811

Figure 5-84 shows the COE value of 137.19 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-7.03 \$/MWh) and the whole circle represents the COE without byproduct credit (144.22 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 144.22 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0811 \$/L in order to reach the COE of 137.19 \$/MWh.

In **Figure 5-85** it can be seen that the performance penalty cost of the direct contact condenser is significantly lower than in the air-cooled condenser case. While the performance penalty cost in the air-cooled condenser case represents 30.2% of the COW, it is only 10.3% in the direct contact condenser case. The fixed operating costs are still the largest contributor to the COW with over 55.5%, whereby, the operational labor for this analysis is equally split between the water recovery system and the water treatment

SOFC-GT Hybrid Systems with Water Recovery

system. The water recovery alone accounts for 49.7% of the COW (14.7% TPC water recovery). The water treatment costs are again dominated by the fixed operating costs and the TPC of the water treatment is responsible for about 6.0% of the COW. A detailed breakdown of the COW can be found in **Figure 5-85**.

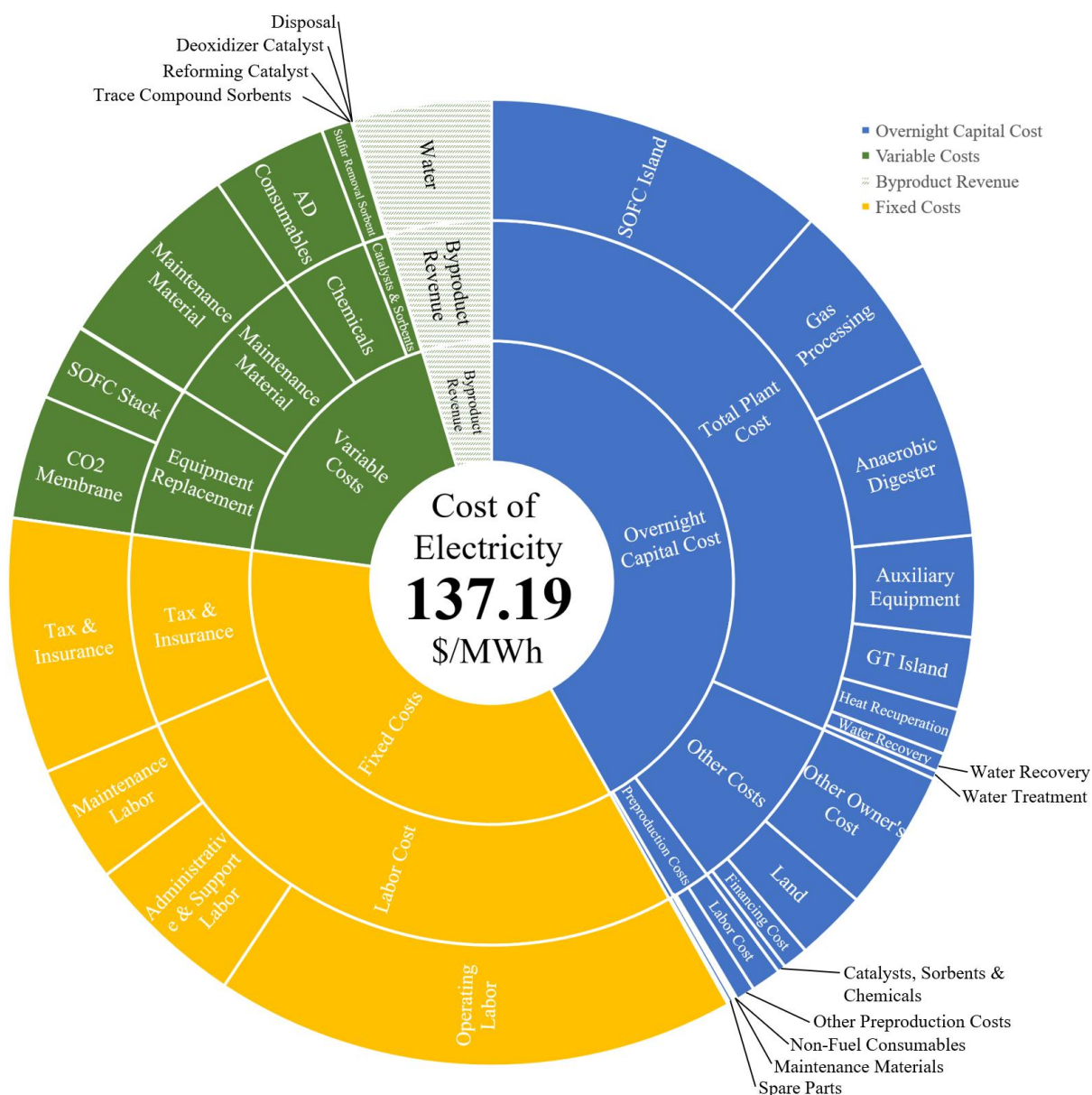


Figure 5-84: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with direct contact condenser for water recovery.

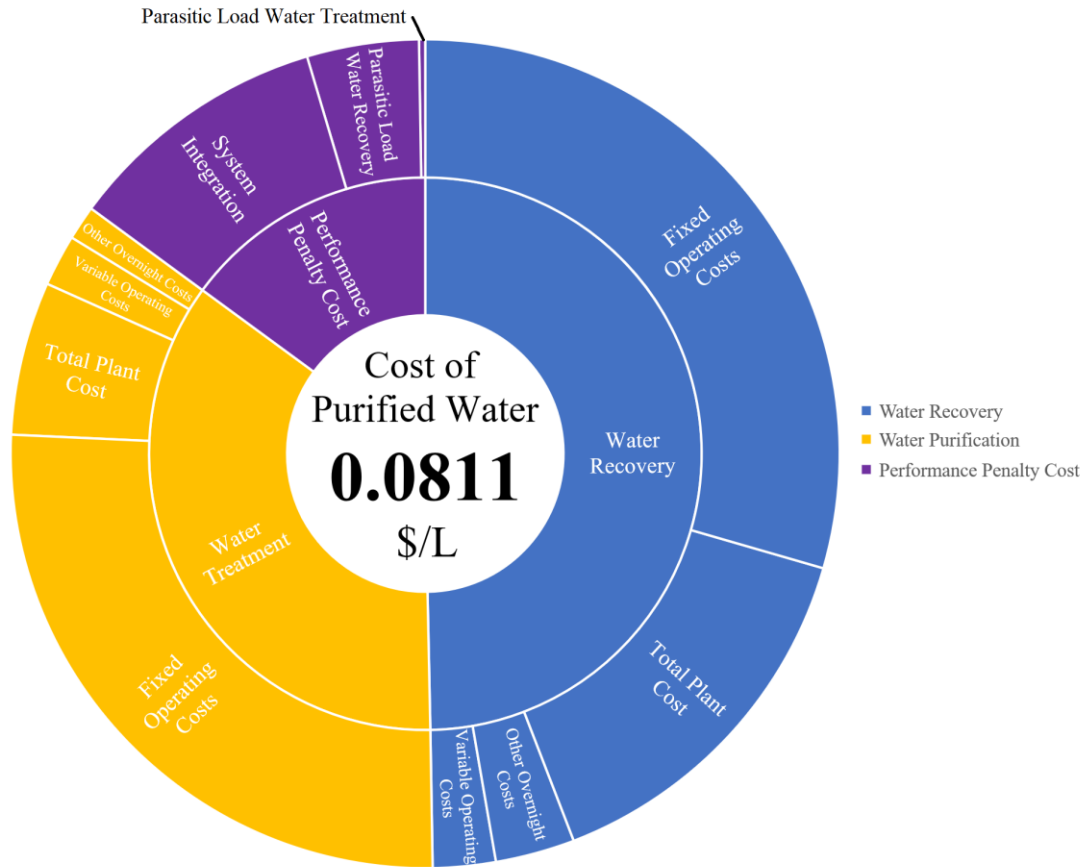


Figure 5-85: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with direct contact condenser for water recovery.

LiBr Absorption System

While the main layout of the SOFC-GT hybrid and its gas processing remain largely identical to the previous biogas plant designs, the heat integration of the anaerobic digester in the LiBr absorber case differs from the biogas cases with air-cooled condenser and direct contact condenser as well as from the natural gas case with LiBr absorber. After the duct burner, the heat exchanger recuperating heat for the anaerobic digester is removed and all the additional heat (heat created by the duct burner) together with the exhaust heat from the GT will be absorbed by the desiccant heater. Thus, the regeneration pressure for the desiccant in the flash regenerator can be increased (equivalent to higher condenser cooling temperature), whilst still achieving the same separation performance. If the flue gas temperature is high enough, the condenser and the regenerated desiccant solution can provide the heat needed in the anaerobic digestion process. In order to maximize the water recovery, the regenerated desiccant is cooled before it is reintroduced into the absorber. The flowsheet of this design can be found in **Figure 5-86**. The corresponding state-point stream data are provided in **Table 5-48**. Alternatively, the regenerated desiccant could be reintroduced into the absorber column without cooling. In this case the water recovery is limited; however, less heating (fuel

flow to the duct burner) is needed for desiccant regeneration and heat recovery can be maximized. In this design the heat recovery is substantially improved, making the duct burner redundant. The flowsheet of the maximum heat recovery case is shown in **Figure 5-87**. State-point stream data are given in **Table 5-49**.

The recovered water in the biogas LiBr cases is cooled against the recirculation medium of the anaerobic digester, which increases the recovered water outlet temperature to 58 °C. This temperature is acceptable with respect to the pH adjustment step; however, needs to be considered in any downstream applications. In the maximum water recovery case, the flue gas leaves the column at a temperature of 49 °C, similarly to the NG case. In the maximum heat recovery case, the flue gas's stack temperature is 71 °C, which is beneficial with respect to plume rise. The dew point temperatures of the two flue gases are 13.5 °C (max. water) and 31.6 °C (max. heat).

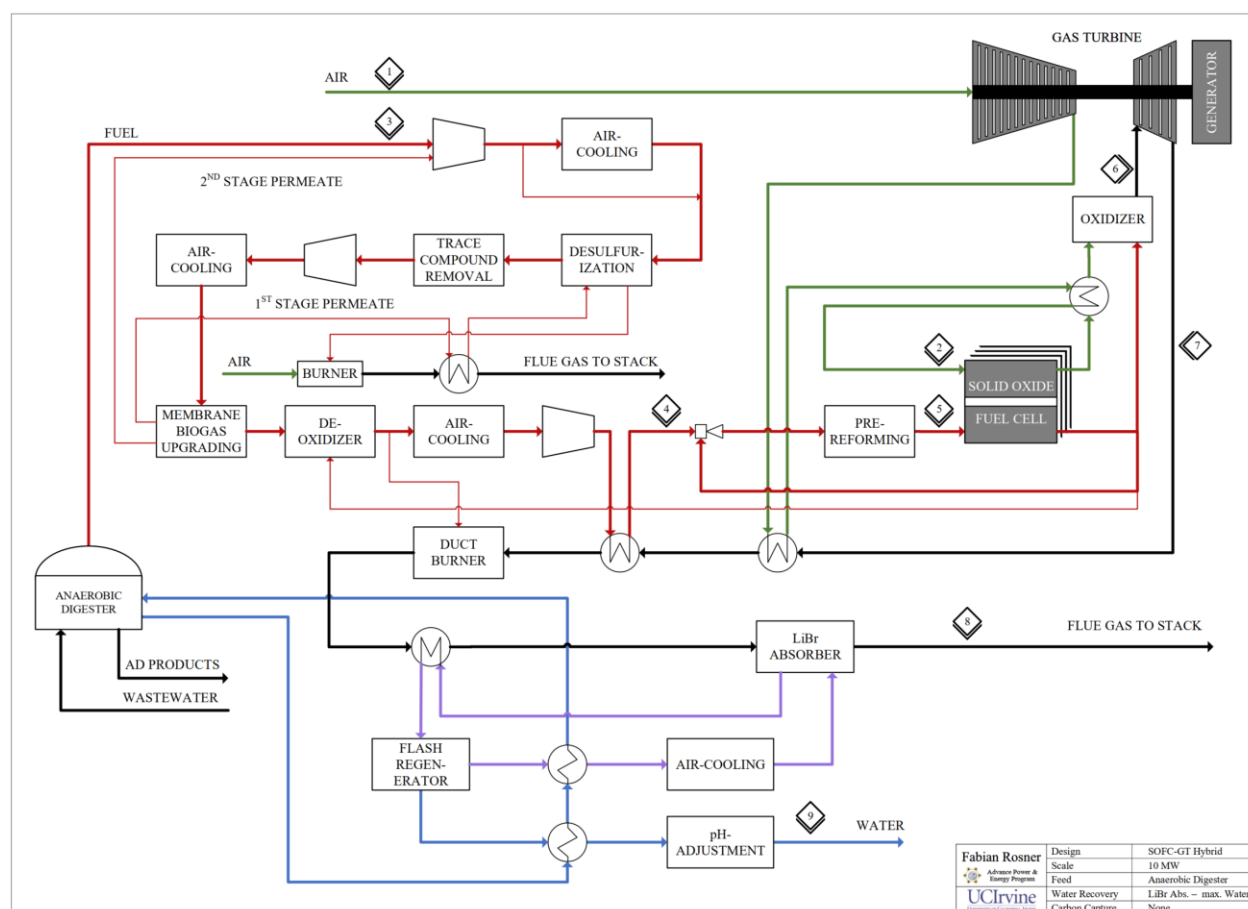


Figure 5-86: Flowsheet of the 10 MW biogas SOFC-GT hybrid system with LiBr absorption system for water recovery designed for maximum water recovery.

Table 5-48: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery Designed for Maximum Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	904	560	49	58
Pressure	bar	1.01	7.19	1.01	25.13	7.19	6.59	1.09	1.01	2.00
Molar Flow Rate	kmol/h	1,075.2	1,075.2	127.4	77.5	332.6	1,141.7	1,141.7	1,012.3	132.0
Mass Flow Rate	kg/h	31,026	31,026	3,379	1,415	7,025	32,159	32,159	29,828	2,378
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07443	0.07443	0.07936	Trace
N ₂	-	0.77315	0.77315	0.00966	0.01618	0.00774	0.72911	0.72911	0.82238	Trace
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00981	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00270	0.19475	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00826	0.07871	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06855	0.21875	0.06269	0.06269	0.07318	Trace
H ₂ O	-	0.00987	0.00987	0.03381	0.01159	0.33344	0.12507	0.12507	0.01526	0.99999
CH ₄	-	0.00000	0.00000	0.58439	0.89262	0.16659	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1	0.99999

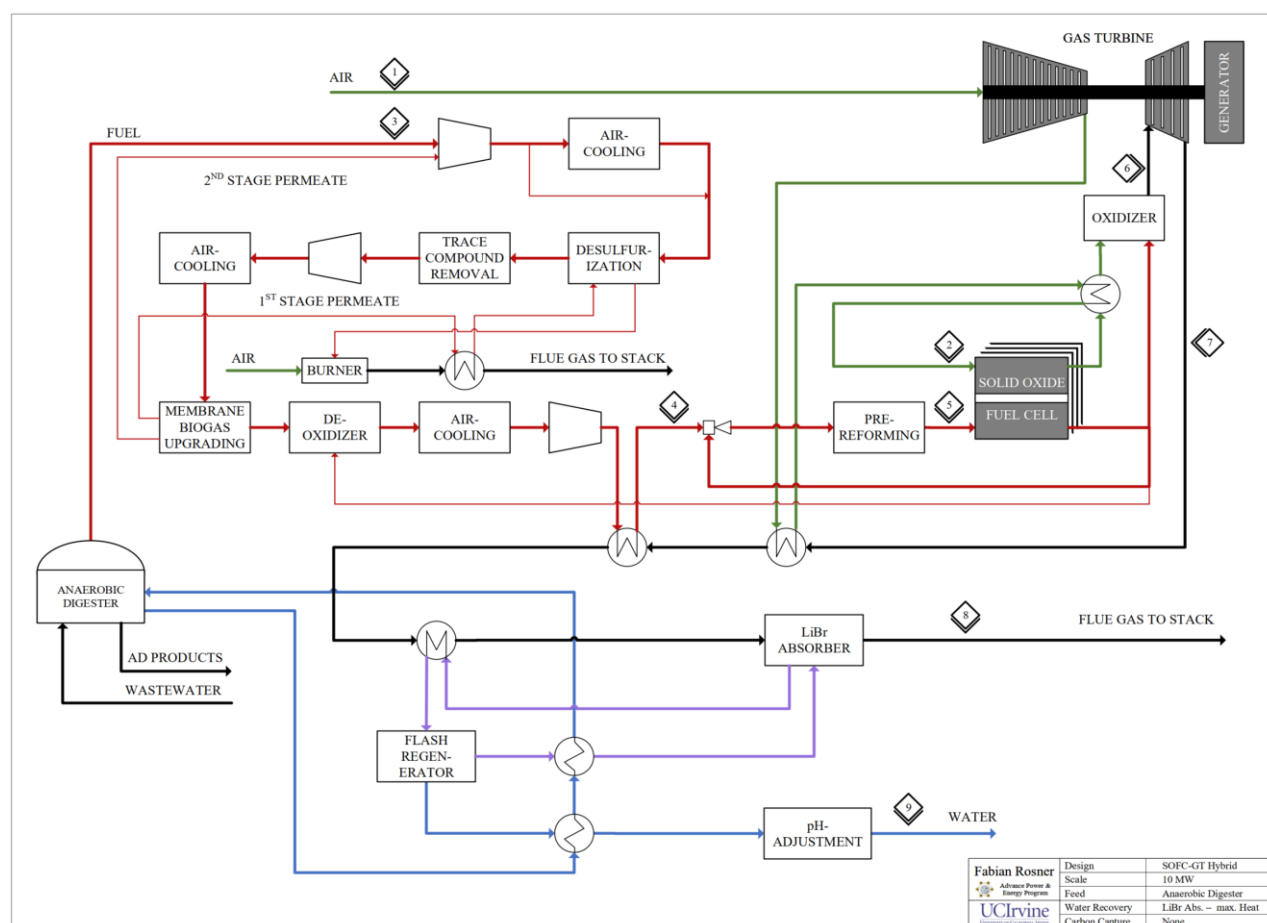
**Figure 5-87:** Flowsheet of the 10 MW biogas SOFC-GT hybrid system with LiBr absorption system for water recovery designed for maximum heat recovery.

Table 5-49: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with LiBr Absorption System for Water Recovery Designed for Maximum Heat Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	904	560	71	58
Pressure	bar	1.01	7.18	1.01	25.07	7.18	6.58	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,075.4	1,075.4	123.3	77.6	332.9	1,142.4	1,142.4	1,047.4	95.0
Mass Flow Rate	kg/h	31,033	31,033	3,271	1,415	7,032	32,176	32,176	30,465	1,711
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07434	0.07434	0.08109	Trace
N ₂	-	0.77315	0.77315	0.00966	0.01617	0.00774	0.72887	0.72887	0.79497	Trace
AR	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00949	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00269	0.19488	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00825	0.07877	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06850	0.21869	0.06280	0.06280	0.06850	Trace
H ₂ O	-	0.00987	0.00987	0.03381	0.01160	0.33340	0.12529	0.12529	0.04596	0.99999
CH ₄	-	0.00000	0.00000	0.58439	0.89268	0.16649	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1	0.99999

Thermodynamic Performance. The net power output of the two configurations with LiBr absorber is very similar with 9,424 kW (max. water) and 9,498 kW (max. heat). The fuel cell island produces 10,790 kW_{DC} (max. water, FU 88.7%) and 10,791 kW_{DC} (max. heat, FU 88.7%). The max. water recovery case has a slightly lower GT power output with 1,292 kW then the max. heat recovery case with 1,301 kW, due to GT backpressure effects associated with the duct burner as well as small changes in the TIT that counteract some of the backpressure effects since it is a recuperated GT design. The GT surge margins in the LiBr absorber cases are 11.1% (max. water) and 11.3% (max. heat). The auxiliary load differs in three key areas from each other: I) the anaerobic digester consumption is reduced in the max. heat case due to lower fuel requirement II) the lower fuel requirement in the max. heat case reduces the power consumption of fuel gas compression and III) the lower water recovery in the max. heat case reduces the power consumption of the water recovery system. As a result, the auxiliary load in the max. heat case is with 2,594 kW about 64 kW lower than in the max. water case.

With the anaerobic digester being integrated into the water recovery system, latent heat from the flue gas can be upgraded and made available to the digester. This is fundamentally different compared to the air cooled-condenser and direct contact condenser cases where the latent heat is released at the flue gas's dew point temperature, which is too low for integration with the anaerobic digester, and the latent heat needs to be rejected to the environment in order to recover the water. As a result of the heat integration in the LiBr absorption cases, the plant efficiency increases instead of going down, as seen in the previous cases. The respective electrical efficiencies of the LiBr absorption cases with max. water recovery and max. heat recovery are 56.72%-LHV and 59.04%-LHV. This is a substantial efficiency increase versus the reference

case with 53.07%-LHV. The GT efficiency and SOFC stack efficiency in the LiBr cases remain virtually constant compared to the reference case, and efficiency gains arise from a lower auxiliary load and lower fuel input (reduction in flue flow to the duct burner).

The carbon negative emission potentials for these LiBr absorber cases are 5,320 kg/h (565 kg/MWh) in the max. water scenario and 5,154 kg/h (543 kg/MWh) in the max. heat scenario. The water recovery from the flue gas in the LiBr absorber case with max. water recovery is 89.5% and 66.4% in the max. heat recovery scenario. At the scale of this nominal 10 MW plant, a total of 57,056 kg of water can be recovered per day, which also creates 189 kg/h of liquid waste (max. water). Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW biogas SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 29,500 m² of farm land (including 90% capacity factor). In the max. heat recovery scenario, the amount of water recovered is reduced to 41,077 kg/day. The plant performance data of both configurations are summarized in **Table 5-50**.

Table 5-50: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrids with LiBr Absorption System for Water Recovery

Gross Power Generation		Unit	LiBr Absorber - max. Water	LiBr Absorber - max. Heat
	Fuel Cell	kW _{DC}	10,789.62	10,791.06
	Gas Turbine	kW	1,292.34	1,300.73
Total Generation		kW	12,081.96	12,091.79
Auxiliary Load				
	DC-AC Inverter Losses	kW	323.69	323.73
	Anaerobic Digester	kW	1399.74	1361.60
	Fuel Gas Compression	kW	757.27	739.15
	Fuel Gas Intercooler	kW	0.24	0.23
	Regen. Sulfur Removal	kW	77.78	75.50
	Water Recovery System	kW	45.39	40.06
	Water Treatment	kW	0.68	0.49
	Zero Liquid Discharge System	kW	N/A	N/A
	Miscellaneous Balance of Plant	kW	7.84	7.85
	Transformer Losses	kW	45.43	45.47
Total Auxiliary Load		kW	2658.07	2594.09
Net Power Generation		kW	9,423.90	9,497.70
Plant Performance				
	Net Efficiency	%-HHV	50.92	53.01
	Net Efficiency	%-LHV	56.72	59.04
Consumables				
	Fuel Gas	kg/h	3,379	3,271
	Thermal Input	MJ _{LHV} /h	59,818	57,912
	Water	kg/h	0	0
Emissions and Waste Streams				
	Carbon Capture	%	N/A	N/A
	Net Carbon Emission (Absolute)	kg/h	0 (5,320)	0 (5,154)
	Net Carbon Emission (Absolute)	kg/MWh	0 (565)	0 (543)
	Liquid Waste	kg/h	189	183
	Solid Waste	g/h	0	0
Water Recovery				
	Flue Gas Water Recovery	%	89.5	66.4
	Flue Gas Water Recovery	kg/d	57,056	41,077

Economics. The TPCs of the 10 MW biogas hybrids with LiBr absorber for water recovery are \$47,786,050 (max. water) and \$47,094,020 (max. heat) resulting in specific plant costs of 5,071 \$/kW (max. water) and 4.958 \$/kW (max. heat). The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The GT island cost remains constant at \$3,651,010 since the same GT is used in all study cases. The anaerobic digester costs have been substantially reduced in the LiBr cases. In the reference case, the digester cost is with \$8,810,340 almost \$900,000 more expensive than in the max. heat case. The higher efficiency is not only beneficial with respect to the operating costs but also reduces the fuel demand and the size of the digester. Similarly, the fuel processing cost is reduced. Ultimately, this helps to reduce the capital cost and offset some of the expenses associated with the water recovery. The water recovery equipment in the max. heat recovery case is about \$245,000 cheaper than in the max. water recovery case, which is primarily due to the size reduction of the desiccant heat exchanger upstream of the absorber column and the elimination of the air-cooled desiccant cooler downstream of the regeneration step. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-88** (max. water) and **Figure 5-89** (max. heat). A more detailed cost summary of all plant equipment is provided in **Table 5-51**.

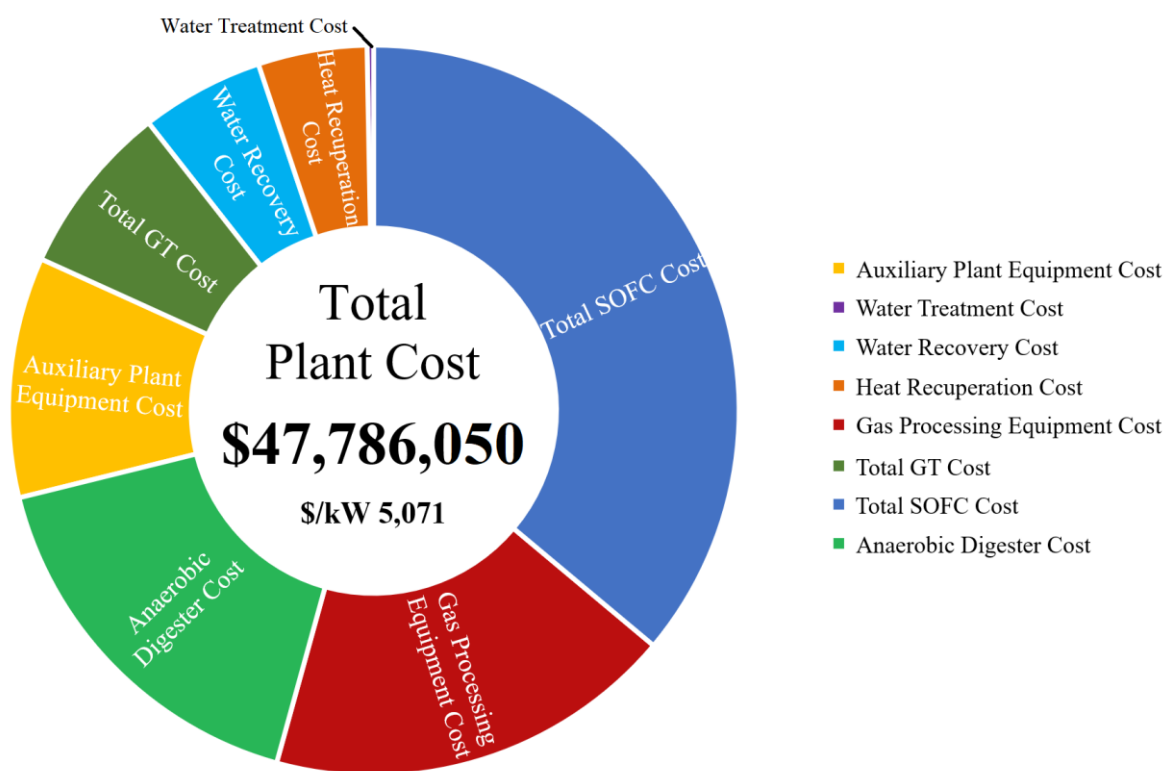


Figure 5-88: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with LiBr absorption system for water recovery designed for maximum water recovery.

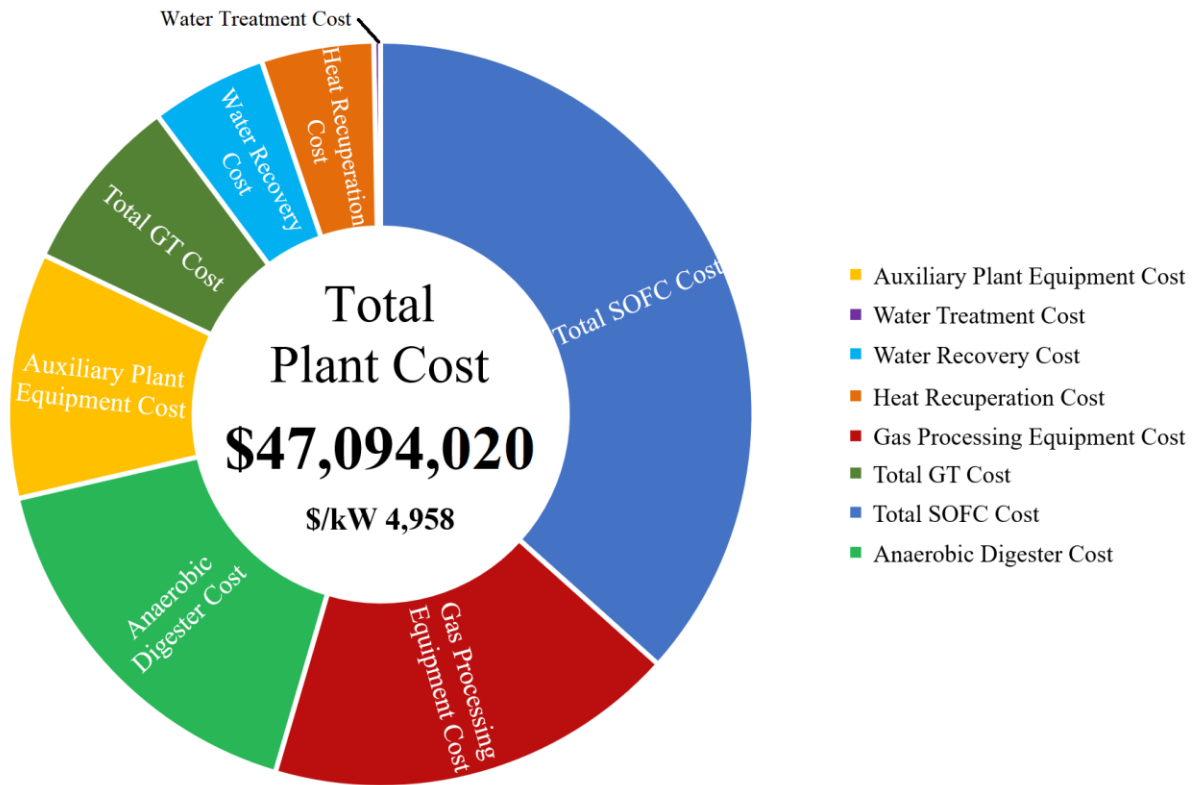


Figure 5-89: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with LiBr absorption system for water recovery designed for maximum heat recovery.

The overnight capital cost in the LiBr absorber case designed for max. water recovery is \$63,377,290, an increase of over 3.7% compared to the reference case. The LiBr case designed for max. heat recovery has a lower overnight capital cost of \$62,494,780. Variable operating costs decrease compared to the reference case mostly due to a reduction in CO₂ membrane costs (part of fuel processing) and anaerobic digester consumables. The 55 wt-% LiBr solution is assumed to cost 2.04 \$₂₀₁₅/kg and the initial fill costs about \$390,150 (max. water) and \$469,670 (max. heat). The LiBr loss is very small and entirely due to entrainment since LiBr is a salt with extremely low vapor pressure. With a mist eliminator, the LiBr losses are expected to be around 10 kg per year. In the LiBr absorber case, water byproduct sales reduce the variable operating costs by \$398,140 per year in the max. water recovery scenario and by \$187,420 in the max. heat recovery scenario.

The fixed operating cost in the LiBr absorber case increases due to tax and insurance costs of the additional equipment (compared to reference case). The operating labor costs in the LiBr absorber case are identical to the air-cooled condenser and direct contact condenser cases; however, when comparing the labor cost on a relative basis, the labor cost in the LiBr absorber case is significantly lower due to the increased water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 137.19 \$/MWh. The

Table 5-51: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrids with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber - max. Water	LiBr Absorber - max. Heat
Fuel Cell Island	\$	17,263,640	17,268,270
SOFC Stacks	\$	6,818,680	6,818,490
Power Conditioning	\$	6,300,610	6,303,010
Pressure Vessel, Housing & Assembly	\$	3,537,660	3,539,990
System Installation	\$	606,690	606,780
Gas Turbine Island	\$	3,651,010	3,651,010
Gas Turbine-Generator Set	\$	3,077,600	3,077,600
Gas Turbine Foundation	\$	44,740	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680	528,680
Anaerobic Digester	\$	8,067,650	7,910,550
Anaerobic Digester	\$	7,769,220	7,618,500
AD Heat Supply	\$	298,430	292,050
Gas Processing and Flue Gas Stack	\$	8,676,230	8,435,610
Fuel Compressors and Intercooler	\$	2,678,430	2,581,760
Fuel Processing	\$	5,867,640	5,723,650
Flue Gas Stack System	\$	130,160	130,200
Heat Recuperation	\$	2,292,450	2,294,100
Cathode Off-gas Heat Exchanger	\$	993,620	998,320
Air Preheater	\$	1,127,140	1,123,780
Fuel Preheater	\$	171,690	171,990
Water Recovery	\$	2,759,230	2,514,410
Water Recovery System	\$	2,609,590	2,371,610
Water Treatment	\$	149,630	142,800
Zero Liquid Discharge System	\$	N/A	N/A
Auxiliary Plant Equipment Cost	\$	5,075,840	5,020,070
Accessory Electric Plant	\$	1,159,790	1,160,740
Instrumentation & Controls	\$	1,754,800	1,729,390
Improvement to Site	\$	456,250	449,640
Buildings & Structures	\$	1,705,000	1,680,310
Total Plant Cost	\$	47,786,050	47,094,020
Specific Plant Cost	\$/kW	5,071	4,958

COW in the LiBr absorber case with max. water recovery is 0.0212 \$/L and 0.0139 \$/L in the max. heat recovery case, indicating that heat recovery is more valuable for the plant operating than water recovery. A summary of the economic key parameters can be found in **Table 5-52**. Detailed breakdowns of the COEs for the two scenarios are shown in **Figure 5-90** (max. water) and **Figure 5-91** (max. heat).

Figure 5-90 and **Figure 5-91** show the COE value of 137.19 \$/MWh in solid colors and the byproduct revenue is shown with striped background, which amounts to -5.36 \$/MWh for the max. water recovery case and to -2.50 \$/MWh in the max. heat recovery case. The whole circle represents the COE without byproduct credit 142.55 \$/MWh in the max. water recovery case and 139.69 \$/MWh in the max. heat recovery case. This means if the water would be provided to the local community at no cost, the COE would increase to 142.55 \$/MWh and 139.69 \$/MWh respectively, due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0212 \$/L and 0.0139 \$/L in order to reach a COE of 137.19 \$/MWh. These

Table 5-52: Economics of the 10 MW Biogas SOFC-GT Hybrids with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber - max. Water	LiBr Absorber - max. Heat
Plant Capacity Factor	%	90	90
Total Overnight Capital Cost			
Total Plant Cost	\$	47,786,050	47,094,020
Preproduction Costs			
6 Months Labor Cost	\$	1,486,520	1,481,740
1 Month Maintenance Materials	\$	68,110	66,930
1 Month Non-Fuel Consumables	\$	114,480	111,770
1 Month Waste Disposal	\$	40	40
25% of 1 month fuel cost at 100% CF	\$	0	0
2% of TPC	\$	955,720	941,880
Inventory Capital Costs			
60 Day Supply Consumables at 100% CF	\$	94,240	91,240
0.5% of TPC (Spare Parts)	\$	238,930	235,470
Other Costs			
Initial Catalyst, Sorbent & Chemicals Cost	\$	794,690	861,440
Land	\$	3,380,370	3,274,610
Other Owner's Costs (15% of TPC)	\$	7,167,910	7,064,100
Financing Costs	\$	1,290,220	1,271,540
Variable Operating Cost for Initial Year of Operation			
Annual Feed Cost at above CF for Initial Year	\$	0	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	1,972,500	1,930,330
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0	0
Annual Byproduct Revenues at above CF for Initial Year	\$	398,140	187,420
Fixed Operating Cost for Initial Year of Operation			
Labor Cost	\$	2,973,040	2,963,470
Property Tax & Insurance	\$	955,720	941,880
Cost of Electricity and Cost of Water			
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A	N/A
First Year Levelized Cost of Water	\$/L	0.0212	0.0139

COWs are substantially lower than in the air-cooled condenser case and direct contact condenser case due to the beneficial heat integration.

Figure 5-92 shows a detailed breakdown of the COW for the max. water recovery case and **Figure 5-93** shows the breakdown of the COW for the max. heat recovery case. For the LiBr absorber cases the total operating labor costs are shown under the water recovery rubric, which is different to the air-cooled condenser case and the direct contact condenser case where the operating labor is split between the water recovery and the water treatment. However, the water treatment in the LiBr absorber case is minor and does not justify the addition of operating labor to its costs. The performance penalty costs in the LiBr absorber cases are 6.8% (max. water) and 6.3% (max heat) and represent only the parasitic loads since the system integration losses are negative (meaning positive effect of water recovery system integration). The fixed operating costs are still the largest contributor to the COW with 46.4% (max. water) and 47.6% (max. heat). The water treatment is responsible for only 4.6% (max. water) and 3.2% (max. heat) of the COW.

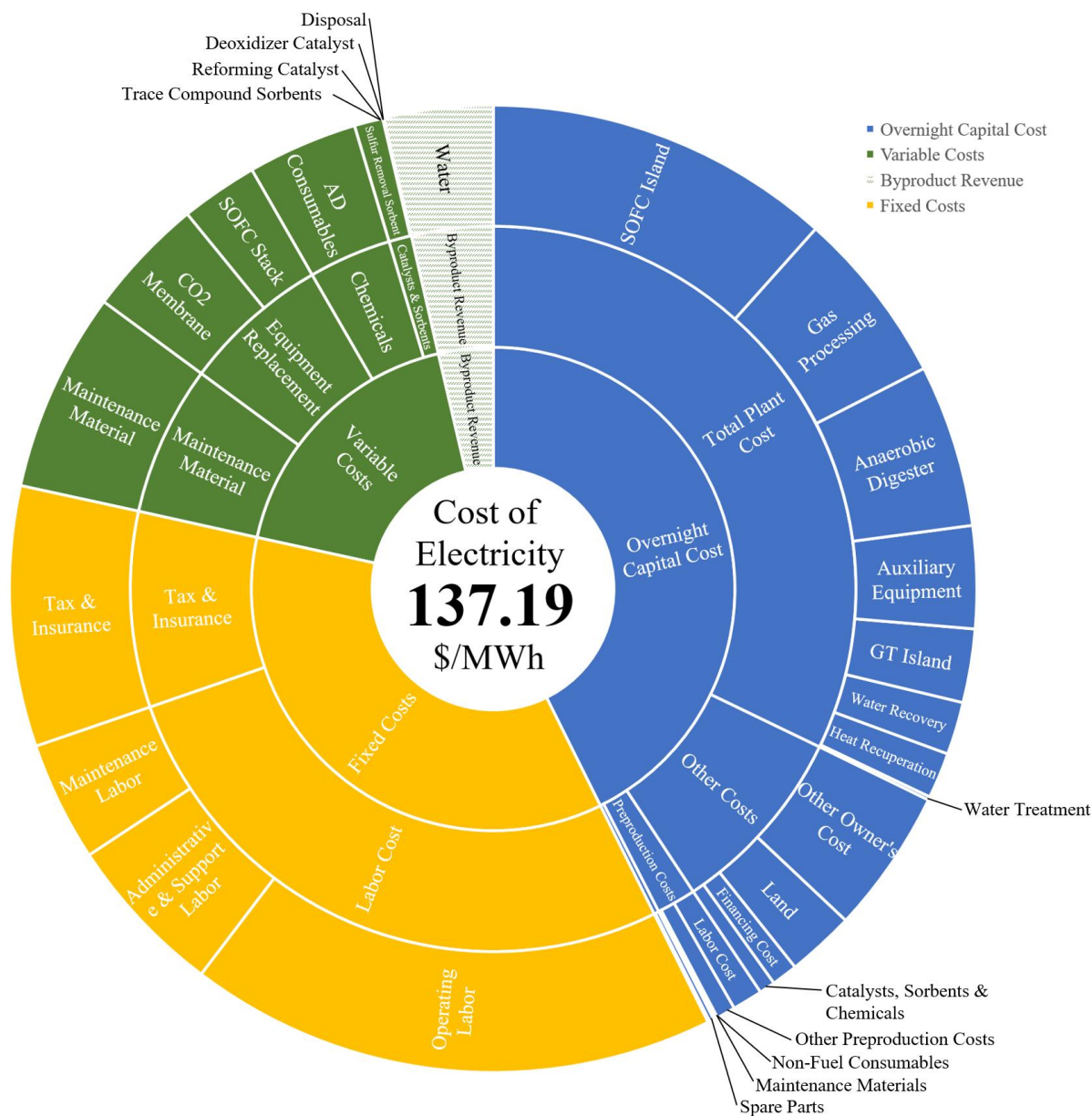


Figure 5-90: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with LiBr absorber for water recovery designed for maximum water recovery.

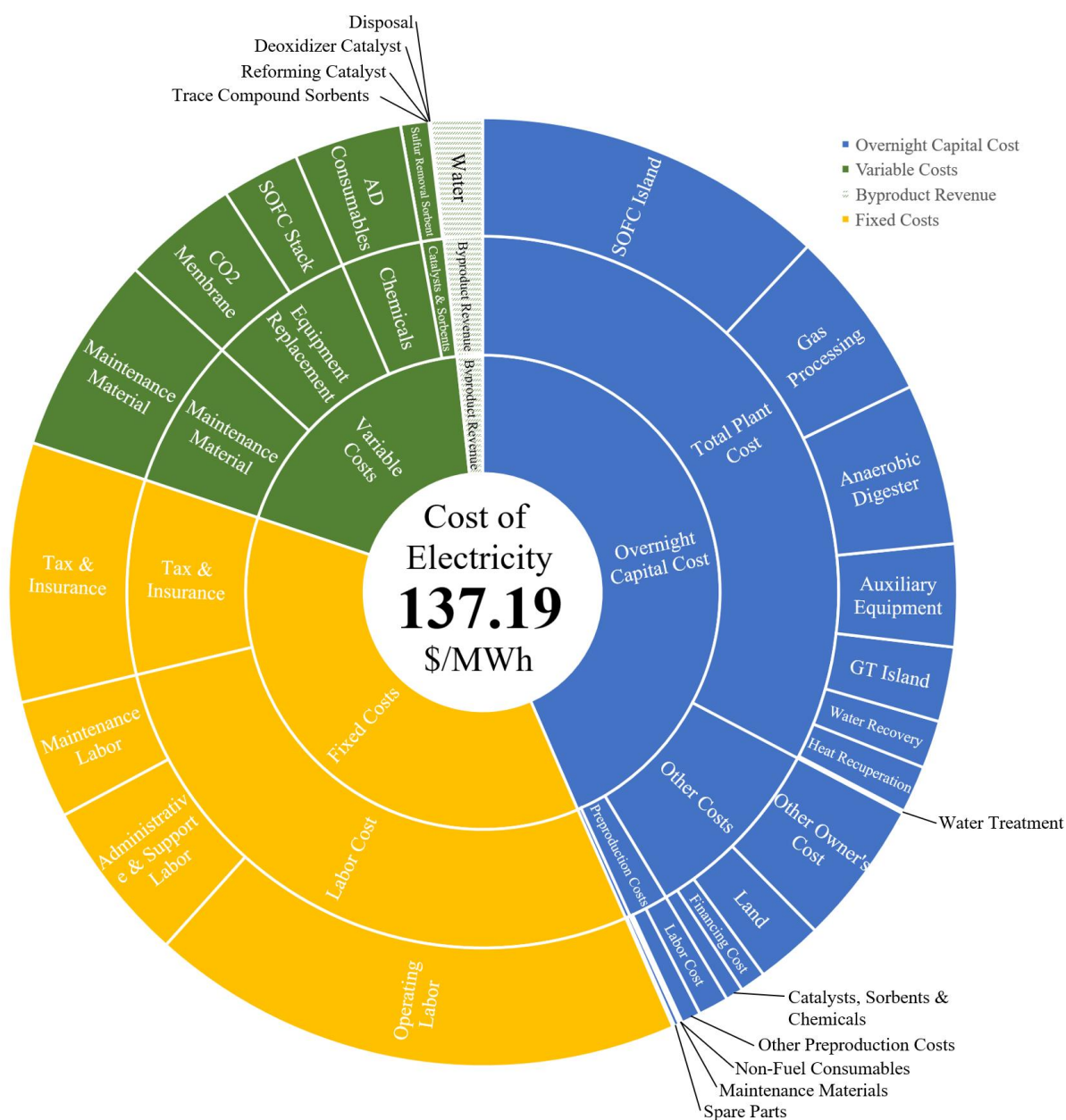


Figure 5-91: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with LiBr absorber for water recovery designed for maximum heat recovery.

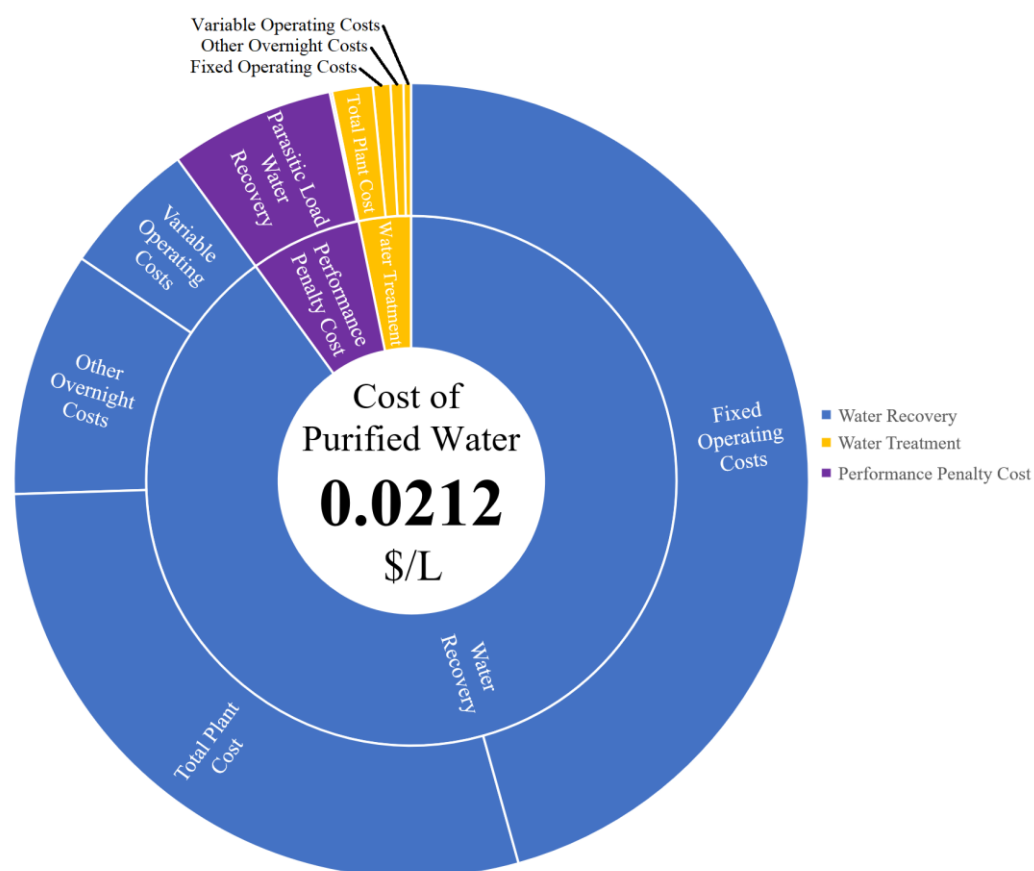


Figure 5-92: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with LiBr absorber for water recovery designed for maximum water recovery.

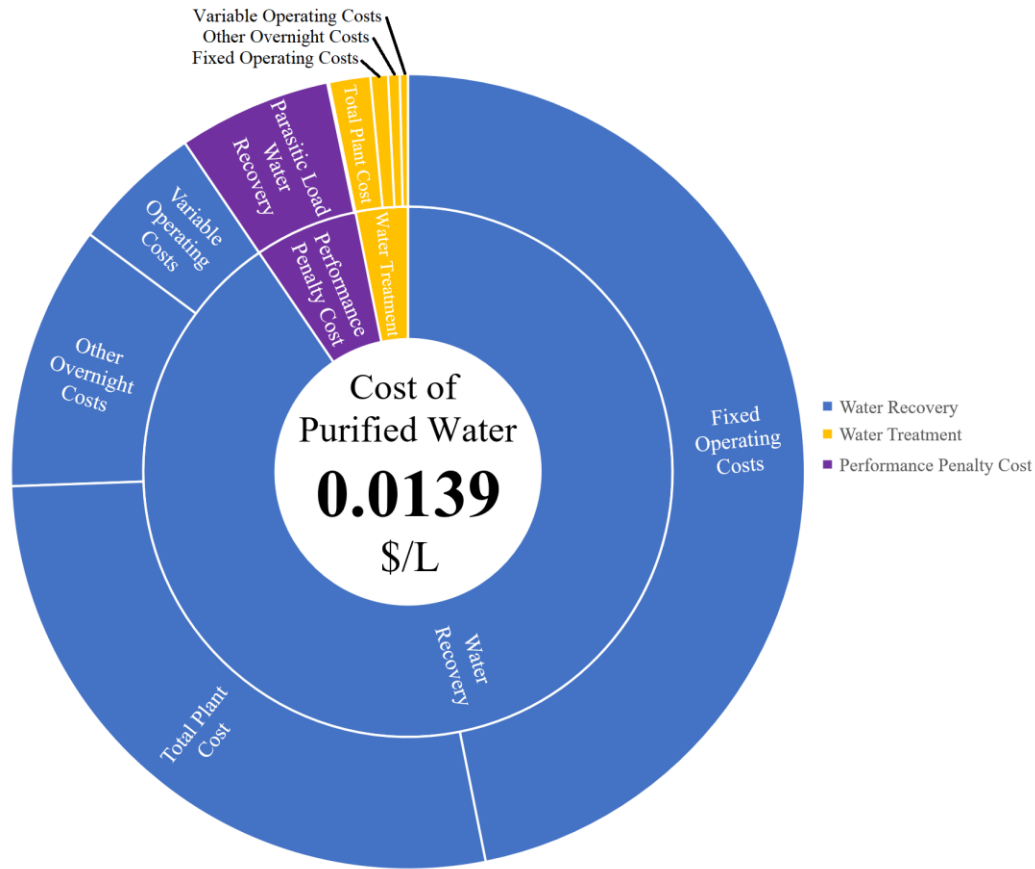


Figure 5-93: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with LiBr absorber for water recovery designed for maximum heat recovery.

Glycol Absorption System

The SOFC-GT hybrid plant designs with MEG absorption system follows the design of the LiBr absorber case. After the duct burner, the heat exchanger used in the air-cooled condenser case and direct contact condenser case for recuperating heat for the anaerobic digester is removed and all the additional heat (heat created by the duct burner) together with the exhaust heat from the GT is recovered by the desiccant heater. It is important to ensure that the hot flue gas entering the desiccant heater does not exceed the decomposition temperature of MEG and sufficient amounts of water are sprayed into the flue gas upstream of the desiccant heat exchanger. In this design, the heat for the digester will be recovered after desiccant regeneration. As a result, the regeneration pressure on the permeate side increases. Again, two configurations are studied: one for maximum water recovery and another one for maximum heat recovery. In order to maximize the water recovery, the regenerated desiccant is cooled before it is reintroduced into the absorber. The flowsheet of this design can be found in **Figure 5-94**. The corresponding state-point stream data are provided in **Table 5-53**. To maximize heat recovery, the regenerated desiccant is reintroduced into the absorber column without cooling. The higher desiccant temperature limits water recovery so that less heating (fuel flow to

SOFC-GT Hybrid Systems with Water Recovery

the duct burner) is needed for desiccant regeneration and heat recovery can be maximized. In this design the heat recovery is substantially improved, however, in the MEG case the duct burner is still needed in the max. heat recovery case. The flowsheet of the maximum heat recovery case using MEG is shown in **Figure 5-95**. State-point stream data are given in **Table 5-54**.

Due to the more intense water treatment required in the MEG absorber case, it is necessary to cool the recovered water to acceptable inlet temperatures for the ultrafiltration unit and reverse osmosis unit. In the maximum water recovery case, the flue gas leaves the column at a temperature of 45 °C, similarly to the NG case. In the maximum heat recovery case, the flue gas's stack temperature is 69 °C, which is beneficial with respect to plume rise. The dew point temperatures of the two flue gases are 21.5 °C (max. water) and 39.8 °C (max. heat).

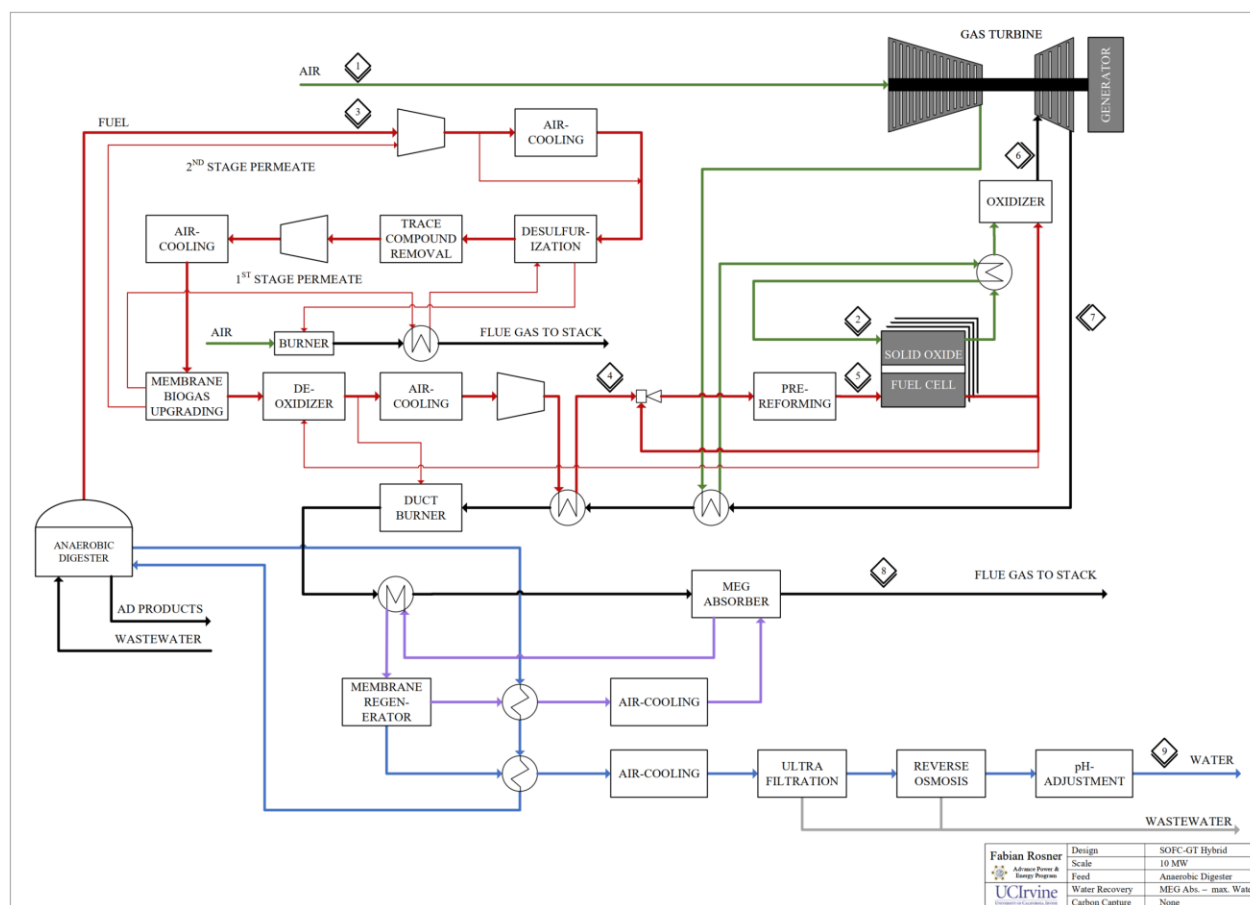


Figure 5-94: Flowsheet of the 10 MW biogas SOFC-GT hybrid system with MEG absorption system for water recovery designed for maximum water recovery.

Table 5-53: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with MEG Absorption System for Water Recovery Designed for Maximum Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	904	560	45	41
Pressure	bar	1.01	7.18	1.01	24.97	7.18	6.58	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,075.6	1,075.6	133.8	77.6	333.2	1,141.7	1,141.7	1,020.1	117.9
Mass Flow Rate	kg/h	31,039	31,039	3,547	1,416	7,037	32,158	32,158	29,989	2,141
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07441	0.07441	0.07181	0.00000
N ₂	-	0.77315	0.77315	0.00966	0.01617	0.00774	0.72943	0.72943	0.81645	0.00000
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00871	0.00871	0.00974	0.00000
H ₂	-	0.00000	0.00000	0.00000	0.00271	0.19493	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00831	0.07881	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06852	0.21868	0.06258	0.06258	0.07625	0.00000
H ₂ O	-	0.00987	0.00987	0.03381	0.01160	0.33334	0.12488	0.12488	0.02529	0.99671
CH ₄	-	0.00000	0.00000	0.58439	0.89259	0.16648	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	0.99954	0.99671

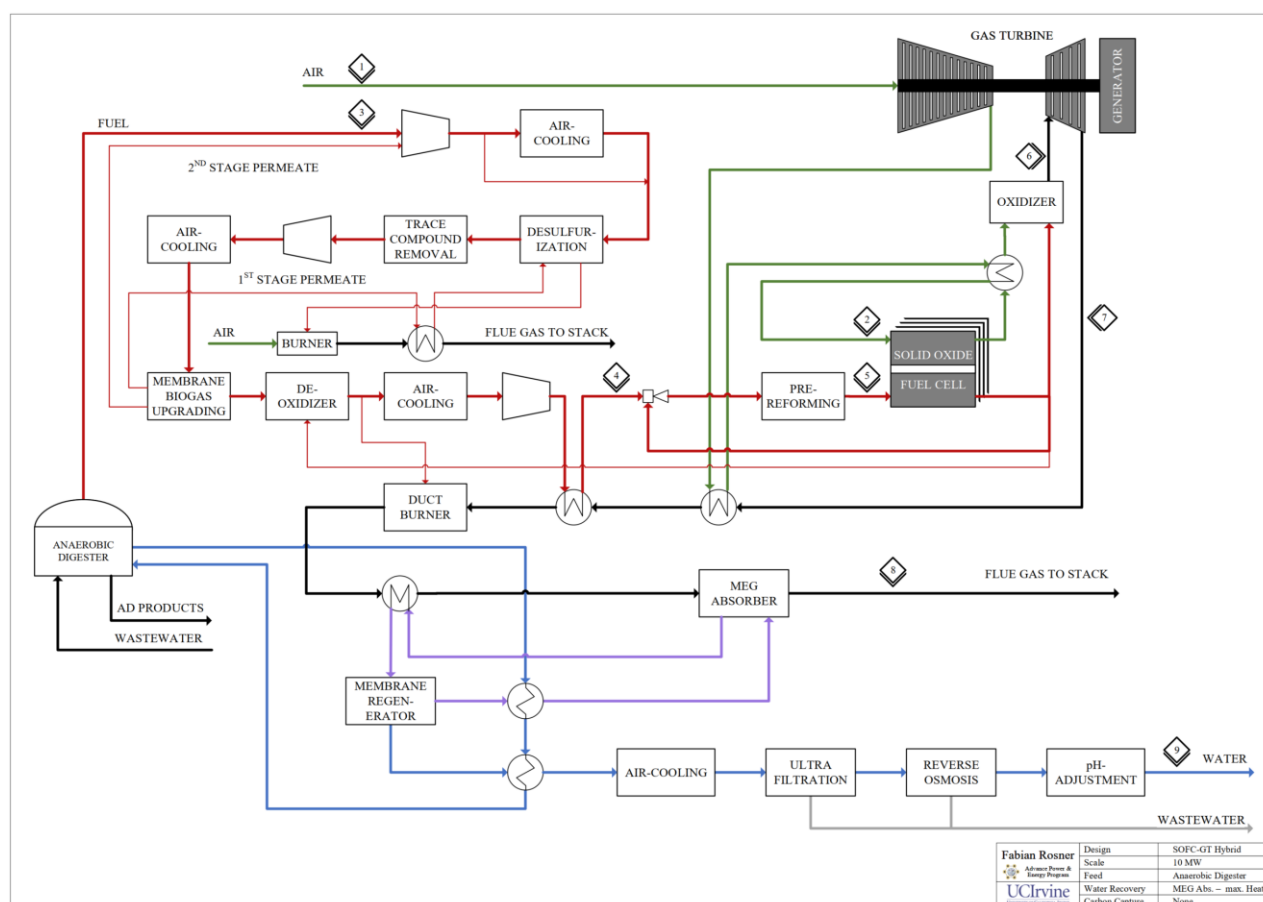
**Figure 5-95:** Flowsheet of the 10 MW biogas SOFC-GT hybrid system with MEG absorption system for water recovery designed for maximum heat recovery.

Table 5-54: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with MEG Absorption System for Water Recovery Designed for Maximum Heat Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	905	560	69	41
Pressure	bar	1.01	7.19	1.01	25.08	7.19	6.58	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,075.3	1,075.3	127.9	77.7	333.5	1,142.0	1,142.0	1,076.1	64.6
Mass Flow Rate	kg/h	31,028	31,028	3,391	1,416	7,041	32,163	32,163	31,064	1,174
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07421	0.07421	0.07418	0.00000
N ₂	-	0.77315	0.77315	0.00966	0.01618	0.00774	0.72903	0.72903	0.77369	0.00000
AR	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00923	0.00000
H ₂	-	0.00000	0.00000	0.00000	0.00270	0.19503	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00826	0.07882	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06829	0.21859	0.06278	0.06278	0.06910	0.00000
H ₂ O	-	0.00987	0.00987	0.03381	0.01159	0.33335	0.12528	0.12528	0.07197	0.99601
CH ₄	-	0.00000	0.00000	0.58439	0.89289	0.16645	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	0.99817	0.99601

Thermodynamic Performance. The net power output of the two configurations with MEG absorber are very similar with 9,338 kW (max. water) and 9,449 kW (max. heat). The SOFC island produces 10,797 kW_{DC} (max. water, FU 88.7%) and 10,802 kW_{DC} (max. heat, FU 88.7%). Due to the small changes in the TIT associated with the fuel flow changes, the max. water recovery case has a slightly lower GT power output with 1,300 kW versus 1,308 kW in the max. heat recovery case. The GT surge margins in the MEG absorber cases are 11.3% (max. water) and 11.2% (max. heat). The auxiliary load of the two cases differs in three key areas from each other: I) the anaerobic digester power consumption is reduced in the max. heat case due to lower fuel requirement II) the lower fuel requirement in the max. heat case reduces the power consumption of fuel gas compression and III) the lower water recovery in the max. heat case reduces the power consumption of the water recovery system. As a result, the auxiliary load in the max. heat case is with 2,662 kW about 96 kW lower than in the max. water case.

With the anaerobic digester being integrated into the water recovery system, latent heat from the flue gas can be upgraded and made available to the digester. This is fundamentally different compared to the air cooled-condenser and direct contact condenser cases where the latent heat is released at the flue gas's dew point temperature, which is too low for integration with the anaerobic digester, and the latent heat needs to be rejected to the environment in order to recover the water. As a result of this heat integration, the efficiency of the MEG absorber cases increases the plant efficiency instead of lowering it. The respective electrical efficiencies of the MEG absorption cases with max. water recovery and max. heat recovery are 53.53%-LHV and 56.66%-LHV. This is a substantial efficiency increase versus the reference case with 53.07%-LHV. The GT efficiency and SOFC stack efficiency in the LiBr cases remain virtually constant

compared to the reference case, and efficiency gains arise from a lower auxiliary load and lower fuel input (reduction in flue flow to the duct burner).

The carbon negative emission potentials for these MEG absorber cases are 5,588 kg/h (598 kg/MWh) in the max. water scenario and 5,344 kg/h (565 kg/MWh) in the max. heat scenario. The water recovery from the flue gas in the MEG absorber case with max water recovery is 82.3% and 47.0% in the max. heat recovery scenario. At the scale of this nominal 10 MW plant, a total of 50,825 kg of water can be recovered per day, which also creates 619 kg/h of liquid waste (max. water), whereby, a substantial portion of the wastewater is created during water treatment. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from a 10 MW biogas SOFC-GT hybrid plant with MEG water recovery system can produce enough water for the irrigation of over 26,200 m² of farmland (including 90% capacity factor). In the max. heat recovery scenario, the amount of water recovered is reduced to 27,817 kg/day. The plant performance data of both configurations are summarized in **Table 5-55**.

Table 5-55: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrids with MEG Absorption System for Water Recovery

Gross Power Generation	Unit	MEG Absorber - max. Water	MEG Absorber - max. Heat
Fuel Cell	kW _{DC}	10,797.26	10,802.40
Gas Turbine	kW	1,298.98	1,307.94
Total Generation	kW	12,096.24	12,110.34
Auxiliary Load			
DC-AC Inverter Losses	kW	323.92	324.07
Anaerobic Digester	kW	1460.19	1402.73
Fuel Gas Compression	kW	785.65	764.79
Fuel Gas Intercooler	kW	0.25	0.24
Regen. Sulfur Removal	kW	81.81	78.12
Water Recovery System	kW	49.80	36.36
Water Treatment	kW	3.47	1.94
Zero Liquid Discharge System	kW	N/A	N/A
Miscellaneous Balance of Plant	kW	7.85	7.86
Transformer Losses	kW	45.49	45.54
Total Auxiliary Load	kW	2758.41	2661.65
Net Power Generation	kW	9,337.83	9,448.68
Plant Performance			
Net Efficiency	%-HHV	48.06	50.86
Net Efficiency	%-LHV	53.53	56.66
Consumables			
Fuel Gas	kg/h	3,547	3,391
Thermal Input	MJ _{LHV} /h	62,798	60,039
Water	kg/h	0	0
Emissions and Waste Streams			
Carbon Capture	%	N/A	N/A
Net Carbon Emission (Absolute)	kg/h	0 (5,588)	0 (5,344)
Net Carbon Emission (Absolute)	kg/MWh	0 (598)	0 (565)
Liquid Waste	kg/h	619	447
Solid Waste	g/h	0	0
Water Recovery			
Flue Gas Water Recovery	%	82.3	47.0
Flue Gas Water Recovery	kg/d	50,825	27,817

Economics. The TPCs of the 10 MW biogas hybrids with MEG absorber for water recovery are \$51,867,150 (max. water) and \$49,769,160 (max. heat) resulting in specific plant costs of 5,555 \$/kW (max. water) and 5,267 \$/kW (max. heat). The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the air-cooled condenser case, which are still valid for this case. The GT island cost remains constant at \$3,651,010 since the same GT is used in all study cases. The anaerobic digester costs have been substantially reduced in the MEG cases. In the reference case, the digester cost is with \$8,810,340 around \$730,000 more expensive than in the max. heat case. The higher efficiency is not only beneficial with respect to the operating costs but also reduces the fuel demand and the size of the digester. Similarly, the fuel processing cost is reduced. Ultimately, this helps to reduce the capital cost and offset some of the expenses associated with the water recovery. The water recovery equipment in the max. heat recovery case is about \$1,735,090 cheaper than in the max. water recovery case, which is primarily due to the size reduction of the membrane-based desiccant regenerator. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-96** (max. water) and **Figure 5-97** (max. heat). A more detailed cost summary of all plant equipment is provided in **Table 5-56**.

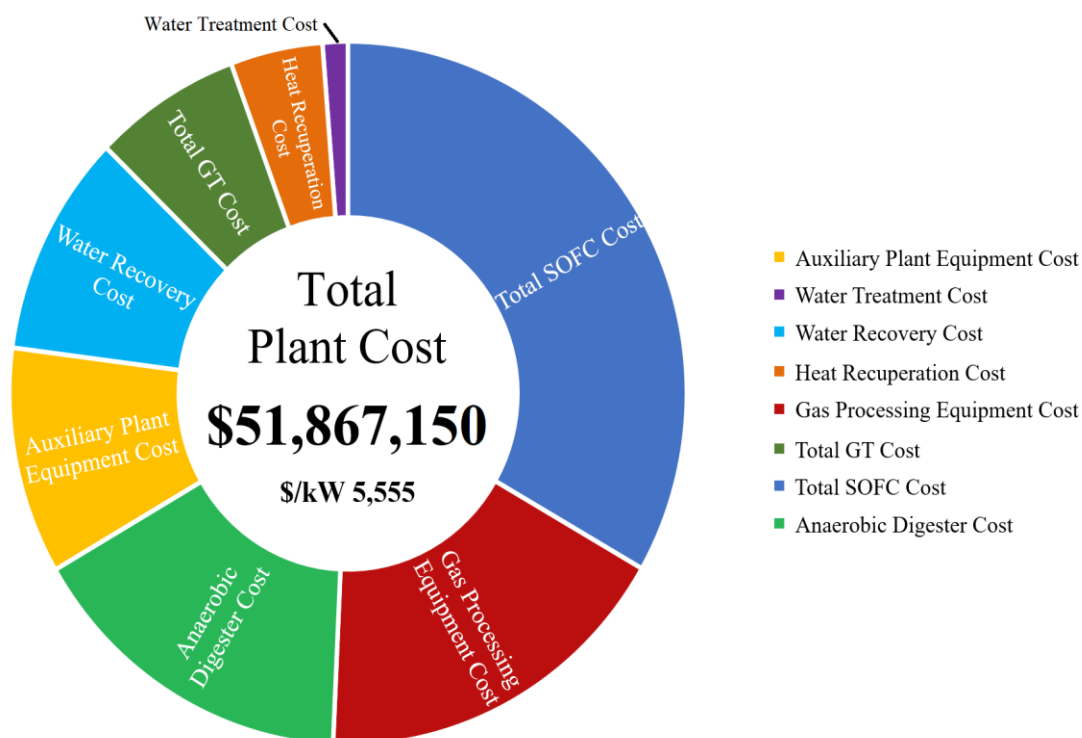


Figure 5-96: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with MEG absorption system for water recovery designed for maximum water recovery.

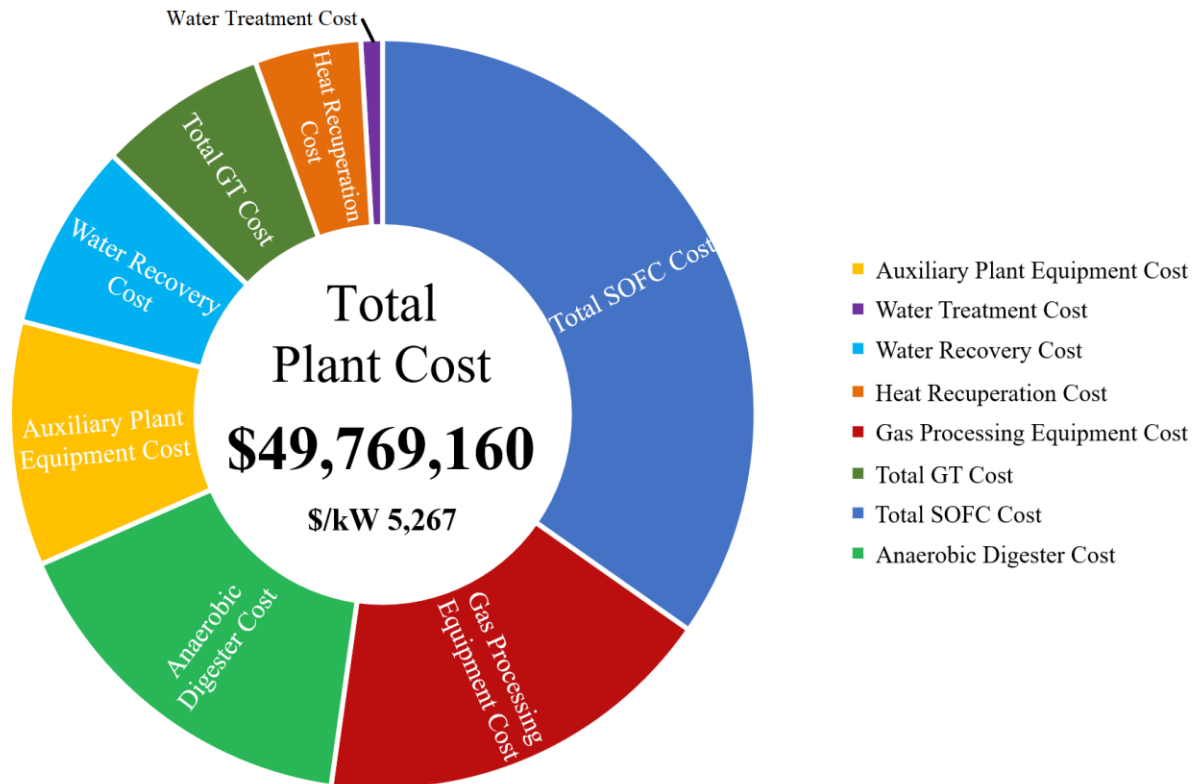


Figure 5-97: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with MEG absorption system for water recovery designed for maximum heat recovery.

The overnight capital cost in the MEG absorber case designed for max. water recovery is \$68,732,070, an increase of over 12.4% compared to the reference case. The MEG case designed for max. heat recovery has a slightly lower overnight capital cost of \$66,026,370. Variable operating costs increase compared to the reference case and this is mostly due to the MEG makeup to compensate for MEG losses during regeneration. Savings in anaerobic digester consumables as well as savings in the fuel processing are small compared to the annual MEG expenses. The 100 wt-% MEG solution is assumed to cost 0.65 \$₂₀₁₆/kg and the initial fill costs about \$138,170 (max. water) and \$157,420 (max. heat). The MEG loss by entrainment in the absorber is small; however, the losses in the regeneration process, associated with MEG's vapor pressure, are significant. In the MEG absorber case, water byproduct sales reduce the variable operating costs by \$2,916,629 per year in the max. water recovery scenario and by \$2,421,055 in the max. heat recovery scenario.

The fixed operating cost in the MEG absorber case increases due to tax and insurance costs of the additional equipment (compared to reference case). The operating labor costs in the MEG absorber case are identical to the air-cooled condenser and direct contact condenser cases. The COW is determined by

Table 5-56: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrids with MEG Absorption System for Water Recovery

Equipment Cost	Unit	MEG Absorber - max. Water	MEG Absorber - max. Heat
Fuel Cell Island	\$	17,264,680	17,270,440
SOFC Stacks	\$	6,818,390	6,818,630
Power Conditioning	\$	6,303,070	6,308,360
Pressure Vessel, Housing & Assembly	\$	3,536,440	3,536,590
System Installation	\$	606,780	606,860
Gas Turbine Island	\$	3,651,010	3,651,010
Gas Turbine-Generator Set	\$	3,077,600	3,077,600
Gas Turbine Foundation	\$	44,740	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680	528,680
Anaerobic Digester	\$	8,270,630	8,078,200
Anaerobic Digester	\$	8,001,150	7,786,590
AD Heat Supply	\$	269,480	291,610
Gas Processing and Flue Gas Stack	\$	9,033,690	8,719,280
Fuel Compressors and Intercooler	\$	2,835,910	2,705,250
Fuel Processing	\$	6,067,620	5,883,860
Flue Gas Stack System	\$	130,160	130,170
Heat Recuperation	\$	2,294,380	2,293,140
Cathode Off-gas Heat Exchanger	\$	994,650	994,040
Air Preheater	\$	1,127,140	1,126,930
Fuel Preheater	\$	172,600	172,170
Water Recovery	\$	5,941,100	4,516,010
Water Recovery System	\$	5,314,700	4,039,040
Water Treatment	\$	626,400	476,970
Zero Liquid Discharge System	\$	N/A	N/A
Auxiliary Plant Equipment Cost	\$	5,411,660	5,241,080
Accessory Electric Plant	\$	1,161,160	1,162,520
Instrumentation & Controls	\$	1,904,670	1,827,620
Improvement to Site	\$	495,210	475,180
Buildings & Structures	\$	1,850,610	1,775,760
Total Plant Cost	\$	51,867,150	49,769,160
Specific Plant Cost	\$/kW	5,555	5,267

calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 137.19 \$/MWh. The COW in the MEG absorber case with max. water recovery is 0.1727 \$/L and 0.2613 \$/L in the max. heat recovery case. A summary of the economic key parameters can be found in **Table 5-57**. Detailed breakdowns of the COEs for the two scenarios are shown in **Figure 5-98** (max. water) and **Figure 5-99** (max. heat).

Figure 5-98 and **Figure 5-99** show the COE value of 137.19 \$/MWh in solid colors and the byproduct revenue is shown with striped background, which amounts to -39.62 \$/MWh for the max. water recovery case and to -32.50 \$/MWh in the max. heat recovery case. The whole circle represents the COE without byproduct credit, 176.81 \$/MWh in the max. water recovery case and 169.69 \$/MWh in the max. heat recovery case. This means, if the water would be provided to the local community at no cost, the COE would increase to 176.81 \$/MWh and 169.69 \$/MWh respectively, due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.1727 \$/L and 0.2613 \$/L in order to reach a COE of

137.19 \$/MWh. These COWs are substantially higher than in the previous case due to large quantities of MEG makeup needed.

Table 5-57: Economics of the 10 MW Biogas SOFC-GT Hybrids with MEG Absorption System for Water Recovery

Item	Unit	MEG Absorber - max. Water	MEG Absorber - max. Heat
Plant Capacity Factor	%	90	90
Total Overnight Capital Cost			
Total Plant Cost	\$	51,867,150	49,769,160
Preproduction Costs			
6 Months Labor Cost	\$	1,517,810	1,499,530
1 Month Maintenance Materials	\$	75,860	71,330
1 Month Non-Fuel Consumables	\$	281,270	276,820
1 Month Waste Disposal	\$	50	40
25% of 1 month fuel cost at 100% CF	\$	0	0
2% of TPC	\$	1,037,340	995,380
Inventory Capital Costs			
60 Day Supply Consumables at 100% CF	\$	415,870	412,500
0.5% of TPC (Spare Parts)	\$	259,340	248,850
Other Costs			
Initial Catalyst, Sorbent & Chemicals Cost	\$	562,720	563,400
Land	\$	3,534,170	3,380,200
Other Owner's Costs (15% of TPC)	\$	7,780,070	7,465,370
Financing Costs	\$	1,400,410	1,343,770
Variable Operating Cost for Initial Year of Operation			
Annual Feed Cost at above CF for Initial Year	\$	0	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	3,857,450	3,760,510
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0	0
Annual Byproduct Revenues at above CF for Initial Year	\$	2,916,630	2,421,060
Fixed Operating Cost for Initial Year of Operation			
Labor Cost	\$	3,035,620	2,999,060
Property Tax & Insurance	\$	1,037,340	995,380
Cost of Electricity and Cost of Water			
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A	N/A
First Year Levelized Cost of Water	\$/L	0.1727	0.2613

Figure 5-100 shows a detailed breakdown of the COW for the max. water recovery case and **Figure 5-101** shows the breakdown of the COW for the max. heat recovery case. For the MEG absorber cases the total operating labor costs are equally split between the water recovery rubric and the water treatment rubric, which is identical to the air-cooled condenser case and the direct contact condenser case. In the MEG cases the fixed operating costs no longer represent the largest COW driving factor and the variable operating costs of the MEG absorption system become the dominant factor contributing over 61.0% (max. water) and 65.7% (max. heat) to the COW. The performance penalty costs in the MEG absorber cases are 1.9% (max. water) and 1.5% (max. heat) and represent only the parasitic loads since the system integration losses are negative (meaning positive effect of water recovery system integration). The fixed operating costs contribute 15.2% (max. water) and 14.5% (max. heat) to the COW. The water treatment is responsible for only 7.9% (max. water) and 7.3% (max. heat) of the COW.

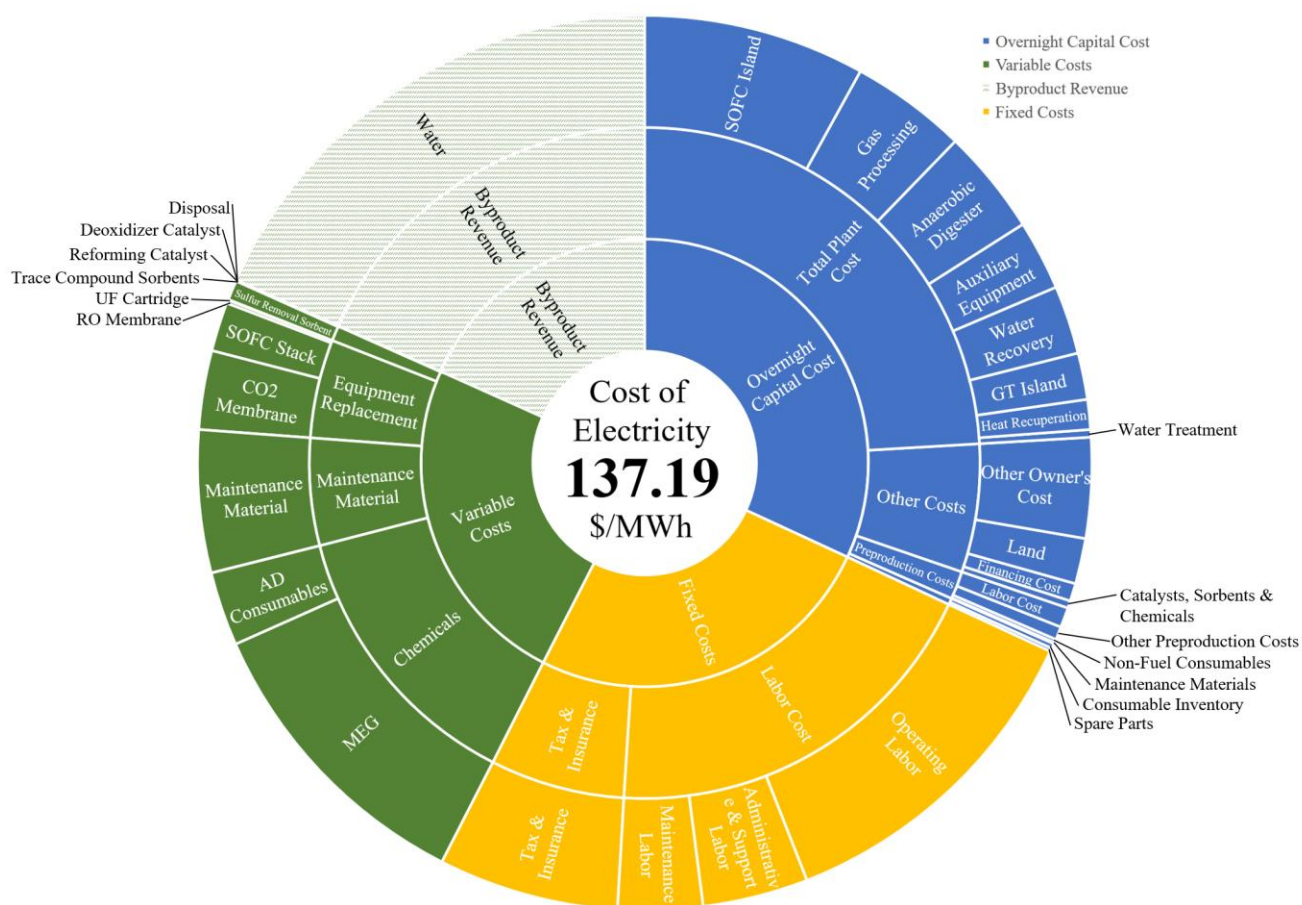


Figure 5-98: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with MEG absorber for water recovery designed for maximum water recovery.

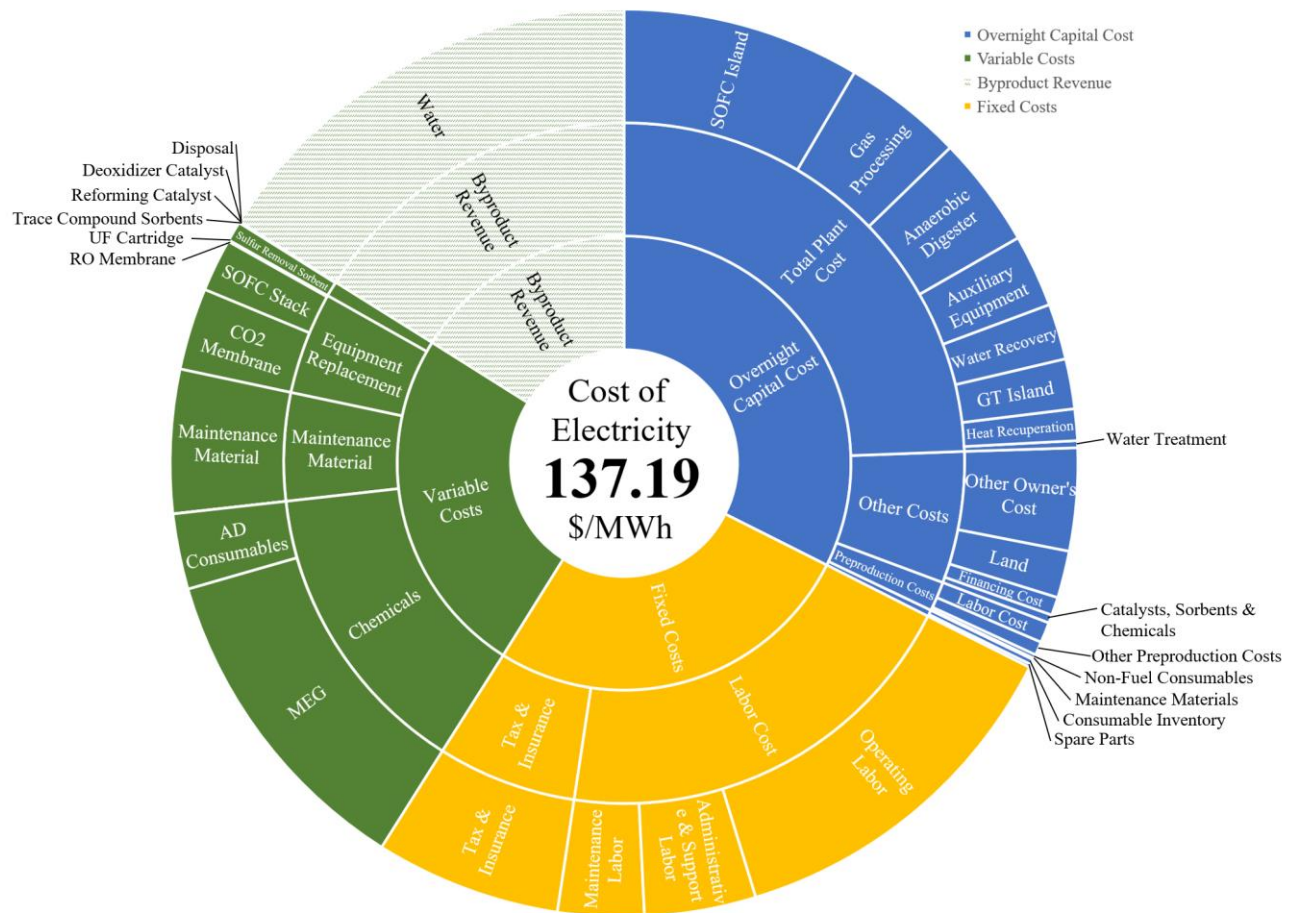


Figure 5-99: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with MEG absorber for water recovery designed for maximum heat recovery.

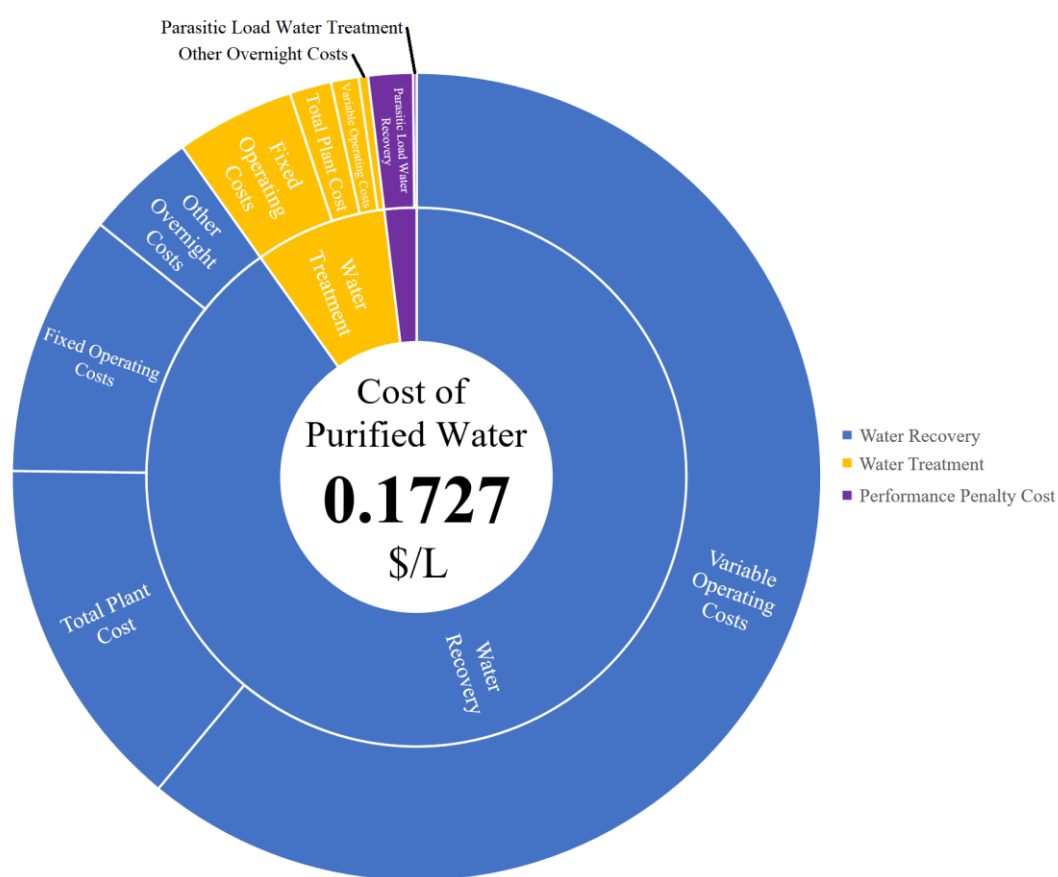


Figure 5-100: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with MEG absorber for water recovery designed for maximum water recovery.

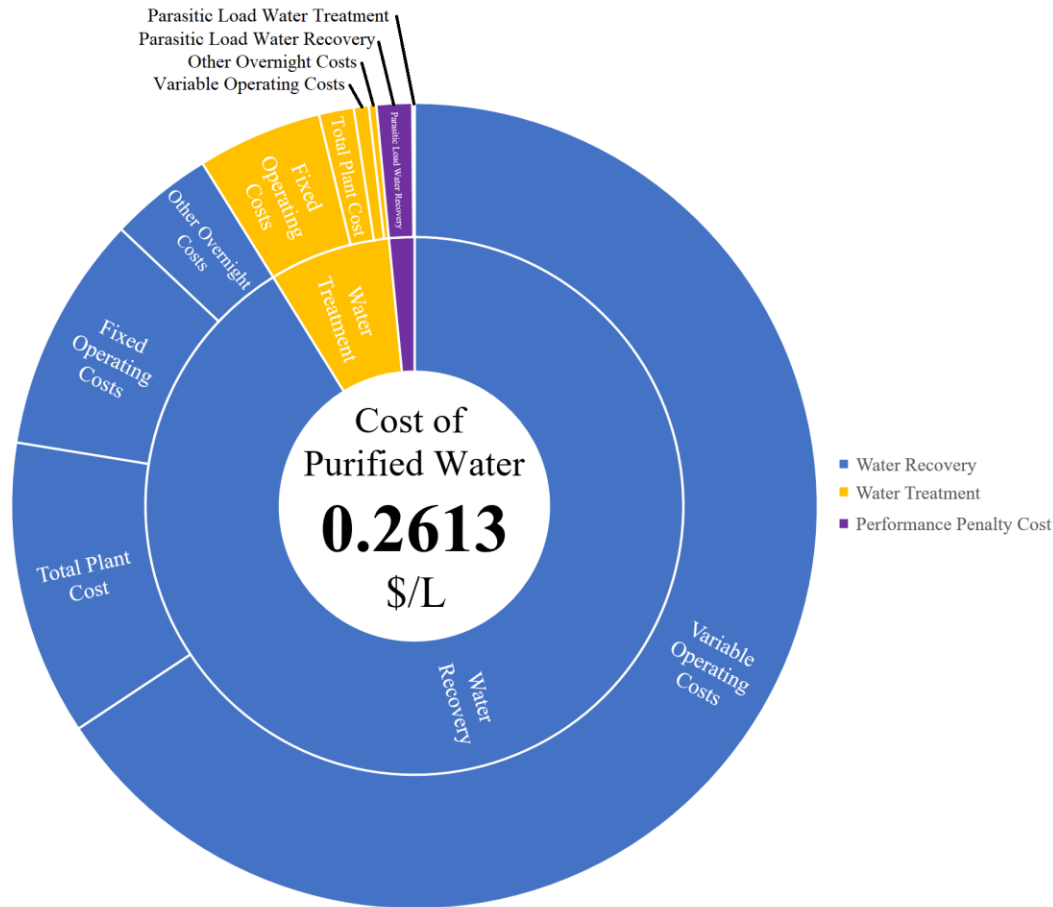


Figure 5-101: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with MEG absorber for water recovery designed for maximum heat recovery.

Transport Membrane Condenser

The plant configuration of the 10 MW biogas SOFC-GT hybrid with TMC is very similar to the air-cooled condenser case and direct contact condenser case. The TMC has a similar function as a water-cooled heat exchanger; however, the membrane allows the water to condense in the pores above its dew point temperature and the condensed water instantly mixes with the cooling water inside the cooling tube. No spray cooling upstream of the TMC is needed in this case. The water stream leaving the TMC (cooling water + recovered water) is cooled in an air-cooled exchanger and the majority of the water is recirculated back to the TMC as cooling water, while the other portion (net water recovered) is prepared for export. Due to the water transport through the membrane, the recovered water is of high quality [198] and only needs minor treatment (pH-adjustment). The water recovery from the flue gas is limited by the cooling medium (ambient air in this case) and does not allow for direct integration with the anaerobic digester. The flue gas leaves the plant with some superheat and a relatively high temperature which is beneficial with respect to plume rise, fog formation and dispersion as discussed earlier. The flowsheet of the TMC case can be found

in **Figure 5-102**. The accompanying state-point stream data as indicated in **Figure 5-102** can be found in **Table 5-58**.

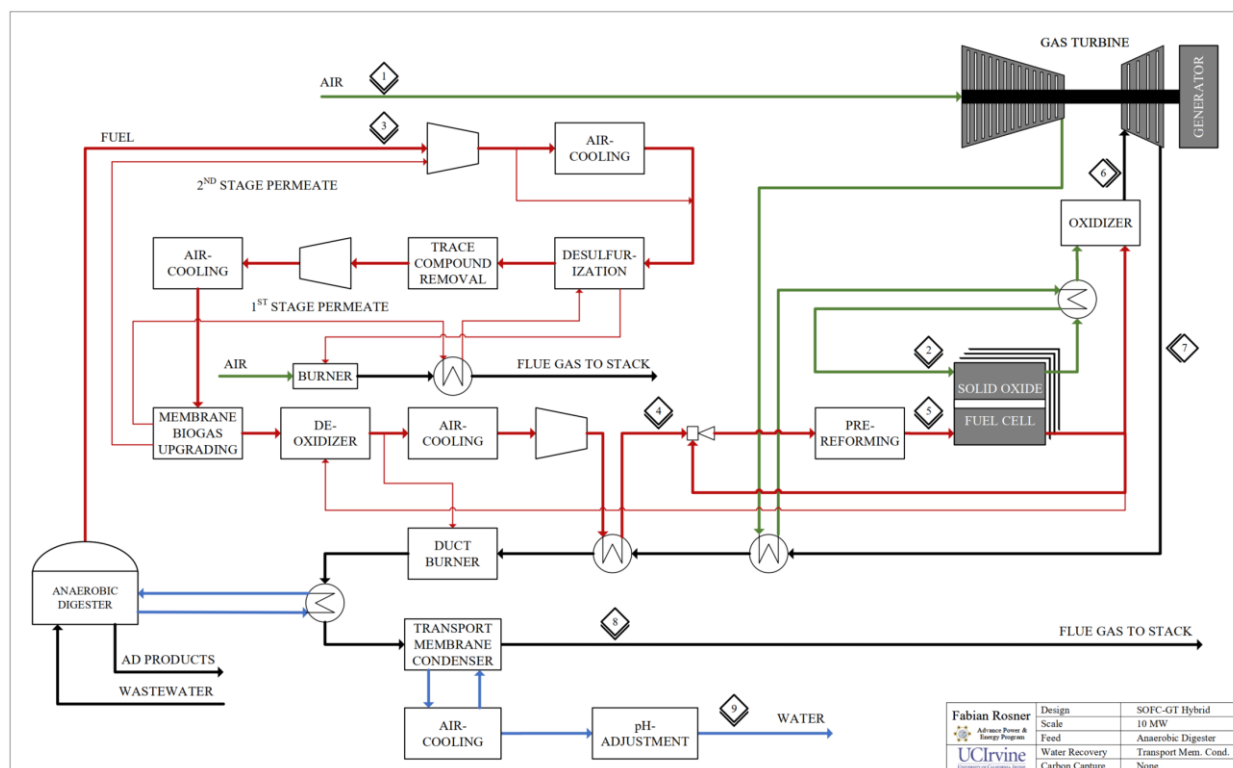


Figure 5-102: Flowsheet of the 10 MW biogas SOFC-GT hybrid system with transport membrane condenser for water recovery.

Table 5-58: Stream Summary of the 10 MW Biogas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	27	322	700	902	558	51	40
Pressure	bar	1.01	7.17	1.01	25.06	7.17	6.57	1.08	1.01	2.00
Molar Flow Rate	kmol/h	1,076.2	1,076.2	135.4	77.8	334.0	1,142.5	1,142.5	1,113.6	36.2
Mass Flow Rate	kg/h	31,057	31,057	3,590	1,419	7,053	32,177	32,177	31,659	652
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00966	0.00000	0.00000	0.07411	0.07411	0.06429	Trace
N ₂	-	0.77315	0.77315	0.00966	0.01618	0.00774	0.72939	0.72939	0.74838	0.00055
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00870	0.00870	0.00893	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00270	0.19498	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00828	0.07878	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.36223	0.06820	0.21857	0.06268	0.06268	0.07067	Trace
H ₂ O	-	0.00987	0.00987	0.03381	0.01162	0.33335	0.12511	0.12511	0.10773	0.99937
CH ₄	-	0.00000	0.00000	0.58439	0.89292	0.16656	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.00004	0.00006	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	Trace	Trace	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	0.00014	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	0.99993	0.99997	0.99998	1	1	1	0.99992

Thermodynamic Performance. The power output of the nominal 10 MW biogas hybrid plant with TMC for water recovery is 9,356 kW, whereby, the SOFC produces a net power of 10,831 kW_{DC} (88.7% FU) and the GT 1,297 kW. Due to the lower backpressure of the GT associated with the lower pressure drop of the TMC, the GT produces 119 kW more than in the air-cooled condenser case. The change in back pressure results in a SOFC operating pressure of 7.17 bar, the lowest among the water recovery cases, and a GT surge margin of 11.5%. The general implications of changes in SOFC operating pressure have been discussed for the 10MW NG case and biogas plant with air-cooled condenser, and also apply here. The auxiliary load accounts for 2,772 kW and the common plant equipment (except water recovery and water treatment) scales with the plant size and plant efficiency. The auxiliary load for the water recovery system is similar to the air-cooled condenser case. On the other hand, water treatment load is reduced when compared to the air-cooled condenser case by 54.3% due to the lower water recovery and less expensive treatment process; however, this auxiliary load reduction is very small compared to the total auxiliary load on an absolute basis.

The efficiency of the TMC case is 52.99%-LHV, which is 0.08% points lower than the reference case. The GT efficiency in this scenario is 57.7%-LHV and the SOFC stack efficiency is 82.1%-LHV.

The negative CO₂ emission potential for this hybrid plant is 5,654 kg/h or 604 kg/MWh. The water recovery from the flue gas using the TMC is 23.1%, which is slightly higher than in the NG case due to a higher water partial pressure in the exhaust gas in the biogas case. At the scale of this nominal 10 MW plant, a total of 15,622 kg of water can be recovered per day. Thus, the hybrid plant with direct contact condenser water recovery system can produce enough water for the irrigation of over 8,000 m² (including 90% capacity factor, 1.74 l/m² of irrigation water per day [341]). Water condensation after biogas compression and intercooling results in a liquid waste stream of 201 kg/h, which is disposed at the on-site wastewater treatment plant. The plant performance data are summarized in **Table 5-59**.

Table 5-59: Thermodynamic Performance of the 10 MW Biogas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Gross Power Generation		Unit	TMC
	Fuel Cell	kW _{DC}	10,830.87
	Gas Turbine	kW	1,297.30
Total Generation		kW	12,128.17
Auxiliary Load			
	DC-AC Inverter Losses	kW	324.93
	Anaerobic Digester	kW	1502.31
	Fuel Gas Compression	kW	799.60
	Fuel Gas Intercooler	kW	0.26
	Regen. Sulfur Removal	kW	82.87
	Water Recovery System	kW	8.55
	Water Treatment	kW	0.10
	Zero Liquid Discharge System	kW	N/A
	Miscellaneous Balance of Plant	kW	7.87
	Transformer Losses	kW	45.61
Total Auxiliary Load		kW	2772.10
Net Power Generation		kW	9,356.07
Plant Performance			
	Net Efficiency	%-HHV	47.57
	Net Efficiency	%-LHV	52.99
Consumables			
	Fuel Gas	kg/h	3,590
	Thermal Input	MJ _{LHV} /h	63,563
	Water	kg/h	0
Emissions and Waste Streams			
	Carbon Capture	%	N/A
	Net Carbon Emission (Absolute)	kg/h	0 (5,654)
	Net Carbon Emission (Absolute)	kg/MWh	0 (604)
	Liquid Waste	kg/h	201
	Solid Waste	g/h	0
Water Recovery			
	Flue Gas Water Recovery	%	23.1
	Flue Gas Water Recovery	kg/d	15,622

Economics. The TPC of the 10 MW biogas hybrid with TMC for water recovery is \$47,238,720, which corresponds to a specific plant cost of 5,049 \$/kW. The SOFC island costs are very similar to the previous cases and the slight variations result from the number of installed spare stacks and SOFC operating pressure effects (see previous explanations for example air-cooled condenser case). The GT island cost remains constant at \$3,651,010 since the same GT is used in all study cases. Other equipment costs such as anaerobic digester, gas processing and flue gas stack scale with the plant size but remain relatively constant. Heat recuperation in the TMC case is shifted towards the air preheater due to the higher GT backpressure, which allows for more heat recovery from the GT exhaust (compared to the reference case). Thus, the cathode off-gas exchanger cost is reduced. The TMC water recovery system costs \$913,280 and is significantly cheaper than in the 10 MW NG case with TMC. Due to the low inlet temperature and the elimination of the water spray upstream of the TMC, the absolute amount of water recovery is reduced resulting in a reduction of the TMC modules needed while achieving a similar net water recovery rate. The water treatment

equipment cost is \$98,720. An illustration of the TPC distribution of the various plant categories is shown in **Figure 5-103**. A more detailed cost summary of all plant equipment is provided in **Table 5-60**.

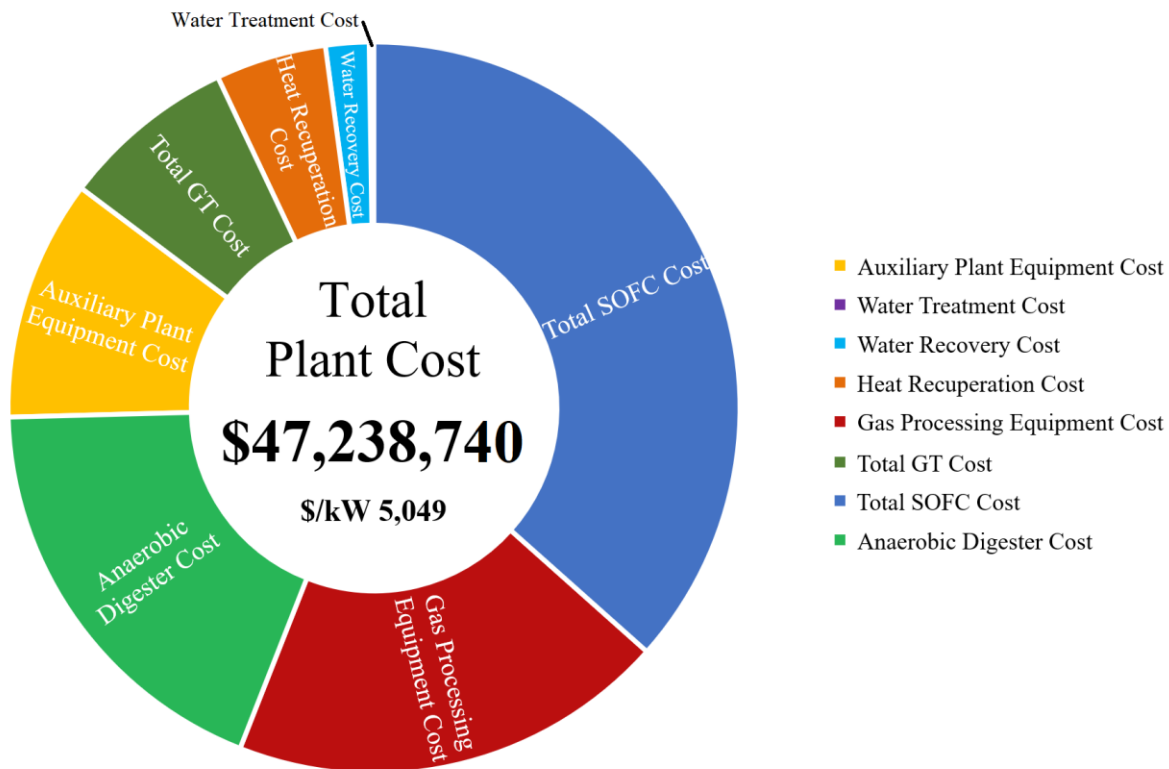


Figure 5-103: Total plant cost and specific plant cost of the 10 MW biogas SOFC-GT hybrid with transport membrane condenser for water recovery.

The overnight capital cost in the TMC case is \$62,539,627. The increase of the overnight capital cost compared to the biogas reference case is mostly due to the TPC increase of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Other variable operating costs increase due to the water recovery and water treatment equipment. In the TMC case, water byproduct sales reduce the variable operating costs by \$439,410 per year.

Table 5-60: Equipment Cost of the 10 MW Biogas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Equipment Cost	Unit	TMC
Fuel Cell Island	\$	17,284,360
SOFC Stacks	\$	6,817,950
Power Conditioning	\$	6,322,460
Pressure Vessel, Housing & Assembly	\$	3,536,620
System Installation	\$	607,330
Gas Turbine Island	\$	3,651,010
Gas Turbine-Generator Set	\$	3,077,600
Gas Turbine Foundation	\$	44,740
Piping, Instrumentation & Electrical Equipment	\$	528,680
Anaerobic Digester	\$	8,800,400
Anaerobic Digester	\$	8,060,020
AD Heat Supply	\$	740,380
Gas Processing and Flue Gas Stack	\$	9,158,390
Fuel Compressors and Intercooler	\$	2,894,380
Fuel Processing	\$	6,133,800
Flue Gas Stack System	\$	130,200
Heat Recuperation	\$	2,297,160
Cathode Off-gas Heat Exchanger	\$	1,005,640
Air Preheater	\$	1,119,350
Fuel Preheater	\$	172,170
Water Recovery	\$	1,012,000
Water Recovery System	\$	913,280
Water Treatment	\$	98,720
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	5,035,430
Accessory Electric Plant	\$	1,164,230
Instrumentation & Controls	\$	1,734,700
Improvement to Site	\$	451,020
Buildings & Structures	\$	1,685,470
Total Plant Cost	\$	47,238,740
Specific Plant Cost	\$/kW	5,049

The fixed operating cost in the TMC case increases due to tax and insurance costs of the additional equipment. The operating labor cost in the previous cases and the TMC case are identical; however, the operating labor in the TMC is completely assigned to the water recovery unit like in the LiBr absorber case due to the relatively simple water treatment. When comparing the labor cost on a relative basis, the labor cost in the TMC case is higher due to the low water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the reference case. Thus, the COE remains constant at 137.19 \$/MWh. The COW in the TMC case is 0.0855 \$/L. A summary of the economic key parameters can be found in **Table 5-61**. A detailed breakdown of the COE is shown in **Figure 5-104**.

Table 5-61: Economics of the 10 MW Biogas SOFC-GT Hybrid with Transport Membrane Condenser for Water Recovery

Item	Unit	TMC
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	47,238,740
Preproduction Costs		
6 Months Labor Cost	\$	1,482,940
1 Month Maintenance Materials	\$	67,220
1 Month Non-Fuel Consumables	\$	119,810
1 Month Waste Disposal	\$	50
25% of 1 month fuel cost at 100% CF	\$	0
2% of TPC	\$	944,770
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	100,140
0.5% of TPC (Spare Parts)	\$	236,190
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	429,620
Land	\$	3,558,880
Other Owner's Costs (15% of TPC)	\$	7,085,810
Financing Costs	\$	1,275,450
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	0
Other Annual Variable Operating Cost at above CF for Initial Year	\$	2,020,510
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	439,410
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	2,965,870
Property Tax & Insurance	\$	944,770
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	137.19
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0855

Figure 5-104 shows the COE value of 137.19 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-5.96 \$/MWh) and the whole circle represents the COE without byproduct credit (143.15 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 143.15 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0855 \$/L in order to reach the COE of 137.19 \$/MWh.

Compared to the NG TMC case, the COW is substantially reduced in the biogas case (from 0.1644 \$/L to 0.0855 \$/L). A mayor factor is the TPC reduction of the TMC, which is now responsible for only 16.7% of the COW (26.8% in the NG TMC case). Simultaneously, this reduces variable operating costs, such as maintenance and TMC replacement costs. In **Figure 5-105** it can be seen that also the parasitic load cost is reduced compared to the NG TMC case due to the lower absolute water recovery (no spray). While the parasitic load cost in the NG TMC case represents 3.6% of the COW, it is only 2.1% in the biogas case. On a relative basis, the fixed operating costs are now more dominant and are responsible for 62.1% of the COW.

SOFC-GT Hybrid Systems with Water Recovery

Overall, the water recovery unit accounts for 85.5% of the COW and the water treatment accounts for 3.5% of the COW.

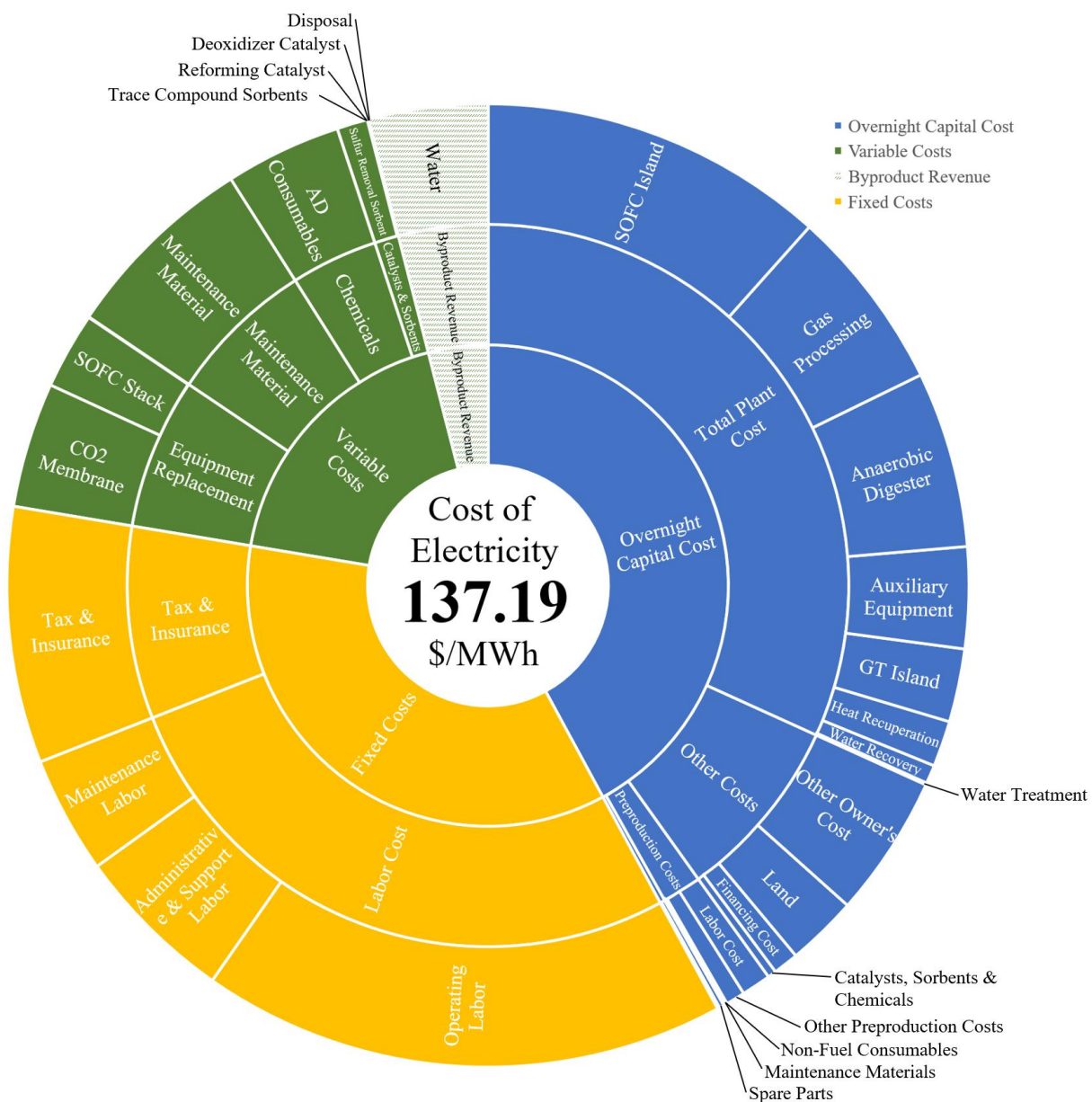


Figure 5-104: Breakdown of the cost-of-electricity of the 10 MW biogas SOFC-GT hybrid with transport membrane condenser for water recovery.

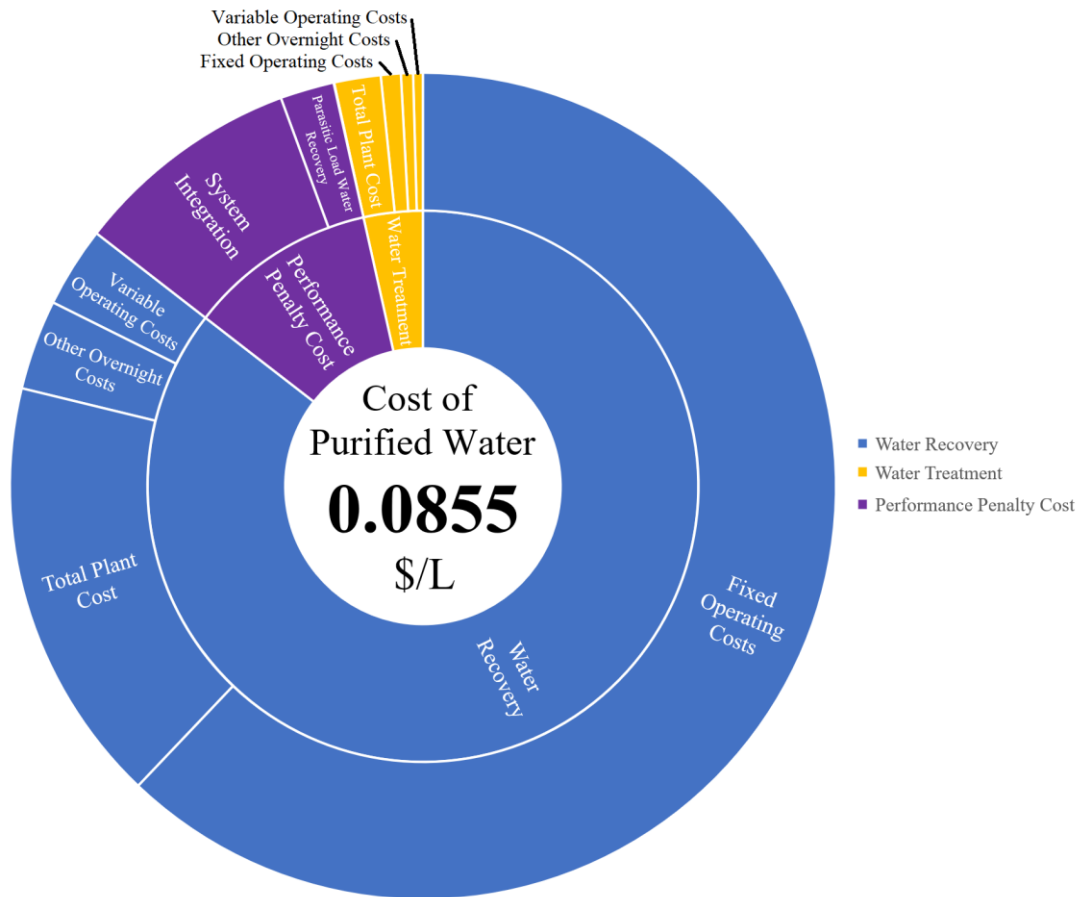


Figure 5-105: Breakdown of the cost-of-water of the 10 MW biogas SOFC-GT hybrid with transport membrane condenser for water recovery.

Summary of the 10 MW Biogas SOFC-GT Hybrids

In total five different water recovery technologies, an air-cooled condenser, a direct contact condenser, a LiBr absorption system, a MEG absorption system and a transport membrane condenser, have been integrated into an SOFC-GT hybrid power plant with anaerobic digester and compared to a reference case without water recovery. The heat integration of the air-cooled condenser, the direct contact condenser and the transport membrane condenser does not allow the latent heat to be recovered for the anaerobic digester. Thus, the latent heat needs to be rejected, similarly to the NG cases, which is associated with a performance penalty between 0.08-0.31% points-LHV. The performance decrease can be traced back to two major effects: the increase of GT backpressure and the parasitic load of the water recovery and treatment. The lowest efficiency is obtained for the air-cooled condenser case, due to the high pressure drop in the air-cooled condenser, despite its low parasitic power consumption. The lowest pressure drop is obtained in the

TMC scenario, which only results in a small efficiency decrease. The back pressure is not just important with respect to efficiency but also has an influence upon the surge margin (high surge margin is maintained at low backpressures). In the LiBr and MEG absorption systems, the latent heat can be upgraded using the hot GT exhaust gas. After upgrading the latent heat, it becomes available at a sufficiently high temperature, which allows for direct heat integration with the anaerobic digester. As a result, the fuel flow to the duct burner (fuel needed to heat the anaerobic digester) can be reduced and the efficiency increases. The absorption systems can be optimized for maximum water recovery or maximum heat recovery. In general, stronger desiccants are able to recover more water and provide more heat to the anaerobic digester. The LiBr case with max. heat recovery is able to satisfy the total heat demand of the anaerobic digester. A comparison of the electrical efficiencies is provided in **Figure 5-106**.

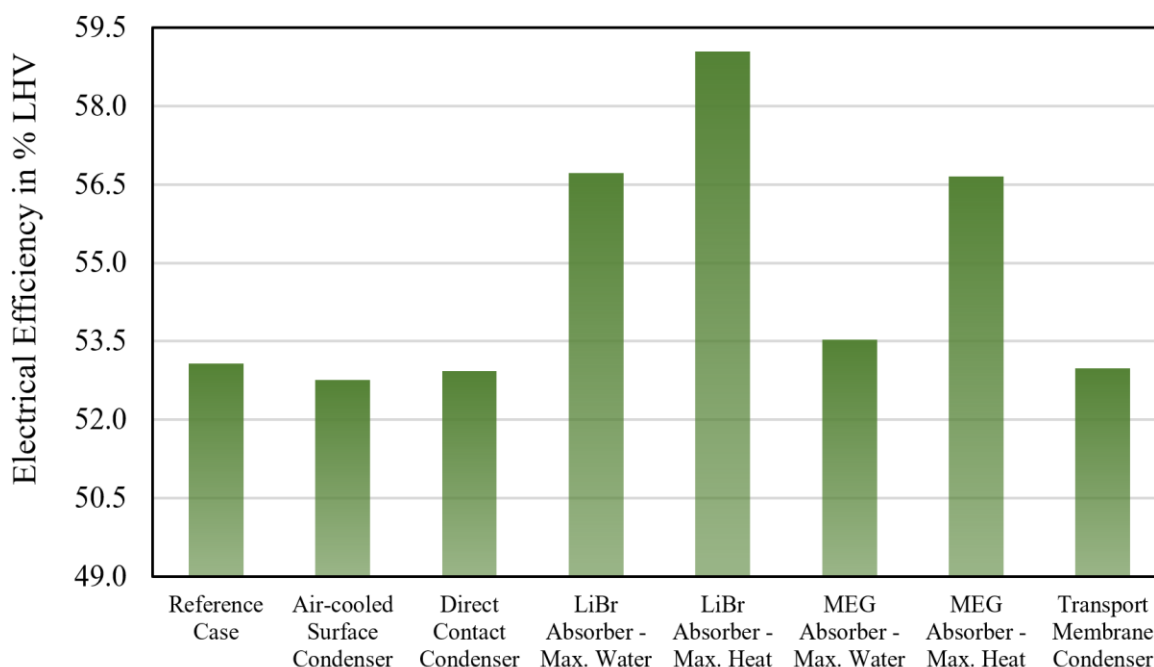


Figure 5-106: Comparison of the electrical efficiencies of the 10 MW biogas SOFC-GT hybrids with water recovery.

The COE is established in the reference case without water recovery and its value is 137.19 \$/MWh. The COE is held constant throughout the water recovery scenarios and the water sales price is adjusted in order to obtain the same COE as in the reference case. Three main factors contribute to the COW: the capital cost, the operating cost and the amount of water recovered. The air-cooled condenser has low capital and operating costs, and the amount of water recovery is the fourth highest among the studied scenarios, which results in the third lowest COW. The direct contact condenser and TMC mostly suffer from low water recovery. The TMC case within the biogas scenarios significantly improved compared to the NG case due

to the TMC size reduction associated with the lower inlet temperature. The MEG absorption system exhibits enormous operating costs due to MEG loss, which are increased with higher regeneration temperatures. As a result, the MEG absorber case designed for max. heat recovery has a higher COW than the max. water recovery case despite its higher efficiency. The LiBr absorption system with its high water recovery and low operating cost is the most economical water recovery technology despite its high investment costs. It particularly benefits from the advantageous heat integration, which helps to lower the COW. The lowest COW is reached in the max. heat recovery case indicating that the latent heat in the water vapor is more valuable than the recovered water itself. A summary of the COE and COW for the various water recovery technologies is presented in **Figure 5-107**.

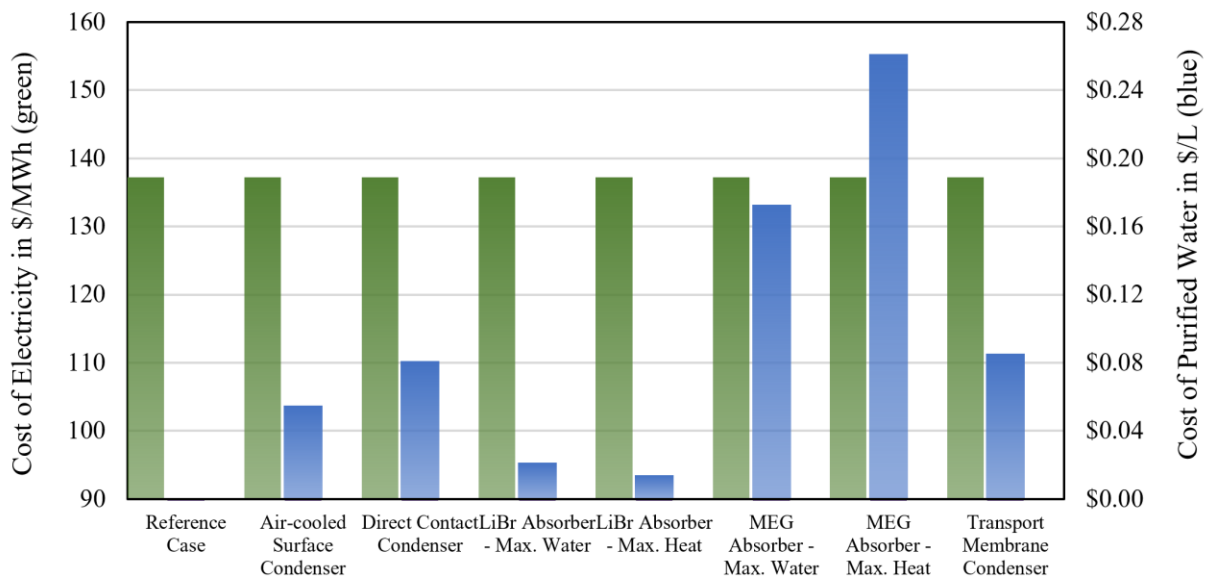


Figure 5-107: Comparison of the cost-of-electricity and cost-of-water of the 10 MW biogas SOFC-GT hybrids with water recovery.

The water recovery technologies use different physical principles to remove the water vapor from the flue gas. The air-cooled condenser as well as the direct contact condenser rely on cooling the flue gas below the gas's dew point temperature. Once the dew point temperature is reached, the water vapor starts condensing. For both, the air-cooled condenser and the direct contact condenser, the ambient air is the cooling medium. In the case of the air-cooled condenser, ambient air directly cools the flue gas; however, a large surface area is needed to facilitate the heat exchange. In a direct contact condenser, this surface area is created by a packed bed with significantly less pressure drop; however, a secondary cooling water loop is needed, which ultimately leads to a higher flue gas outlet temperature (addition of pinch temperatures ambient air-cooling water, cooling water-flue gas). Thus, the air-cooled condenser is able to recover more water than the direct contact condenser. The absorption cases use a liquid desiccant and no longer rely on

the gas's dew point temperature. The desiccant's vapor pressure determines the maximum amount of water that can be recovered from the flue gas. Concentrated aqueous LiBr is an exceptional desiccant and is able to recover significantly more water than the air-cooled condenser. The max. water recovery case uses a lower desiccant inlet temperature for the column and is consequently able to recover more water than the absorber in the max. heat recovery case, which operates at a higher temperature. MEG has similar properties as LiBr and the MEG absorption process also reaches very high levels of water recovery. On the downside, the MEG absorption process produces water at a lower quality and the recovered water needs to be treated, which leads to some loss of water. The TMC works upon the principle of nano-capillary condensation. The curvature of a meniscus in a pore changes the vapor pressure of a liquid above its surface. Using this principle, the TMC is able to condense water above the flue gas dew point temperature. However, current TMC technology uses pore diameters that only lead to dew point temperature increases of around 3.5 °C, which is not enough to compensate for the increased pinch temperature of the TMC. Thus, the TMC case reaches the lowest water recovery. The results of the water recovery rates are plotted in **Figure 5-108**.

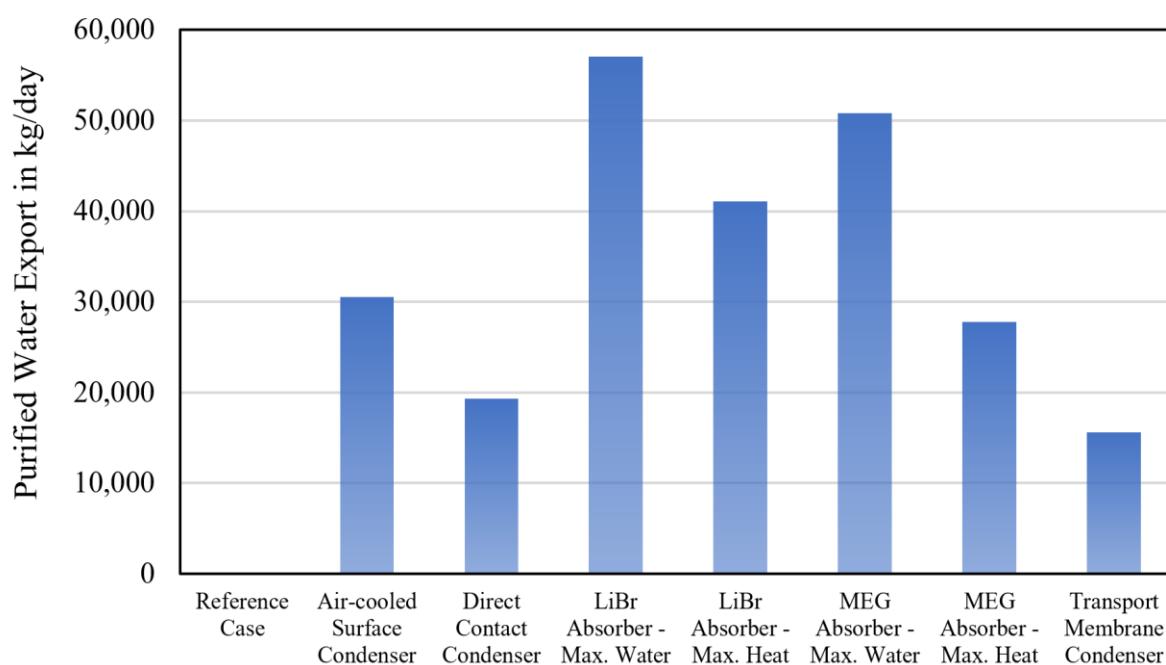


Figure 5-108: Comparison of the water recovery rate of the 10 MW biogas SOFC-GT hybrids with water recovery.

5.2.3 50 MW Natural Gas SOFC-GT Hybrids

Reference Case without Carbon Capture

The design of the 50 MW reference case without carbon capture is based upon the previous 10 MW SOFC-GT hybrid case. The reference case is used to set the baseline performance in order to evaluate the performance of the 50 MW SOFC-GT hybrid with LiBr water recovery. The system analysis is based upon the assumption that the same SOFC modules are used as in the 10 MW cases (to reach mass production) and the 50 MW cases use the same thermal SOFC cell constrains as the previous SOFC-GT hybrid cases, which can be found in **Table 4-13**, together with the other SOFC design parameters. It is challenging to find GTs that match the requirements for SOFC hybridization. Larger engines tend to use axial compressors and exhibit larger pressure ratios; thus, multiple hybrid modules of 10 MW are used to reach the desired scale of 50 MW. Due to this modular approach, performance and economic improvements in the SOFC and GT islands are expected to be minimal; however, economics of scale are relevant to the upstream fuel processing and heat recuperation. An illustration of the 50 MW reference plant design is provided in **Figure 5-109**. The corresponding state-point steam data are shown in **Table 5-62**.

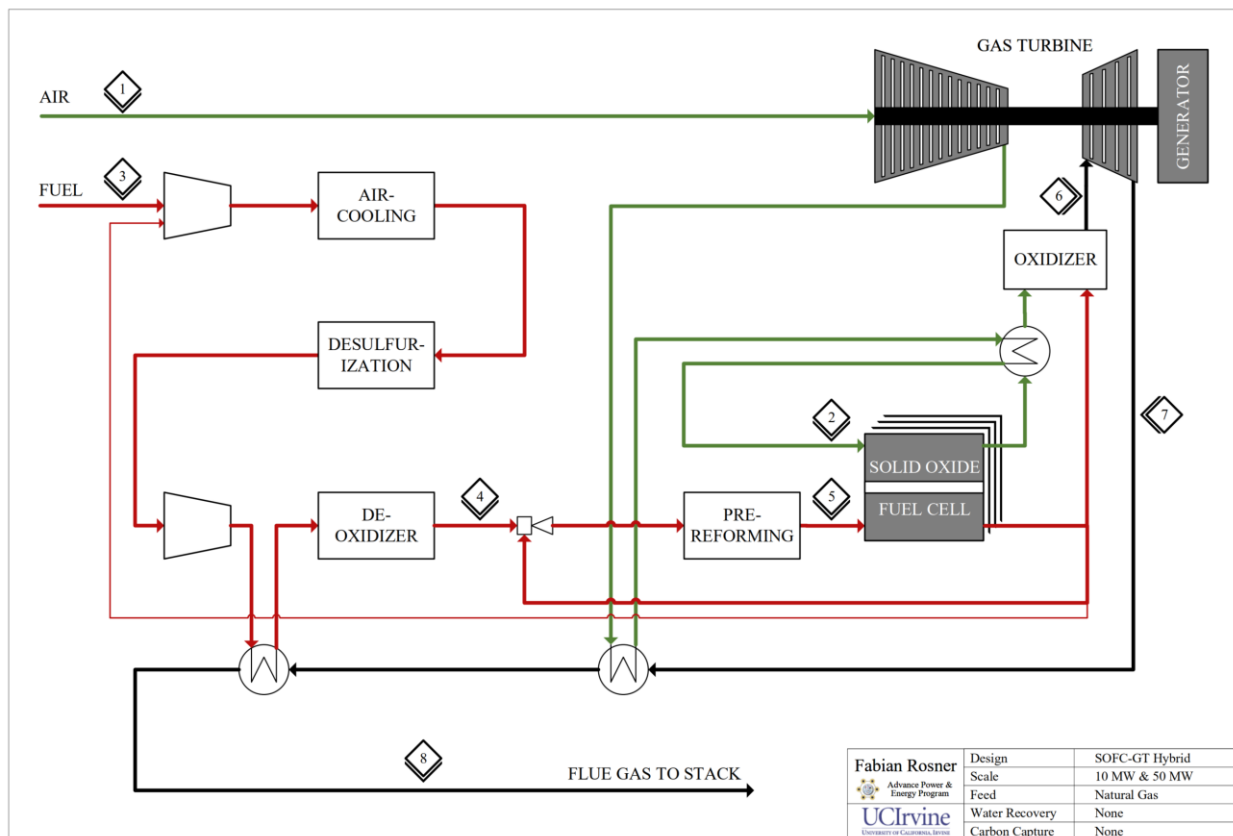


Figure 5-109: Flowsheet of the 50 MW natural gas SOFC-GT hybrid system without carbon capture and without water recovery.

Table 5-62: Stream Summary of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	909	556	326
Pressure	bar	1.01	7.17	4.10	21.25	7.17	6.57	1.05	1.01
Molar Flow Rate	kmol/h	5,382.6	5,382.6	337.9	338.7	1,756.3	5,730.2	5,730.2	5,730.2
Mass Flow Rate	kg/h	155,324	155,324	5,855	5,876	35,734	161,179	161,179	161,179
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07486	0.07486	0.07486
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72718	0.72718	0.72719
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00868
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22862	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09023	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01069	0.20398	0.06174	0.06174	0.06174
H ₂ O	-	0.00987	0.00987	0.00003	0.00165	0.32707	0.12754	0.12754	0.12753
CH ₄	-	0.00000	0.00000	0.93087	0.92860	0.14337	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1

Thermodynamic Performance. The net power output of the nominal 50 MW natural gas reference case is 57,805 kW, whereby, the SOFC produces 53,424 kW of electricity and the five GTs produce another 7,069 kW of electricity. The auxiliary load accounts for 2,687 kW, which includes the DC-AC inverter losses and transformer losses. The DC-AC inverter losses are associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 60% of the total auxiliary load. Second largest auxiliary load is the fuel gas compression, which accounts for 30% of the total auxiliary load, a slight reduction over the 10 MW hybrid due to efficiency gains seen in larger compressors. As a result, the efficiency of the 50 MW reference case is minimally improved and reaches a value of 75.31%-LHV.

The plant design requires no water input as it is equipped with dry cooling technology. Furthermore, no continuous liquid or solid waste streams (does not refer to catalysts and sorbents, which are disposed of after their end-of-life) arise from the operation of this plant. Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. The CO₂ emissions of the 50 MW natural gas hybrid are 269 kg/MWh, which is substantially lower than PC plants with CO₂ emission values of greater than 775 kg/MWh [34], IGCCs (without carbon capture) with values of greater than 750 kg/MWh [34], and state-of-the-art F-class NGCCs (net-efficiency of 59.4%-LHV) with a CO₂ emission value of 336 kg/MWh [34]. The thermodynamic performance results are summarized in **Table 5-63**.

Table 5-63: Thermodynamic Performance of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	53,423.81
Gas Turbine	kW	7,068.56
Total Generation	kW	60,492.37
Auxiliary Load		
DC-AC Inverter Losses	kW	1,602.72
Fuel Gas Compression	kW	817.42
Fuel Gas Intercooler	kW	0.11
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	39.25
Transformer Losses	kW	227.47
Total Auxiliary Load	kW	2,686.97
Net Power Generation	kW	57,805.40
Plant Performance		
Net Efficiency	%-HHV	67.91
Net Efficiency	%-LHV	75.31
Consumables		
Fuel Gas	kg/h	5,855
Thermal Input	MJ _{LHV} /h	276,226
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	15,570
Carbon Emission	kg/MWh	269
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Economics. The economic analysis follows the methodology outlined earlier. The TPC of the 50 MW NG hybrid without water recovery is \$122,403,960 and the specific plant cost is 2,118 \$/kW, a reduction of 196 \$/kW compared to the 10 MW NG reference case. This reduction in specific cost is mostly related to an equipment cost reduction of the gas processing equipment and the heat recuperation equipment since SOFC and GT costs as well as the plant efficiency remain relatively constant. The SOFC island accounts for \$82,809,570 and has a TPC share of 67.7%, a relative increase compared to the 10MW case due to the reduced gas processing and heat recuperation cost. The TPC for the SOFC island includes: stack cost, power conditioning, pressure vessel, housing and assembly, and system installation. The SOFC stack with \$31,654,340 is the single most expensive category of the SOFC island followed by the power conditioning system, which includes the inverters, with \$31,436,800. The GT island contributes \$18,255,070 to the plant's TPC, which is an increase by a factor of five compared to the 10 MW NG case as five of the same Dresser Rand GTs are used in the 50 MW cases. Fuel gas processing represents about 2.0% of the TPC with \$2,419,060 (in the 50 MW NG cases, the NG supply pressure remains at 4.1 bar). Heat recuperation

accounts for 2.5% of the plant's TPC, which is a significant reduction versus the 10 MW NG case. Auxiliary plant equipment costs are estimated with \$15,837,850, which includes accessory electric plant, instrumentation and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-110**. A more detailed overview of the capital cost is provided in **Table 5-64**.

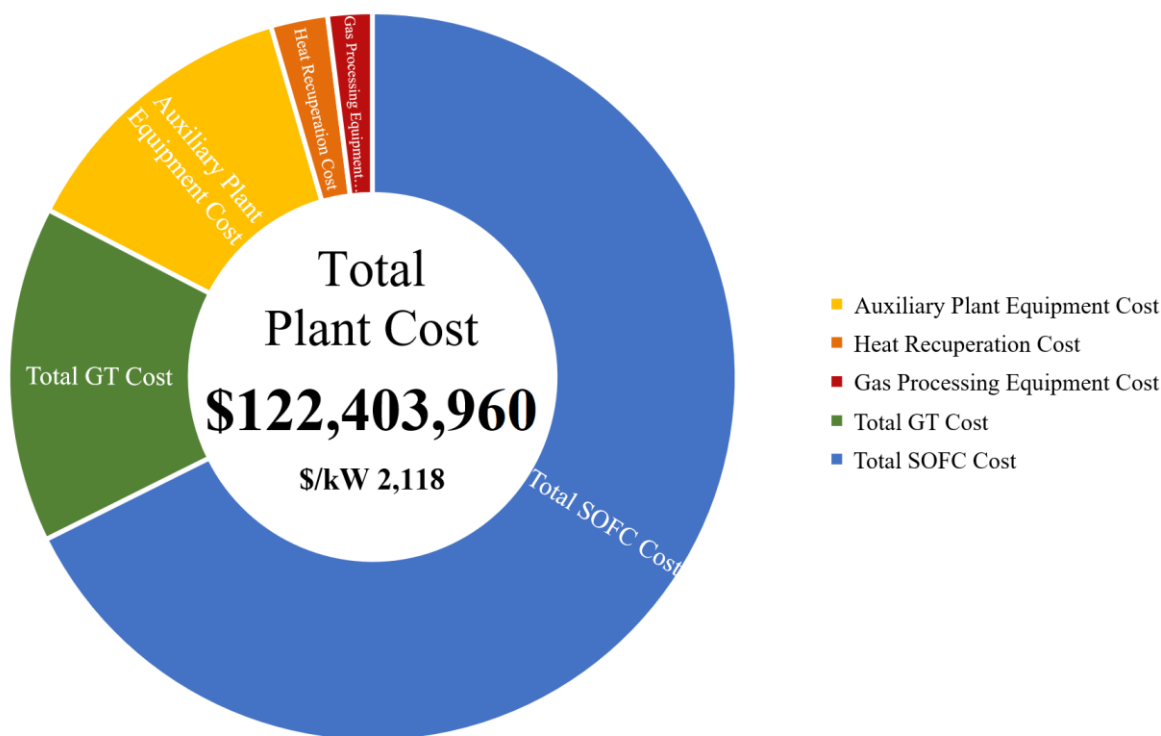


Figure 5-110: Total plant cost and specific plant cost of the 50 MW natural gas SOFC-GT hybrid without carbon capture and without water recovery.

The total overnight capital cost is \$150,996,420 and is comprised of the TPC, the preproduction costs with \$4,933,791, the inventory capital costs with \$633,450 and other costs including financing and land costs with \$23,025,217. Variable annual costs are mostly fuel costs with \$8,680,737. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, and waste disposal costs, which amount to \$2,956,600 per year. Fixed costs are comprised of annual operating labor with \$4,022,019 and costs associated with tax and insurance with \$2,448,079 per year. The resulting COE for the 50 MW SOFC-GT hybrid plant without water recovery is 64.58 \$/MWh. The economic results are summarized in **Table 5-65**. In the detailed breakdown of the COE, provided in **Figure 5-111**, it can be seen that the overnight capital cost is slightly reduced when compared to the 10 MW reference case due to the economies of scale of the balance-of-plant equipment. However, the largest cost saving factor is the operating labor cost. Due to the lower number of operators (on a relative basis) a substantial reduction in the COE can be achieved.

Table 5-64: Equipment Cost of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	82,809,570
SOFC Stacks	\$	31,654,340
Power Conditioning	\$	31,436,800
Pressure Vessel, Housing & Assembly	\$	17,196,570
System Installation	\$	2,521,860
Gas Turbine Island	\$	18,255,070
Gas Turbine-Generator Set	\$	15,388,010
Gas Turbine Foundation	\$	223,690
Piping, Instrumentation & Electrical Equipment	\$	2,643,380
Gas Processing and Flue Gas Stack	\$	2,419,060
Fuel Compressors and Intercooler	\$	1,519,020
Fuel Processing	\$	658,640
Flue Gas Stack System	\$	241,400
Heat Recuperation	\$	3,082,410
Cathode Off-gas Heat Exchanger	\$	1,448,180
Air Preheater	\$	1,337,600
Fuel Preheater	\$	296,640
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	0
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	15,837,850
Accessory Electric Plant	\$	5,806,900
Instrumentation & Controls	\$	4,494,920
Improvement to Site	\$	1,168,680
Buildings & Structures	\$	4,367,350
Total Plant Cost	\$	122,403,960
Specific Plant Cost	\$/kW	2,118

Table 5-65: Economics of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	122,403,960
Preproduction Costs		
6 Months Labor Cost	\$	2,011,010
1 Month Maintenance Materials	\$	137,990
1 Month Non-Fuel Consumables	\$	135,690
1 Month Waste Disposal	\$	90
25% of 1 month fuel cost at 100% CF	\$	200,940
2% of TPC	\$	2,448,080
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	21,430
0.5% of TPC (Spare Parts)	\$	612,020
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	305,210
Land	\$	1,054,510
Other Owner's Costs (15% of TPC)	\$	18,360,590
Financing Costs	\$	3,304,910
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	8,680,740
Other Annual Variable Operating Cost at above CF for Initial Year	\$	2,956,600
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	0
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	4,022,020
Property Tax & Insurance	\$	2,448,080
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	64.58
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	N/A

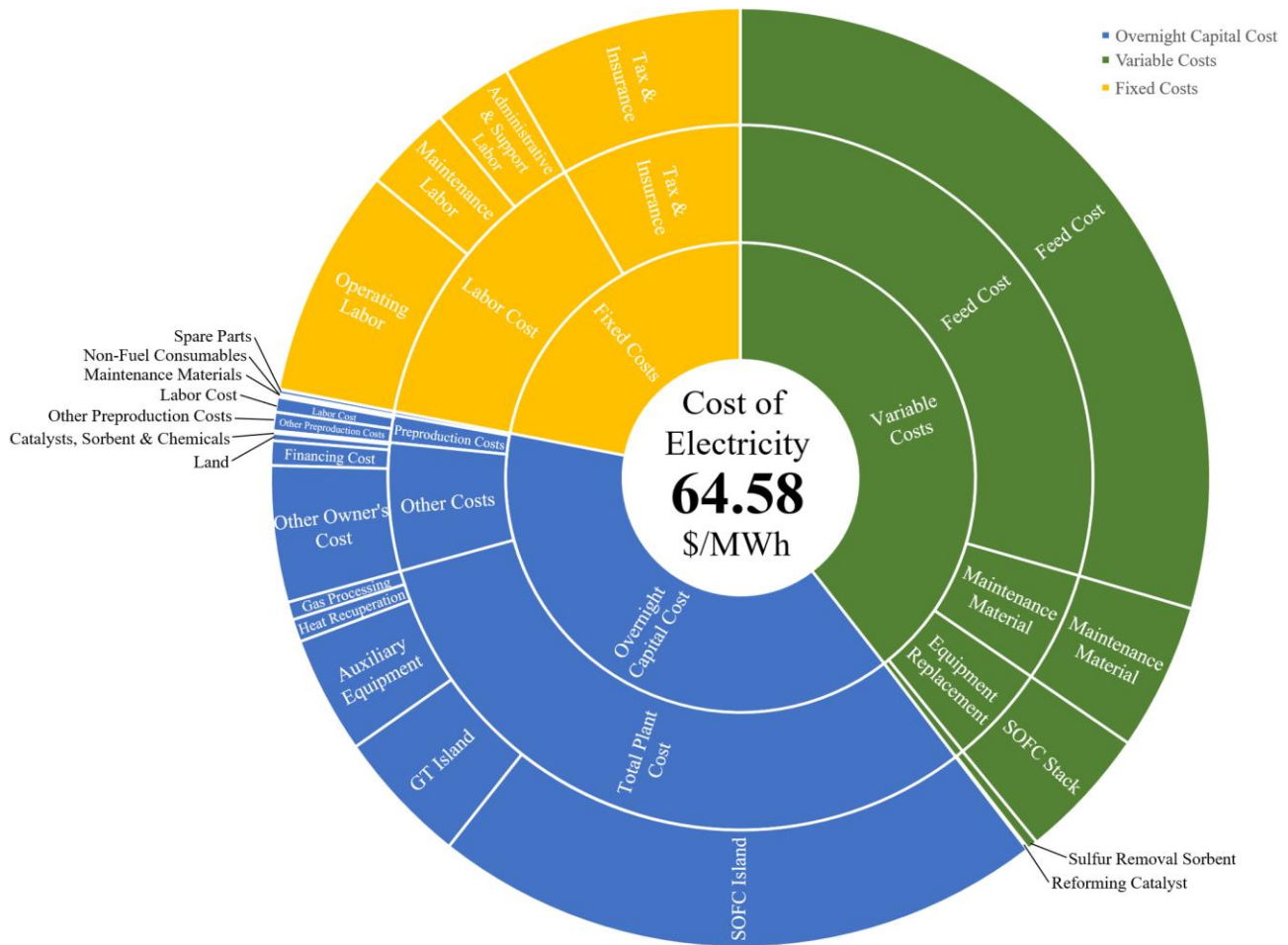


Figure 5-111: Breakdown of the cost-of-electricity of the 50 MW natural gas SOFC-GT hybrid without carbon capture and without water recovery.

LiBr Absorption System without Carbon Capture

The configuration of the 50 MW NG hybrid plant with LiBr absorption system follows the sparing philosophy of the 50 MW NG hybrid reference case. The LiBr absorption system is designed as single train unit and its design is identical to the 10 MW LiBr absorber case. A flue gas heat exchanger and a packed-bed column are installed downstream of the fuel pre-heater. The packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream heat exchanger is used to provide the heat for sorbent regeneration, while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, an air-cooled condenser, supported by a vacuum compressor to remove any non-condensable gases from the

vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is identical to the 10 MW NG LiBr absorber case and a similar plume rise can be expected. The flue gas from the LiBr absorber case differentiates from the air-cooled condenser and direct contact condenser cases as the flue gas in the LiBr condenser case is leaving the stack above its dew point temperature minimizing the risk of plume formation (flue gas dew point temperature around 14 °C). A general flow diagram of the 10MW NG case with LiBr absorption water recovery is shown in **Figure 5-112**. The respective state-point stream data are provided in **Table 5-66**.

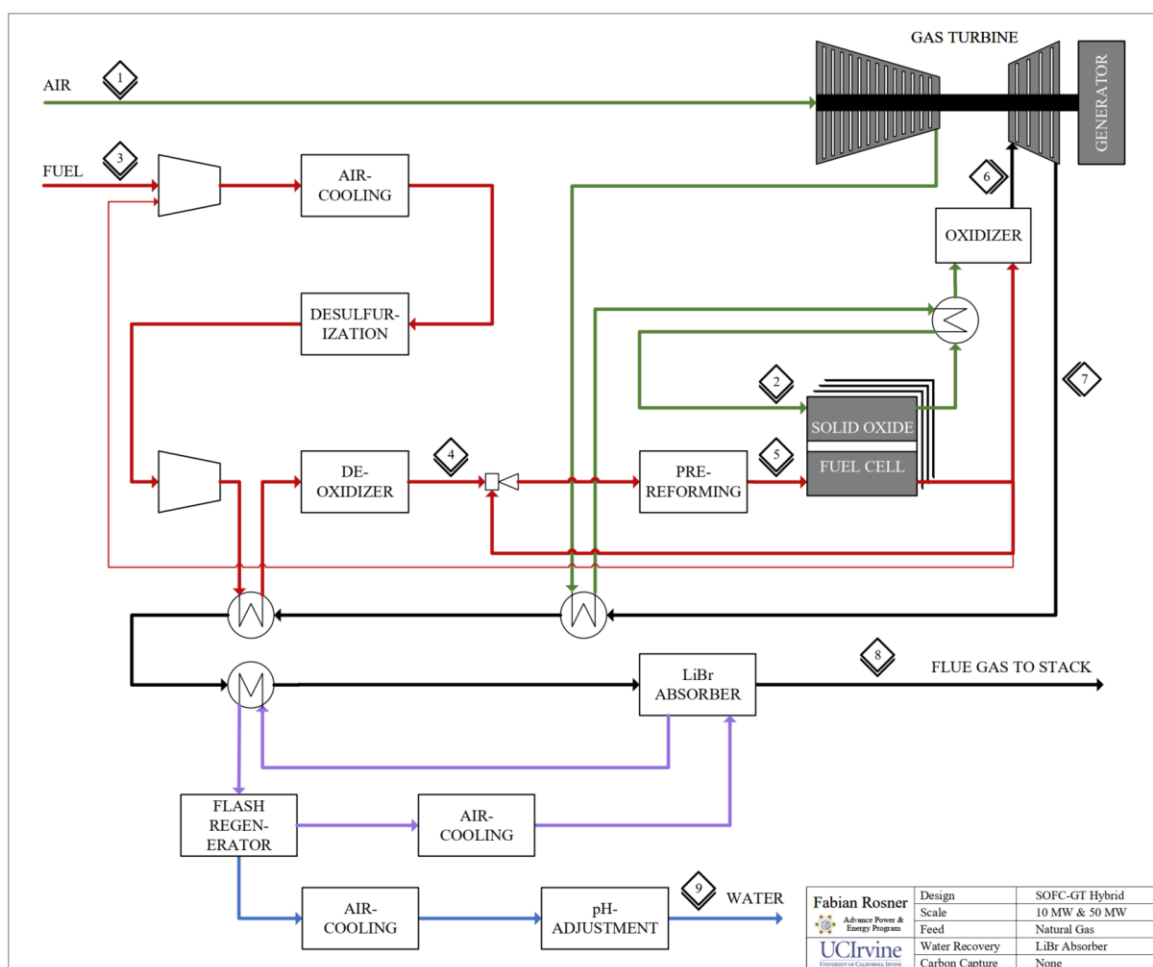


Figure 5-112: Flowsheet of the 50 MW natural gas SOFC-GT hybrid system without carbon capture and with LiBr absorption system for water recovery.

Table 5-66: Stream Summary of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	322	700	910	562	49	40
Pressure	bar	1.01	7.21	4.10	21.33	7.21	6.60	1.08	1.01	2.00
Molar Flow Rate	kmol/h	5,371.9	5,371.9	335.2	336.0	1,745.1	5,716.9	5,716.9	5,070.4	646.5
Mass Flow Rate	kg/h	155,018	155,018	5,809	5,830	35,494	160,827	160,827	149,181	11,648
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07559	0.07559	0.08523	Trace
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.72744	0.72744	0.82018	Trace
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00868	0.00868	0.00979	Trace
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22898	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09042	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01068	0.20386	0.06140	0.06140	0.06923	Trace
H ₂ O	-	0.00987	0.00987	0.00003	0.00164	0.32685	0.12689	0.12689	0.01557	1.00000
CH ₄	-	0.00000	0.00000	0.93087	0.92862	0.14317	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	1	1	0.99999	1	1	1	1

Thermodynamic Performance. The net power output of the nominal 50 MW NG plant with LiBr absorber for water recovery is 56,611 kW. The SOFC island produces 52,989 kW_{DC} and the GT generates an additional 6,721 kW. Due to the pressure drop associated with the LiBr absorber, the GT power output is lower than in the reference case as previously explained. The GT surge margin in the 50 MW LiBr absorber case is identical to the 10 MW case (10.8%) and the SOFC operates at an operating pressure of 7.21 bar at a FU of 88.1%. The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 3,099 kW and represent about 14.0% of the auxiliary load.

The electrical efficiency of the LiBr absorption case decreased by 0.97% points to 74.34%-LHV compared to the 50 MW reference case without carbon capture and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GT due to the backpressure.

The plant with LiBr absorber emits 15,449 kg/h of CO₂, which translates to 273 kg/MWh. The water recovery from the flue gas in the LiBr absorber case is 89.1%. Sorbent-based water recovery systems are independent of the flue gas's dew point temperature and can remove substantial quantities of water. Simultaneously, they are less dependent upon ambient temperatures compared to air-cooled condensers and direct contact condensers. While the ambient temperature still plays a role in the water recovery of the LiBr absorption system by setting the condenser backpressure (pressure in the regenerator), this effect can be moderated by the heat input required for sorbent regeneration. At the scale of this nominal 50 MW plant, a total of 279,545 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S.

SOFC-GT Hybrid Systems with Water Recovery

used 1.74 l/m² of irrigation water per day [341]. Thus, water from this 50 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 144,500 m² of farmland (including 90% capacity factor). The high water quality of the recovered water minimizes water treatment and eliminates any wastewater streams from the plant, which reduces waste disposal costs but also maximizes water export. The plant performance data are summarized in **Table 5-67**.

Table 5-67: Thermodynamic Performance of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	52,988.57
Gas Turbine	kW	6,721.38
Total Generation	kW	59,709.95
Auxiliary Load		
DC-AC Inverter Losses	kW	1589.66
Fuel Gas Compression	kW	811.33
Fuel Gas Intercooler	kW	0.11
Water Recovery System	kW	430.68
Water Treatment	kW	3.53
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	38.74
Transformer Losses	kW	224.53
Total Auxiliary Load	kW	3098.58
Net Power Generation	kW	56,611.38
Plant Performance		
Net Efficiency	%-HHV	67.03
Net Efficiency	%-LHV	74.34
Consumables		
Fuel Gas	kg/h	5,809
Thermal Input	MJ _{LHV} /h	274,060
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	N/A
Carbon Emission	kg/h	15,449
Carbon Emission	kg/MWh	273
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	89.1
Flue Gas Water Recovery	kg/d	279,545

Economics. The TPC of the 50 MW hybrid with LiBr absorber for water recovery is \$127,469,970 resulting in a specific plant cost of 2,252 \$/kW. The SOFC island costs are very similar to the 50 MW reference case and the slight variations result from the plant scale, number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the discussion sections of the 10 MW cases, which are still valid for this case. The GT island cost remains constant at \$18,255,070 since the same GT is used in all study cases. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. The LiBr absorption water recovery system costs \$4,783,440 and can make

use of the economies of scale. Considering that the water recovery is increased by a factor of five (compared to the 10 MW LiBr case) the equipment cost only increased by a factor of 2.16. Similarly, the water treatment equipment cost increased by a factor of 1.63 to \$200,440, while treating five times more water than in the 10 MW LiBr case. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-113**. A more detailed cost summary of all plant equipment is provided in **Table 5-68**.

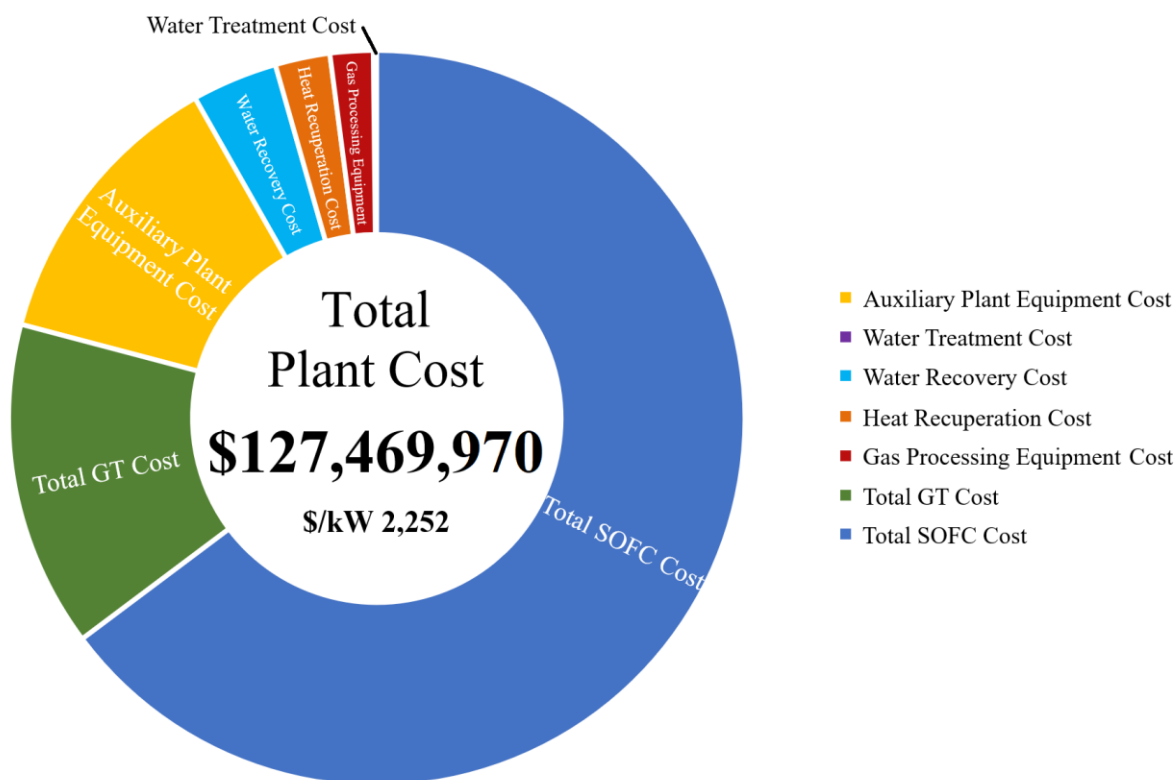


Figure 5-113: Total plant cost and specific plant cost of the 50 MW natural gas SOFC-GT hybrid without carbon capture and with LiBr absorption system for water recovery.

The overnight capital cost in the 50 MW LiBr absorber case is \$159,249,680, an increase of 5.5% compared to the 50 MW reference case. This increase is mostly related to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. Furthermore, the LiBr absorber system requires an initial fill and continuous LiBr makeup. The 55 wt-% LiBr solution is assumed to cost 2.04 \$₂₀₁₅/kg and the initial fill costs about \$1,935,130. The LiBr loss is very small and entirely due to entrainment since LiBr is a salt with extremely low vapor pressure. With a mist eliminator, the LiBr losses are expected to be around 50 kg per year. In the LiBr absorber case, water byproduct sales reduce the variable operating costs by \$1,646,850 per year.

Table 5-68: Equipment Cost of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	82,576,090
SOFC Stacks	\$	31,661,980
Power Conditioning	\$	31,165,120
Pressure Vessel, Housing & Assembly	\$	17,232,010
System Installation	\$	2,516,980
Gas Turbine Island	\$	18,255,070
Gas Turbine-Generator Set	\$	15,388,010
Gas Turbine Foundation	\$	223,690
Piping, Instrumentation & Electrical Equipment	\$	2,643,380
Gas Processing and Flue Gas Stack	\$	2,398,360
Fuel Compressors and Intercooler	\$	1,501,320
Fuel Processing	\$	655,950
Flue Gas Stack System	\$	241,090
Heat Recuperation	\$	3,078,670
Cathode Off-gas Heat Exchanger	\$	1,439,830
Air Preheater	\$	1,341,760
Fuel Preheater	\$	297,080
Carbon Capture	\$	N/A
Combined WGS & Carbon Capture PSA	\$	N/A
CO ₂ Compression and Dehydration	\$	N/A
Water Recovery	\$	4,983,880
Water Recovery System	\$	4,783,440
Water Treatment	\$	200,440
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	16,177,900
Accessory Electric Plant	\$	5,731,790
Instrumentation & Controls	\$	4,680,950
Improvement to Site	\$	1,217,050
Buildings & Structures	\$	4,548,110
Total Plant Cost	\$	127,469,970
Specific Plant Cost	\$/kW	2,252

The fixed operating cost in the LiBr absorber case increases due to tax and insurance costs of the additional equipment. The operating labor costs of the LiBr absorber unit are identical to 10 MW cases; however, when comparing the labor cost on a relative basis, the labor cost in the 50 MW LiBr absorber case is significantly lower due to the five times higher water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the 50 MW reference case. Thus, the COE remains constant at 64.58 \$/MWh. The COW in the LiBr absorber case is 0.0179 \$/L, a reduction of over 44.3% versus the 10 MW LiBr case. A summary of the economic key parameters can be found in **Table 5-69**. A detailed breakdown of the COE is shown in **Figure 5-114**.

Figure 5-114 shows the COE value of 64.58 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-3.69 \$/MWh) and the whole circle represents the COE without byproduct credit (68.27 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 68.27 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at

Table 5-69: Economics of the 50 MW Natural Gas SOFC-GT Hybrid without Carbon Capture and with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
	Plant Capacity Factor	%
		90
Total Overnight Capital Cost		
Total Plant Cost	\$	127,469,970
Preproduction Costs		
6 Months Labor Cost	\$	2,165,290
1 Month Maintenance Materials	\$	146,190
1 Month Non-Fuel Consumables	\$	134,680
1 Month Waste Disposal	\$	80
25% of 1 month fuel cost at 100% CF	\$	199,370
2% of TPC	\$	2,549,400
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	21,300
0.5% of TPC (Spare Parts)	\$	637,350
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	2,238,490
Land	\$	1,125,380
Other Owner's Costs (15% of TPC)	\$	19,120,490
Financing Costs	\$	3,441,690
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	8,612,680
Other Annual Variable Operating Cost at above CF for Initial Year	\$	3,034,370
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	0
Annual Byproduct Revenues at above CF for Initial Year	\$	1,646,850
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	4,330,570
Property Tax & Insurance	\$	2,549,400
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	64.58
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	N/A
First Year Levelized Cost of Water	\$/L	0.0179

0.0179 \$/L in order to reach the COE of 64.58 \$/MWh.

Figure 5-115 shows a detailed breakdown of the COW. The fixed operating cost in the 50 MW LiBr absorber case is substantially reduced compared to the 10 MW case since the water recovery unit in both cases requires the same number of operators but in the 50 MW case, the amount of recovered water is about five times higher reducing the relative operating labor cost to about 24.7% of the COW. Consequently, the performance penalty cost becomes more dominant with 31.3%. System integration losses account for 18.0% and the parasitic load of the water recovery (mostly desiccant and water pumping) accounts for 13.2% of the COW. The TPC's share on the COW increased from 22.7% (10 MW LiBr) to 23.7% in this case despite the capital cost reduction on a per liter-of-recovered-water basis, which is a result of the substantial reduction of the fixed operating cost (shares are displayed on a relative basis). The water treatment is responsible for only 2.0% of the COW.

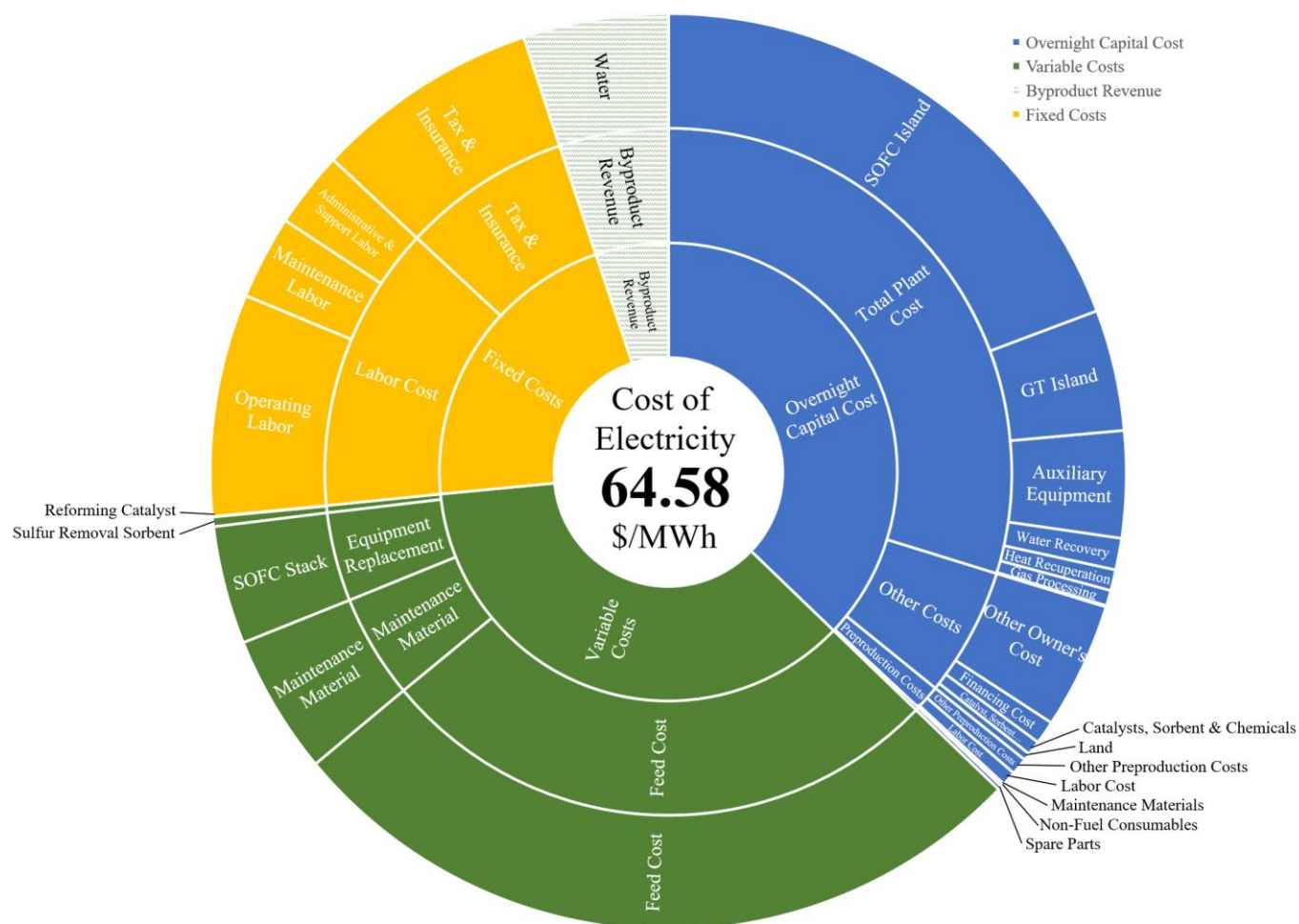


Figure 5-114: Breakdown of the cost-of-electricity of the 50 MW natural gas SOFC-GT hybrid without carbon capture and with LiBr absorber for water recovery.

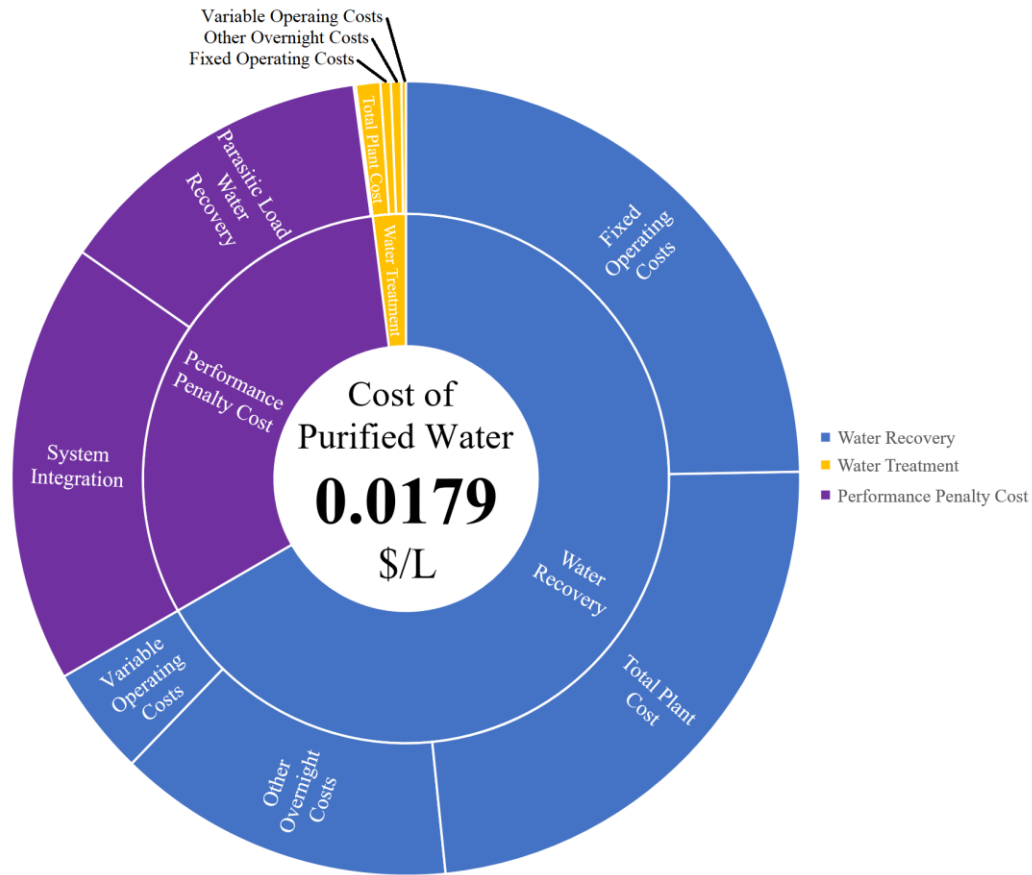


Figure 5-115: Breakdown of the cost-of-water of the 50 MW natural gas SOFC-GT hybrid without carbon capture and with LiBr absorber for water recovery.

Reference Case with Carbon Capture

The design of the 50 MW reference case with carbon capture is similar to the 50 MW SOFC-GT hybrid case without carbon capture; however, additional equipment for carbon removal from the anode of gas is installed in between the SOFC and GT oxidizer. Internal reforming inside the SOFC makes it advantageous to capture the CO₂ from the anode off-gas before the fuel is oxidized and mixed with large quantities of diluents, such as N₂. The anode off-gas is essentially free of methane and all the carbon is present in the forms of CO and CO₂ eliminating the need for a reformer. To increase the amount of CO₂ capture, the remaining CO in the anode off-gas is shifted to CO₂ in a combined WGS-CO₂ PSA reactor, which can operate at temperatures around 200 °C. The sorbent is regenerated via a steam purge and the water vapor is later recovered during compression. Before the CO₂ can be sequestered further dehydration is required in a TEG dehydration unit to avoid pipeline corrosion. The CO₂ export stream still contains some anode-off gas constituents, such as H₂, which could be converted to heat if needed in a catalytic combustor; however, this might only be a viable option for larger applications and is not considered in this work. After the CO₂

SOFC-GT Hybrid Systems with Water Recovery

removal unit, the anode off-gas is rich in H_2 , which is now converted to H_2O in the GT oxidizer. This reference case is used to set the baseline performance in order to evaluate the performance of the 50 MW carbon capture SOFC-GT hybrid with LiBr water recovery. The configuration of the 50 MW reference plant with carbon capture is provided in **Figure 5-116**. The corresponding state-point steam data are shown in **Table 5-70**.

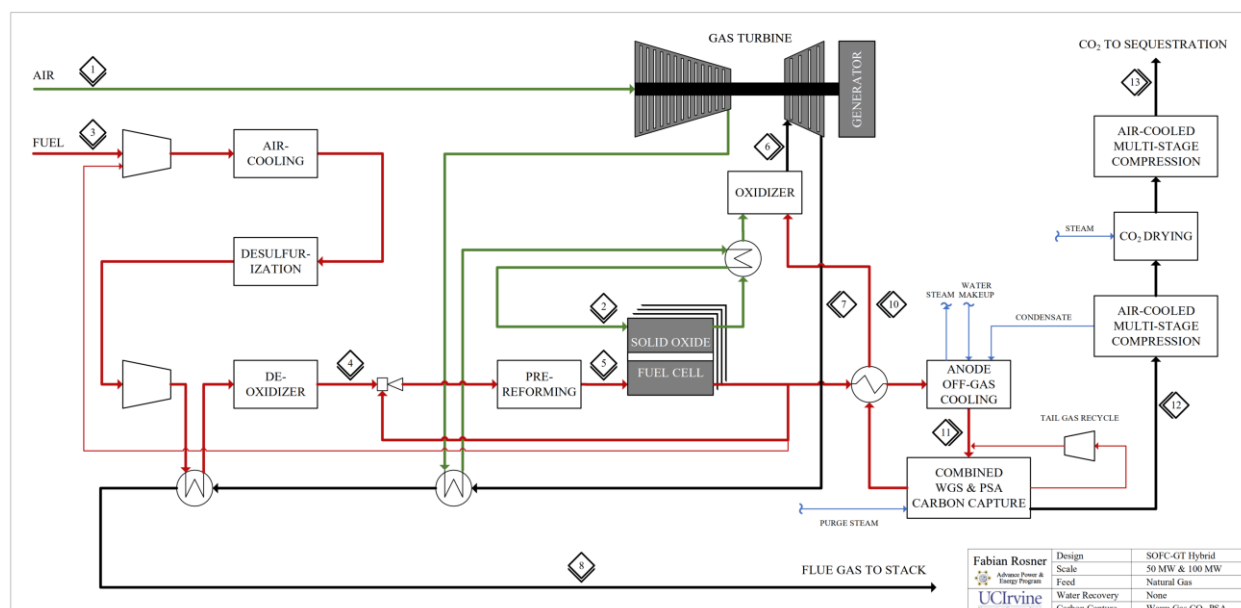


Figure 5-116: Flowsheet of the 50 MW natural gas SOFC-GT hybrid system with carbon capture and without water recovery.

Table 5-70: Stream Summary of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	926	604	322
Pressure	bar	1.01	6.84	4.10	19.87	6.84	5.07	1.05	1.01
Molar Flow Rate	kmol/h	5,475.0	5,475.0	350.5	351.3	1,833.6	5,493.9	5,493.9	5,493.9
Mass Flow Rate	kg/h	157,992	157,992	6,074	6,095	37,169	149,081	149,081	149,081
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07726	0.07726	0.07726
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00671	0.77149	0.77149	0.77149
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00921	0.00921	0.00921
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.23222	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09188	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01064	0.20261	0.00480	0.00480	0.00480
H ₂ O	-	0.00987	0.00987	0.00003	0.00156	0.32408	0.13725	0.13725	0.13725
CH ₄	-	0.00000	0.00000	0.93087	0.92873	0.14249	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	0.00001	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	1	1	1	1
Stream Number	Unit	9	10	11	12	13			
Temperature	°C	-	769	165	209	40			
Pressure	bar	-	5.27	6.16	1.19	151.65			
Molar Flow Rate	kmol/h	-	729.5	1,043.5	673.2	332.9			
Mass Flow Rate	kg/h	-	11,054	25,316	20,941	14,477			
Mass Vapor Fraction	-	-	1	1	1	1			
Composition	mole-basis								
O ₂	-	-	0.00000	0.00000	0.00000	0.00000			
N ₂	-	-	0.00751	0.00522	0.00019	0.00038			
Ar	-	-	0.00000	0.00000	0.00000	0.00000			
H ₂	-	-	0.23212	0.10378	0.00597	0.01201			
CO	-	-	0.00551	0.06151	0.00008	0.00016			
CO ₂	-	-	0.02778	0.27849	0.50721	0.98703			
H ₂ O	-	-	0.72674	0.55076	0.48654	0.00041			
CH ₄	-	-	0.00033	0.00023	Trace	0.00002			
C ₂ H ₆	-	-	0.00001	Trace	Trace	Trace			
C ₃ H ₈	-	-	0.00000	0.00000	0.00000	0.00000			
C ₄ H ₁₀	-	-	Trace	Trace	Trace	Trace			
H ₂ S	-	-	Trace	Trace	Trace	Trace			
C ₄ H ₁₀ S	-	-	Trace	Trace	Trace	Trace			
Total	-	-	1	0.99999	0.99999	1			

Thermodynamic Performance. The net power output of the nominal 50 MW NG reference case with carbon capture is 54,324 kW, whereby, the SOFC produces 55,030 kW of electricity at a FU of 87.8%. This a 3.0% increase in SOFC power over the 50 MW reference case without carbon capture and is due to the GT operating condition, which allows the GT compressor to provide more air to the SOFC and leads to a scaleup of the SOFC island. The five Dresser Rand GTs provide another 4,157 kW of electricity, which is substantially lower than in the 50 MW reference case without carbon capture. The reason for this decrease in GT power output is the reduced mass flow through the turbine section as a consequence of CO₂ removal. This reduction in power output is obtained despite the fact that the TIT in the carbon capture case increased from 909.3 °C to 926.4 °C (lower H₂O+CO₂ mole fraction allows for higher TIT, while maintaining the

same blade metal temperature). Overall, this leads to a 1,305 kW lower gross power generation in the scenario with carbon capture. Although, carbon capture reduces the GT power output, it has a positive impact on the GT's surge margin, which increased to 18.7% in this carbon capture case.

The auxiliary load accounts for 4,863 kW, which includes the DC-AC inverter losses and transformer losses. The DC-AC inverter losses are associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 34.0% of the total auxiliary load. The largest auxiliary load is introduced by the CO₂ removal unit and CO₂ compression, whereby, the CO₂ compression power accounts for 84.2% of the CO₂ Capture and Compression load. Fuel gas compression accounts for 16.8% of the total auxiliary load. As a result, the efficiency of the 50 MW reference case with carbon capture is reduced to 68.22%-LHV.

Table 5-71: Thermodynamic Performance of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	55,029.53
Gas Turbine	kW	4,157.11
Total Generation	kW	59,186.64
Auxiliary Load		
DC-AC Inverter Losses	kW	1,650.89
Fuel Gas Compression	kW	819.15
Fuel Gas Intercooler	kW	0.11
CO ₂ Capture and Compression	kW	2,131.59
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	38.40
Transformer Losses	kW	222.56
Total Auxiliary Load	kW	4,862.70
Net Power Generation	kW	54,323.94
Plant Performance		
Net Efficiency	%-HHV	61.52
Net Efficiency	%-LHV	68.22
Consumables		
Fuel Gas	kg/h	6,074
Thermal Input	MJ _{LHV} /h	286,572
Water	kg/h	126
Emissions and Waste Streams		
Carbon Capture	%	90
Carbon Emission	kg/h	1,688
Carbon Emission	kg/MWh	31
Liquid Waste	kg/h	63
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Although the plant is equipped with dry cooling technology, CO₂ removal is water intensive and requires a net water input of 126 kg/h. At the same time, the treatment of the imported water generates about 63 kg/h of wastewater, which has to be disposed. Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. The CO₂ emissions of the 50 MW natural gas hybrid with 90% carbon capture are 31 kg/MWh. The thermodynamic performance results are summarized in **Table 5-71**.

Economics. The TPC of the 50 MW NG hybrid with carbon capture and without water recovery is \$140,010,970 and the specific plant cost is 2,577 \$/kW, an increase of 459 \$/kW compared to the 50 MW NG case without carbon capture. This increase in specific cost is mostly related to the carbon capture equipment, which is responsible for 9.4% of the TPC, as well as the reduction in net power output. The SOFC island accounts for \$85,770,050 and scales with the plant size (GT air supply). Other factors such as SOFC operating pressure and spare stacks play a small role in this cost increase (compared to 50 MW reference case without carbon capture). The GT island contributes \$18,255,070 to the plant's TPC, which is identical to the 50 MW cases without carbon capture. Fuel gas processing represents about 1.8% of the TPC with \$2,467,740 (in the 50 MW NG cases, the NG supply pressure remains at 4.1 bar) and heat recuperation accounts for 2.2% of the plant's TPC. Auxiliary plant equipment costs are estimated with

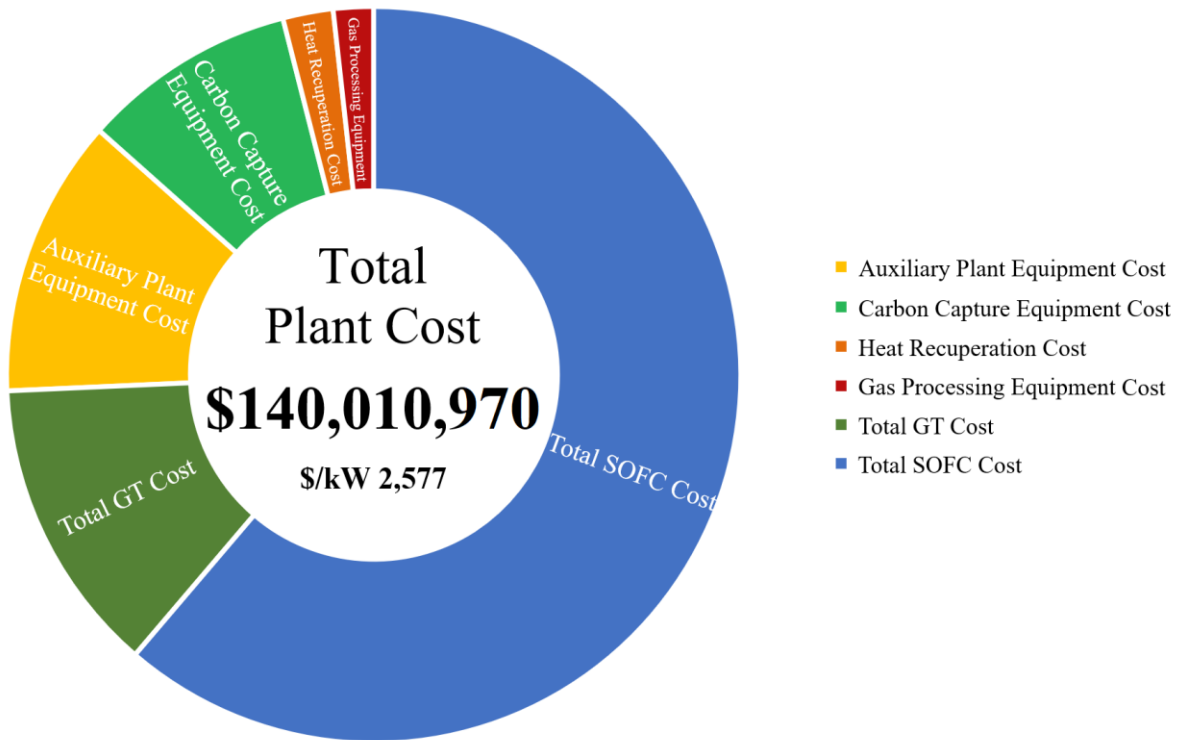


Figure 5-117: Total plant cost and specific plant cost of the 50 MW natural gas SOFC-GT hybrid with carbon capture and without water recovery.

SOFC-GT Hybrid Systems with Water Recovery

\$17,155,400, which includes accessory electric plant, instrumentation and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-117**. A more detailed overview of the capital cost is provided in **Table 5-72**.

Table 5-72: Equipment Cost of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	85,770,050
SOFC Stacks	\$	33,037,060
Power Conditioning	\$	32,269,310
Pressure Vessel, Housing & Assembly	\$	17,847,430
System Installation	\$	2,616,260
Gas Turbine Island	\$	18,255,070
Gas Turbine-Generator Set	\$	15,388,010
Gas Turbine Foundation	\$	223,690
Piping, Instrumentation & Electrical Equipment	\$	2,643,380
Gas Processing and Flue Gas Stack	\$	2,467,740
Fuel Compressors and Intercooler	\$	1,566,030
Fuel Processing	\$	671,350
Flue Gas Stack System	\$	230,360
Heat Recuperation	\$	3,099,160
Cathode Off-gas Heat Exchanger	\$	1,319,340
Air Preheater	\$	1,468,370
Fuel Preheater	\$	311,450
Carbon Capture	\$	13,263,550
Combined WGS & Carbon Capture PSA	\$	8,588,590
CO ₂ Compression and Dehydration	\$	4,674,960
Water Recovery	\$	N/A
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	17,155,400
Accessory Electric Plant	\$	5,681,550
Instrumentation & Controls	\$	5,141,490
Improvement to Site	\$	1,336,790
Buildings & Structures	\$	4,995,570
Total Plant Cost	\$	140,010,970
Specific Plant Cost	\$/kW	2,577

The total overnight capital cost is \$173,632,920 and is comprised of the TPC, the preproduction costs with \$5,793,860, the inventory capital costs with \$765,060 and other costs including financing and land costs with \$27,063,030. Variable annual costs are mostly fuel costs with \$9,005,890. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, and waste disposal costs, which amount to \$3,499,410 per year. Annual CO₂ transporting, storing, and monitoring account for \$3,402,530. Fixed costs are comprised of annual operating labor with \$4,922,300 and costs associated with tax and insurance with \$2,800,220 per year. The resulting COE for the 50 MW SOFC-GT hybrid plant with carbon capture and without water recovery is 85.58 \$/MWh. The economic results are summarized in **Table 5-73**. In the detailed breakdown of the COE, provided in **Figure 5-118**, it can be seen that the variable costs significantly increase. Just the TS&M costs increase the COE by

7.94 \$/MWh. Carbon capture TPC increases the COE by another 2.32 \$/MWh. Other COE driving factors are the additional operating labor and reduced efficiency.

Table 5-73: Economics of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	140,010,970
Preproduction Costs		
6 Months Labor Cost	\$	2,461,150
1 Month Maintenance Materials	\$	159,460
1 Month Non-Fuel Consumables	\$	162,830
1 Month Waste Disposal	\$	1,730
25% of 1 month fuel cost at 100% CF	\$	208,470
2% of TPC	\$	2,800,220
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	65,010
0.5% of TPC (Spare Parts)	\$	700,050
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	1,050,330
Land	\$	1,230,760
Other Owner's Costs (15% of TPC)	\$	21,001,650
Financing Costs	\$	3,780,300
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	9,005,890
Other Annual Variable Operating Cost at above CF for Initial Year	\$	3,499,410
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	3,402,520
Annual Byproduct Revenues at above CF for Initial Year	\$	0
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	4,922,300
Property Tax & Insurance	\$	2,800,220
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.63
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	85.58
First Year Levelized Cost of Water	\$/L	N/A

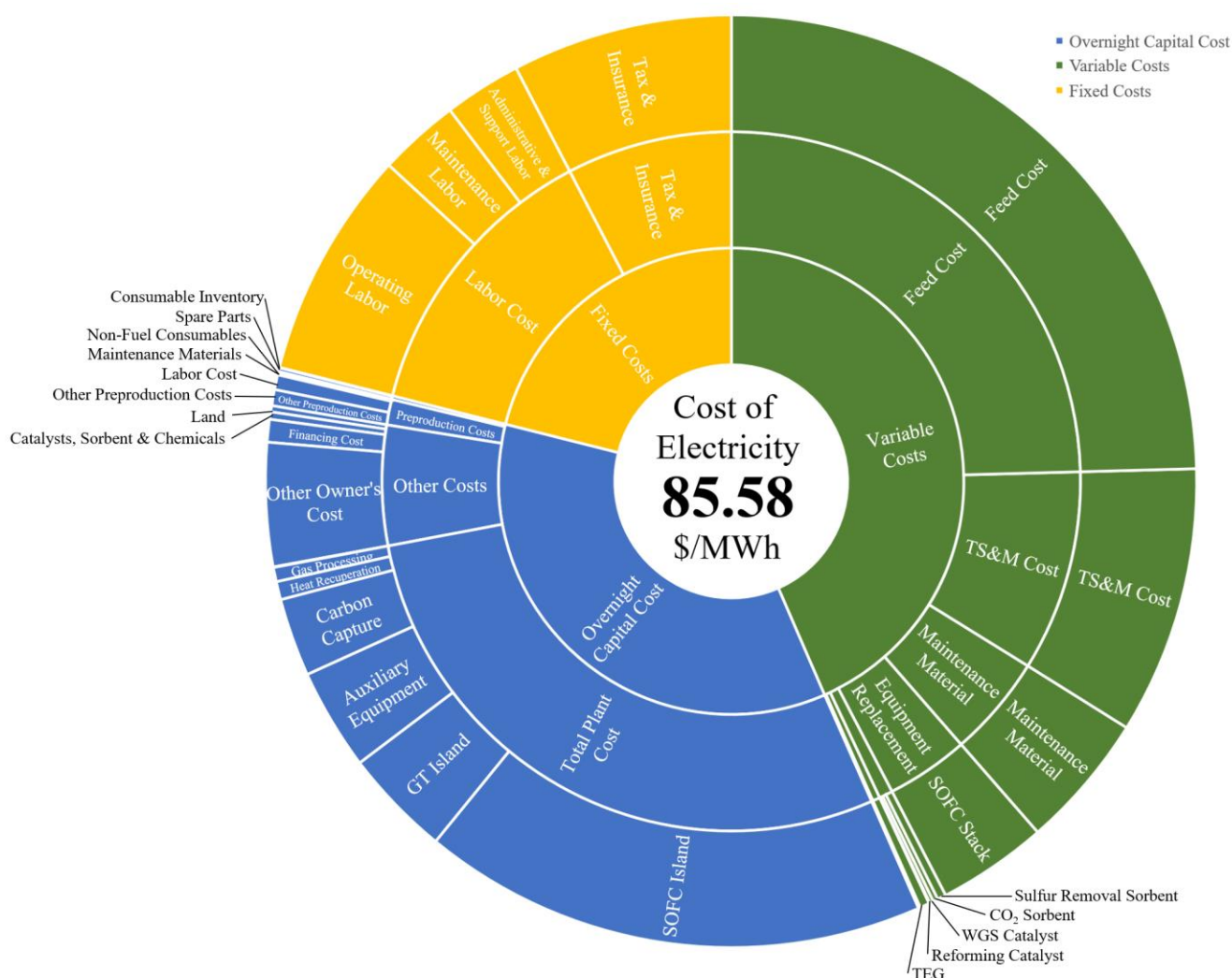


Figure 5-118: Breakdown of the cost-of-electricity of the 50 MW natural gas SOFC-GT hybrid with carbon capture and without water recovery.

LiBr Absorption System with Carbon Capture

The configuration of 50 MW NG hybrid plant with carbon capture and with LiBr absorption system follows the configuration of the 50 MW NG hybrid reference case with carbon capture. The LiBr absorption system is designed as single train unit and its design is identical to the 10 MW LiBr absorber case. A flue gas heat exchanger and a packed-bed column are installed downstream of the fuel pre-heater. The packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream heat exchanger is used to provide the heat for sorbent regeneration while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, set by an air-cooled condenser, supported by a vacuum compressor to remove any non-

condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is identical to the 10 MW NG LiBr absorber case and a similar plume rise can be expected. The flue gas from the LiBr absorber case differentiates from the air-cooled condenser and direct contact condenser cases as the flue gas in the LiBr condenser case is leaving the stack above its dew point temperature minimizing the risk of plume formation (flue gas dew point temperature around 14 °C). A general flow diagram of the 50 MW NG case with carbon capture and LiBr absorption water recovery is shown in **Figure 5-119**. The corresponding state-point steam data are shown in **Table 5-74**.

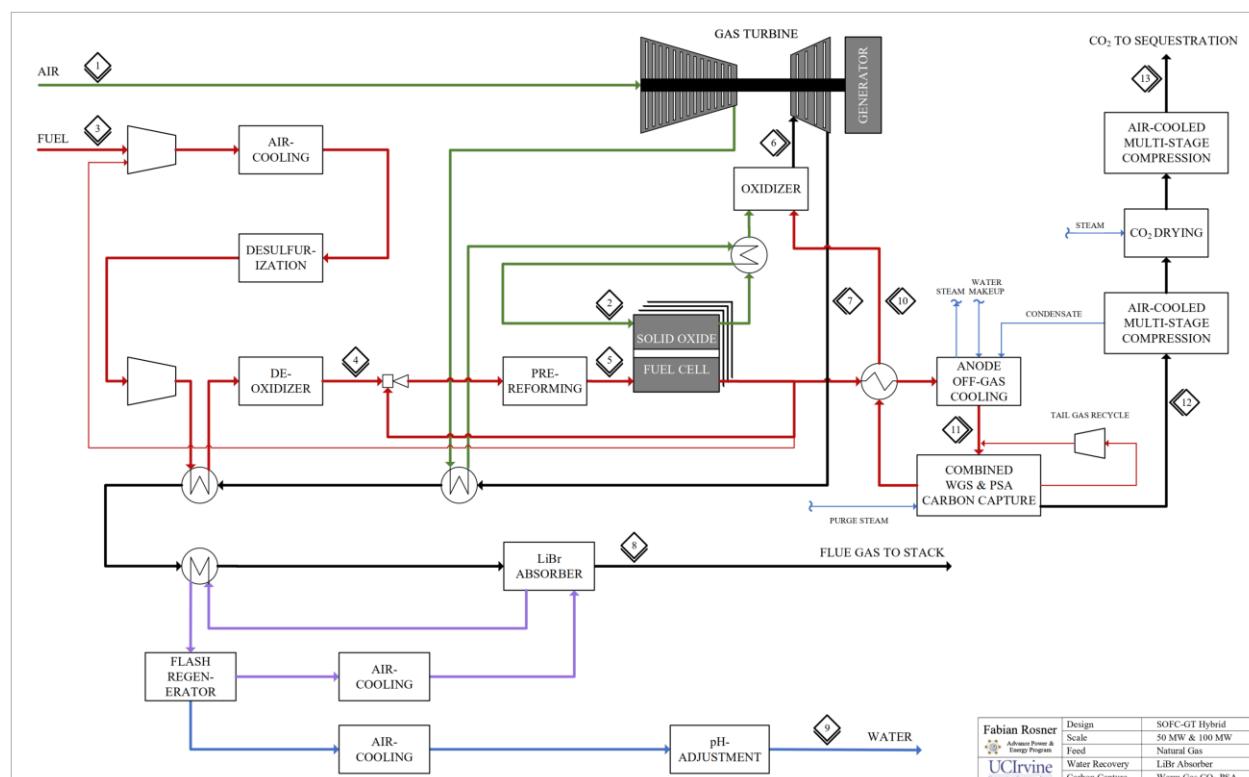


Figure 5-119: Flowsheet of the 50 MW natural gas SOFC-GT hybrid system with carbon capture and with LiBr absorption system for water recovery.

Table 5-74: Stream Summary of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	926	611	49
Pressure	bar	1.01	6.88	4.10	20.14	6.88	5.11	1.08	1.01
Molar Flow Rate	kmol/h	5,461.8	5,461.8	349.6	350.5	1,813.9	5,479.8	5,479.8	4,803.3
Mass Flow Rate	kg/h	157,612	157,612	6,059	6,080	36,848	148,710	148,710	136,522
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07727	0.07727	0.08816
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.77161	0.77161	0.88029
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00921	0.00921	0.01051
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22991	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09077	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01068	0.20335	0.00482	0.00482	0.00549
H ₂ O	-	0.00987	0.00987	0.00003	0.00163	0.32558	0.13710	0.13710	0.01555
CH ₄	-	0.00000	0.00000	0.93087	0.92863	0.14365	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	0.00001	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	1	1	1	1
Stream Number	Unit	9	10	11	12	13			
Temperature	°C	40	774	165	207	40			
Pressure	bar	2.00	5.30	6.20	1.20	151.65			
Molar Flow Rate	kmol/h	676.6	727.0	1,039.9	672.4	332.0			
Mass Flow Rate	kg/h	12,190	11,132	25,341	20,904	14,441			
Mass Vapor Fraction	-	0	1	1	1	1			
Composition	mole-basis								
O ₂	-	Trace	0.00000	0.00000	0.00000	0.00000			
N ₂	-	Trace	0.00751	0.00522	0.00019	0.00039			
Ar	-	Trace	0.00000	0.00000	0.00000	0.00000			
H ₂	-	0.00000	0.22278	0.09912	0.00581	0.01171			
CO	-	0.00000	0.00528	0.05942	0.00007	0.00015			
CO ₂	-	Trace	0.02824	0.28059	0.50638	0.98733			
H ₂ O	-	0.99999	0.73590	0.55545	0.48753	0.00041			
CH ₄	-	0.00000	0.00027	0.00019	Trace	0.00001			
C ₂ H ₆	-	0.00000	0.00001	Trace	Trace	Trace			
C ₃ H ₈	-	0.00000	0.00000	0.00000	0.00000	0.00000			
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace			
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace			
C ₄ H ₁₀ S	-	0.00000	Trace	Trace	Trace	Trace			
Total	-	0.99999	1	0.99999	0.99999	1			

Thermodynamic Performance. The net power output of the nominal 50 MW NG plant with carbon capture and LiBr absorber for water recovery is 53,741 kW. The fuel cell island produces 55,209 kW_{DC} and the GT generates an additional 3,815 kW. Due to the pressure drop associated with the LiBr absorber, the GT power output is lower than in the 50 MW carbon capture reference case. The GT surge margin in the 50 MW carbon capture LiBr absorber case is reduced to 17.8% and the SOFC operates at a pressure of 6.88 bar at a FU of 88.3%. The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 5,282 kW and represent about 8.0% of the auxiliary load.

The electrical efficiency of the LiBr absorption case decreased by 0.56% points to 67.66%-LHV compared to the 50 MW carbon capture reference case and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GT due to the increased backpressure.

The 50 MW carbon capture plant with LiBr absorber emits 1,683 kg/h of CO₂, which translates to 31 kg/MWh. The water recovery from the flue gas in the carbon capture case slightly increased to 90.1% (compared to the 50MW plant without carbon capture) due to the higher H₂O partial pressure in the flue gas. Sorbent-based water recovery systems are independent of the flue gas's dew point temperature and can remove substantial quantities of water. Simultaneously, they are less dependent upon ambient temperatures. While the ambient temperature still plays a part in the water recovery by setting the condenser backpressure (pressure in the regenerator), this effect can be moderated by the heat input required for sorbent regeneration. At the scale of this nominal 50 MW plant, a total of 292,553 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from this 50 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 151,300 m² of farmland (including 90% capacity factor). The high water quality of the recovered water minimizes water treatment and eliminates any wastewater streams from the plant, which reduces waste disposal costs but also maximizes water export. Furthermore, any internal water requirements, e.g., for the CO₂ PSA, can be met with the recovered water. The plant performance data are summarized in **Table 5-75**.

Table 5-75: Thermodynamic Performance of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	55,208.58
Gas Turbine	kW	3,814.98
Total Generation	kW	59,023.56
Auxiliary Load		
DC-AC Inverter Losses	kW	1,656.26
Fuel Gas Compression	kW	822.77
Fuel Gas Intercooler	kW	0.11
CO ₂ Capture and Compression	kW	2,119.45
Water Recovery System	kW	419.87
Water Treatment	kW	3.70
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	38.30
Transformer Losses	kW	221.95
Total Auxiliary Load	kW	5,282.41
Net Power Generation	kW	53,741.15
Plant Performance		
Net Efficiency	%-HHV	61.01
Net Efficiency	%-LHV	67.66
Consumables		
Fuel Gas	kg/h	6,059
Thermal Input	MJ _{LHV} /h	285,851
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	90
Carbon Emission	kg/h	1,683
Carbon Emission	kg/MWh	31
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	90.1
Flue Gas Water Recovery	kg/d	292,553

Economics. The TPC of the 50 MW hybrid with carbon capture and LiBr absorber for water recovery is \$145,373,120 resulting in a specific plant cost of 2,705 \$/kW. The SOFC island cost are very similar to the 50 MW carbon capture reference case and the slight variations result from the plant scale, number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the discussion sections of the 10 MW cases, which are still valid for this case. The GT island cost remains constant at \$18,255,070 since the same GT is used in all study cases. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. The LiBr absorption water recovery system costs \$4,664,460 a slight reduction versus the 50 MW LiBr absorber case without carbon capture due to the smaller exhaust gas volume flow (CO₂ is removed). The water treatment equipment cost is \$202,190. An illustration of the TPC distribution between the various plant categories is shown in **Figure 5-120**. A more detailed cost summary of all plant equipment is provided in **Table 5-76**.

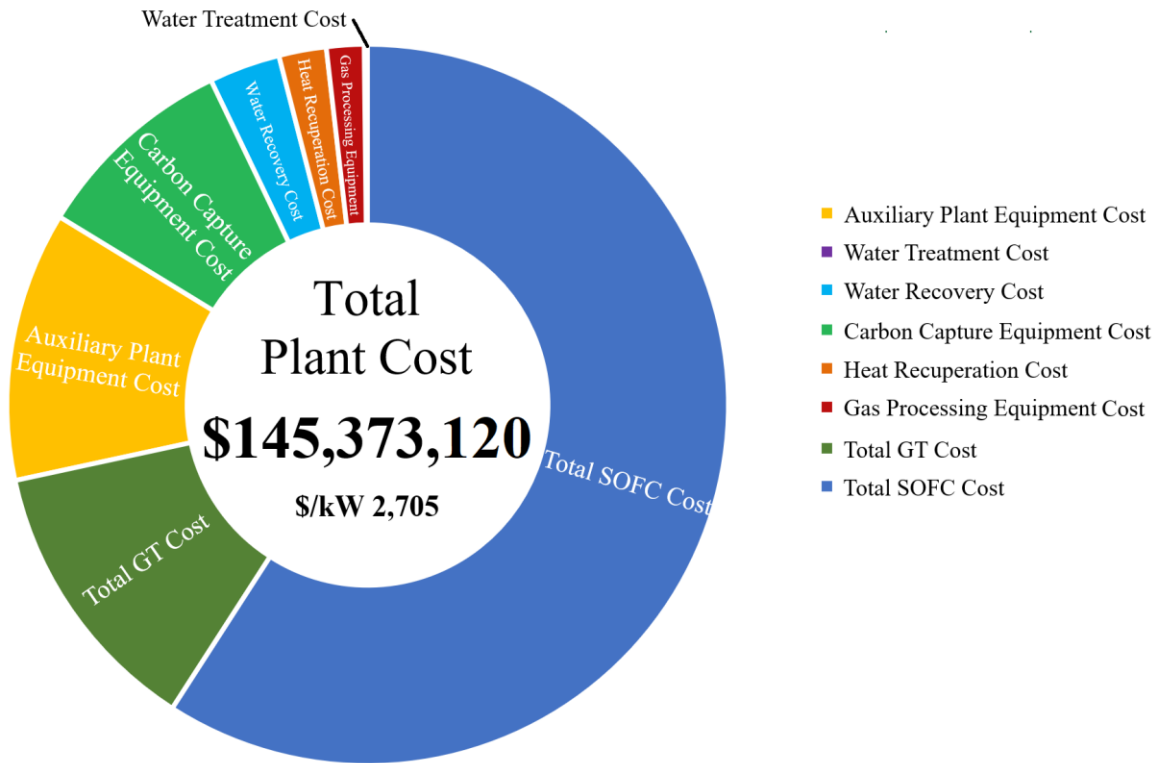


Figure 5-120: Total plant cost and specific plant cost of the 50 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

The overnight capital cost in the 50 MW LiBr absorber case with carbon capture is \$182,245,060, an increase of 5.0% compared to the 50 MW reference case with carbon capture. This increase is mostly related to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. Furthermore, the LiBr absorber system requires an initial fill and continuous LiBr makeup. The 55 wt-% LiBr solution is assumed to cost 2.04 \$₂₀₁₅/kg and the initial fill costs about \$1,914,220. The LiBr loss is very small and entirely due to entrainment since LiBr is a salt with extremely low vapor pressure. With a mist eliminator, the LiBr losses are expected to be around 50 kg per year. In the LiBr absorber case, water byproduct sales reduce the variable operating costs by \$1,494,920 per year.

Table 5-76: Equipment Cost of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	85,903,470
SOFC Stacks	\$	33,045,750
Power Conditioning	\$	32,357,950
Pressure Vessel, Housing & Assembly	\$	17,880,640
System Installation	\$	2,619,130
Gas Turbine Island	\$	18,255,070
Gas Turbine-Generator Set	\$	15,388,010
Gas Turbine Foundation	\$	223,690
Piping, Instrumentation & Electrical Equipment	\$	2,643,380
Gas Processing and Flue Gas Stack	\$	2,468,630
Fuel Compressors and Intercooler	\$	1,568,130
Fuel Processing	\$	670,470
Flue Gas Stack System	\$	230,020
Heat Recuperation	\$	3,078,310
Cathode Off-gas Heat Exchanger	\$	1,281,840
Air Preheater	\$	1,486,560
Fuel Preheater	\$	309,900
Carbon Capture	\$	13,221,830
Combined WGS & Carbon Capture PSA	\$	8,577,840
CO ₂ Compression and Dehydration	\$	4,643,990
Water Recovery	\$	4,866,650
Water Recovery System	\$	4,664,460
Water Treatment	\$	202,190
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	17,579,170
Accessory Electric Plant	\$	5,665,900
Instrumentation & Controls	\$	5,338,390
Improvement to Site	\$	1,387,980
Buildings & Structures	\$	5,186,890
Total Plant Cost	\$	145,373,120
Specific Plant Cost	\$/kW	2,705

The fixed operating cost in the LiBr absorber case increases due to tax and insurance costs of the additional equipment. The operating labor costs of the LiBr absorber unit are identical to 10 MW cases; however, when comparing the labor cost on a relative basis, the labor cost in the 50 MW LiBr absorber case is significantly lower due to the approximately five times higher water recovery. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the 50 MW reference case. Thus, the COE remains constant at 85.58 \$/MWh. The COW in the LiBr absorber case is 0.0156 \$/L, a reduction of over 12.8% versus the 50 MW LiBr case without carbon capture. A summary of the economic key parameters can be found in **Table 5-77**. A detailed breakdown of the COE is shown in **Figure 5-121**.

Figure 5-121 shows the COE value of 85.58 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-3.53 \$/MWh) and the whole circle represents the COE without byproduct credit (89.11 \$/MWh). This means if the water would be provided to the local community at no cost, the COE would increase to 89.11 \$/MWh due to the additional capital and operational costs. If the electricity is sold

Table 5-77: Economics of the 50 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	145,373,120
Preproduction Costs		
6 Months Labor Cost	\$	2,616,380
1 Month Maintenance Materials	\$	167,900
1 Month Non-Fuel Consumables	\$	162,370
1 Month Waste Disposal	\$	280
25% of 1 month fuel cost at 100% CF	\$	207,940
2% of TPC	\$	2,907,460
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	64,730
0.5% of TPC (Spare Parts)	\$	726,870
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	2,960,450
Land	\$	1,326,520
Other Owner's Costs (15% of TPC)	\$	21,805,970
Financing Costs	\$	3,925,070
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	8,983,200
Other Annual Variable Operating Cost at above CF for Initial Year	\$	3,569,920
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	3,393,030
Annual Byproduct Revenues at above CF for Initial Year	\$	1,494,920
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	5,232,770
Property Tax & Insurance	\$	2,907,460
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	77.57
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	85.58
First Year Levelized Cost of Water	\$/L	0.0156

to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0156 \$/L in order to reach the COE of 85.58 \$/MWh.

Figure 5-122 shows a detailed breakdown of the COW. The fixed operating cost in the 50 MW LiBr absorber case with carbon capture is substantially reduced compared to the 10 MW cases since the water recovery unit in both cases requires the same number of operators but in the 50 MW case the amount of recovered water is about 5 times higher reducing the relative operating labor cost to about 25.4% of the COW. The performance penalty cost, in particular the system integration losses, in the carbon capture scenario is significantly smaller than in the 50 MW case without carbon capture. This suggests that after the substantial pressure drop of the CO₂-PSA between the GT compressor and GT expander, which acts similarly to an increased GT backpressure, the system becomes less sensitive to the downstream pressure drop of the water recovery unit. System integration losses account for 6.6% and the parasitic load of the water recovery (mostly desiccant and water pumping) accounts for 19.0% of the COW. The TPC's share on the COW remains relatively constant at 25.4% despite is small capital cost reduction (compared to the 50 MW case without carbon capture). The water treatment is responsible for only 2.3% of the COW.

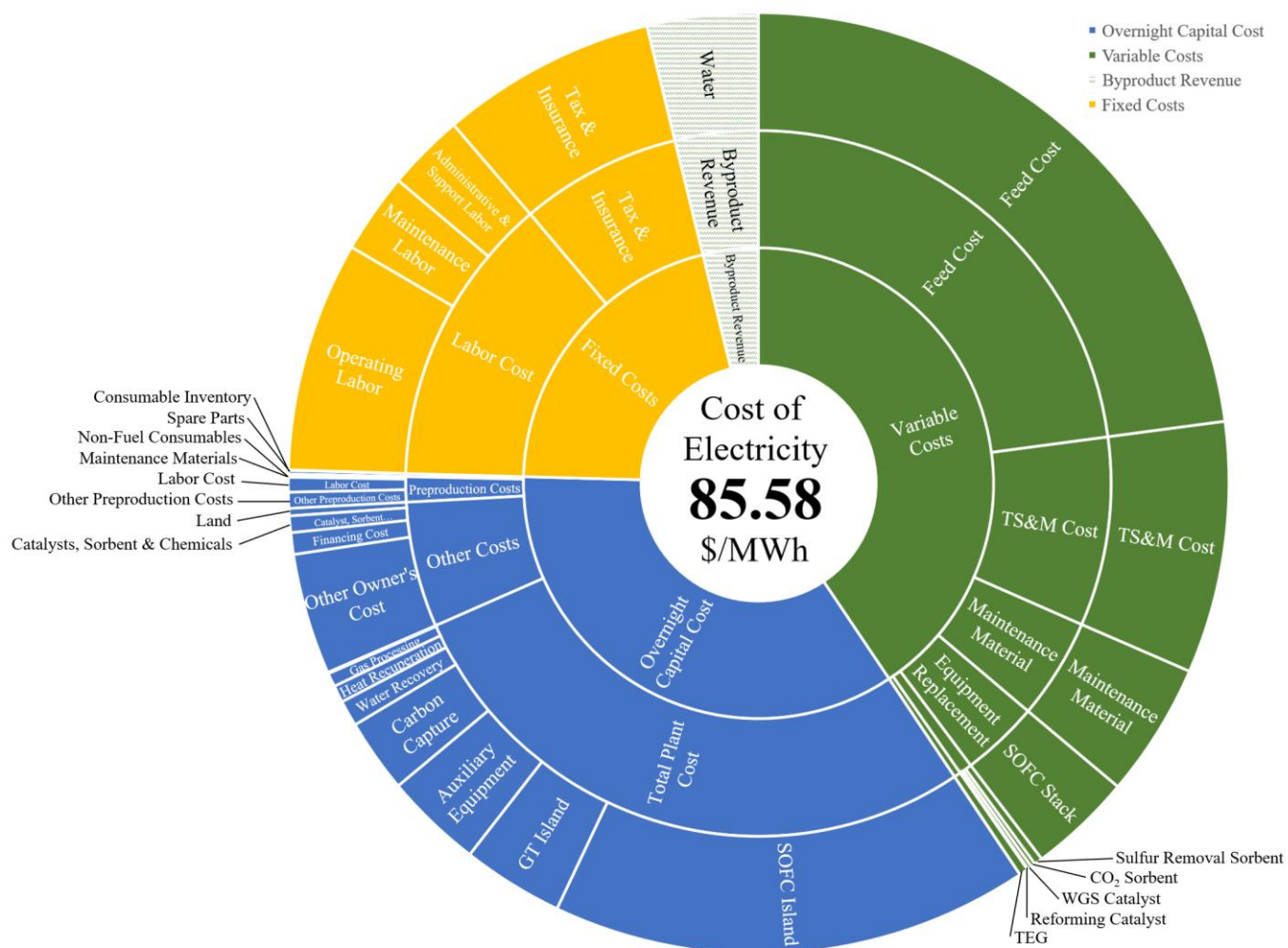


Figure 5-121: Breakdown of the cost-of-electricity of the 50 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

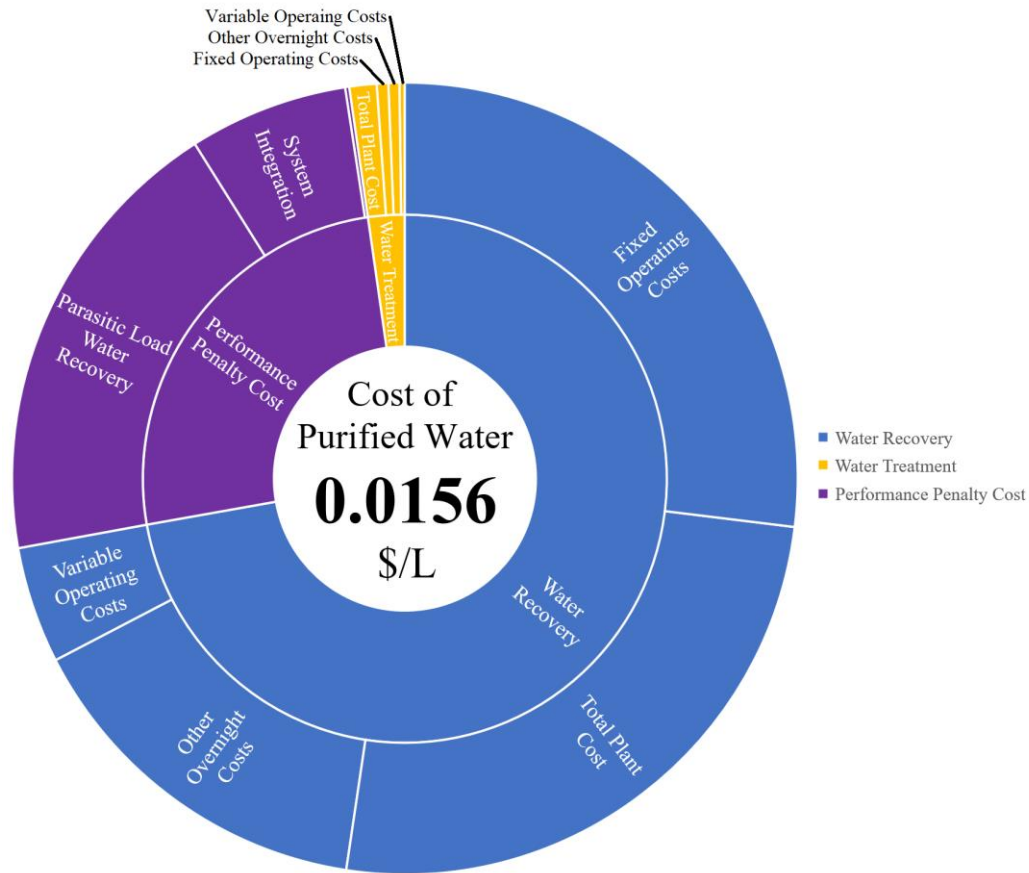


Figure 5-122: Breakdown of the cost-of-water of the 50 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorber for water recovery.

5.2.4 100 MW Natural Gas SOFC-GT Hybrids

Reference Case with Carbon Capture

The design of the 100 MW reference case with carbon capture is identical to the 50 MW SOFC-GT hybrid case with carbon capture. Again, the 100 MW reference case is used to set the baseline performance in order to evaluate the performance of the 100 MW SOFC-GT hybrid with LiBr water recovery. The system analysis is based upon the assumption that the same SOFC modules are used as in the 10 MW cases and 50 MW cases (to reach mass production). Furthermore, the 100 MW cases use the same thermal SOFC cell constrains as the previous SOFC-GT hybrid cases, which can be found in **Table 4-13**, together with the other SOFC design parameters. It is challenging to find GTs that match the requirements for SOFC hybridization. Larger engines tend to use axial compressors and exhibit larger pressure ratios; thus, multiple hybrid modules of 10 MW are used to reach the desired scale of 100 MW. Due to this modular approach, performance and economic improvements in the SOFC and GT islands are expected to be minimal; however, economics of scale are relevant to the upstream fuel processing and heat recuperation, as well as the carbon

capture equipment. The configuration of the 100 MW reference plant with carbon capture is provided in **Figure 5-123**. The corresponding state-point steam data are shown in **Table 5-78**.

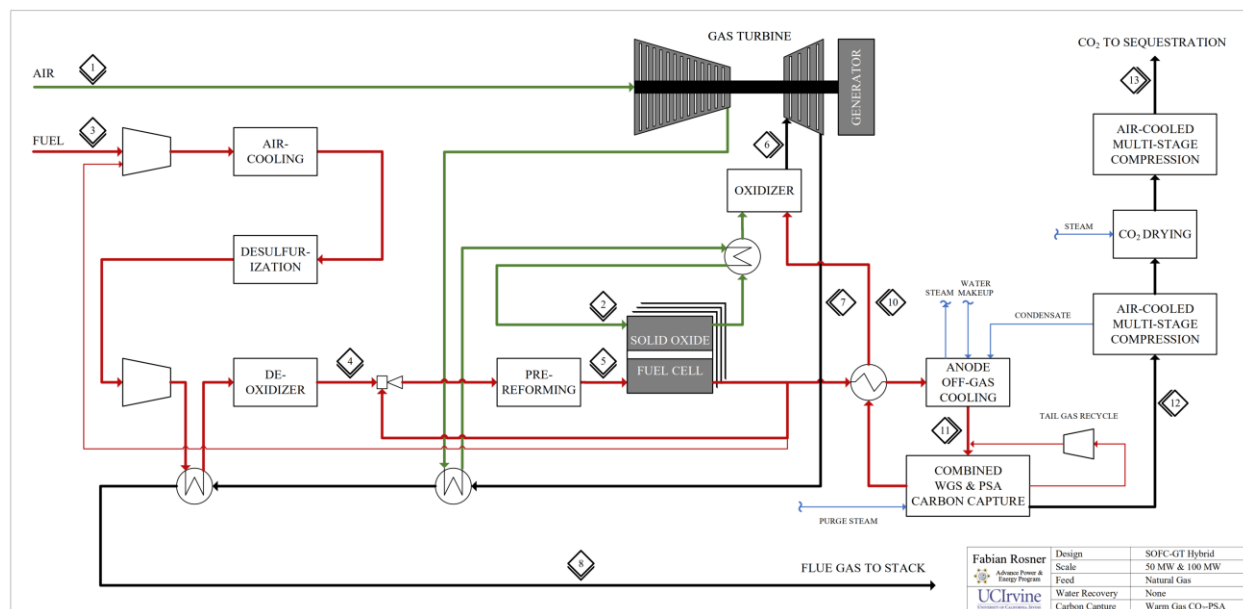


Figure 5-123: Flowsheet of the 100 MW natural gas SOFC-GT hybrid system with carbon capture and without water recovery.

Table 5-78: Stream Summary of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	926	604	322
Pressure	bar	1.01	6.84	13.80	19.82	6.84	5.07	1.05	1.01
Molar Flow Rate	kmol/h	10,950.0	10,950.0	701.0	702.6	3,667.2	10,987.7	10,987.7	10,987.7
Mass Flow Rate	kg/h	315,984	315,984	12,148	12,189	74,338	298,160	298,160	298,160
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07726	0.07726	0.07726
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00671	0.77150	0.77150	0.77150
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00921	0.00921	0.00921
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.23222	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09188	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01064	0.20261	0.00479	0.00479	0.00479
H ₂ O	-	0.00987	0.00987	0.00003	0.00156	0.32408	0.13725	0.13725	0.13725
CH ₄	-	0.00000	0.00000	0.93087	0.92873	0.14249	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	0.00001	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	1	1	1	1
Stream Number	Unit	9	10	11	12	13			
Temperature	°C	-	769	165	208	40			
Pressure	bar	-	5.27	6.16	1.19	151.65			
Molar Flow Rate	kmol/h	-	1,458.9	2,086.9	1,346.4	665.9			
Mass Flow Rate	kg/h	-	22,107	50,631	41,882	28,956			
Mass Vapor Fraction	-	-	1	1	1	1			
Composition	mole-basis								
O ₂	-	-	0.00000	0.00000	0.00000	0.00000			
N ₂	-	-	0.00751	0.00522	0.00019	0.00038			
Ar	-	-	0.00000	0.00000	0.00000	0.00000			
H ₂	-	-	0.23213	0.10378	0.00598	0.01202			
CO	-	-	0.00549	0.06151	0.00008	0.00016			
CO ₂	-	-	0.02779	0.27849	0.50723	0.98701			
H ₂ O	-	-	0.72674	0.55076	0.48651	0.00041			
CH ₄	-	-	0.00033	0.00023	Trace	0.00002			
C ₂ H ₆	-	-	0.00001	Trace	Trace	Trace			
C ₃ H ₈	-	-	0.00000	0.00000	0.00000	0.00000			
C ₄ H ₁₀	-	-	Trace	Trace	Trace	Trace			
H ₂ S	-	-	Trace	Trace	Trace	Trace			
C ₄ H ₁₀ S	-	-	Trace	Trace	Trace	Trace			
Total	-	-	1	0.99999	0.99999	1			

Thermodynamic Performance. The net power output of the nominal 100 MW natural gas reference case with carbon capture is 109,226 kW, whereby, the SOFC produces 110,059 kW of electricity (FU 87.8) and the ten GTs produce another 8,310 kW of electricity. The GT sets the operating pressure of the SOFC, 6.84 bar, and the resulting surging margin is 18.7%. The auxiliary load accounts for 9,144 kW, which includes the DC-AC inverter losses and transformer losses. The DC-AC inverter losses are associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 36.1% of the total auxiliary load. As seen in the 50 MW cases with carbon capture, the largest auxiliary load is CO₂ capture and compression with 4,169 kW (45.6%), a slight reduction over

the 50 MW hybrid with carbon capture due to efficiency gains seen in larger compressors. Fuel gas compression, which accounts for 12.6% of the total auxiliary load, benefits from similar effects as well as the higher natural gas supply pressure assumed for this scale of plant. As a result, the efficiency of the 50 MW reference case is minimally improved and reaches a value of 68.58%-LHV.

Although the plant is equipped with dry cooling technology, CO₂ removal is water intensive and requires a net water input of 250 kg/h. At the same time, the treatment of the imported water generates about 125 kg/h of wastewater, which has to be disposed. Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. The CO₂ emissions of the 100 MW natural gas hybrid with 90% carbon capture are 31 kg/MWh. The thermodynamic performance results are summarized in **Table 5-79**.

Table 5-79: Thermodynamic Performance of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	110,059.48
Gas Turbine	kW	8,310.24
Total Generation	kW	118,369.71
Auxiliary Load		
DC-AC Inverter Losses	kW	3301.79
Fuel Gas Compression	kW	1150.79
Fuel Gas Intercooler	kW	0.08
CO ₂ Capture and Compression	kW	4169.24
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	76.81
Transformer Losses	kW	445.11
Total Auxiliary Load	kW	9,143.82
Net Power Generation	kW	109,225.90
Plant Performance		
Net Efficiency	%-HHV	61.84
Net Efficiency	%-LHV	68.58
Consumables		
Fuel Gas	kg/h	12,148
Thermal Input	MJ _{LHV} /h	573,145
Water	kg/h	250
Emissions and Waste Streams		
Carbon Capture	%	90
Carbon Emission	kg/h	3,374
Carbon Emission	kg/MWh	31
Liquid Waste	kg/h	125
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Economics. The TPC of the 100 MW NG hybrid with carbon capture and without water recovery is \$268,352,970 and the specific plant cost is 2,457 \$/kW, a reduction of 120 \$/kW compared to the 50 MW NG case with carbon capture. This decrease in specific cost is mostly related to the economics of scale of

the carbon capture equipment, which is responsible for 7.5% of the TPC, as well as the fuel processing equipment (1.3%) and heat recuperation equipment (1.3%). The SOFC island accounts for \$171,405,210 and scales with the plant size (GT air supply). Other factors, such as SOFC operating pressure and spare stacks, play a small role in this cost increase. The GT island contributes \$36,510,140 to the plant's TPC, which simply scales by a factor of two compared to the 50 MW hybrids. Auxiliary plant equipment costs are estimated with \$33,354,170, which includes accessory electric plant, instrumentation and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-124**. A more detailed overview of the capital cost is provided in **Table 5-80**.

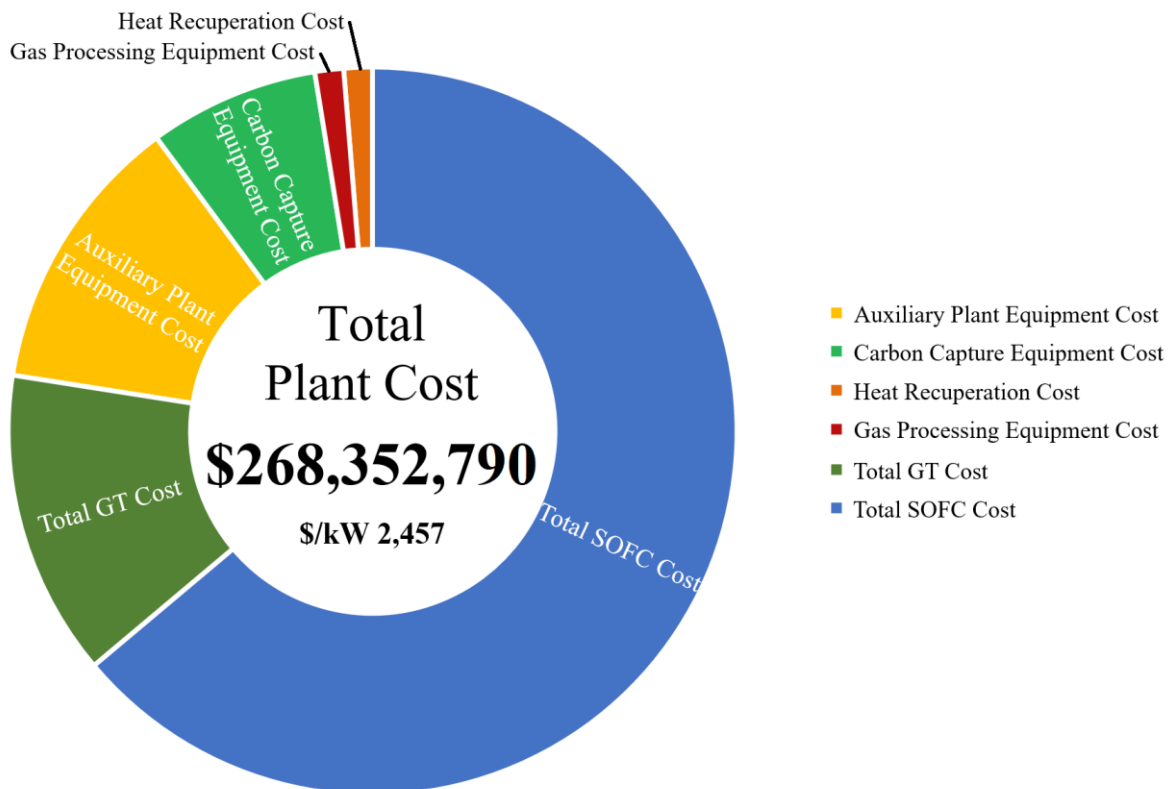


Figure 5-124: Total plant cost and specific plant cost of the 100 MW natural gas SOFC-GT hybrid with carbon capture and without water recovery.

Table 5-80: Equipment Cost of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	171,405,210
SOFC Stacks	\$	65,588,280
Power Conditioning	\$	64,551,300
Pressure Vessel, Housing & Assembly	\$	36,336,530
System Installation	\$	4,929,100
Gas Turbine Island	\$	36,510,140
Gas Turbine-Generator Set	\$	30,776,010
Gas Turbine Foundation	\$	447,380
Piping, Instrumentation & Electrical Equipment	\$	5,286,750
Gas Processing and Flue Gas Stack	\$	3,431,240
Fuel Compressors and Intercooler	\$	2,137,750
Fuel Processing	\$	944,330
Flue Gas Stack System	\$	349,160
Heat Recuperation	\$	3,404,180
Cathode Off-gas Heat Exchanger	\$	1,510,010
Air Preheater	\$	1,590,220
Fuel Preheater	\$	303,950
Carbon Capture	\$	20,247,850
Combined WGS & Carbon Capture PSA	\$	14,236,270
CO ₂ Compression and Dehydration	\$	6,011,580
Water Recovery	\$	N/A
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	33,354,170
Accessory Electric Plant	\$	11,362,770
Instrumentation & Controls	\$	9,854,460
Improvement to Site	\$	2,562,160
Buildings & Structures	\$	9,574,790
Total Plant Cost	\$	268,352,790
Specific Plant Cost	\$/kW	2,457

The total overnight capital cost is \$331,652,050 and is comprised of the TPC, the preproduction costs with \$9,813,450, the inventory capital costs with \$1,471,770 and other costs including financing and land costs with \$52,014,030. Variable annual costs are mostly fuel costs with \$18,011,770. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, and waste disposal costs, which amount to \$6,810,760 per year. Annual CO₂ transporting, storing, and monitoring account for \$6,805,250. Fixed costs are comprised of annual operating labor with \$6,797,660 and costs associated with tax and insurance with \$5,367,060 per year. The resulting COE for the 100 MW SOFC-GT hybrid plant with carbon capture and without water recovery is 80.12 \$/MWh, a reduction of 6.4% compared to the 50 MW reference case with carbon capture. The economic results are summarized in **Table 5-81**. A detailed breakdown of the COE is provided in **Figure 5-125**.

Table 5-81: Economics of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	268,352,790
Preproduction Costs		
6 Months Labor Cost	\$	3,398,830
1 Month Maintenance Materials	\$	301,650
1 Month Non-Fuel Consumables	\$	325,570
1 Month Waste Disposal	\$	3,410
25% of 1 month fuel cost at 100% CF	\$	416,940
2% of TPC	\$	5,367,060
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	130,010
0.5% of TPC (Spare Parts)	\$	1,341,760
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	2,100,260
Land	\$	2,415,330
Other Owner's Costs (15% of TPC)	\$	40,252,920
Financing Costs	\$	7,245,530
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	18,011,770
Other Annual Variable Operating Cost at above CF for Initial Year	\$	6,810,760
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	6,805,250
Annual Byproduct Revenues at above CF for Initial Year	\$	0
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	6,797,660
Property Tax & Insurance	\$	5,367,060
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	72.22
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	80.12
First Year Levelized Cost of Water	\$/L	N/A

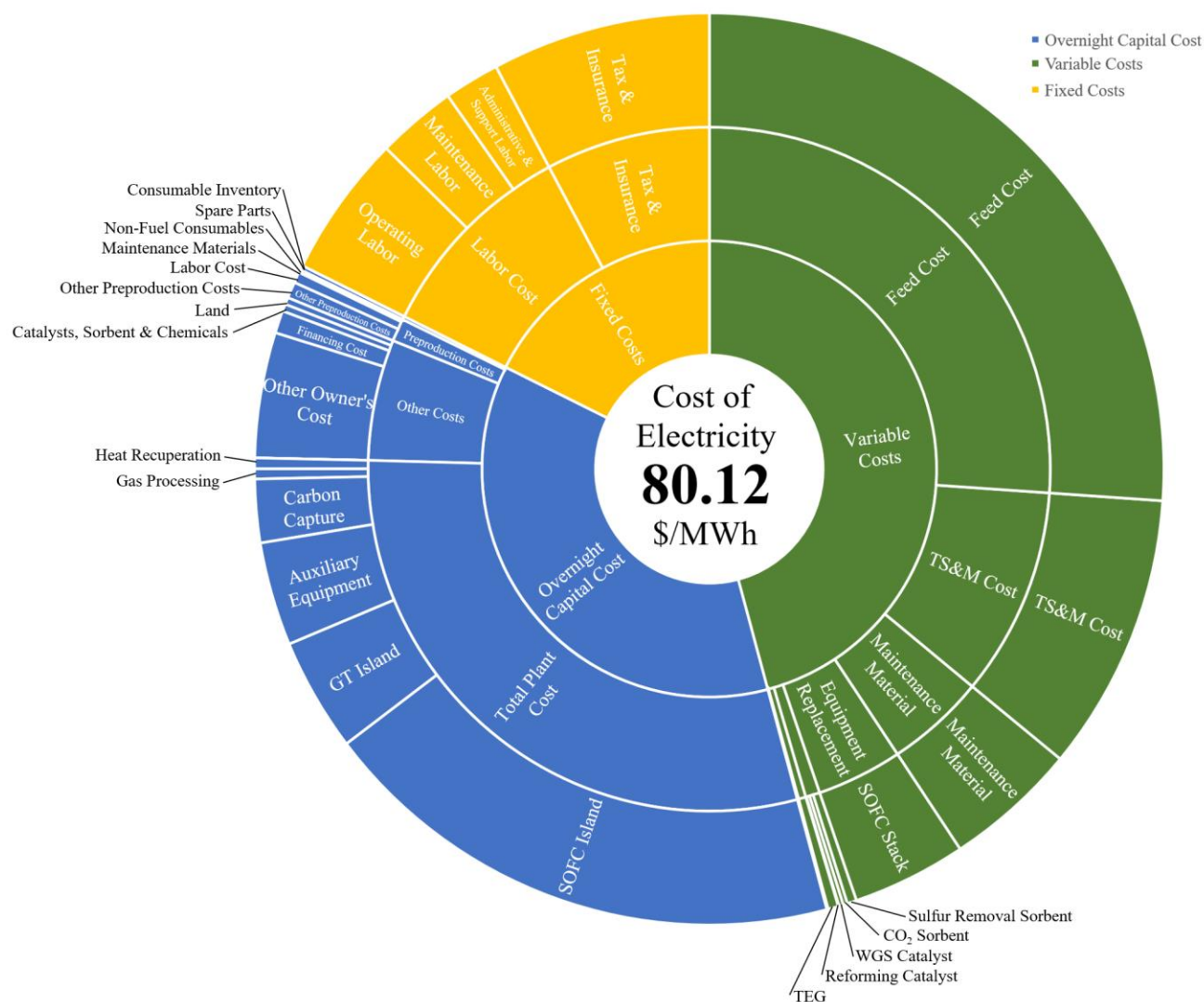


Figure 5-125: Breakdown of the cost-of-electricity of the 100 MW natural gas SOFC-GT hybrid with carbon capture and without water recovery.

LiBr Absorption System with Carbon Capture

The configuration of 100 MW NG hybrid plant with carbon capture and with LiBr absorption system follows the configuration of the 50 MW NG hybrid plant with carbon capture and with LiBr absorption system. The LiBr absorption system is designed as single train unit and its design is identical to the 10 MW LiBr absorber case and the 50 MW LiBr absorption system cases. A flue gas heat exchanger and a packed-bed column are installed downstream of the fuel pre-heater. The packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream

heat exchanger is used to provide the heat for sorbent regeneration while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, set by an air-cooled condenser, supported by a vacuum compressor to remove any non-condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is identical to the 10 MW NG LiBr absorber case and a similar plume rise can be expected. The flue gas from the LiBr absorber case differentiates from the air-cooled condenser and direct contact condenser cases as the flue gas in the LiBr condenser case is leaving the stack above its dew point temperature minimizing the risk of plume formation (flue gas dew point temperature around 14 °C). A general flow diagram of the 100 MW NG case with carbon capture and LiBr absorption water recovery is shown in **Figure 5-126**. The corresponding state-point steam data are shown in **Table 5-82**.

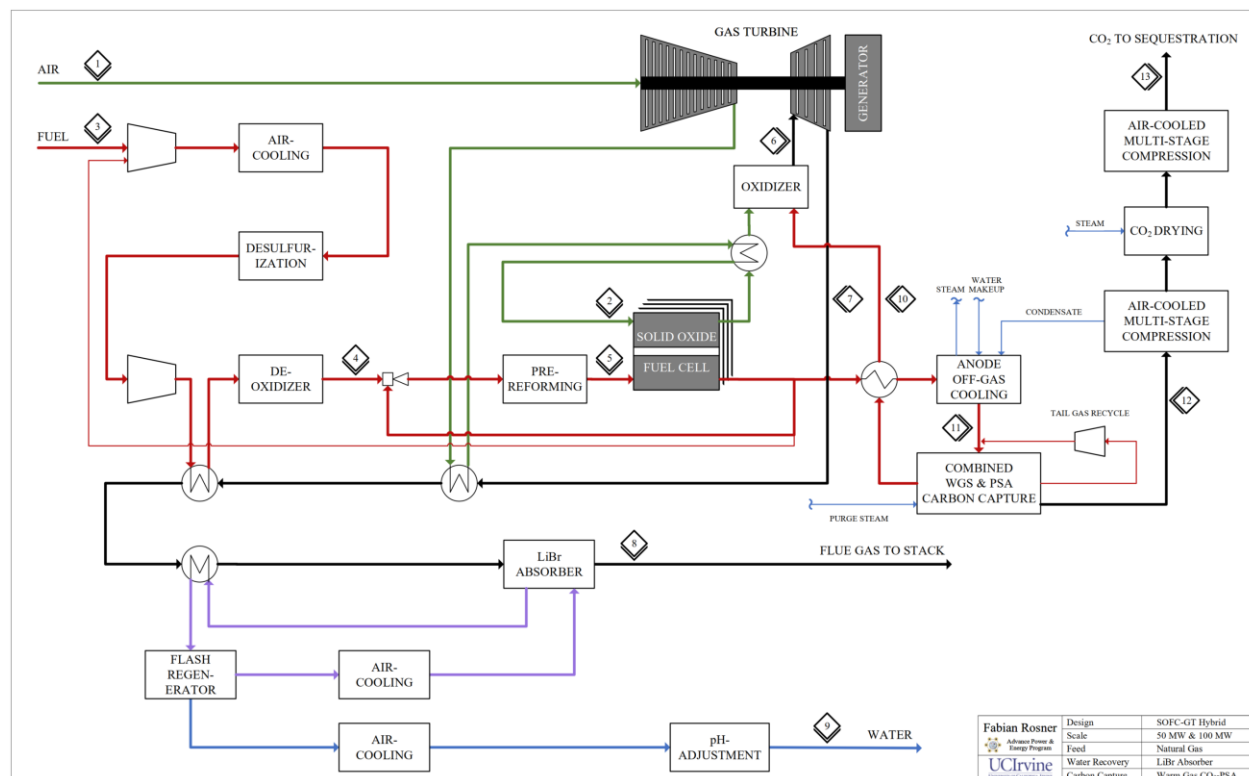


Figure 5-126: Flowsheet of the 100 MW natural gas SOFC-GT hybrid system with carbon capture and with LiBr absorption system for water recovery.

Table 5-82: Stream Summary of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8
Temperature	°C	15	700	15	322	700	926	611	49
Pressure	bar	1.01	6.88	13.80	20.14	6.88	5.11	1.08	1.01
Molar Flow Rate	kmol/h	10,923.7	10,923.7	699.3	701.0	3,627.7	10,959.7	10,959.7	9,606.6
Mass Flow Rate	kg/h	315,224	315,224	12,118	12,160	73,694	297,420	297,420	273,043
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1
Composition	mole-basis								
O ₂	-	0.20741	0.20741	0.00010	0.00000	0.00000	0.07727	0.07727	0.08816
N ₂	-	0.77315	0.77315	0.01600	0.01597	0.00672	0.77161	0.77161	0.88029
Ar	-	0.00924	0.00924	0.00000	0.00000	0.00000	0.00921	0.00921	0.01051
H ₂	-	0.00000	0.00000	0.00000	0.00005	0.22991	0.00000	0.00000	0.00000
CO	-	0.00000	0.00000	0.00000	0.00015	0.09077	0.00000	0.00000	0.00000
CO ₂	-	0.00033	0.00033	0.01000	0.01068	0.20335	0.00481	0.00481	0.00549
H ₂ O	-	0.00987	0.00987	0.00003	0.00163	0.32558	0.13709	0.13709	0.01555
CH ₄	-	0.00000	0.00000	0.93087	0.92863	0.14365	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	0.03200	0.03192	0.00001	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	0.00700	0.00698	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	0.00400	0.00399	Trace	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
C ₄ H ₁₀ S	-	0.00000	0.00000	Trace	Trace	Trace	0.00000	0.00000	0.00000
Total	-	1	1	1	1	1	1	1	1

Stream Number	Unit	9	10	11	12	13
Temperature	°C	40	774	165	207	40
Pressure	bar	2.00	5.30	6.20	1.20	151.65
Molar Flow Rate	kmol/h	1,353.3	1,453.9	2,079.7	1,344.8	664.0
Mass Flow Rate	kg/h	24,380	22,263	50,681	41,808	28,883
Mass Vapor Fraction	-	0	1	1	1	1
Composition	mole-basis					
O ₂	-	Trace	0.00000	0.00000	0.00000	0.00000
N ₂	-	Trace	0.00751	0.00522	0.00019	0.00039
Ar	-	Trace	0.00000	0.00000	0.00000	0.00000
H ₂	-	0.00000	0.22278	0.09912	0.00582	0.01171
CO	-	0.00000	0.00527	0.05942	0.00007	0.00015
CO ₂	-	Trace	0.02824	0.28059	0.50639	0.98733
H ₂ O	-	0.99999	0.73590	0.55545	0.48752	0.00041
CH ₄	-	0.00000	0.00027	0.00019	Trace	0.00001
C ₂ H ₆	-	0.00000	0.00001	Trace	Trace	Trace
C ₃ H ₈	-	0.00000	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace
C ₄ H ₁₀ S	-	0.00000	Trace	Trace	Trace	Trace
Total	-	0.99999	1	0.99999	0.99999	1

Thermodynamic Performance. The net power output of the nominal 100MW NG plant with carbon capture and LiBr absorber for water recovery is 108,060 kW. The SOFC island produces 110,418 kW_{DC} and the GTs generate an additional 7,628 kW. Due to the pressure drop associated with the LiBr absorber, the GT power output is lower than in the 100 MW carbon capture reference case. The GT surge margin in the 100 MW carbon capture LiBr absorber case is reduced to 17.8% and the SOFC operates at a pressure of 6.88 bar at a FU of 88.3%. The auxiliary load is largely the same as in the reference case; however, the water recovery and water treatment systems increase the total auxiliary power consumption to 9,986 kW and represent about 8.5% of the auxiliary load.

The electrical efficiency of the LiBr absorption case decreased by 0.56% points to 68.02%-LHV compared to the 100 MW carbon capture reference case and can be traced back to two dominating factors; the parasitic load associated with the water recovery and treatment, and the reduced power output of the GTs due to the increased backpressure.

The 100 MW carbon capture plant with LiBr absorber emits 3,366 kg/h of CO₂, which translates to 31 kg/MWh. The water recovery from the flue gas in this carbon capture case is 90.1%. Sorbent-based water recovery systems are independent of the flue gas's dew point temperature and can remove substantial quantities of water. Simultaneously, they are less dependent upon ambient temperatures. While the ambient temperature still plays a role in the water recovery, by setting the condenser backpressure (pressure in the regenerator), this effect can be moderated by the heat input required for sorbent regeneration. At the scale of this nominal 100 MW plant, a total of 585,111 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from this 100 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 302,600 m² of farmland (including 90% capacity factor). The high water quality of the

SOFC-GT Hybrid Systems with Water Recovery

recovered water minimizes water treatment and eliminates any wastewater streams from the plant, which reduces waste disposal costs but also maximizes water export. Furthermore, any internal water requirements, e.g., for the CO₂-PSA, can be met with the recovered water. The plant performance data are summarized in **Table 5-83**.

Table 5-83: Thermodynamic Performance of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	110,417.50
Gas Turbine	kW	7,628.42
Total Generation	kW	118,045.92
Auxiliary Load		
DC-AC Inverter Losses	kW	3,312.52
Fuel Gas Compression	kW	1,160.49
Fuel Gas Intercooler	kW	0.08
CO ₂ Capture and Compression	kW	4,145.43
Water Recovery System	kW	839.60
Water Treatment	kW	7.40
Zero Liquid Discharge System	kW	N/A
Miscellaneous Balance of Plant	kW	76.60
Transformer Losses	kW	443.88
Total Auxiliary Load	kW	9,986.01
Net Power Generation	kW	108,059.91
Plant Performance		
Net Efficiency	%-HHV	61.34
Net Efficiency	%-LHV	68.02
Consumables		
Fuel Gas	kg/h	12,118
Thermal Input	MJ _{LHV} /h	571,701
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	90
Carbon Emission	kg/h	3,366
Carbon Emission	kg/MWh	31
Liquid Waste	kg/h	0
Solid Waste	g/h	0
Water Recovery		
Flue Gas Water Recovery	%	90.1
Flue Gas Water Recovery	kg/d	585,111

Economics. The TPC of the 100 MW hybrid with carbon capture and LiBr absorber for water recovery is \$277,799,130 resulting in a specific plant cost of 2,571 \$/kW. The SOFC island costs are very similar to the 100 MW carbon capture reference case and the slight variations result from the plant scale, number of installed spare stacks and pressure effects. Detailed explanations of those effects can be found in the discussion sections of the 10 MW cases, which are still valid for this case. The GT island cost remains constant at \$36,510,140 since the same GT is used in all study cases. Other equipment costs such as gas processing and flue gas stack scale with the plant size but remain relatively constant. The LiBr absorption water recovery system costs \$8,299,920. The water treatment equipment cost is \$217,670. An illustration

of the TPC distribution between the various plant categories is shown in **Figure 5-127**. A more detailed cost summary of all plant equipment is provided in **Table 5-84**.

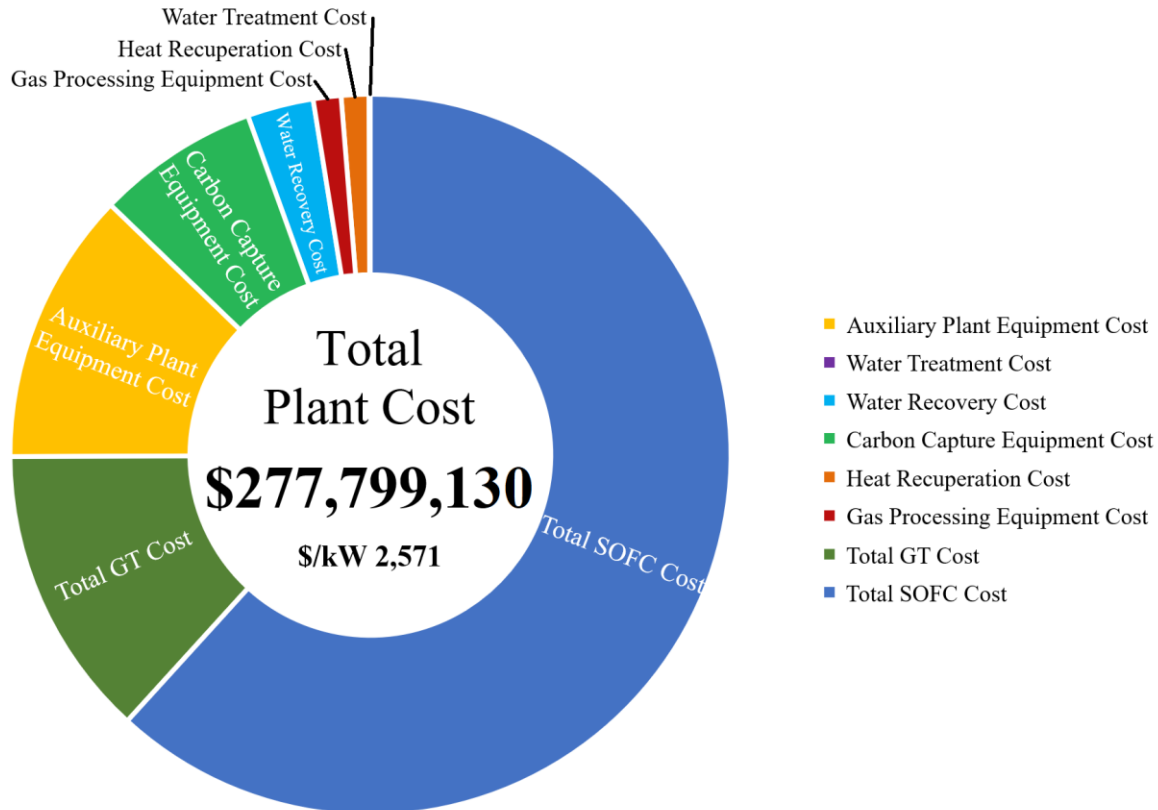


Figure 5-127: Total plant cost and specific plant cost of the 100 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

The overnight capital cost in the 100 MW LiBr absorber case with carbon capture is \$347,169,420, an increase of 4.7% compared to the 100 MW reference case with carbon capture. This increase is mostly related to the TPC of the water recovery system. Another considerable factor is the labor cost, which increases due to additional labor associated with the operation of the water recovery unit. Annual feed costs remain virtually constant. Small changes compared to the reference case are observed due to the slightly lower efficiency and the plant scale. Other variable operating costs increase due to the water recovery and water treatment equipment. Furthermore, the LiBr absorber system requires an initial fill and continuous LiBr makeup. The 55 wt-% LiBr solution is assumed to cost 2.04 \$₂₀₁₅/kg and the initial fill costs about \$3,828,450. The LiBr loss is very small and entirely due to entrainment since LiBr is a salt with extremely low vapor pressure. With a mist eliminator, the LiBr losses are expected to be around 100 kg per year. In the LiBr absorber case, water byproduct sales reduce the variable operating costs by \$2,526,990 per year.

Table 5-84: Equipment Cost of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	171,673,360
SOFC Stacks	\$	65,605,540
Power Conditioning	\$	64,728,530
Pressure Vessel, Housing & Assembly	\$	36,404,710
System Installation	\$	4,934,590
Gas Turbine Island	\$	36,510,140
Gas Turbine-Generator Set	\$	30,776,010
Gas Turbine Foundation	\$	447,380
Piping, Instrumentation & Electrical Equipment	\$	5,286,750
Gas Processing and Flue Gas Stack	\$	3,437,970
Fuel Compressors and Intercooler	\$	2,146,230
Fuel Processing	\$	943,090
Flue Gas Stack System	\$	348,640
Heat Recuperation	\$	3,379,330
Cathode Off-gas Heat Exchanger	\$	1,467,100
Air Preheater	\$	1,609,910
Fuel Preheater	\$	302,320
Carbon Capture	\$	20,183,530
Combined WGS & Carbon Capture PSA	\$	14,216,130
CO ₂ Compression and Dehydration	\$	5,967,410
Water Recovery	\$	8,517,590
Water Recovery System	\$	8,299,920
Water Treatment	\$	217,670
Zero Liquid Discharge System	\$	N/A
Auxiliary Plant Equipment Cost	\$	34,097,210
Accessory Electric Plant	\$	11,331,680
Instrumentation & Controls	\$	10,201,340
Improvement to Site	\$	2,652,350
Buildings & Structures	\$	9,911,830
Total Plant Cost	\$	277,799,130
Specific Plant Cost	\$/kW	2,571

The fixed operating cost in the LiBr absorber case increases due to tax and insurance costs of the additional equipment. The operating labor costs of the LiBr absorber unit are identical to 10 MW cases and the other 50 MW cases; however, when comparing the labor cost on a relative basis, the labor cost in the 100 MW LiBr absorber case is significantly lower due to the approximately ten times higher water recovery than in the 10 MW cases. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the 100 MW reference case. Thus, the COE remains constant at 80.12 \$/MWh. The COW in the LiBr absorber case is 0.0132 \$/L, a reduction of 15.4% versus the 50 MW LiBr case with carbon capture. A summary of the economic key parameters can be found in **Table 5-85**. A detailed breakdown of the COE is shown in **Figure 5-128**.

Figure 5-128 shows the COE value of 80.12 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-2.97 \$/MWh) and the whole circle represents the COE without byproduct credit (83.09 \$/MWh). This means, if the water would be provided to the local community at no cost, the COE

would increase to 83.09 \$/MWh due to the additional capital and operational costs. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0132 \$/L in order to reach the COE of 80.12 \$/MWh.

Table 5-85: Economics of the 100 MW Natural Gas SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
Plant Capacity Factor	%	90
Total Overnight Capital Cost		
Total Plant Cost	\$	277,799,130
Preproduction Costs		
6 Months Labor Cost	\$	3,580,010
1 Month Maintenance Materials	\$	316,520
1 Month Non-Fuel Consumables	\$	324,660
1 Month Waste Disposal	\$	550
25% of 1 month fuel cost at 100% CF	\$	415,890
2% of TPC	\$	5,555,980
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	129,470
0.5% of TPC (Spare Parts)	\$	1,389,000
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	5,920,470
Land	\$	2,567,300
Other Owner's Costs (15% of TPC)	\$	41,669,870
Financing Costs	\$	7,500,580
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	17,966,400
Other Annual Variable Operating Cost at above CF for Initial Year	\$	6,930,650
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	6,786,160
Annual Byproduct Revenues at above CF for Initial Year	\$	2,526,990
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	7,160,020
Property Tax & Insurance	\$	5,555,980
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	72.15
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	80.12
First Year Levelized Cost of Water	\$/L	0.0132

Figure 5-129 shows a detailed breakdown of the COW. The fixed operating cost in the 100 MW LiBr absorber case with carbon capture is substantially reduced compared to the 50 MW carbon capture case since the water recovery unit in both cases requires the same number of operators but in the 100 MW case the amount of recovered water is about two times higher reducing the relative operating labor cost to about 21.2% of the COW. System integration losses account for 7.1% and the parasitic load of the water recovery (mostly desiccant and water pumping) accounts for 20.8% of the COW. The TPC's share on the COW is 27.8% despite the water-specific (\$/L) capital cost reduction (compared to the 50 MW case with carbon capture). The water treatment is responsible for only 1.5% of the COW.

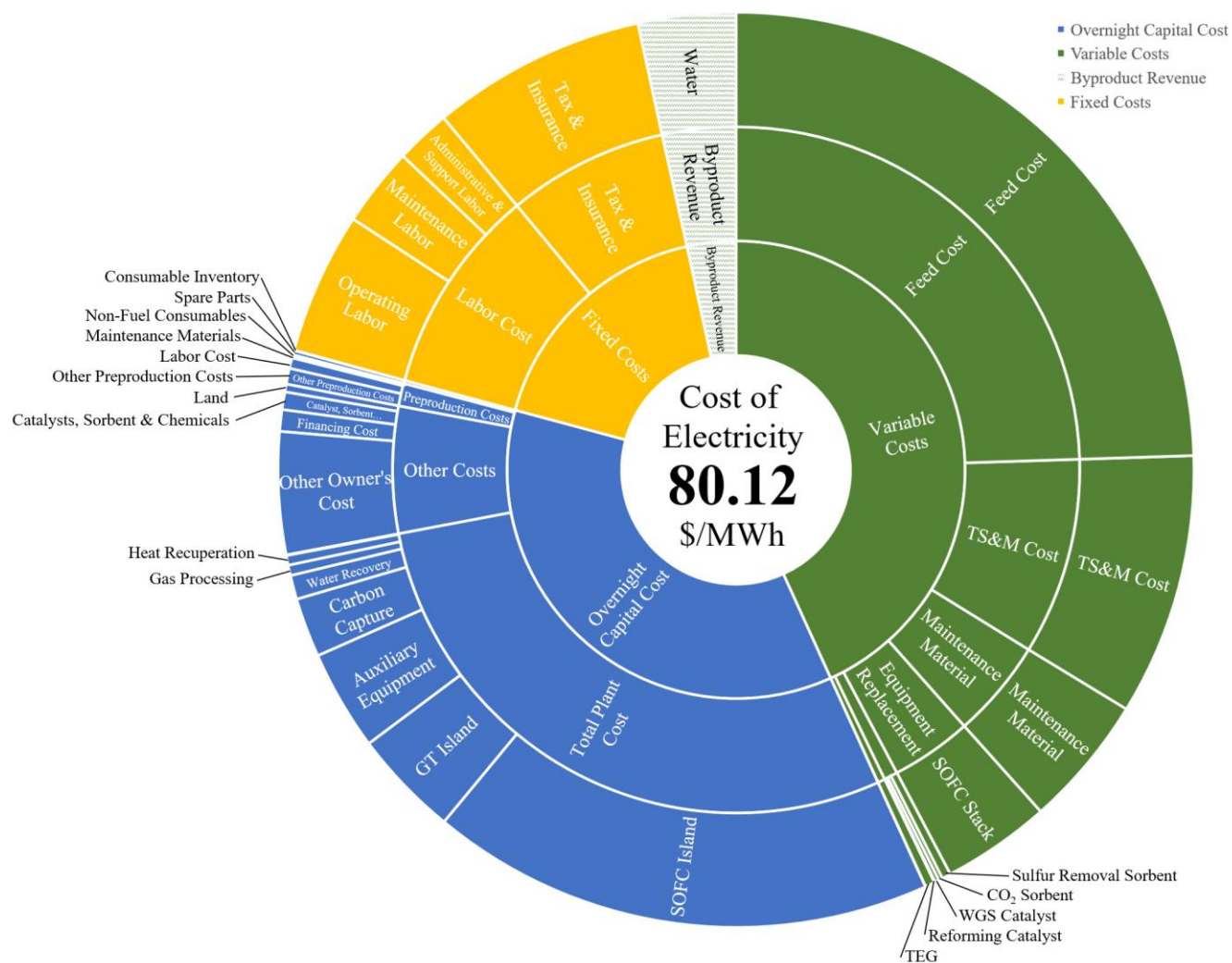


Figure 5-128: Breakdown of the cost-of-electricity of the 100 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

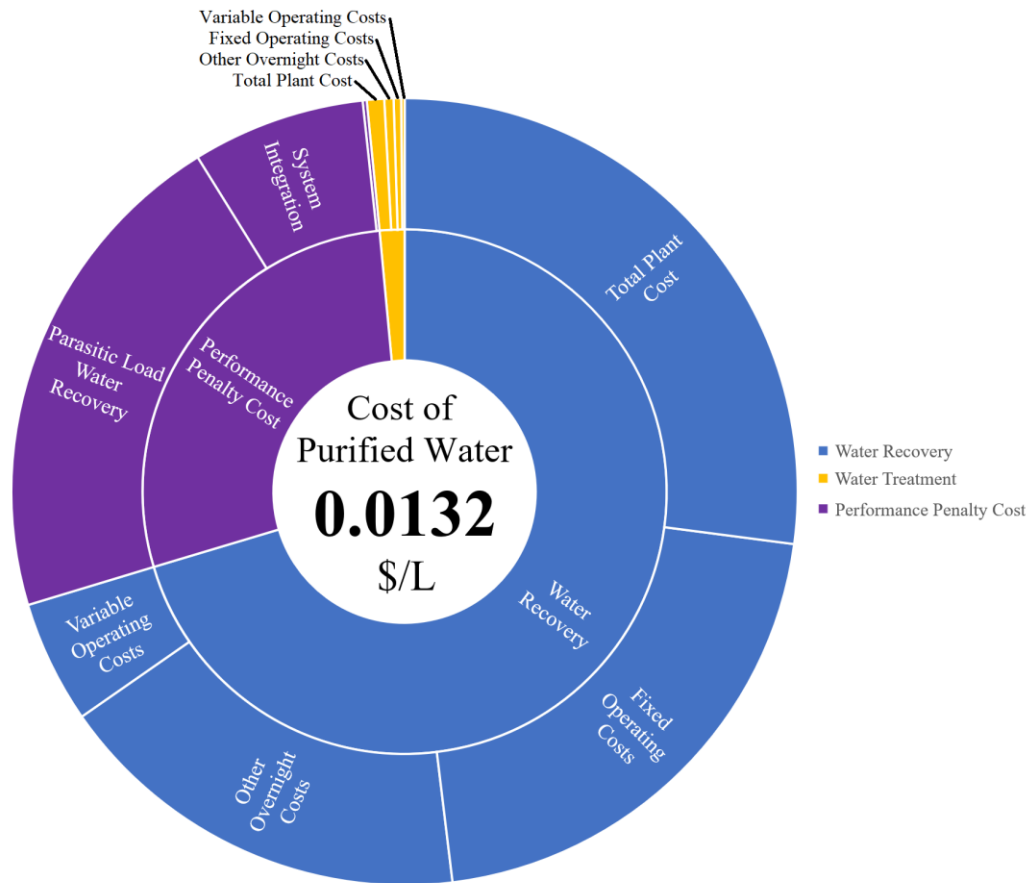


Figure 5-129: Breakdown of the cost-of-water of the 100 MW natural gas SOFC-GT hybrid with carbon capture and with LiBr absorber for water recovery.

Summary of the 50 MW and 100 MW Natural Gas SOFC-GT Hybrids

In total eight different natural gas SOFC-GT hybrid plants at various scales with and without carbon capture, and with and without LiBr absorption system for water recover have been studied. Among the studied scenarios the cases without carbon capture reach the highest efficiencies due to the higher GT power output related to CO₂ mass flow through the GT expander and the lower auxiliary power consumption as no CO₂ needs to be compressed for sequestration. Changes in SOFC operating pressure and TIT have only minor effects on the efficiency. The hybrid plants without carbon capture reach a higher efficiency at larger scale (10 MW versus 50 MW). Due to the modularity of the SOFC and the SOFC-GT hybrid unit, these efficiency gains are small and mostly related to improved component efficiencies, such as the fuel compressor, attainable at larger scales. The efficiency losses due to water recovery (LiBr absorption system) are 0.97% points in the cases without carbon capture and a consequence of the additional parasitic load of the water recovery and treatment units as well as the increased GT backpressure. The studied carbon capture scenarios

exhibit similar behaviors. The larger scale plants with 100 MW show a higher efficiency than the smaller 50 MW plant. When looking closely, it is noticed that the efficiency increases more in the carbon capture scenario when moving from 50 MW to 100 MW than in the scenarios without carbon capture (10 MW to 50 MW), which is related to the improved compressor efficiency of the CO₂ compression train. Also, in the carbon capture scenarios, water recovery reduces the efficiency and results in an efficiency reduction of 0.56% points. A comparison of the electrical efficiencies is provided in **Figure 5-130**.

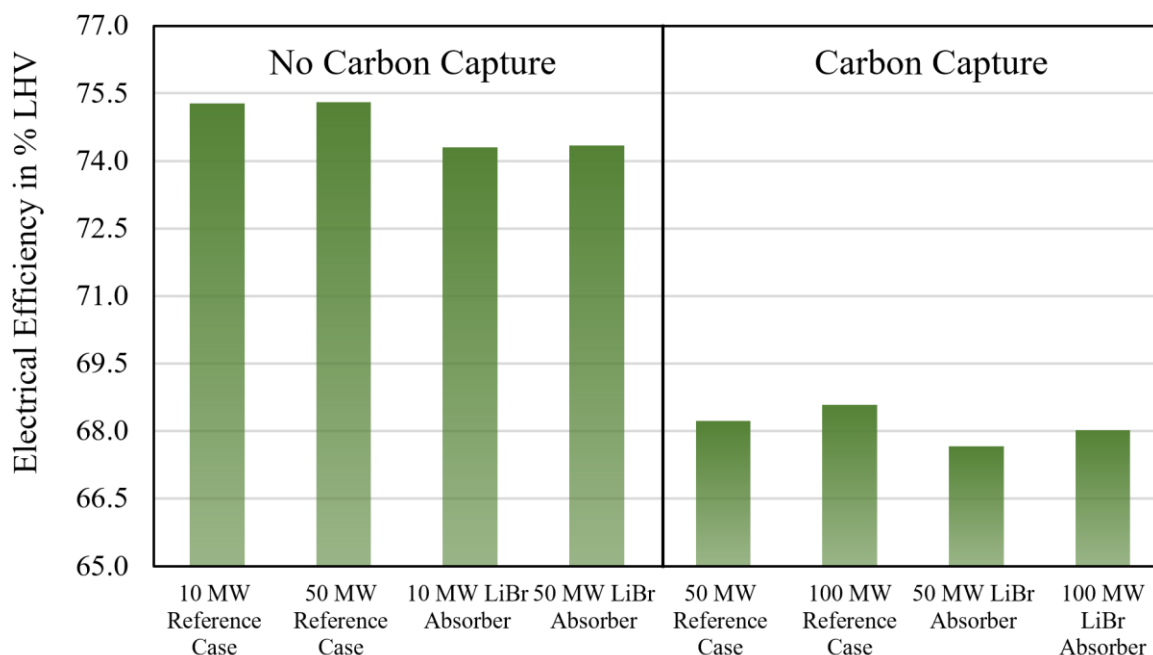


Figure 5-130: Comparison of the electrical efficiencies of the 10-100 MW natural gas SOFC-GT hybrids with and without carbon capture, and with and without LiBr water recovery.

The COE is established in the respective reference cases without water recovery. When moving to larger scales the COE is reduced as a result of the reduced capital cost of the balance-of-plant equipment and reduced operating costs. The increased efficiency has only a moderate impact on the COE. Furthermore, carbon capture increases the COE as additional plant equipment is needed to remove the CO₂ from the anode off-gas and reduces the efficiency. The respective COEs are held constant in the corresponding water recovery scenarios and the water sales price is adjusted in order to obtain the same COE as in the reference case. Three main factors contribute to the COW: the capital cost, the operating cost and the amount of water recovered. While the amount of water recovered remains relatively stable on a relative basis (L/MW), the water recovery benefits from the economies of scale as capital cost and operating labor cost are reduced on a \$/L basis. The lowest COWs are obtained in the carbon capture scenarios due to the larger scale but also

because of the lower flue gas volume flow and higher water vapor concentration in the flue gas after carbon capture. A summary of the COEs and COWs for the scenarios is presented in **Figure 5-131**.

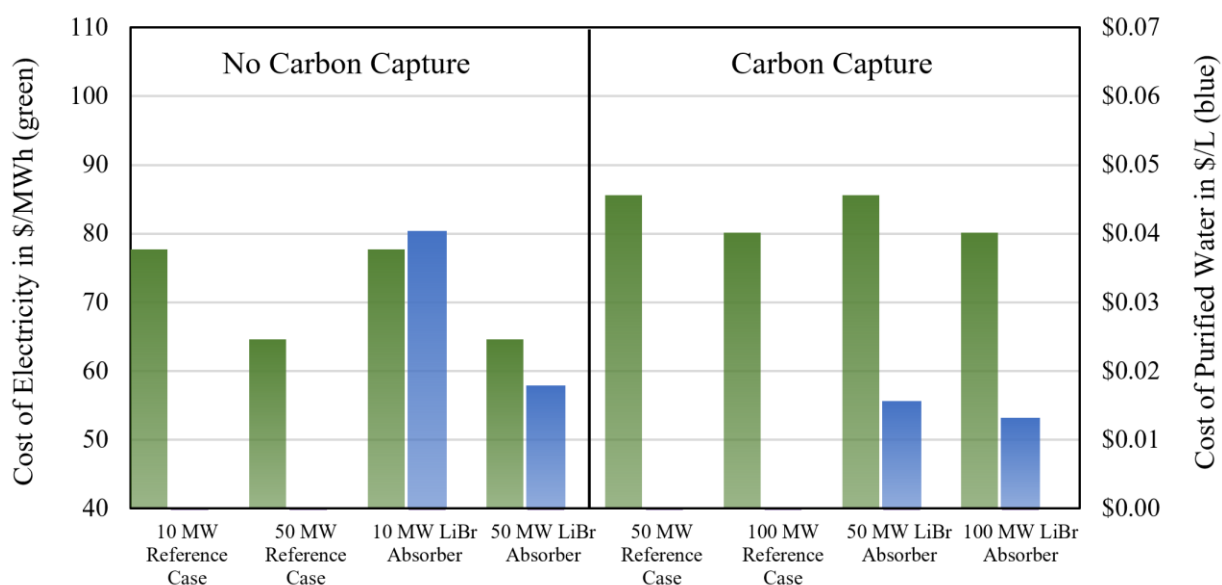


Figure 5-131: Comparison of the cost-of-electricity and cost-of-water of the 10-100 MW natural gas SOFC-GT hybrids with and without carbon capture, and with and without LiBr water recovery.

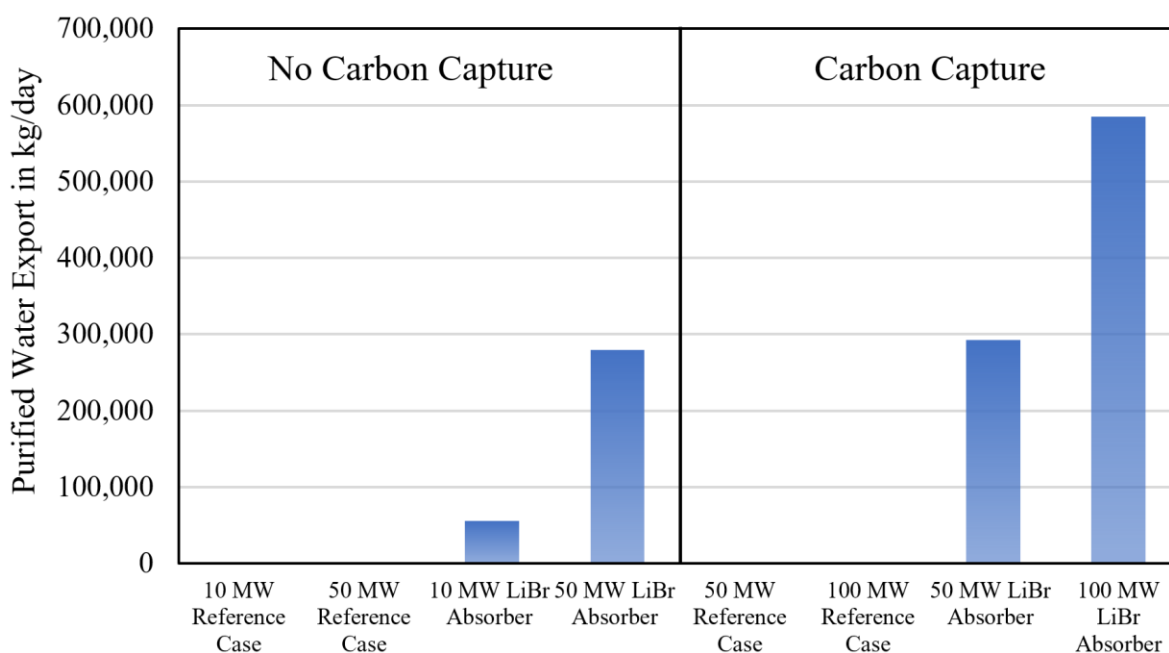


Figure 5-132: Comparison of the water recovery rate of the 10-100 MW natural gas SOFC-GT hybrids with and without carbon capture, and with and without LiBr water recovery.

The water recovery essentially scales with the plant size and the largest amount of water is recovered in the 100 MW plant, which is able to recover over 585,100 liters of water per day. Nevertheless, the carbon capture scenarios are able to achieve slightly higher water recovery rates due to the higher water partial pressure in the flue gas. The results of the amount of water recovery for each plant are plotted in **Figure 5-132**.

5.2.5 100 MW Gasification SOFC-GT Hybrids

Reference Case – Coal with Carbon Capture

In order to enable the use of solid fuels, such as coal, in combination with SOFCs, the solid feedstock needs to be converted into a gaseous fuel rich in H_2 , CO and/or CH_4 . With SOFCs, it is desirable to increase the amount of CH_4 in the syngas in order to take advantage of internal reforming inside the SOFC. The formation of CH_4 is favored under hydrogasification conditions and at lower temperatures, which also improves the gasifier's cold gas efficiency. To further increase the concentration of the syngas (SOFC operates on chemical potential difference), a dual-bed gasifier is used to separate the pyrolysis and gasification reactions (hydrogasification and steam gasification) from the exothermic fuel oxidation reactions with air. This way air can be used directly in the char combustor without the need of an air separation unit. The released heat from the char combustion is transported to the gasification vessel via sand to drive the overall endothermic gasification process. Both vessels of the dual-bed hydrogasifier operate at elevated pressures and requires coal pressurization via lock-hoppers, subsequent to drying and milling. Similarly, air for the char converter needs to be supplied at elevated pressures. In order to utilize the high pressure, high temperature gas leaving the char converter in a downstream expander, the flue gas needs to be free of particulates and corrosive alkali metals. The raw syngas exiting the gasifier is cleaned in multiple steps. First the syngas is cooled, and particulates are removed in candle filters before it is scrubbed in a venturi scrubber using water and NaOH to remove any fine particulates as well as water soluble compounds such as HCl. Downstream of the scrubber, volatile trace metals are removed in an expandable process and a HCl guard bed removes any remaining HCl before the syngas enters the bulk sulfur removal process. Since SOFCs have a very low tolerance for sulfur, a sulfur guard bed is installed downstream of the bulk sulfur removal unit. The recovered sulfur from the bulk sulfur removal unit is mixed with the flue gas from the char converter in the SNOXTM unit where it is ultimately reacted to sulfuric acid, a saleable byproduct. The clean syngas is now preheated and ready to enter the SOFC module. No pre-reforming or recycle are needed in this design as the syngas is rich in moisture and relatively low FU factors are desired to recycle enough hydrogen back to the gasifier. The air supply for the SOFC is separated from the gasifier air supply and the same Dresser-Rand GT as in the previous scenarios is used for the SOFC hybridization. In total nine GTs are needed to reach the nominal 100 MW plant scale. A small portion of

the anode off-gas is sent to the GT oxidizer, while the majority of the syngas is decarbonized and recycled back to the gasifier after adjusting the moisture content. Downstream of the SOFC-GT hybrid, heat is recovered to raise more steam. The plant is able to satisfy its own water demand since substantial amounts of water are recovered during CO₂ compression and the compression of the syngas recycle. Any excess water is used to raise steam and injected into the GT of the SOFC-GT hybrid unit. Water unsuitable for steam generation is used for ash treatment and/or injected into the char converted so that no liquid waste streams are generated by this plant. The configuration of the 100 MW coal gasification reference plant without flue gas water recovery is provided in **Figure 5-133** (a larger version of this illustration can be found in Appendix B). The corresponding state-point steam data are shown in **Table 5-86**.

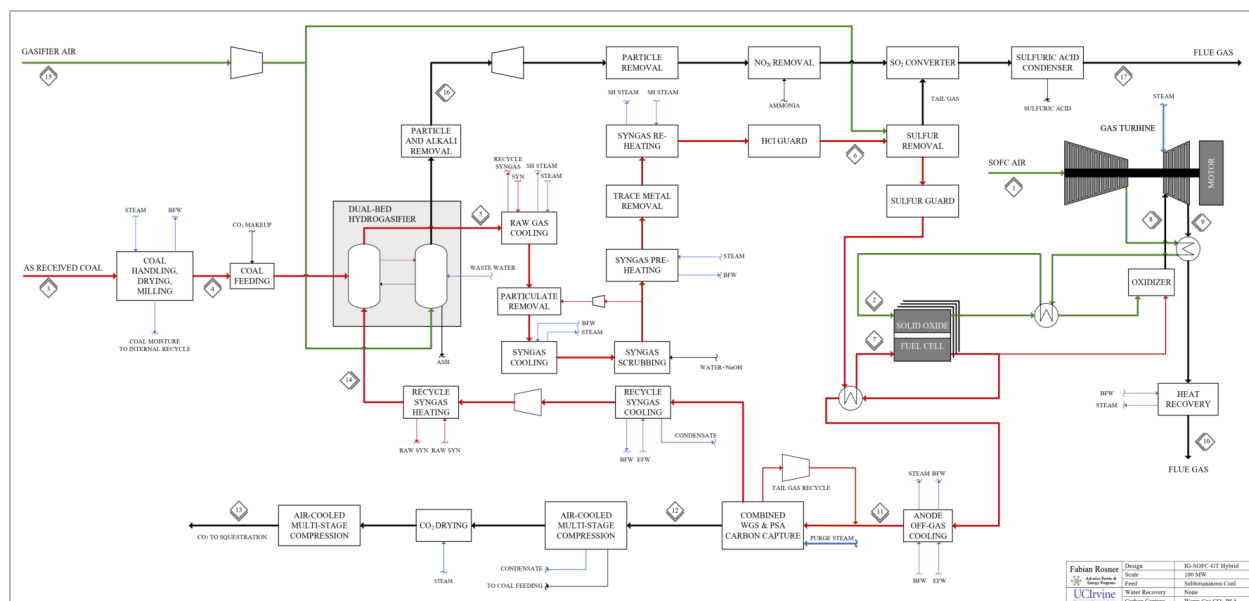


Figure 5-133: Flowsheet of the 100 MW coal gasification SOFC-GT hybrid system with carbon capture and without water recovery.

Table 5-86: Stream Summary of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	71	751	315	700	831	543
Pressure	bar	1.01	6.60	1.01	1.01	15.00	10.05	6.60	5.82	1.05
Molar Flow Rate	kmol/h	10,011.4	10,011.4	-	-	7,370.9	7,715.6	7,715.6	9,469.5	9,469.5
Mass Flow Rate	kg/h	288,897	288,897	41,864	37,897	84,559	90,733	90,605	259,309	259,309
Mass Vapor Fraction	-	1	1	0	0	1	1	1	1	1
Composition	mole-basis									
O ₂	-	0.20741	0.20741	-	-	0.00000	0.00000	0.00000	0.05105	0.05105
N ₂	-	0.77315	0.77315	-	-	0.02375	0.02269	0.02269	0.81838	0.81838
AR	-	0.00924	0.00924	-	-	0.00000	0.00000	0.00000	0.00977	0.00977
H ₂	-	0.00000	0.00000	-	-	0.50348	0.48099	0.48099	0.00000	0.00000
CO	-	0.00000	0.00000	-	-	0.05794	0.05533	0.05533	0.00000	0.00000
CO ₂	-	0.00033	0.00033	-	-	0.03638	0.03459	0.03460	0.00991	0.00991
H ₂ O	-	0.00987	0.00987	-	-	0.24101	0.27539	0.27642	0.11088	0.11088
CH ₄	-	0.00000	0.00000	-	-	0.13590	0.12958	0.12958	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	-	-	0.00109	0.00103	Trace	0.00000	0.00000
SO ₂	-	0.00000	0.00000	-	-	Trace	Trace	Trace	Trace	Trace
HCl	-	0.00000	0.00000	-	-	0.00001	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	-	-	0.99957	0.99959	0.99960	0.99999	0.99999
Stream Number	Unit	10	11	12	13	14	15	16	17	-
Temperature	°C	224	165	206	40	600	15	946	90	-
Pressure	bar	1.01	5.95	0.90	151.65	17.30	1.01	14.70	1.02	-
Molar Flow Rate	kmol/h	9,469.5	8,835.8	4,394.1	1,331.9	6,511.3	1,947.0	1,914.5	1,979.9	-
Mass Flow Rate	kg/h	259,309	129,768	113,198	57,393	54,058	56,185	58,691	60,586	-
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	-
Composition	mole-basis									
O ₂	-	0.05105	0.00000	0.00000	0.00000	0.00000	0.20741	0.02500	0.02534	-
N ₂	-	0.81838	0.01878	0.00025	0.00081	0.02532	0.77315	0.74930	0.76048	-
AR	-	0.00977	0.00000	0.00000	0.00000	0.00000	0.00924	0.00896	0.00909	-
H ₂	-	0.00000	0.39503	0.00625	0.02052	0.63788	0.00000	0.00000	0.00000	-
CO	-	0.00000	0.08386	0.00008	0.00025	0.00766	0.00000	0.00000	0.00000	-
CO ₂	-	0.00991	0.07717	0.30173	0.97701	0.00836	0.00033	0.17551	0.16973	-
H ₂ O	-	0.11088	0.40449	0.69143	0.00051	0.29295	0.00987	0.04012	0.03500	-
CH ₄	-	0.00000	0.02040	0.00027	0.00088	0.02747	0.00000	0.00000	0.00000	-
C ₂ H ₆	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
C ₃ H ₈	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
SO ₂	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	-
HCl	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	-
Total	-	0.99999	0.99973	1	0.99999	0.99964	1	0.99888	0.99963	-

Thermodynamic Performance. The net power output of the nominal 100 MW coal gasification reference case with carbon capture is 109,046 kW, whereby the SOFC produces 128,710 kW of electricity at a FU of 35.8%. The nine Dresser Rand GTs provide another 2,822 kW of electricity, which is substantially lower than in the 100 MW NG reference case. The reason for this decrease in GT power output is the reduced fuel (syngas) flow to the GT oxidizer, since it is more attractive to recycle the syngas back to the gasifier and have it undergo oxidation in the SOFC. This ultimately leads to a lower TIT and the steam injection into the GT further lowers the TIT; however, increases the mass flow through the GT expander. In the gasification scenario, gasifier flue gas expansion provides additional power. By mounting the air supply

compressor and gasifier flue gas expander onto the same shaft, the expander can provide the power needed for compression and an additional 2,653 kW. Overall, this leads to gross power generation of 134,184 kW for this plant.

The auxiliary load accounts for 25,138 kW. The largest auxiliary load with 11,075 kW is related to the syngas recycle compressor, which even exceeds the power consumed for CO₂ compression with 9,061 kW. The DC-AC inverter losses are only the third largest auxiliary load and associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 10.2% of the total auxiliary load. The efficiency of the 100 MW coal gasification reference case is 47.16%-HHV. The SOFC stack efficiency is slightly increased in the syngas scenario due to the low fuel utilization (83.5%-LHV), the GT efficiency on the other hand dropped to 15.0%-LHV. In the syngas scenario, the GT is operating at a different operating point leading to reduced component efficiencies but also the lower TIT reduces the GT efficiency. The surge margin in this case is 29.0% and increased substantially compared to the NG cases.

The plant is equipped with dry cooling technology and as previously mentioned, does not produce any liquid waste streams. Solid waste streams are associated with the gasifier operation and contain ash as well as unconverted feedstock (disposal of catalysts, etc. are handled separately). Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. With the dual-bed gasifier design a carbon capture rate of 74.75% is achieved. The CO₂ emissions of the 100 MW coal gasification SOFC-GT hybrid with 74.75% carbon capture are 181 kg/MWh. The thermodynamic performance results are summarized in **Table 5-87**.

Table 5-87: Thermodynamic Performance of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	128,709.50
Gas Turbine (SOFC-GT Hybrid)	kW	2,821.59
Gas Turbine (Gasifier)	kW	2,653.00
Total Generation	kW	134,184.08
Auxiliary Load		
DC-AC Inverter Losses	kW	2,574.19
Feedstock Handling, Milling and Drying	kW	989.81
Gasifier Feeding System	kW	516.94
Gasifier Ash Handling	kW	90.87
Syngas Cleaning and Scrubbing	kW	137.91
CO ₂ Capture and Compression	kW	9,060.82
Syngas Recycle Compressors	kW	11,075.21
Boiler and Feed Water System	kW	100.91
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Miscellaneous Balance of Plant	kW	87.07
Transformer Losses	kW	504.57
Total Auxiliary Load	kW	25,138.30
Net Power Generation	kW	109,045.78
Plant Performance		
Net Efficiency	%-HHV	47.16
Net Efficiency	%-LHV	48.94
Consumables		
Feedstock	kg/h	41,864
Thermal Input	MJ _{HHV} /h	833,931
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	74.75
Carbon Emission	kg/h	19,700
Carbon Emission	kg/MWh	181
Liquid Waste	kg/h	0
Solid Waste	kg/h	3,448
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Economics. The TPC of the 100 MW coal gasification hybrid with carbon capture and without water recovery is \$664,805,030 and the specific plant cost is 6,097 \$/kW. The gasification and feed preparation section of the plant contributes the largest cost to the TPC with \$247,006,130. The SOFC island accounts for \$177,641,950, whereby, the SOFC stacks do not represent the largest cost anymore. At low fuel utilization factors the number of stacks can be reduced (on a relative basis) and the inverter cost becomes the largest cost item in the SOFC island. The GT island costs are constant throughout the various cases as the same Dresser Rand GT is used for all scenarios. The GT cost of \$32,859,130 accounts for the nine GTs used in this case. Syngas cleanup and flue gas cleanup have similar TPCs and both together have a TPC of \$85,179,410. The carbon capture equipment in this gasification plant has a relatively small share on the plant's TPC with 4.8% and the steam and water system accounts for \$22,996,580. Auxiliary plant equipment costs are estimated with \$67,361,350, which includes accessory electric plant, instrumentation

and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-134**. A more detailed overview of the capital cost is provided in **Table 5-88**.

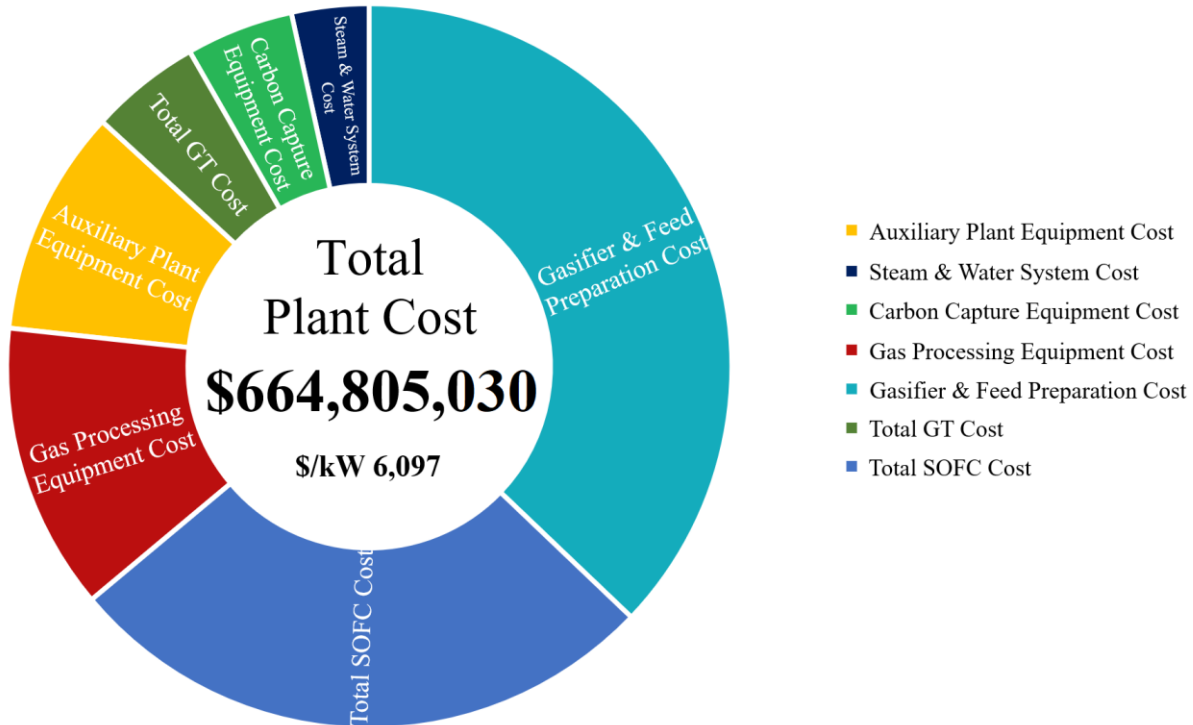


Figure 5-134: Total plant cost and specific plant cost of the 100 MW coal gasification SOFC-GT hybrid with carbon capture and without water recovery.

The total overnight capital cost is \$819,039,890 and is comprised of the TPC, the preproduction costs with \$23,006,720, the inventory capital costs with \$5,089,050 and other costs including financing and land costs with \$126,139,100. Variable annual costs are fuel costs with \$7,248,330 and other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, as well as waste disposal costs, which amount to \$16,259,460 per year. Annual CO₂ transporting, storing, and monitoring account for \$12,477,860. The sulfuric acid sales revenue generates an annual income of \$1,972,700, which is equivalent to a COE reduction of 2.43 \$/MWh. Fixed costs are comprised of annual labor costs with \$15,877,790 and costs associated with tax and insurance with \$13,296,100 per year. The resulting COE for the 100 MW coal gasification SOFC-GT hybrid plant with carbon capture and without water recovery is 158.12 \$/MWh with TS&M (142.75 \$/MWh without TS&M). The economic results are summarized in **Table 5-89**. A detailed breakdown of the COE is provided in **Figure 5-135**.

Table 5-88: Equipment Cost of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	177,641,950
SOFC Stacks	\$	63,052,160
Power Conditioning	\$	75,087,410
Pressure Vessel, Housing & Assembly	\$	34,496,290
System Installation	\$	5,006,100
Gas Turbine Island	\$	32,859,130
Gas Turbine-Generator Set	\$	27,698,410
Gas Turbine Foundation	\$	402,640
Piping, Instrumentation & Electrical Equipment	\$	4,758,080
Gasifier and Feed Preparation	\$	247,006,130
Feed Preparation	\$	49,421,300
Gasifier System	\$	192,136,970
Air Supply and Flue Gas Expansion	\$	5,447,860
Gas Processing and Flue Gas Cleanup	\$	85,179,410
Syngas Cleanup	\$	40,562,120
Flue Gas Cleanup	\$	44,617,290
Carbon Capture	\$	31,760,480
Combined WGS & Carbon Capture PSA	\$	22,952,700
CO ₂ Compression and Dehydration	\$	8,807,780
Steam and Water System	\$	22,996,580
Water Recovery	\$	N/A
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Auxiliary Plant Equipment Cost	\$	67,361,350
Accessory Electric Plant	\$	12,880,850
Instrumentation & Controls	\$	24,412,980
Improvement to Site	\$	6,347,380
Buildings & Structures	\$	23,720,140
Total Plant Cost	\$	664,805,030
Specific Plant Cost	\$/kW	6,097

Table 5-89: Economics of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	85
Total Overnight Capital Cost		
Total Plant Cost	\$	664,805,030
Preproduction Costs		
6 Months Labor Cost	\$	7,938,900
1 Month Maintenance Materials	\$	1,065,880
1 Month Non-Fuel Consumables	\$	433,130
1 Month Waste Disposal	\$	95,050
25% of 1 month fuel cost at 100% CF	\$	177,660
2% of TPC	\$	13,296,100
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	1,765,020
0.5% of TPC (Spare Parts)	\$	3,324,030
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	3,497,680
Land	\$	4,970,930
Other Owner's Costs (15% of TPC)	\$	99,720,750
Financing Costs	\$	17,949,740
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	7,248,330
Other Annual Variable Operating Cost at above CF for Initial Year	\$	16,259,450
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	12,477,860
Annual Byproduct Revenues at above CF for Initial Year	\$	1,972,700
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	15,877,790
Property Tax & Insurance	\$	13,296,100
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	142.75
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	158.12
First Year Levelized Cost of Water	\$/L	N/A

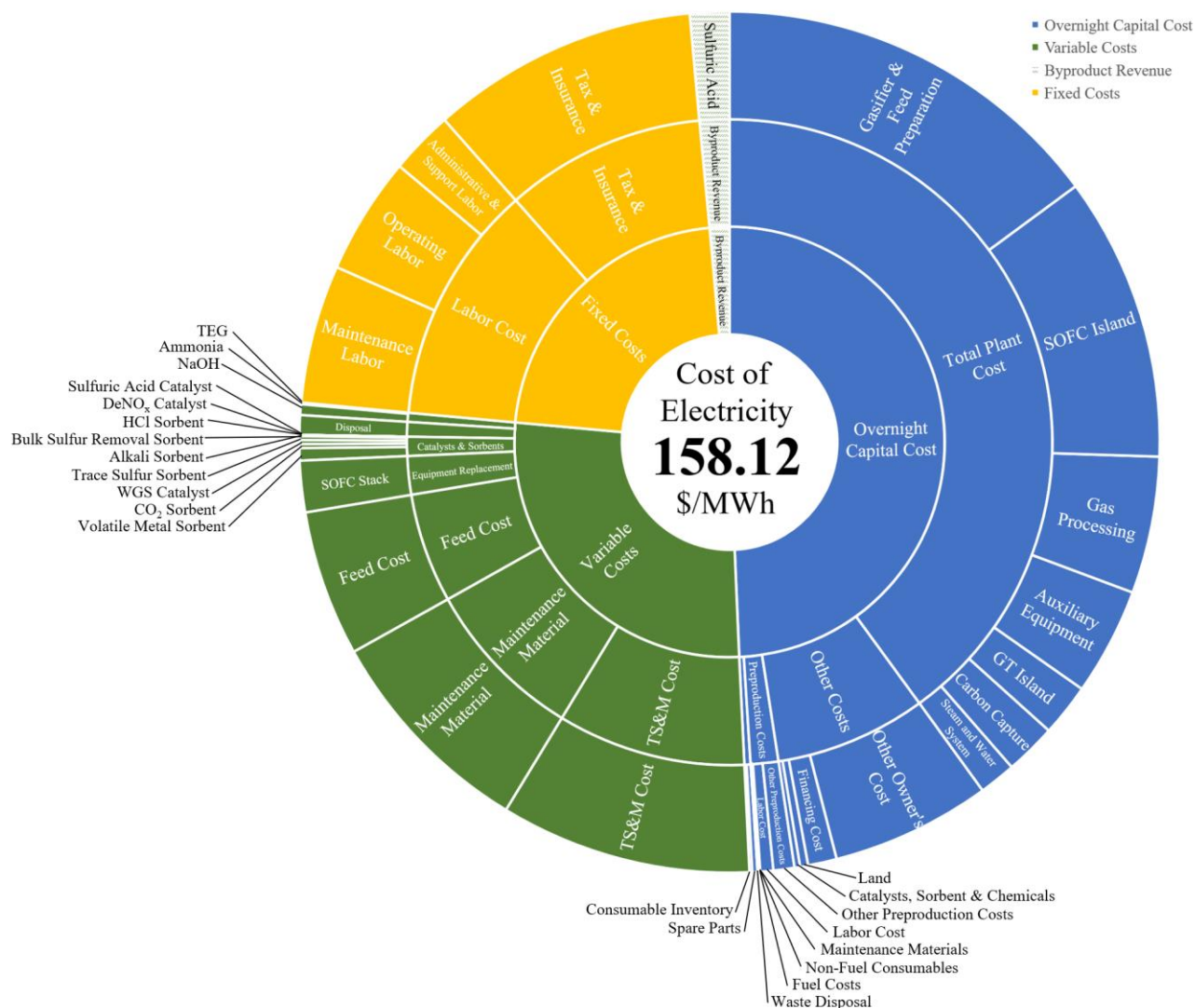


Figure 5-135: Breakdown of the cost-of-electricity of the 100 MW coal gasification SOFC-GT hybrid with carbon capture and without water recovery.

LiBr Absorption System – Coal with Carbon Capture

The coal gasification SOFC-GT hybrid plant with LiBr absorber for water recovery follows the basic configuration described before. The LiBr absorption system is designed as single train unit and its design is identical to the 10 MW LiBr absorber case and the 50MW LiBr absorption system cases. A flue gas heat exchanger and a packed-bed column are installed downstream of the GT and the exhaust gas steam generator. A packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream heat exchanger is used to provide the heat for sorbent regeneration while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, set by an air-cooled condenser, supported by a vacuum compressor to remove any non-condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is identical to previous LiBr absorber cases and a similar plume rise can be expected. The configuration of the 100 MW coal gasification reference plant with LiBr

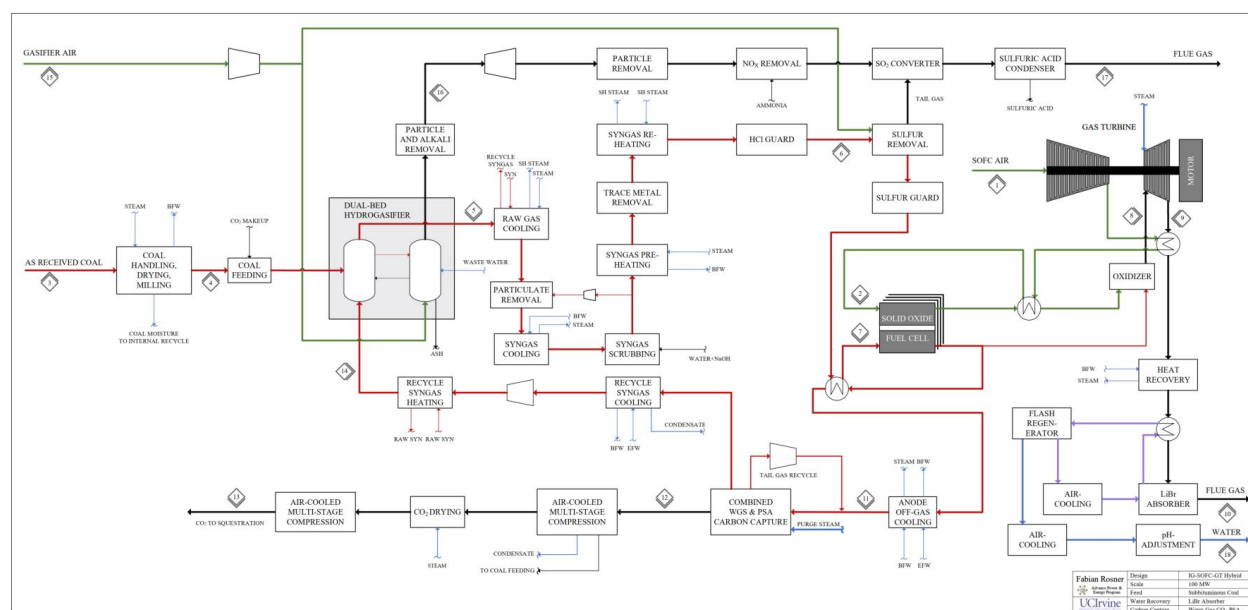


Figure 5-136: Flowsheet of the 100 MW coal gasification SOFC-GT hybrid system with carbon capture and with LiBr absorption system for water recovery.

flue gas water recovery is provided in **Figure 5-136** (a larger version of this illustration can be found in Appendix B). The corresponding state-point steam data are shown in **Table 5-90**.

Table 5-90: Stream Summary of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	71	751	315	700	814	546
Pressure	bar	1.01	6.60	1.01	1.01	15.00	10.05	6.60	5.82	1.08
Molar Flow Rate	kmol/h	10,012.0	10,012.0	-	-	7,371.4	7,716.1	7,716.1	9,470.3	9,470.3
Mass Flow Rate	kg/h	288,917	288,917	41,867	37,900	84,564	90,739	90,611	259,328	259,328
Mass Vapor Fraction	-	1	1	0	0	1	1	1	1	1
Composition	mole-basis									
O ₂	-	0.20741	0.20741	-	-	0.00000	0.00000	0.00000	0.05105	0.05105
N ₂	-	0.77315	0.77315	-	-	0.02375	0.02269	0.02269	0.81837	0.81837
AR	-	0.00924	0.00924	-	-	0.00000	0.00000	0.00000	0.00977	0.00977
H ₂	-	0.00000	0.00000	-	-	0.50348	0.48099	0.48099	0.00000	0.00000
CO	-	0.00000	0.00000	-	-	0.05794	0.05533	0.05533	0.00000	0.00000
CO ₂	-	0.00033	0.00033	-	-	0.03638	0.03459	0.03460	0.00991	0.00991
H ₂ O	-	0.00987	0.00987	-	-	0.24101	0.27539	0.27642	0.11089	0.11089
CH ₄	-	0.00000	0.00000	-	-	0.13590	0.12958	0.12958	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	-	-	0.00109	0.00103	Trace	0.00000	0.00000
SO ₂	-	0.00000	0.00000	-	-	Trace	Trace	Trace	Trace	Trace
HCl	-	0.00000	0.00000	-	-	0.00000	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	-	-	0.99956	0.99959	0.99960	0.99999	0.99999
Stream Number	Unit	10	11	12	13	14	15	16	17	18
Temperature	°C	49	165	207	40	600	15	946	90	40
Pressure	bar	1.01	5.95	0.90	151.65	17.30	1.01	14.70	1.02	2.00
Molar Flow Rate	kmol/h	8,631.5	8,836.4	4,394.3	1,331.8	6,511.8	1,947.1	1,914.7	1,980.0	838.8
Mass Flow Rate	kg/h	244,218	129,777	113,200	57,391	54,062	56,189	58,695	60,590	15,112
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.05601	0.00000	0.00000	0.00000	0.00000	0.20741	0.02500	0.02534	Trace
N ₂	-	0.89789	0.01878	0.00025	0.00081	0.02532	0.77315	0.74930	0.76048	Trace
AR	-	0.01072	0.00000	0.00000	0.00000	0.00000	0.00924	0.00896	0.00909	Trace
H ₂	-	0.00000	0.39503	0.00624	0.02052	0.63788	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.08386	0.00008	0.00025	0.00766	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.01088	0.07717	0.30171	0.97702	0.00836	0.00033	0.17551	0.16973	Trace
H ₂ O	-	0.02449	0.40449	0.69145	0.00051	0.29295	0.00987	0.04011	0.03500	0.99999
CH ₄	-	0.00000	0.02040	0.00027	0.00088	0.02747	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
SO ₂	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCl	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Total	-	0.99998	0.99973	1	0.99999	0.99964	1	0.99888	0.99963	0.99999

Thermodynamic Performance. The net power output of the nominal 100 MW coal gasification hybrid plant with LiBr absorber for water recovery is 106,725 kW. The power generation of the SOFC (128,710 kW, FU 35.8%) is very similar to the reference plant. Similarly, the GT that is providing the air for the gasifier has an identical power output as in the reference case with 2,653 kW. The nine Dresser Rand GTs provide 1,063 kW of electricity, which is substantially lower than in the 100 MW coal gasification reference case and due to the increased backpressure as well as a lower TIT.

The auxiliary load accounts for 25,710 kW. The majority of the auxiliary loads remain relatively constant and the water recovery together with the water treatment add about 579 kW to the plant's auxiliary load. The efficiency of the 100 MW coal gasification reference case with LiBr absorber for water recovery is 46.15%-HHV and is about 1.01% points lower than the reference case. This efficiency decrease is similar to the decrease seen for the NG SOFC-GT hybrid cases without carbon capture. In cases with carbon capture the efficiency decrease is typically around 0.56% points. One reason for the steeper efficiency decrease in the coal gasification case with carbon capture is the heat integration, which can be further improved to achieve higher TITs. The SOFC stack efficiency stays at 83.5%-LHV, while the GT efficiency further decreases to 5.6% due to the lower TIT and a more pronounced decrease in component efficiencies. The surge margin in this case is 29.1%.

The plant is equipped with dry cooling technology and as previously mentioned, does not produce any liquid waste streams. Solid waste streams are associated with the gasifier operation and contain ash as well as unconverted feedstock (disposal of catalysts, etc. are handled separately). Carbon emissions for fossil fuel plants are directly related to the efficiency as well as the fuel. Coal with its high carbon content per unit of energy leads to higher carbon emissions than natural gas. With the dual-bed gasifier design a carbon capture rate of 74.74% is achieved. The CO₂ emissions of the 100 MW coal gasification SOFC-GT hybrid with 74.74% carbon capture and flue gas water recovery are 185 kg/MWh.

At the scale of this nominal 100 MW plant, a total of 362,674 kg of water can be exported per day (note that ash treatment also requires considerable amounts of water, which substantially reduces the amount of exported water especially for high ash feedstocks). Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from this 100 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 177,100 m² of farmland (including 85% capacity factor). The thermodynamic performance results are summarized in **Table 5-91**.

Table 5-91: Thermodynamic Performance of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	128,718.30
Gas Turbine (SOFC-GT Hybrid)	kW	1,063.38
Gas Turbine (Gasifier)	kW	2,653.18
Total Generation	kW	132,434.85
Auxiliary Load		
DC-AC Inverter Losses	kW	2,574.37
Feedstock Handling, Milling and Drying	kW	989.85
Gasifier Feeding System	kW	515.40
Gasifier Ash Handling	kW	90.88
Syngas Cleaning and Scrubbing	kW	137.92
CO ₂ Capture and Compression	kW	9,061.10
Syngas Recycle Compressors	kW	11,076.05
Boiler and Feed Water System	kW	100.92
Water Recovery System	kW	576.13
Water Treatment	kW	3.06
Miscellaneous Balance of Plant	kW	85.93
Transformer Losses	kW	498.00
Total Auxiliary Load	kW	25,709.60
Net Power Generation	kW	106,725.25
Plant Performance		
Net Efficiency	%-HHV	46.15
Net Efficiency	%-LHV	47.90
Consumables		
Feedstock	kg/h	41,867
Thermal Input	MJ _{HHV} /h	833,988
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	74.74
Carbon Emission	kg/h	19,701
Carbon Emission	kg/MWh	185
Liquid Waste	kg/h	0
Solid Waste	kg/h	3,449
Water Recovery		
Flue Gas Water Recovery	%	79.9
Flue Gas Water Recovery	kg/d	362,674

Economics. The TPC of the 100 MW coal gasification hybrid with carbon capture and with LiBr absorber for water recovery is \$672,487,000 and the specific plant cost is 6,301 \$/kW. The gasification and feed preparation section of the plant contributes the largest cost to the TPC with \$247,020,340. The SOFC island accounts for \$177,596,800. Due to the use of the same GTs as in the reference case, the GT island costs remain constant at \$32,859,130. Also, the other plant equipment costs shown only minor changes due to plant scaling. The plant is scaled to the size of the GT (an off-the-shelf item) and due to the increased GT backpressure associated with the water recovery a small change in the plant scale is seen. The TPC increase seen in this case is related to the equipment cost of the water recovery and water treatment, which account for \$7,252,240. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-137**. A more detailed overview of the capital cost is provided in **Table 5-92**.

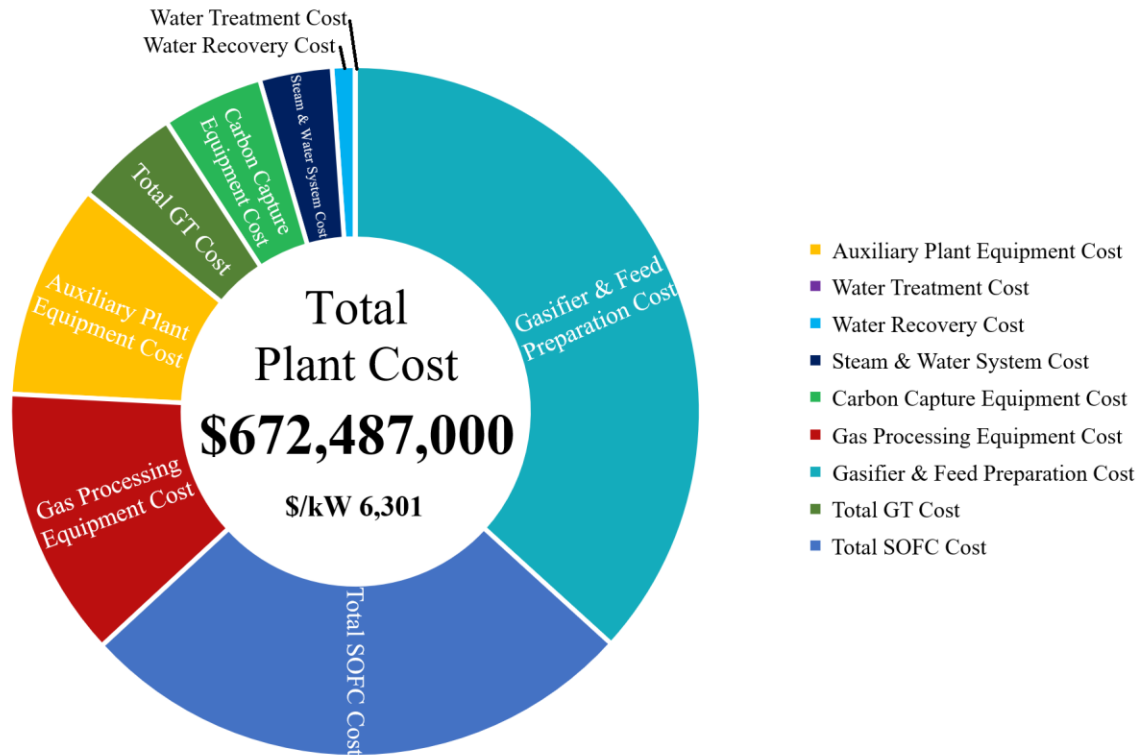


Figure 5-137: Total plant cost and specific plant cost of the 100 MW coal gasification SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

The total overnight capital cost is \$831,845,120 and is comprised of the TPC, the preproduction costs with \$23,343,100, the inventory capital costs with \$5,127,610 and other costs including financing and land costs with \$130,887,410. Variable annual costs are fuel costs with \$7,248,820 and other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, as well as waste disposal costs, which amount to \$16,384,440 per year. Annual CO₂ transporting, storing, and monitoring account for \$12,478,710. The sulfuric acid sales revenue generates an annual income of \$1,972,840, which is equivalent to a COE reduction of 2.48 \$/MWh and the byproduct revenue of the recovered water is \$4,368,300 (-5.50 \$/MWh). Fixed costs are comprised of annual labor costs with \$16,218,760 and costs associated with tax and insurance with \$13,449,740 per year. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the 100 MW coal gasification reference case. Thus, the COE remains constant at 158.12 \$/MWh. The COW in the LiBr absorber case is 0.0388 \$/L. A summary of the economic key parameters can be found in **Table 5-93**. A detailed breakdown of the COE is shown in **Figure 5-138**.

Table 5-92: Equipment Cost of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	177,596,800
SOFC Stacks	\$	63,052,160
Power Conditioning	\$	75,042,420
Pressure Vessel, Housing & Assembly	\$	34,496,040
System Installation	\$	5,006,180
Gas Turbine Island	\$	32,859,130
Gas Turbine-Generator Set	\$	27,698,410
Gas Turbine Foundation	\$	402,640
Piping, Instrumentation & Electrical Equipment	\$	4,758,080
Gasifier and Feed Preparation	\$	247,020,340
Feed Preparation	\$	49,423,330
Gasifier System	\$	192,148,930
Air Supply and Flue Gas Expansion	\$	5,448,090
Gas Processing and Flue Gas Cleanup	\$	85,175,870
Syngas Cleanup	\$	40,564,100
Flue Gas Cleanup	\$	44,611,770
Carbon Capture	\$	31,762,030
Combined WGS & Carbon Capture PSA	\$	22,953,900
CO ₂ Compression and Dehydration	\$	8,808,130
Steam and Water System	\$	22,997,630
Water Recovery	\$	7,252,240
Water Recovery System	\$	7,067,100
Water Treatment	\$	185,140
Auxiliary Plant Equipment Cost	\$	67,822,970
Accessory Electric Plant	\$	12,712,930
Instrumentation & Controls	\$	24,695,080
Improvement to Site	\$	6,420,720
Buildings & Structures	\$	23,994,230
Total Plant Cost	\$	672,487,000
Specific Plant Cost	\$/kW	6,301

Figure 5-138 shows the COE value of 158.12 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-7.98 \$/MWh) and the whole circle represents the COE without byproduct credit (166.10 \$/MWh). This means, if the water would be provided to the local community at no cost, the COE would increase to 163.62 \$/MWh due to the additional capital and operational costs of the water recovery units. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0388 \$/L in order to reach the COE of 158.12 \$/MWh.

Table 5-93: Economics of the 100 MW Coal Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
Plant Capacity Factor	%	85
Total Overnight Capital Cost		
Total Plant Cost	\$	672,487,000
Preproduction Costs		
6 Months Labor Cost	\$	8,109,380
1 Month Maintenance Materials	\$	1,078,100
1 Month Non-Fuel Consumables	\$	433,160
1 Month Waste Disposal	\$	95,060
25% of 1 month fuel cost at 100% CF	\$	177,670
2% of TPC	\$	13,449,740
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	1,765,180
0.5% of TPC (Spare Parts)	\$	3,362,430
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	6,772,800
Land	\$	5,084,410
Other Owner's Costs (15% of TPC)	\$	100,873,050
Financing Costs	\$	18,157,150
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	7,248,820
Other Annual Variable Operating Cost at above CF for Initial Year	\$	16,384,440
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	12,478,710
Annual Byproduct Revenues at above CF for Initial Year	\$	6,341,140
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	16,218,760
Property Tax & Insurance	\$	13,449,740
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	142.42
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	158.12
First Year Levelized Cost of Water	\$/L	0.0388

Figure 5-139 shows a detailed breakdown of the COW. The fixed operating cost in the 100 MW coal gasification hybrid with LiBr absorber is substantially reduced compared to the 10 MW case and the 50 MW cases since the water recovery unit requires the same number of operators independent of the scale; however, in the 100 MW coal gasification case the amount of recovered water is significantly higher, reducing the relative operating labor cost. While the fixed operating cost in the 100 MW coal gasification case looks small compared to the 100 MW NG case on a relative basis (in the diagram), the fixed operating costs are actually greater on a \$/L basis in the coal gasification case because of the lower water recovery. The largest COW cost factor is the performance penalty cost, in particular the system integration losses. The system integration losses include the increased GT backpressure, but also, the heat integration in this specific scenario plays a crucial role in the large system integration losses. As mentioned earlier, the heat integration is currently not maximizing the TIT nor steam injection, which provides space for future system optimization studies to further reduce the COW. The TPC's share on the COW is 14.4% and water treatment is responsible for only 0.8% of the COW.

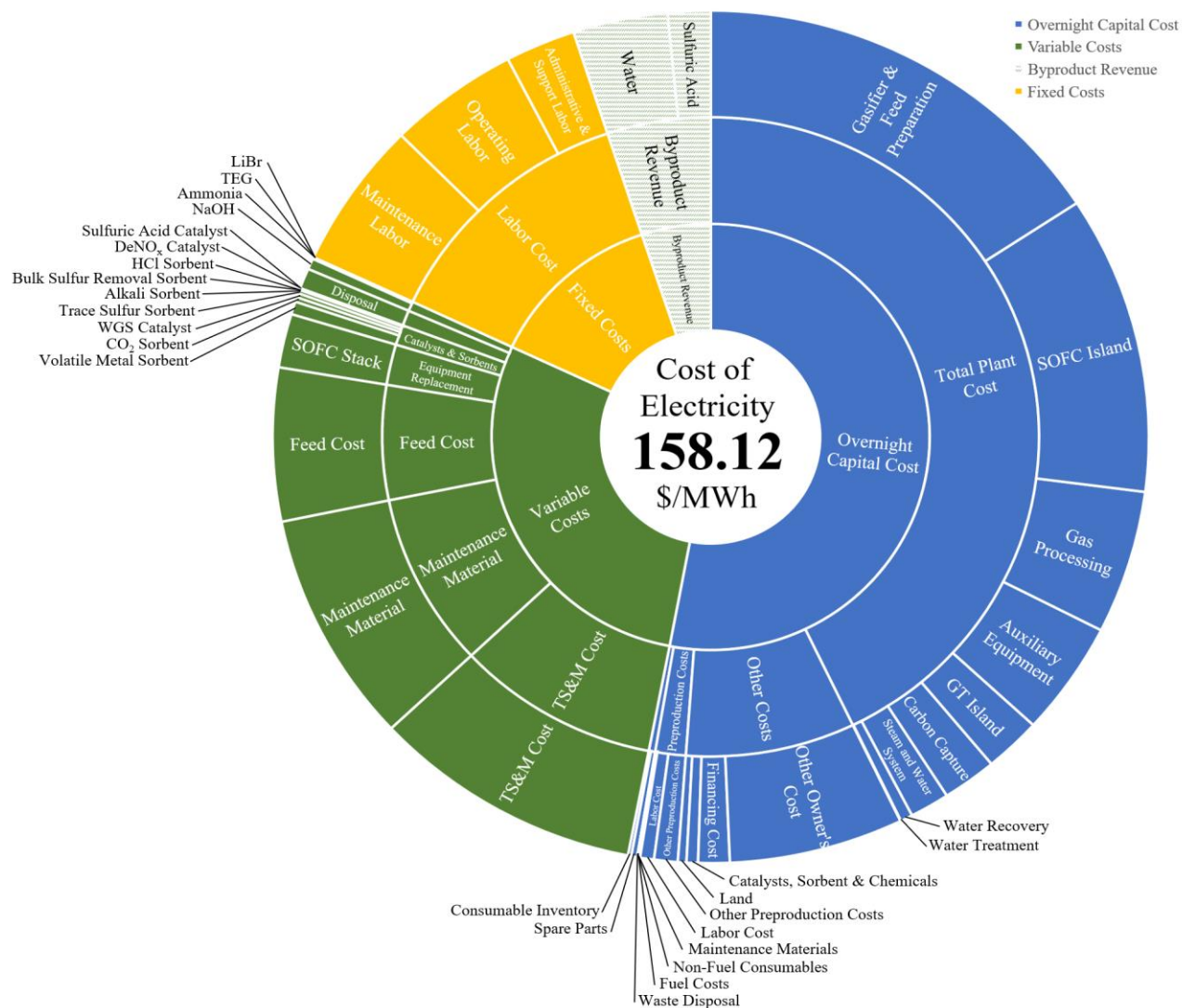


Figure 5-138: Breakdown of the cost-of-electricity of the 100 MW coal gasification SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

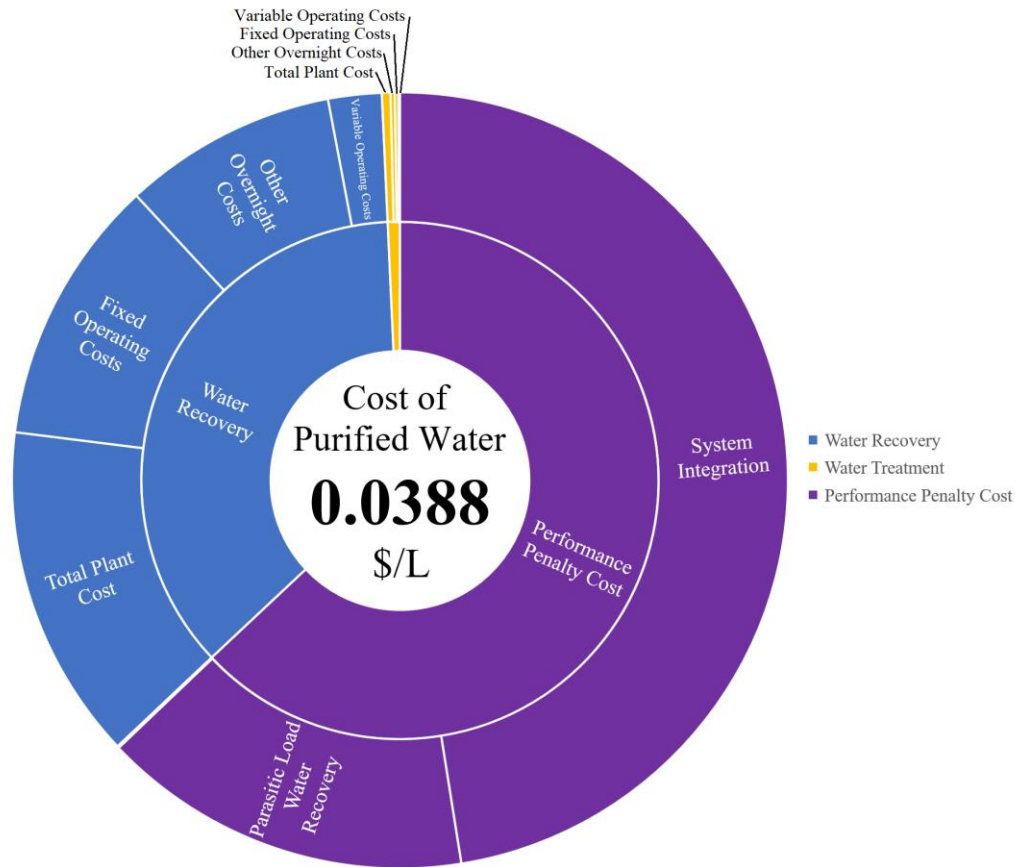


Figure 5-139: Breakdown of the cost-of-water of the 100 MW coal gasification SOFC-GT hybrid with carbon capture and with LiBr absorber for water recovery.

Reference Case – Biomass with Carbon Capture

The design of the biomass gasification SOFC-GT hybrid plant is essentially the same as the coal gasification design. Small modifications are needed in biomass handling, drying and milling upstream up the dual-bed hydrogasifier. The biomass is dried to a very low moisture content in a vapor compression dryer, which is more efficient when removing large amounts of moisture, to increase its friability and reduce milling power. The rest of the plant configuration remains largely unchanged; however, process units are sized based upon the respective requirements. The raw syngas exiting the gasifier is cleaned in multiple steps. First the syngas is cooled, and particulates are removed in candle filters before it is scrubbed in a venturi scrubber using water and NaOH to remove any fine particulates as well as water soluble compounds such as HCl. Downstream of the scrubber, volatile trace metals are removed in an expandable process and a HCl guard bed removes any remaining HCl before the syngas enters the bulk sulfur removal process. Since SOFCs have a very low tolerance for sulfur, a sulfur guard bed is installed downstream of the bulk sulfur removal unit. The recovered sulfur from the bulk sulfur removal unit is mixed with the flue gas from the char

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converter in the SNOX™ unit where it is ultimately reacted to sulfuric acid, a saleable byproduct. The clean syngas is now preheated and ready to enter the SOFC module. No pre-reforming or recycle are needed in this design as the syngas is rich in moisture and relatively low FU factors are desired to recycle enough hydrogen back to the gasifier. The air supply for the SOFC is separate from the gasifier air supply and the same Dresser-Rand GT as in the previous scenarios is used for the hybridization of the SOFC. In total nine GTs are needed to reach the nominal 100 MW plant scale. A small portion of the anode off-gas is sent to the GT oxidizer, while the majority of the syngas is decarbonized and recycled back to the gasifier after adjusting the moisture content. Downstream of the SOFC-GT hybrid, heat is recovered to raise more steam. The plant is able to satisfy its own water demand since substantial amounts of water are recovered during CO₂ compression and the compression of the syngas recycle. Any excess water is used to raise steam and injected into the GT of the SOFC-GT hybrid unit. Water unsuitable for steam generation is used for ash treatment and/or injected into the char converted so that no liquid waste streams are generated by this plant. The configuration of the 100 MW biomass gasification reference plant without flue gas water recovery is provided in **Figure 5-140** (a larger version of this illustration can be found in Appendix B). The corresponding state-point steam data are shown in **Table 5-94**.

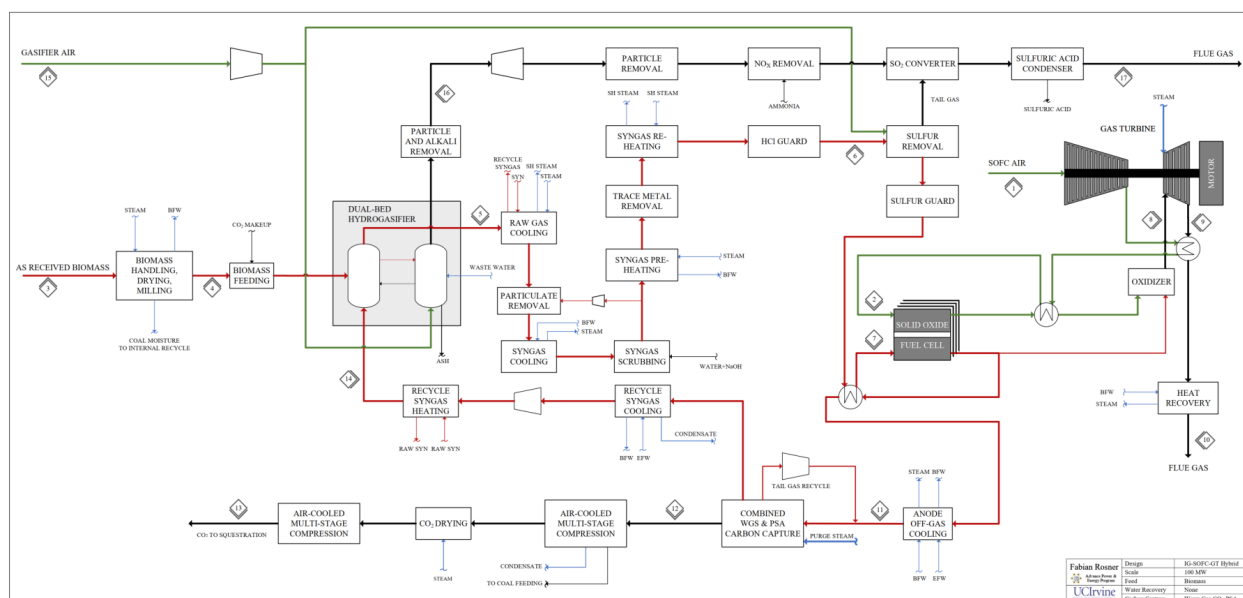


Figure 5-140: Flowsheet of the 100 MW biomass gasification SOFC-GT hybrid system with carbon capture and without water recovery.

Table 5-94: Stream Summary of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	73	761	315	700	772	517
Pressure	bar	1.01	6.60	1.01	1.01	15.00	10.05	6.60	5.82	1.05
Molar Flow Rate	kmol/h	10,013.1	10,013.1	-	-	8,717.1	9,071.1	9,071.1	10,162.5	10,162.5
Mass Flow Rate	kg/h	288,947	288,947	53,909	47,240	101,650	107,959	107,939	271,838	271,838
Mass Vapor Fraction	-	1	1	0	0	1	1	1	1	1
Composition	mole-basis									
O ₂	-	0.20741	0.20741	-	-	0.00000	0.00000	0.00000	0.04976	0.04976
N ₂	-	0.77315	0.77315	-	-	0.01019	0.00979	0.00979	0.76220	0.76220
AR	-	0.00924	0.00924	-	-	0.00000	0.00000	0.00000	0.00910	0.00910
H ₂	-	0.00000	0.00000	-	-	0.50229	0.48269	0.48270	0.00000	0.00000
CO	-	0.00000	0.00000	-	-	0.06096	0.05857	0.05857	0.00000	0.00000
CO ₂	-	0.00033	0.00033	-	-	0.04423	0.04234	0.04234	0.00837	0.00837
H ₂ O	-	0.00987	0.00987	-	-	0.28825	0.31653	0.31666	0.17055	0.17055
CH ₄	-	0.00000	0.00000	-	-	0.09348	0.08971	0.08971	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	-	-	0.00014	0.00013	Trace	0.00000	0.00000
SO ₂	-	0.00000	0.00000	-	-	Trace	Trace	Trace	Trace	Trace
HCl	-	0.00000	0.00000	-	-	0.00000	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	-	-	0.99955	0.99977	0.99977	0.99999	0.99999
Stream Number	Unit	10	11	12	13	14	15	16	17	-
Temperature	°C	265	165	204	40	600	15	934	90	-
Pressure	bar	1.01	5.95	0.89	151.65	17.30	1.01	14.70	1.02	-
Molar Flow Rate	kmol/h	10,162.5	9,912.8	4,832.4	1,423.3	7,056.4	2,042.0	2,108.0	2,195.4	-
Mass Flow Rate	kg/h	271,838	146,927	123,509	61,432	59,242	58,927	63,405	65,920	-
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	-
Composition	mole-basis									
O ₂	-	0.04976	0.00000	0.00000	0.00000	0.00000	0.20741	0.02435	0.03072	-
N ₂	-	0.76220	0.00857	0.00011	0.00037	0.01196	0.77315	0.71553	0.71929	-
AR	-	0.00910	0.00000	0.00000	0.00000	0.00000	0.00924	0.00855	0.00859	-
H ₂	-	0.00000	0.37315	0.00570	0.01928	0.62281	0.00000	0.00000	0.00000	-
CO	-	0.00000	0.07706	0.00005	0.00017	0.00565	0.00000	0.00000	0.00000	-
CO ₂	-	0.00837	0.07498	0.29365	0.97903	0.00799	0.00033	0.16820	0.16151	-
H ₂ O	-	0.17055	0.45196	0.70030	0.00053	0.33166	0.00987	0.08293	0.07981	-
CH ₄	-	0.00000	0.01415	0.00018	0.00061	0.01973	0.00000	0.00000	0.00000	-
C ₂ H ₆	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
C ₃ H ₈	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	-
SO ₂	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	-
HCl	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	-
Total	-	0.99999	0.99986	1	0.99999	0.99980	1	0.99957	0.99992	-

Thermodynamic Performance. The net power output of the nominal 100 MW biomass gasification reference case with carbon capture is 104,647 kW, whereby, the SOFC produces 128,146 kW of electricity at a FU of 35.4%. The nine Dresser Rand GTs provide another 1,668 kW of electricity, which is substantially lower than in the 100 MW coal gasification reference case. The reason for this decrease in GT power output is the reduced TIT and reduced GT component efficiencies, especially of the GT expander. In the gasification scenario, gasifier flue gas expansion provides additional power. By mounting the air supply compressor and gasifier flue gas expander onto the same shaft, the expander can provide the power

needed for compression and an additional 3,174 kW. Overall, this leads to gross power generation of 132,989 kW for this plant.

The auxiliary load increases from 25,138 kW in the coal gasification reference case to 28,342 kW in the biomass gasification reference case. This increase is mostly related to biomass handling, drying and milling but also the syngas recycle increases due to the biomass's fuel composition. The recycle alone is responsible for 42.5% of the auxiliary load and accounts for 12,054 kW. The DC-AC inverter losses are only the third largest auxiliary load and associated with the conversion of the direct current produced by the SOFC to alternating current used in the electricity grid. With an inverter efficiency of 97%, these losses represent 3% of the gross power generated by the SOFC and are responsible for 2,563 kW of the total auxiliary load. The efficiency of the 100 MW biomass gasification reference case is 41.99%-HHV. The SOFC stack efficiency in the biomass gasification case is 79.7%-LHV and slightly reduced versus the coal gasification case due to the lower CH₄ content in the syngas. The GT efficiency dropped to 9.9%-LHV as the TIT and component efficiencies decrease under the present operating condition. The surge margin in this case is 28.9% and substantially increased compared to the NG cases.

The plant is equipped with dry cooling technology and as previously mentioned, does not produce any liquid waste streams. Solid waste streams are associated with the gasifier operation and contain ash as well as unconverted feedstock (disposal of catalysts, etc. are handled separately). By using biomass as feedstock and incorporating a CO₂ removal technology, the plant produces negative CO₂ emissions, meaning it actively removes CO₂ from the atmosphere. With this dual-bed gasifier design a carbon capture rate of 75.62% is achieved, which leads to a CO₂ removal rate of 61,328 kg/h (-586 kg/MWh). The thermodynamic performance results are summarized in **Table 5-95**.

Table 5-95: Thermodynamic Performance of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Gross Power Generation	Unit	Reference Case
Fuel Cell	kW _{DC}	128,146.42
Gas Turbine (SOFC-GT Hybrid)	kW	1,668.43
Gas Turbine (Gasifier)	kW	3,174.49
Total Generation	kW	132,989.33
Auxiliary Load		
DC-AC Inverter Losses	kW	2,562.93
Feedstock Handling, Milling and Drying	kW	2,532.64
Gasifier Feeding System	kW	644.39
Gasifier Ash Handling	kW	18.69
Syngas Cleaning and Scrubbing	kW	164.39
CO ₂ Capture and Compression	kW	9,656.50
Syngas Recycle Compressors	kW	12,054.11
Boiler and Feed Water System	kW	122.07
Water Recovery System	kW	N/A
Water Treatment	kW	N/A
Miscellaneous Balance of Plant	kW	86.29
Transformer Losses	kW	500.08
Total Auxiliary Load	kW	28,342.10
Net Power Generation	kW	104,647.23
Plant Performance		
Net Efficiency	%-HHV	41.99
Net Efficiency	%-LHV	45.15
Consumables		
Feedstock	kg/h	53,909
Thermal Input	MJ _{HHV} /h	897,054
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	75.62
Carbon Emission	kg/h	-61,328
Carbon Emission	kg/MWh	-586
Liquid Waste	kg/h	0
Solid Waste	kg/h	710
Water Recovery		
Flue Gas Water Recovery	%	N/A
Flue Gas Water Recovery	kg/d	N/A

Economics. The TPC of the 100 MW biomass gasification hybrid with carbon capture and without water recovery is \$764,647,440 and the specific plant cost is 7,307 \$/kW. The gasification and feed preparation section of the plant contributes the largest cost to the TPC with \$346,313,230, in particular feedstock preparation and the gasifier costs increased substantially compared to the coal gasification case. The SOFC island accounts for \$165,134,410, whereby, the SOFC stacks do not represent the largest cost anymore. At low fuel utilization factors the number of stacks can be reduced (on a relative basis) and the inverter cost becomes the largest cost item in the SOFC island. Compared to the coal gasification, SOFC stack cost is reduced, which is a consequence of the lower CH₄ concentration in the syngas and leads to improved reaction kinetics inside the SOFC. The GT island cost remains at \$32,859,130. Syngas cleanup and flue gas cleanup account for \$85,179,410, which is similar to the coal case; however, costs for cleanup are shifted away from sulfur components more towards chlorine compounds and alkali metals. The carbon capture

equipment in this gasification plant has a relatively small share on the plant's TPC with 4.4% and the steam and water system accounts for \$26,010,800. Auxiliary plant equipment costs are estimated with \$75,428,700, which includes electric plant accessory, instrumentation and controls, improvement to site, and buildings and structures. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-141**. A more detailed overview of the capital cost is provided in **Table 5-96**.

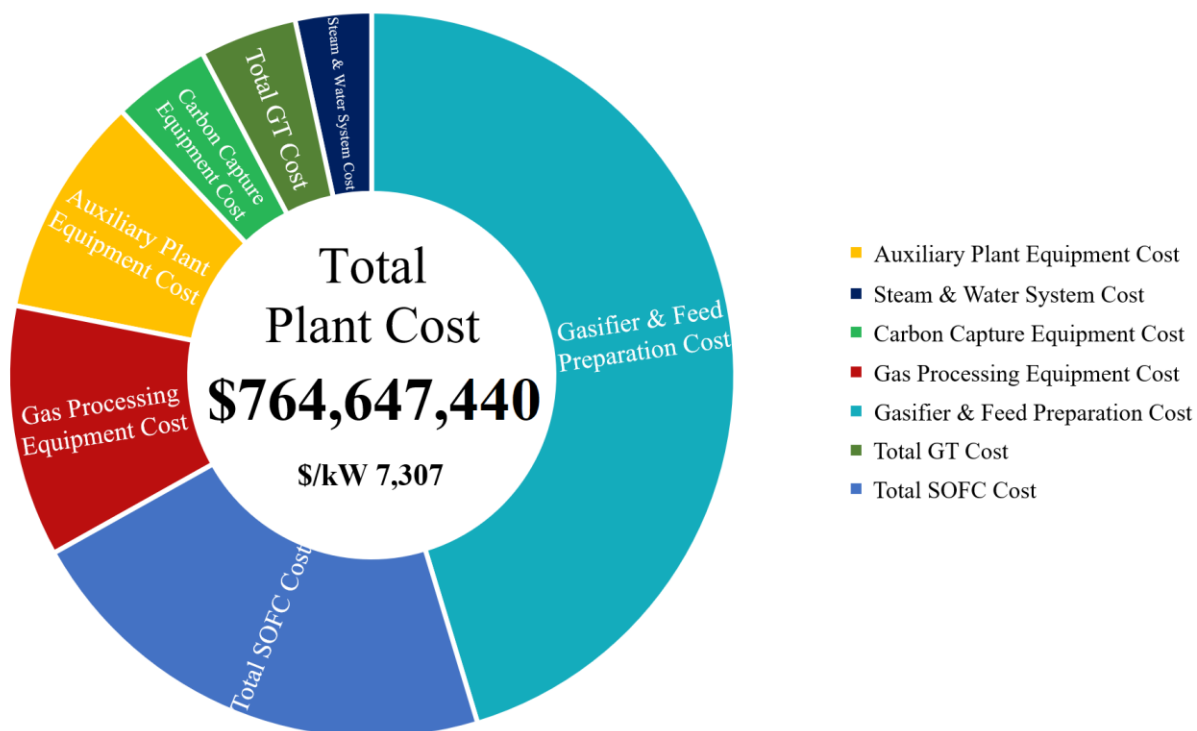


Figure 5-141: Total plant cost and specific plant cost of the 100 MW biomass gasification SOFC-GT hybrid with carbon capture and without water recovery.

The total overnight capital cost is \$951,367,290 and is comprised of the TPC, the preproduction costs with \$27,351,320, the inventory capital costs with \$14,258,700 and other costs including financing and land costs with \$145,109,830. Biomass feedstock costs are substantially higher than coal and the annual feed costs account for \$49,797,570. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, as well as waste disposal costs, and amount to \$20,203,830 per year. Annual CO₂ transporting, storing, and monitoring costs \$13,355,250. With the lower sulfur content found in biomass the sulfuric acid byproduct revenue is only \$303,480, and it is questionable if the capital investment in the bulk sulfur removal unit is justified. Fixed costs are comprised of annual labor costs with \$17,714,150 and costs associated with tax and insurance with \$15,292,950 per year. The resulting COE for the 100 MW biomass gasification SOFC-GT hybrid plant with carbon capture and

without water recovery is 246.13 \$/MWh with TS&M (229.00 \$/MWh without TS&M). The economic results are summarized in **Table 5-97**. A detailed breakdown of the COE is provided in **Figure 5-142**.

Table 5-96: Equipment Cost of the 100 MW Biomass Gasification
SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Equipment Cost	Unit	Reference Case
Fuel Cell Island	\$	165,134,410
SOFC Stacks	\$	55,776,910
Power Conditioning	\$	74,674,480
Pressure Vessel, Housing & Assembly	\$	30,075,160
System Installation	\$	4,607,860
Gas Turbine Island	\$	32,859,130
Gas Turbine-Generator Set	\$	27,698,410
Gas Turbine Foundation	\$	402,640
Piping, Instrumentation & Electrical Equipment	\$	4,758,080
Gasifier and Feed Preparation	\$	346,313,230
Feed Preparation	\$	74,004,220
Gasifier System	\$	266,735,750
Air Supply and Flue Gas Expansion	\$	5,573,260
Gas Processing and Flue Gas Cleanup	\$	85,421,440
Syngas Cleanup	\$	38,380,640
Flue Gas Cleanup	\$	47,040,790
Carbon Capture	\$	33,479,730
Combined WGS & Carbon Capture PSA	\$	24,281,780
CO ₂ Compression and Dehydration	\$	9,197,960
Steam and Water System	\$	26,010,800
Water Recovery	\$	N/A
Water Recovery System	\$	N/A
Water Treatment	\$	N/A
Auxiliary Plant Equipment Cost	\$	75,428,700
Accessory Electric Plant	\$	12,766,160
Instrumentation & Controls	\$	28,079,400
Improvement to Site	\$	7,300,640
Buildings & Structures	\$	27,282,500
Total Plant Cost	\$	764,647,440
Specific Plant Cost	\$/kW	7,307

Table 5-97: Economics of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and without Water Recovery

Item	Unit	Reference Case
Plant Capacity Factor	%	85
Total Overnight Capital Cost		
Total Plant Cost	\$	764,647,440
Preproduction Costs		
6 Months Labor Cost	\$	8,857,070
1 Month Maintenance Materials	\$	1,293,240
1 Month Non-Fuel Consumables	\$	627,580
1 Month Waste Disposal	\$	59,940
25% of 1 month fuel cost at 100% CF	\$	1,220,530
2% of TPC	\$	15,292,950
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	10,435,460
0.5% of TPC (Spare Parts)	\$	3,823,240
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	4,484,580
Land	\$	5,282,660
Other Owner's Costs (15% of TPC)	\$	114,697,120
Financing Costs	\$	20,645,480
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	49,797,570
Other Annual Variable Operating Cost at above CF for Initial Year	\$	20,203,830
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	13,355,250
Annual Byproduct Revenues at above CF for Initial Year	\$	303,480
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	17,714,150
Property Tax & Insurance	\$	15,292,950
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	229.00
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	246.13
First Year Levelized Cost of Water	\$/L	N/A

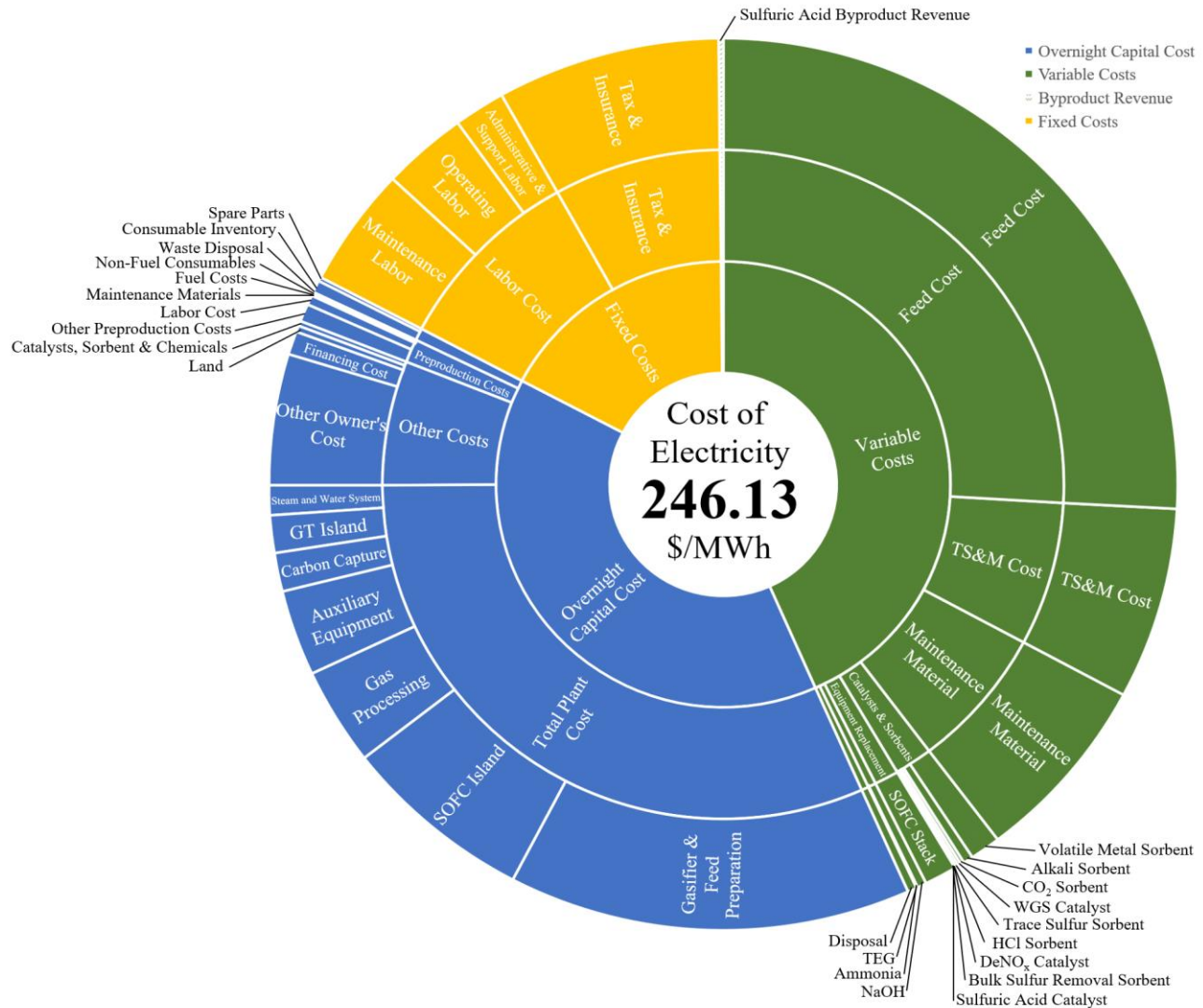


Figure 5-142: Breakdown of the cost-of-electricity of the 100 MW biomass gasification SOFC-GT hybrid with carbon capture and without water recovery.

LiBr Absorption System – Biomass with Carbon Capture

The biomass gasification SOFC-GT hybrid plant with LiBr absorber for water recovery follows the basic configuration of the biomass gasification plant without water recovery. The LiBr absorption system is designed as single train unit and its design is identical to the 10 MW LiBr absorber case and the 50 MW LiBr absorption system cases. A flue gas heat exchanger and a packed-bed column are installed downstream of the GT and the exhaust gas steam generator. A packed-bed column is used to bring the moist flue gas in contact with the liquid LiBr desiccant solution, it acts as an absorber, and the upstream heat exchanger is used to provide the heat for sorbent regeneration while cooling the flue gas down to acceptable column inlet temperatures. The hot LiBr solution leaving the LiBr heater is fed into a low-pressure regenerator flash drum. The regenerator pressure is set by the vapor condensation temperature, set by an air-cooled condenser,

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supported by a vacuum compressor to remove any non-condensable gases from the vessel. The regenerated desiccant is also cooled via an air-cooled exchanger before reintroducing it into the absorber. The recovered water requires cooling before its pH can be adjusted and the water can be exported from the plant site. A portion of the recovered water is used in the mist eliminator at the column outlet to prevent the formation of salt precipitants. Due to the regeneration step, which is similar to distillation, the obtained water is already of high quality [173] and only minimal water treatment is required eliminating the need for ultra-filtration and reverse osmosis. Since the ambient air acts as cooling medium for the liquid desiccant (25 °C pinch temperature) the flue gas leaves the column at a temperature of 49 °C. This flue gas temperature is identical to previous LiBr absorber cases and a similar plume rise can be expected. The configuration of the 100 MW biomass gasification reference plant with LiBr flue gas water recovery is provided in **Figure 5-143** (a larger version of this illustration can be found in Appendix B). The corresponding state-point steam data are shown in **Table 5-98**.

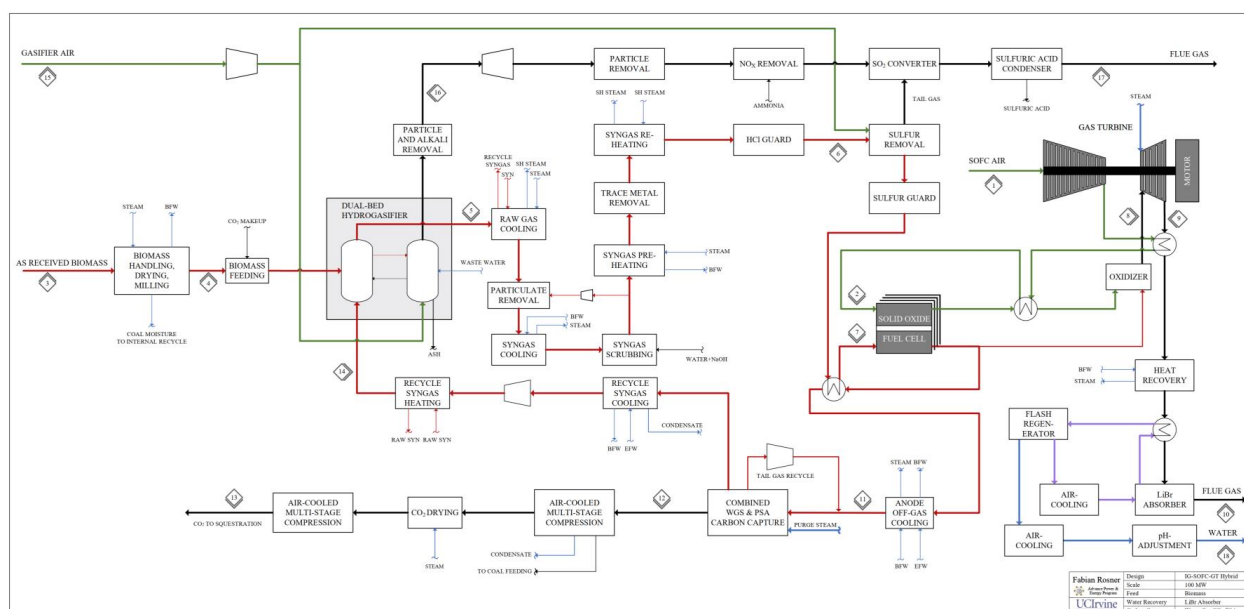


Figure 5-143: Flowsheet of the 100 MW biomass gasification SOFC-GT hybrid system with carbon capture and with LiBr absorption system for water recovery.

Table 5-98: Stream Summary of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Stream Number	Unit	1	2	3	4	5	6	7	8	9
Temperature	°C	15	700	15	73	761	315	700	756	521
Pressure	bar	1.01	6.60	1.01	1.01	15.00	10.05	6.60	5.82	1.08
Molar Flow Rate	kmol/h	10,012.4	10,012.4	-	-	8,716.5	9,070.5	9,070.5	10,161.8	10,161.8
Mass Flow Rate	kg/h	288,927	288,927	53,906	47,237	101,643	107,951	107,932	271,819	271,819
Mass Vapor Fraction	-	1	1	0	0	1	1	1	1	1
Composition	mole-basis									
O ₂	-	0.20741	0.20741	-	-	0.00000	0.00000	0.00000	0.04976	0.04976
N ₂	-	0.77315	0.77315	-	-	0.01019	0.00979	0.00979	0.76220	0.76220
AR	-	0.00924	0.00924	-	-	0.00000	0.00000	0.00000	0.00910	0.00910
H ₂	-	0.00000	0.00000	-	-	0.50229	0.48269	0.48270	0.00000	0.00000
CO	-	0.00000	0.00000	-	-	0.06096	0.05857	0.05857	0.00000	0.00000
CO ₂	-	0.00033	0.00033	-	-	0.04423	0.04234	0.04234	0.00837	0.00837
H ₂ O	-	0.00987	0.00987	-	-	0.28825	0.31653	0.31666	0.17055	0.17055
CH ₄	-	0.00000	0.00000	-	-	0.09348	0.08971	0.08971	0.00000	0.00000
C ₂ H ₆	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₃ H ₈	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	0.00000	-	-	Trace	Trace	Trace	0.00000	0.00000
H ₂ S	-	0.00000	0.00000	-	-	0.00014	0.00013	Trace	0.00000	0.00000
SO ₂	-	0.00000	0.00000	-	-	Trace	Trace	Trace	Trace	Trace
HCl	-	0.00000	0.00000	-	-	0.00000	0.00000	0.00000	0.00000	0.00000
Total	-	1	1	-	-	0.99955	0.99977	0.99977	0.99999	0.99999
Stream Number	Unit	10	11	12	13	14	15	16	17	18
Temperature	°C	49	165	204	40	600	15	934	90	40
Pressure	bar	1.01	5.95	0.89	151.65	17.30	1.01	14.70	1.02	2.00
Molar Flow Rate	kmol/h	8,660.7	9,912.1	4,832.1	1,423.2	7,055.9	2,041.9	2,107.8	2,195.3	1,501.3
Mass Flow Rate	kg/h	244,775	146,917	123,501	61,428	59,237	58,923	63,401	65,916	27,046
Mass Vapor Fraction	-	1	1	1	1	1	1	1	1	0
Composition	mole-basis									
O ₂	-	0.05839	0.00000	0.00000	0.00000	0.00000	0.20741	0.02435	0.03072	Trace
N ₂	-	0.89429	0.00857	0.00011	0.00037	0.01196	0.77315	0.71553	0.71929	Trace
AR	-	0.01068	0.00000	0.00000	0.00000	0.00000	0.00924	0.00855	0.00859	Trace
H ₂	-	0.00000	0.37315	0.00570	0.01928	0.62281	0.00000	0.00000	0.00000	0.00000
CO	-	0.00000	0.07706	0.00005	0.00017	0.00565	0.00000	0.00000	0.00000	0.00000
CO ₂	-	0.00982	0.07498	0.29365	0.97903	0.00799	0.00033	0.16820	0.16151	Trace
H ₂ O	-	0.02681	0.45196	0.70030	0.00053	0.33166	0.00987	0.08293	0.07980	0.99999
CH ₄	-	0.00000	0.01415	0.00018	0.00061	0.01973	0.00000	0.00000	0.00000	0.00000
C ₂ H ₆	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₃ H ₈	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
C ₄ H ₁₀	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
H ₂ S	-	0.00000	Trace	Trace	Trace	Trace	0.00000	0.00000	0.00000	0.00000
SO ₂	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCl	-	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Total	-	0.99999	0.99986	1	0.99999	0.99980	1	0.99957	0.99992	0.99999

Thermodynamic Performance. The net power output of the nominal 100 MW biomass gasification hybrid plant with LiBr absorber for water recovery is 101,896 kW. The power generation of the SOFC (128,138 kW, FU 35.4%) is very similar to the reference plant. Similarly, the GT that is providing the air for the gasifier has an identical power output as in the reference case with 3,174 kW. In this biomass scenario the TIT of the hybrid GT drops to 756 °C, which results in too little power output from the GT expander to power the GT compressor. Thus, rather a motor is needed to drive the compressor than a generator that produces electricity.

The auxiliary load accounts for 29,117 kW. The majority of the auxiliary loads remain relatively constant and the water recovery together with the water treatment add about 786 kW to the plant's auxiliary load. The efficiency of the 100 MW biomass gasification case with LiBr absorber for water recovery is 40.89%-HHV and is about 1.10% points lower than the reference case. This efficiency decrease is similar to the decrease seen for the coal gasification case and is largely related to the heat integration, which can be further improved to achieve higher TITs. The SOFC stack efficiency stays at 79.7%-LHV, while the GT operates in a regime where it does not produce any power. The surge margin in this case is 29.1%.

The plant is equipped with dry cooling technology and as previously mentioned, does not produce any liquid waste streams. Solid waste streams are associated with the gasifier operation and contain ash as well as unconverted feedstock (disposal of catalysts, etc. are handled separately). By using biomass as feedstock and incorporating a CO₂ removal technology, the plant produces negative CO₂ emissions, meaning it actively removes CO₂ from the atmosphere. With this dual-bed gasifier design a carbon capture rate of 75.62% is achieved, which leads to a CO₂ removal rate of 61,323 kg/h (-602 kg/MWh).

At the scale of this nominal 100 MW plant, a total of 649,087 kg of water can be recovered per day. Based on 2015 data, the average farmer in the U.S. used 1.74 l/m² of irrigation water per day [341]. Thus, water from this 100 MW SOFC-GT hybrid plant with LiBr water recovery system can produce enough water for the irrigation of over 317,000 m² of farmland (including 85% capacity factor). The thermodynamic performance results are summarized in **Table 5-99**.

Table 5-99: Thermodynamic Performance of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Gross Power Generation	Unit	LiBr Absorber
Fuel Cell	kW _{DC}	128,137.61
Gas Turbine (SOFC-GT Hybrid)	kW	-299.13
Gas Turbine (Gasifier)	kW	3,174.27
Total Generation	kW	131,012.74
Auxiliary Load		
DC-AC Inverter Losses	kW	2,562.75
Feedstock Handling, Milling and Drying	kW	2,532.46
Gasifier Feeding System	kW	644.34
Gasifier Ash Handling	kW	18.69
Syngas Cleaning and Scrubbing	kW	164.38
CO ₂ Capture and Compression	kW	9,655.83
Syngas Recycle Compressors	kW	12,053.28
Boiler and Feed Water System	kW	122.06
Water Recovery System	kW	781.13
Water Treatment	kW	4.63
Miscellaneous Balance of Plant	kW	85.01
Transformer Losses	kW	492.65
Total Auxiliary Load	kW	29,117.22
Net Power Generation	kW	101,895.52
Plant Performance		
Net Efficiency	%-HHV	40.89
Net Efficiency	%-LHV	43.96
Consumables		
Feedstock	kg/h	53,906
Thermal Input	MJ _{HHV} /h	896,992
Water	kg/h	0
Emissions and Waste Streams		
Carbon Capture	%	75.62
Carbon Emission	kg/h	-61,323
Carbon Emission	kg/MWh	-602
Liquid Waste	kg/h	0
Solid Waste	kg/h	710
Water Recovery		
Flue Gas Water Recovery	%	86.6
Flue Gas Water Recovery	kg/d	649,087

Economics. The TPC of the 100 MW biomass gasification hybrid with carbon capture and with LiBr absorber for water recovery is \$773,378,980 and the specific plant cost is 7,590 \$/kW. The gasification and feed preparation section of the plant contributes the largest amount to the TPC with \$346,292,830. The SOFC island accounts for \$165,070,290. Due to the use of the same GTs as in the reference case, the GT island cost remains constant at \$32,859,130. Also, the other plant equipment exhibits only small changes due to plant scaling. The plant is scaled to the size of the GT (an off-the-shelf item) and due to the increased GT backpressure associated with the water recovery a small change in the plant scale is seen. The plant's TPC increase seen in this case is mostly related to the equipment cost of the water recovery and water treatment, which account for \$8,318,400. An illustration of the TPC breakdown into the major categories is shown in **Figure 5-144**. A more detailed overview of the capital cost is provided in **Table 5-100**.

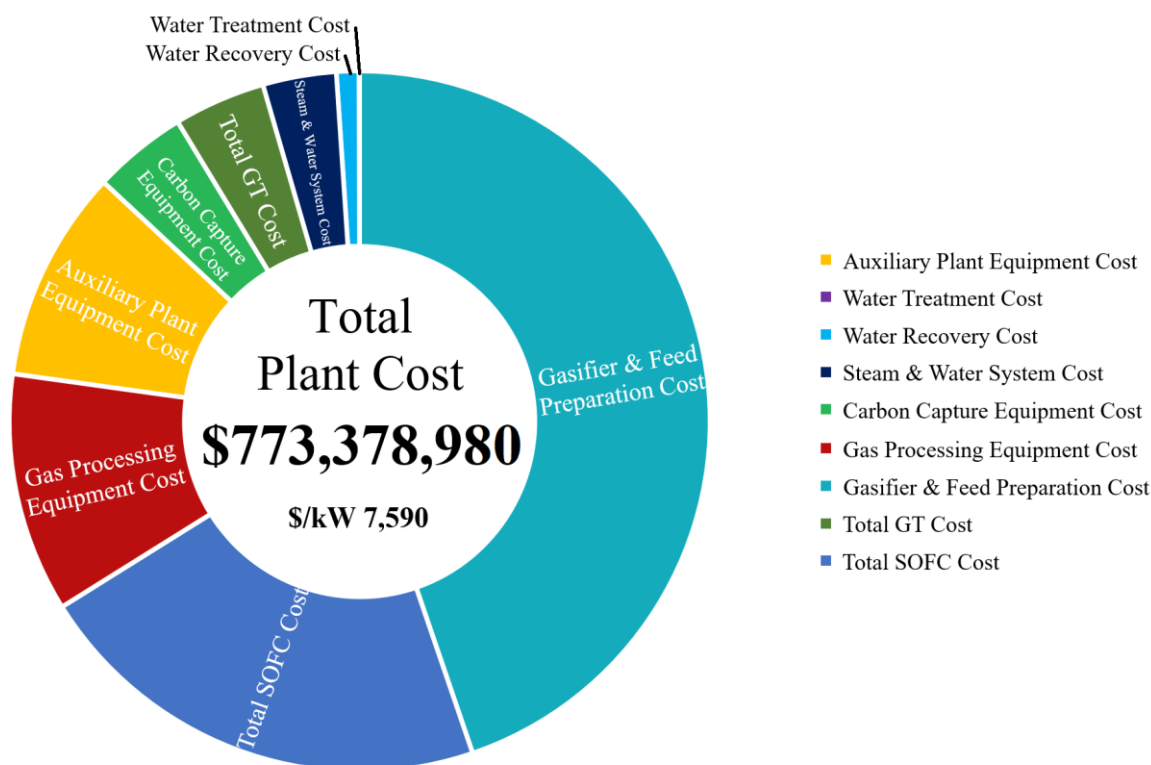


Figure 5-144: Total plant cost and specific plant cost of the 100 MW biomass gasification SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

The total overnight capital cost is \$966,499,720 and is comprised of the TPC, the preproduction costs with \$27,717,150, the inventory capital costs with \$14,301,690 and other costs including financing and land costs with \$151,101,900. Biomass feedstock costs are substantially higher than coal and the annual feed costs account for \$49,794,150. Other variable operating costs include maintenance material costs, equipment replacement costs, catalyst and sorbent costs, as well as waste disposal costs, and amount to \$20,345,750 per year. Annual CO₂ transporting, storing, and monitoring costs \$13,354,340. The sulfuric acid sales revenue generates an annual income of \$303,460, which is equivalent to a COE reduction of 0.40 \$/MWh and the byproduct revenue of the recovered water is \$6,920,200 (-9.12 \$/MWh). Fixed costs are comprised of annual labor costs with \$18,068,890, which includes the additional operating labor for the water recovery, and costs associated with tax and insurance with \$15,467,580 per year. The COW is determined by calculating the sales price necessary in order to achieve the same COE as in the 100 MW biomass gasification reference case. Thus, the COE remains constant at 246.13 \$/MWh. The COW in the LiBr absorber case is 0.0343 \$/L. A summary of the economic key parameters can be found in **Table 5-101**. A detailed breakdown of the COE is shown in **Figure 5-145**.

Table 5-100: Equipment Cost of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Equipment Cost	Unit	LiBr Absorber
Fuel Cell Island	\$	165,070,290
SOFC Stacks	\$	55,776,910
Power Conditioning	\$	74,610,260
Pressure Vessel, Housing & Assembly	\$	30,075,340
System Installation	\$	4,607,780
Gas Turbine Island	\$	32,859,130
Gas Turbine-Generator Set	\$	27,698,410
Gas Turbine Foundation	\$	402,640
Piping, Instrumentation & Electrical Equipment	\$	4,758,080
Gasifier and Feed Preparation	\$	346,292,830
Feed Preparation	\$	74,001,170
Gasifier System	\$	266,718,630
Air Supply and Flue Gas Expansion	\$	5,573,030
Gas Processing and Flue Gas Cleanup	\$	85,396,200
Syngas Cleanup	\$	38,378,730
Flue Gas Cleanup	\$	47,017,460
Carbon Capture	\$	33,478,090
Combined WGS & Carbon Capture PSA	\$	24,280,510
CO ₂ Compression and Dehydration	\$	9,197,580
Steam and Water System	\$	26,009,540
Water Recovery	\$	8,318,400
Water Recovery System	\$	8,118,910
Water Treatment	\$	199,490
Auxiliary Plant Equipment Cost	\$	75,954,510
Accessory Electric Plant	\$	12,576,420
Instrumentation & Controls	\$	28,400,040
Improvement to Site	\$	7,384,010
Buildings & Structures	\$	27,594,040
Total Plant Cost	\$	773,378,980
Specific Plant Cost	\$/kW	7,590

Figure 5-145 shows the COE value of 246.13 \$/MWh in solid colors, the byproduct revenue is shown with striped background (-9.52 \$/MWh) and the whole circle represents the COE without byproduct credit (255.65 \$/MWh). This means, if the water would be provided to the local community at no cost, the COE would increase to 255.25 \$/MWh due to the additional capital and operational costs of the water recovery units. If the electricity is sold to the grid for the same price as in the reference case without water recovery, the water needs to be sold at 0.0343 \$/L in order to reach the COE of 246.13 \$/MWh.

Table 5-101: Economics of the 100 MW Biomass Gasification SOFC-GT Hybrid with Carbon Capture and with LiBr Absorption System for Water Recovery

Item	Unit	LiBr Absorber
Plant Capacity Factor	%	85
Total Overnight Capital Cost		
Total Plant Cost	\$	773,378,980
Preproduction Costs		
6 Months Labor Cost	\$	9,034,440
1 Month Maintenance Materials	\$	1,307,160
1 Month Non-Fuel Consumables	\$	627,580
1 Month Waste Disposal	\$	59,940
25% of 1 month fuel cost at 100% CF	\$	1,220,440
2% of TPC	\$	15,467,580
Inventory Capital Costs		
60 Day Supply Consumables at 100% CF	\$	10,434,790
0.5% of TPC (Spare Parts)	\$	3,866,890
Other Costs		
Initial Catalyst, Sorbent & Chemicals Cost	\$	8,765,140
Land	\$	5,448,680
Other Owner's Costs (15% of TPC)	\$	116,006,850
Financing Costs	\$	20,881,230
Variable Operating Cost for Initial Year of Operation		
Annual Feed Cost at above CF for Initial Year	\$	49,794,150
Other Annual Variable Operating Cost at above CF for Initial Year	\$	20,345,750
Annual CO ₂ Transporting, Storing, and Monitoring Cost at above CF for Initial Year	\$	13,354,340
Annual Byproduct Revenues at above CF for Initial Year	\$	7,191,890
Fixed Operating Cost for Initial Year of Operation		
Labor Cost	\$	18,068,890
Property Tax & Insurance	\$	15,467,580
Cost of Electricity and Cost of Water		
First Year Levelized Cost of Electricity w/o CO ₂ TS&M	\$/MWh	228.53
First Year Levelized Cost of Electricity with CO ₂ TS&M	\$/MWh	246.13
First Year Levelized Cost of Water	\$/L	0.0343

Figure 5-146 shows a detailed breakdown of the COW. Overall, the COW in the biomass gasification scenario is reduced over the coal gasification scenario. This reduction is mostly due to reduced operating costs as well as the lower TPC (on a relative basis), which benefits from the high water recovery in this scenario. The largest COW cost driving factor is the performance penalty cost; in particular the system integration losses, which are even more dominant in the biomass gasification scenario than in the coal gasification scenario. This shows that it is necessary to improve the heat integration and the GT's TIT in the gasification cases with LiBr water recovery together with the steam injection in order to operate the GT (in particular the GT expander) in a region of higher efficiency.

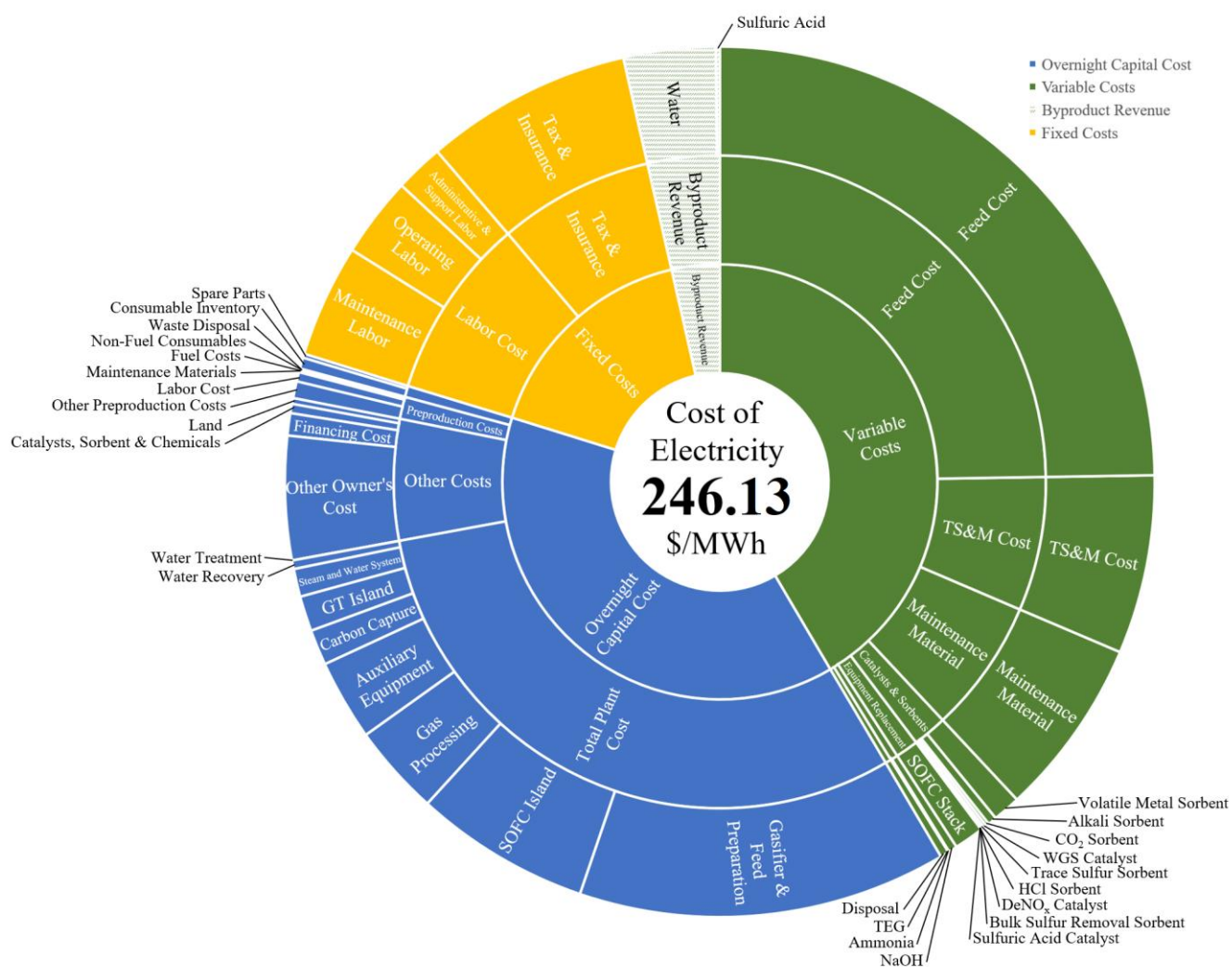


Figure 5-145: Breakdown of the cost-of-electricity of the 100 MW biomass gasification SOFC-GT hybrid with carbon capture and with LiBr absorption system for water recovery.

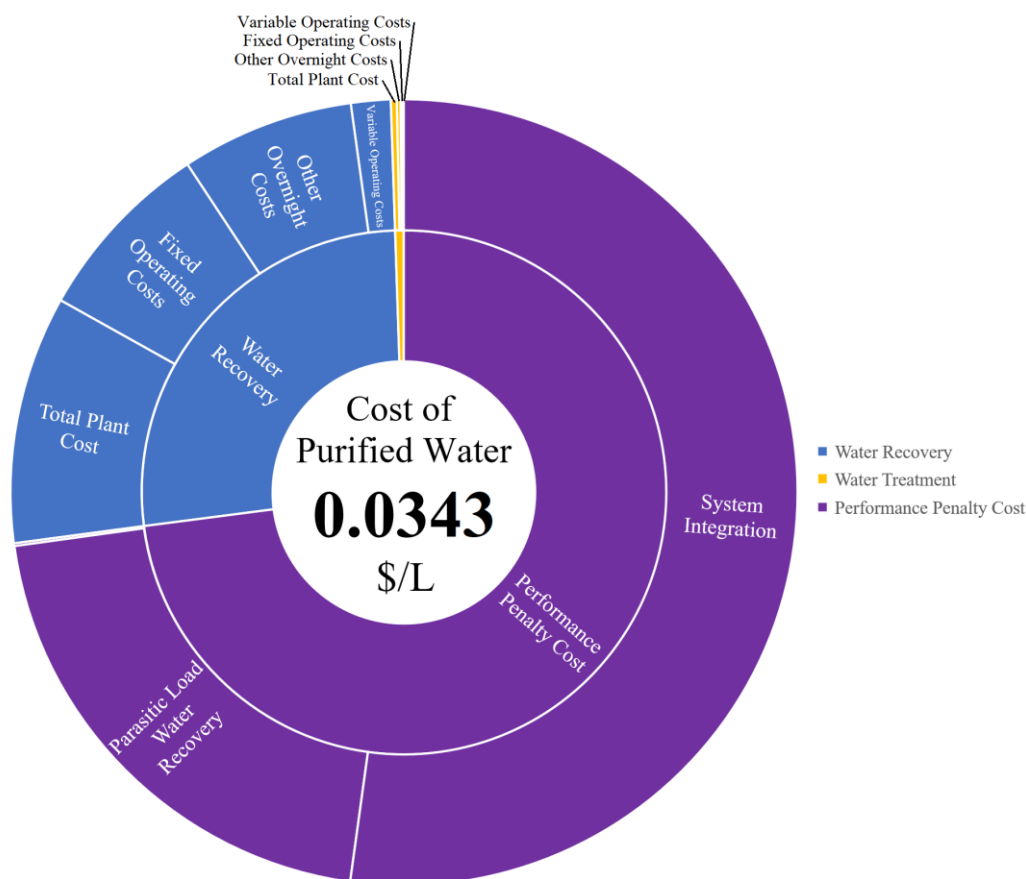


Figure 5-146: Breakdown of the cost-of-water of the 100 MW biomass gasification SOFC-GT hybrid with carbon capture and with LiBr absorber for water recovery.

Summary of the 100 MW Gasification SOFC-GT Hybrids

In total six different 100 MW SOFC-GT hybrid plants with carbon capture, with and without water recovery, have been studied using NG, subbituminous coal and biomass as feedstocks. The highest efficiencies are reached for the NG cases as the feedstock does not have to be dried, milled and/or converted into a gaseous fuel, which is associated with power consumption and conversion losses. The biomass gasification cases have the lowest efficiency as more energy input is required for biomass drying and milling as for coal. Additionally, the syngas recycle in the biogas scenario is increased leading to an even higher auxiliary load. Biomass with its low heating value produces a low heating value syngas, which ultimately leads to a lower firing temperature in the GTs reducing their power output. All these factors contribute to the efficiency trend seen in **Figure 5-147**. As previously discussed, the LiBr water recovery system leads to an efficiency reduction ranging from 0.56-1.10% points. The efficiency drops in the gasification are more severe than in the NG case and related to the GT's steam injection and heat integration, which still offers space for improvement.

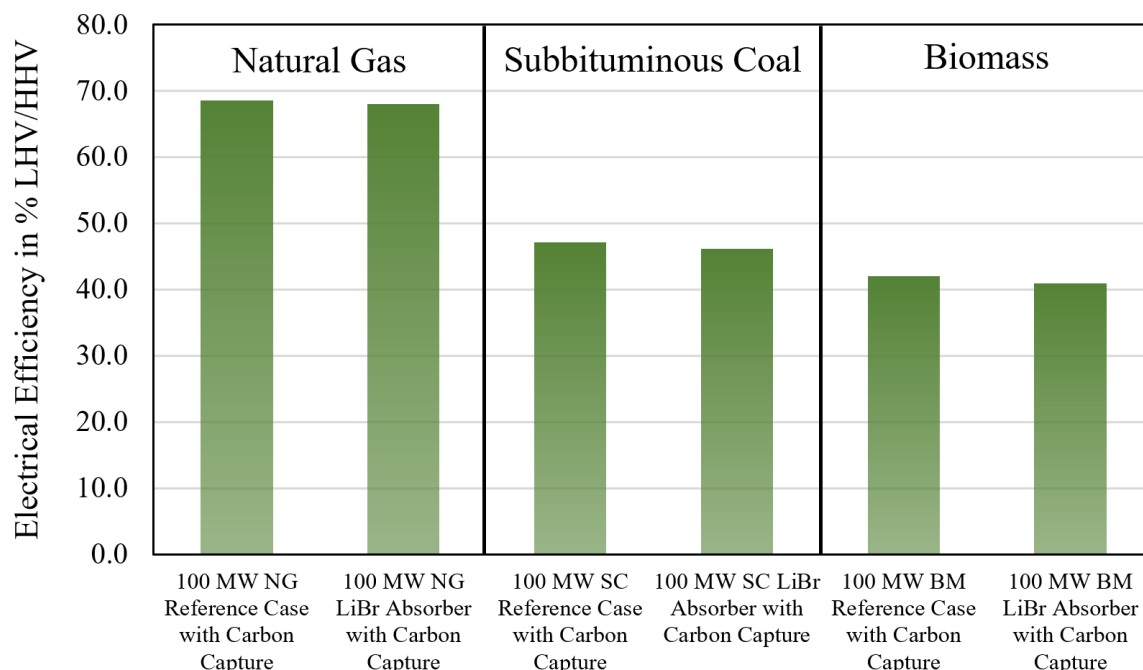


Figure 5-147: Comparison of the electrical efficiencies of the 100 MW SOFC-GT hybrids with carbon capture, and with and without LiBr water recovery.

The COEs are established in the reference cases without water recovery. The lowest COEs are reached in the NG cases, which benefit from the relatively low capital investment needed to build the plant as well as the exceptionally high efficiency, which helps reduce fuel costs. In the coal gasification cases, the fuel costs are the lowest among the studied scenarios; however, the substantial investment needed to gasify the coal and clean the syngas more than offsets these gains. The biomass gasification plant suffers from high investment costs as well as from high fuel costs and has the highest COE. Yet, it may be pointed out that this biomass gasification plant is a negative emission plant that actively removes CO₂ from the atmosphere.

The respective COEs are held constant in the corresponding water recovery scenarios and the water sales price is adjusted in order to obtain the same COE as in the reference case. Three main factors contribute to the COW: the capital cost, the operating cost and the amount of water recovered. The lowest COW is obtained in the NG case due to a better system integration. A summary of the COEs and COWs for the scenarios is presented in **Figure 5-148**.

Due to the economy of scales, larger amounts of water recovery are beneficial for the COW as it scales down investment and operating costs. While coal and biomass carry considerable amounts of surface and inherent moisture, NG exhibits an extremely high H:C ratio. After fuel oxidation any fuel-bound hydrogen will be converted to water and can be recovered from the flue gas. The lowest water recovery is obtained in the coal gasification scenario as coal contains only small amounts of fuel-bound hydrogen. Furthermore, the coals relatively high ash content reduces water exports as water is needed for ash treatment. Natural gas

with its high H:C ratio reaches a substantially higher water recovery only surpassed by the biomass scenario, as biomass possess considerable amounts of fuel moisture, low amounts of ash and a considerable amount of fuel-bound hydrogen. The amount of net water export for each plant is shown in **Figure 5-149**.

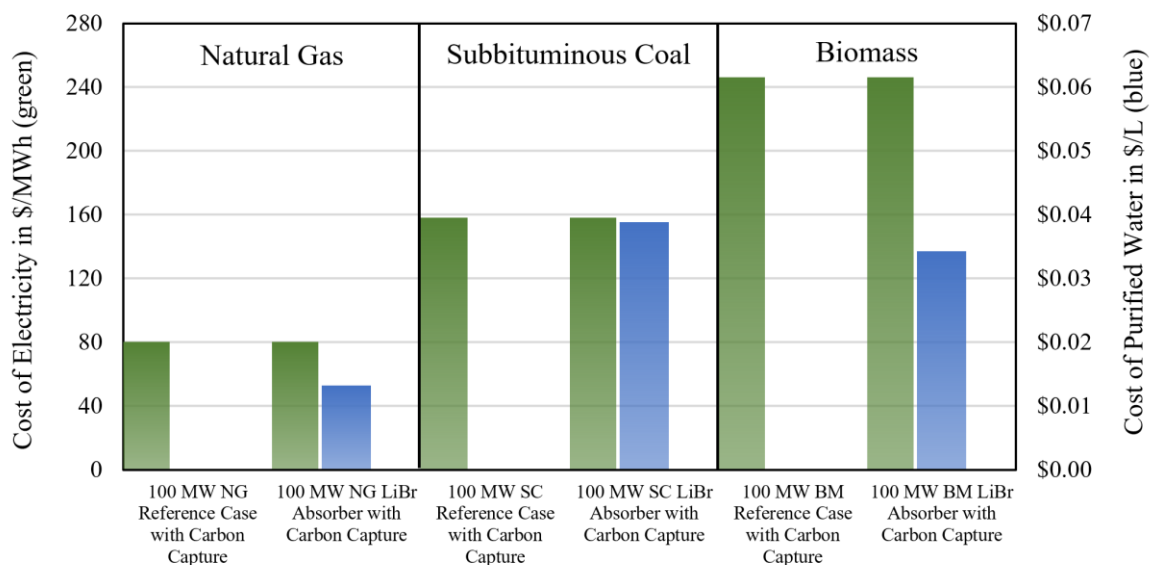


Figure 5-148: Comparison of the cost-of-electricity and cost-of-water of the 100 MW SOFC-GT hybrids with carbon capture, and with and without LiBr water recovery.

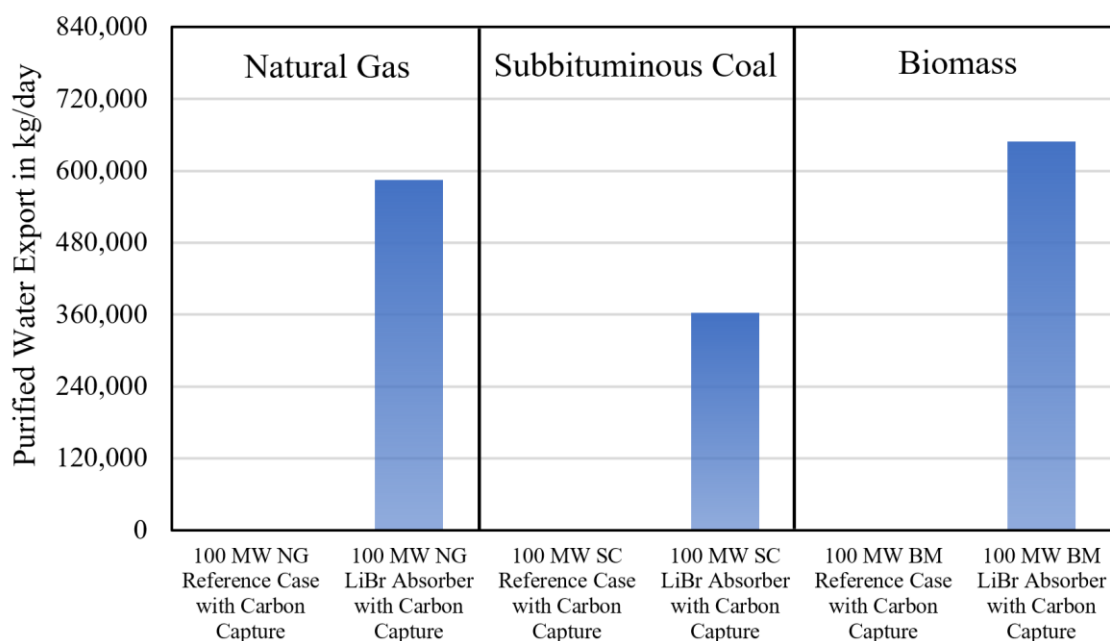


Figure 5-149: Comparison of the water recovery rate of the 100 MW SOFC-GT hybrids with carbon capture, and with and without LiBr water recovery.

Chapter 6

Summary, Conclusions and Recommendations

6.1 Summary

A bottom-up cost approach for the SOFCs has been developed to derive the SOFC stack cost based upon individual cell components, such as electrolyte, anode, cathode, interconnect, frames, seals, and end plates. In combination with a first-principle quasi-2D SOFC model, the cell design has been optimized with respect to thermal SOFC cell gradients and specific stack cost on a \$/kW basis, considering a wide range of geometrical cell design parameters as well as operating parameters. The results provide novel insights into the design of SOFC cells and identified synergistic effects between cell economics and thermal management.

Using this optimized cell design, the SOFC stack has been integrated into a SOFC-GT hybrid system with custom engineered GT to evaluate the full potential of this hybrid technology. The SOFC-GT hybrid system has been investigated at steady state operating condition over a wide range of pressure conditions, fuel utilization factors and operating voltages under consideration of local thermal cell gradients and overall cell temperature differences inside the SOFC. The steady state simulation results served as inputs for the thermo-economic analysis, which provides key economic parameters including the specific plant cost and cost of electricity.

As of today, GTs are primarily offered for high energy density fuels like natural gas and will have to be operated in off-design when integrated into a SOFC-GT hybrid system. This raises concerns about the GT's surge margin. A comparison of constant speed operation, variable speed operation and variable inlet guide vane control (for full load operation), identified feasible operating modes to operate the GT safely below the surge limit and modes that allow for an economical operation using an off-the-shelf GT.

In total, five different water recovery technologies, an air-cooled condenser, a direct contact condenser, a LiBr absorption system, a MEG absorption system and a transport membrane condenser, have been integrated into the optimized SOFC-GT hybrid power plant with off-the-shelf GT and compared to a reference case without water recovery. This analysis has been conducted for natural gas as fuel as well as for biogas, after making the necessary plant design modifications. The results provide insights in how the

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pressure drop of the water recovery unit and the parasitic load impact the plant operation, efficiency, and economics. A detailed analysis of cost driving factors provides a comprehensive understanding of system integration costs, operating costs and capital related expenses to guide future research in the area of flue gas water recovery.

Furthermore, the economies of scale have been studied for the natural gas hybrid systems with and without LiBr absorption system for flue gas water recovery. This study includes the investigation of carbon capture from the anode off-gas as well as the investigation of synergetic effects between carbon capture and flue gas water recovery. Lastly, a coal gasification plant with carbon capture and SOFC-GT hybrid cycle and a negative emission biomass gasification plant with SOFC-GT hybrid system have been investigated with respect to their thermo-economic performance, emissions, and water recovery potential. More detailed summaries about the above-mentioned topics can be found at the end of the respective sections in the main body of this dissertation.

6.2 Conclusions

A SOFC-GT hybrid system has been optimized and studied in the context of various fuels, scales and water recovery technologies. The thermo-economic analyses of these systems revealed the following conclusions:

SOFC CELL DESIGN OPTIMIZATION CAN SIMULTANEOUSLY REDUCE SPECIFIC SOFC COSTS AND THERMAL GRADIENTS.

The cell optimization study showed that thermal gradients inside the SOFC and the specific SOFC stack cost on a \$/kW basis can be effectively reduced by increasing the rib width as well as by adjusting the interconnect thickness or cathode channel height. In the studied scenario, an interconnect thickness of 1 mm has shown to offer the most cost-effective solution. On the cathode side, a channel height of 2 mm has shown to have the greatest cost advantage. In particular, improving the heat transport inside the SOFC from the end of the cell to the front part, where the endothermic reforming reactions take place, helps to accelerate reaction kinetics at the cell inlet and smoothen thermal gradients.

IN A SOFC-GT HYBRID, THE HIGHEST POWER OUTPUT AND THE HIGHEST EFFICIENCY ARE OBTAINED WHEN USING ONLY THE GT EXHAUST TO COMPLETELY PRE-HEAT THE CATHODE AIR.

Using only the GT exhaust to pre-heat the cathode air requires a high GT exhaust temperature. This is achieved at high TITs (low fuel utilization) and/or low pressure ratios. Since the SOFC operates at a significantly higher efficiency than the GT, it is desirable to operate the SOFC at relatively high fuel utilization factors, which means that relatively low pressure ratios are desirable. Operating at low pressure ratios results in a low-temperature compressor discharge, which enables more heat to be recuperated from

the exhaust gas increasing the hybrid system's efficiency. At the same time, no additional heat input from the SOFC off-gases or combustor outlet gas are needed maximizing the GT's efficiency.

IN A SOFC-GT HYBRID THE HIGHEST EFFICIENCY SCENARIO DOES NOT REPRESENT THE LOWEST COST-OF-ELECTRICITY SCENARIO.

The specific SOFC cost and the specific GT cost are crucial economic parameters in a SOFC-GT hybrid system. While the specific SOFC cost increases as operating pressure decreases due to the lower current density, this effect is outweighed by the fact that the specific SOFC costs are increased at higher operating voltage and higher fuel utilization factors (reduction of chemical potential difference at the outlet). Furthermore, the specific GT cost increases with increasing pressure ratio and increasing fuel utilization. Consequently, the lowest cost-of-electricity is achieved at lower operating voltages and lower fuel utilization factors compared to the maximum efficiency case.

VARIABLE SPEED OPERATION OF THE GT IN A SOFC-GT HYBRID PROVIDES THE HIGHEST EFFICIENCY, VARIABLE INLET GUIDE VANE CONTROL PROVIDES THE LARGEST SURGE MARGIN AND CONSTANT SPEED OPERATION OF THE GT SHOWS THE GREATEST ECONOMIC BENEFIT.

Variable speed operation moves the GT operating point into a region of high compressor efficiency, which is able to improve the overall system efficiency. However, this moves the operating point extremely close to the GT's surge limit. Using variable inlet guide vane control, the operating point can be steered away from the surge limit. While this substantially reduces the risk of running into compressor surge, closing the inlet guide vanes also reduces the air flow through the compressor, leading to a downscaling of the entire plant, which substantially increases the specific GT cost and thus, the cost of electricity. The constant speed operation offers a compromise between the variable speed operation and variable inlet guide vane operation, leading to an acceptable surge margin. Due to the highest compressor airflow and its relatively high efficiency, the constant spool speed scenario reaches the lowest cost of electricity.

EVEN THOUGH SOFCs ARE HIGHLY MODULAR AND DO NOT SCALE LIKE CONVENTIONAL BALANCE-OF-PLANT EQUIPMENT, SUBSTANTIAL SAVINGS CAN BE EXPECTED WHEN MOVING TO LARGER PLANT SCALES.

The modular design of SOFCs as well as the SOFC-GT hybrid module, as considered in this work, allow for only marginal savings when moving to larger plant scales. However, the supporting balance-of-plant equipment, such as fuel compressor, vessels, heat exchangers, etc., can still benefit from the economies of scale. In particular, the reduced operating labor, when moving to a larger scale, (on a relative basis) leads to a substantial reduction in the cost of electricity. In the case of carbon capture this economic benefit

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becomes even more apparent as the equipment for carbon capture, compression and dehydration profits from the economies of scales and the additional operating labor reduction associated with the carbon separation unit.

AMONG THE STUDIED WATER RECOVERY TECHNOLOGIES, THE LiBr ABSORPTION SYSTEM IS THE MOST ECONOMICAL TECHNOLOGY FOR WATER RECOVERY.

In total five different water recovery technologies, an air-cooled condenser, a direct contact condenser, a LiBr absorption system, a MEG absorption system and a transport membrane condenser, haven been integrated into a natural gas SOFC-GT hybrid plant and into a biogas SOFC-GT hybrid plant with digester. In both plant configurations, the LiBr absorption system exhibits superior economic performance, while recovering the largest amount of water. The air-cooled condenser suffers from its large pressure drop, which increases the GT backpressure and substantially reduces the power output. On the other hand, the direct contact condenser and transport membrane condenser can only recover a small fraction of the flue gas moisture making it expensive on a \$/L basis. The MEG absorption system mostly suffers from desiccant loss during regeneration leading to high operating costs. The LiBr absorption system has shown to be particularly attractive in combination with the digester in the biogas plant, which requires heat input to drive the chemical reactions in the digester. In the LiBr absorption system, the latent heat can be upgraded and made available to the digester at a temperature level above the flue gas's dew point temperature leading to an overall increase of the plant's efficiency.

WHILE MANY CARBON CAPTURE TECHNOLOGIES ARE WATER INTENSIVE, CARBON CAPTURE IS BENEFICIAL FOR FLUE GAS WATER RECOVERY.

Water recovery is mostly driven by concentration gradients. When removing CO₂ from the anode off-gas or flue gas, the partial pressure of H₂O increases and thus, the driving force for water recovery. While the improved kinetics (heat and mass transfer) certainly lead to a reduction in equipment size, a second effect plays an even more important role. The CO₂ removal from the flue gas leads to a smaller exhaust gas volume flow rate, which helps to reduce equipment size (size of absorption columns is highly dependent on gas flow), reduce capital costs and improve economics.

SOFC-GT HYBRID SYSTEMS OFFER UNUSUALLY ATTRACTIVE PERFORMANCE ON CONVENTIONAL FUELS, TO SUPPORT THE CLEAN ENERGY TRANSITIONS DURING THE INITIAL STAGE, AS WELL AS, ON RENEWABLE FUELS THAT WILL FORM THE BACKBONE OF THE CIRCULAR ECONOMY.

To enable the scaleup of SOFC production, SOFCs and SOFC-GT hybrid systems need to be produced in a modular manner. By using a modular SOFC-GT hybrid design (considering thermal cell gradients,

stack design and aerodynamics of an off-the-shelf GT) and integrating it into a natural gas hybrid, a biogas hybrid, a coal gasification hybrid, and a biomass gasification hybrid, it has been shown that despite their modular design, SOFC-GT hybrid systems are highly fuel flexible. This fuel flexibility, together with their scalability, virtually zero emission and ability to be paired with flue gas water recovery, makes SOFC-GT hybrid systems an attractive technology for power generation in a future circular economy. Regardless of the fuel type, SOFC-GT hybrid systems reach exceptional efficiencies, outperforming conventional power generation systems. In particular, high CH_4 -content fuels result in remarkable efficiencies of greater than 75%-LHV due to internal reforming, which upgrades the fuel using waste heat while cooling the SOFC. Hydrogen-rich fuels cannot make use of this effect; however, reaction kinetics improve with the use of hydrogen reducing the specific SOFC cost. In the case of renewable hydrogen use, produced via electrolysis of water, flue gas water recovery will be of major importance in order to close the water cycle.

6.3 Recommendations

While this work substantially contributed to providing new insights into the design and the economics of SOFC-GT hybrid systems, there are numerous questions that will need to be addressed in future research. Today, many commercial SOFC products are available; however, most of these SOFC applications operate near ambient pressure and the large-scale stack pilot testing under pressurized conditions at high air utilization factors is yet to be accomplished. In particular, it needs to be ensured that the large stacks, 250 cells per stack chosen in this work to reduce cost, are thermally balanced. Also, the SOFC pressure vessel design proposed in this work needs to be tested prior commercialization. In this design, the cathode off-gas is used to pressurize the vessel and oxygen concentration changes during operation as well as explosion limits of the vitiated air and fuel (anode inlet and outlet) need to be evaluated. Furthermore, it will be important to understand the depressurization behavior in case the GT trips.

To provide a platform for easier introduction of SOFC-GT hybrid technology into the market, it is important to have highly fuel flexible GTs that operate around 5 bar. Thus, it is recommended to pursue the development of GTs with radial compressors and expanders that allow for air extraction and the utilization of ultra-low-heating-value fuels. This type of GT technology will not only be needed for SOFC-GT hybrids but will offer a platform for a variety of renewable and unconventional energy conversion technologies.

Other components that need to be studied in more detail include the pre-reformer, which is a pipe with catalyst coated walls. While it is desired to only pre-reform a portion of the methane

entering the pre-reformer, it needs to be ensured that higher hydrocarbons have enough time to diffuse to the pipe's wall and react on the catalyst's surface. Secondly, the heat integration (radiation and convective heat transfer) of the pre-reformer into the SOFC furnace requires a more detailed analysis.

The combined WGS-CO₂ PSA carbon capture technology is currently under development. While several test campaigns for proof-of-concept demonstrations have been successfully completed, it needs to be verified that the reactors can operate on the anode off-gas encountered in this application without liquid water injection without exceeding any temperature limits.

The hydrogasifier is the least developed technology used in this study and a significant amount of work is needed to better understand reaction mechanisms and kinetics of the participating gasification reactions including tar formation mechanisms and tar destruction strategies. Furthermore, dual-bed reactor operation and reactor scaleup are areas of future research.

On the LiBr water recovery side, equipment cost reduction is of major interest. Thus, more research on materials and corrosion is needed for both, column and regenerator. Moreover, it needs to be experimentally quantified how much non-condensable gases are released during the desiccant regeneration process. To further reduce the cost of water, automated control strategies are a research area of interest as well as research on the treatment of the recovered water and potential end users.

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Appendix A: Cell Design

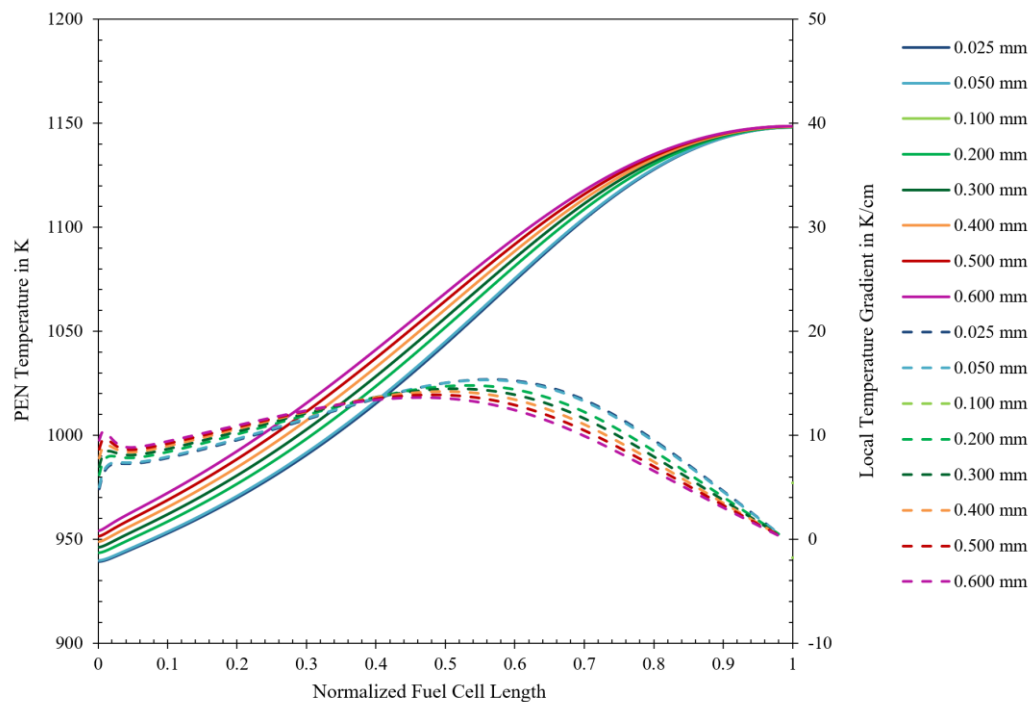


Figure A-1: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various anode thicknesses.

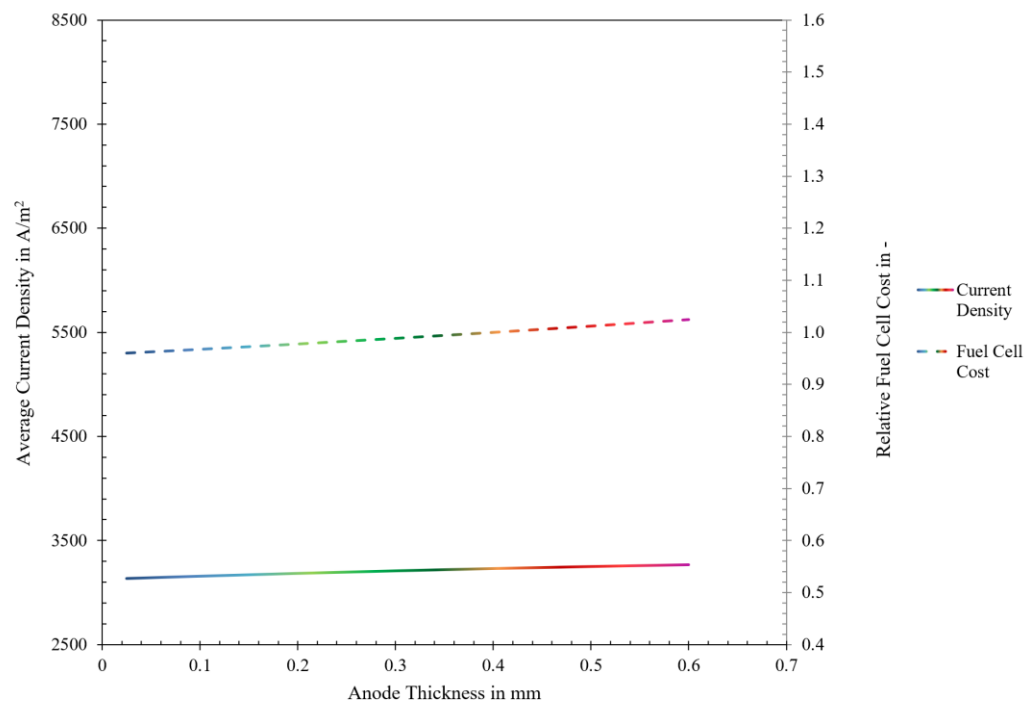


Figure A-2: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of anode thickness.

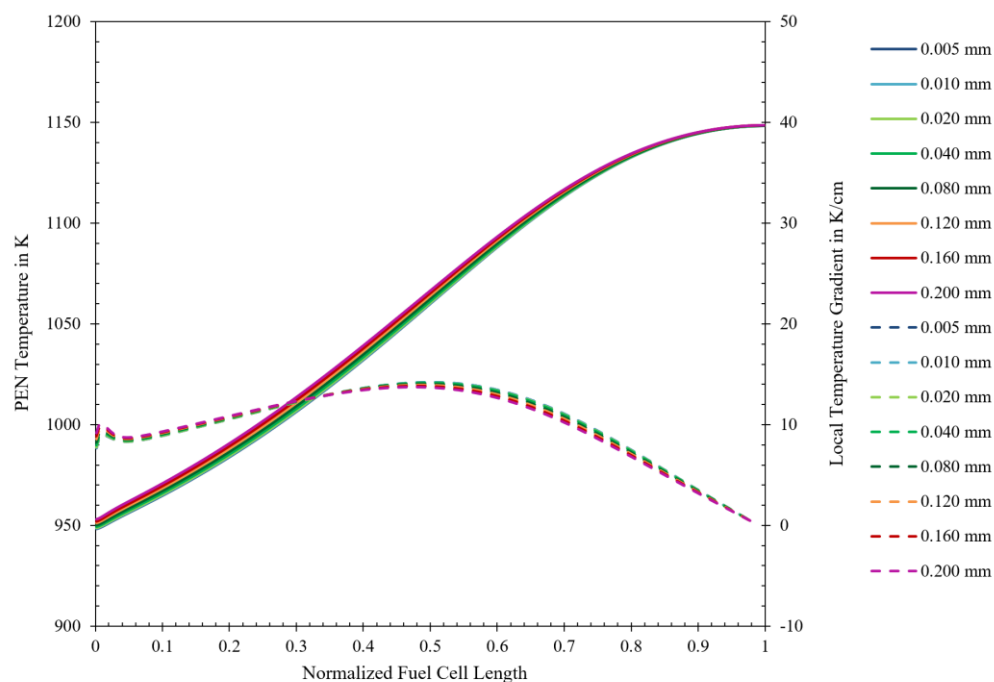


Figure A-3: SOFC PEN temperature profiles (solid lines) and local temperature gradients (dashed lines) along the flow channel for various cathode thicknesses.

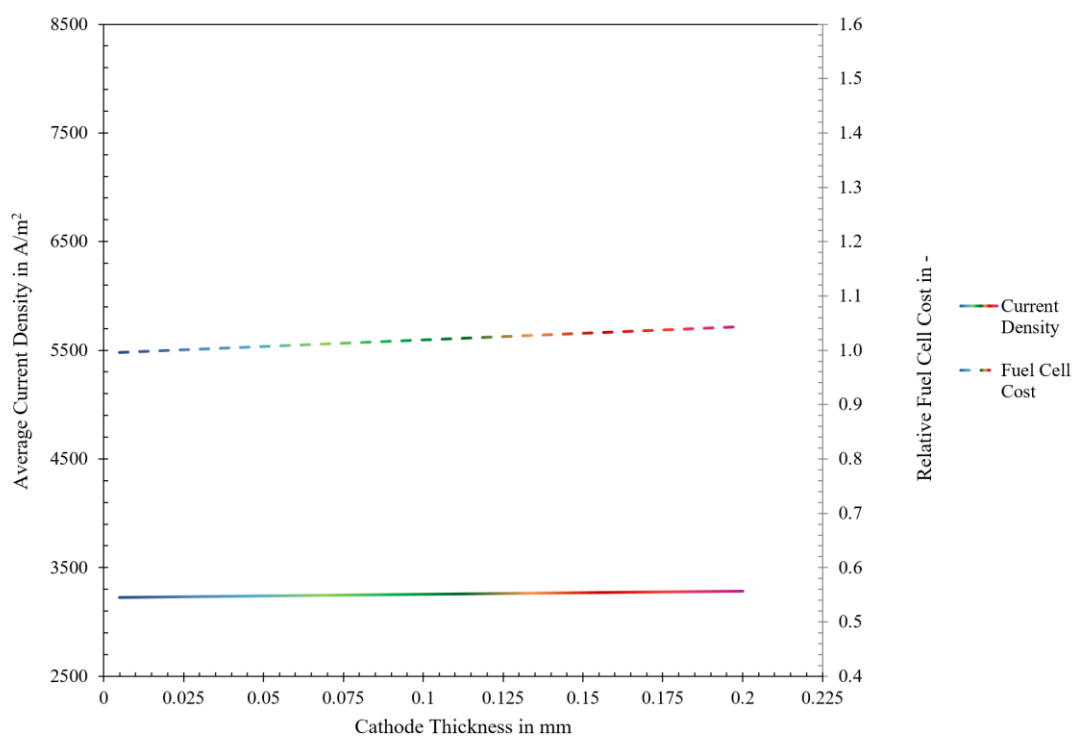


Figure A-4: Cell average current density (solid line) and relative fuel cell cost (dashed line) as functions of cathode thickness.

Appendix B: Flowsheets

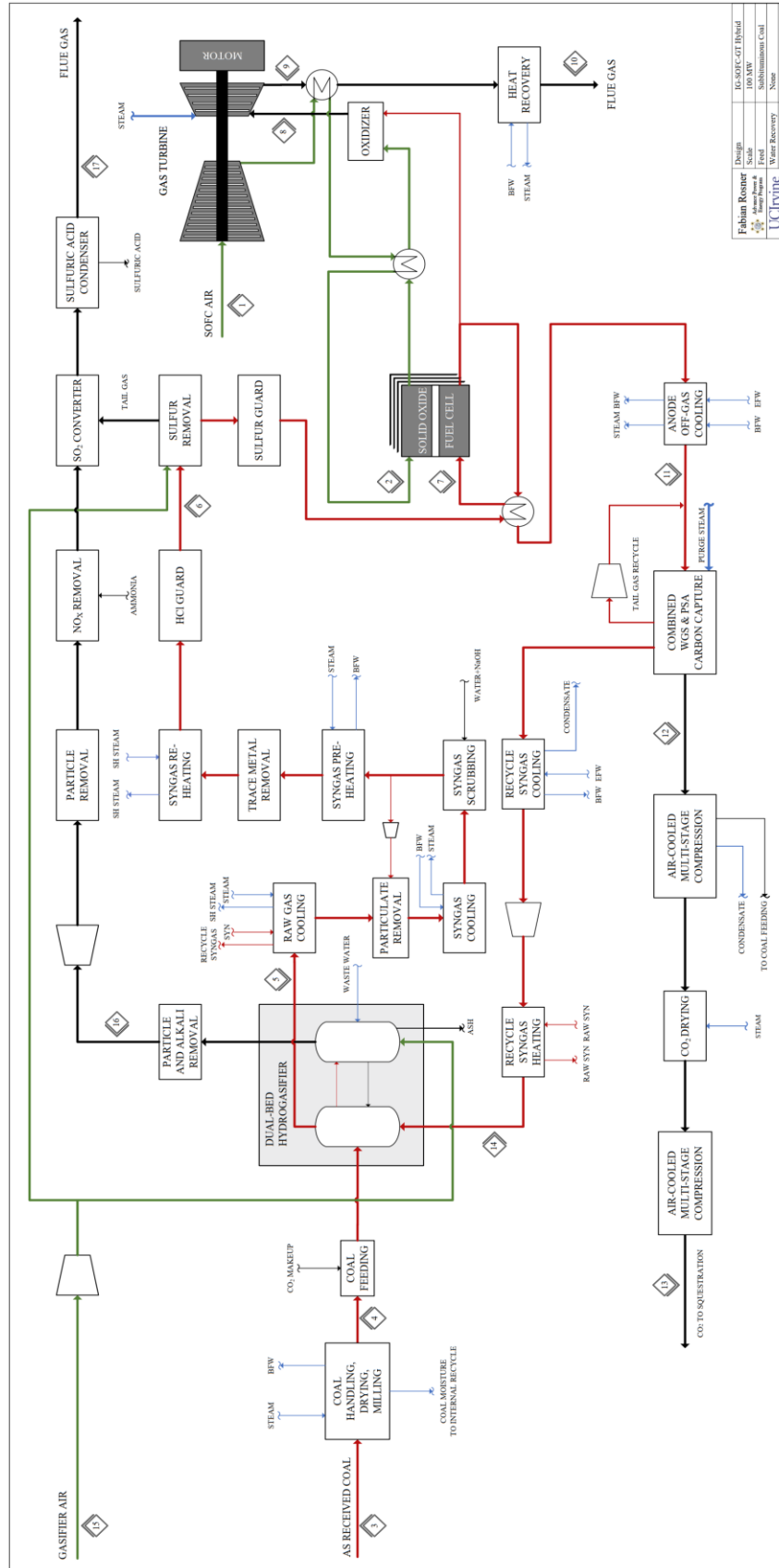


Figure B-1: Flowsheet of the 100 MW coal gasification SOFC-GT hybrid system with carbon capture and without water recovery.



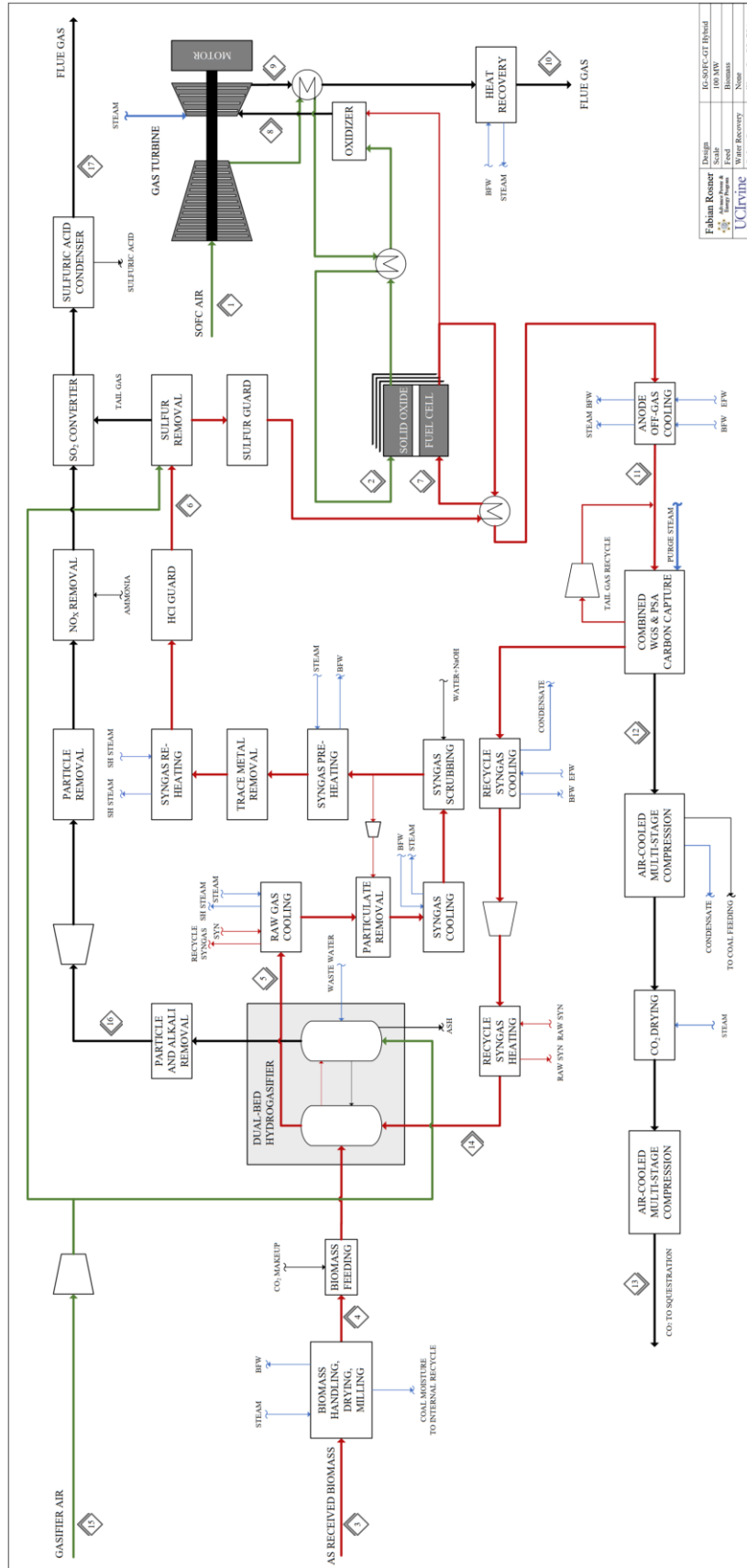


Figure B-3: Flowsheet of the 100 MW biomass gasification SOFC-GT hybrid system with carbon capture and without water recovery.

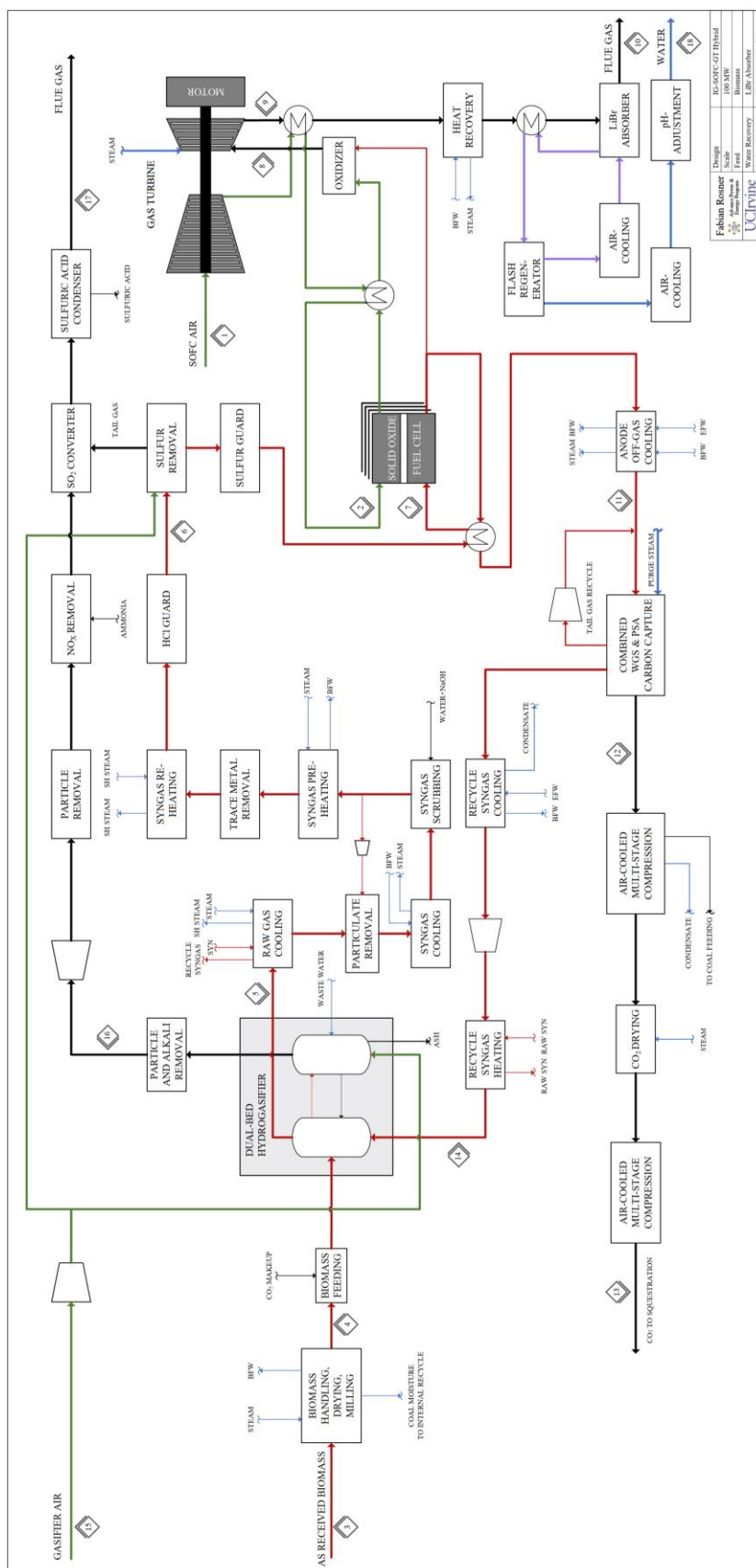


Figure B-4: Flowsheet of the 100 MW biomass gasification SOFC-GT hybrid system with carbon capture and with LiBr absorption system for water recovery.