**Title**
Acid Gas Removal and Dehydration Process Design for Natural Gas Liquids Recovery



**Author**
Gomez, Sonia

**Publication Date**
2019-04-01

# ACID GAS REMOVAL AND DEHYDRATION PROCESS DESIGN FOR NATURAL GAS LIQUIDS RECOVERY

By

Sonia Vanessa Gomez

A capstone project submitted for
Graduation with University Honors

June 6, 2019

University Honors
University of California, Riverside

APPROVED

______________________________________________
Dr. Kawai Tam
Department of Chemical Engineering

______________________________________________
Dr. Richard Cardullo, Howard H Hays Jr. Chair, University Honors

## Abstract

Raw natural gas feeds are commonly purified in the chemical industry into natural gas products for industrial and commercial consumption. The petroleum engineering company, ASABA Technology whose headquarters is the University of California, Riverside (UCR), aimed to design a full-scale natural gas liquid (NGL) recovery plant capable of processing raw natural gas feeds and treating them to the required gas-line specifications set by the state of Texas in the United States of America, where the chemical plant was to be constructed. ASABA Technology focused on three major sub-processes in the overall NGL recovery process: water and acid gas removal (sweetening), use of a demethanizer, and a fractionation train. Herein, the acid gas removal and dehydration process design methods and results were outlined in addition to their relevant operational economics for the chemical plant. The sweetening and dehydration processed were designed on Aspen HYSYS to remove impurities in the raw natural gas feed; particularly, carbon dioxide (CO2), hydrogen sulfide (H2S), methyl mercaptan/methanethiol (CH3SH), and water (H2O). Acid gas removal was achieved by monoethanolamine (MEA) gas scrubbing and the dehydration processes employed triethylene glycol (TEG) due to their efficiency and economical viability.

**Acknowledgments**

Kawai Tam, Ph.D.

Richard Cardullo, Ph.D.

Dhananjay Kumar, M.S.

ASABA Technology:

- Abdullah Abdussamy
- Angelica Franco
- Alexander Shen
- Brian Sullivan

Dennis McIver, M.S.

Kristine Parada, M.S.

Mariah Gomez

**Table of Contents**


**Introduction** **3**

**Methods** **4**

**Design Software** **4**

Acid Gas Removal Process 4

Dehydration Process 7

**Results** **9**

MEA Gas Sweetening Process 9

Flowsheet Calculations 16

Economics 17

**Discussion** **22**

Design Software 22

Acid Gas Removal Process 22

Dehydration Process 23

**Conclusion** **24**

**Table of Nomenclature** **24**

**References** **25**

## Introduction

Natural gas liquids (NGLs) are hydrocarbons such as ethane, propane, butane, isobutane, and pentane that are condensable from gaseous to liquid state and can be typically found in shale basins scattered throughout North America. NGLs are primarily used for petrochemical feedstocks in the chemical industry because NGLs offer a variety of applications from plastic synthesis, chemical manufacturing, and alternative energy sources to coal and crude oil-based power.[3] In order to yield highly purified natural gas products which could be used for industrial and commercial applications, the raw natural gas feedstocks must undergo several steps of processing. Generally, the most common processing steps employed in the chemical industry are acid-gas removal, dehydration, nitrogen rejection, and finally fractionation; however, many companies have slightly different methods in an attempt to optimize their natural gas processing.[4]

Raw feedstocks of NGLs must be processed to remove impurities such as carbon dioxide ($CO_2$), hydrogen sulfide ($H_2S$), nitrogen ($N_2$), methyl mercaptan/methanethiol ($CH_3SH$), and water ($H_2O$). $CO_2$, $H_2O$, and $H_2S$ are of concern because they can corrode the pipelines of a chemical plant.[3] $CO_2$ and $N_2$ are known to decrease the overall heating value of the NGLs which directly damages the product's quality.[5] $H_2O$ can react with NGLs (most commonly with methane) in the pipelines to yield hydrates which are crystalline molecules that often resemble ice and pose problems for efficient NGL processing. High operating pressures, low temperatures, and the presence of $CO_2$ and $H_2S$ promote the formation of hydrates. Additionally, $H_2S$ is toxic at high concentrations and is an environmental pollutant.

The natural gas impurities previously outlined have economic value and could be further processed to be sold. For example, $H_2S$ could be decomposed into elemental sulfur which could be sold to battery manufacturers and hydrogen gas which is widely consumed. If highly purified, $CO_2$ could be sold in the food industry for the carbonation of beverages. The collected $H_2O$ could be recycled by the plant owners or sold to fracking companies who use a high-pressure water mixture in their drilling sites to releases gases from inside rocks. $N_2$ can be sold as a compressed gas (liquid nitrogen) or as an inert gas.

The specific goals of this project were to to design working models for the acid gas removal and dehydration processes simulated on the chemical process simulation software, ASPEN HYSYS. Additionally, the operational economics for the monoethanolamine (MEA) gas scrubbing and the dehydration with triethylene glycol (TEG) were outlined and assessed.

## Methods

### Design Software

The acid gas removal and dehydration process components for the overall natural gas liquids recovery plant were designed on the chemical process simulator, Aspen HYSYS.

## Acid Gas Removal Process

Monoethanolamine (MEA) was employed for gas sweetening (acid gas removal) because it selectively and almost entirely removes the acid gases specified in ASABA technology's raw natural gas feed which are hydrogen sulfide, carbon dioxide, carbonyl sulfide, and methylmercaptan. The primary amine portion of the MEA molecule is a weak base which selectively deprotonates weak acid gases dissolved in the solvent so as to yield their conjugate anions as depicted in reaction scheme I for the removal of hydrogen sulfide gas along with a qualitative estimation of the reaction velocities.

*Reaction Scheme I* [1]

*Step 1 (fast):* $HOCH_2CH_2NH_2 + H_2S \rightleftharpoons HOCH_2CH_2NH_3^+ + SH^-$

*Step 2 (fast):* $HOCH_2CH_2NH_2 + SH^- \rightleftharpoons HOCH_2CH_2NH_3^+ + S^{2-}$

MEA reacts with carbon dioxide in the presence of water as outlined in reaction scheme II because acid gas removal precedes the dehydration process. The weak base MEA first reacts with the carbonyl (carbon double bonded to oxygen) to create a nitrogen - carbon bond. A second MEA molecule then deprotonates the first MEA molecule which bonded to the carbonyl to yield stable ammonium and carboxylate salts. Alternatively, the carbon dioxide could react with water and the MEA in solution to yield an ammonium salt and bicarbonate ion: the bicarbonate then proceeds to react with two MEA molecules to produce ammonium salt and carbonate ion.

*Reaction Scheme II* [6]

*Step 1 (fast):* $CO_2 + 2HOCH_2CH_2NH_2 \rightleftharpoons HOCH_2CH_2NH_3^+ + HOCH_2CH_2NHCOO^-$

*Alternate Step 1 (slow):* $H_2O + CO_2 + 2HOCH_2CH_2NH_2 \rightleftharpoons HOCH_2CH_2NH_3^+ + HCO_3^-$

*Step 2 (slow):* $HCO_3^- + 2HOCH_2CH_2NH_2 \rightleftharpoons HOCH_2CH_2NH_3^+ + CO_3^{2-}$

The removal of carbonyl sulfide is achieved through MEA scrubbing when the COS molecule is first hydrolyzed to produce carbon dioxide and hydrogen sulfide as depicted in reaction scheme III. The carbon dioxide and hydrogen sulfide produced by the hydrolysis of COS then react with MEA as outlined in reaction schemes I and II.

*Reaction Scheme III* [1]*:* $COS + H_2O \rightleftharpoons CO_2 + H_2S$

On ASPEN HYSYS a simulation of acid gas removal by MEA scrubbing was obtained by feeding the raw NG feed (s1) into an absorption column (V-002) where it came into contact with the MEA solution. The MEA solution then reacted with the acid gases to remove them and yield sweetened gases (s22) which leave through the top of the column and while the sour gases (s2) leave through the bottom of the column. The sour gas is designed to be flashed (V-003) before entering another distillation column (V-007) which regenerates the MEA solvent. The regenerated solvent is led to a mixer (T-001) where more of the amine solution is added to maintain the correct and optimal composition for gas sweetening. The overall sweetening process occurs at 1.1 bar, but the absorption column operates at a much higher pressure, so the pump (P-004) is put into place to raise the pressure from 1.1 bar to 50 bars. The sweet gas is also led to a flash tank (V-008) to remove as much water as possible before being processed by the dehydration plant.

*Simplified design calculations*[1]

$$Flow\left(\frac{m^3}{h}\right) = 328\left(\frac{Qy}{x}\right) \qquad (1)$$

0.33 mol acid gas pick-up per mole MEA assumed

Where:

Q = Sour gas to be processed, $MSm^3/day$

y = Acid gas concentration in sour gas, mol%

x = Amine concentration in liquid solution, mass%

$$D_c = 10750 \times \sqrt{\frac{Q}{\sqrt{P}}} \qquad (2)$$

Where:

Q = Sour gas to be processed, $MSm^3/day$

P = Contactor pressure is kPa (abs)

$D_c$ = Contactor diameter in millimeters before rounding up to the nearest 100 mm.

$$D_r = 160 \times \sqrt{\frac{m^3}{h}} \qquad (3)$$

Where:

$m^3/h$ = Amine circulation rate in gallons per minute

$D_r$= Regenerator bottom diameter in millimeters.

## Dehydration Process

*Simplified design calculations*

Solvent recirculation rate was calculated using a ratio of 3 gal TEG/ lb $H_2O$.[1]

Following acid gas removal from the natural gas liquid stream, the products were subject to a water removal (dehydration) process. Triethylene Glycol (TEG) was selected as a dessicant for the NGL stream. The sweetening process was connected to the dehydration process on ASPEN HYSYS by transporting the flashed sweetened gas into a contact/absorption column (V-101) where the TEG solvent physically absorbed water from the gas. Once the TEG has absorbed the water, the TEG sank to the bottom of the absorption column while the natural gas that had been stripped of water was transported to a demethanizer and fractionation train (s38). The wet TEG solution (containing the water stripped from the natural gas), was then processed through a boiler that vaporizes only the water, and is then directed to a mixer (T-104) where more glycol may be added to maintain and correct the amount for optimized dehydration. The regenerated glycol was moved to a pump to be recycled in the contactor and to bring the pressure up to the column's operating conditions.

*Background for the simulation*

*Feed Streams*

The composition of the unprocessed gas stream with a volumetric flow rate of 150 MMscfd was used for the simulation is shown in table 1 below.

| **Component** | **Molar Composition** | **Component** | **Molar Composition** |
|---|---|---|---|
| methane | 0.926 | nitrogen | 0.00563 |
| ethane | 0.0311892 | hydrogen sulfide | 0.00047 |
| propane | 0.0104622 | carbonyl sulfide | 1.4E-05 |
| butane | 0.007888 | methyl mercaptan | 3.1E-05 |
| carbon dioxide | 0.0058233 | water | 0.01248 |

Table 1: The composition of the gas stream.

MEA/Glycol process should produce pipeline quality gas and should contain no more than 2-3 mol% of $CO_2$, 0.25-0.3 g/100 scf of $H_2S$, or total sulfur content of 5-20 g/100 scf, 4.0-7.0 lb/MMscf of $H_2O$, and 3 mol% of $N_2$.

**Results**

## MEA Gas Sweetening Process

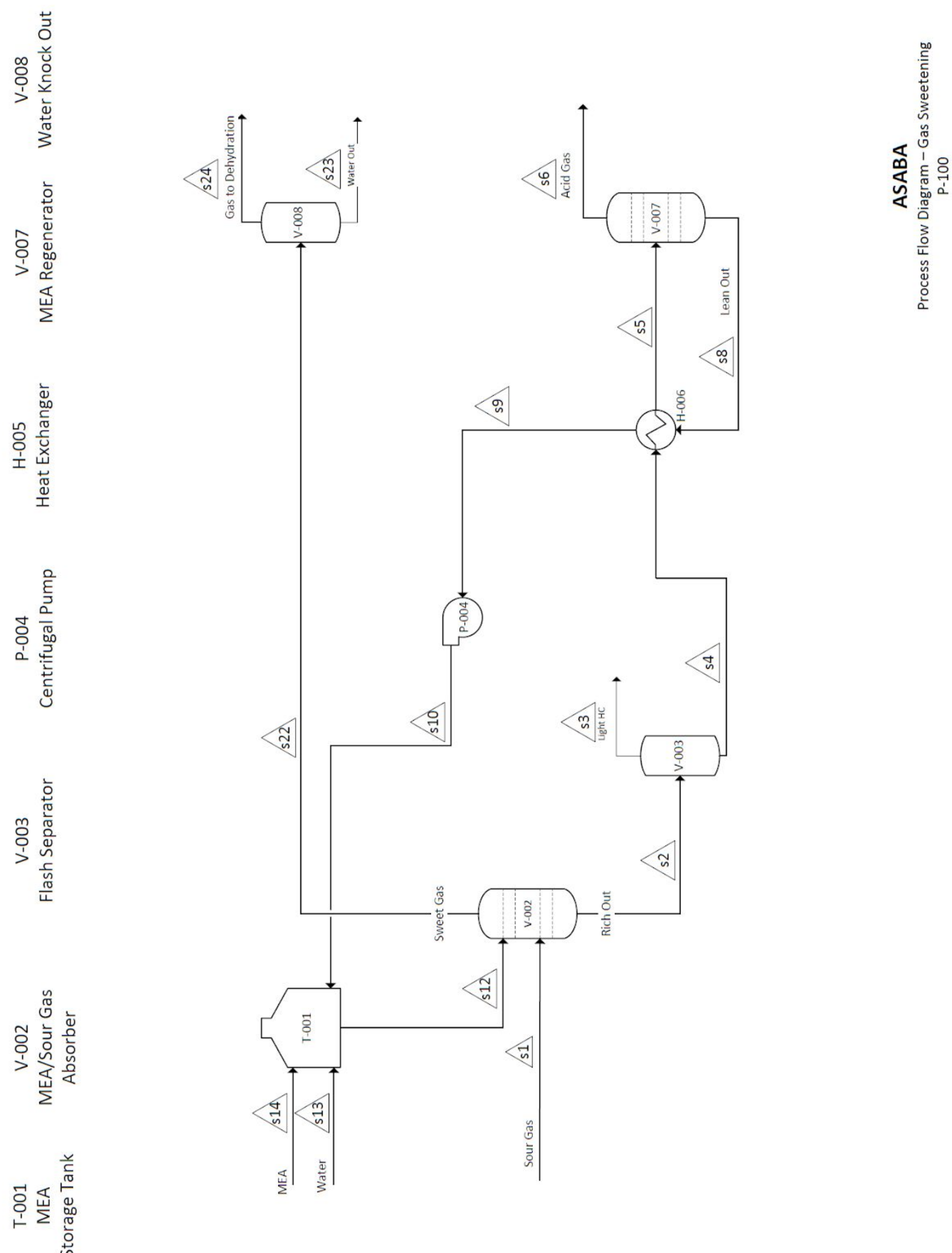


Figure 1: Process flow diagram for the gas sweetening process

Dehydration Process by Triethylene Glycol (TEG)

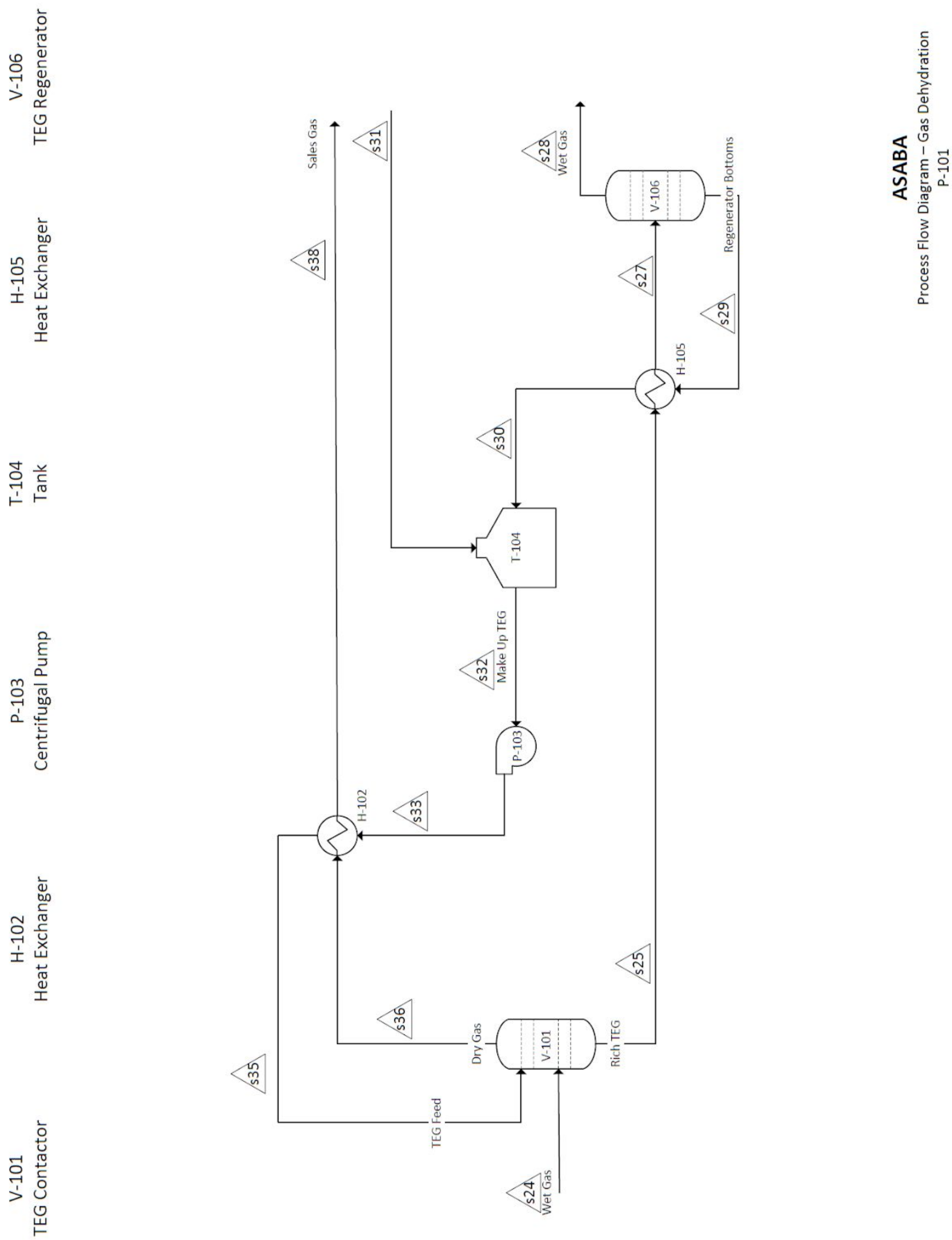


Figure 2: Process flow diagram for the dehydration process

Table 2: Stream summary table

| | Unit | s1 | s2 | s3 | s4 | s5 | s6 |
|---|---|---|---|---|---|---|---|
| Vapor Fraction | | 9.91E-01 | 0.00E+00 | 1.00E+00 | 0.00E+00 | 1.60E-04 | 1.00E+00 |
| Temperature | ℃ | 4.89E+01 | 4.78E+01 | 4.87E+01 | 4.87E+01 | 9.20E+01 | 1.00E+02 |
| Pressure | bar | 6.21E+01 | 5.00E+01 | 1.10E+00 | 1.10E+00 | 1.10E+00 | 1.10E+00 |
| Molar Flow | kgmole/h | 7.47E+03 | 2.23E+03 | 1.66E+00 | 2.23E+03 | 2.23E+03 | 6.34E+02 |
| Mass Flow | kg/s | 3.61E+01 | 1.33E+01 | 7.87E-03 | 1.32E+01 | 1.32E+01 | 3.50E+00 |
| MEAmine | y | 0.00E+00 | 6.55E-02 | 2.22E-05 | 6.56E-02 | 6.56E-02 | 1.54E-07 |
| $H_2O$ | y | 1.25E-02 | 9.11E-01 | 9.97E-02 | 9.12E-01 | 9.12E-01 | 9.26E-01 |
| $CO_2$ | y | 5.82E-03 | 2.11E-02 | 2.21E-03 | 2.11E-02 | 2.11E-02 | 6.81E-02 |
| $H_2S$ | y | 4.74E-04 | 1.59E-03 | 1.61E-03 | 1.59E-03 | 1.59E-03 | 5.58E-03 |
| COS | y | 1.40E-05 | 1.38E-06 | 3.56E-04 | 1.12E-06 | 1.12E-06 | 3.92E-06 |
| M-Mercaptan | y | 3.10E-05 | 1.26E-05 | 7.73E-04 | 1.21E-05 | 1.21E-05 | 4.24E-05 |
| n-Butane | y | 5.26E-03 | 1.43E-06 | 1.85E-03 | 5.31E-08 | 5.31E-08 | 1.86E-07 |
| i-Butane | y | 2.63E-03 | 4.67E-07 | 6.11E-04 | 1.10E-08 | 1.10E-08 | 3.85E-08 |
| Propane | y | 1.06E-02 | 4.67E-06 | 5.94E-03 | 2.35E-07 | 2.35E-07 | 8.24E-07 |
| Ethane | y | 3.12E-02 | 2.00E-05 | 2.52E-02 | 1.24E-06 | 1.24E-06 | 4.35E-06 |
| Methane | y | 9.26E-01 | 6.81E-04 | 8.59E-01 | 3.97E-05 | 3.97E-05 | 1.39E-04 |
| Nitrogen | y | 5.53E-03 | 2.31E-06 | 3.00E-03 | 7.08E-08 | 7.08E-08 | 2.49E-07 |
| TEGlycol | y | *** | *** | *** | *** | *** | *** |

| | Unit | s8 | s9 | s10 | s12 | s13 | s14 |
|---|---|---|---|---|---|---|---|
| Vapor Fraction | | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| Temperature | ℃ | 1.05E+02 | 4.92E+01 | 4.98E+01 | 4.88E+01 | 4.98E+01 | 4.98E+01 |
| Pressure | bar | 1.10E+00 | 1.10E+00 | 5.00E+01 | 5.00E+01 | 5.00E+01 | 5.00E+01 |
| Molar Flow | kgmole/h | 1.59E+03 | 1.59E+03 | 1.59E+03 | 2.12E+03 | 5.29E+02 | 6.55E-02 |
| Mass Flow | kg/s | 9.74E+00 | 9.74E+00 | 9.74E+00 | 1.24E+01 | 2.65E+00 | 1.11E-03 |
| MEAmine | y | 9.17E-02 | 9.17E-02 | 9.17E-02 | 6.88E-02 | 0.00E+00 | 1.00E+00 |
| $H_2O$ | y | 9.06E-01 | 9.06E-01 | 9.06E-01 | 9.29E-01 | 1.00E+00 | 0.00E+00 |
| $CO_2$ | y | 2.36E-03 | 2.36E-03 | 2.36E-03 | 1.72E-03 | 0.00E+00 | 0.00E+00 |
| $H_2S$ | y | 2.82E-10 | 2.82E-10 | 2.82E-10 | 1.88E-10 | 0.00E+00 | 0.00E+00 |
| COS | y | 3.26E-57 | 3.26E-57 | 3.26E-57 | 1.77E-57 | 0.00E+00 | 0.00E+00 |
| M-Mercaptan | y | 1.11E-48 | 1.11E-48 | 1.11E-48 | 8.57E-49 | 0.00E+00 | 0.00E+00 |
| n-Butane | y | 6.20E-103 | 6.20E-103 | 6.20E-103 | 4.65E-103 | 0.00E+00 | 0.00E+00 |
| i-Butane | y | 7.80E-116 | 7.80E-116 | 7.80E-116 | 5.86E-116 | 0.00E+00 | 0.00E+00 |
| Propane | y | 4.91E-98 | 4.91E-98 | 4.91E-98 | 3.68E-98 | 0.00E+00 | 0.00E+00 |
| Ethane | y | 1.88E-94 | 1.88E-94 | 1.88E-94 | 1.40E-94 | 0.00E+00 | 0.00E+00 |
| Methane | y | 2.51E-92 | 2.51E-92 | 2.51E-92 | 1.88E-92 | 0.00E+00 | 0.00E+00 |
| Nitrogen | y | 9.03E-100 | 9.03E-100 | 9.03E-100 | 6.77E-100 | 0.00E+00 | 0.00E+00 |
| TEGlycol | y | *** | *** | *** | *** | *** | *** |

| | Unit | s22 | s23 | s24 | s25 | s27 | s28 |
|---|---|---|---|---|---|---|---|
| Vapor Fraction | | 9.98E-01 | 0.00E+00 | 1.00E+00 | 0.00E+00 | 1.43E-02 | 1.00E+00 |
| Temperature | ℃ | 5.52E+01 | 5.52E+01 | 5.52E+01 | 6.24E+01 | 1.04E+02 | 1.02E+02 |
| Pressure | bar | 5.00E+01 | 5.00E+01 | 5.00E+01 | 6.21E+01 | 1.10E+00 | 1.01E+00 |
| Molar Flow | kgmole/h | 7.37E+03 | 1.28E+01 | 7.35E+03 | 1.37E+02 | 1.37E+02 | 2.42E+01 |
| Mass Flow | kg/s | 3.52E+01 | 6.41E-02 | 3.51E+01 | 4.53E+00 | 4.53E+00 | 1.23E-01 |
| MEAmine | y | *** | *** | *** | *** | *** | *** |
| $H_2O$ | y | 5.05E-03 | 9.99E-01 | 3.32E-03 | 2.26E-01 | 2.26E-01 | 9.30E-01 |
| $CO_2$ | y | 1.52E-05 | 1.72E-07 | 1.52E-05 | 2.27E-06 | 2.27E-06 | 1.29E-05 |
| $H_2S$ | y | 2.00E-08 | 6.02E-10 | 2.00E-08 | 2.10E-08 | 2.10E-08 | 1.19E-07 |
| COS | y | 1.38E-05 | 5.01E-05 | 1.37E-05 | 2.23E-05 | 2.23E-05 | 1.27E-04 |
| M-Mercaptan | y | 2.76E-05 | 2.73E-06 | 2.77E-05 | 4.79E-05 | 4.79E-05 | 2.72E-04 |
| n-Butane | y | 5.33E-03 | 1.70E-06 | 5.34E-03 | 3.74E-04 | 3.74E-04 | 2.13E-03 |
| i-Butane | y | 2.67E-03 | 3.70E-07 | 2.67E-03 | 1.03E-03 | 1.03E-03 | 5.86E-03 |
| Propane | y | 1.07E-02 | 5.14E-06 | 1.08E-02 | 6.16E-04 | 6.16E-04 | 3.50E-03 |
| Ethane | y | 3.16E-02 | 2.13E-05 | 3.17E-02 | 6.42E-04 | 6.42E-04 | 3.65E-03 |
| Methane | y | 9.39E-01 | 7.42E-04 | 9.41E-01 | 9.51E-03 | 9.51E-03 | 5.40E-02 |
| Nitrogen | y | 5.60E-03 | 2.47E-06 | 5.61E-03 | 2.19E-05 | 2.19E-05 | 1.25E-04 |
| TEGlycol | y | 0.00E+00 | 0.00E+00 | 0.00E+00 | 7.62E-01 | 7.62E-01 | 3.26E-05 |

| | Unit | s29 | s30 | s31 | s32 | s33 | s35 |
|---|---|---|---|---|---|---|---|
| Vapor Fraction | | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 | 0.00E+00 |
| Temperature | ℃ | 2.04E+02 | 1.64E+02 | 1.56E+01 | 1.64E+02 | 1.65E+02 | 1.40E+02 |
| Pressure | bar | 1.03E+00 | 1.03E+00 | 1.03E+00 | 1.03E+00 | 6.27E+01 | 6.27E+01 |
| Molar Flow | kgmole/h | 1.13E+02 | 1.13E+02 | 3.74E-02 | 1.13E+02 | 1.13E+02 | 1.13E+02 |
| Mass Flow | kg/s | 4.41E+00 | 4.41E+00 | 1.55E-03 | 4.41E+00 | 4.41E+00 | 4.41E+00 |
| MEAmine | y | *** | *** | *** | *** | *** | *** |
| $H_2O$ | y | 7.52E-02 | 7.52E-02 | 1.00E-02 | 7.51E-02 | 7.51E-02 | 7.51E-02 |
| $CO_2$ | y | 8.78E-11 | 8.78E-11 | 0.00E+00 | 8.77E-11 | 8.77E-11 | 8.77E-11 |
| $H_2S$ | y | 1.56E-11 | 1.56E-11 | 0.00E+00 | 1.56E-11 | 1.56E-11 | 1.56E-11 |
| COS | y | 6.49E-08 | 6.49E-08 | 0.00E+00 | 6.49E-08 | 6.49E-08 | 6.49E-08 |
| M-Mercaptan | y | 6.29E-08 | 6.29E-08 | 0.00E+00 | 6.29E-08 | 6.29E-08 | 6.28E-08 |
| n-Butane | y | 3.28E-08 | 3.28E-08 | 0.00E+00 | 3.27E-08 | 3.27E-08 | 3.27E-08 |
| i-Butane | y | 1.25E-07 | 1.25E-07 | 0.00E+00 | 1.25E-07 | 1.25E-07 | 1.25E-07 |
| Propane | y | 1.55E-08 | 1.55E-08 | 0.00E+00 | 1.54E-08 | 1.54E-08 | 1.54E-08 |
| Ethane | y | 3.13E-09 | 3.13E-09 | 0.00E+00 | 3.13E-09 | 3.13E-09 | 3.12E-09 |
| Methane | y | 1.43E-08 | 1.43E-08 | 0.00E+00 | 1.43E-08 | 1.43E-08 | 1.43E-08 |
| Nitrogen | y | 3.85E-12 | 3.85E-12 | 0.00E+00 | 3.85E-12 | 3.85E-12 | 3.85E-12 |
| TEGlycol | y | 9.25E-01 | 9.25E-01 | 9.90E-01 | 9.25E-01 | 9.25E-01 | 9.25E-01 |

| | Unit | s36 | s38 |
|---|---|---|---|
| Vapor Fraction | | 1.00E+00 | 1.00E+00 |
| Temperature | ℃ | 7.39E+01 | 7.06E+01 |
| Pressure | bar | 6.21E+01 | 4.16E+01 |
| Molar Flow | kgmole/h | 7.33E+03 | 7.33E+03 |
| Mass Flow | kg/s | 3.50E+01 | 3.50E+01 |
| MEAmine | y | *** | *** |
| $H_2O$ | y | 2.62E-04 | 2.62E-04 |
| $CO_2$ | y | 1.52E-05 | 1.52E-05 |
| $H_2S$ | y | 1.97E-08 | 1.97E-08 |
| COS | y | 1.33E-05 | 1.33E-05 |
| M-Mercaptan | y | 2.69E-05 | 2.69E-05 |
| n-Butane | y | 5.35E-03 | 5.35E-03 |
| i-Butane | y | 2.66E-03 | 2.66E-03 |
| Propane | y | 1.08E-02 | 1.08E-02 |
| Ethane | y | 3.18E-02 | 3.18E-02 |
| Methane | y | 9.43E-01 | 9.43E-01 |
| Nitrogen | y | 5.63E-03 | 5.63E-03 |
| TEGlycol | y | 4.94E-06 | 4.94E-06 |

## Flowsheet Calculations

### *Results for Simplified Design Calculation*

The simplified calculations shown above have been used to compute the required circulation rate for the two solvents responsible for acid gas removal and water removal: MEA and TEG. The diameters of the absorption column and regenerator column have also been calculated.

Table 3: Results for Simplified Calculations

| | **MEA calc.** | **TEG calc.** |
|---|---|---|
| **Recirculation[$m^3/h$]** | 43.863 | 11.82 |
| **Absorption column diameter[mm]** | 2493.4 | *2500.0 |
| **Regenerator column diameter[mm]** | 2223.4 | *2500.0 |

*Indicates the default diameter Aspen Hysys utilizes

### *Mass Balance*

Aspen HYSYS and an EXCEL spreadsheet was used to perform the mass balance for the system. The overall mass balance is summarized in the Table 4.

Table 4: Overall Mass Balance for MEA/Glycol process

| Mass In | Mass In | Mass Out | Mass Out |
|---|---|---|---|
| ID | kmol/h | ID | kmol/h |
| s1 | 7471.06 | s3 | 1.663898272 |
| s13 | 529.27 | s6 | 634.4792336 |
| s14 | 0.07 | s23 | 12.8035546 |
| | | s28 | 24.16202634 |

| | | | |
|---|---|---|---|
| | | s31 | 0.037410405 |
| | | s38 | 7329.561159 |
| Total | 8000.40 | Total | 8002.707282 |
| Imbalance | 2.31 | | |
| Relative Imbalance (%) | | 0.028832437 | |

The percentage of error is smaller than 1%, and thus acceptable.

**Economics**

Table 5: Equipment list and Specifications

| **Exchangers** | **Exchanger Type** | **Shell Pressure (barg)** | **Tube Pressure (barg)** | | **MOC** | **Area (square meters)** |
|---|---|---|---|---|---|---|
| H-006 | Fixed, Sheet, or U-Tube | 1.1 | 1.1 | | Stainless Steel / Carbon Steel | 60.3 |
| H-102 | Fixed, Sheet, or U-Tube | 1.5 | 1.5 | | Stainless Steel / Carbon Steel | 5.03 |
| H-105 | Fixed, Sheet, or U-Tube | 63 | 63 | | Stainless Steel / Carbon Steel | 5.03 |
| Mixers/Storage | Type | Power (kW) | # Spares | | | |
| T-001 | Impeller | 0.746 | 1 | | | |

| | | | | | | |
|---|---|---|---|---|---|---|
| T-104 | Impeller | 0.746 | 1 | | | |
| Pump | Type | Power (kW) | # Spares | MOC | Discharge Pressure (barg) | |
| P-004 | Centrifugal | 63.2 | 1 | Stainless Steel | 50 | |
| P-103 | Centrifugal | 36.2 | 1 | Carbon Steel | 62.7 | |
| Towers | Tower Description | Height (m) | Diameter (m) | Tower MOC | Demister MOC | Pressure (barg) |
| V-002 | 20 Stainless Steel Sieve Trays | 4.8 | 2.49 | Stainless Steel | | 50 |
| V-007 | 19 Stainless Steel Valve Trays | 7.63 | 2.22 | Stainless Steel | | 1.1 |
| V-101 | 14 Carbon Steel Sieve Trays | 6.54 | 1.5 | Carbon Steel | | 62.1 |
| V-106 | 18 Carbon Steel Sieve Trays | 7.2 | 4 | Carbon Steel | | 1.01 |
| Vessels | Orientation | Length/Height (m) | Diameter (m) | MOC | Demister MOC | Pressure (barg) |
| V-003 | Vertical | 1.79 | 1.19 | Carbon Steel | | 50 |
| V-008 | Horizontal | 1.95 | 1.95 | Stainless Steel | | 1.1 |

Table 6: Equipment Cost Information

| Exchangers | Purchased Equipment Cost | Bare Module Cost | Base Equipment Cost | Base Bare Module Cost |
|---|---|---|---|---|
| H-006 | $ 55,300.00 | $ 141,000.00 | $ 30,500.00 | $ 100,000.00 |
| H-102 | $ 42,100.00 | $ 108,000.00 | $ 23,300.00 | $ 76,500.00 |
| H-105 | $ 53,100.00 | $ 126,000.00 | $ 23,300.00 | $ 76,500.00 |
| **Mixers/Storage** | **Purchased Equipment Cost** | **Bare Module Cost** | **Base Equipment Cost** | **Base Bare Module Cost** |
| T-001 | $ 66,600.00 | $ 91,900.00 | $ 66,600.00 | $ 91,900.00 |
| T-104 | $ 66,600.00 | $ 91,900.00 | $ 66,600.00 | $ 91,900.00 |
| **Pump** | **Purchased Equipment Cost** | **Bare Module Cost** | **Base Equipment Cost** | **Base Bare Module Cost** |
| P-004 | $ 125,000.00 | $ 224,000.00 | $ 29,300.00 | $ 95,000.00 |
| P-103 | $ 67,500.00 | $ 131,000.00 | $ 21,300.00 | $ 69,100.00 |
| **Towers** | **Purchased Equipment Cost** | **Bare Module Cost** | **Base Equipment Cost** | **Base Bare Module Cost** |
| V-002 | $ 1,390,000.00 | $ 2,460,000.00 | $ 125,000.00 | $ 221,000.00 |
| V-007 | $ 415,000.00 | $ 593,000.00 | $ 194,000.00 | $ 308,000.00 |
| V-101 | $ | $ | $ | $ |

| | | | | |
|---|---|---|---|---|
| | 219,000.00 | 410,000.00 | 47,600.00 | 106,000.00 |
| V-106 | $ 367,000.00 | $ 651,000.00 | $ 344,000.00 | $ 629,000.00 |
| **Vessels** | **Purchased Equipment Cost** | **Bare Module Cost** | **Base Equipment Cost** | **Base Bare Module Cost** |
| V-003 | $ 41,800.00 | $ 91,100.00 | $ 6,660.00 | $ 27,100.00 |
| V-008 | $ 37,400.00 | $ 74,700.00 | $ 12,000.00 | $ 36,200.00 |
| | Purchased Equipment Cost | Bare Module Cost | Base Equipment Cost | Base Bare Module Cost |
| **Totals** | **$ 2,946,400.00** | **$ 5,193,600.00** | **$ 990,160.00** | **$ 1,928,200.00** |

Table 7: Cost of Raw Materials

| **Material** | **kg/h** | **Cost ($/lb)** | **Cost ($/kg)** | **CpH ($/hr)** | **CpA ($/yr)** | **Source** |
|---|---|---|---|---|---|---|
| **Methanolamine (MEA)** | 53174 | 0.59 | 1.3 | 69164.16 | 605,878,013.05 | ICIS |
| **Triethylene Glycol (TEG)** | 512 | $ 0.56 | 1.23 | 632.0041 | 5,536,356.32 | ICIS |
| **Raw Gas Mixture** | 129800 | $ 0.14 | 0.31 | 40061.99 | 350,943,042.91 | |
| **Total** | | | | 109858.15 | 962,357,412.28 | |

Table 8: Economic Options

| | |
|---|---|
| **Cost of Land** | $ 250,000.00 |

| | |
|---|---|
| **Taxation Rate** | 45% |
| **Annual Interest Rate** | 10% |
| **Salvage Value** | 1400000 |
| **Working Capital** | 92900000 |
| **$FCI_L$** | 14000000 |
| **Total Module Factor** | 1.18 |
| **Grass Roots Factor** | 0.5 |

Table 9: Economic Information Calculated From Given Information from tables above.

| | |
|---|---|
| **Revenue From Sales** | 5.91E+09 |
| **$C_{RM}$ (Raw Materials Costs)** | 9.14E+08 |
| **$C_{UT}$ (Cost of Utilities)** | 64800 |
| **$C_{WT}$ (Waste Treatment Costs)** | 0 |
| **$C_{OL}$ (Cost of Operating Labor)** | 1003650 |

Table 10: Factors Used in Calculation of Cost of Manufacturing ($COM_d$)

| | |
|---|---|
| **Multiplying factor for $FCI_L$** | 0.18 |

| | |
|---|---|
| **Multiplying factor for $C_{OL}$** | 2.76 |
| **Factors for $C_{UT}$, $C_{WT}$, and $C_{RM}$** | 1.23 |
| **$COM_d$** | 1.13E+09 |

$$Com_d = 0.18*FCIL + 2.76*C_{OL} + 1.23*(C_{UT} + C_{WT} + C_{RM})$$

**Project Life (Years after Startup) = 10**

**Construction period = 2**

**Distribution of Fixed Capital Investment:**

**End of year One = 60%**

**End of year Two = 40%**

**Discussion**

## Design Software

Aspen HYSYS was selected as the modeling software for ASABA technology over Superpro and Simcentral because ASPEN HYSYS provided more solvent options, more functions, user friendly operations, and has widespread use in the chemical industry. Additionally, ASPEN HYSYS is a specific computational model which has specialized functions for hydrocarbon processing while ASPEN Plus V8.8, Simcentral, and Superpro do not.

## Acid Gas Removal Process

Many different amines are used in gas treating, diethanolamine (DEA), monoethanolamine (MEA), methyldiethanolamine (MDEA), diisopropanolamine (DIPA), and aminoethoxyethanol

(Diglycolamine) (DGA); however, MEA was selected for the acid gas removal (sweetening) process because it is highly efficient in removing the acid gases specified in ASABA technology's raw natural gas feed which were hydrogen sulfide, carbon dioxide, carbonyl sulfide, and methylmercaptan. Within the MEA process design, a flashing the stream was incorporated to remove any hydrocarbons that did not separate in the first absorption column.

The economic parameters of the acid gas removal and dehydration processes outlined in this project were determined using the CapCost spreadsheet provided by Richard Turton et al.[7] The CEPCI (Chemical Engineering Plant Cost Index) value used was 603.1, which is the average value for the year of 2018.[2] The cost of raw materials for the MEA gas scrubbing process (outlined in table 7) was determined to be 0.59 dollars per pound[4] of MEA reagent and is projected to cost ASABA Technology an estimated 605,878,013.05 dollars per year to operate the acid gas removal unit of the natural gas liquids recovery chemical plant.

## Dehydration Process

Triethylene Glycol (TEG) was selected as a dessicant for the NGL stream because over other glycol types because TEG is more easily regenerated to a higher degree of purity, results in higher vapor recoveries, and the operating costs are lower relative to other dehydration methods.[6]

ASABA Technology will source the TEG solution for 0.56 dollars per pound of TEG reagent and and the whole dehydration unit is projected to cost ASABA Technology an estimated

$5,536,356.32 dollars per year to operate. As outlined in table 7, the total operating costs for the sweetening and dehydration processes was estimated by CapCost to be $962,357,412.28.

**Conclusion**

The goal of designing working simulations on ASPEN HYSYS for the acid gas removal and dehydration processes of natural gas liquids was successfully achieved along with an analysis and outline of their economic parameters. The designs constructed in this project will aid ASABA Technology to complete the grand goal of designing a full-scale chemical plant with a processing capability of 150 MMscfd ($ft^3/d$) for raw natural gas processing to generate quality pipeline gas for consumption.

**Table of Nomenclature**

| | |
|---|---|
| MMscfd | Million standard cubic feet per day |
| MMscf | Million standard cubic feet |
| $MSm^3$ | Million standard cubic feet |
| lb/MMscf | Pounds per million standard cubic feet |
| NGL | Natural gas liquids |
| LNG | Liquid natural gas |
| MOC | Materials of Construction |
| y | Mole fraction for gas mixtures |

| | |
|---|---|
| $FCI_L$ | Fixed Capital Investment |
| $C_{RM}$ | Raw Materials Costs |
| $C_{UT}$ | Cost of Utilities |
| $C_{WT}$ | Waste Treatment Costs |
| $C_{OL}$ | Cost of Operating Labor |
| $COM_d$ | Cost of Manufacturing |
| CEPCI | Chemical Engineering Plant Cost Index |
| kg/s | Kg per second |
| kgmole/h | kgmole per hour |

**References**


1. B. Alireza, Natural Gas Processing Technology and Engineering and Design (Gulf Professional Publishing, ed. 3, 2014). pp. 487, 502
2. Chemengonline.com. (2019). *Chemical Engineering Plant Cost Index: 2018 Annual Value - Chemical Engineering | Page 1*. [online] Available at: https://www.chemengonline.com/2019-cepci-updates-january-prelim-and-december-2018-final/ [Accessed 29 May 2019].
3. H. M. King, Uses of Natural. *Geoscience News and Information*. 1, (2013).
4. ICIS Explore. (2019). *Chemicals A-Z*. [online] Available at: https://www.icis.com/explore/chemicals/channel-info-chemicals-a-z/

5. Kidnay, A. and Parrish, W. (2006). *Fundamentals of natural gas processing*. Boca Raton: CRC Press. pp.189-191

6. Kohl, A. and Nielsen, R. (1997). *Gas purification*. 5th ed. Houston, Texas: Gulf, pp.165-166.

7. Speight, J. G. (2017). Properties Processing of Gas From Tight Formations. *Deep Shale Oil and Gas,* 307-347. doi:10.1016/b978-0-12-803097-4.00007-3

8. Turton, R. (2018). *Analysis, synthesis, and design of chemical processes*. Boston: Prentice Hall.