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Dynamic modeling and experimental investigation of a high temperature PEM fuel cell stack

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ABSTRACT

High temperature polymer fuel cells operating at 100 to 200°C require simple fuel processing and produce high quality heat that can integrate well with domestic heating systems. Because the transportation of hydrogen is challenging, an alternative option is to reform natural gas on site. This article presents the development of a dynamic model and the comparison with experimental data from a high temperature proton exchange membrane fuel cell stack operating on hydrogen with carbon monoxide concentrations up to 0.8%, and temperatures from 155 to 175°C. The dynamic response of the fuel cell is investigated with simulated reformat gas. The dynamic response of the fuel cell stack was compared with a step change in current from 0.09 to 0.18 and back to 0.09 A/cm². This article shows that the dynamic model calculates the voltage at steady state well. The dynamic response for a change in current shows that the model compares well with some of the cells in the stack while other cells may have typically lower voltages levels during dynamic operation.

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Introduction

Distributed generation can provide electricity with fewer emissions of greenhouse gases and pollutants compared to traditional centralized power generation. By generating the electricity close to the point of use, both electricity and heat can be used and losses from the transmission and distribution system can be avoided. In developed countries, installing distributed generation provides means to increase the reliability and resiliency of the electrical grid, and allows incremental and pointed investment. In developing countries where the electrical infrastructure is sparse or not reliable,

installing distributed generation is an attractive alternative to provide electricity without huge infrastructure investments.

The technologies available for distributed generation include reciprocating engines, gas turbine engines, and fuel cells. High temperature proton exchange membrane fuel cell systems are the focus of this paper. Currently, most fuel cells installed for stationary power generation are typically operating at steady-state base-load conditions. While this is currently the easiest and most economical way to operate a fuel cell system, the dynamic operation of the fuel cells is necessary for off-grid operation. For example, to accommodate for the intermittent electrical power output from renewable energy sources such as wind and solar, fuel cell

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systems could be dynamically dispatched [1]. In addition, the ability to follow the electrical load will help fuel cells to gain more acceptance from the electric utility companies.

PEM fuel cells can typically change power output settings quickly, especially when operating on pure hydrogen. This ability is one of reasons for the extensive use of PEM fuel cell technology in automotive applications. The PEM fuel cells used in cars operate at temperatures between 20 and 80 °C. A limitation of these so-called “low temperature” PEM fuel cells is the requirement of an input stream almost free of carbon monoxide (CO) and other impurities. In the use of fuel cells for stationary power generation, researchers have developed and deployed integrated low temperature PEM fuel cells and fuel processors [2,3,4]. More recently, PEM fuel cell technology with higher operating temperatures, higher CO levels and better integration with heat recovery has been investigated [5]. Researchers have developed and tested multiple alternative high temperature PEM fuel cell types [6,7]. Among the materials evaluated, the polybenzimidazole (PBI) membrane doped with phosphoric acid allows fuel cells to operate between 160 and 200 °C and have desirable mechanical properties for use in fuel cells. This is the type of high temperature PEM short stack that is studied in this paper.

High temperature PEM fuel cells have demonstrated tolerance of carbon monoxide and other impurities in reformate gases such as methanol and carbon dioxide. Researchers have studied in detail the adsorption of carbon monoxide on the catalyst sites. The common tool used in these studies is electro-chemical impedance spectroscopy (EIS) used to diagnose the effect of CO poisoning on fuel cell performance [8,9,10,11]. Following the EIS results, researchers have developed equivalent circuit models to quantify the effect of impurities such as CO on the fuel cell performance [12], developed a finite two-dimensional model of the electrode structure and of the diffusion process at the interfaces, and fitted the model to the measured polarization curve [13]. Some researchers studied the kinetics of CO coverage as it was adsorbed on the catalyst [14]. These research projects have indicated that CO has a less severe effect on the high temperature PEM fuel cell operation at temperatures between 160 and 200 °C compared to low temperature PEM fuel cells. However, the presence of methanol and carbon dioxide may compound the negative effect on the catalyst [8].

An important aspect of system design for the high temperature PEM fuel cell is thermal management and integration with fuel processing equipment. During a start-up process, the thermal management system of the fuel cell must heat up the fuel cell from the environment temperature to a reasonable fuel cell operating temperature ($T > 100$ °C, boiling point of water). Andreasen and Kær studied the heating of the fuel cell stack by electrical heating and by hot air heating [15]. A dynamic model exploring the heat transfer and the heat generated within the fuel cell was developed and compared with the experimental results. During operation, the thermal management system maintains the operating temperature by rejecting heat produced within the fuel cell to its surrounding, or using it for other processes. Harikishan Reddy and Jayanti studied the cooling options of a fuel cell stack for a scooter by considering the use of cooling plates or forced draft on the system [16]. Supra et al. proposed a different method of using

a heat pipe within the system [17]. In summary, these investigators have studied various thermal management methods for bringing the fuel cell stack to its operating point effectively and maintaining the operating temperature in a safe and desired range.

During operation, voltage and current are the two most important parameters one must know to evaluate the performance of a fuel cell. Increasing temperature tends to improve the performance of the fuel cell as it overcomes activation polarization, but higher temperatures also accelerate degradation processes, as presented in [18,19]. In parallel with the experimental data, models have been developed to study the phenomena of interest such as the influence of the charge double layer [20], or the distribution of current and gases in the flow channels [21]. These models make use of computational fluid dynamics to gain insights into the performance of the fuel cell. Another approach in predicting cell voltage and current is the use of a bulk model or simplified geometry model for fast simulation. Such bulk model can be used for control systems development and can be more readily used to simulate a complete system with all of the balance-of-plant components (e.g. heat exchangers, blowers). First principles are used by many in a bulk control volume to determine the temperature and species concentration (sometimes spatially resolved). For example Park and Min modeled a fuel cell in quasi-three dimensional geometry that included first principles equations solved in several control volumes to determine the local current, temperature and species concentrations in the cell [22]. The investigators have shown that a dynamic model can calculate the dynamic response of rapid demand changes, and can be implemented to model a complete fuel cell system consisting of a fuel processing system, heat exchangers, blowers, off-gas combustor, and a fuel cell stack [23].

Previous models that have been developed have limitations. First, the experimental verification is usually performed at one set of operating conditions, whereas the actual fuel cell will experience different temperatures, for example, during start-up processes and during transient operation. Thus it is important to test the model ability and limits when subjected to changes in temperature and fuel flow composition. Second, only a few models in the literature are dynamic models [24,25,26,27,28]. Understanding the transient behavior of the fuel cell is important for developing control systems and for the dynamic dispatch operation of fuel cell systems that will be required in the future. A dynamic model that can compute accurately and quickly, and can be integrated with other balance-of-plant components is necessary for control system development.

The objectives of this paper are (i) to present a dynamic model of a high temperature PEM fuel cell that can calculate the thermal response of a fuel cell stack accurately, and (ii) to acquire experimental data on a fuel cell short stack for both steady state and transient conditions that are then used for model verification. A recent publication presented a dynamic model with a very similar modeling approach [22]. The current work further extends the dynamic model results by comparing them with experimental data from a high temperature PEM fuel cell short stack. The experiments on steady state operation are used to verify the model at different

temperatures and gas concentrations. Next, the fuel cell short stack was subjected to perturbations comprised of an increase and decrease in current demand. The dynamic stack and cell voltage responses of the fuel cell are then compared with the model.

Experimental setup

A 14 cell stack was tested in a Greenlight fuel cell test setup with an external oil cooler/heater. Details of the experimental setup are presented elsewhere [29]. A setup schematic is presented in Fig. 1 and a picture of the setup is presented in Fig. 2.

This current Greenlight fuel cell test setup is capable of mixing gases on both the anode and the cathode sides. On the anode side, a collection of mass flow controllers can produce mixtures containing H_2 , CO_2 , CO , CH_4 and H_2O . On the cathode side, a mixture of air and H_2O is possible. The system will itself calculate the specific flows based upon input gas compositions for the anode and cathode flows.

The system is connected to a cooling/heating system that is designed to control the input temperature of the oil to the fuel cell. The temperature and pressure of the oil is measured at input and output of the fuel cell stack. The oil coolant used in this system is Paratherm NF heat transfer fluid [30] and this cooling/heating system can be seen on the left-hand side of the photograph in Fig. 2.

The fuel cell stack that was used in the current experiments was produced by Serenergy [31]. The fuel cell stack is comprised of 14 cells. Each cell has a cross-sectional active area of 163.5 cm^2 and a serpentine gas flow geometry. The bipolar plates are designed with parallel oil flow channels.

The fuel cell was tested with a range of different gas compositions and temperatures. For every operating parameter that was manipulated, two polarization curves are created. These polarization curves were produced by increasing the current by 2 A over 5 s, waiting for 30 s, then

Fig. 2 – Fuel cell test setup.

increasing the current again until the voltage of one cell fell below 0.2 V. When this limit was reached, the load was decreased in 4 A decrements until no load was applied. Between the parameter changes, the fuel cell is reset by running at a 33 A load (corresponding to 0.2 A/cm^2) on pure hydrogen. The experiment proceeds after purging pure hydrogen for 600 s and all the cell voltages are above 0.5 V.

The test sequence was fully automated and a range of different parameters were tested. The parameters tested in this setup were operating fuel cell stack temperatures, anode and cathode stoichiometry, and simulated steam reforming gas concentrations with variable CO concentrations. The temperature was varied between 150°C and 175°C and experiments were conducted using both pure hydrogen and reformat gas. On pure hydrogen, the anode stoichiometry was tested in the range from 1.25 to 1.4, and the cathode stoichiometry was tested in the range from 2 to 4. The recommended stoichiometry for this HTPCM fuel cell is 1.35 for the anode and 2–3 for the cathode [31]. The reformat gas test was conducted with a constant stack temperature of 160°C and a constant stoichiometry. The reference reformat gas

Fig. 1 – Schematic of the system setup.

Table 1 – Test matrix. N₂ is used for make-up gas.

Index	Temp. (°C)	Cathode stoic.	Anode stoic.	CO (%)
Anode: hydrogen (100%)				
1–6	150–175	3.5	1.35	0
7–11	160	2–4	1.35	0
12–15	160	3.5	1.25–1.4	0
Anode: reformat gas (60% H ₂ , 15% H ₂ O)				
16–20	155–175	3.5	1.35	0.8
21–23	160	3.5	1.35	0.25–1
24–27	160	3.5	1.4–1.6	0.8
28–31	160	2–3.5	1.35	0.8
40–41	160	3.5	1.35	1.5–2
Anode: reformat gas (41% H ₂ , 8.5% H ₂ O)				
32–39	160–180	3.5	1.35	1

composition is set to 60% H₂, 24.2% CO₂, 0.8% CO, and 15% H₂O. The complete test matrix is shown in Table 1.

Model description

The dynamic model of the fuel cell is developed to identify the electrical, thermal, chemical, and electrochemical response of the fuel cell, using a simplified geometric representation of the fuel cell stack and which could enable a control system development and model-based control. To maintain short simulation time, a model that only resolves the unit cell control volumes (e.g. bipolar plate, cathode gas compartment, membrane electrode assembly, and anode gas compartment) is used. The seven gases considered are CH₄, CO, CO₂, H₂, H₂O, N₂, and O₂. A vector containing the flow rate, gas concentration, and temperature is passed between component models. The fuel cell is divided into nine control volumes in the direction perpendicular to the flow direction. As shown in Fig. 3, the five types of control volumes used in the model are: solid plate, bulk gas, gas diffusion layer (GDL), membrane electrode assembly (MEA),

and coolant. Within these control volumes, the model simulates the heat transfer and mass transport to determine temperature and gas concentrations. The voltage is calculated from the gas concentration at the electrode-electrolyte interface. In the fuel cell, the hydrogen is consumed along the flow channels causing a distribution in current density within the cell. Some models attempt to simulate the change in concentrations by incorporating the spatial discretization along the gas channel [22,27] and using the equipotential constraint to solve for the current distribution. In contrast, this model assumes a uniform current distribution and iteratively determines the voltage from the gas concentrations at the outlet of the cell.

Major assumptions

The modeling method follows the method outlined by Mueller et al. [27]. At the high operating temperature, water is assumed to be in vapor phase in the gas channels and electrodes so that a single phase flow is considered. The water drag coefficient that is used to determine water back-diffusion from cathode to anode in a traditional low temperature PEM is assumed to be zero for this high temperature PEM fuel cell [32]. Finally, because the high temperature fuel cell shows negligible degraded performance effects from carbon monoxide levels below 2%, the poisoning effect of carbon monoxide is not considered in the current model [33]. The experiments conducted in the current work will partially verify whether or not this is a good assumption.

Energy conservation

The fuel cell stack is modeled as comprised of solid control volumes (i.e., bipolar plates, MEA, separator plate) and fluid control volumes (i.e., reformat gases, air and fuel channels, and coolant). The solid control volumes have no mass flow across their boundaries; whereas the fluid control volumes (i.e., gas and coolant) have mass flow across their boundaries. Two separate energy equations are used to determine the temperatures of the components in the fuel cell stack, as presented below.

Temperatures of the solids are calculated using the following ordinary differential equation

$$\rho V C \frac{dT}{dt} = \sum \dot{Q}_{in} + \dot{Q}_{gen} \quad (1)$$

where ρ is the density of the solid, V is the volume, C is the specific heat capacity, measured in kJ/kg-K, $\dot{Q}_{in}$ is the net amount of heat entering the solid volume, $\dot{Q}_{gen}$ is the amount of heat generated in the solid volume.

Temperatures of the gas and coolant are determined by solving the energy conservation equation

$$NC_p \frac{dT}{dt} = \sum \dot{N}_{in} h_{in} - \sum \dot{N}_{out} h_{out} + \sum \dot{Q}_{in} + \dot{Q}_{gen} \quad (2)$$

where N is the number of moles in the control volume, C_p is the specific heat capacity of the gas, measured in kJ/kmol-K, $\dot{N}_{in} h_{in}$ and $\dot{N}_{out} h_{out}$ is the enthalpy flow in and out of the control volume, $\dot{Q}_{in}$ is the net amount of heat entering the solid

Fig. 3 – Control volumes of the fuel cell model.

volume, $\dot{Q}_{gen}$ is the amount of heat generated within the control volume due to chemical reactions.

Species conservation

Species mole fractions at the exit of each control volume are:

$$\frac{d(N\vec{X})}{dt} = \dot{N}_{in}\vec{X}_{in} - \dot{N}_{out}\vec{X}_{out} + \sum \vec{\Phi} \quad (3)$$

Where $\vec{X}$ is the species mole fraction vector, $\vec{\Phi}$ are the species diffusion fluxes from the adjacent volume.

Assuming the control volume is a perfectly stirred reactor and treating all gases as ideal, the model calculates the outlet concentration.

Fuel cell mass transport

Gases in the anode and cathode channels diffuse through the gas diffusion layers (GDL) to and from the catalyst layers. At the catalyst layers (assumed to comprise the interface between the GDL and the electrolyte control volumes), gas molecules participate in electrochemical reactions at this triple-phase interface of the catalyst layer, electrolyte, and gas phase. The model calculates the mass transport of hydrogen, water, oxygen, and nitrogen. The mass transport coefficient in the gas channel and through the GDL is:

$$\vec{g}_m = \frac{Sh \vec{D}_m}{D_H} \quad (4)$$

where Sh is the Sherwood number, $\vec{D}_m$ is the diffusion coefficient, and D_H is the hydraulic diameter of the gas flow channel.

The Sherwood number is determined from the heat and mass transfer analogy:

$$\frac{Nu}{Pr^n} = \frac{Sh}{Sc^n} \quad (5)$$

where Nu is the Nusselt number, Pr is the Prandtl number, Sh is the Sherwood number, and Sc is the Schmidt number.

Rearrange the equation to solve for the Sherwood number, we obtain

$$Sh = Nu \left(\frac{Sc}{Pr} \right)^n = Nu \frac{\alpha}{D_{AB}} \quad (6)$$

where α and D_{AB} are the thermal and mass diffusivities respectively.

The diffusion coefficient is calculated using the Bruggeman correlation with a modification for the effect of porosity and tortuosity in the GDL as follows:

$$\vec{D}_m = \vec{D}_0 \left(\frac{T}{T_0} \right)^{3/2} \left(\frac{P_0}{P} \right) \quad (7)$$

$$\vec{D}_m^{eff} = \varepsilon^{1.5} \vec{D}_m \quad (8)$$

where $\vec{D}_0$ is the species diffusion coefficient at standard pressure and temperature, $\vec{D}_m^{eff}$ is the effective species diffusion coefficient and ε is the GDL porosity.

Components of the diffusion resistance $\vec{R}_{dif}$ are:

$$R_{i,diff} = A \frac{1}{\frac{1}{g_{i,m}} + \frac{t_{gdl}}{D_{i,m}^{eff}}} \quad (9)$$

where $R_{i,diff}$ is the total diffusion resistance of each species, A is the area of diffusion, $g_{i,m}$ is the mass transport coefficient from equation (4), t_{gdl} is the thickness of the GDL, and $D_{i,m}^{eff}$ is the effective diffusion coefficient of the i th species from equation (7).

The species diffusion fluxes $-\Phi$ between the GDL and the gas channels are then:

$$\vec{\Phi} = \vec{R}_{dif} \left(\vec{C}_2 - \vec{C}_1 \right) \quad (10)$$

where $\vec{C}_1$ and $\vec{C}_2$ are the concentration in GDL and the bulk gas.

Fuel cell heat transfer

The convective heat transfer between solids and gases are solved as follows:

$$\dot{Q} = hA(T_2 - T_1) \quad (11)$$

where h is the convective heat transfer coefficient, A is the convective area, and $(T_2 - T_1)$ is the temperature difference.

The convection coefficient h is determined from the Nusselt number (Nu) as follows:

$$h = \frac{Nu \sum x_i k_i}{D_H} \quad (12)$$

where $\sum x_i k_i$ is the conductive heat transfer coefficient of the gas mixture, and D_H is the hydraulic diameter.

Conduction heat transfer is determined from Fourier's law:

$$\dot{Q} = \frac{kA}{L}(T_2 - T_1) \quad (13)$$

where k is the conduction heat transfer coefficient, A is the conduction heat transfer area, L is the thickness, and $(T_2 - T_1)$ is the temperature difference between the two mediums.

Fuel cell electrochemical reactions

The model calculates the Nernst voltage, and the activation and ohmic losses to determine the voltage of the fuel cell. Because the fuel cell operates at relatively low current density, the model does not separately account for a concentration loss. The physics of concentration polarization is rather explicitly accounted for in the GDL diffusion calculations. The cell voltage is thus determined as:

$$V_{cell} = V_{Nernst} - \eta_{act} - \eta_{ohm} \quad (14)$$

where V_{Nernst} is calculated based upon the temperature and concentrations of H_2 , O_2 , and H_2O at the electrode-electrolyte interface.

The activation polarization is calculated from the Tafel equation. The exchange current density at the reference point is expressed as an exponential function in the form show in Equation (15) and determined from previous experimental measurements. The specific parameter values used in the model are given in Table 2. Compared to the exchange current

Table 2 – Model parameters used in the current fuel cell model.

Description	Value	Units
Geometry		
Depth of gas channel (anode and cathode)	1	mm
Depth of cooling channel	2	mm
Thickness of GDL	0.4	mm
Thickness of electrolyte	0.05	mm
Thickness of separator plates	1	mm
Thermodynamic properties		
Separator plate density	2210	kg/m ³
Separator plate specific heat capacity	0.5	kJ/kg-K
Electrolyte dry density	2200	kg/m ³
Electrolyte dry equivalent weight	1000	kg/kmol
Electrolyte solid specific heat capacity	2.179	kJ/kg-K
Heat transfer properties		
Separator plate & bipolar plate conduction coefficient	0.22	kW/m-K
Nusselt number of anode gas & cathode gas	6	
Nusselt number of coolant liquid	15	
Mass transport properties		
GDL void fraction	0.5	
GDL porosity	0.51	
Area of diffusion	225	cm ²
Polarization constant		
GDL electronic conductivity	90	S/m
Membrane proton conductivity	20	S/m
Exchange current density		
A	1.39×10^{-8}	A/m ²
B	0.04	1/K

density measured by [34], the model fit used in the current work calculates a slightly higher exchange current density at low temperature, as shown in Fig. 4. However, the overall shape and magnitude are similar between the current model and those of [34]. The exchange current density is calculated as:

$$j_0 = a \exp(-bT) \quad (15)$$

where T is the temperature of the electrolyte, a and b are the fitting parameters.

Fig. 4 – Comparing the exchange current density in the literature with the model's fit.

Results and discussion

Steady state performance

Fig. 5 shows the experimental results for pure hydrogen and air operation at 160 °C. The anode stoichiometry is 1.35 and the cathode stoichiometry is 3.5. In the plot, the mean and one standard deviation are included, as determined from the measured cells voltages. It can be seen that the deviation grows larger as the current increases. At current density below 0.3 A/cm², the standard deviation is below 30 mV. At a current density of 0.5 A/cm², the standard deviation grows to 80 mV. This large deviation is caused by some weak cells in the stack that cannot sustain voltage when high current density is imposed. The steady-state performance of the model is also presented in Fig. 5. Compared to the experimental data, the model results agree to within 10 mV throughout the current density regime tested. The calculated voltage deviates from the standard deviation band only at very low current density, which is below the practical operation range. Because the values of the standard deviation do not differ much in subsequent measurements, the standard deviations are omitted in subsequent plots to enhance readability.

Fig. 6 shows the effect of short stack operating temperature at 155 and 175 °C for the case of the fuel cell running on hydrogen and air. As temperature increases, the stack shows an improvement in voltage on the order of 30 mV. This is due to an improvement in electrode kinetics and conductivity. The model captures the effect of temperature in Equation (15) with the dependence of the exchange current density on temperature. For current density levels below 0.05 A/cm², the model shows a higher voltage because it does not take into account the species crossover (usually represented as a leakage current) in the electrolyte. At current densities between 0.2 and 0.4 A/cm², the fuel cell voltages predicted by the model very well match those of the experiments for both temperatures evaluated. In the higher current density regime, a linear dependence on current density is observed, which may be facilitated by more water vapor production, resulting in an

Fig. 5 – Voltage and current density of fuel cell on hydrogen and air at 160 °C. Error bar shows one standard deviation.

Fig. 6 – Polarization curve on hydrogen at 155 and 175 °C.

improvement of the membrane conductivity, as stated by [35,22]. For the conditions experimentally investigated, there appears to be no significant concentration polarization, confirming that our major assumption in this regard is reasonable.

Fig. 7 compares the polarization curve measured when the stack was running on hydrogen and reformat gas at 160 °C. The reformat gas composition is similar to reformat obtained from the steam reforming process with 60% H₂, 15% H₂ O, and 0.8% CO. The polarization curve running on reformat gas with CO concentration of 0.25 and 0.5 is identical to the voltage and current of the stack when running reformat with a CO concentration of 0.8%. The voltage of the stack shows that the PBI MEA can tolerate CO concentrations up to 0.8% at 160 °C with negligible performance losses. This conclusion is consistent with data reported in the literature [36]. When running on reformat gas and hydrogen, the observed fuel cell voltage shows slightly higher voltage than the model, especially in the higher current density regime. A possible explanation for the slightly higher experimental voltages observed in these polarization curves is that the higher anode flow rate

Fig. 7 – Polarization curve on hydrogen and reformat gas at 160 °C.

for the reformat case which is required to maintain the same stoichiometry improves the gas diffusion on the anode side, leading to a very slight voltage improvement.

Dynamic operation performance

Fig. 8 shows one of the dynamic current perturbations that were imposed on the fuel cell stack. In this experiment, the current density instantly increases from 0.09 to 0.18 A/cm², stays at this level for 30 s while the voltage settles, and then instantly decreases current density back to 0.09 A/cm². The data is logged at 10 Hz. To maintain sufficient fuel and airflow to the fuel cell stack during the increase in current, the mass flow controllers increase the flow rate of both reactant streams 10 s before the higher current is drawn from the stack. This preparation leads to an increase in the stoichiometric ratio of the cathode and anode during the times of low current density, as shown in Fig. 9.

The concentrations and flow rate of the fuel are shown in Fig. 10. The concentration of CO is maintained at 1% throughout the dynamic experiment. During the ramp-up of fuel flow to prepare for an increase in current, the flow rate has an unexpectedly large overshoot before settling to the higher flow due to the internal control action that is applied by the test stand to manipulate the mass flow controllers. For the decrease in current, the mass flow controllers reduce the flow rate with no overshoot. To compare the dynamic model with the experimental data, we used the experimentally determined flow rates of fuel and air as inputs to the dynamic simulation.

Fig. 11 shows the measured voltage of all 14 cells within the stack as it responded to the dynamic current and flow perturbations shown in Figs. 8–10. During the time from 40 to 70 s, the voltages of the cells have a large variation from 0.19 to 0.54 V. For the step increase in current at 40 s, the two cells that have the lowest voltage also have the slowest recovery time. Further comparisons amongst the cells regarding their time response characteristics are provided in the later sections of this paper.

Fig. 8 – Current density set point for the fuel cell stack dynamic perturbation.

Fig. 9 – Anode and cathode stoichiometric ratios increase to prepare for the increase in current.

Fig. 11 also shows that the relative voltage responses of the individual cells do not seem to be correlated to the location of the cells within the stack. The cells at the edges may have reasonable voltage, whereas the cells in the middle of the stack may have the lowest response. For instance, the cells with numbers 2 and 14 have the lowest voltages, but the two cells located in the adjacent positions, cells number 1 (edge cell) and number 13, have voltages in the mid-range of all those measured. The three cells with numbers 5, 6, and 10

Fig. 11 – Measured voltages of all the cells for steps in current.

have the highest voltages, but they do not correspond necessarily to any particular position in the stack. Further analysis on the cell material and the installation process are needed to explain the large cell-to-cell variations in voltage that are observed in this fuel cell short stack.

The voltage dynamic response of selected cells in the fuel cell short stack is shown in Fig. 12. The recorded voltage dynamics of the three best performing cells number 5, 6, and 10 are presented here. The dynamic response of the fuel cell consists of three primary parts: (i) fuel and air flow dynamic increases while the current remains constant at 0.09 A/cm^2 , (ii) the instant current increase perturbation to 0.18 A/cm^2 , and (iii) instant current decrease perturbation (back down to 0.09 A/cm^2) with corresponding fuel and air flow decrease. During the first part, at time between 31 and 39 s, because the current is held constant and the fuel and air flow rates are increased, the cell voltage increases from around 0.63 to around 0.66 V for all three cells. The overshoot from the fuel and air flows at 39 s results in a 1 s peak voltage that is observed in the model, but, that is not as apparent in the measured results. The 10 Hz resolution of the experimental measurements is the primary reason that short term dynamic responses are not as apparent in the data. Another contribution to slower observed experimental response could be the manifold volume, which reduces the impact of the flow rate overshoot. Compared to the experimental data, the cell voltage shows an increase in the voltage with a similar trend during this fuel and air flow perturbation.

In the second part of the perturbation, where the current instantly increases from 0.09 to 0.18 A/cm^2 and remains there for 30 s, the measured voltages rapidly decline to 0.55 V. The response consists of a fast response due to conduction and electrochemical reactions, and a slower response due to small changes in temperature that occur during the dynamic perturbation. Both experimental data and model show stable voltage after 10–20 s. The experimental data show an undershoot in voltage to about 0.52 V before settling back up to 0.54 V. This undershoot most likely comes from the transient response of the flow within the cell channels that is not captured in the model. After stabilizing at the new current

Fig. 10 – Composition and flow rate of the anode gas inlet.

Fig. 12 – Comparison of the voltage response for cell 5, 6, and 10 with the model.

level, the voltages of the three cells follow closely the value predicted by the model.

In the third part of the dynamic perturbation shown in Fig. 12, where the current decreases to 0.09 A/cm^2 , the experimental data and model both show that a stable voltage is achieved after 3–7 s. The experimental data show a small overshoot that may come from the hydrogen that is effectively stored within the anode flow channel volume when current demand decreases. The dynamic model shows a response with two parts, similar to those observed in the experiment when the current is perturbed. Throughout the two changes in current, the recorded voltage from the three cells and the model are very similar in magnitude (within 20 mV) and dynamic response characteristics.

Fig. 13 shows the voltage response for cells number 1, 2 and 3, which were observed to be quite different in their response characteristics compared to the fuel cell model calculations. The three cells are adjacent to each other, and located at the edge of the fuel cell stack. At the beginning and end of the experiment, the two cells number 1 and 3 have voltages close to the values calculated by the model. But, when the current increases, the two cells have voltage drops that are much lower than the average recorded voltage and certainly much less than cells number 5, 6, and 10, as shown in Fig. 12. Among these three cells, cell number 2 has the worst performance. At the beginning and end of the experiment, the voltage of cell number 2 is lower than the other two cells by 0.1 V. During the increase in current, the cell voltage reduces to 0.2 V. Even when the current is instantly reduced to the initial level, the cell cannot recover to its own initial voltage level. Because the three cells are adjacent to each other, we suggest that there is not likely a significant difference in temperature and flow distribution that could contribute to these observed differences in cell performance. The observed differences may be attributed the variations in the cell MEAs themselves, the membrane and short stack assembly process, or the manifold distributing the fuel to the cells. As more research is dedicated to the fuel cell stack development, we hope that the variations among the cells will be reduced.

Fig. 13 – Comparison of the voltage response of cell 1, 2, and 3 with the model.

Summary and conclusions

Experimental data obtained from a polybenzimidazol-based fuel cell short stack affirmed that for short-term testing, this high temperature PEM fuel cell technology can tolerate the effect of carbon monoxide up to concentrations of 1%. The operating temperature of the fuel cell stack has a prominent effect on the fuel cell voltage. Our experimental results show an increase of about 20 mV for an increase in temperature from 155 to 175 °C. Dynamic operating data for the 14-cell short stack was acquired for significant fuel and air flow rate and current demand perturbations. In addition, a dynamic model for the fuel cell stack was developed for control systems development that assumes that all the cells within the fuel cell stack respond similarly. In the experiment however, significant variations in dynamic response were observed amongst the cells with a select few of the better performing cells responding in a manner that well verified the dynamic model. Our results indicate that for the cells that get sufficient flow and proper temperature and for cells that are not degraded, the dynamic model can well simulate the dynamic response characteristics in good agreement with the experimentally determined dynamic responses.

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