PREDICTION OF CURRENT DISTRIBUTION IN
A MOLTEN CARBONATE FUEL CELL

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ABSTRACT

A mathematical model has been developed to predict the performance of a molten carbonate fuel cell as a function of anode and cathode gas compositions, gas flow rates, and polarization characteristics. The effect of gas flow modes such as crossflow and coflow and the effect of higher pressures on the current distribution are studied. The predicted polarization curves agree well with the experimentally generated polarization curves. Conditions for incorporating a microscopic porous electrode model into the overall model development are briefly outlined.

The objective of this work was to develop a model capable of predicting the current distribution within the molten carbonate fuel cell as a function of gas compositions, gas flow rates, flow geometries, gas utilizations, and operating pressure. The half-cell reactions occurring in this cell are:

\[
\begin{align*}
\text{Anode:} & \quad \text{H}_2 + \text{CO}_3^2- \rightarrow \text{CO}_2 + \text{H}_2\text{O} + 2\text{e}^- \quad (1) \\
& \quad \text{CO} + \text{CO}_3^2- \rightarrow 2 \text{CO}_2 + 2\text{e}^- \quad (2) \\
\text{Cathode:} & \quad 2\text{e}^- + \text{CO}_2 + \frac{1}{2}\text{O}_2 \rightarrow \text{CO}_3^2- \quad (3)
\end{align*}
\]

The CO in the fuel may react electrochemically as shown above, but at the cell operating temperature of \(650^\circ\text{C}\), the water-gas shift reaction —

\[
\text{CO} + \text{H}_2\text{O} \rightarrow \text{CO}_2 + \text{H}_2 \quad (4)
\]

is more rapid, and the CO conversion to \(\text{H}_2\) contributes more to the current than the electrochemical reaction of CO. Hence, in the various stages of the model development, we have assumed that CO is electrochemically inert, and, therefore, Equation 2, as a direct faradaic reaction, is not considered.

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The model development task was carried out in five stages:

1. Predict current distributions for 100-cm² cells with crossflow configuration (where the fuel and oxidant gases flow perpendicular to each other). The water-gas shift reaction was not considered in this stage.

2. Extend the model to include the shift reaction to study its contribution to the current density.

3. Modify the model for the coflow mode of operation and compare the current distributions with that predicted by the crossflow model, to obtain an insight into the effect of various flow configurations.

4. Incorporate the effects of pressures greater than atmospheric on the current distribution predicted by the crossflow model.

5. Incorporate the microscopic porous electrode model in the crossflow model.

**CROSSFLOW MODEL**

**Linear Polarization Model**

In the 100-cm² cells operating at IGT, fuel and oxidant gases flow perpendicular to one another (crossflow). Figure 1 is a schematic diagram of the gas flows to the cell. The anode is porous nickel, which is 30 mils thick and has a mean pore size of around 5μ. The cathode is also porous nickel which oxidizes to NiO. It is 15 mils thick and has a mean pore size of 12 to 14μ. The electrolyte tile is typically a eutectic mixture of 62% Li₂CO₃—38% K₂CO₃ supported on a LiA1O₂ matrix.

In the initial modeling study, Equations 1 and 3 only were considered; the shift reaction (Equation 4) was neglected. The model was generated by formulating mass balances on the anode and cathode sides, which resulted in one equation each for local conversion at the anode and cathode. These two mass balance equations were solved simultaneously with the cell potential balance equation. The potential balance equation was obtained by combining the calculated Nernst potential (local-equilibrium potential) and the assumed linear dependency of electrode overpotential on current density in such a way that the cell terminal potential was a constant.

**Mass Balance Equations**

Figure 1 shows how the cell is divided into sections for computational purposes. Equations 5 through 11 show the dependency of the fuel and oxidant gas components on the gas flow rate, the length of the anode and cathode gas channels, together with current density.
Mass Balance Equations

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Anode

\[ \text{H}_2: \quad \frac{P}{RT} \frac{d}{dx} \left( \frac{\Phi_A}{n} X_H \right) = -\frac{i_A}{2F} = -\frac{i}{2F} \quad (5) \]

\[ \text{CO}_2: \quad \frac{P}{RT} \frac{d}{dx} \left( \frac{\Phi_A}{n} X_{CO} \right) = \frac{i_A}{2F} = \frac{i}{2F} \quad (6) \]

\[ \text{H}_2O: \quad \frac{P}{RT} \frac{d}{dx} \left( \frac{\Phi_A}{n} X_{W} \right) = \frac{i_A}{2F} = \frac{i}{2F} \quad (7) \]

\[ \text{CO}: \quad \frac{P}{RT} \frac{d}{dx} \left( \frac{\Phi_A}{n} X_{CO} \right) = 0 \quad (8) \]

Cathode

\[ \text{CO}_2: \quad \frac{P}{RT} \frac{d}{dy} \left( \frac{\Phi_C}{m} X_{CC} \right) = \frac{i_C}{2F} = -\frac{i}{2F} \quad (9) \]

\[ \text{O}_2: \quad \frac{P}{RT} \frac{d}{dy} \left( \frac{\Phi_C}{m} X_{O} \right) = \frac{i_C}{4F} = -\frac{i}{4F} \quad (10) \]

\[ \text{N}_2: \quad \frac{P}{RT} \frac{d}{dy} \left( \frac{\Phi_C}{m} X_{N} \right) = 0 \quad (11) \]

Conversion of H\(_2\), \(\lambda\), on the fuel side is defined as

\[ \lambda = \frac{\text{moles of H}_2 \text{ consumed}}{\text{moles of H}_2 \text{ in}} = \frac{\phi_A X_H - \phi_A X_H^o}{\phi_A X_H^o} \quad (12) \]

where the \(^o\) superscript refers to the initial conditions; that is,

\[ \lambda = 1 - f_A X_H \quad (13) \]
where
\[ f_A = \frac{\phi_A}{\phi_X}. \]

Conversion of CO\(_2\), \( \mu \), on the oxidant side is defined as
\[ \mu = \frac{\text{moles of CO}_2 \text{ consumed}}{\text{moles of CO}_2 \text{ in}} = \frac{\phi_C X_{CC} - \phi_C X_{CC}^0}{\phi_C X_{CC}^0} \quad (14) \]

that is,
\[ \mu = 1 - f_C X_{CC}^0 \quad (15) \]

where
\[ f_C = \frac{\phi_C}{\phi_C^0}. \]

Combining Equations 12 and 13 for the hydrogen component of the fuel, we obtain Equation 16. Similarly, Equations 17 and 18 can be derived for the CO\(_2\), H\(_2\)O fuel components.

\[ f_A X_H = X_H^0 - \lambda X_H^0 \quad (16) \]
\[ f_A X_{CA} = X_{CA}^0 + \lambda X_H^0 \quad (17) \]
\[ f_A X_{W} = X_{W}^0 + \lambda X_H^0 \quad (18) \]

If Equations 16, 17, and 18 are added, then
\[ f_A = 1 + \lambda X_H^0. \quad (19) \]

In a similar manner for the oxidant components at the cathode,
\[ f_C X_{CC} = X_{CC}^0 - \mu X_{CC}^0 \quad (20) \]
\[ f_C X_{O} = X_{O}^0 - \frac{1}{2} X_{CC}^0. \quad (21) \]

Adding Equations 20 and 21, then
\[ f_C = 1 - \frac{3}{2} \mu X_{CC}^0. \quad (22) \]

If dimensionless distance \( \xi \) is defined as
\[ \xi = \frac{x}{L_A/m} \quad (23) \]
where $L_A/m$ is the length of one section in the $x$-direction, then substituting Equations 13 and 23 into Equations 5, 6, and 7 results in Equation 24:

$$\frac{P_m}{RT L_C L_A} \phi^0_A \chi^0_H \frac{d\lambda}{d\xi} = \frac{i}{2F}.$$  (24)

The current density can be defined as

$$J = \frac{i}{A} = \frac{i_m}{I_{\text{max}}}.$$  (25)

where $I_{\text{max}}$ is the current density that would result if 100% conversion of the anode gas took place and the resulting current was uniformly distributed over the entire cell. Then,

$$A = \frac{2F \phi_A \chi_H m}{RT L_C L_A} \text{ (A/cm}^2)$$  (26)

and therefore

$$\frac{d\lambda}{d\xi} = J.$$  (27)

If dimensionless distance $\eta_d$ is defined as

$$\eta_d = \frac{\gamma}{L_C/n}$$  (28)

where $L_C/n$ is the length of one section in the $y$-direction, then substituting Equations 15 and 28 into Equations 9 and 10 results in Equation 29:

$$\frac{P_m}{RT L_A L_C} \phi^0_C \chi^0_{\text{CC}} \frac{d\mu}{d\eta_d} = \frac{i}{2F}.$$  (29)

Let $S$ denote the ratio of $\text{CO}_2$ flow rate per cathode section to the $\text{H}_2$ flow rate per anode section; that is,

$$S = \frac{\chi^0_{\text{CC}} \phi^0_C \eta}{\chi^0_H \phi^0_A n}.$$  (30)

Substituting Equations 26 and 29 into Equation 30, we obtain

$$\frac{d\mu}{d\eta_d} = \frac{i}{SA} = \frac{J}{S}.$$  (31)
Equations 27 and 31 are two mass balance equations, containing three unknowns, \( \lambda \), \( \mu \), and \( J \). To solve for these unknowns, a third equation is obtained from a potential balance.

**Potential Balance Equation**

Figure 2 is a schematic representation of the potential gradient in a cell for both the open-circuit condition and when under load. The cell potential is given by

\[
V = (V_{CN} - V_{AN}) - i_\text{ohm} + \eta_C - \eta_A
\]  
(32)

The equilibrium potentials at the electrodes are

\[
V_{AN} = V_A^o + \frac{RT}{2F} \ln \left( \frac{X_H X_{CA}}{X_H} \right)
\]  
(33)

\[
V_{CN} = V_C^o + \frac{RT}{2F} \ln \left( \frac{X_{1/2}^o}{X_{CC}} \right)
\]  
(34)

where \( V_A^o \) and \( V_C^o \) are the standard anode and cathode potentials.

Assuming a linear current-overpotential relationship, then

\[
\eta_A = i_A z_A = i_\text{pA}
\]  
(35)

\[
\eta_C = i_C z_C = -i_\text{pC}
\]  
(36)

where \( z_A \) and \( z_C \) are effective electrode impedances (ohm-cm\(^2\)). Substituting Equations 33 through 36 into Equation 32, we obtain

\[
V + V_A^o - V_C^o = \frac{RT}{2F} \ln \left( X_{1/2}^o X_{CC} \right) - \frac{RT}{2F} \ln \left( \frac{X_H X_{CA}}{X_H} \right) - i_A \left( Z_{\text{ohm}} + Z_A + Z_C \right)
\]  
(37)

Using the definition of conversions \( \lambda \) and \( \mu \) from Equations 13 and 15, Equation 37 can be written as

\[
\frac{V + V_A^o - V_C^o}{RT/2F} = \ln \left[ \frac{X_{CC}}{X_H X_{CA}} \right] \left( \frac{X_{1/2}^o X_H X_{CA}}{X_H X_{1/2}^o X_{CA}} \right)

\]

\[
= \ln \left[ \frac{(1 - \mu) (1 - 1/2\mu \frac{X_{CC}}{X_0})^{1/2} (1 - \lambda) (1 + \lambda X_0^2)}{(1 - 3/2\mu X_{CC})^{3/2} (1 + \lambda \frac{X_H}{X_W}) (1 + \lambda \frac{X_H}{X_{CA}})} \right] - \frac{J_{\text{BA}}}{RT/2F}
\]  
(38)
The first term on the right-hand side of Equation 38 is the Nernst loss, and the second term is the overpotential term. Substituting Equation 26 for A in the overpotential term, we obtain

\[
\frac{JZ_A}{RT/2F} = JZ \frac{P \phi_A X_H^o m}{L_C L_A} \left( \frac{2F}{RT} \right)^2
\]

Rearranging Equation 38, we then have

\[
\ln \left[ \frac{(1 - \mu)(1 - \mu A_1)^{1/2} (1 - \lambda)(1 + \lambda A_2)}{(1 - \lambda A_3)^{3/2} (1 + \lambda A_4)(1 + \lambda A_5)} \right] - U - JE = 0
\]

where

\[
A_1 = 0.5 \frac{X_{CC}^o}{X_0^o}
\]
\[
A_2 = X_H^o
\]
\[
A_3 = 1.5 X_{CC}^o
\]
\[
A_4 = \frac{X_H^o}{X_N^o}
\]
\[
A_5 = \frac{X_H^o}{X_{CA}^o}
\]
\[
U = \frac{V + V_A^o - V_C^o}{RT/2F} - \ln \left[ \frac{(X_{CC}^o)(X_H^o)^{1/2} (X_H^o)}{(X_N^o)(X_{CA}^o)} \right]
\]

and

\[
E = \frac{\bar{Z} m P \phi_A X_H^o}{L_C L_A} \left( \frac{2F}{RT} \right)^2.
\]

Equations 27, 31, and 40 were solved simultaneously to give \( \lambda, \mu, \) and the current density at any point on the electrode surface.

A computer program was written to solve these three equations. The inputs to the computer program were 1) anode and cathode gas compositions and flow rates, 2) standard anode and cathode potentials from the Nernst Equation, 3) terminal voltages chosen from representative fuel cell data, 4) length of anode and cathode gas channels, 5) number of sections in the x and y directions across the cell, 6) temperature, 7) pressure and, 8) effective electrode impedance. The effective electrode impedance (input \( \bar{Z} \) to the computer program) represents the sum of the cell resistance and the individual electrode overpotential terms. The cell resistance was obtained from current interruption measurements on 100-cm\(^2\) cells. Representative electrode overpotentials (\( \eta_A \) and \( \eta_C \)) were obtained from 3-cm\(^2\) cells using various fuel and oxidant compositions at 7.5% H\(_2\) utilization and 15% CO\(_2\) utilization.\(^1\)
Figure 3 compares experimental data obtained from 100-cm$^2$ cells and the predicted curve for a fuel composition obtained from reformed natural gas. The theoretical anode conversion was not found to agree exactly with that obtained experimentally (75%). Hence, $Z$ was adjusted to get the experimentally obtained anode conversion. (Refer to section below, Variation in Effective Electrode Impedance, $Z$.)

Inclusion of CO Shift Reaction in the Linear Polarization Model

In the next stage of the linear model development, hydrogen originating from the water-gas shift reaction (Equation 4) was included and the mass and potential balance equations were correspondingly modified. Figures 4 through 6 show experimental polarization curves for three different fuel gases, together with the predicted curves, both with and without consideration of the shift reaction. For fuels having a low initial CO content and a high H$_2$ content (for example, a feed gas obtained by reforming methane, as in Figure 4), the contribution to the current density by the shift reaction is not significant. However, in the case of fuels with high CO contents (for example, fuels obtained by the partial oxidation of heavy oils or by the gasification of coal), the shift reaction contributes significantly to the current density, as can be seen in Figures 5 and 6. Also, the current distribution along the cell is generally more uniform than when no CO conversion took place.

Variation in Effective Electrode Impedance, $Z$

For fuels with low H$_2$ contents, the current densities predicted with and without inclusion of the shift reaction were not in very good agreement with those values obtained experimentally from 100-cm$^2$ cells (Figures 5 and 6). In order to obtain better agreement with experimental data, we adjusted the effective cell impedance, $Z$, given by

$$Z = R \cdot A + \frac{\eta_A}{i} + \frac{\eta_C}{i}$$  

(41)

where $R$ is the cell resistance, $A$ is the area of the cell, $i$ is the current density and $\eta_A$, $\eta_C$ are the anode and cathode overpotentials. For each inlet gas composition, representative cell resistance values, $R$, used were obtained as discussed previously using current interruption techniques on 100-cm$^2$ cells.

Consequently, the variable component of $Z$ is the respective anode ($\eta_A$) and cathode ($\eta_C$) overpotential terms. Given this value of $Z$, along with the other inputs to the computer program, the model predicts the current density and average anode and cathode conversions. The average anode conversion initially obtained was generally different from the experimental 75% conversion achieved in bench-scale cells. Hence, the value of $Z$ was adjusted to obtain the more representative anode conversion of 75%. The predicted average current densities, based on the new values of $Z$, agree to within 4% of the experimental
current densities. Figures 7 and 8 compare the predicted polarization curves (obtained with the adjusted value of $Z$) with the experimental curve for two feed gases.

In general, it was observed that as the value assigned to $Z$ was increased, the current distribution became more uniform. This is illustrated in Table 1.

From the adjusted value of $Z$, the overpotentials $\eta_A$ and $\eta_C$ are calculated at different loads, from Equation 41. In Figure 9 these are compared with the experimental overpotentials, which were extrapolated from 100-cm$^2$ cell data. The assumption of linear current/overpotential relationship appears valid. However, one cannot draw conclusions about the relative contributions of mass transfer and activation overpotentials from Figure 9 because of the presence of an ohmic drop in the pores. (Refer to the section Incorporation of a Microscopic Porous Electrode Model.)

| $Z$, ohm/cm$^2$ | Ratio of Highest to Lowest Current Density |
|----------------|------------------------------------------|
| 0.6            | 4.8                                      |
| 0.8            | 4.2                                      |
| 1.0            | 3.8                                      |
| 1.2            | 3.4                                      |
| 1.4            | 3.1                                      |
| 1.6            | 2.95                                     |
| 5.0            | 1.7                                      |
| 10.0           | 1.4                                      |

COFLOW MODEL

This corresponds to the situation where both fuel and oxidant gases flow in a direction parallel to each other. Here the current distribution is expected to be less uniform compared with the crossflow configuration. The computer model was modified to predict the current distribution across the 100-cm$^2$ fuel cell using this flow configuration.

Figures 10 and 11 compare the average current distribution along the 10-cm cell length for the crossflow and coflow configuration for two representative feed gases. Particularly high current densities are seen at the cell entrance for the coflow configuration, where both fuel and oxidant are at their highest concentrations. Figures 12 and
compare the predicted polarization curves for the coflow and crossflow configurations. Overall average current densities were predicted. However, the values were found to be only slightly lower for the coflow mode. The crossflow configuration would appear to be preferred to the coflow mode because the predicted current distribution is more uniform.

**EFFECTS OF PRESSURE ON CURRENT DISTRIBUTION IN LINEAR MODEL**

Increasing the operating pressure of the molten carbonate fuel cell can be expected to result in performance improvements (Nernst gains) due to the increase in the partial pressures of the reactants. Higher pressures should also produce increases in gas solubilities and mass transport rates in the selected molten carbonate mixture, which should result in improved electrochemical performance.

Enhanced cell performance at pressures above atmospheric have been confirmed in practice with 100-cm² cells. High pressure favors the carbon deposition and methane reactions:

\[
\begin{align*}
2 \text{CO} & \rightleftharpoons C + \text{CO}_2 \\
\text{CO} + 3\text{H}_2 & \rightleftharpoons \text{CH}_4 + \text{H}_2\text{O}
\end{align*}
\]

Equilibrium inlet compositions were calculated at different pressures, at 650°C, taking into account Equations 42 and 43 and the water-gas shift reaction (Equation 4).

The mass and potential balance equations were therefore rewritten to take into account the conversions of CH₄ and CO, both of which are assumed not electrochemically active compared with H₂. A new computer program was developed to predict the current distribution at higher operating pressures.

Figure 14 compares experimental and predicted polarization curves for steam-reformed naphtha at a pressure of 10 atmospheres. The fuel inlet equilibrium composition at 10 atmospheres and 650°C for steam-reformed naphtha corresponds to 31% H₂, 11.1% CO₂, 43.4% H₂O, 4.1% CO and 10.4% CH₄, compared with the composition at 1 atmosphere and 650°C of 52.7% H₂, 10.4% CO₂, 25.2% H₂O and 10.7% CO. The predicted current densities are found to be in very good agreement with the experimental values. The experimental data used in this comparison were obtained at constant flows rather than at constant conversion.

**INCORPORATION OF A MICROSCOPIC POROUS ELECTRODE MODEL**

The development of a porous electrode model is necessary for the integration of electrode kinetic data and the mass transfer and ohmic resistances in the crossflow performance model.
In the work discussed above, a constant effective cell resistance (impedance), $Z$, has been used. This is equivalent to assuming that under probable conditions present at the fuel cell electrodes, to a first approximation the individual electrode overpotentials varied linearly with current density. Such an assumption is reasonable if the local current density inside a pore is small compared to the exchange current density and compared to the limiting current density. It has been estimated earlier that the limiting current densities at the anode and at the cathode are at least 1 A/cm$^2$ or larger.

The majority of work to date on the oxygen-reduction reaction in molten carbonate mixtures has been performed on gold electrodes, where estimated $i_0$ values can be calculated to be 0.01 mA/cm$^2$ for typical oxidant conditions.

As discussed previously, the local current density within the porous electrode is probably low enough to allow that the linear polarization model be used as a good first approximation for predicting the current distribution. However, the porous electrode model should also take into consideration ionic resistivity along the assumed thin film of electrolyte within the pore, and diffusional resistance because of the transport of dissolved reactants across the film. These resistances may cause an apparent increase in the overpotential (between the electrode terminal and the reference electrode), over and above that expected from kinetic relationships.

If the transport of reactants in the gas phase is not rate-limiting, as has been previously shown, then ohmic resistances, kinetics, and mass transfer remain as possible rate-limiting factors. Their relative significance is indicated by the dimensionless parameters —

$$A_1 = \frac{\kappa \delta RT}{I F L^2}$$  \hspace{1cm} \text{kinetic resistance} \hspace{1cm} (44)$$
$$A_2 = \frac{\delta^2 \kappa RT}{D C \eta F^2 L^2}$$  \hspace{1cm} \text{mass transfer resistance} \hspace{1cm} (45)$$
$$A_3 = A_2 / A_1$$  \hspace{1cm} \text{mass transfer resistance} \hspace{1cm} (46)$$

Table 2 presents typical values of these parameters for a 100-cm$^2$ fuel cell.* Table 3 presents the structural data of the electrodes.

* Basis — Fuel composition: 35.16% H$_2$, 23.5% CO$_2$, 34% H$_2$O, 7.25% CO (average of inlet and outlet compositions); Oxidant composition: 60.8% N$_2$, 11.96% O$_2$, 23.9% CO$_2$, 3.4% H$_2$O (average of inlet and outlet compositions). Values of $D_0$, $C_0$, and $\kappa$ are from Reference 4. $i_0$ values of 40 mA/cm$^2$ for the hydrogen reaction and 1 mA/cm$^2$ for the oxygen reaction were used in estimating these parameters.

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This porous electrode model has been integrated with our cross-flow configuration. Preliminary predicted polarization curves from this model, however, require that the exchange current density for the oxygen-reduction reaction on nickel oxide be significantly higher than 0.01 mA/cm$^2$ (as previously reported on a gold electrode).\textsuperscript{5}

Table 2. CHARACTERISTIC DIMENSIONLESS RESISTANCE RATIOS IN A 100-cm$^2$ FUEL CELL

|             | Anode | Cathode |
|-------------|-------|---------|
| $A_1 = \frac{\text{kinetic}}{\text{ohmic}}$ | 0.021 | 3.39    |
| $A_2 = \frac{\text{mass transfer}}{\text{ohmic}}$ | 0.005 | 1.86    |
| $A_3 = \frac{\text{mass transfer}}{\text{kinetic}}$ | 0.24  | 0.55    |

Table 3. STRUCTURAL DATA OF ELECTRODES

|               | Anode | Cathode |
|---------------|-------|---------|
| Pore Radius   | 2.5 $\mu$m | 6.5 $\mu$m |
| Pore Length   | 30 mils | 15 mils |
| Porosity      | 70%    | 80%     |
| Pore Density  | $3.57 \times 10^6$ cm$^2$ | $6.03 \times 10^5$ cm$^2$ |
| Internal/External Area | 427   | 94      |
| Film Thickness| 0.5 $\mu$m | 0.5 $\mu$m |

Assuming a constant exchange current density for both the hydrogen oxidation (40 mA/cm$^2$) and oxygen-reduction reactions along the cell length, preliminary calculations have indicated that $i_0$ values of around 2 mA/cm$^2$ for the latter reaction were necessary to achieve a satisfactory fit with experimental polarization curves.

Refinement of this porous electrode model will include the concentration dependencies of $i_0$ for both the anodic and cathodic reactions, as determined from our electrochemical half-cell data, and will be discussed at a later time.

CONCLUSIONS

The current distribution in 100-cm$^2$ cells with crossflow of fuel and oxidant gases has been modeled assuming linear polarization; the predicted polarization curves are in good agreement with the experimental curves.
The current distribution is more uniform in the crossflow configuration than the coflow mode, although the average current densities of the crossflow and coflow modes differ only slightly.

At higher operating pressures, the methane-reaction equilibrium needs to be considered. The current densities thus predicted agree very well with the experimental values obtained using a 100-cm² cell operated under pressure.

NOMENCLATURE

- $\dot{V}_A$ = total fuel flow rate, cm³/s
- $\dot{V}_C$ = total oxidant flow rate, cm³/s
- $X_k$ = mole fraction of the gas component $k$
- $x$ = distance in the fuel flow direction, cm
- $y$ = distance in the oxidant flow direction, cm
- $i_A$ = anodic current density ($>0$), mA/cm²
- $i_C$ = cathodic current density ($<0$), mA/cm²
- $i$ = current density [$i = i_A = |i_C|$], mA/cm²
- $L_A$ = total length of the fuel gas channel, cm
- $L_C$ = total length of the oxidant gas channel, cm
- $m$ = number of sections in the anode flow direction
- $n$ = number of sections in the cathode flow direction
- $T$ = temperature, K
- $P$ = pressure, atm
- $R$ = gas constant, 82.06 atm·cm³/g-mole·K
- $F$ = faraday constant
- $V$ = terminal potential
- $V_{CN}$ = equilibrium (Nernst) potential at the cathode
- $V_{AN}$ = equilibrium (Nernst) potential at the anode
- $i$ = current density, mA/cm²
- $Z_{ohm}$ = ohmic cell resistance, ohm·cm²
- $iZ_{ohm}$ = ohmic potential drop in cell
- $\eta_C$ = cathodic overpotential ($<0$)
- $\eta_A$ = anodic overpotential ($>0$)
- $\kappa$ = ionic conductivity
δ = film thickness
i₀ = exchange current density
L = pore length
D = diffusivity of gas in electrolyte
C = solubility of gas in electrolyte

Subscripts
A = anode
C = cathode
H = hydrogen
CA = CO₂ in anode gas
CC = CO₂ in cathode gas
W = water
CO = carbon monoxide
O = oxygen
N = nitrogen.

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Figure 1. SCHEMATIC DIAGRAM OF THE FUEL AND OXIDANT FLOWS TO THE CELL

Figure 2. SCHEMATIC REPRESENTATION OF THE POTENTIAL GRADIENT IN A CELL
Figure 3. COMPARISON OF THE EXPERIMENTAL AND THEORETICAL POLARIZATION CURVES FOR A CELL USING REFORMED NATURAL GAS AS THE FUEL.

Figure 4. PREDICTED POLARIZATION CURVES FOR REFORMED NATURAL GAS (With and Without the CO Shift Reaction) COMPARED WITH EXPERIMENTAL DATA.
Figure 5. PREDICTED POLARIZATION CURVES FOR PARTIAL OXIDATION OF HEAVY OILS (With and Without the CO Shift Reaction) COMPARED WITH EXPERIMENTAL DATA

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Figure 10. VARIATION IN CURRENT DENSITY WITH CELL LENGTH FOR REFORMED NATURAL GAS (With and Without the CO Shift Reaction) FOR THE COFLOW AND CROSSFLOW CONFIGURATIONS
Figure 11. VARIATION IN CURRENT DENSITY WITH CELL LENGTH FOR LOW-Btu COAL GAS (With the CO Shift Reaction) FOR THE COFLOW AND CROSSFLOW CONFIGURATIONS

Figure 12. PREDICTED POLARIZATION CURVES FOR REFORMED NATURAL GAS USING COFLOW AND CROSSFLOW CELL CONFIGURATIONS (CO Shift Reaction Not Considered)
Figure 13. PREDICTED POLARIZATION CURVES FOR LOW-\textbf{Btu} COAL GAS USING COFLOW AND CROSSFLOW CELL CONFIGURATIONS

Figure 14. COMPARISON OF EXPERIMENTAL AND PREDICTED POLARIZATION CURVES FOR REFORMED NAPHTHA AT 10 atm PRESSURE