Use of different renewable fuels in a steam reformer integrated into a solid oxide fuel cell: Theoretical analysis and performance comparison

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Article info
Article history:
Received 27 August 2012
Received in revised form 7 December 2012
Accepted 17 December 2012
Available online 21 January 2013

Keywords:
Renewable fuels
Steam reformer
Hydrogen production
Solid oxide fuel cell
Integrated system

Abstract
Hydrogen production from renewable energy resources has received significant attention with advances in fuel cell technology. The fuel type and operational reforming conditions directly affect fuel cell electricity generation. This study analyzes the theoretical performance of a solid oxide fuel cell (SOFC) integrated with a steam reforming process using three different renewable fuels: ethanol, glycerol and biogas. The effects of key steam reformer operating parameters on the hydrogen production for SOFCs are investigated. The performances of SOFC systems run on different fuels are compared in terms of electrical and thermal efficiencies. It is found that the biogas-fueled SOFC system requires the most energy, whereas the ethanol-fueled SOFC system achieves the highest electrical and thermal efficiencies.

1. Introduction

Fuel cells can generate electricity via an electrochemical reaction using hydrogen as fuel and oxygen as an oxidant. Thus, hydrogen production technology has developed in parallel with advances in fuel cell technology. Traditionally, hydrogen has been produced by reforming methane derived from natural gas, which mostly comes from fossil resources. As fossil fuel is in limited supply and causes environmental problems, hydrogen production from renewable and environmentally friendly fuels should be explored.

Among such renewable fuels, ethanol, biogas and glycerol have received considerable attention [1-4]. Ethanol is one of the most attractive raw materials for use in fuel cell hydrogen production because of its non-toxicity and liquid form. Many studies have been devoted to hydrogen production from ethanol [5-9]. Sun et al. [10] studied hydrogen production from ethanol using steam reforming, autothermal reforming and partial oxidation. They found that ethanol steam reforming provides the highest hydrogen yield. A steam-to-ethanol ratio above 6 and temperatures greater than 900 K produce a high hydrogen yield with low carbon monoxide (CO) and carbon formation in the steam reforming environment.

Biogas is also a promising fuel, especially for developing countries. Biogas can be produced by the anaerobic digestion of animal and human waste, agricultural residues, aquatic weeds and other organic matters and mainly consists of methane (CH₄) and carbon dioxide (CO₂), containing CH₄ up to 60 vol.% [11]. To date, use of biogas for hydrogen production process has been widely studied; however, works on the integration of a biogas-based fuel system and a high-temperature fuel cell, such as molten carbonate fuel cells (MCFCs) and solid oxide fuel cells (SOFCs), are quite limited.

Glycerol is another potential fuel for hydrogen production. It is a byproduct of the production of biodiesel, whose consumption is continually increasing. Generally, crude glycerol always contains impurities; however, the purifying process for crude glycerol has a high operational cost and is uneconomic [12]. The utilization of glycerol for hydrogen production is considered promising [13-20].
The use of an expensive catalyst can be avoided by the use of high temperatures, which showed that, in addition to the traditional use of natural gas, reforming agent when biogas is used as fuel. Douvatidze et al. [26] showed that, in addition to the traditional use of natural gas, ethanol is an attractive fuel for SOFC operation. Tsiakaras et al. [27] revealed that the ethanol-fueled SOFCs exhibit the highest theoretical efficiency when operated at 800–1200 K and that the ethanol steam reforming is the most attractive fuel processing system. Farhad et al. [28] studied biogas-fueled SOFC systems and observed that the best system thermal and electrical efficiencies can be achieved when the heat generated by an afterburner is used to supply other heat-requiring units in the SOFC system. Because the thermal and electrical efficiencies of SOFC systems depend on the fuel type, the thermal management and performance of SOFC systems using different fuels should be studied in detail to select the best fuel for different SOFC applications.

The aim of this study is to theoretically analyze a SOFC system integrated with a steam reforming process and compare several important renewable resources for hydrogen production: biogas, ethanol and glycerol. The effect of operating conditions on the reformer performance is investigated in terms of hydrogen yield, product distribution and carbon formation. The energy requirements of each unit in the SOFC system fed by different fuels revealed that the ethanol-fueled SOFCs exhibit the highest theoretical efficiency when operated at 800–1200 K and that the ethanol steam reforming is the most attractive fuel processing system. Farhad et al. [28] studied biogas-fueled SOFC systems and observed that the best system thermal and electrical efficiencies can be achieved when the heat generated by an afterburner is used to supply other heat-requiring units in the SOFC system. Because the thermal and electrical efficiencies of SOFC systems depend on the fuel type, the thermal management and performance of SOFC systems using different fuels should be studied in detail to select the best fuel for different SOFC applications.

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are also examined. In addition, the electrical efficiency, thermal efficiency and cell-stack efficiency of the SOFC integrated system for power generation is investigated. Finally, the amount of CO₂ released from the SOFC system using different fuels is investigated to address environmental concerns.

2. Fuel processor and SOFC integrated system

Fig. 1 shows the external reformer and SOFC integrated system, which comprises heat exchangers, fuel processor, SOFC and afterburner. Steam and fuel are mixed, then preheated at operating reforming temperatures and sent to the steam reformer, where a synthesis gas (hydrogen-rich gas) is produced. Next, the synthesis gas is preheated at the desired temperature and then fed to the SOFC. The SOFC produces electrical power and steam via the electrochemical reaction of hydrogen and oxygen in air. In general, SOFCs cannot be operated at complete fuel utilization; thus, the residual fuel is combusted in an afterburner to generate heat for use in other heat-requiring units in the SOFC system, such as a reformer and fuel and air pre-heaters (dash line in Fig. 1).

In this study, biogas, ethanol and glycerol are considered the potential fuels to produce hydrogen for SOFC systems. Biogas is composed of 60 mol% methane and 40 mol% carbon dioxide. To analyze the theoretical performance of the SOFC system, it is assumed that the system is run under steady-state conditions and that all gases behave as ideal gases. Furthermore, heat losses from individual components in the SOFC system are negligible and the operating pressure and temperature of the reformer and the SOFC are constant. The considered SOFC is planar and described by a one-dimensional model, which is operated at a constant cell voltage along the cell coordinate.

2.1. Fuel processor

Hydrocarbon fuels are reformed to produce a synthesis gas via a steam reforming reaction. The final composition of the produced synthesis gas under equilibrium conditions is determined from the minimization of the Gibbs free energy. The main products of each fuel processing system are hydrogen, methane, carbon dioxide, carbon monoxide and water. Their equilibrium compositions depend on the operating temperature and pressure of the steam reformer. Eq. (1) gives the total Gibbs free energy of the system [29]:

\[
G = \sum_i \eta_i G_i^0 + \sum_i \eta_i T \sum_j \ln \frac{\nu_j \phi_j P_j}{P} \quad \text{(1)}
\]

where \(G\) is the total Gibbs free energy, \(\eta_i\) is the number of moles of species \(i\), \(G_i^0\) is the standard Gibbs free energy of species \(i\), \(\nu_i\) is the gas constant, \(T\) is the reforming temperature, \(P_i\) is the operating pressure of the reformer, \(\phi_i\) is the mole fraction of species \(i\) and \(\phi_i\) is the fugacity coefficient of a gas mixture (\(\phi_i = 1\) for ideal gas).

The equilibrium compositions obtained by minimizing the Gibbs free energy must satisfy the following constraints:

\[
\sum_i a_{ij} n_i = A_j; \quad j = 1, \ldots, k \quad \text{(2)}
\]

where \(a_{ij}\) is the number of atoms of element \(j\) in species \(i\) and \(A_j\) is the total number of atoms of element \(j\) in the feed stream. The Lagrange multiplier method is applied to find the product composition of the reactive system [30].

The overall steam reforming reactions of ethanol, glycerol and biogas are represented by Eqs. (3)–(5), respectively.

\[
\begin{align*}
C_2H_5OH + 3H_2O & \leftrightarrow 2CO_2 + 6H_2 \quad \text{(3)} \\
C_3H_6O_3 + 3H_2O & \leftrightarrow 3CO_2 + 7H_2 \quad \text{(4)} \\
CH_4 + 2H_2O & \leftrightarrow CO_2 + 4H_2 \quad \text{(5)}
\end{align*}
\]

Regarding carbon formation in the reforming system, a stoichiometric approach is applied to determine the boundary of the carbon formation. The reactions that are most likely to lead to carbon formation in the reforming system with hydrocarbon fuels are as follows:

\[
\begin{align*}
2CO & \leftrightarrow CO_2 + C \quad \text{(6)} \\
CH_4 & \leftrightarrow 2H_2 + C \quad \text{(7)} \\
CO + H_2 & \leftrightarrow H_2O + C \quad \text{(8)} \\
CO_2 + 2H_2 & \leftrightarrow 2H_2O + C \quad \text{(9)}
\end{align*}
\]

From a thermodynamic viewpoint, the Boudouard reaction (Eq. (6)) is considered the major carbon generation reaction because it has the lowest Gibbs free energy. The possibility of carbon formation can be calculated from the value of carbon activity [31], defined as

\[
a_c = \frac{K_{BR} x_{CO}^2 P_t}{X_{CO_2}} \quad \text{(10)}
\]

where \(a_c\) is the activity coefficient of carbon and \(K_{BR}\) represents the equilibrium constant of the Boudouard reaction. For carbon activities greater than unity, the system is not in equilibrium and carbon formation occurs. When the carbon activity equals unity, the system is in equilibrium. Finally, for carbon activities less than unity, carbon formation is thermodynamically impossible. Note that the carbon activity is only an indicator of the presence of carbon in a system and does not indicate the quantity of carbon generated.

The heat required for the operation of the steam reformer under isothermal conditions can be computed from the following energy balance equation:

\[
Q_{SR} = \left( \sum_i n_{i,x}^{out} h_{i,x}^{out} \right) - \left( \sum_i n_{i,x}^{in} h_{i,x}^{in} \right) \quad \text{(11)}
\]

where \(n_{i,x}\) is the molar flow rate of species \(i\) and \(h_{i,x}\) is the enthalpy of species \(i\).
The enthalpy can be calculated by:

$$h = h^0 + \int_{298}^{T_{\text{act}}} C_p dT$$

(12)

where $h^0$ is the enthalpy at the standard condition [29].

To evaluate the steam reforming performance of the fuel, the hydrogen yield is defined as

$$H_2 \text{ yield (Ethanol)} = \frac{n_{\text{out}}^{H_2,f}}{6(n_{\text{C}_2H_5OH,f})}$$

(13)

$$H_2 \text{ yield (Glycerol)} = \frac{n_{\text{out}}^{H_2,f}}{7(n_{\text{C}_1H_2O_5,f})}$$

(14)

$$H_2 \text{ yield (Biogas)} = \frac{n_{\text{out}}^{H_2,f}}{4(n_{\text{C}_1H_s,f})}$$

(15)

The factors in the denominator of Eqs. (13)–(15) depend on the number of moles of hydrogen obtained from the steam reforming reactions of ethanol, glycerol and biogas as shown in Eqs. (3)–(5).

2.2. SOFC

Reformate gas consisting of CH₄, H₂O, CO, H₂, and CO₂ is fed into the fuel channel of the SOFC, and air consisting of O₂ and N₂ is fed into the air channel. In general, hydrocarbon fuels can be further converted into hydrogen and carbon dioxide by steam reforming and water gas shift reactions (Eqs. (16) and (17)) within the SOFC because of its high temperatures and the use of an anode catalyst to accelerate these reactions.

Steam reforming reaction:

$$\text{CH}_4 + \text{H}_2\text{O} \rightarrow 3\text{H}_2 + \text{CO}$$

(16)

Water gas shift reaction:

$$\text{CO} + \text{H}_2\text{O} \rightarrow \text{H}_2 + \text{CO}_2$$

(17)

In SOFC operation, hydrogen is consumed in an electrochemical reaction to produce electricity, whereas the oxygen in air is reduced into oxygen ions at the cathode side. The electrochemical reactions occurring at the SOFC fuel and air channels are given in Eqs. (18)–(20). The kinetic equations of all reactions occurring within the SOFC are given in Table 1 [32,33].

Hydrogen oxidation reaction:

$$H_2 + O^{2-} \rightarrow H_2O + 2e^-$$

(18)

Oxygen reduction reaction:

$$1/2\text{O}_2 + 2e^- \rightarrow O^{2-}$$

(19)

Overall cell reaction:

$$H_2 + 1/2\text{O}_2 \rightarrow H_2O$$

(20)

Table 1

<table>
<thead>
<tr>
<th>Reaction Type</th>
<th>Rate Expression</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam reforming reaction</td>
<td>$R_{\text{SR}} = \kappa_{\text{SR}}P_{\text{CH}<em>4,f}\exp\left(-\frac{E</em>{\text{SR}}}{RT}\right)$</td>
</tr>
<tr>
<td>Water gas shift reaction</td>
<td>$R_{\text{WGSR}} = \kappa_{\text{WGSR}}P_{\text{CO}<em>2,f}\left(1 - \frac{P</em>{\text{H}<em>2O,f}}{P</em>{\text{H}<em>2O}\kappa</em>{\text{WGSR}}P_{\text{CO}_2,f}}\right)$</td>
</tr>
<tr>
<td>Electrochemical reaction</td>
<td>$R_{\text{elec}} = j \frac{F}{P_{\text{H}_2,f}}$</td>
</tr>
</tbody>
</table>

The component concentrations in fuel and air channels along the flow direction can be described by the following equations:

Fuel channel: ($i = \text{CH}_4, \text{H}_2\text{O}, \text{CO}, \text{H}_2$ and $\text{CO}_2$)

$$\frac{dC_{i,f}}{dx} = \frac{1}{u_i} \kappa_{ki} \sum \nu_{i,k} R_k \frac{1}{h_i}$$

(21)

$$C_{i,f}|_{x=0} = C_{i,f}^0$$

Air channel: ($i = \text{O}_2$ and $\text{N}_2$)

$$\frac{dC_{i,a}}{dx} = \frac{1}{u_i} \nu_{i,v} R_{v,i} \frac{1}{h_a}$$

(22)

$$C_{i,a}|_{x=0} = C_{i,a}^0$$

The theoretical open-circuit potential can be expressed by the Nernst equation:

$$E_{\text{OCV}} = E^0 - \frac{RT}{2F} \ln \left(\frac{P_{\text{H}_2O}}{P_{\text{H}_2}P_{\text{O}_2}}\right)$$

(23)

where $E^0$, the open-circuit voltage at standard pressure, is a function of the operating temperature:

$$E^0 = 1.253 - 2.4516 \times 10^{-4}T$$

(24)

Due to internal voltage losses, the actual voltage ($V$) is lower than the open-circuit voltage (Eq. (25)).

$$V = E - \eta_{\text{ohmic}} - \eta_{\text{conc}} - \eta_{\text{act}}$$

(25)

where $\eta_{\text{ohmic}}, \eta_{\text{conc}}, \eta_{\text{act}}$ are the ohmic, concentration, and activation losses, respectively.

The ohmic losses can be explained by the linear relation between the voltage drop and current density:

$$\eta_{\text{ohm}} = j R_{\text{ohm}}$$

(26)

where $R_{\text{ohm}}$ is the internal electrical resistance, which depends on the conductivity and thickness of individual fuel cell layers as follows:

$$R_{\text{ohm}} = \frac{\tau_{\text{anode}}}{\sigma_{\text{anode}}} + \frac{\tau_{\text{electrolyte}}}{\sigma_{\text{electrolyte}}} + \frac{\tau_{\text{cathode}}}{\sigma_{\text{cathode}}}$$

(27)

where $\tau_{\text{anode}}, \tau_{\text{cathode}}$ and $\tau_{\text{electrolyte}}$ are the thickness of the anode, cathode and electrolyte layers, respectively. $\sigma_{\text{anode}}$ and $\sigma_{\text{cathode}}$ are the electronic conductivity of the anode and cathode, respectively, and $\sigma_{\text{electrolyte}}$ is the ionic conductivity of the electrolyte.

The concentration overpotentials are due to insufficient hydrogen at the reaction sites [34,35], as described by Eq. (28):

$$\eta_{\text{conc}} = \eta_{\text{conc, anode}} + \eta_{\text{conc, cathode}}$$

$$= \left(\frac{RT}{2F} \ln \left(\frac{P_{\text{H}_2O,TPB}P_{\text{H}_2,f}}{P_{\text{H}_2,O,TPB}}\right)\right)_{\text{anode}} + \left(\frac{RT}{4F} \ln \left(\frac{P_{\text{O}_2,a}}{P_{\text{O}_2,TPB}}\right)\right)_{\text{cathode}}$$

(28)

The partial $\text{H}_2$, $\text{H}_2\text{O}$ and $\text{O}_2$ pressures at the three-phase boundaries can be determined using a gas transport model in porous media [34], as given by

$$P_{\text{H}_2,TPB} = P_{\text{H}_2,f} - \frac{\alpha_{\text{anode}}}{2FD_{\text{eff, anode}}}$$

(29)
\[ p_{H_2OTPB} = p_{H_2O} + \frac{RT \tau_{anode}}{2FD_{eff, anode}} \]  \hfill (30)

\[ p_{O_2TPB} = P - \left( P - p_{O_2,a} \right) \exp \left( \frac{RT \tau_{cathode}}{4FD_{eff, cathode}} \right) \]  \hfill (31)

where \( D_{eff, anode} \) is the effective gaseous diffusivity through the anode (considering a binary gas mixture of \( H_2 \) and \( H_2O \)) and \( D_{eff, cathode} \) is the effective oxygen diffusivity through the cathode (binary gas mixture of \( O_2 \) and \( N_2 \)).

The activation overpotentials can be determined from the nonlinear Butler—Volmer equation, which relates the current density to the activation overpotential:

\[ j = j_{0, anode} \left[ \frac{p_{H_2OTPB}}{p_{H_2O}} \exp \left( \frac{\alpha nF}{RT} \eta_{act, anode} \right) - \frac{p_{H_2O}}{p_{H_2O}} \exp \left( - \left( 1 - \alpha \right) nF \right. \eta_{act, anode} \right) \]  \hfill (32)

\[ j = j_{0, cathode} \left[ \exp \left( \frac{\alpha nF}{RT} \eta_{act, cathode} \right) - \exp \left( - \left( 1 - \alpha \right) nF \right. \eta_{act, cathode} \right) \]  \hfill (33)

where \( \alpha \) is the transfer coefficient (0.5), \( n \) is the number of electrons transferred in the single elementary rate-limiting reaction step, and \( j_{0, anode} \) and \( j_{0, cathode} \) are the exchange current densities at the anode and cathode (Eqs. (34) and (35)), respectively.

\[ j_{0, cathode} = \frac{RT}{nF} k_{cathode} \exp \left( - \frac{E_{cathode}}{RT} \right) \]  \hfill (34)

\[ j_{0, anode} = \frac{RT}{nF} k_{anode} \exp \left( - \frac{E_{anode}}{RT} \right) \]  \hfill (35)

where \( E_{cathode} \) and \( E_{anode} \) represent the activation energies of the cathode and anode exchange current densities and are 137 and 140 kJ mol\(^{-1}\), respectively. \( k_{cathode} \) and \( k_{anode} \) represent the pre-exponential factors and are 2.35 \times 10^{11} and 6.54 \times 10^{11} s\(^{-1}\) m\(^{-2}\), respectively [35].

The average current density \( \langle j_{ave} \rangle \) can be computed from the current density distribution along the cell length, which is obtained by solving the fuel cell model:

\[ \langle j_{ave} \rangle = 1 \int_0^L j(z)\,dz \]  \hfill (36)

To evaluate the performance of the SOFC, the overall fuel utilization (\( U_f \)) and the electrical power output (\( P_{sofc} \)) are determined by

\[ U_f = \frac{j_{ave}LW}{2F(4n_{in}^{\text{CH}_4} + n_{in}^{\text{H}_2} + n_{in}^{\text{CO}})} \]  \hfill (37)

\[ P_{sofc} = j_{ave}VA_{c} \]  \hfill (38)

where \( n_{in}^{\text{species}} \) is the inlet molar flow rate of gas species \( i \) and \( V \) represents the actual potentials.

As the SOFC operation is operated under isothermal conditions, excess air is fed into the SOFC to maintain the operating temperature at a desired value. The difference in the inlet and outlet fuel cell temperatures is limited to 100 K [36,37]. As a result, the amount of the inlet air can be calculated from the energy balance around a fuel cell control volume as

\[ \left( \sum_i n_{in}^{\text{species}} h_i^{\text{in}} \right) + \left( \sum_i n_{in}^{\text{species}} h_i^{\text{in}} \right) - \left( \sum_i n_{out}^{\text{species}} h_i^{\text{out}} \right) - P_{sofc} = 0 \]  \hfill (39)

where \( n_i \) is the inlet and outlet molar flow rates of species \( i \) through the SOFC and \( h_i \) is the inlet and outlet enthalpies of formation of species \( i \) which is a function of temperature.

The performance of the SOFC system is considered in terms of the stack electrical efficiency, the system electrical efficiency (\( \eta_{el} \)) and the thermal efficiency (\( \eta_{th} \)), which are defined as

| Table 3 | Value of operating conditions and parameters used for model validation. |
|-----------------|-------------------------|-----------------|
| Cell length, \( L \) (m) | 0.1 | 
| Cell width, \( W \) (m) | 0.1 | 
| Fuel channel height, \( h_f \) (mm) | 1 | 
| Air channel height, \( h_a \) (mm) | 1 | 
| Anode thickness, \( T_{\text{anode}} \) (\( \mu \text{m} \)) | 500 | 
| Cathode thickness, \( T_{\text{cathode}} \) (\( \mu \text{m} \)) | 20 | 
| Electrolyte thickness, \( T_{\text{electrolyte}} \) (\( \mu \text{m} \)) | 1023 | 
| Inlet fuel and air temperature, \( T_{\text{inlet, fuel}} \) (K) | 1 | 
| Operating pressure, \( P_{\text{sofc}} \) (bar) | 21% \( O_2, 79% N_2 \) | 
| Fuel composition | 33% \( CH_4, 67% H_2O \) | 
| Fuel utilization | 0.8 | 
| Operating voltage (V) | 0.6582 | 

| Table 4 | Validation of the SOFC model. |
|-----------------|-------------------------|-----------------|
| Kanga et al. [38] | Model | Error (%) |
| **Gas composition at the outlet of the fuel channel (mole fraction)** | | |
| Methane | 0.0001 | 0.0001 | 0 | |
| Water | 0.6772 | 0.6871 | 1.46 | |
| Hydrogen | 0.0409 | 0.1215 | 19.4 | |
| Carbon monoxide | 0.0349 | 0.0336 | 3.72 | |
| Carbon dioxide | 0.1639 | 0.1577 | 3.78 | |
| **Gas composition at the outlet of the air channel (mole fraction)** | | |
| Oxygen | 0.1885 | 0.1885 | 0.69 | |
| Nitrogen | 0.8128 | 0.8115 | 0.16 | |
| **SOFC performance parameters** | | |
| Air ratio | 7.5 | 7.5670 | 0.89 | |
| Average current density (A cm\(^{-2}\)) | 0.5 | 0.4988 | 24 | |
| Power density (W cm\(^{-2}\)) | 0.3291 | 0.3283 | 0.24 | |
Table 5
Values of operating conditions for SOFC system under nominal conditions.

<table>
<thead>
<tr>
<th>Component</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Reformer</strong></td>
<td></td>
</tr>
<tr>
<td>Operating temperature (K)</td>
<td>1073</td>
</tr>
<tr>
<td>Operating pressure (bar)</td>
<td>1</td>
</tr>
<tr>
<td>S/C</td>
<td>1.5</td>
</tr>
<tr>
<td><strong>Solid oxide fuel cell</strong></td>
<td></td>
</tr>
<tr>
<td>Cell operating temperature (K)</td>
<td>1173</td>
</tr>
<tr>
<td>Cell operating pressure (bar)</td>
<td>1</td>
</tr>
<tr>
<td>Net power output (kW)</td>
<td>150</td>
</tr>
<tr>
<td>Air composition</td>
<td>21% O₂, 79% N₂</td>
</tr>
<tr>
<td>Cell fuel utilization factor (%)</td>
<td>0.7</td>
</tr>
<tr>
<td>Fuel cell active area (m²)</td>
<td>55.2</td>
</tr>
<tr>
<td>Anode thickness, ( t_{\text{anode}} ) (μm)</td>
<td>500</td>
</tr>
<tr>
<td>Cathode thickness, ( t_{\text{cathode}} ) (μm)</td>
<td>50</td>
</tr>
<tr>
<td>Electrolyte thickness, ( t_{\text{electrolyte}} ) (μm)</td>
<td>20</td>
</tr>
<tr>
<td>Fuel channel height, ( h_{\text{f}} ) (mm)</td>
<td>1</td>
</tr>
<tr>
<td>Air channel height, ( h_{\text{a}} ) (mm)</td>
<td>1</td>
</tr>
<tr>
<td>DC-AC inverter efficiency</td>
<td>96</td>
</tr>
<tr>
<td><strong>Afterburner</strong></td>
<td></td>
</tr>
<tr>
<td>Afterburner combustion efficiency (%)</td>
<td>98</td>
</tr>
</tbody>
</table>

\[ \eta_{\text{el,sofc}} = \frac{P_{\text{sofc}}}{n_{H_2}\text{LHV}_H} + n_{CO}\text{LHV}_{CO} + n_{CH_4}\text{LHV}_{CH_4} \] (40)

\[ \eta_{\text{el}} = \frac{P_{\text{sofc}}}{n_i\text{LHV}_i} \] (41)

\[ \eta_{\text{th}} = \frac{Q_{\text{rec}} - Q_{\text{use}}}{n_i\text{LHV}_i} \] (42)

where \( n_i \) is the inlet molar flow rate of component \( i \), \( \text{LHV}_i \) is the lower heating value of component \( i \), \( Q_{\text{rec}} \) is the total thermal energy used in the SOFC system and \( Q_{\text{use}} \) is the thermal energy obtained from the afterburner. Note that the reference temperature of 100 °C is given to determine an amount of heat that can be recovered from the afterburner. The thermal efficiency defined as Eq. (42) indicates the maximum theoretical thermal efficiency that the SOFC system can achieve and no heat loss is assumed.

All of the system model equations describing the SOFC system are coded and solved using Matlab. Table 2 shows the material property parameters of the SOFC. To verify the SOFC model used here, the modeling results are compared with the simulation data of Kanga et al. [38] under the same operating conditions, as shown in Table 3. The results show that the model prediction agrees very well with the published data (Table 4). The maximum error of around 4% is found for predicting the fractions of CO and CO₂ at the outlet of the fuel channel. The predicted parameters of the SOFC performance, such as average current density and power density, are less than 1% different from the reported values.

The models used for other components of the SOFC system, such as vaporizers, pre-heaters, the mixer and the afterburner, are reported elsewhere [39]. In this study, the SOFC system is designed for a net power output of 150 kW and a fuel utilization of 0.7. The operating conditions for the SOFC system under nominal conditions are given in Table 5.

3. Results and discussion

3.1. Hydrogen production from various fuels

In this section, hydrogen production via steam reforming using ethanol, biogas and glycerol as fuels is investigated. Fig. 2 shows the effect of operating temperature on hydrogen production from the steam reforming of these three fuels. The results show a similar trend for each fuel type; this behavior is also observed for other fuels, such as methane and methanol [40,41]. An increase in the reforming...
temperature enhances the hydrogen yield due to the endothermicity of the steam reforming reaction. Ethanol and glycerol provide a higher hydrogen yield than biogas does. At atmospheric pressure, glycerol provides a slightly higher hydrogen yield than ethanol at temperatures below 973 K. When the steam reformer is operated at high pressure, glycerol produces more hydrogen than ethanol at temperatures below 1033 K. The operating pressure does not affect the hydrogen yield at temperatures above 1073 K.

Fig. 3 shows the effect of the steam-to-carbon (S/C) ratio on hydrogen yield at a pressure of 1 bar. At a low S/C ratio, glycerol produces more hydrogen than ethanol does. However, ethanol gives the highest hydrogen yield for higher S/C ratios. The use of biogas provides the lowest hydrogen yield throughout the operational range studied. Fig. 4 shows the ratio of the amount of gaseous product to that of hydrogen produced from the steam reforming of glycerol, ethanol and biogas at a temperature of 973 K and an S/C ratio of 2. The glycerol reforming process yields the highest CO2 concentration because glycerol has a hydrogen-to-carbon ratio lower than other fuels. The results also indicate that when ethanol is used as a fuel, less CO is produced. The presence of CO may cause the formation of carbon in the steam reforming process and the SOFC.

Fig. 5 shows the carbon formation boundary when glycerol, ethanol and biogas are used to produce hydrogen. The carbon formation tendencies of the different fuels are similar, being lower for high-temperature operation and high S/C ratios. Because the steam reforming of biogas generates more carbon monoxide, shifting the Boudouard reaction towards the products, this process must be performed at a higher temperature and S/C ratio to avoid carbon formation. Although the amount of carbon monoxide generated by the steam reformer fueled by ethanol is lower than that fueled by glycerol, ethanol reforming exhibits a higher carbon formation tendency than glycerol reforming because glycerol has a higher oxygen-to-carbon ratio than ethanol. Thus, glycerol is more likely than ethanol to be reformed to CO and CO2 rather than carbon.

3.2. SOFC system

Next, the performance of the SOFC fed by the synthesis gas obtained from the steam reforming of ethanol, glycerol and biogas is investigated. The power output of the SOFC system is specified at 150 kW. Fig. 6 presents the fraction of the heat duty required for each unit in the SOFC systems. The air preheater requires the most heat input, 79–81% of the overall heat consumption of the SOFC system, compared to other heat-requiring units. SOFCs are usually operated at high temperatures, and the electrochemical reaction is an exothermic reaction. These factors cause a temperature gradient in the SOFC stack. Therefore, excess air is required to maintain the SOFC temperature at a suitable level. The steam reformer also requires a large amount of energy.

The energy consumptions for each unit of the SOFC system are compared in Fig. 7 for different fuels. The results indicate that the SOFC system run on biogas requires the most external energy, whereas the ethanol-fueled SOFC system requires the least energy. Biogas reforming requires the most energy (77.9 kW) because the high CO2 content in biogas promotes a reverse water gas shift reaction, which is an endothermic reaction. Although the energy requirement of the glycerol steam reforming is lower, more heat is needed for the evaporator due to the high boiling point of glycerol (561.9 K).

Fig. 8 shows the SOFC stack efficiency and the thermal and electrical efficiencies of the SOFC systems supplied by different fuels. As observed in the figure, the SOFC system fueled by ethanol gives the best performance in terms of the SOFC stack efficiency.
The SOFC system fed by biogas shows the lowest thermal efficiency of the SOFC stack is fairly independent of fuel type (42–43%). The use of biogas as a fuel for the SOFC system provides the worst electrical efficiency because biogas mainly consists of carbon dioxide, which dilutes the hydrogen fuel fed into the SOFC stack. This dilution lowers the reversible cell voltage and increases the hydrogen production shows the least tendency toward carbon dioxide emission of the ethanol-fed SOFC system; it has the highest electrical and thermal efficiencies. Ethanol is a suitable fuel for the SOFC system; it has the least carbon dioxide emission.

Fig. 9 presents the amount of carbon dioxide released from the SOFC system at a power output of 150 kW. The SOFC system fed by ethanol minimizes the emission of carbon dioxide, whereas the glycerol-fueled SOFC system produces the most carbon dioxide. The carbon dioxide emission of the ethanol-fed SOFC system is less than that of the other systems by 21.76% (glycerol) and 19.97% (biogas). The simulation study indicates that ethanol is a superior performance of the ethanol steam reformer compared that of other fuels, which translates into the greatest hydrogen production. When biogas is used as a fuel for SOFC, the use of a purification unit is suggested to remove carbon dioxide from the biogas, thereby improving both the thermal and electrical system efficiencies.

4. Conclusions

The performance and efficiency of a steam reforming process and solid oxide fuel cell (SOFC) integrated system was theoretically analyzed in this study. Different renewable fuels, i.e., glycerol, ethanol, and biogas, were applied to generate hydrogen for SOFC. The results showed that ethanol can produce the most hydrogen at an operating temperature of 1073 K. The utilization of glycerol for hydrogen production shows the least tendency toward carbon formation. The SOFC system supplied by biogas requires the most energy and has the lowest electrical and thermal efficiencies. Ethanol is a suitable fuel for the SOFC system; it has the highest electrical and thermal efficiencies of the three fuels, and it has low carbon dioxide emission.

Acknowledgments

Support from the Thailand Research Fund and the Special Task Force for Activating Research (STAR), Chulalongkorn University Centenary Academic Development Project, is gratefully acknowledged.

D. Saebea would like to thank the Office of the Higher Education Commission, Thailand for their grant support under the program “Strategic Scholarships for Frontier Research Network for the Joint Ph.D. Program Thai Doctoral degree” for this research.

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