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Size and exergy assessment of solid oxide fuel cell-based H₂-fed power generation system with alternative electrolytes: A comparative study

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ABSTRACT

This work aims to analyse and compare the thermodynamic performance and size of two types of solid oxide fuel cell (SOFC)-based plants. The former is the conventional H₂-fed plant based on SOFC with an oxygen-ion conducting electrolyte (SOFC-O), and the latter is based on SOFC using a proton-conducting electrolyte (SOFC-H). Thermodynamic analysis reveals that in the SOFC-H system, due to H₂O formation at the cathode side, not only the anode concentration losses decreases, but also the partial pressure difference between H₂ and H₂O increases which leads to an increase in Nernst voltage compared to the SOFC-O system. Due to this, SOFC-H and SOFC-O based plants exhibit different performance in terms of the cell voltage, power, efficiency, stack outlet temperature and size of heat exchangers used for preheating the fuel and air. The results indicate, for current densities less than around 3,000 A/m², the energy and exergy efficiencies of SOFC-H-based system are more than those of the SOFC-O-based plant. This results in reduced area of heat exchangers per unit power used in the SOFC-H-based plant as compared with the SOFC-O-based plant. In addition, the sensitivity analysis demonstrates that using thin cells in the SOFC stack is favourable for the SOFC-H-based plant.

1. Introduction

Fossil fuel emissions have shown to be associated with accelerating anthropogenic climate change. Therefore, incorporation of alternative power generation systems are deemed necessary. Solid oxide fuel cell (SOFC)-based power generation systems have immense potential to produce clean electricity owing to the following advantages [1–3]:

- Has a high chemical-to-electrical energy conversion efficiency
- Operates at higher temperatures and is capable of being integrated with bottoming cycles and processes
- Has fast electrochemical reaction rates due to high temperature operation, eliminating the need for expensive noble metal catalysts
- Various hydrocarbon-based fuels, e.g., natural gas, can be used directly without external reforming processes.

Conventionally, SOFC, which is based on an oxygen ion-conducting electrolyte (together abbreviated as ‘SOFC-O’), is made from a solid

layer of yttria-stabilised zirconia. The solid layer tolerates high temperatures (approximately 1,000 °C), which enables the SOFC-O to conduct a feasible internal reforming process during a long lifetime [4]. Numerous studies have been conducted to evaluate the performance of the SOFC-O fed by alternative fuels. Thermoeconomic analysis and optimization of a multi-generation system based on SOFC-O plant was performed by Sadeghi et al. [1]. The optimization results showed that the system performance is impressively influenced by the two key parameters of fuel cell inlet temperature and fuel utilization factor. Chitsaz et al. [2] investigated the effects on the performance of SOFC-O based cogeneration plant of coupling with external reforming. The results demonstrated that coupling the proposed system with external reforming is not recommended from the viewpoints exergy and environment. Thermodynamic evaluation of small-scale power generation systems based on SOFC-O integrated with gasifier and gas turbine were carried out by Aravind et al. [5]. They reported that steam addition does not have a significant effect on the system electrical efficiency. A comprehensive study on fuel recycling in SOFC-O systems fed by alternative fuels was conducted by Rokni [6]. It was observed that the fuel recycling

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Nomenclature			
<i>Symbols</i>		U	Overall heat transfer coefficient (kW/ m ² K)
A	Area (m ²)	U _F	Fuel utilisation factor
A _c	Cell active area (m ²)	U _A	Air utilisation factor
$\dot{E}_x$	Exergy rate (kW)	v	Voltage (V)
$\dot{E}_x^{ch}$	Chemical exergy rate (kW)	$\dot{W}$	Power (kW)
$\dot{E}_D$	Exergy destruction rate (kW)	x	Overall reaction molar conversion rate
$\dot{E}_{loss}$	Energy loss (kW)	y	Mole fraction
E^{ph}	Thermophysical exergy rate (kW)	<i>Abbreviations</i>	
$\overline{e}_x^{ch,0}$	Standard chemical exergy per mole rate (kJ/mol)	AB	Afterburner
F	Faraday constant (C/mol)	ABl	Air blower
h	Specific enthalpy (J/kg)	FBl	Fuel blower
$\bar{h}$	Molar specific enthalpy (J/kg)	HEX	Heat exchanger
J	Current density (A/m ²)	LHV	High heat value (kJ/kg)
$\dot{m}$	Mass flow rate (kg/s)	PD	Power density (kW/m ²)
N _c	Number of cells	<i>Subscripts</i>	
$\dot{n}$	Mole flow rate (mol/sec)	e	Plant exit
P	Pressure (Pa)	en	Energy
$\dot{Q}$	Heat transfer rate (kW)	is	Isentropic
$\bar{R}$	Universal gas constant (J/mol.K)	ex	Exergy
T	Temperature (K)	<i>Greek symbols</i>	
		η	Efficiency (%)

has significant effect on the SOFC efficiency and the amount of anode off-gas recirculation strongly depends on the fuel type. A. S. Mehr et al. [7] proposed a novel solar-assisted biogas fed SOFC-O plant for wastewater treatment applications in Italy. The results indicated that installing solar collectors is beneficial for both energy and economic perspectives; so that installing 300 m² solar collectors could cover 25% of required natural gas from the grid. A SOFC-O based cogeneration system for power and steam production integrated with Stirling engine was proposed and analyzed by Hosseinpour et al. [8]. The thermodynamic analysis revealed that the exergy efficiency of 56.44% is achievable for the system. Van Biert et al. [9] compared the thermodynamic performance of various SOFC-O combined cycles. It was observed that under the operating condition of moderate cell voltages and high stack temperatures, the pressurized SOFC-gas turbine combined cycle attains the highest electrical efficiency. While, at high cell voltages and low temperatures, the SOFC-steam turbine combined cycle is more efficient. Potentials and challenges of solid oxide fuel cells fueled with biogas was investigated by Saadabadi et al. [10]. Utilization of high temperature residual heat from SOFCs for biogas production was taken account as a potential characteristic of SOFCs to integrate with wastewater treatment applications. Mojaver et al. [11] conducted a comparative thermodynamic and economic analyses on the performance of integrated methane fed SOFC-O plant and Organic Rankine Cycle (ORC). The exergy analysis demonstrated that under the optimized condition, the exergy efficiency of the proposed system is achieved as 60.20%. Considering the total exergy gained, CO₂ emission and product unit cost as the criteria, multi-objective optimization of a syngas fed SOFC-O plant integrated with a downdraft gasifier was carried out by Sadeghi et al. [12]. The optimization results showed that minimizing the product unit cost of the system aligns with an increase in the CO₂ emission and a reduction in exergy efficiency. Applying four alternative gasification agents, Hosseinpour et al. [13] investigated and compared the performance of an internally reformed SOFC-O power plant from the viewpoints of thermodynamics, economics and environment. It was found that the plant using oxygen as the gasification agent achieves the highest exergy efficiency and is the most cost efficient system.

In addition, the proton-conducting electrolyte can be integrated into the SOFC's structure (SOFC-H). These fuel cells normally use a barium

oxide-based ceramic as the electrolyte [4]. The main advantages of SOFC-H are as following [3–4]:

- Possibility of complete hydrogen utilisation by the anode of the fuel cell, because of proton migration from anode to cathode
- Less concentration losses on the anode site, due to H₂O formation at the cathode outlet.

In view of the advantages of a proton-conducting electrolyte in SOFCs, researches have evaluated the performance of SOFC-H-based systems. A thermodynamic assessment of a hydrogen-fed SOFC-H was performed by Demin et al. [14]. They investigated the effects of hydrogen purity, air humidity, hydrogen as well as oxygen utilisation factor, and SOFC operating temperature on the fuel cell efficiency. The results demonstrated that, in the stack operating temperature range of 1,000 to 1,300 K, the maximum electrical efficiency of SOFC-H is more than that of the SOFC-O. Maffei et al. [15] investigated the performance of a hydrogen- and/or ammonia-fed intermediate-temperature (700 °C) fuel cell using a proton-conducting BaCeO₃ solid electrolyte doubly doped with gadolinium and praseodymium. They observed that water and nitrogen were the only chemical by-products of the cell, making it an environmentally friendly device. Ni et al. [3] compared the performances of a methane-fed SOFC with an oxygen ion-conducting electrolyte and that with a proton-conducting electrolyte. They reported a lower anodic concentration overpotentials in the SOFC-H compared with the SOFC-O, because of H₂O formation on the cathode side of the SOFC-H. In addition, Ni et al. [16] conducted a thermodynamic analysis to compare the performance of an ammonia fed SOFC-H and SOFC-O. The results revealed that, under fuel and oxygen utilisation factors of 80% and 20%, respectively, the efficiency of the ammonia fed SOFC-H is 11% more than that of the SOFC-O due to the higher partial pressure of hydrogen and lower partial pressure of steam. In another work, an improved electrochemical model for the ammonia-fed SOFC-H was formulated by Ni et al. [17]. It was found that drawing more current had slightly increased the rate of NH₃ decomposition, but considerably decreased the hydrogen partial pressure at the anode-electrolyte interface. A direct ammonia-fed SOFC based on a thin proton-conducting electrolyte composed of BaCe_{0.8} Gd_{0.2}O_{3-δ} (BCGO) was tested at

600 °C and 650 °C, by Zhang et al. [18]. The results revealed that the maximum power density of 47 mW/cm² is achieved when the cell is operated at 600 °C. The performance of a planar SOFC-H was investigated by Patcharavorachot et al. [19] through a detailed electrochemical model for calculating the cell voltage losses, including ohmic, activation, and concentration losses. The results revealed that the system exhibited the best performance under an operating temperature of 800 °C and a pressure of 1 bar. In addition, they reported that, due to the relatively low conductivity of the proton-conducting electrolyte, the ohmic losses are the major overpotentials in an anode-supported SOFC-H. Arpornwichanop et al. [20] analysed the performance of a methane-fed SOFC-H with direct internal reforming. As a result, the SOFC-H achieved a maximum efficiency of 47% at the optimal condition. The performance of combined methane-fed SOFCs with various electrolytes including the proton-conducting and oxygen ion-conducting electrolytes was evaluated by Patcharavorachot et al. [21]. The simulation results demonstrated that, for a stack operating temperature below than 700 °C, the electrical efficiency of the SOFC-H system was greater than that of the SOFC-O. Ishak et al. [22] studied and compared the thermodynamic performances of an anode-supported ammonia-fed SOFC with oxygen-ion conducting electrolyte and proton-conducting electrolyte, respectively. For a stack operating temperature of 800 °C, the maximum power produced by the SOFC-H was 20–30% more than that of the SOFC-O. In another work, Arpornwichanop et al. [23] analysed the performance of a methane-fed SOFC-H with auto-thermal reforming. According to their report, the operating temperature had a strong effect on the performance of the SOFC-H, carbon monoxide formation, and heat production. Hossain et al. [24] reviewed studies that used proton-conducting electrolytes in intermediate-temperature SOFCs. They concluded that electrolytes prepared from barium cerate and barium zirconate were the most appropriate electrolytes for intermediate-temperature SOFC applications. Kalinci et al. [25] analysed the thermodynamic performance of a NH₃- and H₂-fed SOFC-H. From the results, the concentration voltage losses on the anode side was negligible, because H₂O was produced on the cathode side. However, the activation and ohmic losses were considerable. Lin et al. [26] examined the performance of several types of multiple-doped barium cerate proton-conducting electrolytes in SOFCs. From the results, a maximum power density of 138 mW/cm² was achieved by the hydrogen-fed SOFC-H at 700 °C. A thermal and electrochemical modelling of a power and syngas cogeneration system based on a tubular SOFC-H was conducted by Chen et al. [27]. They found that

the integration of SOFC-H at 0.7 V could effectively enhance syngas production and generate electricity (1.5 W) simultaneously. Mathematical modelling of an SOFC based on a proton-conducting electrolyte with current leakage was developed by Zhang et al. [28]. The simulation showed reduced energy and Faraday efficiencies with increasing operating temperature due to the existence of current leakage. Lei et al. [29] performed a theoretical and experimental study on the thermodynamic performance of a methane-fed proton-conducting SOFC with internal steam reforming. The results indicated that a single cell can operate stably with steam reforming of methane at 550 °C. Finally, the performance of a reversible SOFC based on a proton-conducting electrolyte was analysed by Danilov et al. [30]. According to the study, the reversible SOFC is cost-effective and durable in the range of 0.40–1.6 V.

From the literature review, the performance of SOFC-H and SOFC-O mainly depends on the fuel type and operating conditions, especially the SOFC operating temperature, current density, and fuel utilisation factor. Depending on the system operating temperature and current density for higher values of fuel utilisation factor, e.g., $U_F = 0.85$, the SOFC-H shows better performance with more electrical efficiency. However, the existing literature does not compare the performance of the SOFC-H system with that of the SOFC-O-based plant in terms of exergy. In addition, the fuel cell as a key component of an SOFC-based plant significantly affects the overall system characteristics, especially thermal management of heat exchangers used in the system. Therefore, under the same thermodynamic conditions, power plants with different types of SOFCs must employ heat exchangers with different effective areas. This study aims to investigate the influence on both the exergy efficiency and the size of the system of different types of SOFCs. To this end, electrochemical modelling of a hydrogen-fed anode-supported SOFC system with a proton-conducting electrolyte and alternatively an oxygen ion-conducting electrolyte was developed, and the energy and exergy in both systems were analysed.

2. System description and assumptions

2.1. System configuration

A schematic diagram of the proposed atmospheric SOFC-based power plant is shown in Fig. 1. The system consists of air and fuel blowers, heat exchangers, SOFC stack, and afterburner. The steps of operation of the system is as follows:

Fig. 1. Schematic of the proposed SOFC power plant.

Table 1

Input data for modelling of the proposed SOFC-based power generation system [1].

System	Parameter	Value
SOFC system	Stack inlet temperature	750 °C
	Stack operating pressure	1 bar
	Fuel utilisation factor, U_F	0.85
	Air utilisation factor, U_A	0.1
	Current density, J	1–20,000 A/m ²
	Number of cells, N_c	10,000
	Cell active area, A_c	0.01 m ²
	Anode thickness	500 μ m
	Cathode thickness	50 μ m
	Electrolyte thickness	20 μ m
	Blowers isentropic efficiencies, η_{is}	0.85
	Pressure drop in heat exchangers	2%
	Pressure drop in SOFC	2%
	Pressure drop in afterburner	5%
Ambient condition	Temperature, T_0	25 °C
	Pressure, P_0	1 bar

- The fuel (hydrogen) and air are pressurized by the fuel blower (FBI) and air blower (ABI), respectively. Then, utilizing the thermal energy of exhausted gases from the afterburner (AB), they are preheated up to the stack operating temperature of 750 °C through the corresponding heat exchangers.
- In the SOFC stack, depending on the electrolyte type, an oxygen ion or a proton migrates through the electrolyte spurring an electrochemical reaction. Unreacted gases and air exiting the SOFC stack enter the afterburner and undergo complete combustion, producing a high-temperature stream of gases to feed the heat exchangers.

2.2. Assumptions

For ease of modelling, the following assumptions were made:

- The system operates under steady state.

- Changes in the kinetic and potential energies are negligible.
- The system is at thermodynamic equilibrium.
- All gases in the system follow ideal gas laws.
- The SOFC stack and heat exchangers are isolated, and there is no heat exchange between them and the environment.
- Contact resistances are neglected in the SOFC stack.
- The pressure drop in pipes is neglected.
- The exchange current densities of the anode and cathode in the SOFC-H at an SOFC operating inlet temperature of 750 °C are known [19].

3. SOFC modelling

3.1. Data

The SOFC system was modelled using the engineering equations solver (EES). All input data used in the modelling are summarised in Table 1.

3.2. Fundamentals and mole balance equations

The schematic of a SOFC with an oxygen ion-conducting electrolyte is shown in Fig. 2. From the figure, H₂O forms on the anode side of the SOFC-O due to the oxygen-ion migration from cathode to anode. The term “x” in Fig. 2a denotes the molar conversion rate of the overall reaction, which is defined by the following equation [1,2,12]:

$$x \rightarrow \{H_2 + \frac{1}{2}O_2 \rightarrow H_2O\} \quad (1)$$

The reaction in Eq. (1) consumes $\times$ moles of H₂ and $x/2$ mol of O₂ to produce $\times$ moles of H₂O, which is depicted in Fig. 2a. x is derived from the relationship between the molar conversion rate of the overall reaction and the cell characteristics defined by Eq. (2) [1,2,12]:

$$x = \frac{J N_c A_c}{2 F} \quad (2)$$

Fig. 2. Molar balance equations for the SOFC with oxygen ion-conducting electrolyte: **a)** based on the molar conversion rate of the overall reaction (x), **b)** based on the fuel utilisation factor (U_F).

Fig. 3. Molar balance equations for the SOFC with proton-conducting electrolyte: **a)** based on the molar conversion rate of the overall reaction (x), **b)** based on the fuel utilisation factor (U_F).

Table 2
Mole balance relationships for two types of SOFCs.

SOFC-O Anode	Cathode	SOFC-H Anode	Cathode
$\dot{n}_{H_2,7} = \dot{n}_{H_2,3} - x$	$\dot{n}_{O_2,8} = \dot{n}_{O_2,6} - \frac{x}{2}$	$\dot{n}_{H_2,7} = \dot{n}_{H_2,3} - x$	$\dot{n}_{O_2,8} = \dot{n}_{O_2,6} - \frac{x}{2}$
$\dot{n}_{H_2O,7} = x$	$\dot{n}_{N_2,8} = \dot{n}_{N_2,6}$		$\dot{n}_{N_2,8} = \dot{n}_{N_2,6}$
			$\dot{n}_{H_2O,8} = x$

where J , N_c , A_c and F are the current density, number of cells, cell active area, and the Faraday constant (of 96,485C/mol) [1,2,12]:

In addition, according to the definition of fuel utilisation factor presented in Eq. (3) [1,2,12], the mole balances can be expressed in terms of U_F , which is illustrated in Fig. 2b.

$$U_F = \frac{\dot{Fuel}_{consumed}}{\dot{Fuel}_{supplied}} = \frac{x}{\dot{n}_{H_2,3}} \quad (3)$$

Fig. 3 represents the mole balance in SOFC-H. From the figure, in this type of SOFC, H_2O is formed at the cathode side because protons migrate from the anode to the cathode and react with O_2 . The mole balance relations for both SOFCs are summarised in Table 2.

For both cases, the fuel mass flow rate can be determined by Eq. (3); however, the air mass flow rate can be calculated based on the air utilisation factor defined by Eq. (4) [1,2,12]:

$$U_A = \frac{\dot{Air}_{consumed}}{\dot{Air}_{supplied}} = \frac{\left(\frac{x}{2}\right)}{\dot{n}_{O_2,6}} \quad (4)$$

3.3. Energy balance equation

After solving the mole balance equations and determining the species

concentration on the two electrodes, the energy balance equation of the adiabatic SOFC stack can be written as following [1,2,12]:

$$\dot{n}_{Fuel} \bar{h}_{Fuel, anode \ inlet} + \dot{n}_{Air} \bar{h}_{cathode \ inlet} = \dot{W}_{SOFC, Stack} + \dot{n}_{anode \ offgas} \bar{h}_{anode \ offgas, anode \ outlet} + \dot{n}_{Air} \bar{h}_{cathode \ outlet} \quad (5)$$

Here, $\dot{W}_{SOFC, Stack}$ is the power produced by the SOFC stack, which is calculated as follows [1,2,12]:

$$\dot{W}_{SOFC, stack} = N_c A_c J V_c \quad (6)$$

where V_c shows the cell voltage, which equals the various voltage losses subtracted from the Nernst voltage, and is given as follows [1,2,12]:

$$V_c = V_N - V_{loss} \quad (7.a)$$

$$V_{loss} = V_{act} + V_{conc} + V_{ohm} \quad (7.b)$$

3.4. Voltage analysis

All the equations used for calculating the Nernst voltage and voltage losses of SOFC-O and SOFC-H are listed in Tables 3 and 4, respectively. The terms g_i^0 , y_i , $J_{0,i}$, $E_{act,i}$, σ_{anode} , $D_{eff,i}$, K_i , TBP and $\bar{R}$ in Tables 3 and 4 denote Gibbs standard function, mole fraction of species, exchange current density of the electrodes, activation energy on the electrodes, pre-exponential factor of the exchange current density, diffusion coefficient, electric conductivity of electrodes, three-phase boundary, and universal gas constant (8.314 J/mol.K).

4. Thermodynamic analysis

Under the steady-state condition, and by neglecting the changes in kinetics and potential energies, the energy and exergy balance equations

Table 3

Electrochemical equations applied for Nernst voltage and voltage loss calculations of SOFC-O [9].

Nernst voltage	$V_N = -\frac{\Delta \bar{g}^0}{2F} + \frac{\bar{R} T_{stack, outlet}}{2F} \ln \left[\frac{y_{H_2, anode, outlet} y_{O_2, cathode, outlet}^{0.5} \left(\frac{P_{stack, outlet}}{P_0} \right)^{0.5}}{y_{H_2O, anode, outlet}} \right] \quad (8)$ $\Delta \bar{g}^0 = \bar{g}_{H_2O}^0 - \bar{g}_{H_2}^0 - 0.5 \bar{g}_{O_2}^0, \quad \bar{g}_i = \bar{h}_i (T = T_{stack, outlet}) - T_{stack, outlet} \bar{s} (T = T_{stack, outlet}, P = P_0)$
Activation voltage losses	$V_{act} = V_{act, anode} + V_{act, cathode} \quad (9)$ $V_{act, anode} = \frac{\bar{R} T_{stack}}{F} \operatorname{arcsinh} \left(\frac{J}{2J_{0, anode}} \right), \quad J_{0, anode} = \frac{\bar{R} T_{stack}}{2F} \sigma_{anode} \exp \left(\frac{-E_{act, anode}}{\bar{R} T_{stack}} \right)$ $V_{act, cathode} = \frac{\bar{R} T_{stack}}{F} \operatorname{arcsinh} \left(\frac{J}{2J_{0, cathode}} \right), \quad J_{0, cathode} = \frac{\bar{R} T_{stack}}{2F} \sigma_{cathode} \exp \left(\frac{-E_{act, cathode}}{\bar{R} T_{stack}} \right)$ $E_{act, anode} = 137 \text{ kJ/mol}, \quad \sigma_{anode} = 6.54 \times 10^{11} \frac{1}{\Omega \cdot m^2}, \quad J_{0, anode} = 2.83 \times 10^{10} \frac{A}{m^2}$ $E_{act, cathode} = 140 \text{ kJ/mol}, \quad \sigma_{cathode} = 2.35 \times 10^{11} \frac{1}{\Omega \cdot m^2}, \quad J_{0, cathode} = 1.02 \times 10^{10} \frac{A}{m^2}$
Concentration voltage losses	$V_{conc} = V_{conc, anode} + V_{conc, cathode} \quad (10)$ $V_{conc, anode} = \frac{\bar{R} T_{anode, outlet}}{2F} \ln \left(\frac{P_{H_2O, TPB} P_{H_2, anode, outlet}}{P_{H_2, TPB} P_{H_2O, anode, outlet}} \right), \quad V_{conc, cathode} = \frac{-\bar{R} T_{cathode, outlet}}{4F} \ln \left(\frac{P_{O_2, TPB}}{P_{O_2, cathode, outlet}} \right)$ $P_{H_2O, TPB} = P_{H_2O, anode, outlet} + \frac{J L_{anode} T_{anode, outlet} \bar{R}}{2 F D_{eff, anode}}, \quad P_{H_2, TPB} = P_{H_2, anode, outlet} - \frac{J L_{anode} T_{anode, outlet} \bar{R}}{2 F D_{eff, anode}}$ $P_{O_2, TPB} = P_{cathode, outlet} - (P_{cathode, outlet} - P_{O_2, cathode, outlet}) \exp \left(\frac{J L_{cathode} \bar{R} T_{cathode, outlet}}{4 F D_{eff, cathode} P_{cathode, outlet}} \right)$
Ohmic voltage losses	$D_{eff, anode} = 3.66 \cdot 10^{-5} \frac{m^2}{s}, \quad D_{eff, cathode} = 1.37 \cdot 10^{-5} \frac{m^2}{s} \quad (11)$ $V_{ohm} = J \left(\frac{L_{cathode}}{k_{cathode}} + \frac{L_{anode}}{k_{anode}} + \frac{L_{elec}}{k_{elec}} \right)$ $k_{cathode} = 84000 \frac{1}{\Omega \cdot m}, \quad k_{anode} = 80000 \frac{1}{\Omega \cdot m}, \quad k_{elec} = 33400 \exp \left(\frac{-10300}{T_{stack}} \right) \frac{1}{\Omega \cdot m}$

Table 4

Electrochemical equations applied for Nernst voltage and voltage loss calculations of SOFC-H [19].

Nernst voltage	$V_N = -\frac{\Delta \bar{g}^0}{2F} + \frac{\bar{R} T_{stack, outlet}}{2F} \ln \left(\frac{y_{H_2, anode, outlet} y_{O_2, cathode, outlet}^{0.5} \left(\frac{P_{stack, outlet}}{P_0} \right)^{0.5}}{y_{H_2O, cathode, outlet}} \right) \quad (12)$ $\Delta \bar{g}^0 = \bar{g}_{H_2O}^0 - \bar{g}_{H_2}^0 - 0.5 \bar{g}_{O_2}^0, \quad \bar{g}_i = \bar{h}_i (T = T_{stack, outlet}) - T_{stack, outlet} \bar{s} (T = T_{stack, outlet}, P = P_0)$
Activation voltage losses	$V_{act} = V_{act, anode} + V_{act, cathode} \quad (13)$ $V_{act, anode} = \frac{\bar{R} T_{stack}}{F} \operatorname{arcsinh} \left(\frac{J}{2J_{0, anode}} \right), \quad J_{0, anode} = 800 \text{ A/m}^2$ $V_{act, cathode} = \frac{\bar{R} T_{stack}}{F} \operatorname{arcsinh} \left(\frac{J}{2J_{0, cathode}} \right), \quad J_{0, cathode} = 800 \text{ A/m}^2$
Concentration voltage losses	$V_{conc} = V_{conc, anode} + V_{conc, cathode} \quad (14)$ $V_{conc, anode} = \frac{\bar{R} T_{anode, outlet}}{2F} \ln \left(\frac{P_{H_2, anode, outlet}}{P_{H_2, TPB}} \right), \quad V_{conc, cathode} = \frac{-\bar{R} T_{cathode, outlet}}{2F} \ln \left(\frac{P_{O_2, cathode, outlet} P_{H_2O, TPB}}{P_{O_2, TPB} P_{H_2O, cathode, outlet}} \right)$ $P_{O_2, TPB} = P_{O_2, cathode, outlet} - \frac{J L_{cathode} T_{cathode, outlet} \bar{R}}{2 F D_{eff, cathode}}, \quad P_{H_2O, TPB} = P_{H_2O, cathode, outlet} + \frac{J L_{cathode} T_{cathode, outlet} \bar{R}}{4 F D_{eff, cathode}}$ $P_{H_2, TPB} = P_{anode, outlet} - (P_{anode, outlet} - P_{H_2, anode, outlet}) \exp \left(\frac{J L_{anode} \bar{R} T_{anode, outlet}}{4 F D_{eff, anode} P_{anode, outlet}} \right)$
Ohmic voltage losses	$D_{eff, anode} = 8.984 \times 10^{-5} \frac{m^2}{s}, \quad D_{eff, cathode} = 4.748 \times 10^{-5} \frac{m^2}{s} \quad (15)$ $V_{ohm} = J \left(\frac{L_{cathode}}{k_{cathode}} + \frac{L_{anode}}{k_{anode}} + \frac{L_{elec}}{k_{elec}} \right)$ $k_{cathode} = 84000 \frac{1}{\Omega \cdot m}, \quad k_{anode} = 80000 \frac{1}{\Omega \cdot m}, \quad k_{elec} = 225.92 \exp \left(\frac{-6300}{T_{stack}} \right) \frac{1}{\Omega \cdot m}$

for each component of the system as the control volume are expressed as follows [31,32]:

$$\dot{Q}_{cv} + \sum \dot{n}_i \bar{h}_i = \dot{W}_{cv} + \sum \dot{n}_e \bar{h}_e \quad (16)$$

$$\sum \dot{n}_i e \bar{x}_i + \sum \dot{Q}_j \left(1 - \frac{T_0}{T_j} \right) = \sum \dot{n}_e e \bar{x}_e + \dot{W}_{cv} + \dot{E}_{XD} \quad (17)$$

where, subscripts i and e denote the inlet and outlet streams, respectively, and ex is the unit exergy flow rate of each stream, which is the sum of the thermo-physical and chemical exergies defined as follows [33]:

$$e \bar{x}_k = e \bar{x}_k^{th} + e \bar{x}_k^{ch} \quad (18)$$

$$e \bar{x}_k^{th} = (\bar{h} - \bar{h}_0) - T_0 (\bar{s} - \bar{s}_0) \quad (19)$$

$$e \bar{x}_k^{ch} = \sum y_k e \bar{x}_k^{ch,0} + \bar{R} T_0 \sum y_k \ln(y_k) \quad (20)$$

Here, $e \bar{x}_k^{ch,0}$ signifies the standard chemical exergy of each species [33]. The applied energy and exergy balance equations for each component of the system are presented in Table 5. In addition, the energy and exergy efficiencies are defined as follows:

$$\eta_{en} = \frac{\dot{W}_{SOFC, Stack} - \dot{W}_{FC} - \dot{W}_{AC}}{\dot{m}_{H_2} LHV_{H_2}} \quad (28)$$

$$\eta_{ex} = \frac{\dot{W}_{SOFC, Stack} - \dot{W}_{FC} - \dot{W}_{AC}}{\dot{m}_{H_2} e x_{ch, H_2}} \quad (29)$$

In order to calculate the heat exchangers' area, A_{HEX} , the following energy equation is applied for each heat exchanger [34]:

Table 5

Governing energy and exergy balance equations for the system components.

Component	Energy and exergy balance equations
Fuel blower	$\dot{m}_1 \bar{h}_1 + \dot{W}_{FC} = \dot{m}_2 \bar{h}_2 \dot{m}_1 ex_1 + \dot{W}_{FC} = \dot{m}_2 ex_2 + \dot{E}x_{D,FC}$ (21.a) (21.b)
Air blower	$\dot{m}_4 \bar{h}_4 + \dot{W}_{AC} = \dot{m}_5 \bar{h}_5 \dot{m}_4 ex_4 + \dot{W}_{FC} = \dot{m}_5 ex_5 + \dot{E}x_{D,AC}$ (23.a) (23.b)
Heat exchanger 1	$\dot{m}_2 \bar{h}_2 + \dot{m}_9 \bar{h}_9 = \dot{m}_3 \bar{h}_3 + \dot{m}_{10} \bar{h}_{10} \dot{m}_2 ex_2 + \dot{m}_9 ex_9 = \dot{m}_3 ex_3 + \dot{m}_{10} ex_{10} + \dot{E}x_{D,HEX1}$ (24.a) (24.b)
Heat exchanger 2	$\dot{m}_4 \bar{h}_4 + \dot{m}_{10} \bar{h}_{10} = \dot{m}_5 \bar{h}_5 + \dot{m}_{11} \bar{h}_{11} \dot{m}_4 ex_4 + \dot{m}_{10} ex_{10} = \dot{m}_5 ex_5 + \dot{m}_{11} ex_{11} + \dot{E}x_{D,HEX2}$ (25.a) (25.b)
Afterburner	$\dot{m}_7 \bar{h}_7 + \dot{m}_8 \bar{h}_8 = \dot{m}_9 \bar{h}_9 \dot{m}_7 ex_7 + \dot{m}_8 ex_8 = \dot{m}_9 ex_9 + \dot{E}x_{D,AB}$ (26.a) (26.b)
Exergy loss through the system	$\dot{E}x_{loss} = \dot{m}_{11} ex_{11}$ (27)

$$\dot{Q}_{HEX,i} = A_{HEX,i} U_{HEX,i} LMTD_{HEX,i} \quad (30)$$

Here, $\dot{Q}_{HEX,i}$ and $A_{HEX,i}$ are the heat transfer rate and area of each heat exchanger. $U_{HEX,i}$ denotes the overall heat transfer coefficient between the hot and cold streams in heat exchangers, which is taken as 50 W/m²K from a previous study [7]. $LMTD_{HEX,i}$ is the logarithm mean temperature difference between the hot and cold streams, which is defined by the following equation [35]:

$$LMTD_{HEX,i} = \frac{(T_{hot,in} - T_{cold,out}) - (T_{hot,out} - T_{cold,in})}{\ln \left(\frac{T_{hot,in} - T_{cold,out}}{T_{hot,out} - T_{cold,in}} \right)} \quad (31)$$

5. Results

5.1. SOFC-O modelling validation

As presented by the authors in previous studies [1,2,12], using the experimental data reported by Tao et al. [36] and Raymond a George [37] as well as the theoretical results reported by Akkaya et al. [38], the modelling conducted for the SOFC with the oxygen ion-conducting electrolyte is validated, which is presented in Table 6. Referring to Table 6, it is observed that there is a good agreement between the results obtained from the present work and those reported in the literature [36–37]; so that the maximum error is 5.19%, which is calculated at high current density of 6000 A/m².

Table 6

Comparison of the results obtained from the present modelling implemented for the SOFC-O with those experimental data and theoretical results available in literature [36–38].

Current density (A/m ²)	Cell voltage (V)		Error (%)
	Present work	Literature	
2000	0.742	0.76 (Experimental data reported by Tao et. al [36])	2.37
3000	0.684	0.68 (Experimental data reported by Tao et. al [36])	0.59
4000	0.634	0.62 (Experimental data reported by Tao et. al [36])	2.26
5000	0.582	0.57 (Experimental data reported by Tao et. al [36])	2.1
6000	0.547	0.52 (Experimental data reported by Tao et. al [36])	5.19
3200	0.6096	0.610 (Experimental data reported by Raymond a George [37])	0.066
3200	0.6096	0.609 (Theoretical results reported by Akkaya et. al [38])	0.099

5.2. Comparison of SOFC-H modelling

We compared the results of this model with those of Patcharavachot et al. [19] for a hydrogen-fed SOFC-H, and analysed the cell's performance. The results of the analysis are presented in Fig. 4a and 4b, as the variation in various voltage losses and cell voltage, as well as power density, respectively. The fuel cell's operating temperature and pressure were taken as 1,073 K and 1 atm. Fig. 4 indicates that the calculated voltage losses and cell voltage as well as the power density for SOFC-H aligns with the reference data [19]. Therefore, the modelling presented for SOFC-H is reliable.

5.3. Parametric study

In this section, the effects on the fuel cell and system performance of current density is investigated for current densities up to 20,000 A/m².

Fig. 5.a shows the influence of electric current density on the activation voltage losses (V_{act}) for anode and cathode sides of both the SOFCs. From Fig. 5a, as the current density increases, the activation voltage losses on the two electrodes increases for both types of fuel cells. This trend can be explained by Eqs. 9 and 13. Moreover, from Fig. 5a, the activation voltage losses of the SOFC-H is more than those of the SOFC-O, indicating that the electrochemical reactions in the SOFC-H do not favour current generation as much as those in the SOFC-O. This is mainly because of the large difference between the values of exchange current densities calculated/reported for each cell [9,19]. Furthermore, from Fig. 5a, as the current density increases, the difference between the activation losses of the two SOFCs increases such that, for higher values of current density, approximately 3,000 A/m², V_{act} for the SOFC-H is almost three times more than that of the SOFC-O.

The effects of electric current density on the concentration voltage losses for both FCs are depicted in Fig. 5b. From this figure, for all ranges of J, V_{conc} on the anode side of the SOFC-H and V_{conc} on the cathode side of the SOFC-O are negligible. This is because H₂O in the SOFC-H and the SOFC-O is formed on the cathode and anode sides, respectively. In addition, from Fig. 5b, as the current density increases, V_{conc} for both cases increased, which is mainly due to the higher rate of the electrochemical reactions at higher currents.

Variations in the ohmic (V_{ohm}) and total voltage losses (V_{loss}) with the current density are shown in Fig. 5c. Referring to this figure, as the current density increases, V_{ohm} in both the cases increases, which is explained by Eqs. 11 and 15. In addition, from Fig. 5c, V_{ohm} of the SOFC-H is higher than V_{ohm} of SOFC-O, which is mainly because the proton-conducting electrolyte in the SOFC-H has a higher resistance than the oxygen ion-conducting electrolyte in the SOFC-O (Eqs. 11 and 15). Moreover, Fig. 5c also shows that V_{loss} of the SOFC-H is greater than V_{loss} of the SOFC-O, and the difference becomes wider for electric current densities of more than around 3,000 A/m².

Fig. 5d illustrates the influences of current density on the Nernst (V_{Nernst}) and cell voltages (V_{cell}) for both systems. According to Fig. 5d, the Nernst voltage of the SOFC-H is higher than that of the SOFC-O. This is justified by the fact that the partial pressure difference between H₂ and H₂O at the SOFC-H outlet is more than of the SOFC-O. Regarding the fundamentals and mole balance relationships provided in Section 3.1 and Table 2, H₂ is the only species on the anode side of the SOFC-H. Thus, H₂ mole fraction at the anode outlet is 1, which is much higher than that of H₂O formed at the cathode side. This large difference between the y_{H_2} and y_{H_2O} leads to higher values of the term

$$\left[\frac{y_{H_2, anode, outlet} y_{O_2, cathode, outlet}^{0.5}}{y_{H_2O, cathode, outlet}} \left(\frac{P_{stack, outlet}}{P_0} \right)^{0.5} \right] (>1) \text{ in the Nernst equation (Eq. 12). To conclude, the term } \frac{\bar{R} T_{stack, outlet}}{2 F} \ln$$

$$\left[\frac{y_{H_2, anode, outlet} y_{O_2, cathode, outlet}^{0.5}}{y_{H_2O, cathode, outlet}} \left(\frac{P_{stack, outlet}}{P_0} \right)^{0.5} \right] \text{ in Eq. 12, will be a positive value}$$

Fig. 4. Comparison between the results obtained by the present work and Patcharavorachot et al. [19] regarding the **a)** various voltage losses and **b)** cell voltage as well as power density.

and sums up with the term with the Gibbs free energy ($-\frac{\Delta G^0}{2F}$) in Eq. 12. However, in the case of SOFC-O, considering that H_2O is formed at the anode outlet, both y_{H_2} and y_{H_2O} are lower than 1. This means the value of the term $\left[\frac{y_{H_2, \text{anode outlet}}^{0.5} y_{O_2, \text{cathode outlet}}^{0.5}}{y_{H_2O, \text{anode outlet}}} \left(\frac{P_{\text{stack outlet}}}{P_0} \right)^{0.5} \right]$ is < 1 . Consequently, the term $\frac{\bar{R} T_{\text{stack outlet}}}{2F} \ln \left[\frac{y_{H_2, \text{anode outlet}}^{0.5} y_{O_2, \text{cathode outlet}}^{0.5}}{y_{H_2O, \text{anode outlet}}} \left(\frac{P_{\text{stack outlet}}}{P_0} \right)^{0.5} \right]$ is achieved as a negative value, which leads to a decrease in the total value of the Nernst voltage (Eq. 8).

From Fig. 5d, for current density values of less than around 3,000 A/m^2 , V_{cell} of the SOFC-H is more than that of the SOFC-O. However, at current densities more than around 3,000 A/m^2 , the SOFC-O's cell voltage is higher. This is because, as shown in Fig. 5a and 5c earlier, for current densities of more than around 3,000 A/m^2 , the activation, ohmic, and total voltage losses of the SOFC-H are much higher than those of the SOFC-O. In other words, in the open circuit mode, the Nernst voltage of the SOFC-H is more than that of the SOFC-O. However, by drawing current and getting away from the open circuit mode, the

voltage losses, particularly those due to the activation and ohmic resistances, increase faster in the SOFC-H. Eventually, at a current density of around 3,000 A/m^2 , the overall cell voltage of the SOFC-O improves over that of the SOFC-H. In addition, from Fig. 5d, as current density increases, the cell voltages of both FCs decrease. This is because, while drawing more current causes a drop in the electrochemical potential, it results in a surge in voltage losses, as depicted in Fig. 5c.

The influence of the electric current density on the power density (PD) of both SOFC systems is demonstrated in Fig. 5e. As in Fig. 5d, PD of the SOFC-H up to current densities of approximately 3,000 A/m^2 , in Fig. 5e, is greater than that of the SOFC-O. However, for current densities more than around 3,000 A/m^2 , the opposite trend is observed. This follows a similar trend as V_{cell} vs. J (Fig. 5d). In addition, in the case of SOFC-H, by increasing the current density up to approximately 7,500 A/m^2 , the current density increment overcomes the cell voltage drop (Fig. 5d) and the power density increases. However, beyond this optimal point, the influence of cell voltage reduction surpasses the current density increment effects, and the power is reduced. In addition, a similar trend is observed for the SOFC-O system; however, the optimal values of PD are achieved for a current density of approximately 18,500

Fig. 5. Variations in the a) activation voltage losses, b) concentration voltage losses, c) ohmic and total voltage losses, d) Nernst and cell voltages, and e) power density with the current density of at $T_{in,stack} = 750$ °C, $U_F = 0.85$ and $U_A = 0.1$ for both SOFC-H and SOFC-O.

A/m².

Changes in the fuel and air mass flow rates with increasing current density for both systems are shown in Fig. 6a. From Fig. 6a, under the same thermodynamic and operating conditions, both systems consume the same amounts of fuel and air. This can be justified by Eq. (2), where the same values of J , N_c and A_c applied for each SOFC lead to the same molar conversion rate of the overall reaction (x). Thus, since both systems have the same, constant fuel and air utilisation factors ($U_F = 0.85$, $U_A = 0.1$), by Eqs. (3) and (4), the hydrogen and air molar flow rates in SOFC-H are equal to those in SOFC-O. In addition, from Fig. 6a, the air and fuel mass consumption rates in both systems increase as the current densities increase. This is because, from Eq. (2), higher values of electric current density leads to greater values of x. Therefore, from Eqs. (3) and (4), and by considering that U_F and U_A remain constant, $\dot{n}_{H_2}$ and $\dot{n}_{O_2}$ increase as the current density increases.

Identical fuel and air mass flow rates (see Fig. 6a)—and given that the stack inlet temperature in both systems are the same and constant ($T_{in,stack} = 750$ °C)—lead to the same amount of thermal energy flowing

into both the SOFCs. Therefore, from Fig. 5e, and the energy balance equation in Eq. (5), for current densities of less than around 3,000 A/m², the SOFC-H achieves a lower stack outlet temperature than the SOFC-O. Thus, in this range of current density, the stack temperature difference and the exhaust gas temperature of the SOFC-H will be lower than those of the SOFC-O (Fig. 6b). However, for current densities greater than around 3,000 A/m², the reverse trend is observed. In addition, Fig. 6b shows that, by increasing the current density, the stack temperature difference as well as the exhaust gas temperatures can be increased in both cases. From Fig. 6a, as the current density increases, the fuel and air mass flow rates increase, which leads to an increased energy flow at the stack inlet. Thus, since the stack is adiabatic, despite the power density has been increased (see Fig. 5e), it leads to an increase in the stack outlet temperature. Indeed, as the current density increases both the hydrogen and air inlet energy flow rates (stack inlet energy) as well as the stack power increase. Therefore, referring to the energy balance equation given in Eq. (5), the stack outlet temperature could either increase or decrease. However, as illustrated in Fig. 6c, the rate of increase in the

Fig. 6. Variations in a) the fuel and air mass flow rates and b) the stack temperature difference and exhausted gas temperature c) rate of increase in the stack inlet energy flow rate and power with the current density of at $T_{in,stack} = 750$ °C, $U_F = 0.85$ and $U_A = 0.1$ for both SOFC-H and SOFC-O.

stack inlet energy flow is more than the power increment rate. Thus, as current density increases, the stack outlet temperature raises; which indicates on more thermal energy losses at high current densities.

Although in both systems for the wide ranges of current density, by increasing the J , PD raises (see Fig. 5e), the fuel mass flow rate increment (see Fig. 6a) overcomes the power growth and as a result referring to Eqs. (28) and (29), the energy and exergy efficiencies decreases; which is illustrated in Fig. 7a. As in Fig. 5d and 5e, Fig. 7a reveals that the energy and exergy efficiencies of the SOFC-H are higher than those of the SOFC-O up to current densities of approximately 3,000 A/m², and for densities greater than around 3,000 A/m², the efficiencies of the SOFC-O are higher. Therefore, the supremacy of SOFC-H system compared to the SOFC-O is only valid for the low values of current

Fig. 7. Variations in a) the energy and exergy efficiencies and b) the total exergy destruction rate and exergy loss with the current density at $T_{in,stack} = 750$ °C, $U_F = 0.85$ and $U_A = 0.1$ for both SOFC-H and SOFC-O.

density which has not been clarified before through similar studies [14,16,21]. In addition, the obtained result is based on such assumptions as $J_0, cathode = J_0, anode = 800$ A/m² [19], which these values are extremely lower than those calculated for SOFC-O [9]. Therefore, even if in future, the values of exchange current density for the SOFC-H get updated; the obtained result about the supremacy of SOFC-H system compared to the SOFC-O, not only still remains the same and reliable, but also the advantageous of SOFC-H can be extended for current densities > 3,000 A/m². This is justified because; as the exchange current densities become larger, the activation losses decrease which leads to an increase in the cell voltage.

Fig. 6b and 7a demonstrate the increase in the exhaust gas temperature and the decrease in the exergy efficiency with the current density, respectively. This trend leads to an increase in the exergy loss as well as the total exergy destruction rate for both systems, respectively (Fig. 7b). In addition, as in the previous figures, for current densities less than around 3,000 A/m², SOFC-H shows better performance in terms of exergy. However, for current densities greater than around 3,000 A/m², the exergy destruction rate and exergy loss values of the SOFC-H system are higher than those of the SOFC-O plant.

The number of cells (N_c) and effective area of each cell (A_c) in both cases are the same; therefore, the size of the SOFC stack of both the systems are also the same. In addition, from Fig. 6a, under the same thermodynamic conditions, the fuel and air mass flow rates flowing through the blowers are also same in both systems. Therefore, since both systems operate under atmospheric pressure, the blowers' pressure ratio, and consequently, their size, in both SOFC-H and SOFC-O will be the same. Moreover, because the total mass flow rates through the afterburner in both plants are the same, the size of the combustors are also equal. To conclude, the sizes of the blowers, SOFC stack, and

afterburner of the SOFC-H are similar to the sizes of those of the SOFC-O.

However, due to the varying temperatures of the exhaust gases (see Fig. 6b), the areas of the heat exchangers will not be equal. Therefore, the only components responsible for a difference in the system size are the fuel and air heat exchangers. This difference between the two systems and the influence of current density on the heat exchangers' area is illustrated in Fig. 8a. From Fig. 8a, for $J < 3,000 \text{ A/m}^2$, the area of the heat exchangers of the SOFC-O-based plant is less than that of the SOFC-H. However, for higher current densities ($J > 3,000 \text{ A/m}^2$), the reverse trend is observed. This can be justified by Fig. 6b. From the figure, for $J < 3,000 \text{ A/m}^2$, temperature of the exhaust gases in the hot stream flowing into the heat exchangers is higher in the SOFC-O-based plant. Therefore, considering the same, constant stack inlet temperatures for both systems, the area of the heat exchangers in the SOFC-O-based power plant will be smaller. Similarly, for $J > 3,000 \text{ A/m}^2$, the size of the heat exchangers of the SOFC-H-based power plant is smaller compared to those in the SOFC-O-based one. In addition, Fig. 8a shows that, by increasing the current density, the size of the heat exchangers in both cases can be increased. This is because, as previously demonstrated in Fig. 6a, drawing more currents brings about more consumed fuel and air in both systems.

Finally, change in the normalised area of the heat exchangers per kW ($\frac{A_{HEX}}{W_{net}}$) with the current density is shown in Fig. 8b for both types of SOFCs. From Fig. 8b, as the current density increases, $\frac{A_{HEX}}{W_{net}}$ of heat exchangers in both power plants increases. This is mainly because of the increase in the fuel and air mass flow rates, which has been discussed in Fig. 6a and 8a. In addition, Fig. 8b demonstrates that, for $J < 3,000 \text{ A/m}^2$, $\frac{A_{HEX}}{W_{net}}$ in the SOFC-H system is lower than that of the SOFC-O system,

Fig. 8. Variations in a) heat exchangers' areas and b) normalised heat exchangers' areas per unit power production ($\frac{A_{HEX}}{W_{net}}$) with the current density at $T_{in, stack} = 750 \text{ }^\circ\text{C}$, $U_F = 0.85$ and $U_A = 0.1$ for both SOFC-H and SOFC-O.

and for $J > 3,000 \text{ A/m}^2$, the reverse trend is observed. This is because, although A_{HEX} of the SOFC-H for $J < 3,000 \text{ A/m}^2$ is more than that of the SOFC-O (Fig. 8a), from Fig. 5e, the power density of the SOFC-H plant is also higher overcoming the conflicting effects arising from A_{HEX} . The same reasoning can be applied for observations at $J > 3,000 \text{ A/m}^2$.

6. Sensitivity analysis

6.1. Effects of SOFCs physical characteristics

The parametric study performed for both H_2 -fed SOFC-based plants revealed that under the thermodynamic condition of $T_{in, SOFC} = 750 \text{ }^\circ\text{C}$ and for higher fuel utilisation factors e.g. $U_F = 0.85$, there is a critical value for current density (around $3,000 \text{ A/m}^2$) below which the performance of SOFC-H-based plant is more promising than that of the SOFC-O-based plant. However, this value could be changed as the physical characteristics of the SOFCs such as the thicknesses of electrodes and electrolyte vary. In this regard, as the thicknesses of electrodes and electrolyte decreases, the critical current density increases, which is shown in Fig. 9. It means that using thin cells in the SOFC stack is favourable for the SOFC-H-based plant, because it extends the range of current density for which the performance of SOFC-H-based plant is more advantageous than that of the SOFC-O-based plant.

6.2. Effects of fuel utilisation factor

Effects on the SOFCs energy and exergy efficiencies of fuel utilisation factor at SOFC inlet temperature of $750 \text{ }^\circ\text{C}$ and current density of 1000 A/m^2 is outlined in Fig. 10a. Referring to Fig. 10a, as the U_F increases, both SOFCs achieve higher efficiencies; which is due to a decrease in the hydrogen molar flow rate according to Eq. (3).

In addition, Fig. 10b reveals that under the condition of $U_F = 0.65$, the critical current density value is achieved around 2000 A/m^2 . Therefore, compared to the results obtained from Fig. 7a, it is concluded that for the higher values of fuel utilization factor, higher values of critical current density are calculated. Thus, it can be inferred that employing a system with higher values of fuel utilization factor allows the SOFC-H type show a better performance in wider ranges of current density.

7. Applications

The parametric study presented in the previous section revealed that the resistances, especially those from the activation and ohmic losses, increase more rapidly with the increasing current density in the case of SOFC-H. However, from Fig. 5e and 7, for the higher fuel utilisation factor, e.g. $U_F = 0.85$, when $J < 3,000 \text{ A/m}^2$ (depending on the thicknesses of electrodes and electrolyte) the H_2 -fed SOFC-H shows better

Fig. 9. Influence on the critical current density of a decrease in the electrodes and/or electrolyte thicknesses.

Fig. 10. Effects on the SOFCs energy and exergy efficiencies of **a)** fuel utilisation factor at $T_{in, stack} = 750$ °C, $J = 1000$ A/m² and $U_A = 0.1$ **b)** current density at $T_{in, stack} = 750$ °C, $U_F = 0.65$ and $U_A = 0.1$.

performance with higher values of power density, energy, and exergy efficiencies. Therefore, for applications which use H₂ as the fuel and operate at low current densities ($J < 3,000$ A/m²), e.g. lightweight stationary application, the SOFC-H system is more efficient. In addition, for this range of current density ($J < 3,000$ A/m²), the SOFC-H power plant is more compact. It is worthy to mention, for the mobile applications operating at higher current densities, using SOFC-H system is not recommended.

8. Conclusion

In this study, the thermodynamic performance of an H₂-fed SOFC power plant with an oxygen ion-conducting electrolyte (SOFC-O), and alternatively with a proton-conducting electrolyte (SOFC-H), was analysed and compared. Under the condition of high fuel utilisation ($U_F = 0.85$), the efficiency and size of both systems are significantly affected by the current density. The important results of the study are highlighted as follows:

- Due to the higher difference in partial pressure between H₂ and H₂O at the SOFC-H outlet, the corresponding Nernst voltage is higher than that of the SOFC-O.
- All polarisation losses including activation, concentration and ohmic losses of the SOFC-H are higher than those of the SOFC-O. The only exception is the anode concentration loss, which is negligible in the SOFC-H because of H₂O formation on the cathode side.
- With increasing current density, the resistances, especially those from the activation and ohmic losses, increase more rapidly in the case of SOFC-H.

- A comparative evaluation of the proposed systems under the same thermodynamic conditions and fuel utilisation of 0.85 revealed that there is a critical value for current density (around 3,000 A/m²); so that for $J < 3,000$ A/m², the SOFC-H-based power plant showed better thermodynamic performance and is more compact. However, for $J > 3,000$ A/m², the SOFC-O is superior.
- The critical current density depends on the thicknesses of electrodes and electrolyte. It is found that using thin cells in the SOFC stack is favourable for the SOFC-H-based plant.
- Employing a system with higher values of fuel utilisation factor allows the SOFC-H type show a better performance in wider ranges of current density.
- In both types of SOFCs, with increasing current density, the cell voltage losses, exergy destruction and the exergy losses through the system and heat exchanger size increased, and consequently the cell voltage and system efficiencies decreased.

CRediT authorship contribution statement

Mohsen Sadeghi: Conceptualization, Methodology, Software, Modelling, Validation, Writing - original draft, Investigation, Supervision. **Moharram Jafari:** Supervision. **Yashar S. Hajimolana:** Review, Advising, Editing. **Theo Woudstra:** Review, Advising, Editing. **P.V. Aravind:** Review, Supervision.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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