A novel hybrid liquefied natural gas process with absorption refrigeration integrated with molten carbonate fuel cell

Mehdi Mehrpooya1,2, Parimah Bahramian1, Fathollah Pourfayaz1, Hadi Katooli1 and Mostafa Delpisheh2

1Department of Renewable Energies and Environment, Faculty of New Sciences and Technologies, University of Tehran, Tehran 1439957131, Iran; 2Hydrogen and Fuel Cell Laboratory, Faculty of New Sciences and Technologies, University of Tehran, Tehran 1439957131, Iran

Abstract

The production of liquefied natural gas (LNG) is a high energy-consuming process. The study of ways to reduce energy consumption and consequently to reduce operational costs is imperative. Toward this purpose, this study proposes a hybrid system adopting a mixed refrigerant for the liquefaction of natural gas that is precooled with an ammonia/water absorption refrigeration (AR) cycle utilizing the exhaust heat of a molten carbonate fuel cell, 700°C and 2.74 bar, coupled with a gas turbine and a bottoming Brayton super-critical carbon dioxide cycle. The inauguration of the ammonia/water AR cycle to the LNG process increases the cooling load of the cycle by 10%, providing a 28.3-MW cooling load duty while having a 0.45 coefficient of performance. Employing the hybrid system reduces energy consumption, attaining 85% overall thermal efficiency, 53% electrical efficiency and 35% fuel cell efficiency. The hybrid system produces 6300 kg mol h−1 of LNG and 146.55 MW of electrical power. Thereafter, exergy and sensitivity analyses are implemented and, accordingly, the fuel cell had an 83% share of the exergy destruction and the whole system obtained a 95% exergy efficiency.

Keywords: LNG production; dual mixed refrigerant; absorption refrigeration; molten carbonate fuel cell; super-critical carbon dioxide cycle; hybrid system

1. INTRODUCTION

Natural gas is one of the most important energy carriers in the 21st century and functions as a valuable fuel in being clean and widely available. Owing to the vast and accessible availability of natural gas resources and development of new technologies in the past few decades and the reduction in cost and execution time, the economic efficiency of projects centralized on developing and transferring gas has improved and prodigious capital is invested [1]. Also, a global effort to reduce greenhouse gases (GHGs), especially carbon dioxide, has put err on the side of using natural gas compared to other fuels [2].

Natural gas exports are a cardinal approach in financing natural gas-rich countries, which has not been properly exploited so far. When in the liquid phase, natural gas is more economical and less hazardous to transfer over long distances, and this is the main notion underscoring the development of the liquefied natural gas (LNG) industry.

There are many arrangements proposed for producing LNG and associated refrigeration cycles in patents and the literature: the pre-cooled liquefaction process [3], the double mixed refrigerant (DMR) liquefaction process [4] and the process of dual mixed refrigerant with carbon dioxide preheating cycle [5]. Also, combined heating and cooling power (CCHP) systems have been addressed in many studies [6, 7]. The number of equipment and the amount of power needed in refrigeration cycles are among the chief factors in considering the proposed arrangements. Energy consumption is an important subject that must be minimized in order to reach an efficient energy conversion system, which, in fact, is one of the most important challenges concerning LNG.
production technologies [8]. The major share of this energy consumption is pertinent to the compressors of a refrigeration cycle; therefore, promoting energy-efficient liquefaction processes yields higher energy efficient LNG chains, resulting in the reduction of fuel consumption and carbon emissions [8].

In refrigeration processes used for the production of LNG, high-pressure compressors, and hence great energy consumption, are employed to cool the refrigerant and increase the cost of LNG production. It is conspicuous that the appropriate approach to reduce the costs of LNG production and increase the cost-effectiveness of this energy carrier is to discover appropriate methods for reducing energy consumption [8]. The liquefaction process can be integrated with upstream processes, and through this, the overall required refrigeration decreases accordingly [9–11]. In recent years, various studies have been performed to reduce the required power and initial costs. For instance, operating conditions and refrigeration system process configurations were modified to optimize system performance and reduce costs [12, 13].

Absorption refrigeration (AR) systems, in lieu of vapor compression (VC) refrigeration cycles, can be considered as an approach to decrease the required power in a process. AR cycles have many advantages over VC cycles, such as less energy consumption, resulting in lower operating costs and the potential of exploiting waste energy. Mehrpooya et al. [14] inaugurated this approach; they reported a 30% reduction in power consumption and a 0.48 coefficient of performance (COP). A study by Ghorbani et al. [15] followed a similar path, whereby replacing the precooling stage of DMR and C3MR liquefaction cycles with an AR cycle; the adopted procedure reduced the power consumption by ~18.4% and 12.5%, respectively, compared with the mentioned processes. Kalinowski et al. [16] studied AR cycles in LNG processes, focusing on reducing the required power. In their study, substitution of propane chillers with AR systems was evaluated and found that the recovering waste heat from a 9-MW electricity generation process provisioned 5.2 MW of waste heat that was capable to provide the cooling needed for the LNG plant, and thereby saving 1.9 MW of electricity consumption. The authors asserted that the application of integrated cooling, heating and power is an outstanding energy saving opportunity for the oil and gas industry. Srikhirin et al. [17] addressed that the working fluids commonly used in ARs are water/ammonia (H₂O/NH₃) and Li/H₂O. Different modifications are performed on ARs to improve their efficiency. Ansarinasab and Mehrpooya [18] investigated a new process to produce LNG using AR through advanced exergoeconomic analysis; the COP for the system was 0.49 and the avoidable exogenous destruction of the cycle was very low. Gong and Boulama [19] investigated a single-effect lithium bromide/water (LiBr/H₂O) AR system using advanced exergy analysis. They indicated that the evaporator and condenser exergy destruction were endogenous and avoidable and had low sensitivity to the cycle operating parameters. Mehrpooya et al. [14] studied a novel AR system that employed a mixed fluid cascade and reported that the COP of the H₂O/NH₃ cycle is 0.48 and the specific power consumption is 0.172 kWh/kgLNG, showing a 30% reduction in the power consumption. Sun and Xie [20] investigated a combined microgas turbine with AR. Niasar et al. [21] appraised a CCHP hybrid system that adopted an ammonia–water AR cycle that provisioned 5.2 MW of refrigeration at a COP of 0.45.

Fuel cell systems are suitable options for common power generation systems, especially in the micro-sized distribution systems, due to their relatively high efficiency and low environmental impact [22]. High-temperature fuel cells, for instance, molten carbonate and solid oxide fuel cells, are considered as one of the future electrical power generation systems due to their environmental friendliness and high-efficiency benefits. The high temperature and operating pressure of MCFCs make them suitable for mixed heat and power applications [23, 24]. In addition, high operating temperatures enable MCFCs to use fuels such as natural gas and landfill gas, without the need for an external refinement [25]. Many studies have addressed the conflation of fuel cells and LNG process. Zhang et al. [26] investigated the integration of MCFC and AR systems for combined power and cooling purposes. They reported that maximum power density and efficiency increased by more than 3.0%. An alternative approach to maximizing the efficiency of MCFC systems is to combine fuel cells with power generators such as gas turbines (GTs), turboexpanders or gas microturbines. Based on this, the combined system of molten carbonate fuel cells and GTs has attracted a lot of attention due to higher efficiency [23, 27, 28]. The optimum design of the SOFC-GT hybrid process was studied [29], which addressed that the optimum current density and number of cells are 9000 A·m⁻² and 2122. Another example of hybrid SOFC, NH3/water and ARS Rankine cycle process is studied [30], which revealed that the electrical efficiency of the combined system is 62.4% and heat recovery is 22.1 kW. A combined PEM fuel cell generator and AR system [31] and LNG process [32] was also studied in the literature.

The integration of MCFC and SO₂ was addressed and an exergy analysis of the hybrid system was put through [33]; the fuel cell and the reformer had the highest exergy destruction, respectively. Sánchez et al. [34] explored SOFC and MCFC hybrid systems coupled with SO₂ bottoming cycle in part-load operation. They reported that heat exchanger efficiency has a significant impact on system efficiency. On similar lines, Baronci et al. [35] compared the organic Rankine cycle (ORC) with SO₂ as bottoming cycle for MCFC, expressing that the latter obtains an electrical efficiency twice as the former.

Many studies have been done on LNG production methods and finding a way to increase energy efficiency in these processes. Vatani et al. [36] was investigated five common processes of LNG with energy and exergy analyses and reported that the mixed fluid cascade process has the maximum exergy efficiency (51.82%). Rodgers et al. [37] developed a strategy for using dissipated heat from the GT by using in an AR system to increase the propane cycle cooling capacity and, as a result, to improve the energy efficiency of the process of liquefaction in an LNG plant.

In this paper, a brainchild hybrid integration of mixed refrigerant liquefaction process with AR precooling, molten carbonate fuel cell, GT and Brayton supercritical CO₂ power cycle is proposed and investigated. The proposed system has not been studied.
in the literature, and the framework aims at producing LNG with regard to reduction in power consumption and efficient energy utilization while decreasing the environmental effect of the power plant, which are the novelty in this study. The cooling load of the pre-cooling cycle is applied through an AR cycle that provisions the required thermal load by a molten carbonate fuel cell coupled with a GT and Brayton cycle, which is an innovative approach addressed in this study. Energy and exergy analyses of the hybrid system are carried out to appraise the system’s structure and efficiency and assessing the degree of irreversibility of the process components. The operating conditions are evaluated through a sensitivity analysis and the effect of different variables on the system performance are analyzed.

2. SYSTEM DESCRIPTION AND METHODOLOGY

Herein, the components of the hybrid system are described in detail. The assumptions used for simulation and modeling of the

Table 1. Specifications of the DMR process.

| Parameter | Value |
|-----------|-------|
| Pre-cooling cycle refrigerant temperature in condenser outlet | 9.36°C |
| Main cycle refrigerant in condenser outlet | 31.9°C |
| LMTD in multi-streams heat exchangers | 3°C |
| Natural gas pressure | 65 bar |
| Pressure drop in heat exchangers | 0 bar |
| Cooling temperature in pre-cooling cycle | −32.2°C |
| Pump adiabatic efficiency | 90% |
| Compressor adiabatic efficiency | 80% |

Table 2. Molar components of DMR process flow.

| Components (mol%) | Pre-cooling refrigerant | Condensing/sub-cooling refrigerant |
|-------------------|-------------------------|----------------------------------|
| Nitrogen          | 7.0                     | 4.8                              |
| Methane           | 24.82                   | 29.9                             |
| Ethane            | 64.16                   | 21.3                             |
| n-Butane          | 11.03                   | 13.5                             |
| Hydrogen          | 0                       | 6.9                              |
| CO₂               | 0.95                    | 1.65                             |
| CO                | 0.81                    | 0.45                             |
| Hydrogen          | 0.53                    | 0.4                              |
| Oxygen            | 0                       | 0.4                              |
| H₂O               | 0                       | 0.4                              |

Table 3. Molar component of natural gas.

| Mole fraction (%) | Methane | Ethane | Nitrogen | CO₂ | CO | Hydrogen | Oxygen | H₂O |
|-------------------|---------|--------|----------|-----|----|----------|--------|-----|
| Methane           | 92.77   | 3.43   | 0.95     | 0.81| 0.53| 0.53     | 0      | 0   |
| Flow properties   | Temperature, °C | 30 | Pressure, kPa | 300 | Molar Flow, kgmol.hr⁻¹ | 912 | Density, kg.m⁻³ | 0.731 | Relative density | 0.586 |
|                   | LHV, Kj.kg⁻¹ | 47.13 |
|                   | HHV Kj.kg⁻¹ | 52.21 |

...
processes are presented and the relations applied in the modeling are defined.

Figure 1 demonstrates the block flow diagram of the proposed system. The system embraces five main sections: DMR liquefaction cycle, pre-cooling AR cycle, MCFC, GT and Brayton supercritical CO₂ power cycle. Natural gas is fed into the DMR liquefaction process at 29°C and 650 kPa and it is also fed into the MCFC at 20°C and 300 kPa. The required water and air for electrochemical reactions are fed at ambient pressure into the MCFC at 20°C and 15°C, respectively. The power for the DMR liquefaction process is provisioned by the MCFC, the GT and the Brayton supercritical CO₂ power cycle. Exhaust heat from the MCFC-GT system outlet is recovered by the Brayton cycle and an AR system. Ammonia water is used in the ARS system and the produced refrigeration is used as the required precooling for natural gas liquefaction. In the following subsections, each of the blocks of the proposed system is scrutinized and delineated technically and the modeling procedures are conducted.

Figure 2. Schematic of the DMR process simulated in the Aspen-HYSYS [53].

Table 4. Simulation assumptions of the AR cycle (H₂O/NH₃).

| Parameter                                      | Value       | Unit             |
|------------------------------------------------|-------------|------------------|
| Low pressure                                   | 1.2         | Bar              |
| High pressure                                  | 13.0        | Bar              |
| NH₃ concentration in concentrated solution     | 25.0        | mass%            |
| NH₃ concentration in diluted solution          | 6.28        | mass%            |
| Cooling load of absorber                       | 345.54      | kJ/kg of strong solution |
| Cooling load of condenser                      | 240.00      | kJ/kg of strong solution |
| Heating load of generator                      | 501.63      | kJ/kg of strong solution |
| Heating load of evaporator                     | 227.46      | kJ/kg of strong solution |
| Cooling load of rectifier                      | 145.19      | kJ/kg of strong solution |
| NH₃ temperature in evaporator                  | -29.55      | °C               |
| Pump power consumption                         | 1.56        | kJ/kg of strong solution |
| NH₃ purity                                     | 99.9        | mass%            |
| Concentrated solution flow/diluted solution flow | 0.8        | Kg.hr⁻¹/kg.hr⁻¹  |
| COP                                            | 0.452       |                  |
Figure 3. (a) The schematics of the AR process [15] and (b) effect of NH3 concentration on the temperature of ammonia–ammonia-saturated liquids in adsorbents [15].

2.1. Dual mixed refrigerant process

The DMR process is commonly alluded to as the modification of the C3-MR process, which uses a mixed fluid cycle for precoothing of natural gas instead of propane [51]. The precoothing cycle of the DMR process functions at two compression levels. The mixed refrigerant of the pre-cooling cycle, which comprises ethane, propane and isobutane, along with the natural gas and a stream from the main cycle, enters the first heat exchanger of the precooling cycle (C1a) as the hot sides. These three streams enter the first heat exchanger and the required cooling load to reduce the temperature of them to 0°C is provided through reduction of a portion of the pressure of the precooling cycle stream in a pressure-reducing valve.

In modeling and simulation, the Peng–Robinson equation is implemented, with regard to its applicability in a wide assortment of temperature and pressure conditions in defining the properties of light hydrocarbons employed in LNG processes [8]. Table 1 shows the conditions and assumptions of the DMR process. Table 2 presents the molar component of refrigerant streams of the DMR process and Table 3 outlines the conditions and properties of natural gas. Figure 2 expresses the process flow diagram of the DMR process. Two mixed refrigerant cycles, precooling and subcooling cycles, are drawn by red and blue lines, respectively.

| Parameter | Value | Unit |
|-----------|-------|------|
|          |       |      |

**Table 5.** Kinetics parameters of external reforming reactions [30].

| Parameter | Value | Unit |
|-----------|-------|------|
| kreforming | $5.922 \times 10^{8} \times \exp(-\frac{209}{RT})$ | kmol | kg cat | kPa | s |
| kWGS | $6.028 \times 10^{-4} \times \exp(-\frac{26830}{RT})$ | kmol | kg cat | kPa | s |
| Kp(reforming) | $1.98 \times 10^{17} \times \exp(-\frac{26830}{RT})$ | 1 | kPa | s |
| Kp(WGS) | $1.767 \times 10^{-2} \times \exp(-\frac{4400}{RT})$ | kmol | kg cat | kPa | s |
| Kco | $5.127 \times 10^{-13} \times \exp(140000/RT)$ | 1 | kPa |
| $K_{H_2}$ | $5.6 \times 10^{-10} \times \exp(-\frac{23400}{RT})$ | $\frac{1}{(kPa)^2}$ |
| $K_{H_2O}$ | $9.251 \times \exp(-\frac{15900}{RT})$ | $\frac{1}{(kPa)^2}$ |

| Parameter | Value | Unit |
|-----------|-------|------|
|          |       |      |

**Table 6.** MCFC performance parameters used in equations [37].

| Parameter | Value | Unit |
|-----------|-------|------|
| LiNaCO$_3$ | $3.28 \times 10^{-9}$ | 1.97 $\times 10^{-6}$ |
| LiKCO$_3$ | $3.39 \times 10^{-6}$ | 2.2 $\times 10^{-3}$ |
| C1 | $2.04 \times 10^{-3}$ | 1.39 $\times 10^{-6}$ |
| C2 | $1.12 \times 10^{-2}$ | 0.0128 |
| $\Delta H_{C1}$(Jmol$^{-1}$) | 132000 | 83400 |
| $\Delta H_{C2}$(Jmol$^{-1}$) | 67100 | 22800 |
| $\Delta H_{H_2}$(Jmol$^{-1}$) | 23700 | 77800 |
| $\Delta H_{H_2O}$(Jmol$^{-1}$) | 23000 | 25200 |

| Input | Value | Unit |
|-------|-------|------|
| Natural gas (fuel) inlet temperature | 20 | °C |
| Air inlet temperature | 15 | °C |
| Water inlet temperature | 25 | °C |
| Methane conversion | 40% | |
| reformer S/C | 3.2 | |
| Fuel utilization | 0.9 | |
| Nominal cell temperature | 699.1 | °C |
| Air gas inlet composition | 21% O2, 79% N2 | |
| Pressure | 2.71 | Bar |
| Fuel cell pressure drop | 0 | kPa |
| Reformer pressure drop | 28.6 | kPa |
| Steam temperature | 134 | °C |
| Cell current density | 0.3 | Acm$^{-2}$ |
| Cell voltage | 1.1 | V |
| Active cell area | 320 | cm$^2$ |
| Number of cells | 215000 | |

**Table 7.** MCFC assumptions for simulation [38].

| Input | Value | Unit |
|-------|-------|------|
| Natural gas (fuel) inlet temperature | 20 | °C |
| Air inlet temperature | 15 | °C |
| Water inlet temperature | 25 | °C |
| Methane conversion | 40% | |
| reformer S/C | 3.2 | |
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| Steam temperature | 134 | °C |
| Cell current density | 0.3 | Acm$^{-2}$ |
| Cell voltage | 1.1 | V |
| Active cell area | 320 | cm$^2$ |
| Number of cells | 215000 | |
2.2. AR process

Many industrial processes generate large volumes of thermal energy from burning fossil fuels to generate steam or heat to be used in their energy supply chain. At the end of the processes, the heat is dispelled to the environment. This waste heat can be utilized in AR cycles that operate with heat duty. As aforementioned, there are two main refrigeration cycles: VC and AR. The performance of AR systems is highly dependent on the chemical and thermodynamic properties of the operating fluid [38]. Many operating fluids have been proposed in different studies. The study of absorption fluids by Marcris and Gutraj [39] illustrates that there are 40 combinations of refrigerants and 200 absorbent compounds. However, the most common and incorporated fluids are H$_2$O/NH$_3$ and LiBr/H$_2$O [40]. Thermodynamic properties of H$_2$O/NH$_3$ [41–45] and LiBr/H$_2$O [46,47] are reported in the literature.

The use of AR systems reduces global environmental footprints, such as the effects of GHG emissions from the burning of fossil fuels in industrial plants. To extend the application of AR systems, their performance needs to be enhanced and their costs reduced. Refrigeration systems in LNG processes consume a high amount of power to produce the required refrigeration in the process [48]. In the LNG process, natural gas must be liquefied to about $-162^\circ$C; cascade refrigeration systems are a promising approach used to provide refrigeration in different stages [49]. Figure 3a depicts the simulated AR cycle.

2.2.1. Integration of the AR cycle with the DMR

After designing the DMR process for LNG production, and design of the H$_2$O/NH$_3$ AR cycle, this section addresses the integration of the AR system in the LNG production process. It should be noted that in assimilating the H$_2$O/NH$_3$ absorption cycle into the DMR process, the available temperature in this absorption cycle is $-30^\circ$C, and given one of the goals of this research, the investigation of the possibility of using absorption cooling cycles in high production capacity plants, DMR process is selected among the two-cycles LNG production methods (Liquefin, DMR, APX, C3-MR).

First, ARS is adjusted to the desired temperature conditions, and then the necessary changes are made to the DMR process in order to integrate the ARS into the process. Ammonia water absorption cooling cycles are available for cooling to temperatures up to $-60^\circ$C [50]. However, for refrigeration at temperatures below $-33^\circ$C, the pressure of the evaporator after pressure reducing valves must be lower than 1 bar; to clarify, parts of the ammonia–water cycle including evaporator, absorber and pump should operate in relative vacuum conditions. Working in vacuum conditions requires special provisions, including the use of...
of a vacuum pump and nitrogen purging in order to prevent the entry of air into the system, as well as supplying equipment with resistance to vacuum conditions. Such a condition increases the cost of equipment and plumbing and thus the total cost of the unit. In this study, working under atmospheric pressure conditions is avoided and the pressure of ammonia flow is considered to be 1.2 bar. The heat exchanger used as the evaporator is of shell and tube types. Inlet ammonia stream temperature to the evaporator is $-25^\circ$C.

Figure 3b illustrates the effect of ammonia concentration on the temperature of ammonia–water saturated liquids. Due to the temperature limitation, the ammonia concentration in the rich mixture of water–ammonia was considered to be 0.25 mass fraction, which causes a 31.9$^\circ$C temperature on the absorbent.

The pump outlet pressure is considered to be 13 bar, which results in a condensing temperature of 34$^\circ$C for ammonia with 99.9% purity. The water vapor temperature utilized in the tower boiler is 180$^\circ$C. By considering 5$^\circ$C as the minimum temperature approach in the boiler, the output temperature of the boiler cannot reduce more than 5$^\circ$C. Then again, by reducing the ratio of the bottom product flow rate to the feed, the water component of the bottom increases, and as a result, the temperature of the bottom product increases too. Consequently, owing to the temperature limitation at the bottom of the tower, the ratio of the bottom product to the feed was considered to be 0.8. The assumptions for the AR system inaugurated in the DMR cycle are summarized in Table 4. Also, this table presents the optimized design characteristics of the water–ammonia absorption cooling cycle. Figure 4 outlines the inauguration of the AR water–ammonia cycle into the DMR process. The initial cost of AR systems is higher than that of VC systems; nevertheless, attributed to the new hybrid integration, the required surface area of multi-flow heat exchangers is reduced and, accordingly, their manufacturing costs are also reduced, which in part compensates the high initial costs of the employment of AR system.

2.3. Molten carbonate fuel cell
Having access to natural gas, the hybrid system uses natural gas as the input fuel for the MCFC. The conditions for natural gas are given in Table 3. Different kinetics are presented for SMR and WGS reactions. These kinetics are subjected to operating conditions and the type of catalyst used. The kinetics parameters of the external reforming reactions are given in Table 5. These
kinetics are derived from the Langmuir–Hinshelwood–Hougen–Watson (LHHW) model and the Freundlich adsorption equation. The approach of the LHHW model is implemented for the surface concentrations of the species in the reaction, whereby relating partial pressures of surface species to observed species in the gas phase, which can be fitted to the kinetic data [51]. Freundlich adsorption equation is an empirical expression that represents the adsorption isothermal variation of a gas quantity that is adsorbed by a unit mass of solid adsorbent with pressure, defined as \( \frac{x}{m} = kP^{1/n} \); \( x \) is the mass of the gas adsorbed over mass \( m \) of the adsorbent at the pressure of \( P \); also, \( k \) and \( n \) are constants that are based on adsorbent and gas at a specific temperature [52]. The steam-to-carbon ration, S/C, is defined as

\[
\frac{s}{C} = \frac{H_2O}{CO + CH_4 + 2C_2H_6 + \ldots}
\]  

The natural gas in the reformer is converted to hydrogen through the following reactions [33]:

\[
CH_4 + H_2O \leftrightarrow CO + 3H_2
\]  

\[
CO + H_2O \leftrightarrow CO_2 + H_2
\]  

In the anode part of the MCFC, the electrochemical reaction ensues the following [33]:

\[
H_2 + CO_3^{2-} \rightarrow CO_2 + H_2O + 2e^-
\]  

\[
CO + CO_3^{2-} \rightarrow 2CO_2 + 2e^-
\]  

And finally, in the reduction reaction in the cathode [33]:

\[
\frac{1}{2}O_2 + CO_2 + 2e^- \rightarrow CO_3^{2-}
\]  

In order to increase the MCFC exhaust voltage, several fuel cells are aligned in series, which is called a fuel cell stack set. The area of a single fuel cell in this study is considered 330 cm². The power of the MCFC stack is defined as [6]

\[
Power_{stack} = N_{cell} \times A_{cell} \times j_{cell} \times V_{cell},
\]  

where \( N \) is the number of cells used in it, \( A \) is the area of each cell, \( j \) is the current density and \( V \) is the produced voltage of each cell.

The term utilization factor (UF) in a fuel cell is defined as the amount of fuel consumed in the fuel cell is divided by the inlet fuel to the fuel cell. If fuel other than hydrogen is used, the term is amended as

\[
n_{H_2, equivalent} = n_{H_2} + n_{CO} + 4n_{CH_4} + 7n_{C_2H_6} + \ldots
\]  

\[
UF = 1 - \frac{n_{H_2, anode exhaust}}{n_{H_2} + n_{CO} + 4n_{CH_4} + 7n_{C_2H_6} + \ldots}
\]  

| Table 8. | Modeling settings for process components [38]. |
|-----------------|-----------------------------------------------|
| **Heat exchanger** | **Modeling settings** | **Solution method** | **Heat loss Tolerance** | **Maximum iteration** |
| **Heat exchanger** | **Solution method** | **Heat loss Tolerance** | **Maximum iteration** |
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| **Heat exchanger** | **Solution method** | **Heat loss Tolerance** | **Maximum iteration** |
| **Heat exchanger** | **Solution method** | **Heat loss Tolerance** | **Maximum iteration** |
Table 9. Simulation results of MCFC system.

| Output                          | Value          |
|--------------------------------|----------------|
| Fuel feed                      | 912.4 kgmol·h⁻¹|
| Air feed                       | 1.467 × 10⁶ kgmol·h⁻¹|
| Cell power                     | 132 MW         |
| Cell voltage                   | 1.1 V          |
| Steam rate                     | 2960 kgmol·h⁻¹|
| Exhaust gas temperature (before heat recovery) | 700 °C |

where \( n \) is the molar flow rate. This equation is used to calculate how much each part is equivalent to how much hydrogen. The operating voltage of the MCFC is obtained from the following [22]:

\[
V = E - \eta_{\text{Nernst}} - jR_{\text{tot}}, \quad (10)
\]

where \( E \) is the maximum potential theoretically achievable; \( \eta_{\text{Nernst}} \) is the loss of Nernst; \( J \) is the current density; and \( R_{\text{tot}} \) is the summation of the irreversibility in anode, cathode and electrode. The highest potential of recyclable fuel cells is gained from the summation of their reversibility in anode, cathode and electrode. The Gibbs free energy can be stated in terms of temperature [22]:

\[
\Delta G = -242000 + 45.8T,
\]

where \( T \) is the stack temperature in Kelvin. The standard potential of the cell is obtained from the free energy of the Gibbs. The loss of Nernst is defined as follows:

\[
\eta_{\text{Nernst}} = \frac{RT}{2f} \ln \frac{P_{\text{H}_2} \left( \frac{1}{P_{\text{O}_2,\text{ca}}} \right) P_{\text{CO}_2,\text{ca}}}{P_{\text{H}_2,\text{an}} P_{\text{CO}_2,\text{an}}}, \quad (13)
\]

where \( P \) represents the average partial pressure of the components in the cell. The total irreversible losses, \( R_{\text{tot}} \), consist of total losses:

\[
R_{\text{tot}} = R_{\text{an}} + R_{\text{ca}} + R_{\text{ir}}, \quad (14)
\]

where \( R_{\text{an}} \) and \( R_{\text{ca}} \) are irreversible losses of anode and cathode, individually, and \( R_{\text{ir}} \) is the inner resistance of the cell. The losses for the MCFC can be calculated from the following equations [22]:

\[
R_{\text{an}} = C_a e^{-\frac{\Delta H}{RT}} P_{\text{H}_2,\text{an}}^{-0.5} \quad (15)
\]

\[
R_{\text{ca}} = \left( C_1 e^{-\frac{\Delta H}{RT}} P_{\text{CO}_2,\text{ca}}^{0.75} P_{\text{CO}_2,\text{ca}}^{0.5} \right) + C_2 e^{-\frac{\Delta H}{RT}} P_{\text{CO}_2,\text{ca}}^{-1} \quad (16)
\]

\[
R_{\text{ir}} = C_{\text{ir}} e^{\frac{\Delta H}{RT}}, \quad (17)
\]

where \( R \) is the resistivity in terms of \( \Omega \text{cm}^{-2} \), \( C_1 \) is the dependent variable of electrodes and electrolytes, \( P \) is the pressure in terms of atm and \( \Delta H \) is the enthalpy in terms of \( \text{Jmol}^{-1} \). The current in each unit is achieved from the current density multiplied by the surface area of the cell as

\[
i = j \times A_{\text{unit}}. \quad (18)
\]

The numerical value of the variables used in the MCFC voltage calculation is given in Table 6. With the help of these equations and the MATLAB software environment, voltage and current values are calculated and the results are used in the HYSYS simulator. In the next step, the data that meet the requirements of the designed system are used in the final simulation. Table 7 shows the parameters used in the MCFC system and the assumptions for the simulation.

Figure 5 illustrates the scheme of the MCFC fuel cell system and the bottoming Brayton cycle. In the MCFC, 40% of the natural gas in the reformer is transformed to hydrogen and the rest is converted into the fuel cell. To improve the system’s efficiency, part of the output stream of the cathode is returned and united with the output of the anode and burned in the catalytic combustion chamber before entering the cathode. Table 8 shows the settings used in the simulation of the components in the HYSYS simulator.

2.4. Super-critical CO₂ Brayton cycle

In this study, a closed-loop recuperated supercritical CO₂ Brayton cycle is designed inaugurated as bottoming cycle of the MCFC. Figure 5 displays the CO₂ bottoming cycle coupled with the MCFC, whereby the energy stream is supplied using the exiting stream from the catalytic burner, exchanged through the E101 heat exchanger, which then enters the cathode of the MCFC. The critical conditions of CO₂ occur at 30.98°C and 7.38 MPa [53]. The CO₂ has a higher efficiency compared to the ORC in the corresponding temperatures of the catalytic burner of the MCFC system.

2.5. Hybrid integration

Figure 6 exhibits the process flow diagram of the simulated hybrid system in the Aspen-HYSYS software environment. It is possible to define different efficiencies for the hybrid system. Typically, the overall system efficiency is defined as electrical efficiency output over chemical efficiency input. The net overall thermal efficiency, grounded on lower heating value (LHV) and higher heating value (HHV), is defined as

\[
\eta_{\text{Overall,LHV}} = \frac{W_{\text{Net}}}{m_{\text{fuel,in}} \times LHV_{\text{fuel}} - Q_{\text{recover heat}}}, \quad (19)
\]

where \( W_{\text{Net}} \) is the total net power produced by

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Figure 6. Process flow diagram of the proposed simulated hybrid system.
Figure 7. The simulated model results in comparison with experimental data from [59].

Figure 8. (a) The composite curves of the three main converters (C1, C2 and C3) in the AR MRI process and (b) composite curve of the whole process [17].
Table 10. Stream properties of the simulated system.

| Stream | Pressure (kPa) | Temperature (°C) | Mass flow (kg/h) | Vapor/phase fraction | Molar enthalpy (kJ/kgmole) |
|--------|---------------|------------------|------------------|----------------------|-----------------------------|
| NG     | 300           | 20               | 15505            | 1                    | -77264                      |
| water  | 101.4         | 20               | 53225            | 0                    | -286756                     |
| 4      | 300           | 124.3            | 68830            | 0.98                 | -200355                     |
| air    | 101.4         | 15               | 423249           | 1                    | -298.5                      |
| 3      | 271.4         | 138.2            | 423249           | 1                    | 3331                         |
| 5      | 300           | 298              | 68830            | 1                    | -193029                     |
| 7      | 271.4         | 445.4            | 68830            | 1                    | -144640                     |
| 6      | 271.4         | 600              | 68830            | 1                    | -138427                     |
| 9      | 271.4         | 907.7            | 96353            | 1                    | -32213                      |
| 1      | 300           | 20.01            | 53225            | 0                    | -286751                     |
| 2      | 300           | 134              | 53225            | 1                    | -238297                     |
| 12     | 271.4         | 885.9            | 0                | 0.98                 | -200355                     |
| 8      | 271.4         | 907.7            | 96353            | 1                    | -32213                      |
| Cathode in |        | 878.6            | 1386603          | 1                    | -21813                      |
| 13     | 271.4         | 699.1            | 1386605          | 1                    | -22447                      |
| 10     | 271.4         | 813.7            | 1386605          | 1                    | -17290                      |
| 1b     | 20730         | 650.02           | 132029           | 1                    | -238297                     |
| 5b     | 22500         | 118.8            | 132029           | 1                    | -395973                     |
| 2b     | 8131          | 557.8            | 132029           | 1                    | -369801                     |
| 4b     | 7500          | 35               | 132029           | 1                    | -398453                     |
| 14     | 271.4         | 699.1            | 0                | 0                    | -22447                      |
| Turbine outlet | 101.4   | 824              | 461553           | 1                    | -49466                      |
| 17     | 101.4         | 598.8            | 461553           | 1                    | -57502                      |
| 16     | 271.4         | 1001.3           | 0                | 0                    | -42918                      |
| Turbine inlet | 271.4    | 1001.3           | 461553           | 1                    | -42918                      |
| 15−    | 271.4         | 708.3            | 459327           | 1                    | -42279                      |
| 15+    | 271.4         | 700              | 927277           | 1                    | 498.6                       |
| 11a    | 1300          | 32.02            | 745200           | 0                    | -230800                     |
| 12a    | 1300          | 123.2            | 745200           | 0.02                 | -222875                     |
| 8a     | 1300          | 37.02            | 655775           | 0                    | -252847                     |
| 1a     | 1300          | 45.48            | 89425            | 1                    | -45883                      |
| 7a     | 1300          | 148.5            | 655775           | 0                    | -243790                     |
| 8a'    | 120           | 37.27            | 655775           | 0                    | -252847                     |
| 2a     | 1300          | 33.97            | 89425            | 0                    | -66321                      |
| 3a     | 1300          | -10              | 89425            | 0                    | -69949                      |
| 6a     | 130           | 28.97            | 89425            | 1                    | -45826                      |
| 9a     | 120           | 47.09            | 745200           | 0.095                | -226944                     |
| 10a    | 120           | 31.91            | 745200           | 0                    | -230829                     |
| w2     | 1000          | 25               | 4976957          | 0                    | -285817                     |
| w1     | 1000          | 30               | 4976957          | 0                    | -286205                     |
| w4     | 100           | 7559225          | 0                | 0                    | -285832                     |
| w3     | 100           | 25               | 7559225          | 0                    | -286221                     |
| 4a     | 130           | -27.90           | 89425            | 0.06                 | -69949.3                    |
| 5a     | 130           | -27.61           | 89425            | 0.93                 | -49454.6                    |
| 13a    | 1300          | 123.2115         | 745200           | 0.02                 | -222875                     |
| Burner inlet | 271.4  | 20               | 2226.2           | 1                    | -77258.7                    |
| outlet | 101.3         | 235.3            | 461553           | 1                    | -69704.3                    |
| bb     | 101.5         | 512              | 461553           | 1                    | -60505.5                    |
| 3b     | 7813          | 134.7            | 132029           | 1                    | -391280                     |
| 6b     | 20730         | 475.4            | 132029           | 1                    | -374494                     |
| 11     | 271.4         | 886.0            | 1386603          | 1                    | -20978.2                    |

Table 11. Efficiency of the hybrid system.

| Parameter                           | Value |
|-------------------------------------|-------|
| Net overall thermal efficiency (LHV) | 85    |
| Electrical efficiency (LHV)         | 53    |
| MCFC efficiency (LHV)               | 35    |

System components, $m_{\text{fuel,in}}$ is the mass flow rate of natural gas to the system and $Q_{\text{recover, heat}}$ is the recovered heat in the system components. Also, the MCFC efficiency, based on LHV and HHV, is defined as [32]

$$\eta_{\text{MCFC,LHV}} = \frac{W_{\text{DC}}}{(m_{\text{fuel,in}} \times LHV_{\text{fuel}})_{\text{Anode inlet}}},$$

(20)
Table 12. Definition of exergy destruction in components and corresponding exergy efficiency.

| Component          | Exergy destruction                                                                 | Exergy efficiency                                                                 |
|--------------------|------------------------------------------------------------------------------------|-----------------------------------------------------------------------------------|
| Compressor and pump| $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) + W - \sum (m_0 \cdot e_0)$               | $\epsilon = \frac{\sum (m_i \cdot e_i)}{W} - \sum (m_0 \cdot e_0)$             |
| Turbine            | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - W - \sum (m_0 \cdot e_0)$               | $\epsilon = \frac{\sum (m_i \cdot e_i)}{W} - \sum (m_0 \cdot e_0)$             |
| Expansion valve    | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i)}{W} - \sum (m_0 \cdot e_0)$             |
| Heat exchangers    | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i)}{W} - \sum (m_0 \cdot e_0)$             |
| Tower              | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i)_{\text{output}} - \sum (m_0 \cdot e_0)_{\text{input}}}{\sum (m_i \cdot e_i)}$ |
| Mixer and Separator| $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i)_{\text{output}} - \sum (m_0 \cdot e_0)_{\text{input}}}{\sum (m_i \cdot e_i)}$ |
| Fuel cell          | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i)_{\text{output}} - \sum (m_0 \cdot e_0)_{\text{input}}}{\sum (m_i \cdot e_i)}$ |
| Air cooler         | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) + e_{ai} + W - \sum (m_0 \cdot e_0) + e_{ao}$ | $\epsilon = \frac{\sum (m_i \cdot e_i)_{\text{output}} + e_{ai} + W}{\sum (m_i \cdot e_i)}$ |
| Reformer           | $I = Ex_i - Ex_0 = \sum (m_i \cdot e_i) - \sum (m_0 \cdot e_0)$                  | $\epsilon = \frac{\sum (m_i \cdot e_i) + e_{ao}}{\sum (m_i \cdot e_i)}$          |
| Hybrid system*     | $\epsilon = 1 - \frac{\text{Total system irreversibility}}{\text{Total system power consumption}}$ | $\epsilon = \frac{\sum (m_i \cdot e_i)_{\text{output}} + e_{ai} + W}{\sum (m_i \cdot e_i)_{\text{input}}}$ |

*The sum of the irreversibility in all components.

Figure 9. (a) V-I and P-I curves of the MCFC, (b) current density effect on MCFC power density and efficiency and (c) the current density effect on molar flow rate of fuel and air.
Table 13. Exergy rates of process streams and flows.

| Stream | Total exergy (kW) | Physical exergy (kW) | Chemical exergy (kW) | Stream | Total exergy (kW) | Physical exergy (kW) | Chemical exergy (kW) |
|--------|------------------|---------------------|---------------------|--------|------------------|---------------------|---------------------|
| 1      | 2572.0           | 6.7                 | 2565.3              | 11a    | 11273.0          | 394.2               | 11269.1            |
| 2      | 12260.1          | 9694.8              | 2565.3              | 12a    | 114139.2         | 14479.5             | 11269.1            |
| 3      | 12461.9          | 12035.1             | 426.8               | 13a    | 114139.2         | 1479.5              | 11269.1            |
| 4      | 221838.2         | 8824.3              | 213014.0            | 8a     | 631751.8         | 202.2               | 631549.6          |
| 5      | 224728.1         | 532228.6            | 1421677.7           | 01     | 1580882.4        | 202.2               | 1580882.4         |
| 6      | 242867.9         | 18633.6             | 224234.3            | 02     | 1581446.0        | 1710.3              | 1581446.0         |
| 7      | 237946.6         | 13712.3             | 224234.3            | 03     | 1313596.2        | -3.1                | 1313599.4         |
| 8      | 2153905.1        | 532228.6            | 1421677.7           | 04     | 1595566.9        | 33981.2             | 1595566.9         |
| 9      | 1953906.3        | 532228.6            | 1421677.7           | 05     | 1594610.6        | 33981.2             | 1594610.6         |
| 10     | 1935956.1        | 532228.6            | 1421677.7           | 06     | 1590864.8        | 33981.2             | 1590864.8         |
| 11     | 1892639.9        | 532228.6            | 1421677.7           | 07     | 1581351.7        | 19369.5             | 1581351.7         |
| 12     | 1965174.5        | 532228.6            | 1421677.7           | 08     | 1581446.0        | 1710.3              | 1581446.0         |
| 13     | 2111483.7        | 680.1               | 210468.2            | 09     | 1581351.7        | 19369.5             | 1581351.7         |
| 14     | 224728.1         | 532228.6            | 1421677.7           | 10     | 1576913.1        | 394.2               | 1576913.1         |

where \( W_{DC} \) is the power produced in the fuel cell and \((m_{fuel,in} \times LHV_{fuel})_{Anode \ inlet}\) is the mass flow of fuel fed into the anode side of the fuel cell. The total electrical efficiency of the hybrid system is defined as

\[
elec_{final}(LHV) = \frac{W_{MCFC} + W_{Brayton} + W_{GT} - W_{Com}}{m_{fuel} \times LHV_{fuel}}, \tag{21}
\]

where \( W_{MCFC}, W_{Brayton} \) and \( W_{GT} \) stand for the power produced in the fuel cell, S-CO\(_2\) Brayton cycle and GT and \( W_{Com} \) is the power used by the compressor.

2.6. Exergy analysis

The main approach in this investigation is to reduce power consumption and increase overall system efficiency. The first step to reducing the energy needed is to check the locale and the amount of energy lost due to component and process irreversibility. The analysis of the first law of thermodynamics (energy analysis) is extensively employed to appraise thermodynamic systems. The exergy analysis method is used to measure the irreversibility in the process and the component, which is entailed in the second law of thermodynamics.

Exergy is defined as the highest acquired useful work, derived from a stream as it is brought to reference environmental conditions. In a system, exergy is classified into four main components: physical, chemical, potential and kinetic exergy. Herein, the potential and kinetic exergy is considered negligible as the system...
| Component            | Exergy destruction rate (kW) | Exergy efficiency |
|----------------------|------------------------------|-------------------|
| Heat exchanger       |                              |                   |
| E-100                | 18043                        | 0.816             |
| E-101                | 794                          | 0.9996            |
| E-102                | 902                          | 0.965             |
| E-103                | 1194                         | 0.980             |
| Regenerative HX      |                              |                   |
| E-105                | 33849                        | 0.499             |
| E1                   | 654                          | 0.9995            |
| E2                   | 2568                         | 0.999             |
| E3                   | 1696                         | 0.999             |
| E4                   | 6441                         | 0.987             |
| RHX                  | 510                          | 0.999             |
| Turbine              |                              |                   |
| GT                   | 3919                         | 0.892             |
| K-100                | 637                          | 0.864             |
| Compressor           |                              |                   |
| Air comp.            | 2783                         | 0.812             |
| K-101                | 397                          | 0.808             |
| Generator            |                              |                   |
|                      | 47087                        | 0.988             |
| Splitter             |                              |                   |
|                      | 15663                        | 0.992             |
| Combustion chamber   |                              |                   |
|                      | 12090                        | 0.916             |
| Mixer                |                              |                   |
| mix100               | 1570                         | 0.993             |
| mix102               | 30412                        | 0.985             |
| p1                   | 0                            | 0.979             |
| Pump                 |                              |                   |
| p1-2                 | 3                            | 0.991             |
| Anode                |                              |                   |
|                      | 85427                        | 0.958             |
| Cathode              |                              |                   |
|                      | 78438                        | 0.960             |
| Reformer             |                              |                   |
|                      | 18138                        | 0.925             |
| Catalytic burner     |                              |                   |
|                      | 43316                        | 0.978             |
| MCFC                 | 100674                       | 0.530             |

3. RESULTS AND DISCUSSION

The proposed hybrid system is simulated in Aspen HYSYS, and herein, the results of the analysis are expressed and thereafter delineated through a sensitivity analysis and an exergy analysis. Throughout the study, an attempt was made to measure the validity of the results by comparing the simulation results with the actual results, and effort was put through to attain the necessary information for the operation of the system from reliable sources. The simulated MCFC in the present study is compared with the experimental data from a study by Milewski et al. [55] and presented in Figure 7. As observed, the results of the simulated model are consistent with experimental results, indicating that the model for electrochemical calculations is validated in this study. With regard to the inputs expressed for the system, the simulation results obtained from the MCFC performance are expressed in Table 9. The stream properties of the hybrid system simulation are shown in Table 10. Figure 8 illustrates the composite curves of the heat exchangers (C1, C2 and C3) in the AR-DMR process as well as the combined curve of the entire process. Due to the fact that the temperature difference is the impetus force of heat transfer, by reducing the temperature difference, the driving force of the heat transfer decreases and accordingly the required heat transfer area increases. Therefore, the effects of both factors must be addressed to achieve the optimal condition. The combined composite curve also shows the thermal design of the whole process according
to the number of available heat exchangers and the process of temperature changes during the process. The AR cycle provisions a 28.3 MW of cooling load for the DMR cycle.

The fuel entering the reformer participates 40% in the reaction and produces the hydrogen needed at the anode. The output flow from the anode and the inlet air to the system are burned before entering the cathode to produce carbon dioxide and meet the required temperature of the cathode input flow. On the other hand, since the output flow from the combustion chamber has a higher thermal load than the cathode requires, it is used to generate power and intensify the system's overall efficiency in the SCO\textsubscript{2} cycle.

The most significant factor affecting the performance of the MCFC include its temperature and current density. As the temperature of the cell increases, its performance increases correspondingly. A suitable range for MCFC performance is between 920\degree K and 1020\degree K; increasing the temperature of the cell to more than 1020 K improves efficiency but leads to depreciation and an increase in costs. At temperatures lower than 880\degree K, the performance of the cell decreases dramatically. Increasing the density reduces the output voltage and reduces the efficiency of the cell. Although a decrease in density leads to an increase in cell efficiency, its excessive reduction also results in a significant reduction in power density and, consequently, a sharp increase in the MCFC surface area, underscoring the increase in economic costs. The temperature of the exhaust gases from the combustion chamber and the system are 830\degree C and 195\degree C, respectively. The efficiencies of the hybrid system are tabulated in Table 11 as delineated in hybrid integration.

3.1. Sensitivity analyses
In order to reach an optimal design, it is essential to examine the effect of each variable on the system's performance through a sensitivity analysis. Sensitivity analysis scrutinizes the impact of various variables and the way they affect the output data of

![Figure 10](https://example.com/image10.png)

(a) Voltage and power density effect on MCFC performance and (b) UF effect on MCFC voltage.
the simulated system. The effects of various factors such as temperature, pressure and electric current are investigated on power, voltage and efficiency. The sensitivity analyses of some variables are presented herein.

Figure 9a displays the MCFC voltage and power density performance in terms of current density, based on the surface and temperature of the fuel cell. With the increase in current density, the voltage of the fuel cell reduces due to irreversible losses. The power density rises by increasing the current density and reaches a maximum value and then reduces. The maximum power density value is 0.76 W.cm$^{-2}$, which occurs at a current density of 1.3 Acm$^{-2}$. For a 0.6 Acm$^{-2}$ current density, the power density and voltage are 0.56 W.cm$^{-2}$ and 0.88 V, individually.

Figure 9b and c denotes the changes in the efficiency, the power density and the flow rate of fuel and air entering the fuel cell in terms of current density. Through the increase in current density, the cell efficiency is reduced due to irreversible losses, and as a result, a lower portion of the heat generated via the electrochemical reaction is converted to electricity, and the rest is released as free heat. This phenomenon leads to a reduction in power output, and consequently, more air is needed to remove heat from the cell and cool it down.

Figure 10a shows the temperature effect on cell performance, namely the voltage and the power density, assuming UF = 0.9. As observed, the performance of the cell is greatly reduced at lower temperatures, where the resistance of ion movement from the electrolyte is increased and, as a result, yields a decrease in fuel cell performance.

Figure 10b shows the fuel UF effect on the operating voltage of the molten carbonate fuel cell at 699.1°C. The more fuel is used, the greater the loss of activation and concentration loss, which leads to a decrease in voltage and consequently a decrease in fuel cell performance. The temperature and fuel UF are central
parameters that should be prudently selected. By aggregating the temperature of the cell, the cell efficiency increases, but consequently, the cost increases too. At first, the higher the number of fuels, the higher the efficiency of the cell, but at UFIs higher than 0.85, the concentration decreases, resulting in a decrease in the efficiency of the MCFC. In addition, an excessive increase in the fuel UF will result in higher heat production in the fuel cell and, as a result, increases the required air for cooling the cell. An increase in air leads to an increase in compressor power consumption and a reduction of the hybrid system’s thermal efficiency.

The results of the sensitivity analysis of the reformer temperature are shown in Figure 11a. The graph displays that hydrogen and carbon dioxide production increase with increasing temperature.

As evident, the gradient of the changes is high at first, and at a temperature of \(\sim 700^\circ C\), these changes decelerate and ultimately reach a constant molar flow. The reaction of hydrogen production is endothermic, so increasing its production speed is predictable with increasing temperature. In this study, concerning the conversion rate required for methane in the reformer, the temperature of 600\(^\circ C\) was selected for the reformer output.

As shown in Figure 11b, the simulation MCFC pressure has little effect on cell power density, electrical efficiency and voltage; only at initial pressures do these increase through increasing pressure. Nevertheless, this trend is slightly different from the thermodynamic theory of the pressure effect on fuel cell performance, attributable to the fact that with the increase in pressure, the voltage of the fuel cell should rise. This difference between experimental and simulated results can be discerned through the limitations of the simulator.

As manifested in Figure 12a, by reducing the flow rate of the bottom product to the feed, the required feed flow rate decreases immensely. As the ratio decreases from 0.94 to 0.78, the required feed flow rate is reduced to about one-fourth of the initial value. This typifies that a significant block of the cooling cycle, containing generator, absorber, pump, feed preheater heat exchanger and its pertinent equipment will be considerably reduced and, consequently, the total cost of the cycle will be meaningfully reduced. Alternatively, as shown in Figure 12b, the changes in the required heating (\(Q_h\)) and cooling (\(Q_c\)) energy consumption are insignificant compared with the feed flow reduction. Indeed, the heating and cooling energies reduce with a decrease in the bottom product flow ratio of the tower to the feed of 0.88 with a slight slope and then increases slightly.

By considering the minimum temperature being equal to 5\(^\circ C\) for the reboiler, the output temperature of it cannot exceed 5\(^\circ C\). In contrast, by decreasing the ratio of the bottom product flow of the tower to feed, the water component in the bottom product of the tower increases, and, as a result, the temperature of the bottom product of the tower increases. Henceforth, caused by the temperature limitation at the bottom of the tower, and according to Figure 12c, which shows the effect of bottom feed ratio on reboiler temperature, the ratio of the bottom product of the tower to feed is considered equal to 0.8.

3.2. Results of exergy analysis

Table 13 shows the aforementioned exergy classifications for flows, namely process streams, heat and work flows. Table 14 illustrates the exergy destruction and efficiency in the components of the hybrid system. In correspondence, Figure 13a illustrates the share of each component from the overall exergy destruction of the hybrid system, while Figure 13b depicts the exergy efficiencies in each block of the hybrid system. As evident, the MCFC has the largest share of the hybrid system exergy destruction and has the lowest exergy efficiency compared with other blocks. The highest exergy efficiency is pertinent to the E1 heat exchanger that is in
concord with the DMR block efficiency (0.996), and the fuel cell has the lowest, being 0.53. The SCO\textsubscript{2} cycle has the highest exergy efficiency among the blocks of the hybrid system. According to the results, the fuel cell must be improved based on the exergy analysis and needs improvement.

4. CONCLUSION

This study proposes a hybrid system comprising the DMR LNG production process, AR precooling cycle, MCFC and SCO\textsubscript{2} bottoming cycle. The investigation fulfilled the purpose of reducing power consumption and effectively utilizing energy. The inauguration of the AR cycle in the DMR process, through conveying a 28.3 MW of cooling load, provisions the $-33^\circ$C temperature needed in the pre-cooling stage. This increased the cooling load of the second cycle by 10% due to the limitation of the refrigerant temperature of the AR system due to the limitation of the present study for pressures above the atmospheric pressure. The most important parameter affecting the performance of the MCFC include its temperature and current density. Due to the coupling of the MCFC system with the AR cycle, the operating temperature of the cell must be considered in such a way as to provide the required thermal load of the AR cycle. Grounded on the
sensitivity analysis, the appropriate range of the MCFC output voltage is 0.8–1.2 V.

The electrical efficiency of the hybrid system at the temperature of 700°C, S/C of 3 and fuel density of 0.35 Acm⁻² is 53% (HHV). The COP of the AR cycle is 0.45. The amount of electricity consumed in the hybrid system is reduced and the amount of water used for cooling is increased. Using the exergy analysis, the most important locales in which losses occur were identified that were cathode, anode, combustion chamber and AR cycle generator.

In sum, the hybrid system produces 6300 kg mol⁻¹ of LNG, 146.55 MW of electrical power, 110 MW from the MCFC, 32.5 MW from the GT and 4.05 MW from the SCO₂ cycle. The overall energy and exergy efficiencies of the hybrid system were 85% and 95%, individually. To conclude, this study is an effective step in designing and evaluating the integration of the LNG process, AR cycles, MCFCs and SCO₂ bottoming cycle, and the results show that the proposed hybrid system is operable and feasible.

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