

Research Paper

Investigation of a solid oxide fuel cell integrated into an internal combustion engine with carbon capture for maritime applications

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ABSTRACT

The significance of limiting greenhouse gas emissions has grown in the maritime sector and regulations in this regard have become increasingly stringent. This leads to the need for new and more environmentally friendly power sources, and technologies to be used in ships. Fuel cells and carbon capture systems can enable ships to both meet the International Maritime Organization's emissions regulations and contribute to a sustainable environment by reducing greenhouse gas emissions. In this study, the solid oxide fuel cell is electrochemically, and thermodynamically modeled and integrated into the internal combustion engine for maritime applications. Reformed natural gas is used as fuel, so the steam reforming and the carbon capture and storage systems are also integrated into the solid oxide fuel cell and internal combustion engine power system. Integrated systems are simulated and energy, exergy, economy, and environmental analyses are carried out. The effects of solid oxide fuel cell operating temperature, current density, operating time due to degradation, and the use of hydrogen instead of natural gas on analyses results are investigated. The findings obtained in this study showed that the proposed integration could reduce carbon dioxide emissions by 76.1 per cent, while the hydrogen-fuelled alternative system could reduce carbon dioxide emissions by 96.9 per cent at an operating lifetime of 35,000 h for the solid oxide fuel cell. Given the fuel cell degradation, the average overall energy efficiencies of the systems were 44.1 per cent and 57.5 per cent, respectively, while the cost of the proposed system was 41.1 per cent lower than the hydrogen-fuelled alternative. Hydrogen-fuelled alternative system was closer to the net zero target in emission and had higher efficiencies, the proposed system might be a more effective solution in the transition to the net zero target because of its lower volume and mass requirements and lower costs.

1. Introduction

The International Maritime Organization has significantly accelerated its work since 2008 and has set rules and regulations in several phases to limit the CO₂ emissions from international maritime activities. As a result of the 72nd Marine Environment Protection Committee meeting held in 2018 agreed that CO₂ emissions emitted from ships will be reduced by 50 % by 2050 compared to 2008 [1]. As a result of the meetings held later, it has been decided to take this implementation date even earlier than 2050 in various phases for different ship types and to classify the ships according to the CO₂ rate ships emit [2]. At the 80th MEPC meeting in July 2023, it was decided to reduce CO₂ emissions per transport work from international maritime transport by at least 40 % in 2030 compared to 2008 [3]. It was also decided to adopt technologies, fuels and/or energy sources with zero or near-zero greenhouse gas

emissions, accounting for at least 5 % and ideally 10 % of the energy consumed by international maritime transport by 2030, and to achieve net zero greenhouse gas emissions in or around 2050.

Despite their high cost, fuel cells are more efficient, cleaner, and quieter than conventional internal combustion ship engines. For these reasons, the use of fuel cells is promising in the fight against climate change and has started to attract worldwide attention in recent years for ships [4]. The first applications related to the use of fuel cells on ships were carried out for submarines, and since the 2000 s, researchers have begun to examine the use of fuel cells in power systems for commercial ships. Although the use of various types of fuel cells in ships has been practiced and recommended, it is predicted that solid oxide fuel cells (SOFC) will be a better option for ships due to their high waste heat utilization potential and high efficiency when compared the performance of the cells [5]. A study showed that the proton exchange membrane fuel cell type, which is the most widely used, does not

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Nomenclature**Acronyms**

4E	Energy, exergy, economy, and environmental
BO-NG	Boil-off natural gas
CAPEX	Capital expenditure
CCS	Carbon capture and storage
CEPCI	Chemical engineering plant cost index
DWT	Deadweight ton
EEDI	Energy efficiency design index
FW	Freshwater
HEX	Heat exchanger
ICE	Internal combustion engine
LCC	Life cycle cost
LNG	Liquefied natural gas
MCFC	Molten carbonate fuel cells
MCR	Maximum continuous rate
MDEA	Methyl diethanolamine
MESH	Mass, thermodynamic equilibrium, stoichiometry relations, and heat balance
NG	Natural gas
OPEX	Operational expenditure
PCC	Post-combustion capture
PZ	Piperazine
RMSE	Root mean square error
SGC	Specific gas consumption
SOFC	Solid oxide fuel cell
SPOC	Specific pilot oil consumption
SR	Steam reformer
SW	Seawater
TC	Turbocharger
<i>C</i>	Cost, \$
<i>D</i>	Diffusion coefficient, m ² /s
<i>E</i> ⁰	Reversible potential at standard conditions, V
<i>F</i>	Faraday's constant, C/mol
<i>f</i>	Factor
<i>h</i>	Voltage loss coefficient
<i>i</i>	Current density, A/cm ²

<i>in</i>	Interest rate
<i>K</i>	Equilibrium constant
<i>M</i>	Molecular weight, g/mol
<i>m</i>	Mass rate, kg/s
<i>n</i>	Electron number of fuel
<i>ne</i>	Economic life of the ship, year
<i>P</i>	Partial pressure, kPa
$\bar{R}$	Universal molar gas constant, J/(K.mol)
<i>r</i>	Pore diameter, μm
<i>T</i>	Temperature, K
<i>t</i>	Time, h
<i>V</i>	Potential, V
$\Delta\bar{g}$	Gibbs free energy, J
δ	Thickness, mm
ϵ	Porosity
ζ	Tortuosity
ρ	Specific resistivity, Ω.m
<i>v</i>	Fuller diffusion volume

Subscripts

0	Exchange
<i>act</i>	Activation
<i>an</i>	Anode
<i>cat</i>	Cathode
<i>con</i>	Concentration
<i>con.</i>	Conventional power system
<i>eff</i>	Effective
<i>el</i>	Electrolyte
<i>fr</i>	Freight
<i>ini.</i>	Initial
<i>int</i>	Interconnection
<i>K</i>	Knudsen
<i>new</i>	Proposed power system
<i>O</i>	Ordinary
<i>ohm</i>	Ohmic
<i>P</i>	Equilibrium number
<i>pur</i>	Purchase
<i>sol</i>	Solvent

support long sea voyages as power generation equipment. Instead, SOFC power systems effectively reduce emissions and meet mission requirements for months at a satisfactory level [5]. In addition, the SOFC power system can achieve higher efficiencies with bottoming and combined cycles [6]. Kistner et al. [7] investigated the challenge of providing an uninterruptible power supply on board a ship by considering the power gradient limitation of SOFC modules used in ships with hybrid energy storage. In addition to the oversized calculation of aggregates for safety reasons, the biggest challenges in the study were reported to be the variable ship power system load profiles due to the unique nature of the ships and the lack of operational data. Van Veldhuizen et al. [8] investigated the effects of different fuel utilization and cathode-off gas utilization on net electricity and combined heat and power efficiencies in marine SOFC systems. In the study, it was found that without cathode-off gas recirculation, the available waste heat was insufficient to generate some type of fuels to the saturated vapor phase and the recirculation increased the heat efficiencies between 11.9–105 % for different fuels. In all these studies, there are disadvantages arising from using only SOFC as prime mover. Unlike onshore installations, power and load changes on board can be abrupt and are more dynamic. In SOFC, the load change is slower than in an internal combustion engine (ICE). It is also more challenging to control the load variation in SOFC and it is possible to increase the degradation due to changing parameters. It is possible to overcome or reduce these disadvantages

using SOFC and ICE together on board and there are some recent studies in the literature that take this into account. A study for an Aframax-type crude oil tanker showed that using direct coupled hybrid SOFC and gas-fired ICE reduced both CO₂ emissions and exergy destruction when compared to indirect coupled of these systems [9]. In another study, the effects of dual fuel engines' and SOFC's contributions to total power generation in varying amounts on greenhouse gas emissions and power generation system costs were investigated for a tanker and a cruise ship [10]. Haseltalab et al. [11] proposed innovative energy and power management and component sizing for the use of a hybrid SOFC-battery-gas engine as a power source in liquefied natural gas (LNG)-fueled ships. This paper has shown that emissions can be reduced by a maximum of 53 % with an LNG-fueled hybrid power system instead of a conventional power supply on ships and with the proposed energy and power management approaches. In a study belong to Sapra et al. [12] on a combined power cycle using SOFC with gas-injected ICE on ships, it was proposed that the injection of unburned hydrocarbons in the anode gas into the ICE with the gaseous fuel of ICE and the effect of the proposed system on emissions was investigated. In addition to the various disadvantages overcome by ICE-SOFC integration, CO₂ emissions are also significantly reduced. However, if a carbon-free fuel is not used, it is not possible to reduce emissions to near zero with this integration alone. Moreover, none of these studies have considered the degradation of SOFC.

Another effective solution in fighting climate change is using carbon capture and storage (CCS) systems. CCS systems are widely used in land facilities and they are also recommended to be used for ships in the near future due to the CO₂ emission-reducing capability of up to 90 %. Luo and Wang [13] presented the first systematical study in applying solvent-based carbon capture for ships. Awoyomi et al. [14] suggest using ammonia as a solvent in a ship based CCS system and the combined capture of CO₂ and SO_x emissions. Feenstra et al. [15] examined the carbon capture feasibility and the cost of the ship CCS system, with the change in the amount of emissions due to the power demand and the use of LNG or diesel as fuel. Güler and Ergin [16] examined the effect of hydraulic design parameters of CCS systems and compared CCS systems' feasibility with different carbon control methods for ships. Ji et al. [17] also examined the effect of design parameters of CCS systems for ships and proposed using piperazine (PZ)-methyl diethanolamine (MDEA) blended amine as the solvent. Long et al. [18] investigated the effects of different absorber configurations and stripping pressure on marine CCS systems for a diesel fueled ship and also proposed using PZ-MDEA amine blend as solvent. CO₂ and SO_x capture effectiveness, energy efficiency design index (EEDI) value and energy penalty of a handymax tanker are analyzed by Stec et al. [19] at different ambient conditions. A novel methodology for calculating EEDI that considers the CCS system for ships is proposed by [20]. Another study proposed using membranes instead of solvent based CCS systems to reduce the equipment size for ships [21]. Einbu et al. [22] investigated the thermal assessment of the CCS system onboard and proposed to use an afterburner to meet the thermal requirements of the ship based CCS system. Ros et al. [23] discuss the results of the DerisCO₂ project which focused on de-risking of the ship based CCS system in their study. Güler and Ergin [24] investigated the integration of cooling, heating, and power systems of LNG carriers with the CCS system and showed that the integration of the CO₂ liquefaction system into the boil-off natural gas (BO-NG) reliquefaction system reduces the total life cycle cost (LCC) up to 17 %. In all these studies using CCS systems on ships, only ICE has been considered as the CO₂ source. However, the importance given to the reduction of CO₂ emissions and stricter emission regulations has recently increased the interest in alternative power system selection. Therefore, there is a gap in the literature on the use of CCS system with power system other than ICE for ships.

It has been observed that the integration of the CCS system with high-temperature fuel cells, such as SOFC and molten carbonate fuel cells (MCFC) gives promising results. Therefore, SOFC-CCS integration has been studied extensively since the water in the anode gas can be easily removed and CO₂ can be easily captured due to the increased CO₂ concentration [25]. Only the MCFC-CCS integration has been analyzed until today for ships, but the exergy and economic analyses of the integration are missing in this study [26]. In another study, SOFC-CCS integration has been suggested for ships but only the system's storage volume and mass requirements have been examined [27].

Analyses of SOFC and ICE integration for ships have been performed by various studies until today, but not all energy, exergy, economy, and environmental (4E) analyses have been performed in any study except this paper. Besides, the integration of SOFC, ICE, and CCS has not been studied in the literature until today for either ships or land facilities. This study is the first to propose this integration and to perform 4E analyses of it. The effect of SOFC current density and operating temperature on 4E analyses has been investigated in the literature before, but the degradation effect was not considered in this study [28]. On the other hand, model-based simulation applications that take into account the degradation have been carried out only for SOFC and gas turbine integration until today [29]. This study is the first model-based simulation study that considers long-term cell degradation and examines the results with 4E analysis for SOFC and ICE integration.

In this paper, the SOFC is electrochemically and thermodynamically modeled and integrated into the ICE for ships. Reformed natural gas (NG) is used as fuel, so the steam reforming (SR) and the CCS systems are

also integrated into the SOFC and ICE power systems. Integrated systems' thermodynamic models are simulated by Aspen HYSYS, and 4E analyses are carried out. The effects of SOFC operating temperature, current density, operating time due to degradation, and using hydrogen instead of NG on 4E are investigated. According to the results of the 4E analysis, the most suitable fuel cell operating temperature, current density, and operating time parameters are determined, and the effect of the losses in the fuel cell depending on the current density change is examined for the most suitable operating temperature condition. Finally, the 4E analysis results of the proposed system are compared with the results obtained when H₂ is used instead of methane as fuel. The process diagrams of the proposed SOFC-ICE-CCS system and alternative SOFC-ICE-steam turbine (ST) models are shown in Fig. 1.

2. Modeling of conventional power systems for studied ships

The efficiency of direct oxidation of fuels in SOFC is considerably reduced; therefore, the use of reformed fuel is considered in this study. On the other hand, as the carbon content of the fuel increases, the energy requirement, equipment size, and cost of the reforming process increase. Therefore, using NG as fuel instead of diesel fuel is more advantageous in reformation. However, using NG as fuel in ships has some problems such as cargo volume loss, increasing weight, and equipment cost due to storage and cooling. These disadvantages are eliminated in LNG carriers since these types of ships can use the LNG or BO-NG in cargo tanks as fuel. Furthermore, the ICE-CCS-SOFC integration will be more feasible for LNG carriers due to the possibility of CO₂ liquefaction with cold LNG utilization in CCS systems [30]. Finally, in a previous study by the authors, it was shown that the cost of using the CCS systems on LNG carriers is lower than other carbon control methods, as opposed to displacement-type ships like crude oil tankers [16]. Particulars of selected ships in line with this information are shown in Table 1 [31]. It is stated the power, specific gas consumption (SGC), and specific pilot oil consumption (SPOC) values are valid for Tier III NO_x emission conditions, ambient air and scavenging air coolant temperatures of 25 °C, exhaust back pressure of 3 kPa, lower calorific value of fuel gas of 50 MJ/kg, and lower calorific value of pilot oil of 42 MJ/kg by the engine manufacturer [32]. The mean effective pressures of engines are given as 21 bar at full load condition by the manufacturer [32]. Since the SGC and SPOC values at 75 % maximum continuous rate (MCR) of the main engine are considered in the EEDI calculation, the values at this load are given in Table 1. The molar component of the LNG is modeled as 90.1 % methane, 6.2 % ethane, 2.3 % propane, 1 % butane, and 0.4 % nitrogen as a typical Qatari gas [33].

The power requirements of the studied ships are provided by the gas-injected main engines, and their characteristics are given in Table 1 and the model of the main engines is shown in Fig. 2. According to Fig. 2, the LNG from the cargo tank is pressurized up to the injection pressure with a pump, gasified in the cold box, and injected into the engine's cylinders with pilot oil. On the other side, scavenging air is first pressurized in the turbocharger (TC) compressor, cooled with seawater, compressed in cylinders, and reacted with the fuel in the combustion chamber. The formed high-pressure flue gas expands in the cylinder, then flue gas is expanded again in the turbine of the TC and is released from the funnel to the atmosphere. In ICE, the engine cylinder and the lube oil must be cooled. For this purpose, this cooling process is modeled between the cylinder and the TC turbine, which is in line with the literature as seen in Fig. 2 [13].

The components and properties of the flue gas formed after combustion were determined using the Gibbs free energy equation [13]. The thermodynamic properties of all streams in the gas phase are estimated with the Peng-Robinson equation state [34]. The isentropic efficiencies of compression and expansion are assumed to be 85 % and 90 % for TC, 90 % and 95 % for main engine [13]. In line with these assumptions, the difference between the results obtained from the model and the manufacturer's data is analysed for the main engine of the Q-Flex vessel at 75

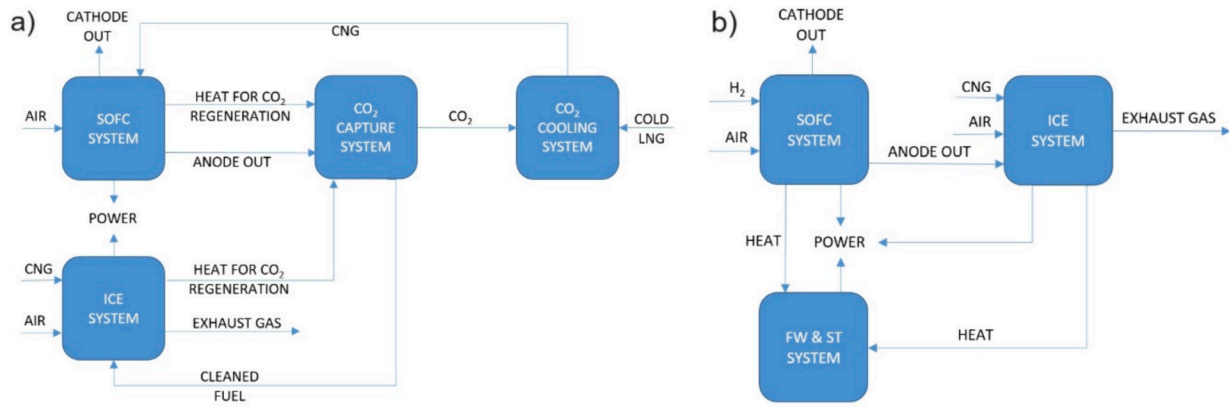


Fig. 1. Process diagrams of SOFC-ICE-CCS integration (a) and SOFC-ICE-ST integration (b).

Table 1

Main dimensions and particulars of studied ships.

Description	Unit	CLNG carrier	Q-Flex	Q-Max
Scantling cargo capacity	cbm	150,000	210,000	265,000
Scantling draught	m	12.3	12.7	12.7
Length overall	m	288	315	345
Length between perpendicular	m	275	303	332
Breadth	m	44.2	50	54
Design draught	m	11.6	12	12
Design ship speed at 100 % MCR	kn	20	20	20
Selected main engine		6S60ME-C10.6	7G60ME-C10.5	8G60ME-C10.5
Total main engine power at 100 % MCR	kW	2x14,940	2x19,880	2x22,720
SGC at 75 % MCR	g/kWh	135	135	135
SPOC at 75 % MCR	g/kWh	3.06	3.06	3.06
Jacket water heat dissipation at 75 % MCR	kW	2x1660	2x2120	2x2420
Lube oil heat dissipation at 75 % MCR	kW	2x850	2x970	2x1110
Dry engine weight	t	2x344	2x498	2x554

% MCR. It was calculated that the resulting power and exhaust gas temperature in Kelvin differ by 4.5 % and 0.4 %, respectively [32]. Thus, it seems that the results are in acceptable agreement.

3. Modeling of solid oxide fuel cell system

The system is basically studied in two parts: electrochemical and thermodynamic modeling of the SOFC.

The electrochemical model of the SOFC system is established using AVL Cruise software [35]. All equations used in the electrochemical model of SOFC are shown in Table 2. According to this table, Equations (1–4) show the reaction of hydrogen on the anode side, the reaction of oxygen on the cathode side, the reaction of carbon monoxide on the anode side, and the total reaction of hydrogen. The net potential of the fuel cell (V_{cell}) is determined by Equation (5). Here, V_{Nernst} , V_{act} , V_{ohm} , and V_{con} represent Nernst potential, activation losses, ohmic losses, and concentration losses, respectively. In Equation (6), E^0 , $\bar{R}$, n , F , P , and T show the reversible potential at standard conditions, universal molar gas constant, electron number of fuel, Faraday's constant, partial pressure, and temperature respectively. a_n and a_c indexes represent the anode and cathode sides. E^0 is defined by the Gibbs free energy ($\Delta\bar{g}$) and is expressed according to Equation (7). Gibbs free energy can be defined as a function of temperature as in Equation (8). The Tafel equation (9) is used to calculate the activation losses. Here, i and i_0 are the design and exchange current densities. The empirical expressions of the exchange

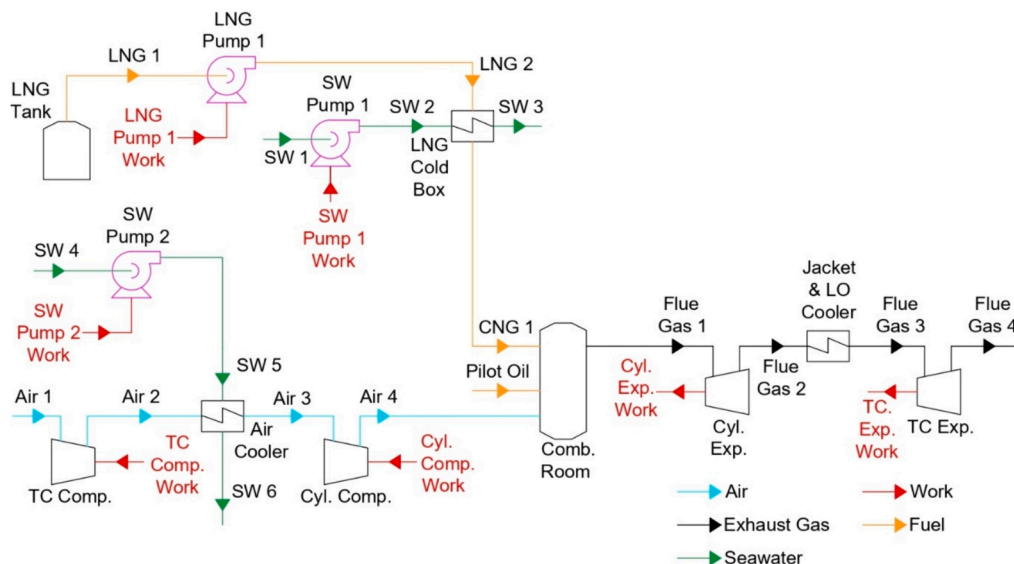


Fig. 2. Schematic model of internal combustion engine.

Table 2
Equations used in electrochemical modeling of SOFC.

Equation	Unit	References	Eq. no
$2H_2 + 2O^{2-} \rightarrow 2H_2O$	—	—	(1)
$O_2 + 4e^- \rightarrow 2O^{2-}$	—	—	(2)
$2CO + 2O^{2-} \rightarrow 2CO_2 + 4e^-$	—	—	(3)
$2H_2 + O_2 \rightarrow 2H_2O$	—	—	(4)
$V_{cell} = V_{Nernst} - V_{act} - V_{ohm} - V_{con}$	V	[38]	(5)
$V_{Nernst} = E^0 - \frac{\bar{R}T}{nF} \ln \left(\frac{P_{H_2O,an} P_{CO_2,an}}{P_{H_2,an} P_{O_2,cat} P_{CO,an}} \right)$	V	[38]	(6)
$E^0 = \frac{\Delta \bar{g}}{nF}$	V	[38]	(7)
$\Delta \bar{g} = 0.002474T^2 + 48.996T - 243730$	J/mol	[38]	(8)
$V_{act} = \frac{\bar{R}T}{nF} \ln \left(\frac{i^2}{i_{0,an} i_{0,cat}} \right)$	V	[39]	(9)
$i_{0,cat} = 2.9 \times 10^9 \left(\frac{P_{O_2}}{P_{cat}} \right)^{0.25} \exp \left(\frac{-1.2 \times 10^5}{\bar{R}T} \right)$	A/cm ²	[36]	(10)
$i_{0,an} = 7 \times 10^9 \left(\frac{P_{H_2}}{P_{an}} \right) \left(\frac{P_{H_2O}}{P_{an}} \right)^{-0.5} \exp \left(\frac{-1.2 \times 10^5}{\bar{R}T} \right)$	A/cm ²	[36]	(11)
$V_{ohm} = i(\rho_{an} \delta_{an} + \rho_{cat} \delta_{cat} + \rho_{el} \delta_{el} + \rho_{int} \delta_{int})$	V	[36]	(12)
$\rho_{an} = 2.98 \times 10^{-5} \exp \left(\frac{-1392}{T} \right)$	$\Omega.m$	[36]	(13)
$\rho_{cat} = 8.114 \times 10^{-5} \exp \left(\frac{600}{T} \right)$	$\Omega.m$	[36]	(14)
$\rho_{el} = 2.94 \times 10^{-5} \exp \left(\frac{10350}{T} \right)$	$\Omega.m$	[36]	(15)
$\rho_{int} = 1.257 \times 10^{-3} \exp \left(\frac{4690}{T} \right)$	$\Omega.m$	[36]	(16)
$V_{con,cat} = -\frac{\bar{R}T}{4F} \ln \left(1 - \frac{i}{i_{O_2}} \right)$	V	[40]	(17)
$V_{con,an} = \frac{\bar{R}T}{2F} \ln \left[\frac{\left(1 - \frac{i}{i_{H_2}} \right) \left(1 - \frac{i}{i_{CO}} \right)}{\left(1 + \frac{i}{i_{CO_2}} \right) \left(1 + \frac{i}{i_{H_2O}} \right)} \right]$	V	[40]	(18)
$i_{O_2} = \frac{4FPD_{cat,eff}}{\bar{R}T\delta_{cat}} \ln \left(\frac{p}{p - P_{O_2}} \right)$	A/m ²	[40]	(19)
$i_{j,an} = \frac{2FPD_{an,eff}}{\bar{R}T\delta_{an}}$	A/m ²	[40]	(20)
$D_{O,k-j(eff.)} = \frac{1 \times 10^{-7} T^{1.75} \epsilon \left(\frac{1}{M_j} + \frac{1}{M_k} \right)^{0.5}}{P \zeta \left(v_j^{1/3} + v_k^{1/3} \right)^2}$	m ² /s	[36]	(21)
$D_{K,j(eff.)} = \frac{97reT^{0.5}}{\zeta M_j^{0.5}}$	m ² /s	[36]	(22)
$D_{O,j(eff.)} = \frac{\left(1 - \frac{P_j}{P_{an}} \right)}{P_k \frac{P_l}{P_{an} D_{O,k-j(eff.)}} + \frac{P_l}{P_{an} D_{O,l-j(eff.)}} + \dots}$	m ² /s	[41]	(23)
$\frac{1}{D_{j(eff.)}} = \frac{1}{D_{O,j(eff.)}} + \frac{1}{D_{K,j(eff.)}}$	m ² /s	[36]	(24)
$D_{an,eff} = \left(\frac{P_j}{P_{an}} \right) D_{j(eff.)} + \left(\frac{P_k}{P_{an}} \right) D_{k(eff.)} + \dots$	m ² /s	[36]	(25)
$D_{cat,eff} = D_{O_2(eff.)}$	m ² /s	[36]	(26)
$V_{cell,t} = V_{cell,ini} (1 - h)^{\frac{t}{1000}}$	V	[37]	(27)

current densities of the SOFC for cathode and anode obtained from the experimental results are shown in Equations (10–11). Ohmic losses are caused by the internal resistance of the electrodes, electrolyte, and interconnectors and they are expressed as in Equations (12–16). Here, el and int indices represent electrolyte and interconnection elements, ρ and δ parameters represent the specific resistivity and thickness of a component. The thicknesses of the components are shown in Table s1 [36]. Concentration losses are dominant in regions with very high current density, and they are calculated as in Equations (17–18) for cathode and anode. Here, i_{O_2} , i_{H_2} , i_{H_2O} , i_{CO} , and i_{CO_2} represent the limit current densities for oxygen, hydrogen, water, carbon monoxide, and carbon dioxide, respectively. The limit current densities for these molecules are determined with Equations (19–20). In these equations, $D_{an,eff}$ and $D_{cat,eff}$ represent the effective diffusion coefficients of the anode and cathode

side, respectively. The diffusion coefficient is calculated by the harmonic average of ordinary and Knudsen diffusion coefficients. These calculations are shown in Equations (21–26). In these equations, M , ϵ , ζ , v , and r represent molecular weights, porosity, tortuosity, Fuller diffusion volume, and pore diameter, respectively. O and K indices are used for ordinary and Knudsen diffusion coefficients, j , k , and l indices are used for materials. The parameters used in the calculation of the diffusion and potential value at time t are shown in Equation (27). The ini index represents the initial value, h parameter represents voltage loss and this value is assumed to be %0.75/1000 h [37].

The electrochemical model was verified by an experimental study in the literature and the results with uncertainties are shown in Fig. 3 [42]. The experiment was carried out at 800 °C and atmospheric pressure. The fuel flow rate was 140 ml/min and the air flow rate was 550 ml/min and the fuel contained 56 % H₂ and 44 % CO. The composite of 50 wt% strontium-doped lanthanum cobaltite and 50 wt% Sm-doped CeO₂ interlayer on the cathode side and the Ni + YSZ interlayer on the anode side were both 20 μ m thick. The root mean square error (RMSE) value was calculated as 7.129 %. According to Fig. 3, it is seen that the error increases gradually after a value in the regions where concentration losses start to dominate. In practice, the operating current density of the SOFC is lower than the maximum power density. When the difference between the results and the model is analyzed for the region before the maximum power density, the RMSE value is calculated as 2.477 %.

The thermodynamic model of the SOFC system is established using Aspen HYSYS software, and the schematic diagram of the model is shown in Fig. 4 [43]. According to Fig. 4, compressed NG from liquefaction system preheated by the anode gas outlet. The freshwater (FW) taken from the tank is pressurized with a pump and vaporized with two-stage heating. The heating in the first stage is provided by the air coming from the cathode side of the SOFC, and the heating in the second stage is provided by the gas outlet from the anode side of the SOFC. The mixed steam and NG enter the SOFC from the anode after reforming. On the cathode side, the air, which is pressurized with a fan and heated by a regenerative heat exchanger (HEX), enters the SOFC. The power is generated in the SOFC due to reactions with fuel and air. The anode gas from the SOFC outlet, which is at a very high temperature, is first distributed by a flow-controlled tee. The anode gas then heats the fuel, FW going to the SR, and the CO₂-rich solvent from the CCS system to separate the CO₂ from the solvent. The anode gas contains a high amount of water vapor. If this water is removed from the circuit, it will be much easier to capture CO₂ due to the increased CO₂ concentration in the anode gas, and the costs and dimensions of the CCS system will be significantly reduced. For this reason, after the cooled anode gas passes through the collector, it is also cooled with seawater and passes through the separator to purify the anode gas from water to a large extent. The separated water is transferred to the FW storage tank and is recycled.

The zero-dimensional approach is used in the modelling of the SOFC and reformer. The reformer and SOFC are assumed to be a “blackbox” with defined conditions at the boundaries. It is assumed that the fuel and air outlet temperatures are equal to the SOFC operating temperature. The reformer is simulated as an equilibrium reactor and the SOFC as a conversion reactor. The energy balance is obtained by considering heat entering the reactors, the heat withdrawn from the reactors, generated power and the heat lost. The heat transfer from the hot anode gases to the reforming of steam and NG mixture is obtained using the energy balance given in Equation (s24).

The thermodynamic equations including mass, energy, entropy, exergy balances, and efficiencies are shown in Table s3. The thermodynamic balance equations of system components are shown in Table s4.

The temperature and pressure of ambient air are assumed to be 25 °C and 1 atm as specified in the conventional power system [32]. The FW and seawater temperature is also assumed to be 25 °C. The seawater outlet temperature is taken as 37 °C, which is the maximum temperature allowed by a classification society [44]. The pump efficiencies are

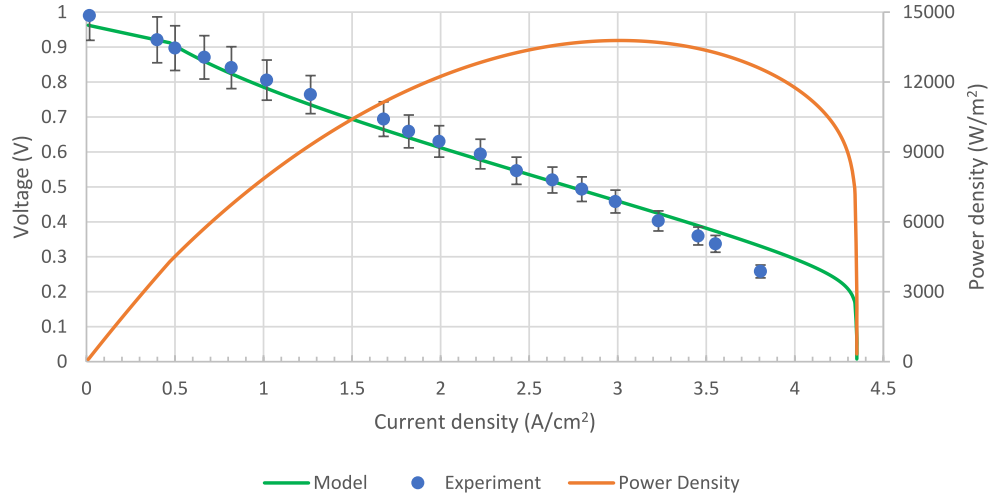


Fig. 3. Validation of the electrochemical model for SOFC.

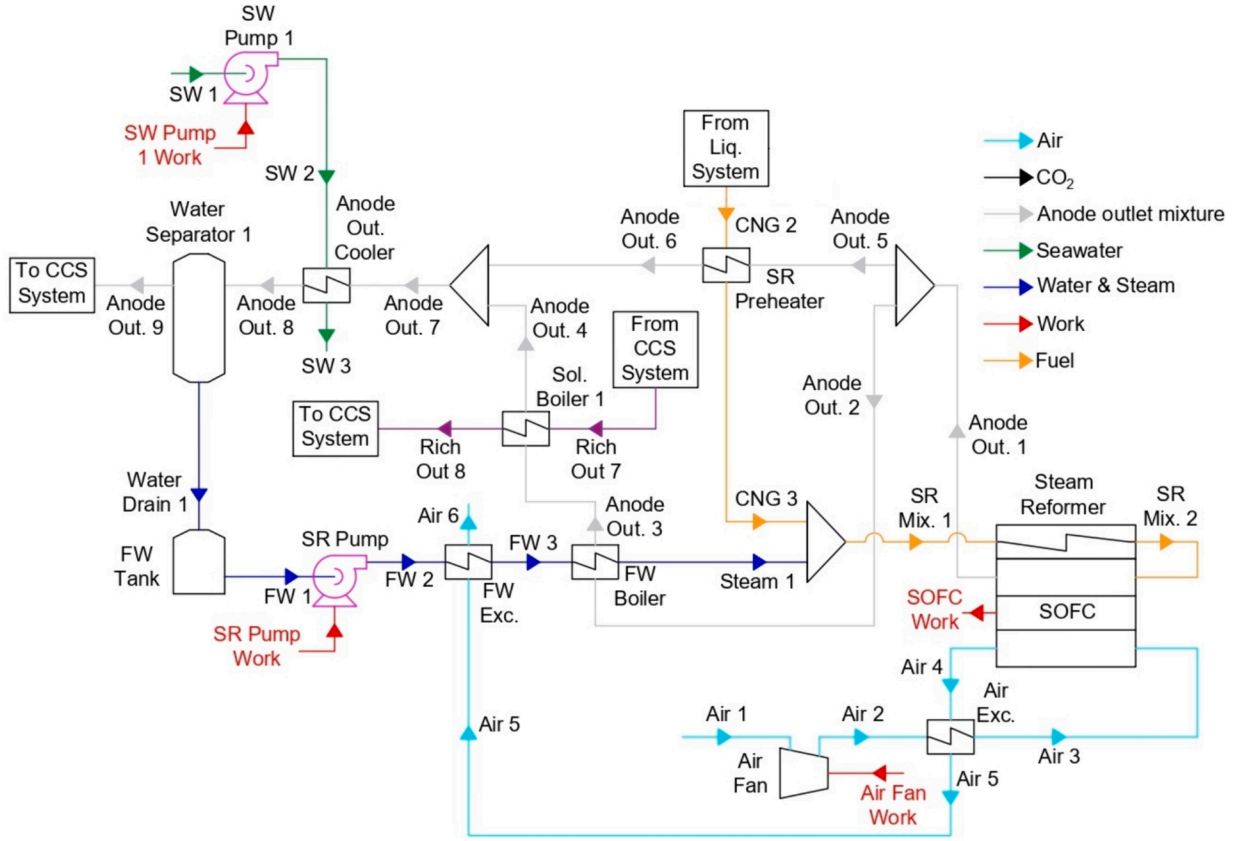


Fig. 4. Schematic model of SOFC system.

assumed to be 90 % for all systems in this study [36]. It is also assumed that all HEXs are adiabatic for all systems in this study [38]. In all HEXs, the pressure loss for the gas phase is assumed to be 5 kPa; for the liquid phase, boiling liquid and condensing steam is assumed to be 50 kPa. Inverter efficiency (η_{DC-AC}) is assumed to be 95 % [9]. SR and water gas shift reactions are modeled as equilibrium reactions and shown in Table s5. The K_p equilibrium constants for SR reactions are shown in Table s6 for 873–1173 K temperatures [45]. Also, the K_p equilibrium constant for the water gas shift reaction is shown in Table s7 for various temperatures [46].

The K_p values in the intermediate temperature values were deter-

mined with the following equation, and the relative maximum error is calculated as 0.7 % for all K_p values in Table s5.

$$\ln(K_p) = A + \frac{B}{T} + C \cdot \ln(T) + D \cdot T \quad (78)$$

The steam to carbon ratio is taken as 3 [38]. The amounts of CH_4 , CO , and H_2 formed due to SR at various temperatures were validated with the values in the literature, and the molar differences are shown in Table 3 [47]. It could be said that the results are in acceptable agreement.

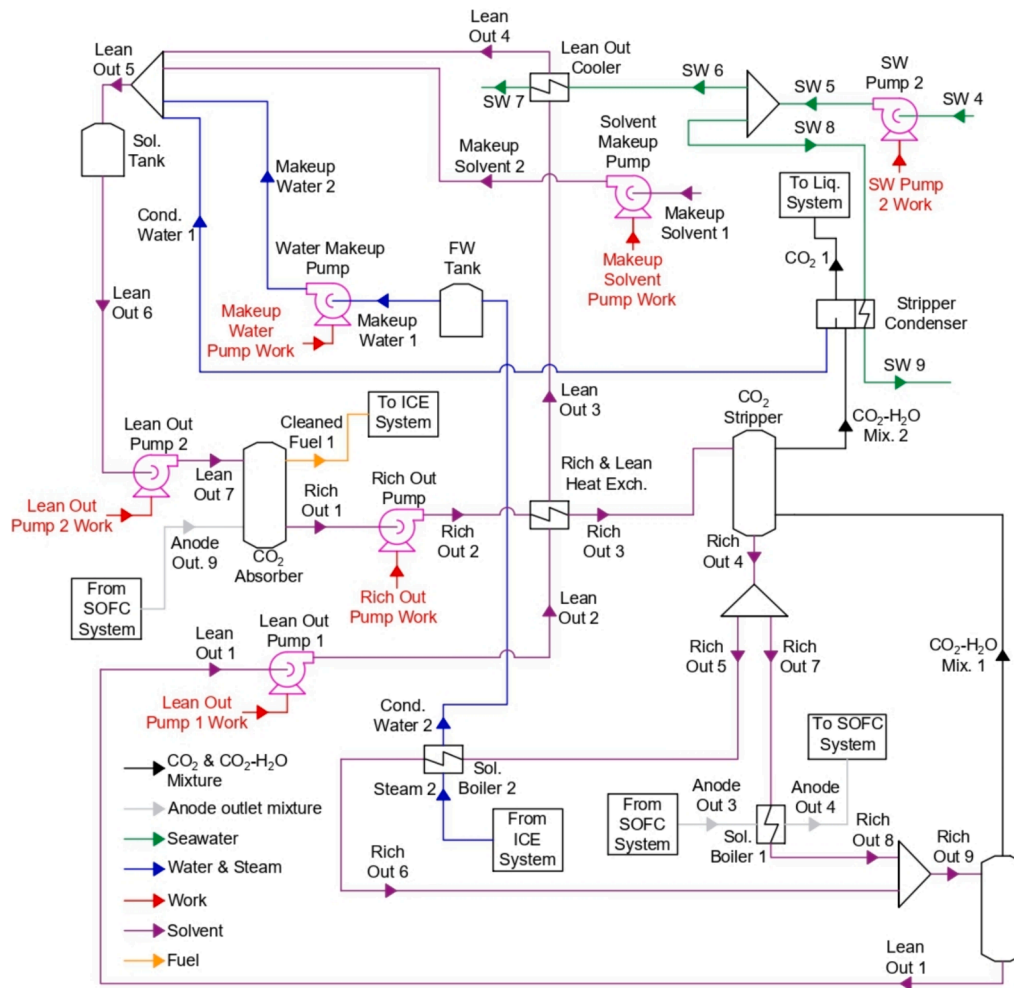


Fig. 5. Schematic model of the CO₂ capture system.

Table 3
Validation of SR results [47].

Temperature (°C)	CO (%)	CH ₄ (%)	H ₂ (%)
500	4.66	0.83	2.34
700	2.41	4.02	0.66
900	0.69	2.9	1.46

4. Modeling of carbon capture and storage system

Using the CCS system for ships has certain difficulties. The ships' restricted space is one of the main obstacles. Column heights of the CCS system may reach up to 20–30 m on land facilities. In addition to the arrangement, column heights are even more critical in terms of ship stability. However, these heights would be reduced due to the lower CO₂ emission rate onboard compared to land-based facilities. Moreover, with the integration proposed in this study, the anode gas is cooled at the SOFC outlet, so the water content is largely removed, and the CO₂ composition entering the absorber becomes quite high. This would reduce the column dimensions considerably. Also, this system has a large number of components. Additionally heavy and costly in terms of freight loss is the CCS system. It is possible to reduce these negative effects using CCS systems on ships with large cargo capacity and dimensions, as in this study. Despite all these disadvantages, the CCS system can be an important option for reducing emissions for ships due to its ability to reduce emissions to a vast extent and its considerable cost.

The CCS system is modeled in two parts: carbon capture and liquefaction-storage. The carbon capture section includes the part capturing the CO₂ from the CO₂-rich anode gas at the SOFC system outlet to its separation in the stripper column and condenser. The liquefaction-storage section includes CO₂ pressurization, cooling, liquefaction, and storage in a tank.

In this study, carbon capture is carried out using solvents. Using solvents are commercialized, most common, and most promising method for carbon capture according to the literature [48]. Besides, this method is based on “end-of-pipe” technology, which is the easiest way in terms of handling during the operation of ships. The carbon capture part of the CCS system is modeled as shown in Figure 45. According to Figure 45, the CO₂-rich anode gas enters the CO₂ absorber column at the bottom of it. The aqueous solvent flows from the top of the column and absorbs the CO₂ in the anode gas by chemical reaction and mass transfer. The aqueous solvent enriched in CO₂ content at the absorber outlet is pressurized with a pump and preheated by a regenerative HEX. The CO₂-rich solution that enters the CO₂ stripper from the top of the column desorbs some CO₂ into the CO₂-water vapor mixture coming from the bottom of the column due to the increased temperature. The CO₂-rich solution at the outlet of the stripper column is separated with a tee and partially evaporated by passing through two different boilers. Heating in these boilers is carried out with anode gas from the SOFC system and steam from the ICE system. After these heated solutions are mixed, the water vapor and CO₂ that pass into the gas phase are separated from the solution by passing through a separator. The CO₂-lean solution passing through the separator is pressurized with a pump, preheats the CO₂-rich

solution in a regenerative HEX, is cooled to the absorber inlet temperature with the seawater, and transferred to a dosing tank. Finally, it is taken from this tank, pressurized with a pump, and then recycled into the system by transferring to the CO₂ absorber. During this process, the CO₂-water vapor mixture coming from the CO₂ stripper is passed through a condenser and is significantly separated from the water it contains. This mixture with increased CO₂ fraction is transferred to the liquefaction system. The water separated from the condenser is transferred to the solvent dosing tanks. The dose of the solvent in this tank is adjusted using streams from the FW and pure solvent tank. On the other hand, the cleaned gas from the top of the absorber becomes a fuel with a high calorific value and is sent to the ICE instead of being released into the atmosphere.

The mass transfer calculations in absorption and stripping processes are performed using the equilibrium-based method. This method is implemented by applying the conservation of mass, thermodynamic equilibrium, stoichiometry relations, and heat balance (MESH) equations. These sets of equations are applied for each theoretical stage in the absorption and stripping columns. Inlet and outlet streams of a theoretical stage are shown in Fig. 6. HYSIM inside-out method with adaptive damping is used to improve the convergence of the numerical solution of the MESH equations [49]. The modified MESH equations according to the HYSIM inside-out method used in this study are given in Table s8.

In this study, 30 % aqueous PZ is used as the solvent. PZ has an excellent reaction rate when compared to conventional amine-based alkanolamine, shows high thermal stability and less volatility, and is not corrosive to stainless steel [48]. Table s10 shows that reversible reactions occur for the absorption of CO₂ in the aqueous PZ solution.

The equilibrium-based model used in the mass transfer calculation assumes that the liquid and vapor phases entering the stage are in thermal equilibrium. Murphree vapor phase efficiency is defined to obtain realistic results for the equilibrium model. In this study, stage efficiencies are determined using the AIChE method [50]. Mellapak 250Y is used as packing material for the columns [13]. BRF-85 method is used to estimate the interface area of the columns [51]. Thermodynamic properties of the liquid phase are estimated by the electrolyte NRTL model and the Peng-Robinson equation state is used for the gas phase [52]. The carbon capture part of the CCS system is validated by a study in the literature. Only a 0.08 % difference in rich loading CO₂ solubility is calculated for the same ambient conditions and absorption column parameters [13]. It could be said that the results are in good agreement.

The captured CO₂ can be stored in the carbamate form in solid, liquid, or gas phases. Storing CO₂ in the gas phase requires a very high level of pressurization and volume. In the carbamate form, the amount

of weight and cargo loss would be high and it is challenging to handle solids in ships. Thus it is considered to be more advantageous to liquefaction of CO₂. In addition, due to the “end-of-pipe” technology makes it easy to transfer and store liquids on ships.

The pressure at the triple phase point of CO₂ is 5.1 bar and −56 °C. As the storage pressure of CO₂ increases, compressor power consumption and purchasing costs increase, but the requirement for cooling and the cost of the cooling cycle decreases. Seo et al. [53] investigated the effects of storage conditions on the feasibility of ship-based CCS, and they found that the LCC is lower between 6–25 bar. On the other hand, BO-NG in the cargo tanks of LNG carriers is used as fuel and has great cooling potential. Therefore, it is recommended to use this cold energy for cooling the CO₂ [30]. The authors modeled a cooling system that utilized the cold energy of BO-NG used as the fuel and showed that the optimum CO₂ storage pressure is related to the amount that this cold energy can provide in their previous publication with a multiobjective optimization method [24]. In the previous study, the optimum CO₂ pressure was calculated in the range of 12.8–23.1 bar for various cases. Similarly, in this study, the CO₂ liquefaction-storage system is modeled so that cold energy is obtained from NG which is sent to SOFC as fuel as shown in Fig. 7. The CO₂ from the stripper condenser is pressurized, intercooled, and separated from its water content in 2 stages (Fig. 7). Then, CO₂ passes through the LNG cold box and it is liquefied until storage conditions by the LNG cold energy. Finally, it is transferred and stored in a tank during the ship's operation. After the pressurization of the CO₂, the intercooling process is carried out with seawater. The moisture separated from the CO₂ in the separators is drained to the FW tank. When the amount of cooled CO₂ is compared for validation, a 0.52 % difference is calculated with the literature and it could be said to be in good agreement [54].

When Qatar's LNG export is analyzed by destination, it is seen that the European union, India, Japan, South Korea, and China come first [55]. The distance from Ras Laffan to the LNG terminals in these destinations is examined [56]. The economical cruising speed of LNG carriers is 16.5 knots, and it is considered that the CO₂ tank with 400 h of storage will be sufficient.

5. Modeling of internal combustion engine integration to the solid oxide fuel cell

In addition to the relatively high cost, the use of SOFC has disadvantages, such as complex fabrication, difficulties in thermal control, cell degradation and limited experience in shipboard use. SOFC-ICE integration requires an extra, more complex power supply structure and control. Besides, the volume in the engine room is very limited and the use of two systems together may make the layout more difficult or the cargo volume of the ship may have to be reduced in the design. However, in addition to these disadvantages, integrating the system will reduce and limit the disadvantages arising from the use of SOFC alone and increase energy security. Apart from these, as mentioned in the introduction, energy efficiency will be increased using unburned hydrocarbons in the anode gas as a fuel for the ICE. The power supplied from the fuel cell will decrease over time due to degradation. The required power can be obtained by increasing the current density and fuel flow rate transferred to the SOFC. However, increasing in fuel rate and current density increase the degradation rate of SOFC over time and reduce the life of the cell [57]. In line with this information, integration, and modeling are carried out in such a way the power lost due to degradation is supplied by ICE in this study. Moreover, ICEs are currently used on almost all ships and the use of SOFC on ships is relatively new. With this integration, energy security will be increased using the ICE when SOFC has a problem in operation.

The ICE system to be integrated into the SOFC is modeled as shown in Fig. 8. According to Fig. 8, the clean anode gas is intercooled, separated from the moisture and other contaminations in its content. It is pressurized by the compressors in three stages until it reaches the ICE

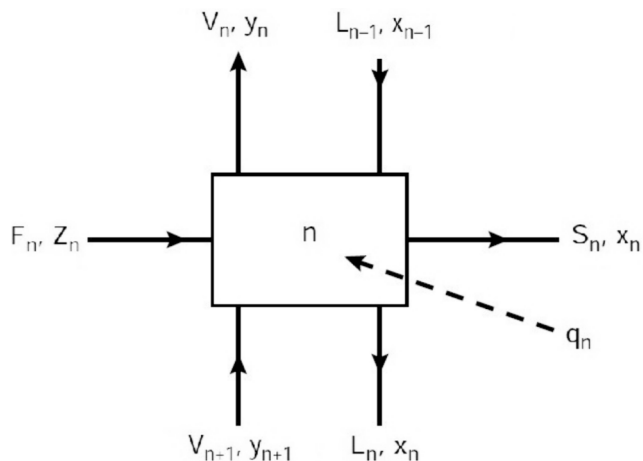


Fig. 6. Inlet and outlet streams of a theoretical stage in absorption and stripping columns [16].

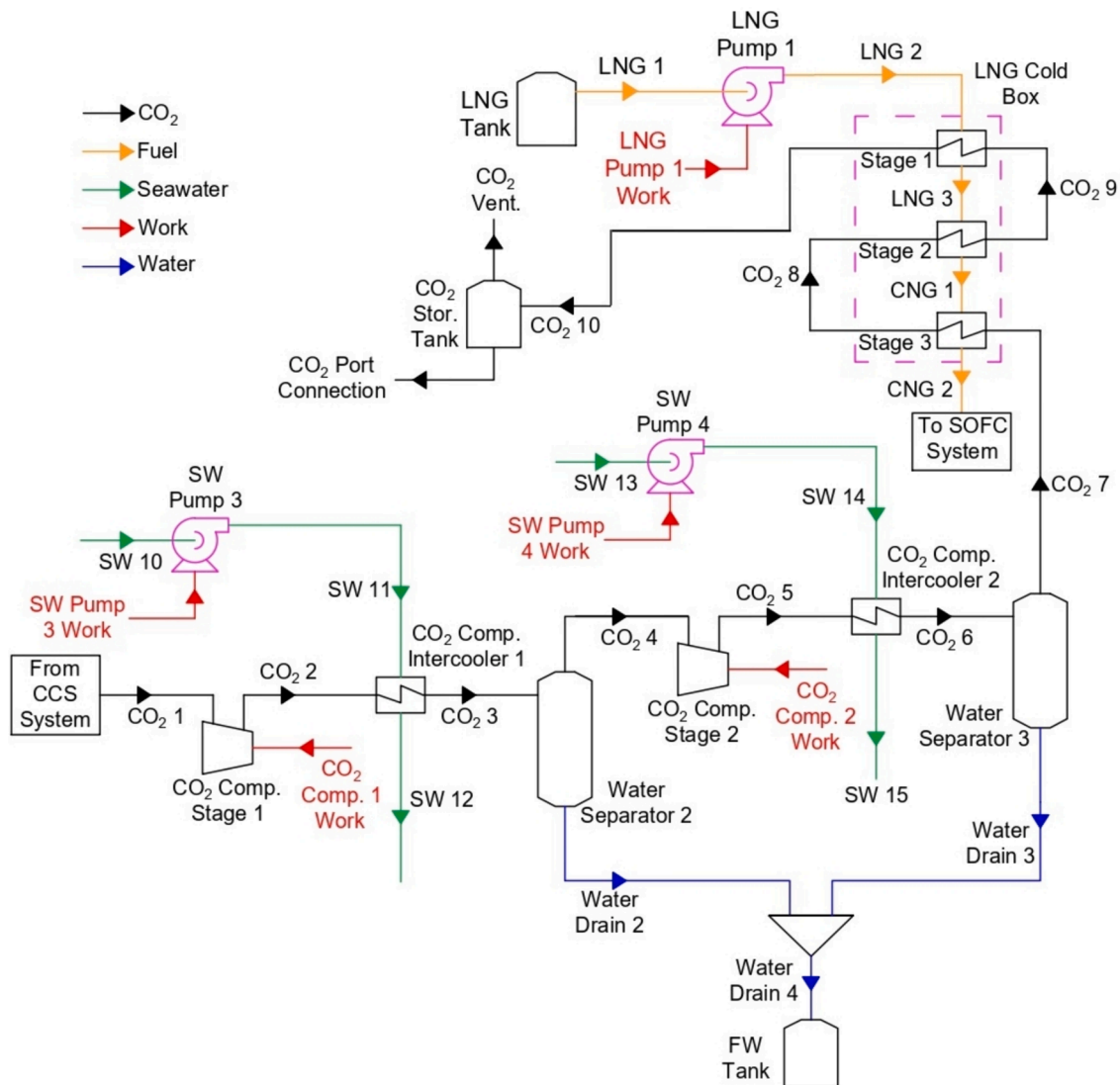


Fig. 7. Schematic model of the CO₂ liquefaction-storage system.

injection pressure, respectively. The remainder of the ICE system is modeled similarly to the conventional ship power systems described in [section 2](#). However, the amount of LNG fuel in this model increases as the power supplied from the SOFC decreases due to the degradation in the SOFC. Furthermore, the flue gas from the TC is released into the atmosphere by passing through a boiler containing the economizer and evaporator parts. The hot FW heated by the TC air cooler and fuel cooler passes through this boiler and becomes steam. This steam is used to separate CO₂ by passing through the solution boiler, as shown in [Fig. 5](#).

6. Modeling of hydrogen fuelled solid oxide fuel cell integration to the internal combustion engine and steam turbine power systems

H₂ as a fuel does not contain CO₂ emissions, therefore it will have a significant place in sustainable energy supply in the near future. On the other hand, methane reforming is a process that requires a large amount of heat, a high initial investment cost and occupies a significant additional volume. Therefore, there are several advantages to using hydrogen instead of reformed methane as a fuel. Besides, there is no need for the CCS system due to the using H₂. The CCS system also requires a large amount of heat, a certain LCC, and occupies a significant additional volume. Also, the using H₂ instead of reformed methane in

SOFC increases cell efficiency. For these reasons, it is considered important to examine how the results will change in terms of 4E if hydrogen is used as the fuel for LNG carriers. Thus, an alternative system is modeled as seen in Fig. 9, simulated, and 4E analyses are performed. The SOFC system is integrated into the ICE due to its previously mentioned advantages like the proposed system.

According to Fig. 9, the cooled anode gas is injected into the combustion chamber of an ICE after 3-staged pressurization, intercooling, and drainage of water without passing through the carbon capture column. Unlike the proposed system, the waste heat from the system is quite large and at high temperatures since it is not used in the CCS system and SR. A Rankine cycle with an ST is also integrated into the system to utilize this waste heat. According to Fig. 9, the FW coming from the tank is divided into various branches. The water in the first branch is pressurized by a pump and preheated by the hot gas coming from the cathode side. It is then heated up to the evaporation temperature by passing through the economizer of the exhaust gas boiler. It is combined with another hot water branch and becomes superheated steam by passing through the evaporator and superheater, respectively. The second branch is pressurized by a pump, passed through the HEX in the third stage of the fuel, which is sent to the ICE and combined with the first branch. The third branch is pressurized by a pump again and passed through the HEX in the second stage of the fuel, which is sent to the ICE

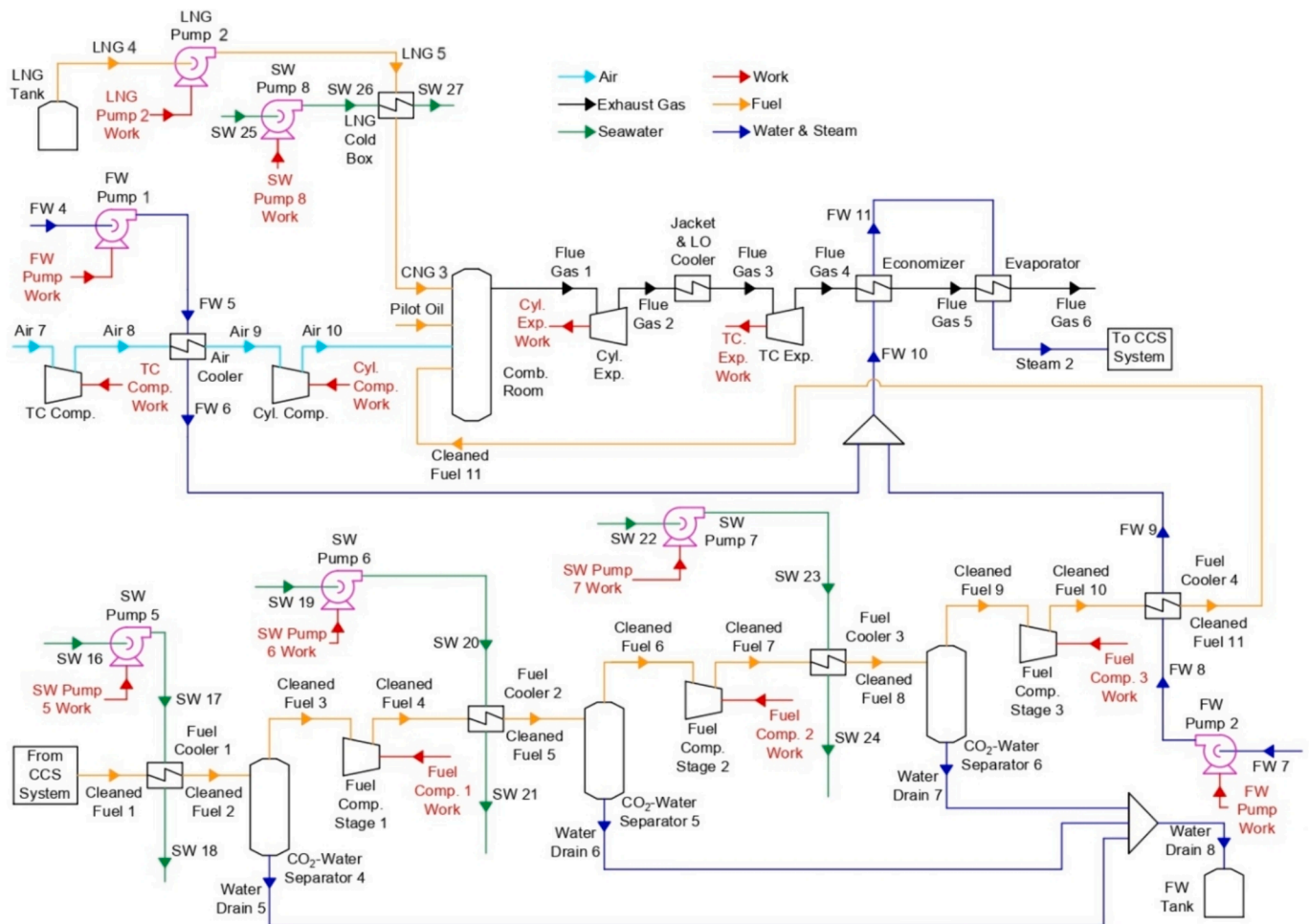


Fig. 8. Schematic model of ICE integrated into SOFC.

and combined with the fourth branch. The fourth branch is also pressurized by a pump and preheated by the HEX in the first stage of the fuel, which is sent to the ICE and then reaches the evaporation temperature at the outlet of the third stage of the anode out boiler. These two combined branches, pass through the other two stages of the anode out boiler, so the FW is evaporated and superheated respectively. Another branch is used to control the temperature of the SOFC after it is pressurized, and the high temperature of the SOFC causes it to enter the superheated vapor phase. Finally, all branches that become superheated steam phase combine, pass through the ST, and produce additional power to the system. The steam at the turbine outlet is condensed with seawater and sent to the tank using a pump, so it is recycled.

In addition to the advantages of H_2 compared to reformed methane, there are also some disadvantages, such as storage difficulties, high cost of fuel per unit of energy, and safety. In this study, it is assumed that hydrogen will be stored as compressed gas in a 350-bar tank, which is the most common option [58]. In a previous study, it has been showed that the initial investment cost of liquefaction is more than two times higher than compressed gas storage option [59]. Also, storage below $-253\text{ }^{\circ}\text{C}$ is required for the liquefaction. It is considered that reaching under this temperature and insulating cryogenic tank is challenging. The problem with high pressure storage is that the tanks are quite heavy and lead to greater volume loss than liquefaction in addition to the safety-relevant hazards. A 350 bar type IV tank can store 5.6 kg of usable H_2 , while its total weight is 104.4 kg and occupies 316.4 L [60]. In addition to all these disadvantages, it has a high cost of about 30 $\$/\text{kg}$ per tank weight even though it is cheaper than liquefaction [60]. In addition, the service life of these tanks is only 10 years.

7. Economic analyses

In this study, the total LCC per captured CO₂ (C_{CO_2}) is used to measure economic evaluation to compare the alternative emission-reducing technologies. The cost per unit of reduced CO₂ is calculated by subtracting the total LCC of the suggested system (LCC_{new}) from the total LCC of the conventional power system ($LCC_{con.}$) and dividing by the amount of reduced CO₂ during the ship life cycle for studied ships. The capital expenditure (CAPEX) and operational expenditure (OPEX) constitute the LCC. The parameters and equations used in the calculations are shown in [Tables 4 and 5](#). According to [Table 5](#), CAPEX is a function of the purchasing cost of equipment (C_{pur}), and the loan is considered for this calculation. However, inflation, taxes, and insurance fees are ignored. Here, f_{CAPEX} represents the initial investment factor and considers engineering services, construction management, commissioning, owner, soft, and other costs. The purchasing cost of the equipment except for the absorber and stripper column are estimated by the equations shown in [Table s11](#). The purchasing costs of the absorber and stripper columns are estimated by Aspen HYSYS [\[43\]](#). Freight loss cost ($C_{fr,loss}$) due to the weight of the systems, fuel costs (C_{fuel}), and solvent makeup costs ($C_{sol,loss}$) constitute the OPEX. Maintenance, availability, and personnel costs are considered similar for all cases, so they are ignored [\[36\]](#). $C_{fr,loss}$ is calculated by the reduction in the amount of cargo on the ship due to the additional weight for the investigated systems on ships. In this study, both the weights of the conventional main power system and the equipment of the suggested systems are considered. The weight of the conventional power system was assumed to be the dry weights of each ICE and its components, as shown in [Table 1](#) [\[32\]](#). SOFC

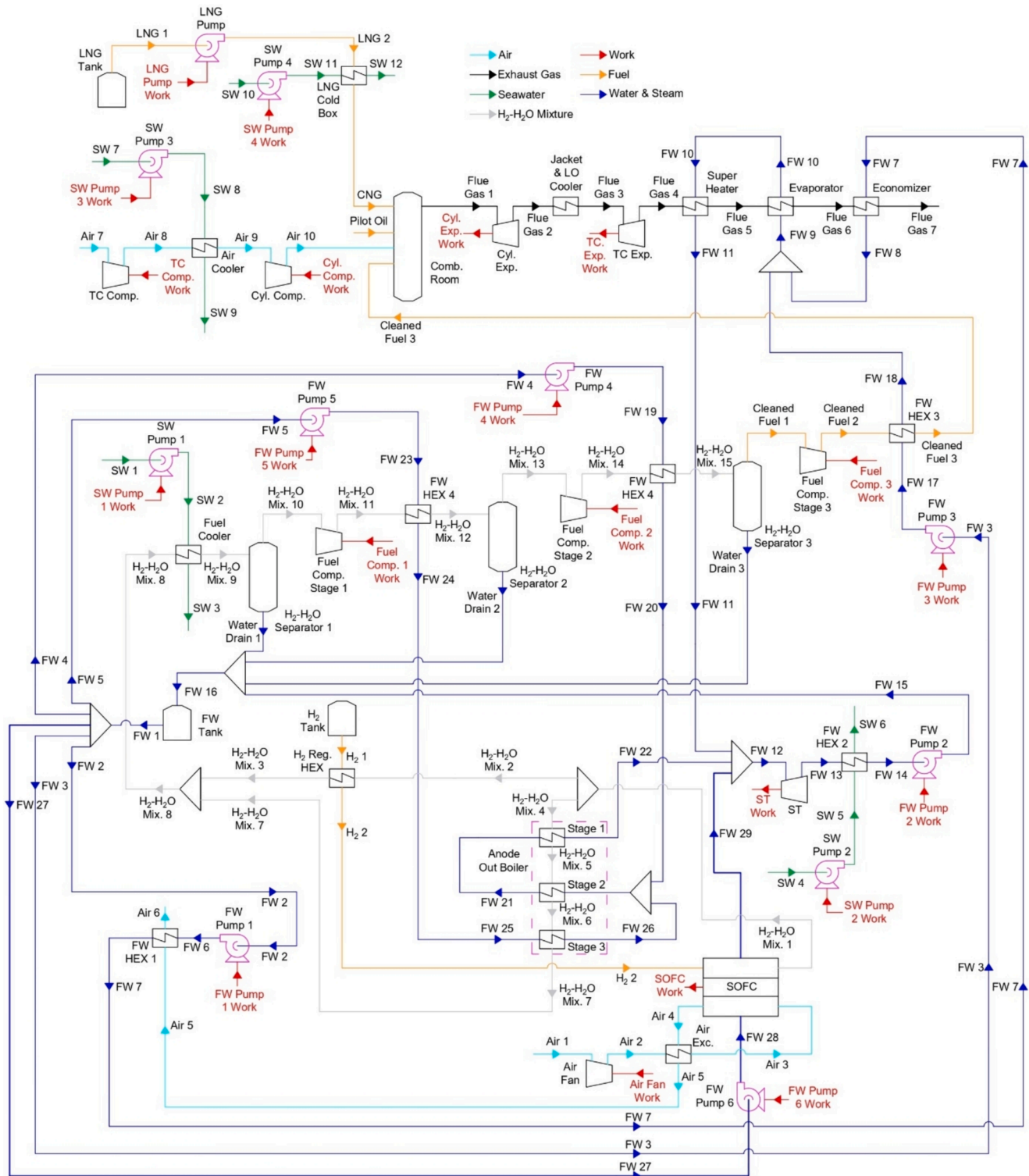


Fig. 9. Schematic model of H₂ fuelled SOFC integration to the ICE and ST power systems.

system weight is estimated with SOFC power density data and power obtained. The weight of the remaining equipment is estimated using Aspen HYSYS [43]. In addition to the equipment, a large amount of water is used for both reforming and CCS systems in the suggested system. The weight of stored process water is included in the $C_{fr,loss}$ calculation. Besides, the weight of the stored CO₂ onboard is more than the consumed fuel, so this additional weight is also considered in the calculation.

8. Results and discussion

Around 700 °C, almost all of the methane is reformed, according to the reformation results. In this study, it was assumed that the fuel was reformed at 700 °C [38]. The molar composition of the reformed fuel as a result of reforming is shown in Table s16. The stream and equipment duty data of the models at the start of the operation are shown in Tables s14 to s22.

Table 4
Economic analysis parameters.

Description	Value	Unit	References
Initial investment factor (f_{CAPEX})	2.5	—	[61]
Interest rate (in)	4	%	[61]
Economic life of the ship (ne)	25	year	[13]
SOFC system weight	50	W/kg	[62]
Process water storage duration	12	hour	—
The average freight rate (C_{fr})	0.466	\$/ (day.m ³)	[63]
Cost of pilot oil (VLSFO)	814.5	\$/ton	[64]
LNG fuel price	945.5	\$/ton	[64]
CEPCI	808.7	—	[65]
CEPCI _{r,pump}	584	—	[66]
CEPCI _{r,comp}	342	—	[66]
CEPCI _{r,sep}	708	—	[66]
CEPCI _{r,HEX}	369	—	[66]
CEPCI _{r,ref}	542	—	[66]
CEPCI _{r,JCE}	577	—	[66]
CEPCI _{r,I-CO₂}	576.1	—	[66]
CEPCI _{r,I-H₂}	607.5	—	[66]
CEPCI _{r,I}	567.3	—	[66]
CEPCI _{r,SOFC}	567.3	—	[66]

Table 5
Economic analysis equations.

Equations	Unit	Eq. no
$C_{CO_2} = (LCC_{new} - LCC_{con.}) / \dot{m}_{CO_2}$	\$	(28)
$LCC = CAPEX + OPEX$	\$	(29)
$CAPEX = C_{pur} \times f_{CAPEX} \times \frac{in.ne.(in+1)^{ne}}{(in+1)^{ne}-1}$	\$	(30)
$OPEX = C_{sol,loss} + C_{fr,loss} + C_{fuel}$	\$	(31)
$C_{sol,loss} = \dot{m}_{sol,makeup} \times C_{sol}$	\$	(32)
$C_{fr,loss} = \frac{\Delta m_{Eq} \times C_{fr}}{DWT}$	\$	(33)
$C_{fuel} = \dot{m}_{LNG} \times C_{LNG} + \dot{m}_{pilotoil} \times C_{pilotoil}$	\$	(34)

Various parameters and equipment dimensions regarding the design of the CCS system are given in Table 6. The results are given for the conditions of 900 °C temperature and 2.5 kA/m² current density at the start of operation. According to Table 6, the dimensions of the absorber and stripper column are quite small compared to the literature despite the higher CO₂ capture amount [13,15,70]. For the Q-Flex ship type with the same particulars, the authors previously reported a much lower capture rate of 34.2 %, with absorber column packing height and diameter of 3.45 and 2.8 m, and stripper column packing height and diameter of 1.1 and 3.2 m, respectively [24]. In the integration proposed in this study, the fact that the CO₂ content of the gas to the absorber column inlet is much higher than the post-combustion capture (PCC) has considerably reduced the dimensions.

The effect of SOFC operating temperature and current density on cell potential and power density is shown in Fig. 10 for reformed NG-fueled SOFC. The cell potential decreased gradually with increasing current density due to the effect of ohmic and activation losses in accordance with the literature and decreased suddenly when the limit current density was approached [37]. At low current densities, the value of the cell potential is higher for fuel cells operating at low-temperature values. However, as the current density increases, fuel cells operating at high temperatures are more advantageous regarding of cell potential. This is

Table 6
The process parameters and equipment sizes of the CCS systems for LNG carriers.

Description	Q-Max	Q-Flex	CLNG	Unit
Lean CO ₂ loading	0.42	0.42	0.42	mol CO ₂ /mol solvent
Rich CO ₂ loading	0.66	0.66	0.66	mol CO ₂ /mol solvent
Absorber diameter	1.45	1.4	1.2	m
Absorber packing height	2.8	2.6	2.6	m
Stripper diameter	1.9	1.8	1.6	m
Stripper packing height	5.5	4.5	3.5	m
Pressure drop in the absorber	0.404	0.315	0.344	kPa
Pressure drop in the stripper	0.718	0.606	0.464	kPa
Stripper column pressure	1.5	1.5	1.5	bar
Flooding rate of the absorber	78.1	72.6	75.0	%
Flooding rate of the stripper	78.8	78.1	76.1	%
CO ₂ capture rate	92.5	92.5	92.5	%
Stripper reboiler duty	16.09	14.45	11.23	MW

because activation losses require higher current density than low temperatures to be effective at high temperatures. For example, the current density at which the activation loss starts to be effective in the fuel cell with an operating temperature of 900 °C is around 2.5 kA/m², while it is around 0.6 kA/m² at 750 °C operating temperature. The power density, on the other hand, peaks with increasing current at all temperature values up to the region where the loss of concentration begins to be effective and then decreases suddenly, like the cell potential.

The effect of SOFC operating temperature and current density on cell potential and power density for H₂-fuelled SOFC is shown in Fig. 11. Figs. 10 and 11 show that the efficiency of SOFC using reformed NG fuel is approximately 20.6–41 % lower than that of H₂-fuelled SOFC. When the temperature and current density increase, it is seen that the efficiency difference increases even more. The basis of this difference is understood when loss mechanisms are analyzed. Fig. 12 shows the loss mechanisms at 750 °C for both fuel types, and the difference in activation, ohmic, and concentration losses is quite small. The main difference is due to parasitic losses, and while the efficiencies decrease with increasing current density, the parasitic losses are constant. When Figs. 10 and 11 are examined, it is seen that the polarisation curves form a kink in the high-voltage region. Concerning this, in Fig. 12, activation losses started to increase only after a specific current density value. This is due to calculating activation losses with the Tafel equation given in Equation (9). While this equation does not take into account losses at low current densities, it is valid in the current density region examined in this study.

The polarization curves shown in Figs. 10 and 11 give crucial clues about the energy efficiency of the systems at the beginning of operation. However, it should be considered that the efficiency will decrease over time due to the effect of degradation. Fig. 13 shows the changes in CO₂ emission reduction amount, the fuel cell, and the overall system efficiency over time due to the degradation effect for both systems examined. The analysis is performed for the conditions of 900 °C temperature and 2.5 kA/m² current density.

As shown in Fig. 13, the fuel cell efficiencies for both systems show an almost linear decrease, but the loss in overall system efficiencies is much lower compared to the cell. Especially in the SOFC-ICE-CCS system with reformed fuel, this loss is even smaller and can fulfill the expectations if a longer SOFC lifetime is demanded. The energy efficiency of the current power system is 52.78 % and the efficiency of the SOFC-ICE-CCS system is below this efficiency including the start of operation. The efficiency of the H₂-fuelled system is above this efficiency up to about 30,000 h. Also, the average efficiency of this system is higher than the conventional power system for the entire feasible operation period. When the CO₂ emission reduction potentials are analyzed, the system, which initially reduced the emissions by 100 % due to the use of only H₂,

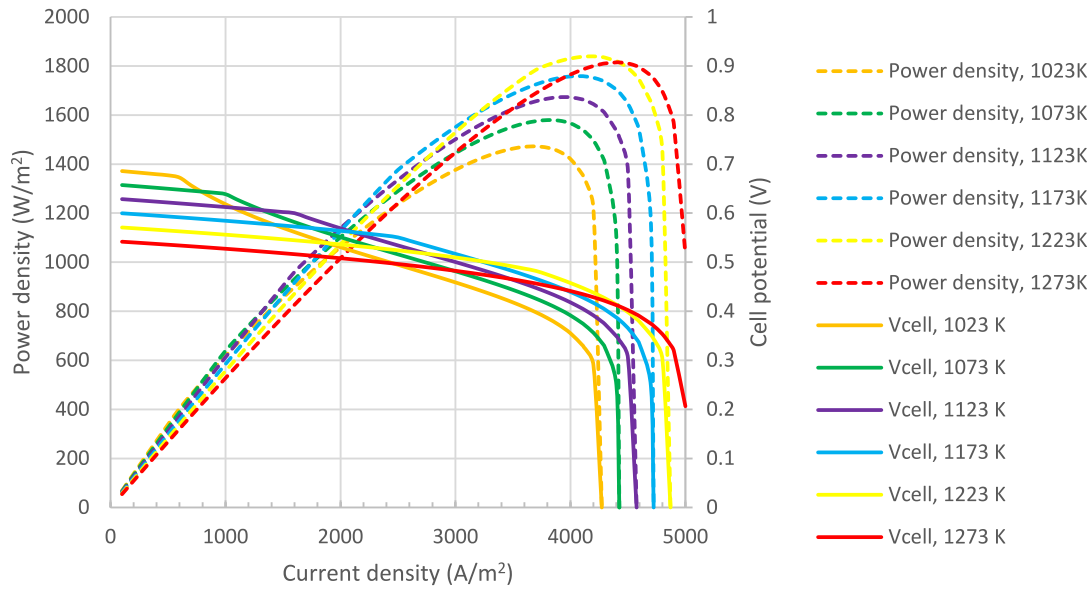


Fig. 10. Effect of operating temperature and current density on potential and power density for reformed NG-fueled SOFC.

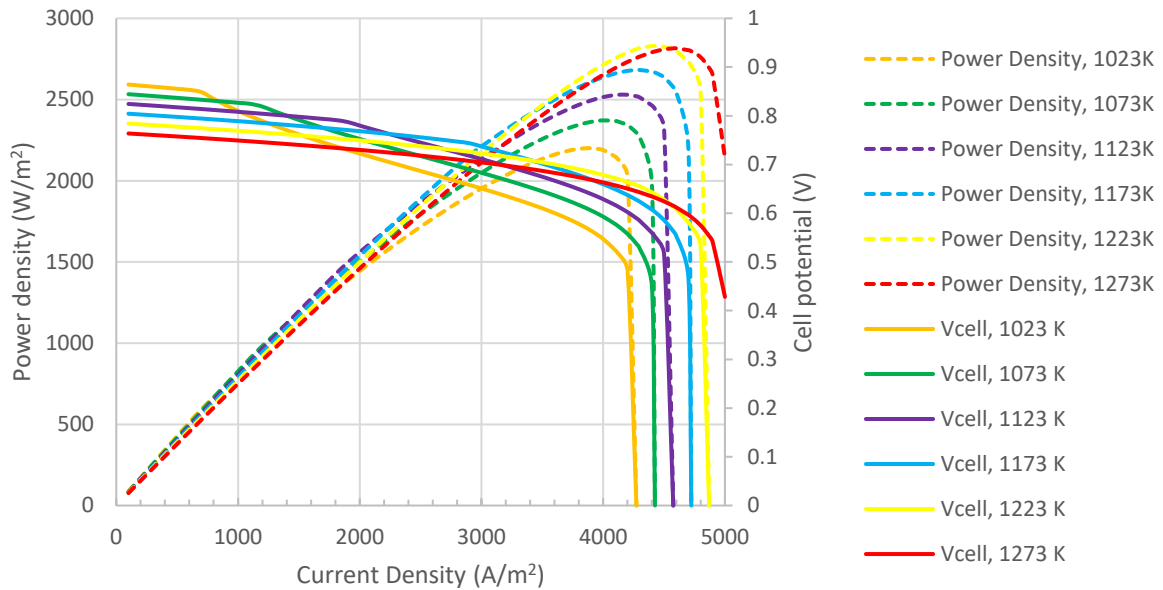


Fig. 11. Effect of operating temperature and current density on potential and power density for H₂ fueled SOFC.

decreased up to 90 % in time due to the provision of the required energy from ICE. Although the carbon capture rate of the proposed SOFC-ICE-CCS system exceeds 90 %, the amount of emission reduction of this system is around 80 % at the beginning of operation due to reasons, such as some of the CO gas in the fuel cell not reacting in the cell, the overall system efficiency is lower, and the additional power requirement of the CCS system. Since only CO₂ from the anode gas is captured, this rate falls below 70 % after approximately 40,000 h of operation. When an operating life of 35,000 h is selected, the average CO₂ reduction rate of reformed and H₂-fuelled systems were 76.1 % and 96.9 %, the average SOFC energy efficiencies were 38.8 % and 45.3 %, and the average overall energy efficiencies were 44.1 % and 57.5 %, respectively. The results in Fig. 13 are valid for all three ship types examined in this study.

The cell potential affects the efficiency of the system, hence the amount of fuel to be used and the fuel cost. The power density determines the area of the SOFC and therefore the initial investment cost. Accordingly, while increasing the current density will increase the fuel

cost, it will decrease the initial investment cost of the SOFC until the maximum power density value. Thus, there will be a current value that minimizes the cost for the selected operating temperature. The situation for temperature is slightly different. The required cell area and initial investment cost will decrease again with increasing temperature, but the fact that the materials used are more resistant to temperature and more difficult to produce will increase the initial investment cost. The effect of temperature on the initial investment cost is shown in Equation (s76). In addition, the effect of temperature on the potential that can be obtained varies depending on the current density selected. The operating life of the SOFC affects the initial investment cost of the cell, the efficiency of the cell, and the integrated system over time due to degradation. As the operating life increases, the initial investment cost of the cell decreases, but the operating cost of the integrated system increases as the efficiency decreases more over time.

As a result of these findings, the SOFC current density, operating life, and temperature that minimize the cost of the system with CCS-ICE-

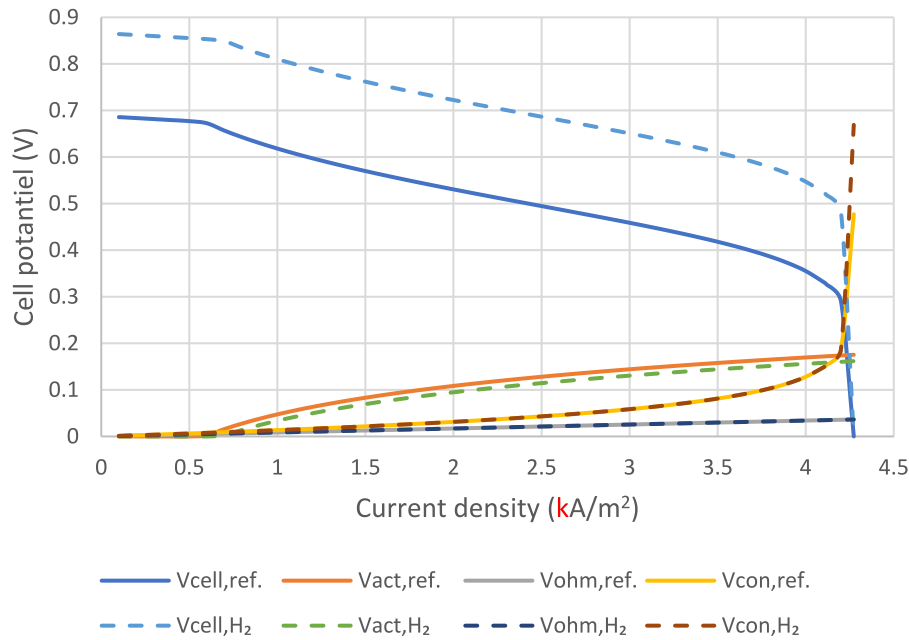


Fig. 12. Loss mechanisms of SOFC and variation of net potential with current density.

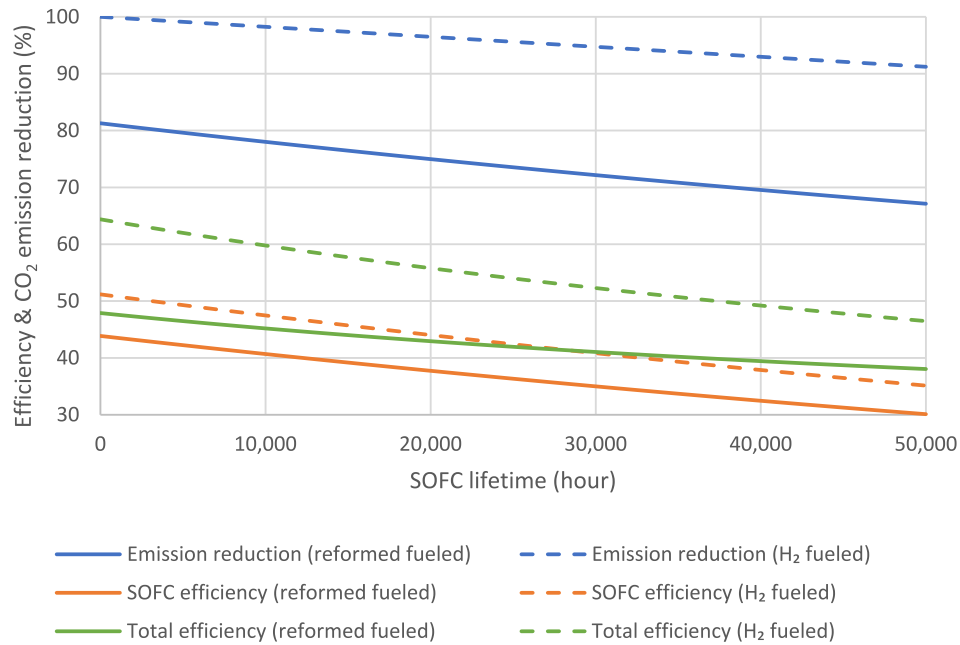


Fig. 13. CO₂ reduction rates, SOFC energy efficiencies, and overall system efficiencies for SOFC-ICE-CCS system with reformed fuel and SOFC-ICE-ST system with H₂ fuel.

SOFC integration were determined by a parametric study. The effect of current density on CO₂ reduction cost with operating life is shown in Fig. 14 and the effect of temperature on CO₂ reduction cost with operating life is shown in Fig. 15.

The range of 2–4 kA/m² is analyzed for current density, and according to Fig. 14, the most cost-effective current value is 2.5 kA/m². The SOFC operating life, which minimizes the cost, is calculated at around 35,000 h. The operating temperature selected here is 900 °C.

For the temperature, the range of 750–1000 °C is examined and according to Fig. 15, the optimum temperature value that minimizes the cost is calculated around 900 °C. The current density selected here is 2.5 kA/m². Unlike the current density, here, the unit CO₂ abatement cost showed a sudden change after 900 °C. This is due to the variation of the

required SOFC surface area, material value, and the potential obtained with temperature variation, as mentioned before. The cost-minimising SOFC operating time is around 36,000 h, similar to that in Fig. 14. As a result, depending on the operating life under the conditions studied, the choice of current density reduced the CO₂ abatement cost up to 27.3 % and the choice of temperature reduced the CO₂ abatement cost up to 31.5 %. The results in both figures are obtained for the Q-Flex type ship. The price of LNG can vary considerably over time. Thus, the results of the effect of the LNG price and operating life on CO₂ abatement cost for the SOFC-ICE-CCS system are shown in Fig. 16.

The cost analysis results for the case of using hydrogen instead of reformed fuel are shown in Fig. 17 depending on the SOFC lifetime. Here, the analysis was carried out at 2.5 kA/m² current density and

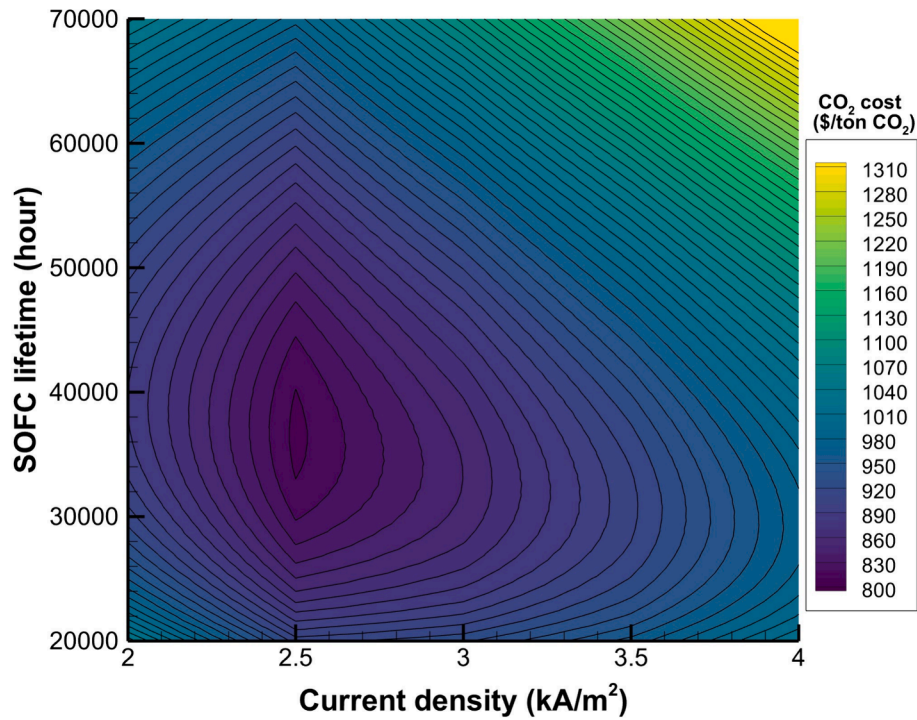


Fig. 14. Effect of SOFC current density and operating time on CO₂ abatement cost for SOFC-ICE-CCS system.

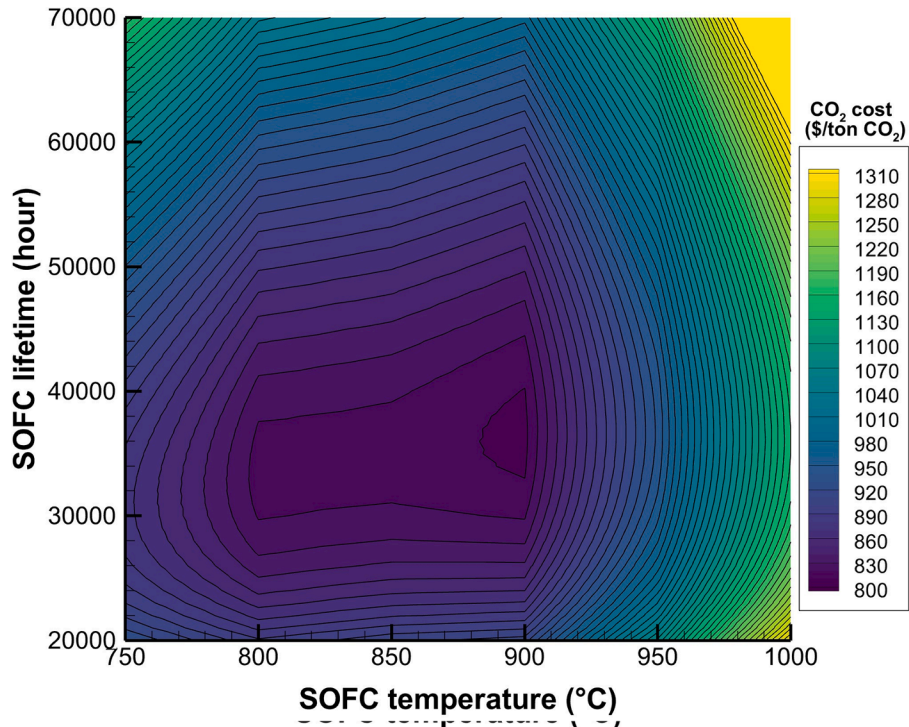


Fig. 15. Effect of SOFC temperature and operating time on CO₂ abatement cost for SOFC-ICE-CCS system.

900 °C temperature conditions. Normally, fuel prices vary greatly over time. However, when H₂ and LNG fuel prices are accepted as approximately 3500 and 950 \$/ton, the lowest CO₂ abatement cost is approximately 1310 and 771.1 \$/ton CO₂ for H₂ and reformed fueled SOFC integrated systems, respectively [39,64]. For Q-Max and CLNG type ships, the cost of CO₂ reduction is 754.5 and 799.3 \$/ton CO₂. The cost for the reformed fueled case is 41.1 % lower than the case where pure

hydrogen is used as fuel. In the literature, in the case of using hydrogen as fuel in ships with fuel cells, CO₂ abatement cost value is slightly above 1000 \$/ton CO₂ for an interest rate of 1 % and can exceed 2100 \$/ton depending on the interest rate [67]. In line with these results, it is seen that the use of the CCS system reduces costs to a great extent. On the other hand, in the previous studies of the authors, it has been determined that if only the CCS system is used for the same ships, the costs are

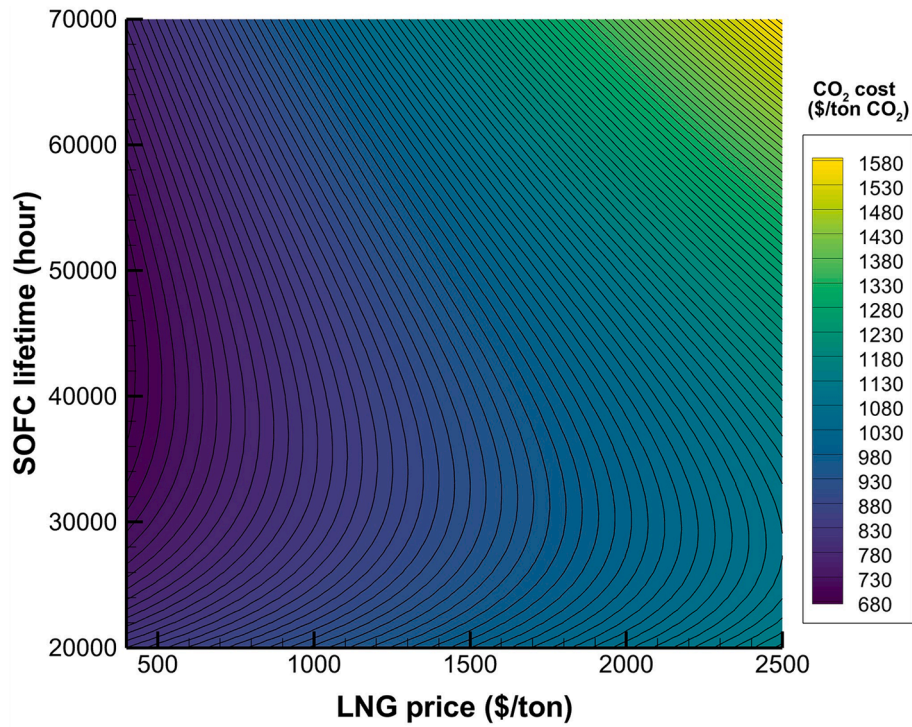


Fig. 16. Effect of LNG price and SOFC lifetime on CO₂ abatement cost for SOFC-ICE-CCS system.

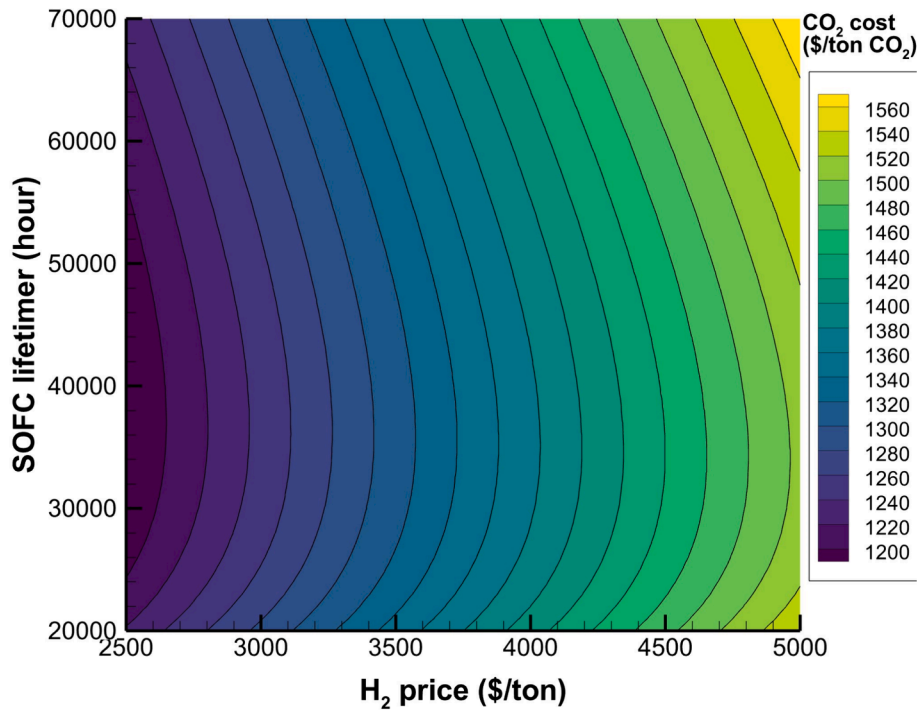


Fig. 17. Effect of H₂ price with SOFC lifetime on CO₂ abatement cost for SOFC-ICE-ST system.

around 59.5–73.7 \$/ton CO₂, and with the integration of BO-NG and CO₂ liquefaction systems of the ships, these costs can be reduced to around 43.1–49.7 \$/ton CO₂ [24]. However, in these studies, it was stated that CO₂ was only captured around 30–35 % and the costs increased suddenly if the heat required for CO₂ regeneration could not be met from the waste heat from ships. In this study, it should be taken into consideration that CO₂ is captured/reduced up to around 90 %.

Cost reduction seems to be an important priority to increase the

feasibility of the integrated systems presented in the study. Analyzing the sources of the cost will also serve this purpose significantly. Fig. 18 shows the distribution of the annual costs of cost sources for ships with both conventional power systems and integrated systems analyzed in the study, in a million \$. Fig. 19 shows the distribution of the purchasing costs of the studied systems for various cost items.

As shown in Fig. 18, the initial investment cost of the conventional power systems of the ships is quite small compared to studied systems

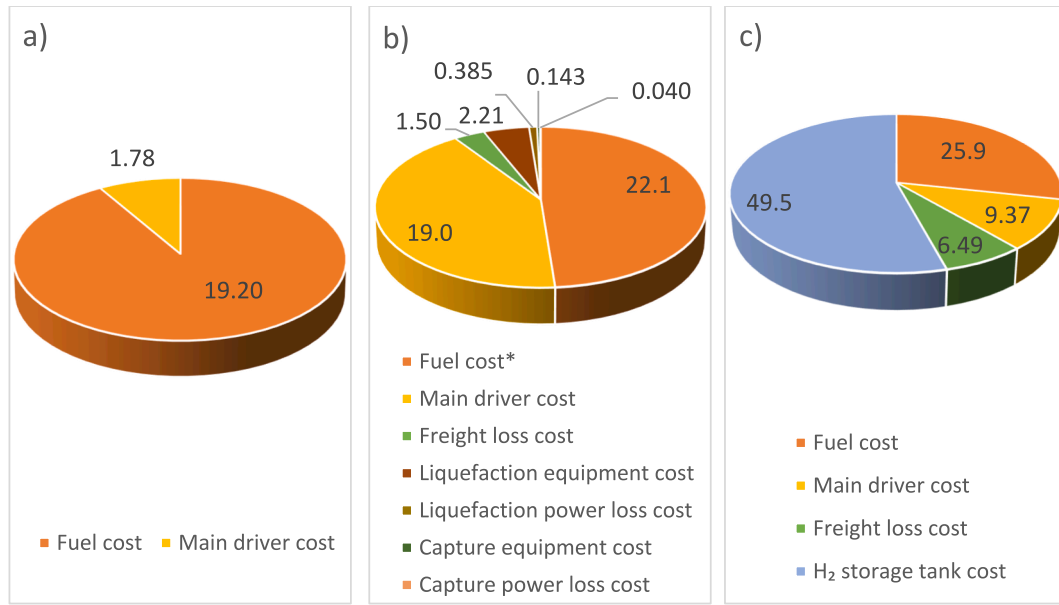


Fig. 18. Annual cost of cost sources for conventional power systems (a), SOFC-ICE-CCS integrated system with reformed fuel (b), and SOFC-ICE-ST integrated system with H₂ fuel (c), in million \$.

and the fuel constitutes the majority of the LCC. According to Figs. 18 and 19, the biggest cost source for SOFC-ICE-CCS integration is the SOFC cost, which is part of the main driver cost. In particular, the low efficiency and power densities due to the use of reformed fuel have greatly increased the SOFC initial investment cost. The decrease in efficiency has also increased the fuel cost and compared to the conventional power system, this difference accounts for approximately 13.6 % of the LCC per captured CO₂. The other largest cost item is freight loss and other costs associated with the CCS system are quite small compared to the total cost. When the cost of the H₂-fuelled SOFC-ICE-ST power system is analyzed, it is seen that the biggest cost belongs to the H₂ storage tanks. The main reason for this cost is that these tanks are quite heavy

compared to the amount of hydrogen stored and the relatively high tank cost per unit weight. Although the cost of the main driving system is also higher than the conventional power system, this cost is much lower than the SOFC-ICE-CCS integration due to the high efficiency and power density brought by the use of pure H₂. The freight loss cost is higher compared to SOFC-ICE-CCS integration due to the weight of the fuel tanks. The weight of the proposed SOFC-ICE-CCS system was estimated to be approximately 5971 tonnes heavier than the conventional onboard system, and this loss was estimated to be approximately 10,821 tonnes in the SOFC-ICE-ST integration. In the SOFC-ICE-ST integration, almost all of this weight loss is due to the weight of the H₂ tanks and it is estimated that the tanks occupy approximately 31256 m³ in volume. In

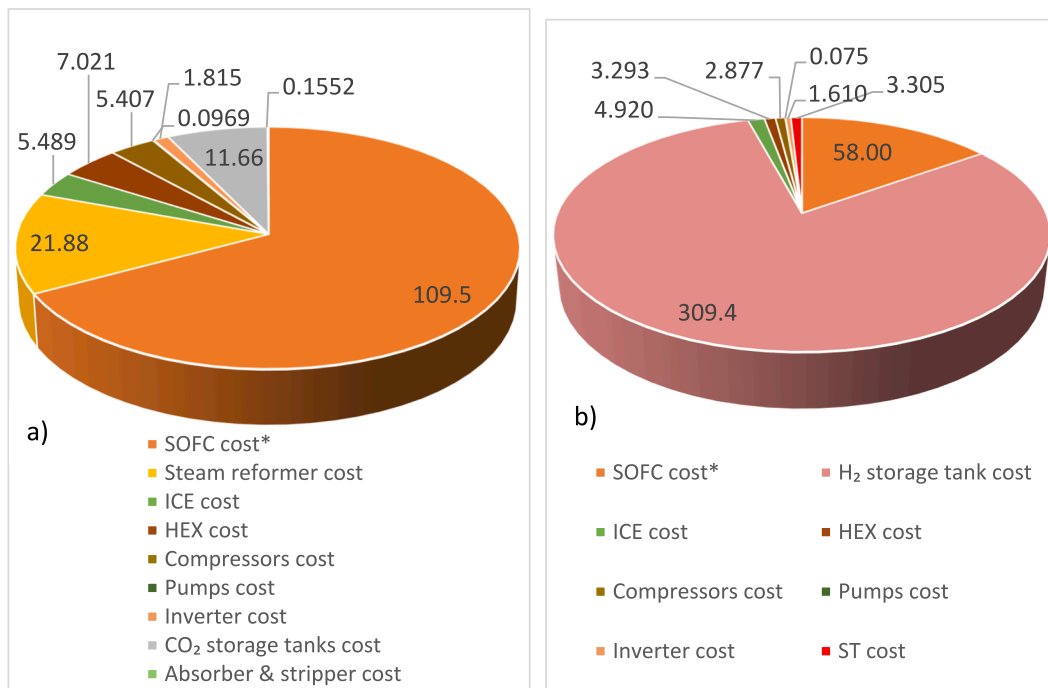


Fig. 19. Purchasing cost distribution for SOFC-ICE-CCS integrated system with reformed fuel (a) and SOFC-ICE-ST system with H₂ fuel (b), in million \$. * While calculating the SOFC purchasing cost, the overall SOFC cost used during the ship's lifetime is taken into account.

addition, the SOFC efficiency, which decreased over time due to degradation, caused more fuel consumption and increased cost compared to the conventional power system. According to Fig. 19, when the common cost items in both integrations are compared, it is seen that the H₂-fuelled system is much cheaper than the other system. H₂-fuelled system is 47 % cheaper for SOFC, 10.4 % cheaper for ICE, 53.1 % cheaper for HEX, 39.1 % cheaper for compressors, 22.6 % cheaper for pumps and 11.3 % cheaper for inverter. In line with these results, if the problem of the cost of storing hydrogen on board is solved, the H₂ fuelled SOFC-ICE-ST system will become very feasible in reducing CO₂ emissions on ships.

Exergy analyses of the proposed systems and the conventional power systems of the examined ships are performed and the results are shown with Sankey diagrams. The results of the conventional power systems are shown in Fig. 20, the results of SOFC-ICE-CCS integration are shown in Fig. 21 and the results of SOFC-ICE-ST integration are shown in Fig. 22. The results of the integrated systems are given for start-up conditions with no degradation at 2.5 kA/m² current density and 900 °C temperature conditions. From Figs. 20, 21, and 22, it is seen that the exergy efficiency of the conventional power system is much higher than the SOFC-ICE-CCS integration but lower than the SOFC-ICE-ST integration. As mentioned before, using reformed combusted fuel has greatly reduced the Nernst potential and cell efficiency, which has greatly reduced the exergy efficiencies. Unlike SOFC-ICE-ST integration, here the waste heat is used to meet the needs of the CCS system and SR, the rest of the waste heat is not utilized. This results in exergy loss and destruction. In addition, the additional cooling and other power requirements of the CCS system also reduced the exergy efficiency.

When all the analysis results are evaluated, the recommended operating conditions for the proposed SOFC-ICE-CCS system are 900 °C temperature and 2.5 kA/m² current density, approximately 35,000 h SOFC operation time. It is suggested that the SOFC operation time should be reduced as the fuel cost increases. The biggest challenge of the proposed system is cell degradation and SOFC cost. If the voltage loss as a result of cell degradation was 0.125 % instead of 0.75 %, the cost under the operating conditions proposed in this study would be 429.5 \$/ton for the Q-Flex type ship, thus a 43 % reduction. The average CO₂ emission reduction would increase from 76.1 % to 80.3 %. Another challenge is the low efficiency of reformed fuel in SOFC. This significantly increases the fuel cost of the system. The use of alternative fuels that will increase

SOFC efficiency can solve this problem.

In a study where SOFC was directly coupled with ICE for an LNG fuelled crude oil tanker, the electrical efficiency of SOFC was 50.8 % and the total electrical efficiency of the system was 54.3 % [9]. In this study, in the SOFC-ICE-CCS integration, the electrical efficiency of the SOFC at the beginning of operation is 45.6 %, and the total electrical efficiency of the system is 50.2 %. Although the efficiencies are low in this study, the heat and power requirements of the CCS system should be considered. In addition, the current density is chosen to minimize the cost in this study. It should be taken into account that higher efficiencies can be achieved at lower current densities. In the literature study, the effects of cost and current density are not taken into account. On the other hand, while it is stated in the literature that CO₂ emissions can be reduced up to 16.1 % with the direct couple system, 80.8 % of CO₂ emissions can be reduced at the beginning of operation with the CCS system integration proposed in this study. In another study, in the SOFC-ICE integration with reformed LNG-fuelled SOFC, the efficiency of the integration at 3 kA/m² and 800C temperature is approximately 42 %, while the efficiency is 47.36 % under the conditions given in this study [12]. In this study, the maximum efficiency of the integration can slightly exceed 45 %, which is still lower than the integration efficiency in this study. It is also reported that CO₂ can be reduced up to a maximum of 20.7 % in the reference study, which is much higher in this study. In the literature, the cell potential and power density values of a SOFC-GT integrated system with the same SOFC operating temperature, fuel utilisation factor and reformed fuel at a current density of 2.5 kA/m² are 0.568 V and 1426 kW/m², respectively [68]. In this study, the cell potential and power density values under the same conditions are 0.551 V and 1376 kW/m², respectively, and the results are close to the literature. The system duration of 40,000 h is considered a reasonable target for SOFCs [69]. In line with the literature, it is calculated that the SOFC operating time of 35,000 h minimizes the cost under the conditions examined in this study.

9. Conclusions

In this study, an innovative SOFC-ICE-CCS integrated system using reformed NG as fuel for LNG carriers is proposed. Thermodynamic models of integrated systems are simulated by Aspen HYSYS, and 4E analyses are carried out. The SOFC is also modeled electrochemically,

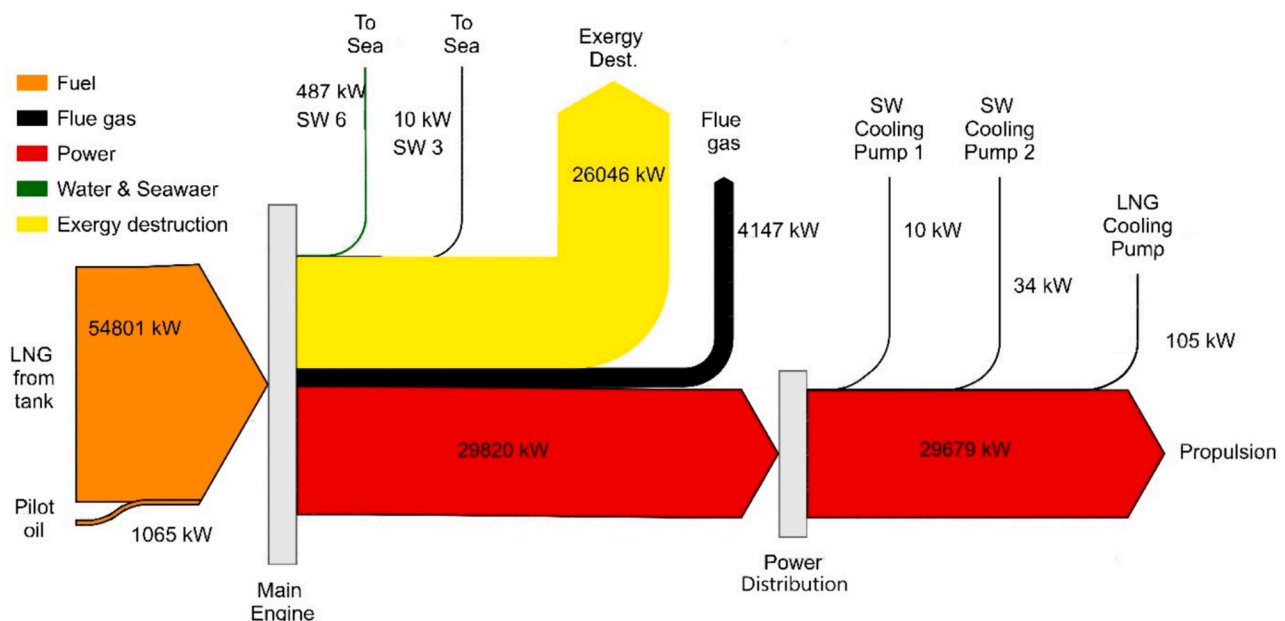


Fig. 20. Sankey diagram of the conventional power system showing the exergy flow and destruction.

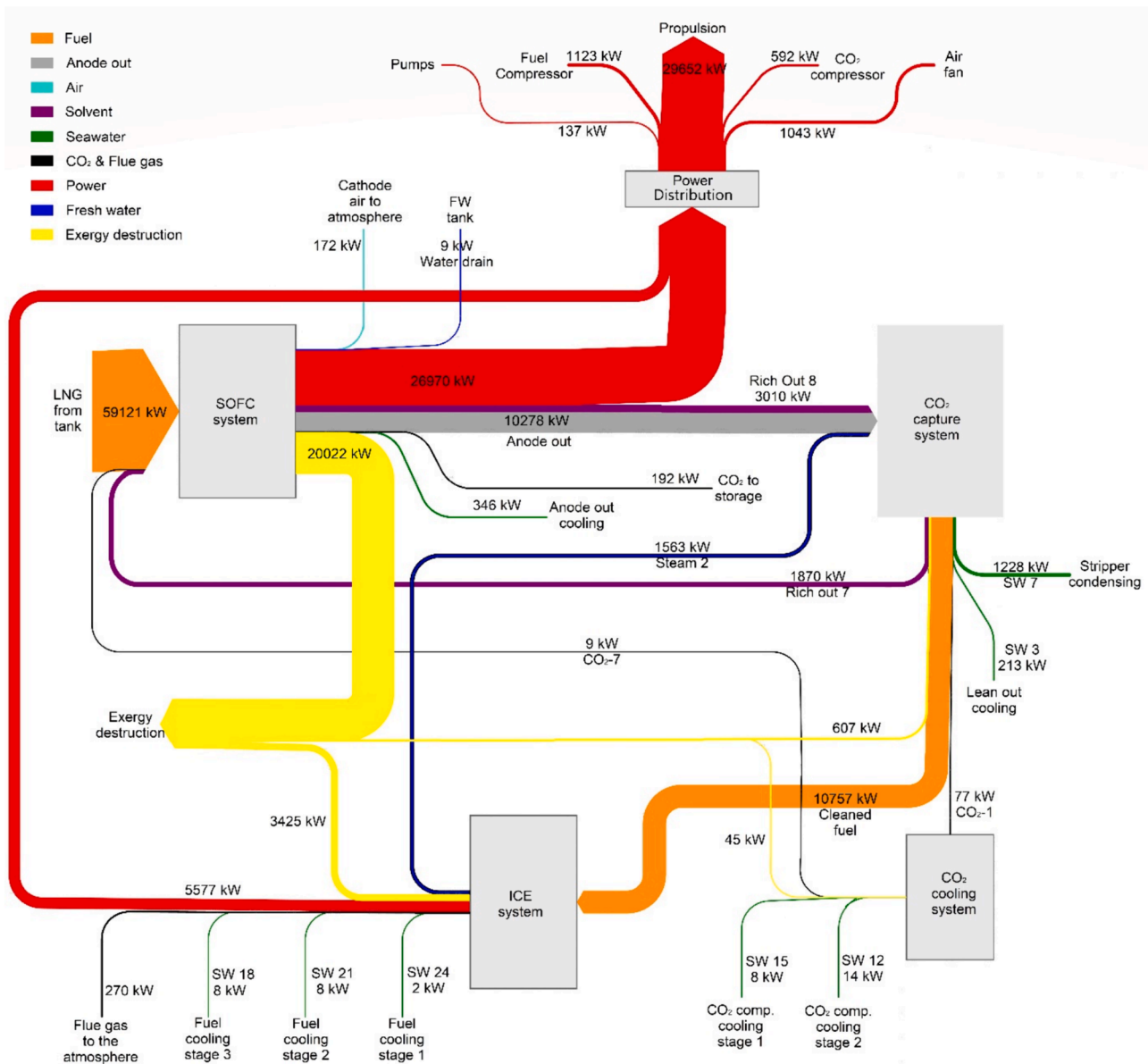


Fig. 21. Sankey diagram of the SOFC-ICE-CCS integrated system with reformed fuel showing the exergy flow and destruction.

and the effects of SOFC operating temperature, current density, and operating time due to degradation are investigated. According to the results of the 4E analysis, the most suitable fuel cell operating temperature, current density, and operating time parameters are determined, and the effect of the losses in the fuel cell depending on the current density change is examined. As an alternative to the proposed system, H₂-fuelled another system with SOFC-ICE-ST integration is modeled, and 4E analyses of these two systems and the power systems currently available on ships were performed, and the results were compared.

Although SOFC-ICE integration has been previously investigated on ships and some of the energy, exergy, and environmental analyses have been performed in these studies, to our knowledge, 4E analyses have been performed for the first time in this study. This study has contributed to the literature by considering cell degradation in SOFC-ICE integration for the first time. Although the effects of SOFC current density and operating temperature on the results have been examined before, this is the first study in which these parameters are analysed by considering degradation. The integration of the CCS system into the SOFC-ICE system is proposed and achieved first time in both ships and land facilities with this study.

The key conclusions can be summarised as follows:

- SOFC efficiency is at least 20.6 % lower for reformed fueled SOFC than H₂ fueled. When the loss mechanisms are analyzed, it is seen that the difference is largely related to parasitic losses and the Nernst potential obtained.
- CO₂ reduction rate, energy, and exergy efficiencies decrease over time as the SOFC operation time increases due to the degradation effect. The cost-minimising SOFC operation time is calculated to be around 35,000 h depending on the selected conditions. The initial investment cost of SOFC, cell design that reduces degradation and the creation of more comprehensive degradation models should be given importance in the future.
- When the proposed SOFC-ICE-CCS integration is compared with the alternative SOFC-ICE-ST integration, the average overall energy efficiencies of the systems at 35,000 h SOFC lifetime are calculated as 44.1 % and 57.5 %, and CO₂ reduction rates as 76.1 % and 96.9 %, respectively. The energy efficiency of the conventional system is 52.8 %. With the integration of H₂-fuelled SOFC-ICE-ST, it is possible

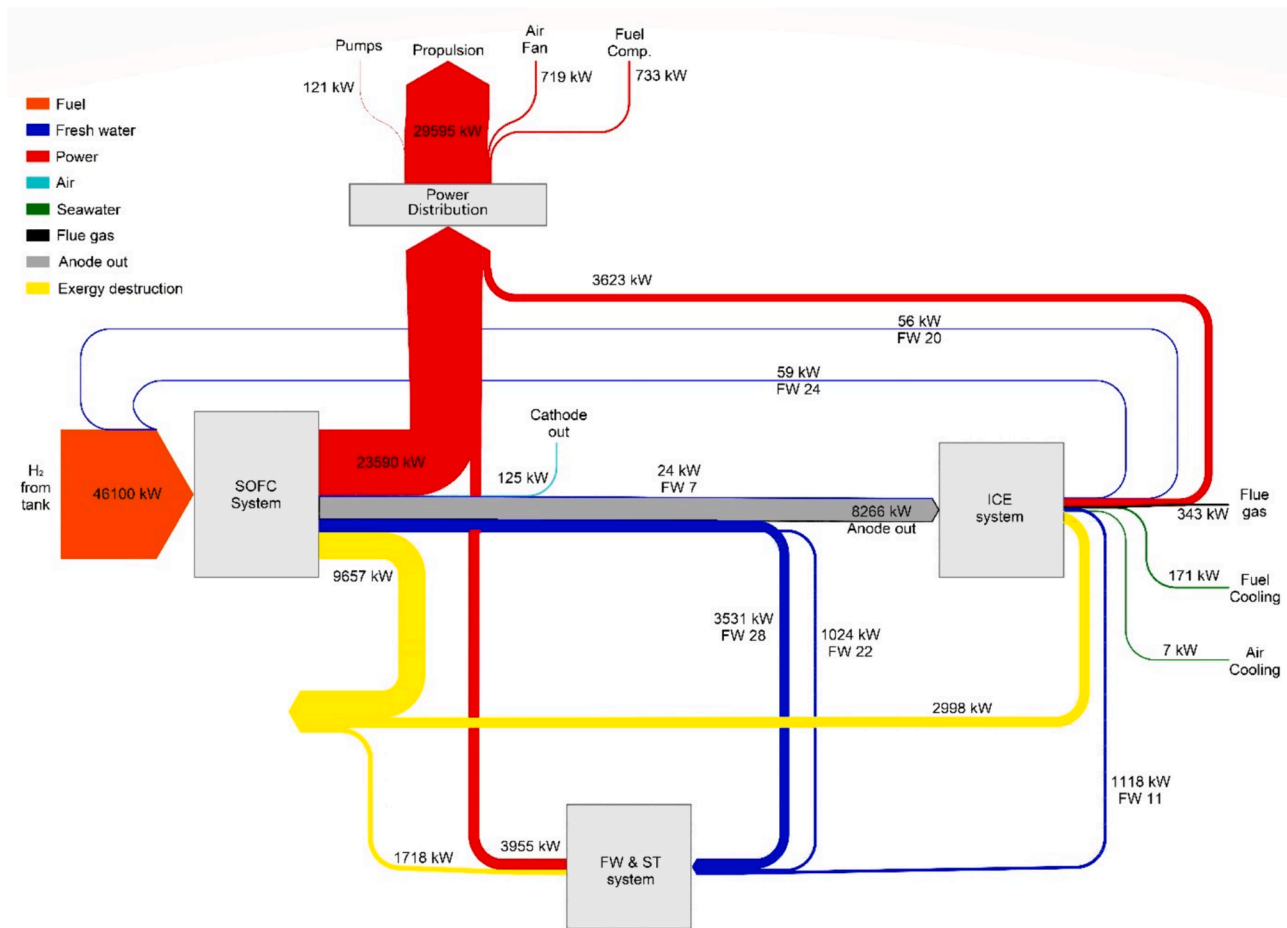


Fig. 22. Sankey diagram of the SOFC-ICE-ST integrated system with H₂ fuel showing the exergy flow and destruction.

to both increase energy efficiency and achieve almost zero emission transport at sea despite cell degradation.

- Despite its low efficiency, CCS-SOFC-ICE integration can reduce CO₂ by 76.1 % with waste heat due to its ability to operate at high temperatures and higher heat potential, while in the authors' previous studies, only 30–35 % of CO₂ could be captured with waste heat in CCS-ICE integration for the same ships.
- With SOFC-ICE-CCS integration, the absorber and stripper columns are considerably smaller than with ICE-CCS integration. In a Q-Flex type vessel, the SOFC-ICE integration captures only 34.6 % of CO₂, while the diameters of the absorber and stripper columns are 2.8 and 3.2 m, whereas in the SOFC-ICE-CCS integration, the diameters are 1.4 and 1.8 m respectively. Reducing the dimensions is important as it will reduce weight loss, facilitate the layout of the ship, and improve stability.
- When the exergy flow and destruction of the proposed systems are examined, it is seen that the exergy loss in the SOFC system is low compared to the alternative system, especially due to the use of reformed fuel and less utilization of waste heat potential.
- The conditions that minimize the cost of the proposed system are found to be around 900 °C for temperature and 2.5 kA/m² for current density. The CO₂ reduction cost under these conditions is calculated as approximately 711.1 \$/ton, while the cost under the same conditions in alternative SOFC-ICE-ST system integration is calculated as 1310 \$/ton.
- When the cost components of the studied systems are analyzed, it is found that the initial investment cost of both systems is much larger than the operating cost, the majority of the cost for the SOFC-ICE-

CCS system is due to the SOFC, for the SOFC-ICE-ST system is due to the H₂ storage tank.

Both systems are feasible to use in future ship power systems. Although the SOFC-ICE-ST integration is closer to the net zero target in emission and has higher efficiencies, the SOFC-ICE-CCS system may be a more effective solution in the transition to the net zero target due to its lower volume and mass requirements and lesser costs. To increase the feasibility of the proposed system, it is prioritized to reduce the initial investment cost of SOFC, and to examine and develop different storage options that reduce the initial investment cost for the alternative H₂-fuelled system in the future.

CRediT authorship contribution statement

Engin Güler: Writing – original draft, Visualization, Validation, Software, Investigation, Formal analysis, Data curation. **Selma Ergin:** Writing – review & editing, Supervision, Resources, Methodology, Investigation, Funding acquisition, Conceptualization.

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.

Data availability

Data will be made available on request.

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Appendix A. Supplementary data

Supplementary data to this article can be found online at <https://doi.org/10.1016/j.enconman.2024.118660>.

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