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## Fuel cell integrated carbon negative power generation from biomass

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#### HIGHLIGHTS

## GRAPHICAL ABSTRACT

- The conceptualization of a novel system consisting of a steam gasification facility, fuel-cell modules, and organic Rankine cycle.
- Detailed investigation of the steam gasification facility for hydrogen rich syngas production.
- $\bullet$  Results shows up to 99.2%  $\rm CO_2$  was capture using molten carbonate fuel cell.
- Thermodynamic, economic and life cycle carbon emission analyses of the novel system.

#### ARTICLE INFO

Keywords: Biomass Carbon Capture and Storage SOFC MCFC Techno-economic Analysis Life Cycle Analysis



## ABSTRACT

Combining biomass-fuelled power plant with carbon capture and storage allows  $CO_2$  to be removed from the atmosphere, considering biomass a carbon–neutral fuel. In the present study, a biomass-based  $CO_2$  negative system has been proposed, which combines a biomass steam gasification facility with a solid oxide fuel cell, a post-combustion carbon capture facility with a molten carbonate fuel cell (MCFC), and waste heat recovery using the organic Rankine cycle. A techno-economic analysis of two scenarios, namely a) with MCFC-based  $CO_2$  capture and b) without  $CO_2$  capture, was conducted. Integration of MCFC and carbon capture system was able to capture 99.2% of  $CO_2$ . It was found that the energy efficiency of the system was decreased by 9.43%, with the incorporation of  $CO_2$  capture acalculated as 62% and 70.22%, respectively. The economic analysis reveals that the levelized cost of electricity (LCOE) for the configuration with  $CO_2$  capture is estimated to be 0.062 \$/kWh and 0.052 \$/kWh, respectively. Finally, life cycle  $CO_2$  emissions for both the scenarios have been performed, and the analysis reveals that the proposed  $CO_2$  negative system is able to capture 1100 tonnes of  $CO_2$  per year.

## 1. Introduction:

global temperature rise to be well below  $2^{\circ}$  above pre-industrial levels [1]. Every emission-producing sector must decarbonize at a rate of 7.5 % annually to meet this target. [2]. This requires a huge amount of investment in technology and infrastructure as well as a shift away from

To combat climate change, the 2016 Paris Agreement targets the

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Nomenclature:		R Rk	Area specific resistance, $\Omega.m^2$ Thermal resistance
А	Area, m <sup>2</sup>	RMSE	Root Mean Square Error
AB	Afterburner	SB	Steam to biomass ratio
ASU	Air separation unit	SOFC	Solid oxide fuel cell
BW	Blower	Т	Temperature, °C
C <sub>biomass</sub>	Biomass cost, \$/GJ	TD	Transmission and distribution loss
CEPCI	Chemical Engineering Plant Cost Index	UF	Fuel utilization factor
CCS	Carbon capture and storage	V	Voltage, V
GCD	Gas cleaning device	Ŵ	Power, kW
HRSG	Heat recovery steam generator	у	Mole fraction
Ė	Exergy, kW		
F	Faraday's constant, 96,485C/mol	Greek syr	nbols
Н	Hours of operation, hr	η	Efficiency
HEX	Heat exchanger	ζ	Adjustment factor
HHV	Higher heating value, MJ/kg	Subscript	
i <sub>n</sub>	Discount rate, %	a	anode
Ι	Current, A	c	cathode
j	Current density, A/m <sup>2</sup>	ch	Chemical
LHV	Lower heating value, MJ/kg	ex	Exergy
LCOE	Levelized cost of electricity, \$/kWh	inv	Invertor
ṁ	Mass flow rate, kg/s	PC	Procurement cost
MCFC	Molten carbonate fuel cell	nh	Physical
MF	Multiplication factor	SVS	System
Ν	Number of cells	TOC	Total overnight cost
ORC	Organic Rankine cycle	TPC	Total plant cost
Р	Pressure, bar		F

current habits of consumption without considering the consequences [3].

Biomass is considered as one of the most promising renewable energy sources [4] due to its salient features such as near zero emissions, wide availability, and relatively low cost compared to other prominent energy sources [5]. Gasification is a typical method for generating syngas, a combustible gaseous mixture, from various biomass resources, such as municipal solid waste, wood, rice husks, paper, etc. Methane, hydrogen, carbon monoxide, nitrogen, and carbon dioxide are the typical components of the syngas, produced in the biomass gasification process [6]. In contrast to other traditional techniques, such as pyrolysis, combustion, and fermentation, gasification produces a higher amount of syngas with lower carbon emissions [7]. Air, O2, and steam are the typical gasifying agents used in the biomass gasification process [8]. Among the various gasifying agents, steam is thought to be the most effective in producing hydrogen-rich syngas with an optimum heating value via the biomass gasification process [9]. Although steam gasification is an efficient technique compared to other biomass conversion methods, the syngas still contains undesirable impurities such as tar, char, particulates, etc. These contaminants in the syngas must be reduced to acceptable levels for further utilisation depending upon the end-user application [10].

A solid oxide fuel cell (SOFC) is a highly efficient energy conversion device that is able to directly convert the chemical energy of fuel to electrical energy [11]. The main advantages of using SOFC are fuel flexibility, noise-free operation, low environmental emissions, and fewer corrosion problems compared to other types of fuel cells [12]. The SOFC operates at a high temperature, similar to that of a biomass gasifier, making it suitable for integration. Furthermore, as a by-product, SOFC generates high-quality heat, which may be effectively utilised in cogeneration, tri-generation, or multi-generation systems.

The molten carbonate fuel cell (MCFC) also operates at high temperatures and has the unique ability to operate with CO<sub>2</sub>-containing gas mixtures in both anode and cathode streams [13]. At the exhaust of the cathode stream, MCFCs can concentrate CO<sub>2</sub> while also producing electricity at the same time. Thus, MCFC can act as a  $CO_2$  separator. This feature expands the possibility of incorporating MCFC into a system that emits a  $CO_2$ -rich gas stream [14]. Barckholtz et al. [15] showed that by employing MCFC, close to 90 % of  $CO_2$  can be avoided from various flue gas sources while generating additional electrical power.

Previously, a broad range of studies have been carried out on the analysis and assessment of various biomass-based SOFC configurations. However, in most of the previous studies, air was employed as a gasifying agent. Yuksel et al. [16] investigated a SOFC-based combined system integrated with an air gasification facility. The combined system's overall energy and exergy efficiencies were reported to be 56.17 % and 52.83 %, respectively. Wu et al. [17] proposed a novel combined system integrating an air gasification facility, SOFC, and homogeneous charge compression ignition (HCCI) engine. They reported that the combined system's energy and exergy efficiencies could reach up to 68 % and 51 %, respectively. Peng et al. [18] investigated the technoeconomic analysis of a combined system integrating SOFC, plasma gasification facility, gas turbine, and supercritical CO<sub>2</sub> cycle. The energy and exergy efficiencies of the system were estimated to be 57.56 % and 59.30 %, respectively. Furthermore, the economic analysis revealed that the net present value of 110 M\$ can be attained by the system during its 20-year lifespan. In another work, the SOFC module was proposed to be integrated with a steam gasification plant, and heat pipes were employed to supply heat to the gasifier by Mojaver et al. [19]. They reported that the system could provide an exergy efficiency of 44.22 % and produce a net power of 535 kW. Papurello et al. [20] investigated direct coupling between biomass energy and SOFC tubular cells. The proposed system used 25 cells, with a maximum electrical power output of 225 W. Cao et al. [21] investigated a combined heat and power system integrating a proton conducting SOFC, an air gasification unit, and a supercritical CO<sub>2</sub> Brayton cycle. The system generates a maximum of 138 kW of electricity and 195 kW of heat.

From the previous discussion, it is evident that very few studies can be found on the integration of steam gasification technology with SOFC. Thus, in this paper, a detailed investigation of an SOFC-based system integrated with a steam-only gasification facility has been performed. Furthermore, MCFC has been integrated for the  $CO_2$  separation process as well as to get some additional power. The major contributions to the work are listed below:

- The conceptualization of a novel system integrated with biomass gasifier, SOFC, MCFC, and ORC.
- Detailed investigation of the steam gasification facility for hydrogenrich syngas production.



Fig. 1. Layout of the proposed biomass-fuel cell combined system. where, AB: Afterburner; ASU: Air Separation Unit; BW: Blower; CCS: Carbon Capture and Storage; GCD: Gas Cleaning Device; HEX: Heat Exchanger; MCFC: Molten Carbonate Fuel Cell; ORC: Organic Rankine Cycle; OVT: Organic Vapour Turbine; P: Pump; SOFC: Solid Oxide Fuel Cell; VG: Vapour Generator.

- Up to 99.2 % CO<sub>2</sub> capture using MCFC.
- Thermodynamic, economic, and life cycle carbon emission analyses of the proposed system.

#### 2. System description

Fig. 1 shows the schematic outline of the proposed integrated biomass-based power generation system. The hybrid system consists of a gasifier and gas cleaning unit, SOFC sub-system, MCFC based CO<sub>2</sub> separation sub-system, and an organic Rankine cycle (ORC) for waste heat recovery. Steam gasification of biomass takes place at the gasifier unit, producing raw syngas. The necessary heat for the steam gasification process is supplied to the gasifier using a heat pipe from the afterburner unit. The required steam is produced using a heat recovery steam generator (HRSG) and then supplied to the gasifier unit. A hot gas cleaning device (GCD) has been installed to remove undesired elements such as char, tar, etc. Furthermore, the cleaned producer gas is fed into the anode channels of SOFC and MCFC units using a blower (BW1). Fresh air is heated up by a heat exchanger (HEX3) and then supplied to the cathode channel of the SOFC unit using a blower (BW3). The unutilized fuel in the SOFC unit is completely burned using a primary afterburner (AB1) unit. A CO2-rich stream from AB1 exhaust is supplied to the cathode channel of the MCFC. The unutilized fuel in the MCFC unit is completely burned at the secondary afterburner (AB2) unit using pure oxygen (98 %). An air separation unit (ASU) has been installed to get the desired pure oxygen. The high temperature exhaust from the AB2 unit is utilised to heat up the incoming air to the SOFC. The major constituents of the AB2 exhaust stream are H<sub>2</sub>O and CO<sub>2</sub>. This stream is further fed to the CO<sub>2</sub> capture and storage (CCS). The CCS unit contains a moisture separator and some CO2 compressors. The H2O is separated from the exhaust stream of AB2 by using a moisture separator. Furthermore, pure CO<sub>2</sub> is compressed to 110 bar and stored as liquid CO2. In the bottoming cycle, ORC has been used to generate excess power by utilising waste heat from the topping cycles. The working fluid is toluene in this study. The global warming potential (GWP) of toluene is minimal, which is 3, and the ozone depletion potential (ODP) of toluene is zero, which is also encouraging for the environment [22].

#### 3. Methodology

#### 3.1. Hydrogen-rich syngas production from various biomass feedstocks

Hydrogen is critical to meeting net-zero greenhouse gas objectives. Methods for producing low-cost, greener, and long-lasting hydrogen have gained a lot of interest recently [23]. Although hydrogen is often obtained from fossil fuels or the electrolysis of water, steam gasification of renewable energy sources such as biomass and solid waste can be deemed as a more sustainable and environmentally friendly solution [24]. The key benefits of this technology are its adaptability to various incoming biomass resources, production capacity flexibility, and potential to minimise greenhouse gas emissions [25,26].

To this end, a model for steam gasification of biomass was developed using the ASPEN Plus software package. The schematic of the model is shown in Fig. 2. Five different biomass feedstocks were used individually to produce syngas. They are rice husk, wood residue, raw food waste, sugarcane bagasse, and algal biomass. The properties of the biomass feedstocks are tabulated in Table 1. The gasification temperature, pressure, and the steam to biomass ratio were kept constant at 800 °C, 1.06 bar, and 0.8, respectively, for all cases of operation. Heat pipes are used to supply heat to the gasifier unit, allowing it to generate hydrogen-rich syngas. In heat pipes, sodium is considered as the working fluid. The heat from the primary afterburner is used to bring the sodium to its gaseous state. Saturated liquid is converted to saturated vapour by absorbing the latent heat of vaporisation. On the other hand, the thermal resistance of the heat pipes operates as an impediment to supplying all the absorbed heat to the gasifier reactor. Each heat pipe is made up of nine different thermal resistances. The following equations given below can be used to calculate the amount of heat transfer to the gasifier unit [12].

$$\dot{Q}_{transfer} = \frac{\Delta T}{\sum_{k=1}^{9} R_k} N_{HP} \tag{1}$$

where,  $\dot{Q}_{transfer}$  is the required heat rate for biomass gasification;  $N_{HP}$  is the number of heat pipes and  $R_k$  is the thermal resistance.

$$\Delta T = T_{AB1} - T_{Gasifier} \tag{2}$$

where,  $\Delta T$  is temperature difference.

Initially, the thermal resistance of each heat pipe was estimated and then summed over for the total number of heat pipes used here in this study. To estimate the thermal resistance, a detailed heat transfer analysis has been carried out by Mojaver et al. [12] and Panopoulus et al. [27]. The same methodology has been employed to determine the individual resistances. Considering the convection and radiation modes of heat transfer, the thermal resistances of the evaporator ( $R_1$ ) and the condenser part ( $R_9$ ) have been estimated based by equations written below e.

$$R_{1,9} = \frac{1}{2 \times h_{c,e} \times A_{es}}$$
(3)

$$A_{es} = \pi \times D_o \times l_{c,e} \tag{4}$$

where,  $A_{es}$  is the external surface of the heat pipe tube,  $h_{c,e}$  is the heat transfer coefficient,  $D_o$  is the outer diameter of the heat pipe, l is the length and subscript "c" represents condenser and subscript "e"



Fig. 2. Process simulation and evaluation using Aspen Plus software package.

#### Table 1

Properties of Biomass feedstocks.

	Composition (%)	Rice Husk [30]	Wood Residue [31]	Raw Food Waste [32]	Sugarcane Bagasse [33]	Algal Biomass [34]
Ultimate analysis	Ash	21.68	0	0	2.94	0
	Carbon	38.43	50.26	45.71	46.96	35.27
	Hydrogen	2.97	6.72	6.72	5.72	4.71
	Nitrogen	0.49	0.16	2.91	0.27	4.44
	Sulphur	0.07	0	0	0.02	0.73
	Oxygen	36.36	42.66	44.66	44.05	54.85
Proximate analysis	Moisture	9.95	5.01	9.6	9.1	4.8
	Fixed Carbon	14.99	16.94	13	16.36	11.5
	Volatile Matter	55.54	77.71	73.78	71.87	65.6
	Ash	19.52	0.34	3.62	2.67	18.1
	HHV (MJ/kg)	12.69	21.05	19.21	18.51	12.12
	LHV (MJ/kg)	11.86	19.57	17.63	17.14	11.06

represents evaporator.

The conduction thermal resistances inside the heat pipe shell  $(R_2 \text{ and } R_8)$  have been estimated by the following relation.

$$R_{2,8} = \frac{\ln(D_o/D_{in})}{2\pi \times k_{hp} \times l_{c,e}}$$
(5)

where,  $k_{hp}$  is the thermal conductivity of the heat pipe shell and  $D_{in}$  is the internal diameter of the heat pipe.

The thermal resistances  $\mathsf{R}_3$  and  $\mathsf{R}_7\,$  have been estimated by the following relation

$$R_{3,7} = \frac{\ln(D_{in}/D_{w,in})}{2\pi \times k_w \times l_{c,e}}$$
(6)

where,  $k_w$  is the thermal conduction of solid and liquid of a wetted mesh wick and  $D_{w,in}$  is the heat pipe wick internal diameter.

Then the internal resistances  $(R_4 \text{ and } R_6)$  due to liquid and steam have been estimated by the following relation.

$$R_{4,6} = \frac{\overline{R} \times T^2 \times (2 \times \pi \times \overline{R} \times T)^{0.5}}{H_{lv} \times p_v \times 2 \times \pi \times D_{w,in} \times l_{c,e}}$$
(7)

where,  $\overline{R}$  is the ideal gas constant,  $H_{iv}$  is the latent heat of vaporisation, and  $p_v$  is the vapour pressure of the working fluid.

As suggested by the earlier researchers [12,27], the resistance ( $R_5$ ) caused by the drop in temperature of the steam transferred from the condenser and evaporator is neglected. Finally, the total thermal resistance has been estimated by taking the sum of all the thermal resistances as shown below.

$$\mathbf{R}_{\text{Total}} = \sum_{k=1}^{9} \mathbf{R}_k \tag{8}$$

The higher and lower heating values for all the feedstocks were calculated using the following equations [28].

$$HHV_{fuel} = 349.1 \times C + 1178.3 \times H - 103.4 \times O - 15.1 \times N - 21.21 \times Ash$$
(9)

$$LHV_{fuel} = HHV_{fuel} - h_{fg}(M + 9 \times H)$$
(10)

where,  $h_{\rm fg}$  represents enthalpy of vaporization of water (2260 kJ/kg). The steam to biomass ratio is an important parameter in gasification process and it can be expressed as follows

$$SB = \frac{Steam(kg) + Fuel Moisture(kg)}{Dry Biomass(kg)}$$
(11)

The important chemical reactions taking place in the gasification process are follows [29]

 $C + O_2 \rightarrow CO_2 \Delta H = -393.0 kJ/mol \tag{12}$ 

$$C + 0.5O_2 \rightarrow CO\Delta H = -112.0kJ/mol \tag{13}$$

 $C + CO_2 \rightarrow 2CO\Delta H = +172.0 kJ/mol \tag{14}$ 

$$C + H_2 O \rightarrow CO + H_2 \Delta H = +131.0 kJ/mol \tag{15}$$

$$CO + H_2 O \rightarrow CO_2 + H_2 \Delta H = -41 kJ/mol$$
(16)

$$C + 2H_2 \rightarrow CH_4 \Delta H = -74.0 kJ/mol \tag{17}$$

$$H_2 + 0.5O_2 \rightarrow H_2O\Delta H = -242.0kJ/mol \tag{18}$$

$$CH_4 + H_2O \rightarrow CO + 3H_2\Delta H = +206.0kJ/mol$$
<sup>(19)</sup>

$$H_2 + S \rightarrow H_2 S \Delta H = -20.2 kJ/mol \tag{20}$$

Fig. 3 shows the producer gas composition for five different biomass feeds. The highest hydrogen percentage is found in wood residue (49.21 %), followed by raw food waste (47.13 %), sugarcane bagasse (46.21 %), rice husk (40.21 %), and algal biomass (34.12 %). Thus, in our analysis, wood residue has been chosen as a biomass feed.

#### 3.1.1. Gasifier model validation

Table 2 compares the results of the present Aspen Plus equilibrium model with the experimental and numerical results of Loha et al. [30]. The composition of syngas has been compared on a dry and N<sub>2</sub> free basis. The comparison indicates excellent agreement with both the numerical and experimental data, with a Root Mean Square Error (RMSE) of 2.473 and 3.318, respectively. The simulated CH<sub>4</sub> data in this study is lower compared to the experimental value. This is due to the fact that the equilibrium gasification models generally underpredict CH<sub>4</sub> compositions [35].

## 3.2. Solid oxide fuel cell

The SOFC runs at a high temperature (700–1100 °C) and has a high chemical energy to electricity conversion potential [36,37]. As the operating temperature is high, various types of fuel, such as methane, producer gas, biogas, NH<sub>3</sub> etc., can be internally reformed without the use of a separate reformer. In this study, an internal reforming type solid oxide fuel cell model was chosen. The major chemical reactions that are taken into account in the SOFC model are listed below [38].

$$CH_4 + H_2O \rightarrow CO + 3H_2\Delta H = +206.0kJ/mol$$
<sup>(21)</sup>

$$CO + H_2 O \rightarrow CO_2 + H_2 \Delta H = -41 kJ/mol$$
<sup>(22)</sup>

$$H_2 + 0.5O_2 \rightarrow H_2O\Delta H = -242.0kJ/mol \tag{23}$$

Current flow though SOFC can be determined by the following relation [39]

$$I_{FC} = \frac{\dot{m}_{a,in} \times \left(y_{H_2} + y_{CO} + y_{CH_4}\right) \times 2 \times F}{M_{mol,a}}$$
(24)



Fig. 3. Properties of Syngas produced from steam gasification of Biomass.

Table 2	
Casifier model validation	

Rice Husk	Simulation(Current study)	Simulation [30]	Experimental [30]
CO (%)	26.1	29.37	27.5
CO <sub>2</sub> (%)	19.84	18.3	19.5
H <sub>2</sub> (%)	53.88	50.79	48.8
CH4 (%)	0.18	1.54	4.2
RMSE		2.473	3.318

where,  $y_{H_2}$ ,  $y_{CO}$ ,  $y_{CH_4}$  are the concentrations of  $H_2$ , CO and CH<sub>4</sub> at the inlet; F is the Faraday constant;  $M_{mol,a}$  is the mol mass of anode gas and  $\dot{m}_{a,in}$  is the anode inlet mass flow rate.

In practice, only a portion of the syngas gets converted at the SOFC unit. The fuel utilization factor (UF) denotes the ratio of actual to maximum feasible conversion and it is represented by the following equation

$$UF = \frac{I}{I_{FC}}$$
(25)

where, I denote actual current flow.

The current density is considered to be proportional to the voltage loss. The proportionality constant ( $R_{PC}$ ) is indicated with the equivalent cell resistance by the analogy with Ohm's law.

The SOFC voltage can be estimated by the following relation [40]

$$V_{SOFC} = \frac{\Delta G}{2F} + \frac{RT_{SOFC}}{2F} ln \left( \frac{y_{O_2}^{0.5} \times y_{H_2}}{y_{H_2O}} \times P_{SOFC}^{0.5} \right) - I \times R_{SOFC}$$
(26)

where,  $R_{SOFC}$  is the area specific resistance for SOFC;  $\Delta G$  is the standard Gibbs free energy;  $T_{SOFC}$  is operating temperature of SOFC;  $P_{SOFC}$  is the operating cell pressure;  $y_{H_{2}O}$  represents mole fraction of  $H_2O$ ;  $y_{O_2}$  is the mole fraction of  $O_2$ , and R is the universal gas constant.

The power generated by SOFC module can be estimated as follows

$$\dot{W}_{SOFC} = N_{SOFC} \times j \times A_{SOFC} \times V_{SOFC} \times \eta_{inv}$$
(27)

where,  $N_{SOFC}$  is the number of cells; j is the current density;  $A_{SOFC}$  is the cell area;  $\eta_{inv}$  is the efficiency of inverter.

## 3.2.1. SOFC model validation

The results of the SOFC model used in this study have been validated

with the experimental results of Singhal et al. [41]. It can be observed from Fig. 4 that the simulated results match well with the experimental results, with a maximum error of 3.7 %.

#### 3.3. Molten carbonate fuel cell

Generally, in fuel cells hydrogen or hydrogen-containing gas is used as fuel. With the help of an electrochemical reaction, the chemical energy of the fuel is converted into electrical energy without direct combustion of the fuel. Unlike other fuel cells, the MCFC has a unique feature of taking CO<sub>2</sub> containing gas into the cathode side. Whereas like the other fuel cells, it takes hydrogen or hydrogen-containing gas as fuel input into the anode side. In the cathode channel, the CO<sub>2</sub> forms carbonate ions with the help of oxygen present in the gas stream entering the cathode side. The carbonate ions flow through the electrolyte present in the MCFC and reach the anode side. In the anode channel, the carbonate ions react with hydrogen and generate CO<sub>2</sub> and water. The carbonate ions of the molten carbonate salt (electrolyte) act as the charge carriers. There is no extra CO<sub>2</sub> generation because the amount of carbonate ions produced on the cathode side has been consumed by the anode side, apart from CO<sub>2</sub> separation from the cathode side to the anode side. This gives a lucrative opportunity to separate CO<sub>2</sub> from an after-combustion gas stream and generate electricity simultaneously. As an outcome of this thermo-chemical process at the anode exhaust CO2 concentrated gas stream is produced with some amount of moisture and excess fuel. On the other side, comparatively more clean gas with a very low concentration of CO<sub>2</sub> is generated at the cathode exhaust. Direct pure hydrogen can be used as the fuel input in the MCFC. Steam reforming is utilised to produce hydrogen when methane used as fuel. If CO-containing gas is used as fuel, the water gas shift reaction is used to produce hydrogen. The gas containing both CO and  $\mathrm{CH}_4$  (such as syngas) utilises both the steam reforming reaction and the water gas shift reaction to generate hydrogen, which is the primary fuel used in MCFC. The electrochemical reactions occurring at the MCFC are schematically represented in Fig. 5.

$$H_2 + \frac{1}{2}O_2 + CO_2 \to H_2O + CO_2$$
(28)

$$CH_4 + H_2 O \rightarrow H_2 + CO \tag{29}$$

$$H_2O + CO \rightarrow H_2 + CO_2 \tag{30}$$



Fig. 4. Simulation and experimental results of a single cell at 1000  $^\circ$ C with 89 % H<sub>2</sub> and 11 % H<sub>2</sub>O as fuel.



Fig. 5. Molten carbonate fuel cell.

The MCFC voltage was estimated by the following relation

$$V_{MCFC} = \frac{\Delta G}{2F} + \frac{RT_{MCFC}}{2F} ln \left( \frac{y_{O_2,c}^{0.5} \times y_{H_2,a} \times y_{CO_2,c}}{y_{H_2O,a} \times y_{CO_2,a}} \times P_{MCFC}^{0.5} \right) - I \times R_{MCFC}$$
(31)

where,  $R_{MCFC}$  is the area specific resistance for MCFC;  $\Delta G$  is the standard Gibbs free energy;  $T_{MCFC}$  is operating temperature of MCFC  $P_{MCFC}$  is the operating cell pressure;  $y_{H_2,a}$  represents mole fraction of  $H_2$  in the anode channel;  $y_{O_{2,c}}$  is the mole fraction of  $O_2$  in the cathode channel;  $y_{CO_{2,c}}$  is the mole fraction of  $CO_2$  in anode channel;  $y_{CO_{2,c}}$  is the mole fraction of  $CO_2$  in anode channel;  $y_{CO_{2,c}}$  is the mole fraction of  $CO_2$  in cathode channel and R is the universal gas constant.

The power generated by MCFC module can be estimated as follows

$$\dot{W}_{MCFC} = N_{MCFC} \times j \times A_{MCFC} \times V_{MCFC} \times \eta_{inv}$$
(32)

where,  $N_{MCFC}$  is the number of cells; j is the current density;  $A_{MCFC}\,$  is the cell area;  $\eta_{inv}$  is the efficiency of inverter.

## 4. Performance analysis of the system

Exergy associated with the material flow can be estimated as the summation of physical exergy and chemical exergy as expressed below [42]

$$\dot{E} = \dot{E}_{ph} + \dot{E}_{ch} \tag{33}$$

where,

$$\dot{E}_{ph} = \sum_{i} \dot{n}_{i} ((\overline{h}_{i} - \overline{h}_{o}) - T_{o}(\overline{s}_{i} - \overline{s}_{o}))$$
(34)

$$\dot{E}_{ch} = \dot{n}_i \left( \sum_i y_i \overline{e}_{ch}^o + RT_0 \sum_i y_i ln y_i \right)$$
(35)

The specific chemical exergy of the biomass can be estimated as follows [43]

$$\overline{e}_{ch}^{biomass} = \zeta \times LHV_{biomass} \tag{36}$$

where,  $\zeta$  is the "adjustment factor "and it is estimated by the following equation [43]

$$\zeta = \frac{1.044 + 0.0160(\frac{H}{C}) - 0.3493(\frac{O}{C})(1 + 0.0531(\frac{H}{C}) + 0.0493(\frac{N}{C}))}{1 - 0.4124(\frac{O}{C})}$$
(37)

where, the mass percentages of hydrogen, carbon, and oxygen in the biomass are represented by H, C, and O, respectively.

The total power generated by the system can be estimated by the following equation.

$$\dot{W}_{sys} = \dot{W}_{SOFC} + \dot{W}_{MCFC} + \dot{W}_{ORC} - \dot{W}_{Auxilary}$$
(38)

Exergetic efficiency  $(\eta_{ex})$  of the system can be determined by the following relation

$$\eta_{ex} = \frac{\dot{W}_{sys}}{\dot{E}_{sys,in}} \tag{39}$$

where, input exergy to the system is denoted by  $\dot{E}_{sys,in}$ .

#### Table 3

Input parameters for technical analysis.

Blocks	Components	Parameters	Values	Units
Producer gas	Gasifier	Pressure	1.06	bar
generation and gas		Temperature	800	°C
conditioning block		Pressure loss	2	%
[8,44]	Gas cleaning	Pressure loss	1	%
	device	Exit temperature	30	°C
MCFC block [45-47]	MCFC	Current density	1500	A/m <sup>2</sup>
		Total area	110	m <sup>2</sup>
		Operating	650	°C
		temperature		
		CO <sub>2</sub> utilization	0.8	-
		factor		
		Fuel utilization	0.85	-
		factor		
		Oxygen	0.18	-
		utilization factor		
	DC-AC	Conversion	98	%
	inverter	efficiency		
SOFC block [48,49]	SOFC	Current density	1500	A/m <sup>2</sup>
		Total area	250	m <sup>2</sup>
		Area specific	$2.8 \times$	$\Omega.m^2$
		resistance	$10^{-05}$	
		Operating	750	°C
		temperature		
		Fuel utilization	0.85	
		factor		
		Oxygen	0.18	
		utilization factor		
	DC-AC	Conversion	98	%
	inverter	efficiency		
ORC block [50,51]	Turbine	Inlet temperature	280	°C
		Isentropic	85	%
		efficiency		
	Pump	Pressure ratio	20	-
		Isentropic	85	%
		efficiency		
	Condenser	Temperature	40.36	°C
		Pressure	0.08	bar
		Pinch point	10	°C
		temperature		
		difference		

The input parameters used in the analysis are provided in Table 3.

The technical performance of two distinct scenarios, one with  $CO_2$  capture and the other without  $CO_2$  capture, has been examined in this section. Only SOFC has been employed as power generation equipment in the system without  $CO_2$  capture. Fig. 6 compares the power outputs from the two scenarios. The system with a  $CO_2$  capture facility produces a net power output of 527.3 kW, of which the SOFC, MCFC, and ORC blocks generate 376.39 kW, 131.64 kW, and 19.27 kW, respectively. On the other hand, a net power generation of 597 kW is produced by the system without a  $CO_2$  capture facility.

Exergy and energy efficiencies of the system with and without  $CO_2$  capture are compared in Fig. 7. It is found that the energy efficiency of the system is decreased by 9.43 %, with the incorporation of  $CO_2$  capture facilities. Furthermore, the exergetic efficiencies of the system with and without  $CO_2$  capture are estimated to be 62 % and 70.22 %, respectively. The exergy flow diagram of the proposed system is depicted in Fig. 8.

#### 4.1. Economic analysis

The input parameters for the economic analysis are listed in Table 4. The levelized cost of electricity has been estimated by the equation provided below[60]

$$LCOE = \frac{YE}{UF_{CAP} \times H \times \dot{W}_{sys} \times (1 - TD)}$$
(40)

where, H is the annual operating hours,  $UP_{CAP}$  is the capital utilisation parameter, TD is the transmission and distribution loss and YE is the yearly expenditure of the system.

Yearly expenditure (YE) can be determined as follows

$$YE = TACC + O\&M + YRC + ACF$$
(41)

where, TACC: total annual capital cost, O&M: operation and maintenance cost, and ACF: cost of fuel.

The total annual capital cost (TACC) can be calculated by the following equation

$$TACC = NCAP \times (1 + MF_{PC}) \times (1 + MF_{TPC}) \times (1 + MF_{TOP}) \times CRF$$
(42)

where,  $MF_{TOC}$ : multiplication factor for total overnight cost;  $MF_{TPC}$ : multiplication factor for total plant cost;  $MF_{PC}$ : multiplication factor for procurement, construction and engineering cost; NCAC: net capital cost; and CRF: capital recovery factor.

The net capital cost (NCAC) is estimated by adding the capital cost of all the components of the proposed system as illustrated below.

$$NCAC = \sum_{i} CAP_i \tag{43}$$

The capital recovery factor (CRF) can be estimated by the following relation

$$CRF = \frac{i_n (1+i_n)^{yr}}{(1+i_n)^{yr} - 1}$$
(44)

where,  $i_n$  and  $y_r$  are "annual interest rate" and "operational years", respectively.

Table 5 shows the cost correlations of all the equipment.

The CO<sub>2</sub> that was captured has been proposed to be transported after the compression process at the CO<sub>2</sub> drying and compression unit. Saline aquifers, improved oil recovery, improved coal bed methane, improved gas recovery, etc. are often used for the sequestration of the segregated CO<sub>2</sub>. Due to its enormous capability for CO<sub>2</sub> storage, aquifers are taken into consideration for CO<sub>2</sub> storage in this study. For large-scale and longdistance CO<sub>2</sub> transfer, the pipeline is regarded as one of the most affordable and popular options [69]. Additionally, a 50 km pipeline is presumed to be advised for the local storage site. According to the literature, Table 5 summarises all the fundamental data related to cost input for the proposed CO<sub>2</sub> capture and sequestration facility considered



Fig. 6. The power output of two different scenarios.



Fig. 7. Exergy and energy efficiencies of two different scenarios.

in this study.

The equipment cost functions are updated by employing the Chemical Engineering Plant Cost Index (CEPCI). The present study considers the reference year as 2020. The following equation was used to update the equipment cost of the i<sup>th</sup> component [70].

$$CAP_{i,2020} = CAP_i \times \frac{CEPCI_{2020}}{CEPCI_{OY}}$$
(45)

where,  $CEPCI_{2020}$  and  $CEPCI_{OY}$  are the cost indices for the reference year and the year in which the cost correlation was created, respectively.

The proposed power generation system's operation and maintenance cost (O&M) is estimated to be 2.5 percent of NCAC. The system's yearly operation and maintenance cost (O&M) is calculated by the following equation

$$O\&M = 0.025 \times NCAC \tag{46}$$

To compensate for the shorter lifetime of fuel cell modules, an extra replacement cost (5 % of component cost in every-five years) has been considered. The total replacement cost (TRC) over the lifetime of the system is estimated as follows



Fig. 8. Exergy flow of the system.

#### Table 4

Input parameters for economic analysis.

Parameter	Value	Unit	Ref.
year	30	years	[52]
Н	8000	hours	[53]
UF <sub>CAP</sub>	0.85	-	[54]
TD	4	%	[14]
C <sub>biomass</sub>	2	\$/GJ	[55]
i <sub>n</sub>	3	%	[56,57]
MF <sub>TOP</sub> :	20.20	%	[58]
MF <sub>TPC</sub>	52.5	%	[58]
MF <sub>PC</sub>	9	%	[58]
Electricity selling price	0.1514	£/kWh	[59]

$$TRC = (0.05 \times 5) \times CAP_{SOFC} + (0.05 \times 5) \times CAP_{MCFC}$$
(47)

The yearly replacement cost (YRC) can be estimated by the following equation [71]

$$YRC = TRC \times \frac{i_n (1 + i_n)^{yr}}{(1 + i_n)^{yr} - 1}$$
(48)

The annual fuel cost (ACF) is determined as follows

$$ACF = C_{biomass} \times \dot{m}_{biomass} \times LHV_{biomass} \times H \times 3600$$
<sup>(49)</sup>

where  $\dot{m}_{biomass}$ : the fuel flow rate and  $C_{biomass}$ : biomass cost. Here in this section, economic performances of two different scenarios have been investigated, viz. with CO<sub>2</sub> capture and without CO<sub>2</sub> capture.

The Fig. 9 compares the economic parameters of the two scenarios. The annual capital cost (NCAC), annual operation and maintenance cost (O&M), yearly replacement cost (YRC), and annual fuel cost (AFC) are estimated to be US\$150296, US\$3757, US\$12424 and US\$46631, respectively for the proposed system with CO<sub>2</sub> capture. On the other hand, the NCAC, O&M, YRC, and AFC are estimated to be US\$139352, US\$3484, US\$13809 and US\$46631, respectively for the system without CO<sub>2</sub> capture.

The Fig. 10 compares the levelized cost of electricity of the two different scenarios. The LCOE of the system without  $CO_2$  capture and with  $CO_2$  capture is calculated to be 0.052 \$/kWh and 0.062 \$/kWh, respectively. The estimated levelized cost of electricity is on the lower side, and it is very competitive to other proposed systems and even fossil-fuel based systems with carbon capture and storage (CCS). According to Singh et al. [72], the LCOE for coal-fired facilities with CSS in the United States and China is 0.1479 \$/kWh and 0.0943 \$/kWh, respectively. In a recent study by Akrami et al. [73], LCOE of a biomass-based power system with MCFC based CO<sub>2</sub> capture was reported to be 0.0859 \$/kWh. The estimated cost of electricity for the proposed  $CO_2$  negative system is majorly contributed by the capital cost and replacement cost of the plant, as the system is majorly dominated by the SOFC

and MCFC units. However, the proposed system will be much more costcompetitive once the capital costs of the SOFC and MCFC are reduced, and their life span will improve with advancements in material research in the future.

## 4.2. Sensitivity analysis

A sensitivity analysis has been performed for estimating LCOE with the variation interest rates, projected year of operation and cost of fuel. Fig. 11(a) depicts the sensitivity analysis of the LCOE with different interest rates for two separate scenarios (with and without CO<sub>2</sub> capture). In the predicted line, it is observed that the LCOEs of both scenarios increase when interest rates rise. The maximum levels of LCOE are reached at a 12 % interest rate and are 0.1098 \$/kWh and 0.1312 \$/kWh for systems without and with CO<sub>2</sub> capture, respectively. Fig. 11 (b) illustrates the LCOE sensitivity analysis with projected plant life for two distinct scenarios (with and without CO<sub>2</sub> capture). In the anticipated line, the LCOE of both scenarios reduces as the projected year of operation increases. At a projected year of operation of 40 years, the minimal values of LCOE are 0.04605 \$/kWh and 0.05456 \$/kWh for systems without and with  $CO_2$  capture, respectively. Fig. 11(c) depicts the LCOE sensitivity analysis for two alternative scenarios with increasing biomass costs (with and without CO<sub>2</sub> capture). In the expected line, the LCOEs of both scenarios rise as fuel prices increase. The minimum LCOE values for systems without and with CO<sub>2</sub> capture are 0.05216 \$/kWh and 0.062 \$/kWh, respectively.

#### 4.3. Payback period

The Fig. 12 depicts a simple payback model applied to the proposed system with  $CO_2$  capture. The break-even time for the configuration with  $CO_2$  capture is estimated to be 4.2 years. Similarly, Fig. 13 depicts a simple payback model applied to the proposed system with  $CO_2$  capture. The break-even time for the configuration without  $CO_2$  capture is estimated to be 3.5 years. The payback period for the system with  $CO_2$  capture is longer due to the fact that the  $CO_2$  negative system is capital intensive because the system is dominated by MCFC and SOFC. However, it outperforms other biomass-based power systems with MCFC-based  $CO_2$  capture. For example, Akrami et al. [73] reported a payback period of 6.3 years for a biomass-based power system with MCFC-based  $CO_2$  capture.

#### 5. Life cycle analysis

The input parameters required for life cycle analysis (LCA) of different scenarios are provided in Table 6.

The system's life-cycle CO<sub>2</sub> emissions are influenced by several steps in the biomass supply chain (from cultivation to power production). Total CO<sub>2</sub> emissions from the plant harvesting and transport phase (in Fig. 14, mentioned as plant growth, photosynthesis, and associated transport) are estimated to be 1.741 t/year. The CO<sub>2</sub> emissions from the wood processing plant are estimated to be 42 t/year. In this work, the wood pellets, produced at the wood processing plant, are transported by truck to the shipping port, which is about 150 km away in Georgia, the United States. The life cycle CO<sub>2</sub> emissions at this stage are estimated to be 19.58 t/year. Then the pellets are considered to travel about 7,500 km by ship to a port in the United Kingdom. Consequently, the pellets are again transported to the proposed power system, which is about 50 km away from the port in the UK, using heavy-duty trucks [69]. The lifecycle CO2 emissions for the wood pellet transportation phase through ship are estimated to be 32.64 t/year. Similarly, the life cycle CO2 emission for the wood pellets transportation from the port to the power plant is estimated to be 3.264 t/year. During the transportation phase, there are requirements for the handling and storage of the pellets at the shipyard. The life cycle CO<sub>2</sub> emissions for the handling and storage of the pellets at the shipyard are estimated to be 0.9638 t/year. A certain

## D. Roy et al.

### Table 5

Cost correlations of all the equipment.

Equipment	Cost functions	Description	Ref.
Gasifier	$CAP_G = 1600 \times \left(\dot{m}_{fuel} \times 0.9 \times 3600\right)^{0.67}$	$CAP_G$ in $,\dot{m}_{fuel}$ in kg/hr	[61]
Gas cleaning unit	$CAP_{GCE} = 0.05 \times CAP_{G}$	$CAP_{GCE}, CAP_G$ in \$	[62]
Blower	$CAP_{BW} = \frac{39.5 \times \dot{m}_{air}}{0.9 - n_{mu}} \times \left(\frac{P_{out}}{P_{m}}\right) \times \ln\left(\frac{P_{out}}{P_{m}}\right)$	$CAP_{BW}$ in \$, $\dot{m}_{air}$ in kg/s, $P_{in}$ and $P_{out}$ in bar	[14]
SOFC	$CAP_{SOFC} = A_{SOFC} (2.96T_{cell} - 1907)$	$CAP_{SOFC}$ in \$, $A_{SOFC}$ in m <sup>2</sup> , $T_{cell}$ in K	[63]
SOFC inverter	$CAP_{inverter} = 10^5 \left(\frac{\dot{W}_{SOFC,DC}}{500}\right)^{0.7}$	$CAP_{inverter}$ in \$, $\dot{W}_{SOFC,DC}$ in kW	[63]
SOFC auxiliary	$CAP_{SOFCAUX} = 0.1 \times CAP_{SOFC}$	CAP <sub>SOFC.AUX</sub> in \$,CAP <sub>SOFC</sub> in \$	[63]
Afterburner	$CAP_{AB} = \frac{46.08 \times \dot{m}_{oxydont}}{0.995 - \frac{P_{out}}{n}} [1 + exp(0.018 \times T_{out} - 26.4)]$	$CAP_{AB}$ in \$, $P_{in}$ and $P_{out}$ in bar, $T_{out}$ in K	[14]
Heat exchanger	$P_{in}$ $CAP_{HEX} = 8500 + 409 \times A_{HEX}^{0.85}$	$CAP_{HEX}$ in D\$, $A_{HEX}$ in m <sup>2</sup>	[64]
Vapour generator/HRSG	$CAP_{HRSG/VG} = 6570  imes \left( rac{\dot{Q}}{LMTD}  ight)^{0.8} + 21276  imes \dot{m}_{steam} + 1184.4  imes \dot{m}_{gas}^{1.2}$	$\mathit{CAP}_{\mathit{HRSG/VG}}$ in \$, Q in kW,LMTD in K, $\dot{m}_{\mathit{steam}}$ and $\dot{m}_{\mathit{gas}}$ in kg/s	[61]
ASU	$CAP_{ASU} = 36.63 \times 10^{6} \times \left(\frac{\dot{m}_{oxygen}}{28.9}\right)^{0.7}$	$\dot{m}_{oxygen}$ in kg/s, $CAP_{ASU}$ in \$,	[46]
MCFC	$CAP_{MCFC} = A_{MCFC}(2.96T_{cell} - 1907)'$	$CAP_{MCFC}$ in \$, $A_{MCFC}$ in $m^2$ , $T_{cell}$ in K	[65]
MCFC auxiliary	$CAP_{MCFC,AUX} = 0.1 \times CAP_{MCFC}$	$CAP_{MCFC.AUX}$ in \$, $CAP_{MCFC}$ in \$	[66]
Heat Pipe	$\text{CAP}_{\text{Heat Pipe}}  = 33 \times \dot{Q}$	$CAP_{Heat Pipe}$ in \$, $\dot{Q}$ in kW.	[67]
Condenser	$CAP_{Condenser} = 1773  imes \dot{m}_{ORC}$	$\dot{m}_{ORC}$ in kg/s, $CAP_{Condenser}$ in \$,	[68]
Pump	$CAP_{Pump} = 3540 \times (W_P)^{0.71}$	$CAP_{Pump}$ in \$, $W_P$ in kW	[68]
Organic vapour turbine	$CAP_{OVT} = 1.5  imes (225 + 170  imes V_{OVT})$	$CAP_{OVT}$ in \$, $V_{OVT}$ in m <sup>3</sup> /s	[68]
$\mathrm{CO}_2$ drying and compression system	$CAP_{CCS} = 682870 \times \left(\frac{\dot{m}_{CO_2}}{24}\right)^{0.6}$	$CAP_{CCS}$ in \$, $\dot{m}_{CO_2}$ in kg/s	[45]
$\mathrm{CO}_2$ transport and storage System	$CAP_{CTS} = \begin{pmatrix} spcific \\ cost \end{pmatrix} \times \begin{pmatrix} total \ ton \ of \\ CO_2 \ captured \end{pmatrix}$	CAP <sub>CTS</sub> in \$	[69]



Fig. 9. Annual expenditure for the two scenarios.



Fig. 10. Levelized cost of electricity for the two scenarios.



Fig. 11. Sensitivity of LCOE with respect to (a) interest rate, (b) plant life and (c) biomass cost.



Fig. 12. Simple payback model for the system with CO<sub>2</sub> capture.



Fig. 13. Simple payback model for the system without CO<sub>2</sub> capture.

amount of  $CO_2$  is emitted through the cathode exhaust gas from the MCFC unit of the proposed plant. The life cycle  $CO_2$  emissions directly emitted from the system are estimated to be 10.26 t/year. The total amount of  $CO_2$  captured directly from the plant is estimated to be1268 t/ year. After the capture process, the  $CO_2$  has been considered to be compressed up to 110 bar or more and transported through a pipeline before its geological storage at a suitable site near the proposed power plant. There is a certain amount of  $CO_2$  emitted during the compression, transport, and from the underground storage as well as via the fugitive emission process. The emissions during the transportation and sequestration of the captured  $CO_2$  are also considered in this study. The

methodology used in the IPCC report estimates that pipeline transportation accounts for 0.2-23.2 t/km/y of the pipeline's CO<sub>2</sub> emissions, while compressor emissions range from 7.0 to 116.1 t CO<sub>2</sub>/MW/y [69]. The overall CO<sub>2</sub> emission factor of aquifer storage is 7.01 kg CO<sub>2</sub>/t CO<sub>2</sub> [74]. For the simplicity and robustness of the whole energy analysis, the total final pressure of the dry compressed CO<sub>2</sub> stream is kept at such an elevated value so that all the pressure loss during the transmission storage process could have been taken care of before injection into the storage site. The life cycle CO<sub>2</sub> emissions for CO<sub>2</sub> compression, transport, and geological storage are estimated to be 12.23 t/year, 1.223 t/ year, and 8.849 t/year, respectively. Finally, the total positive emissions

#### Table 6

Input parameters for life cycle analysis [69].

Parameter	Value	Unit
Wood production harvest and transport	9.9	MJ/MW <sub>Biomass</sub>
Wood processing in pellets plant	573.3	MJ/MW <sub>Biomass</sub>
Handling and storage	3.8	MJ/MW <sub>Biomass</sub>
Ocean transport of wood pellets	0.03	MJ/t km
Wood pellets transport to power plant (by truck)	2.3	MJ/t km
Compression of CO <sub>2</sub>	111	kWh/tCO2
CO <sub>2</sub> storage	7	kWh/tCO2
Fugitive CO <sub>2</sub> emissions from compression process	23.2	t/MW-year
Fugitive CO <sub>2</sub> emissions during transport through pipeline	2.32	t/MW-year
Fugitive $\mathrm{CO}_2$ emissions from the storage site	7.01	kg/t of sequestrated CO <sub>2</sub>

from the system are estimated to be 132.8 t/year, taking into account all the life cycle  $CO_2$  emissions values for the individual stages of the whole life cycle of the proposed system (as shown in Fig. 14). However, as a huge amount of  $CO_2$  is captured from the proposed system, the overall net life cycle emissions from the plant become negative. The overall net life cycle emissions are estimated by the difference between the amount of  $CO_2$  captured and the amount of  $CO_2$  emitted in the whole life cycle process of the proposed system, taking into account the plant growth with respect to the power generation with the CCS process. Fig. 15 shows the life cycle  $CO_2$  emissions are estimated to be - 1129.2 t/year for the system with  $CO_2$  capture.

To do a comparative study with the system without capture, a similar process for the estimation of the life cycle  $CO_2$  emission has been applied

to the system without CO<sub>2</sub> capture. The required stages of the whole life cycle of the system without capture remain the same. Two things are different here in this case during the estimation of lifecycle emission calculation of the plant without CO<sub>2</sub> capture. Firstly, there is no amount of net CO<sub>2</sub> capture here in the whole life cycle of the system without capture. Secondly, the amount of CO2 emitted from the biomass-based power generation system without CO<sub>2</sub> capture is consumed by the wood plants during their growth through the photosynthesis process, fulfilling the criteria of woody biomass as a carbon neutral fuel. By this method, the net life cycle CO<sub>2</sub> emission from the system without capture comes out to be a positive one at 105.9 t/year. Bioenergy with carbon capture and storage (BECCS) presents a special possibility for the simultaneous net removal of atmospheric CO2 and the provision of energy generation. Thus, BECCS stands out as a potential and successful strategy for lowering CO<sub>2</sub> levels in the atmosphere. The direct CO<sub>2</sub> emissions into the atmosphere may be reduced by integrating CCS technology with any traditional power generation process. However, this does not guarantee that the overall net CO<sub>2</sub> emissions from the power generation method in its whole life cycle will be zero, even if the energy source is renewable.

LCA is a potent and widely used system analysis tool for examining the environmental factors and potential effects of a system of goods or services over the course of its life. It is a well-established technique for systematically examining environmental implications throughout the life cycle. LCA studies play a significant role in understanding the net energetic, economic, and environmental effects of power production systems based on renewable or fossil fuels. Hence, a methodical approach for each stage of the proposed plant with CCS technology is executed here to make sure that no positive emission is happening and there is simultaneous removal of  $CO_2$  from the atmosphere. This will



Fig. 14. CO<sub>2</sub> life cycle for the current study.



Fig. 15. Life cycle CO<sub>2</sub> emissions from the systems.

make this proposed carbon negative system a sustainable industrial method for net  $\rm CO_2$  removal from the atmosphere along with positive net power generation in the true sense.

#### 6. Conclusions

In the present study, we propose a first-of-its-kind biomass-based CO<sub>2</sub> negative power generation system combining a steam-biomass gasification facility with a solid oxide fuel cell, a post-combustion carbon capture facility with a molten carbonate fuel cell, and waste heat recovery using the organic Rankine cycle. A techno-economic analysis of two different scenarios, viz., a) with CO<sub>2</sub> capture and b) without CO<sub>2</sub> capture, was investigated in detail. Five different biomass fuels, depending on the availability in the location, have been chosen for the gasification process to generate hydrogen rich syngas. For the overall performance evaluation of the proposed system, the biomass fuel is chosen based on the highest percentage of hydrogen yield in the syngas. Integration of the MCFC and carbon capture facility was able to capture 99.2 % of CO<sub>2</sub>. However, the energy efficiency of the system decreased by 9.43 % with the incorporation of  $CO_2$  capture facility. The exergy efficiencies for the configurations with CO<sub>2</sub> capture and without CO<sub>2</sub> capture are estimated to be 62 % and 70.22 %, respectively. It has been estimated that the proposed systems without CO<sub>2</sub> capture and with CO<sub>2</sub> capture are able to generate electricity at the rates of 0.052 \$/kWh and 0.062 \$/kWh, respectively. Furthermore, a simple payback model suggests that the break-even time for the CO<sub>2</sub> capture configuration is estimated to be 4.2 years. In comparison to other competing CO<sub>2</sub> capture technologies, the proposed system has a competitive LCOE. However, the proposed system might offer much better cost competitiveness once the production costs of the SOFC and MCFC decrease with the gradual improvement of material research in the future. Lastly, the life cycle analysis reveals that the proposed CO<sub>2</sub> negative system is able to capture 1100 tonnes of CO2 per year.

## CRediT authorship contribution statement

**Dibyendu Roy:** Conceptualization, Methodology, Software, Investigation, Writing – original draft. **Samiran Samanta:** Conceptualization, Methodology, Software, Investigation, Writing – original draft. **Sumit**  **Roy:** Conceptualization, Software, Supervision, Writing – review & editing. **Andrew Smallbone:** Supervision, Project administration, Resources, Writing – review & editing. **Anthony Paul Roskilly:** Supervision, Project administration, Funding acquisition.

#### **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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#### D. Roy et al.

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#### D. Roy et al.

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