

Techno-economic-environmental assessment of biomass oxy-gasification staged oxy-combustion for negative emission combined heat and power

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Abstract

Climate change mitigation requires developing low-carbon technologies capable of achieving CO₂ emission reductions at the gigatonne scale and affordable cost. Biomass gasification, coupled with carbon capture and storage, offers a direction to atmospheric CO₂ removal. To compensate for the issues associated with the high-investment requirement of CO₂ removal unit and lower efficiency compared to fossil-based power cycles, this study proposed a conceptual system for combined heat and power, based on biomass oxy-gasification integrated with staged oxy-combustion combined cycle (BOXS-CC). Aspen Plus[®] is used to develop the process model of the proposed cycle. The results obtained in the techno-economic analysis showed that the net power efficiency of the proposed concept with 50.2 kg/s biomass flowrate was 41.6%, and the heat efficiency was 27.4%, leading to a total efficiency of 69.0%, including CO₂ compression. Moreover, the economic assessment of BOXS-CC revealed that it can achieve a levelised cost of electricity of €21.4/MWh, considering

the heat and carbon prices of €46.5/MWh and €40/tCO₂, respectively. Such economic performance is superior compared to fossil fuel power plants without CO₂ capture. The environmental assessment shows that BOX-CC system results in net negative emissions of 766 kg CO₂ eq./MWh_e.

Keywords: Biomass gasification; oxy-combustion cycle; techno-economic analysis; combined heat and power; Greenhouse gas emissions

1. Introduction

Reaching the CO₂ emission reduction targets of the Paris Agreement to limit global warming below 2°C is unlikely to be met since energy consumption is currently growing by 2% per year, and this growth is highly dependent on fossil fuels [1]. Carbon capture and storage (CCS) is a cost-effective solution to limit greenhouse gas emissions [2]. If CCS was excluded from the decarbonisation scenario, the cost of achieving the emission reduction targets would increase by up to 140% due to the considerably higher costs of alternative decarbonisation pathways [3–5].

CCS is often associated with the decarbonisation of fossil fuel-based combustion systems. Yet, it can also be integrated with bio-energy conversion technologies and, similarly to direct air capture technologies, achieve negative CO₂ emissions [6]. Therefore, bio-energy with CCS (BECCS) is currently being regarded as a promising negative-emission technology that can help meet the CO₂ emission reduction targets by 2050 [7,8]. Among the thermochemical options for biomass conversion (direct combustion, torrefaction, pyrolysis, gasification and hydrothermal liquefaction), gasification is one of the most efficient options been widely applied for biomass conversion [9]. Several studies have been conducted on the integration of biomass gasification with CCS. Dinca et al. [10] performed a techno-economic analysis of biomass gasification power plant with amine-based CO₂ capture. The overall efficiency and levelised cost of electricity (LCOE) of the power plant were around 47% and £65/MWh, respectively. Zang et al. [11] performed a techno-economic comparison of different biomass integrated gasification combined cycles. They revealed that the electric efficiencies of the power plants with CCS vary between 23.9–27.1%, while their LCOE is in the range of 188–228 €/MWh. Xiang et al. [12] analysed the effect of the gasification temperature and pressure on the performance of the biomass integrated gasification combined cycle with oxy-fuel combustion. They revealed that the maximum electric efficiency of 35.4% is achievable at gasification temperature and pressure of 1000°C and 3.5 MPa, respectively.

Combined heat and power (CHP) systems achieve higher total efficiencies than the power generation systems by enabling the supply of electric power and hot water for district heating [13,14]. Few works can be found regarding the integration of biomass gasification with CHP. Perna et al. [15] investigated the effect of biomass gasification with CHP. They showed that the electric and total efficiencies for biomass gasification integrated with conventional combined cycle are 17.3% and 69.7%, respectively.

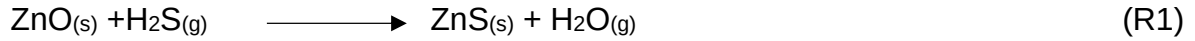
Pérez et al. [16] performed a techno-economic analysis on a lab-scale CHP system integrated with biomass gasification. They revealed that the system's net electric and total efficiencies are 23.4% and 55.8%, respectively, while the LCOE is £16.8/MWh.

Among all different kinds of gasifying agents (O_2 , air, steam, CO_2 , etc.), high-purity O_2 exploitation results in higher heating value syngas [17]. Yet, in this case, the air separation unit (ASU) operation is necessary that increases the capital cost of the plant [18], which is a disadvantage to its commercialisation. The concept of staged combustion has been applied to the natural gas-fired oxy-combustion combined cycle by Khallaghi et al. [19]. Such an oxy-combustion cycle was shown to have a comparable efficiency (51.2%) to the state-of-the-art NG-fired oxy-combustion cycles, such as the Allam cycle (55%) [20]. Importantly, due to its less complex layout and lower volumetric flow rates throughout the system, the levelised cost of electricity for the proposed cycle (£33/MWh) was shown to be competitive to that of the Allam cycle (£50/MWh) [18,21]. This was primarily due to lower capital cost, regardless of the net power efficiency being lower by 3%. Therefore, to compensate for the higher capital cost associated with biomass oxy-gasification, this work aims to evaluate the techno-economic feasibility of a biomass oxy-gasification integrated with a staged oxy-combustion combined cycle (BOXS-CC) for CHP generation with negative CO_2 emissions. Aspen Plus[®] is used to develop the process simulation and then, the effect of split fraction and gasification and combustion temperature on the cycle performance were investigated. Moreover, the results obtained from the optimal thermodynamic performance of the proposed cycle are used for economic feasibility evaluation of the BOXS-CC.

2. Process description and simulation

Figure 1 presents a schematic of the proposed BOXS-CC process. It is assumed that a complete conversion of biomass takes place in the presence of high-purity oxygen, generated in the ASU, as a gasifying agent, in a reactor (Gasifier). Syngas from the gasification consists mainly of H_2 , N_2 , CO , CO_2 , CH_4 , H_2S and H_2O ; solid residues are removed in a separator. The syngas is then cooled down, and the extracted heat is utilised for district heating.

The acceptable level of H₂S is up to 8 ppm at the inlet of a gas turbine [22]. Therefore, H₂S removal with ZnO sorbent (Reaction R1), which is among the widely used methods to reduce the H₂S concentration, is considered for the BOXS-CC process.



The purified syngas is then pressurised to 300 bar for combustion (Figure 1). The rest of the O₂ stream produced in the ASU, which comprises the stoichiometric amount of O₂ required for syngas combustion and 5% excess O₂ to ensure complete combustion, is heated to 200°C before being sent to the first combustion stage. To prevent the excessive temperature in the combustor, the combustion products from the first stage along with the unreacted O₂ are fed to the second combustion stage, where they are acting as a coolant. This process continues until the syngas is entirely consumed in the last combustion stage. Moreover, further cooling at each combustion stage occurs as heat is extracted to pre-heat the CO₂ stream in the supercritical CO₂ (sCO₂) cycle. The exhaust gas is then cooled down by transferring heat for pre-heating the oxygen streams and district heating. The water vapour in the cold exhaust gas is condensed and separated. The remaining stream that contains high purity CO₂ is sent to the CO₂ purification unit (CPU) to be conditioned for storage (Figure 1). In the sCO₂ cycle, as shown in Figure 1, the CO₂ stream is pressurised to 300 bar by pump and compressor. Before entering the sCO₂ turbine, it is heated in a low-temperature recuperator (LTR) and high-temperature recuperator (HTR), as well as the multi-stage combustor.

2.1. Model development

The process model for the BOXS-CC process was developed in Aspen Plus[®]. The equation of state used for thermodynamic property estimation was the Peng Robinson. The biomass and solid residues were defined as the non-conventional components. Biomass enters the yield reactor, which was simulated by using the *RYIELD* block where decomposition occurs to form conventional components such as C, H₂, O₂, H₂O, N₂ and S. Thereafter, components from decomposition along with O₂ entered the gasifier block. The heat required for decomposition was supplied through partial oxidation with the sub-stoichiometric amount of oxygen supplied. The gasifier and all combustion chambers were simulated using the Gibbs reactor (*RGibbs*), which utilises the principle of Gibbs energy minimisation to reach the equilibrium composition under specified conditions [23]. The gasification temperature was set at 750°C [24] by adjusting the O₂ flow rate to the gasifier. The *SSPLIT* block was used after the gasifier for the separation of solid residues from gas. The *Sep* unit was used for the syngas purification. The temperature and pressure of all combustion stages were set at 950°C and 300 bar, respectively. The *MHeatX* block was used with the minimum temperature approach of 5°C to represent the recuperators in the sCO₂ cycle [25,26]. All other heat exchangers, including heater and cooler, were modelled using the *Heater* block. The *Compr* block was used for all turbines and single-stage compressors. The water separator was modelled using *Flash2* block. The stream after condensation was sent to the CPU and then pressurised to 110 bar. The sCO₂ turbine inlet temperature was set at 700°C by implementing the design specification that adjusted the CO₂ mass flow rate in the sCO₂ cycle to arrive at the desired temperature. A cryogenic ASU was simulated using a standard double-column cryogenic unit with the high- and low-pressure columns operated at 5.6 and 1.3 bar, respectively, and filled with 350Y structured packing [27]. The O₂ purity in the ASU was set at 95%, associated with the minimum energy consumption for a cryogenic ASU [28]. It was assumed that cooling water at 15°C is available at the plant site and that heat losses are negligible. The properties of the biomass feedstock are presented in Table 1. Moreover, the key assumptions used to model the turbomachinery and the initial operating parameters for the BOXS-CC are presented in Table 2 and Table 3, respectively.

Table 1. Biomass feedstock composition

Parameter	Value
Biomass composition (wood flour) [29]	
Moisture (wt%, wet)	4.9
Carbon (wt%, dry)	47.0
Hydrogen (wt%, dry)	6.9
Nitrogen (wt%, dry)	3.4
Oxygen (wt%, dry)	42.2
Sulphur (wt%, dry)	0.1
Ash (wt%, dry)	0.4
Proximate analysis	
Volatile matter (wt%, dry)	77.3
Fixed carbon (wt%, dry)	17.7
Ash (wt%, dry)	0.4
Lower heating value (MJ/kg)	15.3

Table 2. Turbomachinery design conditions

Parameter	Value
Isentropic efficiency of pump (%) [30]	90
Isentropic efficiency of the turbine (%) [30]	93
Isentropic efficiency of the compressor (%) [30]	89
Mechanical efficiency of compressors and pump (%) [31]	99.6
Electrical efficiency of the generator (%) [32]	98.5
Combustor and gasifier pressure drop (mbar)	150
Pressure drop in heat exchangers (%)	1

Table 3. Initial BOXS-CC operating parameters

Parameter	Value
Combustor pressure (bar)	300
Combustor temperature (°C)	950
Gasifier temperature (°C)	750
Turbine backpressure (bar)	35
sCO ₂ turbine inlet temperature (°C)	700
sCO ₂ turbine inlet pressure (bar)	300
sCO ₂ turbine backpressure (bar)	75
Recompression split fraction (-)	0.2

2.2. Model validation

The considered BOXS-CC process consists of a syngas gasification section and two power generation sections: (1) the staged oxy-combustion power cycle and (2) the sCO₂ cycle. The power generation sections have been previously validated and investigated extensively by Khallaghi et al. [19,21]. The accuracy of the gasification model used in this study is validated based on the wood chip oxy-gasification reported by Dimitriou et al. [33]. The prediction of the model developed in this study was in good agreement with the literature data, Table 4.

Table 4. Benchmark of the syngas composition

Component (vol%)	Dimitriou et al. [33]	This study
H ₂ O	12.6	12.2
H ₂	28.3	28.5
CO	26	26.4
CO ₂	21.2	22.2
CH ₄	10.5	10.3
N ₂	0.56	0.39
H ₂ S	0.02	0.01

3. Technical assessment

To evaluate the thermodynamic performance of the BOXES-CC, the net power efficiency (η_{net}), heat efficiency (η_{heat}) and total efficiency (η_{total}) of the considered system are defined in Eqs. (1), (2) and (3), respectively. The net power output ($\dot{W}_{net}$) is calculated as the gross power output less the system's parasitic load. The amount of heat available for district heating (Q_{heat}) calculated as the sum of the heat from cooling the syngas and exhaust gas streams. The chemical energy input is defined as the product of the fuel consumption rate ($\dot{m}_{fuel}$) and the lower heating value (LHV).

$$\eta_{net} = \frac{W_{net}}{\dot{m}_{fuel} \cdot LHV} \quad (1)$$

$$\eta_{heat} = \frac{Q_{heat}}{\dot{m}_{fuel} \cdot LHV} \quad (2)$$

$$\eta_{total} = \eta_{net} + \eta_{heat} \quad (3)$$

3.1. Thermodynamic performance

Table 5 presents the thermodynamic assessment based on the initial input parameters from Tables 1-3. It reveals that this concept has a net power output of 283.5 MW and the heat output of 211 MW available for district heating. Such output corresponded to net power and heat efficiencies of 36.9% and 27.4%, respectively, leading to the total efficiency of 64.3%. It can be seen that the sCO₂ compression imposes the main parasitic load on the system, as it accounts for more than 22% of the total thermal energy input in the system.

Table 5. Performance summary of BOXS-CC based on initial operating parameters

Component	Value
Thermal energy input (MW)	768.3
High-pressure turbine (topping cycle) (MW)	60.3
sCO ₂ turbine power output (MW)	559.2
Air separation unit power consumption (MW)	56.9
O ₂ compression power consumption (MW)	35.8
Syngas compression (MW)	74.8
sCO ₂ compression stage power consumption (MW)	173.8
CO ₂ compression power consumption (MW)	4.7
Net power output (MW)	283.5
Net electric efficiency (%)	36.9
Heat output (MW)	211.0
Heat efficiency (%)	27.4
Total efficiency (%)	64.3

To improve the performance of the BOXS-CC system, a sensitivity analysis was performed on the split fraction (SF) in the sCO₂ cycle, gasification temperature and combustion temperature. It is worth noting that the effects of the pressure and temperature in the sCO₂ cycle have been extensively analysed in the literature [19,21]. Therefore, the sensitivity analysis of these parameters is not included in the scope of this work. It needs to be emphasised that when a single recuperator is used in the sCO₂ cycle, a pinch problem occurs due to an imbalance in the specific heat of the sCO₂ streams [25]. This is due to a large variation in the specific heat of CO₂ with temperature and pressure, especially in the vicinity of the critical point [34]. To compensate for this imbalance, the sCO₂ stream is split after the low-temperature recuperator. The diverted CO₂ stream is recompressed without heat rejection before merging the main CO₂ stream between both recuperators, as shown in Figure 1. This split is characterised by the SF defined as the fraction of the total CO₂ flow entering the high-temperature recuperator.

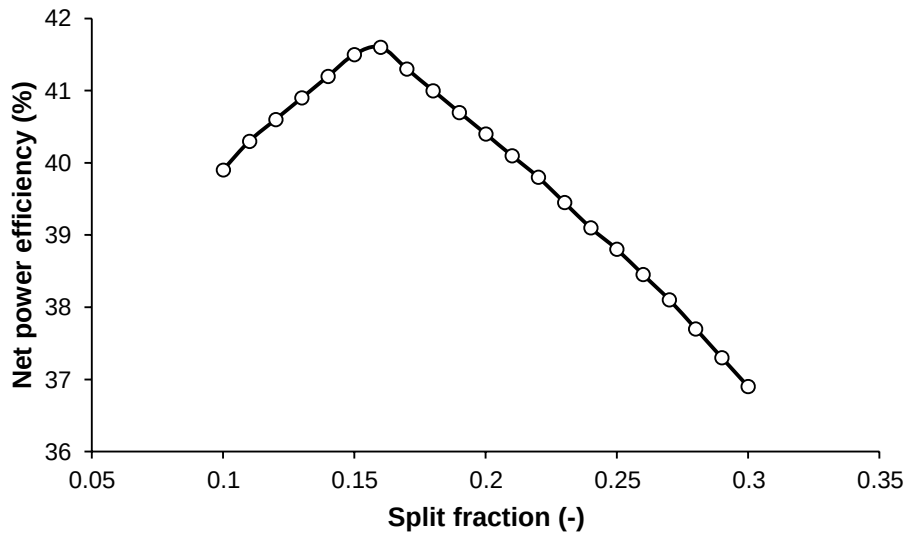


Figure 2. Effect of the split fraction on the net electric efficiency

Analysis of the effect of SF on the thermodynamic performance (Figure 2) has revealed that the value of 0.16 would result in the highest net power efficiency of 41.6%. This is because, at this SF rate, the approach temperature is minimised at both ends of the low-temperature recuperator, maximising the heat transfer rate. Moreover, Figure 3 represents the effect of the gasification temperature on the BOXS-CC performance. The gasification temperature was adjusted by controlling the O₂ mass flow rate, while other parameters, such as the sCO₂ turbine inlet temperature (700°C) and the combustion temperature (950°C) were kept constant. The results revealed an increase in the gasification temperature from 750°C to 900°C, resulting in an increase in the heat efficiency from 27.4% to 28.8%. This was because more heat is available for district heating (Figure 4). On the one hand, the net electric efficiency decreased from 41.6% to 40.1%. This can be explained by the fact that more heat from the syngas stream before compression is transferred to district heating. This balances the drop in the net power output. As a result, the total efficiency of the cycle remained around 69%.

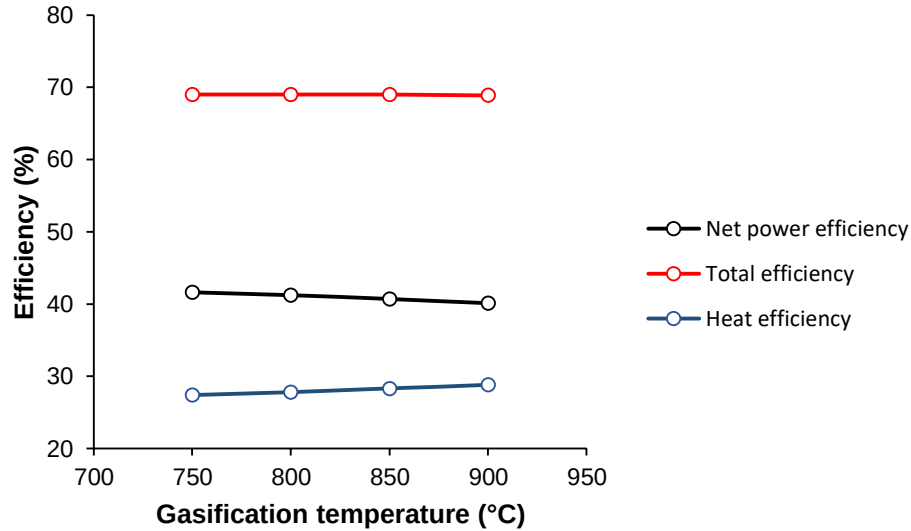


Figure 3. Effect of gasification temperature on the cycle performance

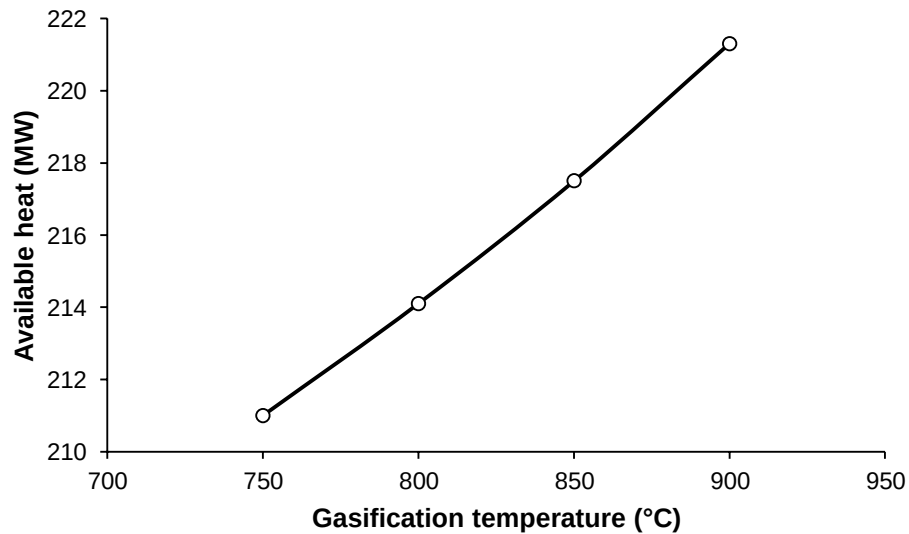


Figure 4. Effect of gasification temperature on heat available for district heating

Furthermore, Figure 5 showed the effect of the combustion temperature on the BOXS-CC performance at the gasification temperature of 750°C. It can be seen that although the net power efficiency drops from 41.6% at 950°C to 39.5% at 1150°C, the total efficiency increases from 69.0% to 70.3%. This was a result of less heat being transferred to the sCO₂ cycle at each combustion stage at higher combustion temperatures. Therefore, to keep the sCO₂ turbine inlet temperature at 700°C, the CO₂ mass flow rate reduced, leading to the subsequent reduction in the power output of the sCO₂ turbine by 6.2% (35 MW) (Figure 6). On the contrary, due to a higher inlet

temperature, the power output of the high-pressure turbine was increased by 17.2% (10.4 MW). This figure was less than the drop in the power output of the sCO₂ turbine, leading to a reduction in the net power efficiency. Nevertheless, an increase in the high-pressure turbine inlet temperature resulted in more heat for district heating (Figure 7), increasing the total efficiency.

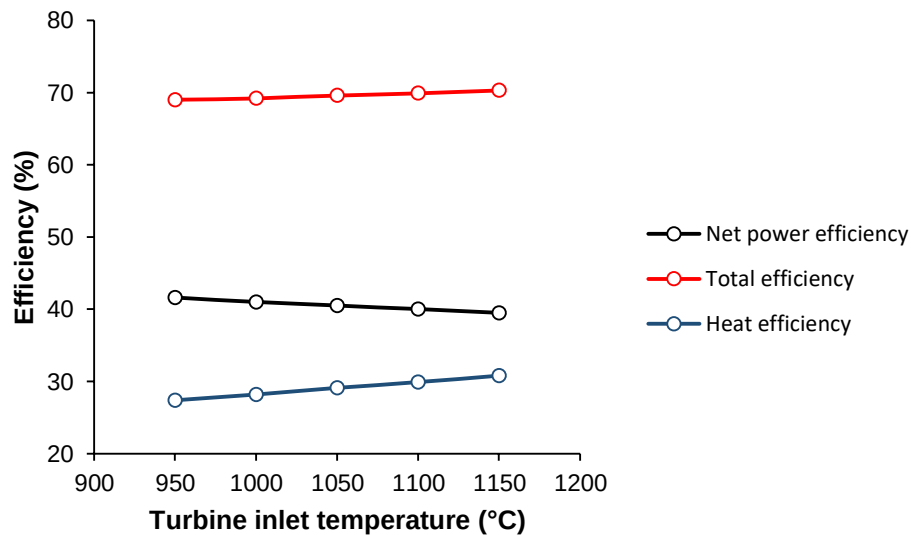


Figure 5. Effect of the combustion temperature on the BOXS-CC performance

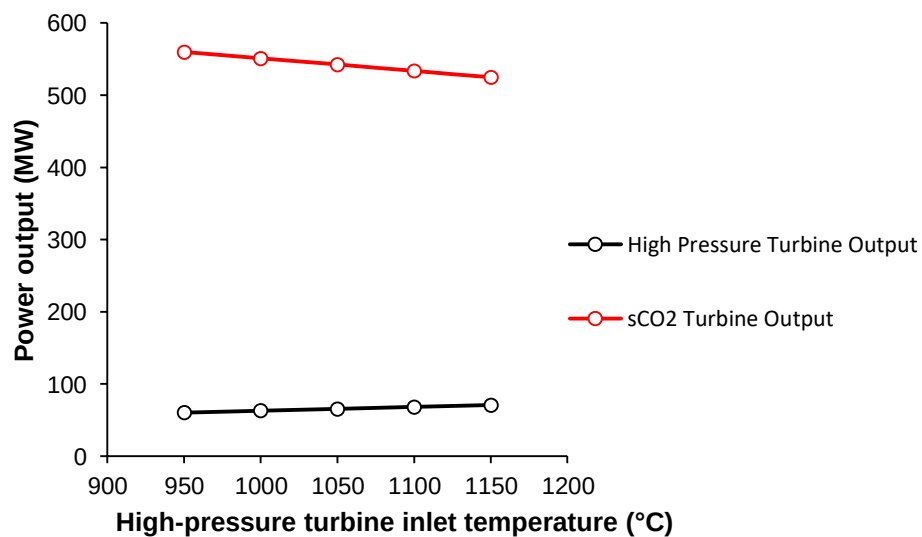


Figure 6. Effect of the combustion temperature on power output

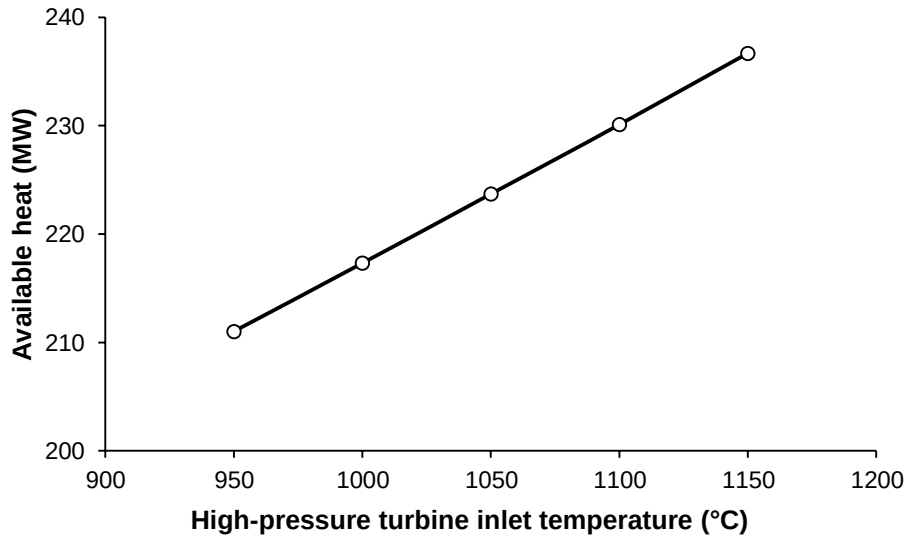


Figure 7. Effect of the combustion temperature on heat available for district heating

However, it needs to be highlighted that without EGR being used to lower the combustion temperature in the BOXS-CC, working under high temperature (1150°C) is intolerable. Moreover, the results indicated that the difference in the total efficiency between 950°C and 1150°C at 300 bar is only 1.3%. Therefore, the thermodynamic performance of the BOXS-CC under 300 bar and 950°C is summarised in Table 6. The analysis shows that the proposed concept has a net power output of 319.6 MW with net power and total efficiencies of 41.6% and 69.0%, respectively.

Table 6. Thermodynamic performance of BOXS-CC based on revised operating parameters

Component	Value
Thermal energy input (MW)	768.3
High-pressure turbine (topping cycle) (MW)	60.3
sCO ₂ turbine power output (MW)	559.6
Air separation unit power consumption (MW)	56.9
O ₂ compression power consumption (MW)	25.8
Syngas compression (MW)	74.8
sCO ₂ compression stage power consumption (MW)	138.3
CO ₂ purification and compression power consumption (MW)	4.5
Net power output (MW)	319.6
Net power efficiency (%)	41.6
Heat output (MW)	211.0
Heat efficiency (%)	27.4
Total efficiency (%)	69.0

Compared to the state-of-the-art NGCC with a net efficiency above 62% [35] and an NGCC with CO₂ capture, which results in a net efficiency penalty of more than 8% [36], the BOXS-CC is characterised by a lower net power efficiency (41.6%). However, the lower power efficiency of the BOXS-CC is compensated by the generation of heat for district heating, which results in a total efficiency of 69.0%.

4. Economic assessment

The economic evaluation of the BOXS-CC was performed using the levelised cost of electricity (*LCOE*), Eq. (4). To do so, the total annualised cost (*TAC*) that comprise the annualised total plant cost (*TPC*), the fuel cost (*FC*), the carbon revenue (*C_{carbon}*), variable (*V_{O&M}*) and fixed (*F_{O&M}*) operating and maintenance costs, Eq. (5).

$$LCOE = \frac{TAC}{8760 \times W_{net} \times CF} \quad (4)$$

$$TAC \left[\frac{M\text{€}}{y} \right] = TPC \times FCF + FC + C_{carbon} + V_{O\&M} + F_{O\&M} \quad (5)$$

TPC equals the sum of direct capital cost (purchase cost), indirect capital cost. Indirect Capital Cost (ICC) comprises project contingency and the owner's cost. The annualised *TPC* equals the *TPC* multiply by the fixed charge factor (*FCF*), which is defined using Eq. (6), considering the project interest rate (*r*) and project lifetime (*T*).

$$FCF = \frac{r(1+r)^T}{(1+r)^T - 1} \quad (6)$$

1. The assumptions used in the economic analysis are summarised in Table 7.

Table 7. Assumptions for the economic analysis

Parameter	Value
Indirect capital cost (<i>ICC</i>) as a fraction of total purchase cost (%) [37]	45
Variable operating cost (<i>VOM</i>) as a fraction of total capital cost (%) [31]	2.0
Fixed operating cost (<i>FOM</i>) as a fraction of total capital cost (%) [31]	1.0
Wood flour price (€/t _{dry}) [38]	68.5
Plant lifetime (<i>T</i>) (years) [31]	25
Project interest rate (<i>r</i>) (%) [31]	8.75

The equipment purchase costs (C_B) for gasifier, sulphur removal, CPU, gas turbine, heat exchanger and compressor were calculated based on reference cost data from the literature (Table 8) using Eq. (7) where C_A is the cost of the reference component with the capacity of Q_A and f is the scaling factor. It is worth noting that although a simple separator is considered for the BOXES-CC modelling, an entire sulphur removal plant with a recovery section is considered for the cost analysis.

$$C_B = C_A \left(\frac{Q_B}{Q_A} \right)^f \quad (7)$$

Table 8. Scaling parameters for the component purchase cost

Component	Scaling factor	C_A (M€)	Q_A	f	Ref.
Gasification reactor	Mass flow (t/h)	48.53	54.4	0.75	[11]
Sulphur removal and recovery	Mass flow (t/h)	43.33	452.6	0.67	[39]
CPU	Mass flow (t/h)	55.98	266.6	0.6	[40]
Compressor	Power (kW)	0.44	413	0.68	[40]
Gas turbine	Power (MW)	49.4	272.1	0.3	[41]
Heat exchanger	Heat transfer (MW)	6.1	828	0.67	[42]

The correlations used to estimate the capital costs of other components are summarised in Table 9. The following correlations relate the purchasing cost of each component (C_i), to their operating parameters such as mass flow rate ($\dot{m}_i$), pressure ratio (β), isentropic efficiency (η_i), turbine inlet temperature (T_T).

Table 9. Capital cost estimation correlation

Equipment [scaling parameter]	Correlation
sCO ₂ Turbine [mass flow at the inlet to the turbine $\dot{m}_T$ (kg/s), Isentropic efficiency η_T (-), Pressure ratio β (-), Turbine inlet temperature $T_{T,i}$ (C)] [43]	$C_{Turbine} = \dot{m}_T \cdot \beta \cdot \left(\frac{392.2}{1 - \eta_T} \right) \cdot \ln(\beta) (1 + \exp(0.036 T_{T,i} - 65.66))$
Air separation unit [O ₂ production rate, $\dot{m}_{O_2}$ (kg/s)] [44]	$C_{ASU} = 2.926e^7 \left(\frac{\dot{m}_{O_2}}{28.9} \right)^{0.7}$
Generator [Break power output, $W_{T,BRK}$ (kW)] [45]	$C_{Gen} = 26.18 (W_{T,BRK})^{0.95}$

4.1. Economic performance

The economic feasibility of the proposed concept was evaluated by considering the LCOE at different heat and carbon trading prices. The breakdown of the capital cost for the considered case is shown in Table 10. The total plant cost of the BOXS-CC was estimated to be €585.4M. It has been shown that the gasifier has the highest contribution to the total capital cost.

Table 10. Economic performance of the BOXS-CC

	Unit	Value
ASU	M€	54.9
CPU	M€	59.3
Gasification reactor	M€	119
Gas turbine	M€	28.8
sCO ₂ turbine	M€	37.8
Compressors	M€	24.2
Generators	M€	6.9
Heat exchangers	M€	30.5
Total equipment cost	M€	391.6
Total plant cost (TPC)	M€	585.4
Annualised TPC	M€/y	52.8
Fuel cost	M€/y	86.4
Fixed O&M	M€/y	5.8
Variable O&M	M€/y	11.7
Total annualised cost	M€/y	156.7
LCOE	€/MWh	70

Table 10 reveals that when there is no revenue from negative CO₂ emission, the LCOE (€70/MWh) is higher than the electricity prices reported for fossil fuel power plants without CO₂ capture (€32.5/MWh–€63.8/MWh) [31]. However, Figure 8 shows that the LCOE of the BOXS-CC was highly sensitive to changes in the carbon price. At the carbon price of €40/tCO₂, the LCOE is around €33/MWh, which is in the range of the conventional NG-fuelled power cycles without CO₂ capture. Considering the carbon price of €120/tCO₂ in some European countries such as Sweden in 2017 [46], the carbon price of €40/tCO₂ is reasonable.

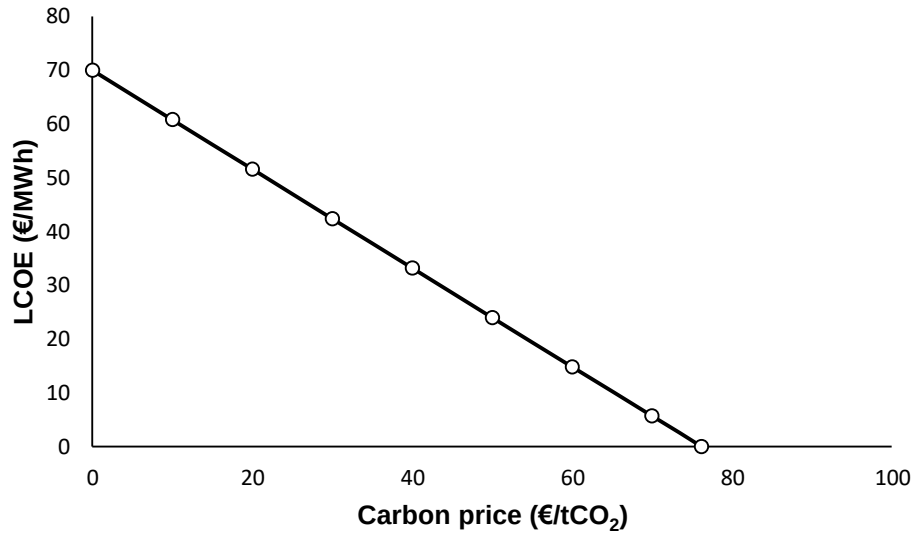


Figure 8. Effect of the carbon price on the economic performance of the BOXS-CC

The effect of the heat price on the LCOE at different values of the carbon price is presented in Figure 9. Moreover, some technical constraints such as energy resource availability or economic reasons may result in different capacity factor. Therefore, the effect of a different capacity factor on the LCOE is presented in Figure 10 while the heat and carbon prices are kept at €46.5/MWh [47] and €40/tCO₂, respectively.

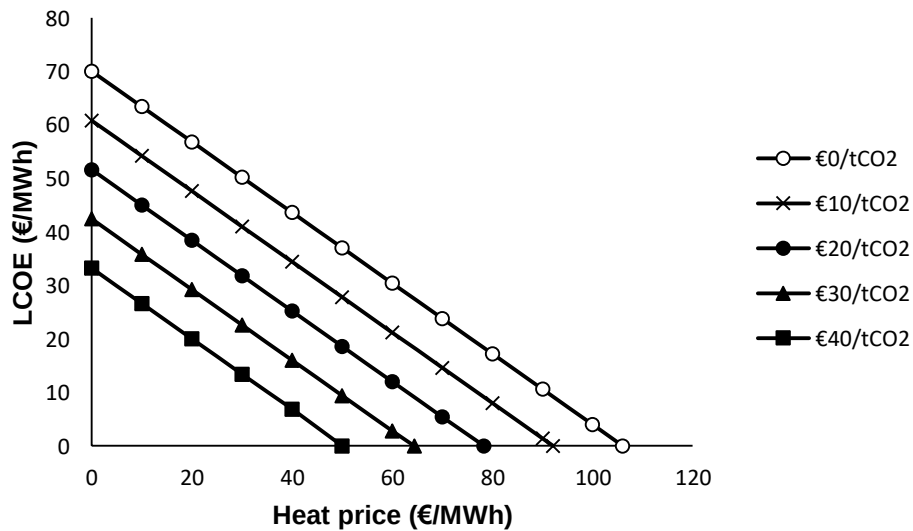


Figure 9. Effect of heat price on the economic performance of the BOXS-CC.

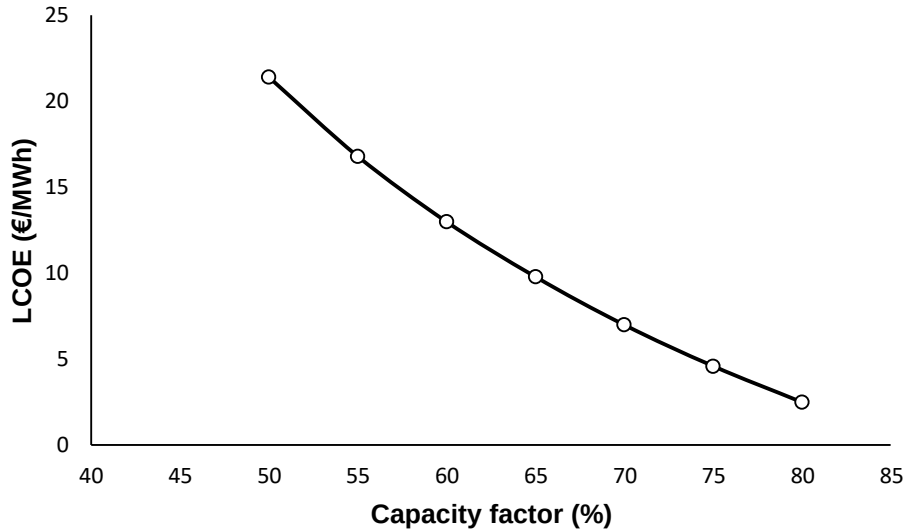


Figure 10. Effect of the capacity factor on the economic performance of the BOXS-CC

It has been shown that the economic performance of the BOXS-CC is susceptible to changes in both heat price (Figure 9) and capacity factor (Figure 10). Figure 10 reveals that at a capacity factor of 50%, the LCOE would be €21.4/MWh. This LCOE is superior to other fossil fuel power plants with CO₂ capture plants (€45/MWh-€90.5/MWh) [31]. Therefore, further development of the BOXS-CC would contribute to the decarbonisation of the power sector, and even negative emissions can be achieved at an affordable cost.

5. Environmental assessment

The environmental assessment focuses on quantifying the net greenhouse gas (GHG) emissions associated with BOXS-CC system. The assessment has been carried out following the life cycle assessment (LCA) methodology as per the ISO 14040 [48] and 14044 standards [49]. The global warming potential (GWP) impact is estimated as per IPCC -AR5 [50]. The plant is assumed to be situated in the UK. The scope of the study is from the production of feedstock to the compression of CO₂ ready for storage. The system boundary includes the construction of the BOXS-CC plant, production of sawdust, its transport, production of electricity and heat in BOXS-CC plant, treatment of H₂S, and carbon capture for storage. It is assumed that the feedstock (sawdust) for the BOXS plant is imported from Canada. Ash and ZnS waste are assumed to be landfilled. The functional unit is defined as the 'generation of 1 MWh of electricity'.

Table 11 provides a summary of input and output data for the system. The background LCA data for the feedstock, equipment, transport, waste disposal and other relevant processes have been sourced from Ecoinvent v3.7 [51]. The GHG impact of manufacturing the plant components (i.e. gasifier, gas turbine, and compressor) has been estimated by upscaling the Ecoinvent data using the scaling factor (f) in Table 8. Since the plant also provides heat as a co-product, exergy allocation has been used to allocate the impact between electricity and heat.

Table 11. Life cycle inventory data for the BOXS-CC system

Data	Value
Feedstock (kg/MWh_e)	658
Sea transport (Canada to the UK) ($t.km/MWh_e$)	4260
Road transport (Canada) ($t.km/MWh_e$)	132
Road transport (UK) ($t.km/MWh_e$)	132
Catalyst (ZnO) (kg/MWh_e)	0.4
Waste (ash and catalyst) (kg/MWh_e)	3.1
Wastewater (ash and catalyst) (m^3/MWh_e)	0.3
Heat (co-product) (MWh_{th}/MWh_e)	0.66
Exergy allocation factor (heat) (%)	10
CO ₂ captured (before allocation for heat) (kg/MWh_e)	921

5.1. Environmental performance

As shown in Figure 11, the BOXS-CC system results in net negative emissions of 766 kg CO₂ eq./MWh_e. The background emissions associated with the supply chain and the plant components are estimated at 62.5 kg CO₂ eq./MWh_e. The main contributor to the background GHG emissions is transporting, which contributes 90%, followed by the production of feedstock and other inputs, which accounts for 8%. The contributions of plant construction and waste treatment and disposal are very small.

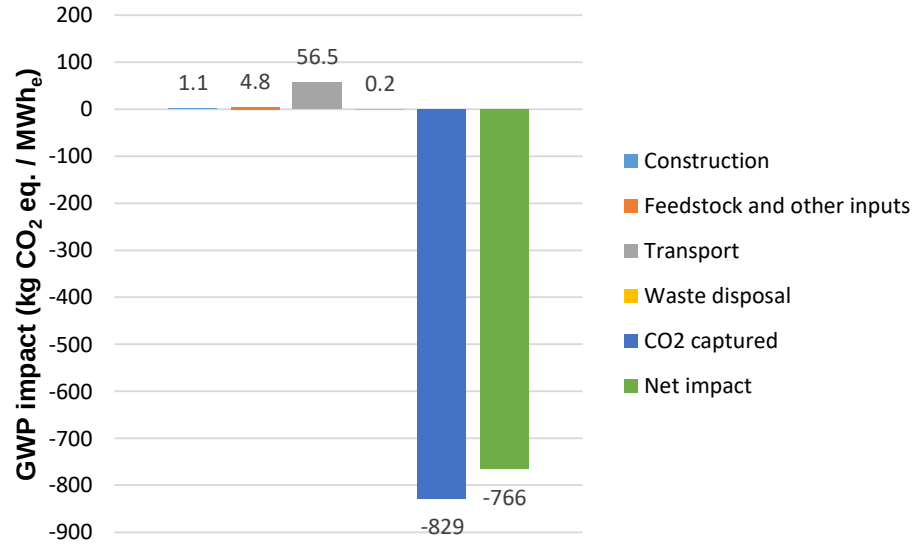


Figure 11. Greenhouse gas emission for generation of 1 MWh of electricity from the BOXS-CC (after 10% allocation for heat)

6. Conclusions

This study presented a CO₂-negative emission concept of biomass oxy-gasification integrated with the combined heat and power system based on staged syngas oxy-combustion combined cycle (BOXS-CC). The techno-economic feasibility analysis revealed that the integration of staged-combustion concept with biomass gasification has the potential to compensate the energy penalty associate with biomass oxy-gasification. The results revealed that the proposed cycle was characterised with net power efficiency and total efficiency of 41.6% % and 69.0%, respectively. The economic assessment revealed that at the capacity factor of 50%, the levelised cost of electricity (LCOE) was €21.4/MWh, considering a heat price of €46.5/MWh and carbon price of €40/tCO₂. Such economic performance was competitive compared to fossil fuel power plants with and without CO₂ capture. The environmental assessment shows that BOX-CC system results in net negative emissions of 766 kg CO₂ eq./MWh_e. Overall, the BOXS-CC was shown to have the potential to contribute to significant cost reduction in decarbonising the power and heat sector while achieving negative CO₂ emissions to offset the CO₂ emissions from other sources.

List of Abbreviations

BOXS-CC	Biomass oxy-gasification staged combustion combined cycle
CCS	Carbon capture and storage

BECCS	Bio-energy with carbon capture and storage
CHP	Combined heat and power
NG	Natural gas
SOF-NGCC	Staged oxy-fuel natural gas combined cycle
ASU	Air separation unit
NGCC	Natural gas combined cycle
LTR	Low-temperature recuperator
HTR	High-temperature recuperator
sCO ₂	Supercritical CO ₂
CPU	Carbon purification unit
LHV	Lower heating value
TIT	Turbine inlet temperature
LCOE	Levelised cost of electricity
SOFC	Solid oxide fuel cell

Nomenclature

A_{HE}	Heat exchanger surface area [m ²]
C_j	Capital cost of equipment j [£]
CF	Capacity factor [-]
FCF	Fixed charge factor [-]
FOM	Fixed operating and maintenance cost [£]
k	Correction coefficient
LCOE	Levelised cost of electricity [£/MWh]
LHV	Lower heating value [kJ/kg]
$\dot{m}_{fuel}$	Fuel consumption rate [kg/s]
$\dot{m}_{O_2}$	O ₂ production rate in air separation unit [kg/s]
SFC	Specific fuel cost [£/MWh]
TCR	Total capital requirement [£]
T_{Ti}	Turbine inlet temperature [°C]
VOM	Variable operating and maintenance cost [£/MWh]
$\dot{W}_{j,BRK}$	Break power output/requirement of equipment j [MW]
W_{net}	Net power output of the entire system [MW]
η_{heat}	Heat efficiency
η_j	Isentropic efficiency of equipment j [%]
η_{net}	Net efficiency [%]
η_{total}	Total efficiency
β	Pressure ratio [-]

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Techno-economic-environmental assessment of biomass oxy-gasification staged oxy-combustion for negative emission combined heat and power

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