Techno-economic evaluation of biomass-to-methanol production via circulating fluidized bed gasifier and solid oxide electrolysis cells: A comparative study

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ABSTRACT

Methanol is considered a promising solution for decarbonizing the transportation and chemical industry sectors, being a worldwide traded commodity that can be synthesized from biomass, renewable electricity, CO_2 and other carbon-rich gases. This study investigates the potential of Solid Oxide Electrolysis Cells (SOEC) in enhancing the performance of bio-methanol production from biomass gasification. The research explores three distinct biomass-to-methanol plant configurations, incorporating an oxygen-blown Circulating Fluidized Bed Gasifier (CFBG) and different SOEC systems, namely: (i) steam electrolysis for hydrogen generation, (ii) co-electrolysis of steam and CO_2 separated from syngas and (iii) direct supply of purified bio-syngas to the SOEC. The study reveals that, although the choice of SOEC type and system configuration could impact energy conversion efficiency and carbon efficiency, all plants show similar performance. In terms of Levelized Cost of Fuel (LCOF) and total efficiency, the syngas-electrolysis configuration exhibits the lowest LCOF, 21.56 €/GJ, and comparable total efficiency of around 80 % to the steam-electrolysis configuration. On the other hand, the CO_2-H_2O-electrolysis configuration showed the highest LCOF due to higher electricity consumption and capital investment.

1. Introduction

Biomolecules are increasingly considered a viable alternative to fossil fuels in hard-to-electrify transportation sectors, such as aviation, maritime and heavy-duty trucking, and in the chemical industry. Importantly, methanol serves today as a crucial commodity in the chemical sector, predominantly used in producing other chemical substances like formaldehyde, acetic acid, and various types of plastics [1] and may become a building block of a future sustainable chemical industry [2]. Thermochemical conversion of biomass (BtX) is an efficient method for biofuel production. However, biomass resources are limited and insufficient to meet all the transportation, and chemical industry demands sustainably. With the global energy demand growing, finding efficient and sustainable methods to produce biofuels and exploit most of the biogenic carbon from this limited resource is of the utmost importance. To make the most of biomass resources, renewable electricity from wind and solar sources can be used [3] to enhance biomass-derived fuels in "Power and Biomass-to-X" (PBtX) systems, where green

hydrogen is supplied to remove excess oxygen from the syngas, increase the carbon efficiency of the process [4] and ultimately increase the production per unit of biomass [4–8].

Previous studies showed that for PBtX processes to be competitive, the electrolysis plant need to operate with high-capacity factors, possibly switching-off or reducing the load in periods with high electricity prices [9]. Also, the cost of the bio/e-fuel produced with PBtX processes is highly sensitive to the electricity price [5,6,9–12]. These two facts, make the use of high temperature solid-oxide electrolysis cells (SOEC) particularly interesting in PBtX plants, thanks to their high energy efficiency (i.e. low electricity consumption per unit of hydrogen produced), despite the high Capex compared to more mature low temperature electrolysis. Thanks to their high efficiency, SOEC technology and manufacturing capacity are expected to scale-up and deliver plants with hydrogen production capacity of the order of 10,000 Nm^3/h within a decade [13].

Three PBtX process configurations based on SOEC are assessed in paper based on the conceptual block diagrams in Fig. 1. In the first integration approach (depicted in Fig. 1 (a)), a steam-fed electrolyzer is

utilized to generate hydrogen. This hydrogen is then blended with the syngas, resulting in the necessary syngas composition required for synthesis of the bio-product. In the second integration strategy, a stream of clean syngas is fed to the electrolysis system, where the necessary CO and H$_2$ are produced, creating the appropriate syngas composition. Another integration strategy involves a system where only CO$_2$ and H$_2$O are introduced into the electrolyzer to produce conditioned syngas (Fig. 1 (c)).

The literature studies mostly examined the integration of steam electrolysis (Fig. 1a) in PBtX plants.

Sun et al. [14] conducted a study on integrating municipal solid waste gasification and solar electrical energy to produce Methanol. They compared three system configurations: municipal solid waste-to-methanol with carbon capture and storage, municipal solid waste-to-methanol with a SOEC unit, and municipal solid waste incineration for power generation. The results showed that the municipal solid waste-to-methanol with a SOEC configuration had the highest efficiency (61 % exergy efficiency).

Similarly, Zhang et al. [15] optimized a biomass-to-Methanol process coupled with SOEC. First, they sized the electrolyzer based on two concepts: (1) full conversion of carbon in biomass and (2) zero power exchange. A sharp trade-off between the methanol production cost rate and the system efficiency in the PBtX system was found. For example, although the case of full conversion of carbon showed higher energy efficiency (64.5 % vs. 53 %), it also resulted in significantly higher methanol production cost (540 €/t vs. 383 €/t).

Butera et al. [16] considered a biomass to methanol process based on a low-temperature circulating fluid bed gasifier. Five scenarios were considered, differing in the gasifying and oxidation agents (air, O$_2$/CO$_2$ or O$_2$/H$_2$O), methods for setting H$_2$ to CO ratio (SOEC or water gas shift reactor), and pure oxygen supply (air separation unit or SOEC). According to the results, SOEC-based systems outperformed systems using water–gas-shift (WGS) reactor to adjust H$_2$ to CO ratio. The SOEC-based system using an O$_2$/H$_2$O mixture as the gasifying agent achieved a higher energy efficiency of 53.8 %, compared to the WGS-based scenario (with the same gasifying agent), which yielded an energy efficiency of 42.6 %. For the former system the carbon efficiency was calculated 45.5 % versus carbon efficiency of 32.2 % for the latter case.

Direct electrolysis of syngas in a power and biomass to MeOH plant (Fig. 1b) was only assessed by Butera et al. [17] who considered a plant
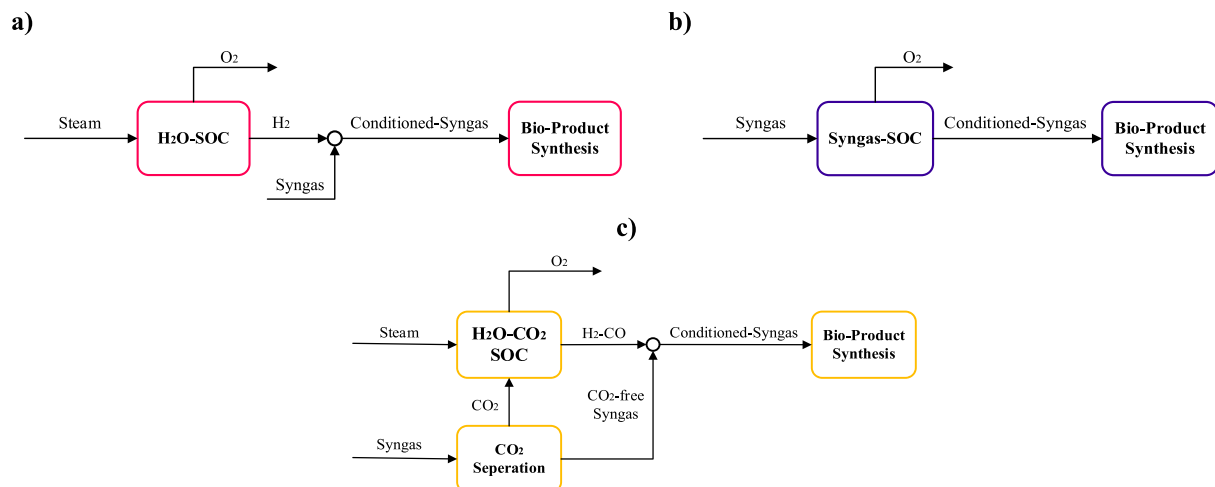

Fig. 1. Different ways of integrating SOEC in BtX plant: (a) steam-feed, (b) syngas-feed, (c) CO$_2$-H$_2$O-feed.

based on a Two-Stage Electro-gasifier in which the SOEC is fed with gas from a pyrolysis reactor. Also, CO_2 from the downstream acid gas removal unit is recirculated before the SOEC to increase the overall carbon conversion and methanol production. The economic viability of the proposed plant was taken into account by Butera et al. [10].

In another study, Pozzo et al. [18] designed a plant coupling DME synthesis unit with a biomass gasification process and a high-temperature co-electrolysis unit fed with syngas from a biomass gasification/pyrolysis process. It was found that the proposed plant yielded twice as much DME as the conventional BtX plant.

The third option to integrate a co-electrolysis unit in a PBtX plant is to feed the electrolyzer with steam and separated CO_2 (Fig. 1c).

Upon an evaluation of existing literature, it becomes evident that the PBtX plant consistently outperforms the BtX plant in key performance areas, specifically with regard to carbon efficiency. The differences in the obtained carbon efficiencies are largely dependent on the system configuration, in particular, the specific type of gasifiers utilized and the chosen strategies for sizing the SOEC into the BtX plant. For instance, decisions around achieving complete carbon conversion [18], supplying the O_2 to the gasifier [19], establishing zero net power exchange [15], converting only a portion of captured CO_2 [17], all significantly impact the carbon efficiency, which ranges between 40 %, for the case in which the electrolyzer is sized to produce the required O_2 for the gasifier, and 97 %, when the electrolysis unit is sized to achieve full carbon conversion. As for the cost of the product, in the examined literature, it varies in a rather wide range, between 20.16 €/GJ and 40 €/GJ.

Even though different power and biomass to methanol process integrated with SOEC units have been assessed in the literature, there is a gap on comparing from techno-economically perspective the different integration options with consistent methodology. Therefore, the novelty of this study is:

- Investigation of SOEC integration in a biomass-to-methanol plant, based on an oxygen-blown Circulating Fluidized Bed Gasifier (CFBG). The main reason for selecting a CFBG technology is that it is a mature gasification process, suitable for medium-large scale plants and can achieve the highest carbon efficiency by retaining all the biogenic carbon in the syngas [12]. In the previous studies, entrained flow gasification or TwoStage gasifier were commonly coupled with SOEC.

- Conducting a techno-economic evaluation of distinct system configurations, encompassing steam-fed, CO_2-steam-fed, and syngas-fed SOEC applications, with consistent methodology, marking the first comprehensive comparative analysis of its kind.

- Placing specific emphasis on essential aspects such as oxygen requirements, methane reformation, steam demands, and heat integration systems within the various system configurations which usually were simplified in the similar studies.

Furthermore, a comparative analysis is conducted between the proposed system and a similar PBtX plant coupled with a low-temperature electrolyzer, highlighting their respective advantages and disadvantages.

The results can be utilized to guide the selection of the most suitable SOEC type and system configuration for biofuel production, leading to enhanced carbon efficiency and biogenic resource utilization.

2. Plant description

The diagrams in Figs. 2–4 display the block diagrams for three Power and Biomass to Methanol plants analyzed and compared in this paper. The plants are categorized based on the type of SOEC: Steam-Electrolysis (StE), Syngas-Electrolysis (SyE), and Steam-CO_2-Electrolysis (StCE). The role of SOEC devices in all plants is to convert most of the biogenic carbon from CO and CO_2 molecules into biomethanol, by achieving the proper syngas module $M = (H_2 - CO_2)/(CO + CO_2)$ without any carbon species removal.

All plant configurations incorporate identical essential conversion stages, namely biomass dehydration, gasification, syngas cleaning, SOEC, syngas conditioning and compression, methanol synthesis and purification, and MeOH loop tail gas combustion in an internal combustion engine.

The plant is assumed to be fed with 100 MW_{LHV} of woodchips with 45 % initial moisture. The as-received biomass is dried in a belt dryer to decrease its moisture content, resulting in dried biomass sent to the gasification island, based on a pressurized circulating fluidized bed generating syngas at 4 bar and 870 °C. The scalability of a CFBG gasifier to a capacity in the hundreds of megawatts is widely acknowledged, making it the favored option for large-scale biomass and waste combustors. After the gasification process, most carbon input remains in the
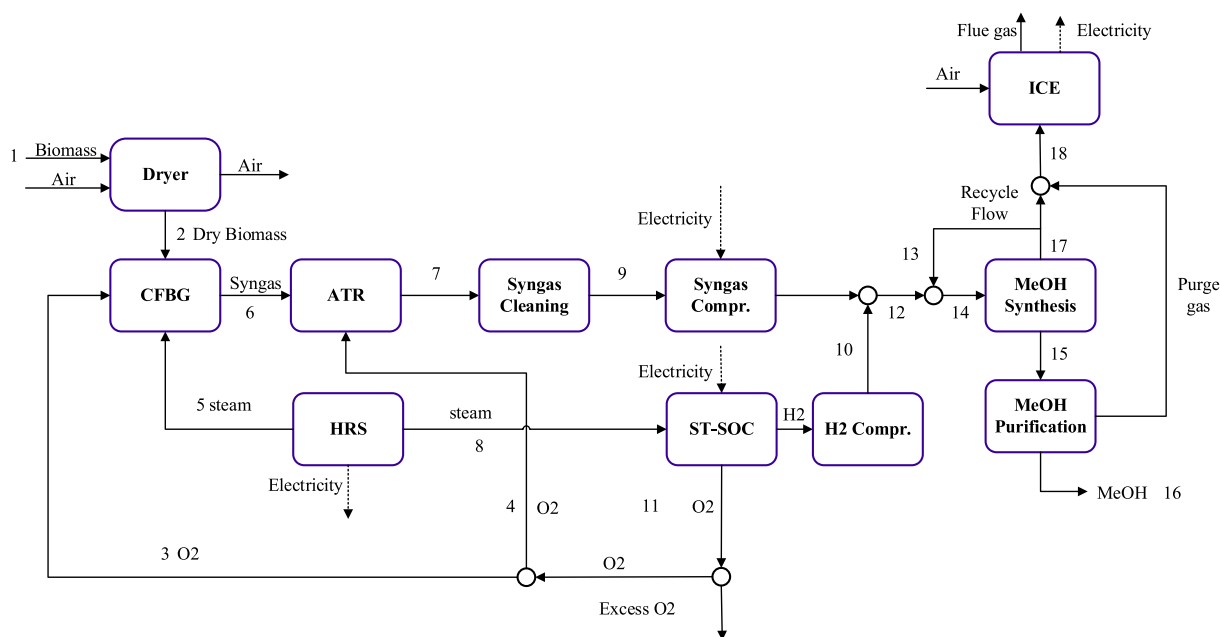

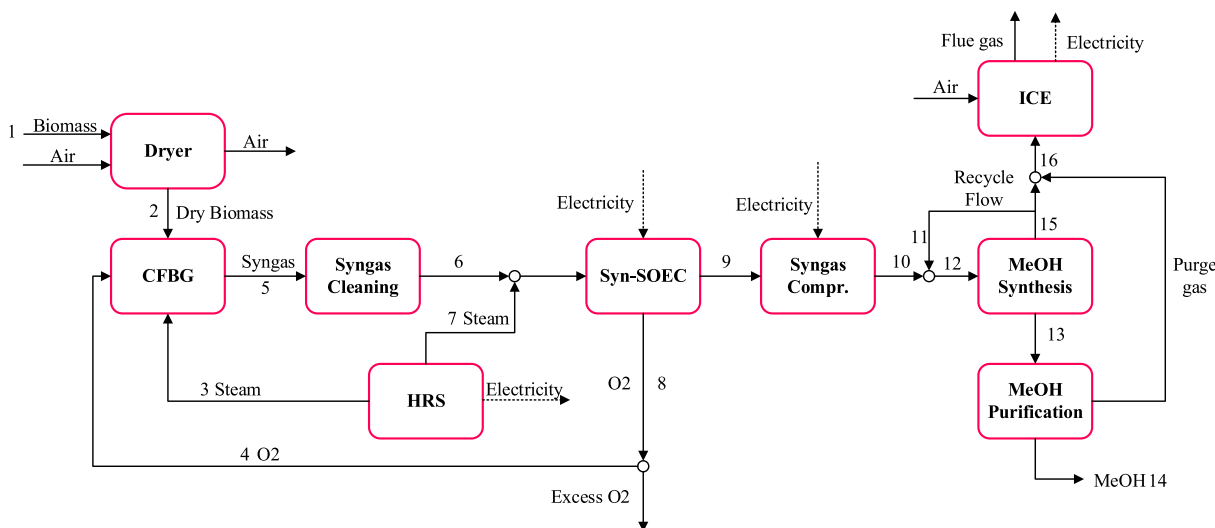**Fig. 2.** Block diagram of the Steam-Electrolysis (StE) plant.

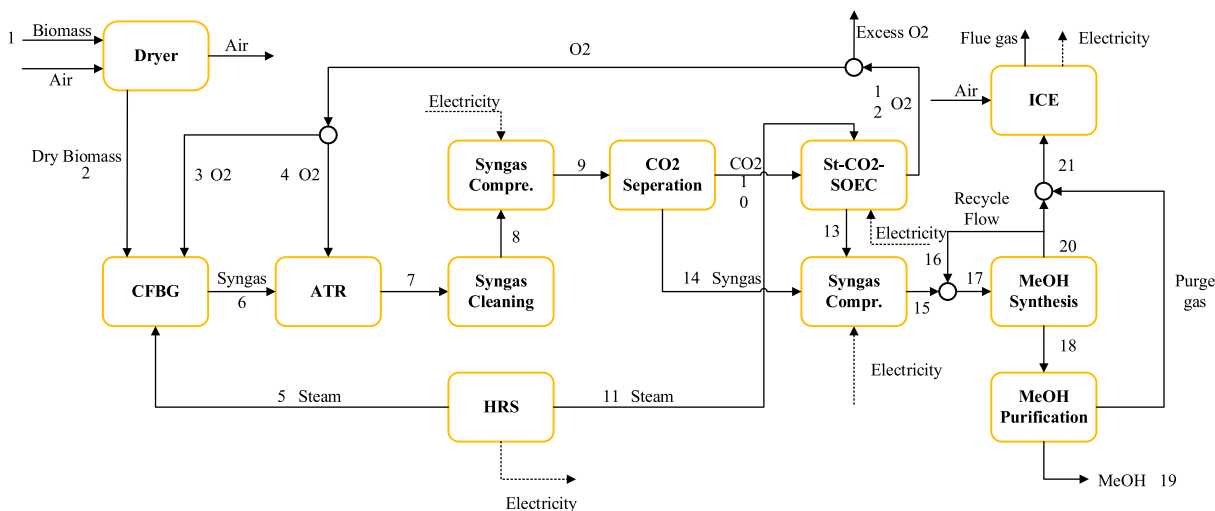**Fig. 3.** Block diagram of the Syngas-Electrolysis (SyE) plant.

Fig. 4. Block diagram of the Steam-CO$_2$ Electrolysis (StCE) plant.

syngas as CO, CO$_2$, CH$_4$, and higher hydrocarbons and a minor amount of tar, while a small portion is extracted as unconverted char from the fluidized bed.

The StE and StCE plant include a catalytic auto-thermal reformer unit downstream the gasifier and high temperature filtration to convert methane and tar into useful reactants for MeOH synthesis (i.e. CO and H$_2$) [20,21]. Oxygen from electrolysis is supplied to the reformer to reach a temperature of 915 °C. In the SyE plant, the reformer is not utilized because the methane produced undergoes steam-reforming reactions within the SOEC.

The syngas must undergo further purification, conditioning, and compression to be fed to the methanol synthesis process. First, the syngas is cleaned to remove contaminants such as sulfur, chlorine, and ammonia, that would poison the SOEC and catalytic reactor downstream. So, in the syngas cleaning island, a water scrubber, a liquid Redox "LO-CAT" unit, and activated carbon beds are employed for all plants.

In the StE plant (Fig. 2), the desired syngas mixture is achieved by adding hydrogen generated by a high-temperature solid oxide steam-electrolysis device to the syngas. In the SyE plant (Fig. 3), a mix of steam and preheated syngas at 750 °C is sent to a solid oxide syngas-electrolysis device to co-electrolyze CO$_2$ and H$_2$O to produce H$_2$ and

CO, resulting in the desired syngas mixture for the biofuel synthesis. In the StCE plant (Fig. 4), the SOEC is fed with steam and CO$_2$ separated by a MDEA-based absorption process [22].

In all plants, the conditioned syngas is pressurized to 90 bar at the methanol synthesis section inlet. This is done using a five-stage inter-cooled compressor. In the Steam-Electrolysis plant, also, an additional intercooled compressor is utilized to increase the pressure of the hydrogen, which is then mixed with the syngas before it reaches the methanol synthesis section.

Methanol synthesis is based on a conventional boiling water reactor (BWR). The syngas flows through tubes containing catalysts and is surrounded by boiling water at 238 °C. The per-pass methanol yield is restricted by thermodynamic equilibrium, so most of the unconverted reactants are recycled back to the reactor. The crude Methanol is cooled to 40 °C and separated from the light gases in a flash unit before being throttled to 2 bar for purification. The purification process involves distillation columns that remove the light gases from the crude Methanol and separate water from Methanol to achieve a final purity of 99.85 % wt.

In all plant configurations, a cogenerative internal combustion engine (ICE) is utilized to generate electricity from the purge gases from the methanol synthesis loop and the purification units. Furthermore,

high-pressure steam generated from syngas and MeOH reactor cooling is expanded in a steam turbine to generate electricity. The resulting low-pressure steam is then directed to the gasifier and SOEC sections. Additionally, the low-temperature waste heat may be recovered and sold to the district heat network, if available.

The integrated plant mass and energy balances are computed through process simulations using Aspen Plus®. For the thermodynamic properties, the general model is the RKS-BM complemented with the SRK model in the methanol synthesis section, the NRTL model in the methanol purification section, and the ELECNRTL model in the water scrubber.

The main calculation assumptions are resumed in Table 1. It should be noted that since this is a conceptual study, the simulation relies on models from individual components that have been validated in the original works or balances from experimental equipment directly such as gasifier. The main assumptions for the gasification, syngas cleaning, and methanol synthesis units are consistent with the works by Poluzzi et al. [11,23]. In addition, the SOEC is designed and modeled based on previous similar studies such as [14,17]. A description of the various components of the plant and the modelling methods are outlined in the following sections.

2.1. Biomass Pre-Treatment

A belt dryer is used to reduce the moisture content of the as-received woody biomass from 45 % to 15 %. The heat for moisture evaporation is provided by the drying air. The required thermal energy (1 MWh/t_{H2O} evaporated) is supplied by a hot water loop with temperatures ranging from 90 to 30 °C. Moreover, 32 kWh/t of dry feedstock is considered for the electrical energy consumption of the belt dryer [24].

2.2. Circulating fluidized bed gasification

The gasification process used is based on a pressurized circulating fluidized bed gasifier, that converts woody biomass to syngas. CFBG operates at 870 °C and 4 bar, and a mixture of steam and oxygen is used as gasifying agent. Hence, the biomass partial oxidation by oxygen from SOEC makes the gasification process thermally sustained. The amount of steam fed to the CFBG is specified to yield the steam-to-carbon (S/C) ratio of 1 at the outlet of the gasifier. Moreover, for sealing purposes, additional steam and air are used in the biomass feeder, filter cleaning, and solid purge.

The composition of the produced syngas is determined according to the gasification process parameters and assumptions (Table 1) calibrated to replicate the syngas composition from the Varkaus Demo plant [25]. Minor species are calculated by assuming that all the chlorine in biomass is transformed to HCl, all sulfur to H_2S, and 10 % of the input nitrogen to ammonia.

2.3. Syngas cleaning

In the StE and StCE systems, the raw gas passes through a high-temperature filtration before the catalytic auto-thermal reformer (ATR) [26]. The reformer is fed with oxygen generated as a by-product from SOEC to achieve an exit temperature of 915 °C. A restricted equilibrium calculation approach is employed for the ATR. It is assumed that 90 % of methane content and all higher hydrocarbons are converted into H_2 and CO and that the WGS reaction reaches equilibrium.

Downstream the ATR in the StE and StCE plants and downstream of the gasifier in the SyE plant, the syngas undergoes cooling to a temperature of 250 °C and is then subjected to water scrubbing. The scrubber removes ammonia and chlorine present in the gas. For bulk

Table 1
Process design parameters.

As-received biomass											
LHV, MJ/kg$_{AR}$	Moisture, %wt	Proximate analysis, %$_{wt,dry}$			Ultimate analysis, %$_{wt,dry}$						
9.74	45	FC	V	Ash	C	H	N	Cl	S	O	Ash
		18.84	80.0	1.16	51.19	6.08	0.2	0.05	0.02	41.3	1.16

Belt Dryer				
T$_{biomass,out}$ [°C]	80	Moisture out [wt%]		15

Gasifier						
T$_{Gasifier out}$ [°C]	870	P$_{Gasifier}$ [bar]	4	S/C^1 [-]		1
x$_{CH4}$ [kg$_{CH4}$/kg$_{bio,dry}$]	0.07	x$_{C2H4}$ [kmol$_{C2H4}$/kmol$_{CH4}$]	0.45	Char conversion [%]		95.5
T$_{steam,in}$ [°C]	200	Sealing gas [kg/kg$_{bio,dry}$]	H_2O=0.12Air = 0.03	Q$_{loss}$ [kW]		1000

Syngas cleaning, conditioning and compression						
T$_{reformer,out}$ [°C]	915	CH$_4$ conversion in ATR [%]	90	S/C $_{reformer inlet}$[-]		1
Syngas compressor number of stages	6	Hydrogen compressor number of stages	5	Syngas compressor number of stages (before CO$_2$ separation unit)		4
MDEA process operating pressure [bar]	30	MDEA regeneration thermal duty [MJ/kg$_{CO2,removed}$]	1	MDEA electric consumption, [kWh/ kg$_{CO2,removed}$]		0.012
T$_{intercooler,out}$ [°C]	40	η$_{is, comp}$ [%]	75	η$_{mech,scrubber pump}$ [%]		90
η$_{hyd,scrubber pump}$ [%]	72	η$_{mech, comp}$ [%]	92			

SOEC				
P$_{SOC}$ [bar]	3.2	T$_{SOCin}$[°C]	750	T$_{SOCout}$[°C]	850

Methanol production						
P [bar]	90	T$_{Boiling water}$ [°C]	238	L$_{Tube}$ [m]		6
d$_{Tube}$ [mm]	40	ρ$_{catalyst}$ [kg/m^3]	1712	L$_{catalyst}$ [m]		3.5
d$_{catalyst}$ [mm]	3.5	Bed voidage degree	0.39	GHSV [h^{-1}]		5000
RR	5	η$_{lmech, comp}$ [%]	94	η$_{is, comp}$ [%]		80

Methanol purification						
P$_{Stabilizing column}$ [bar]	1.3	Stage number $_{Stabilizing column}$	20	d$_{Stabilizing column}$ [mm]		0.9
P$_{Concentration column}$ [bar]	1	Stage number $_{Concentration column}$	40	d $_{Concentration column}$ [mm]		2.5

Heat Pump (for concentration column)						
T$_{Source}$ [°C]	64	T$_{Sink}$[°C]	98	COP$_{Real}$/COP$_{Carnot}$ [-]		0.5

sulfur removal, a liquid Redox unit (LO-CAT process) is employed, which involves the conversion of H_2S into elemental sulfur and water. The simulation of this unit is based on data provided by Kazemi et al. [27], treating it as a black box. Finally, a bed of activated carbon is utilized for scavenging of residual H_2S and other impurities.

2.4. Solid oxide electrolysis cell

A high-temperature solid oxide electrolysis cell is employed to adjust the gas composition for methanol synthesis. The SOEC operates at a temperature of 850 °C and a pressure of 3.2 bar, which matches the pressure of clean syngas. The power consumption can be determined by the operational voltage and by estimating the required current based on the amount of oxygen separated by the SOEC [28].

In the StE plant, only steam is fed to the SOEC. The steam utilization is 80 %, with a hydrogen recirculation of 20 % [29], determining the required input steam. The produced hydrogen is then cooled, compressed and mixed to the syngas [30] before the methanol synthesis section to yield the required module.

In the StCE plant, the clean syngas pressure is raised to 30 bar by a four-stage intercooled compressor. The high-pressure syngas then proceeds to a CO_2 removal unit based on MDEA solvent to separate 90 % of the CO_2 in the syngas. The CO_2 is mixed with steam at a pressure of 3.2 bar and a temperature of 440 °C, along with cathode recirculation. The recirculation ensures a specific amount of H_2 (usually 10 vol%) in the feed flow to prevent re-oxidation of Ni-YSZ. The SOEC reactant conversion is defined to yield an appropriate syngas module of 2.05 for methanol production.

Moreover, to achieve the syngas composition yielding the highest possible amount of MeOH, the mass flow rate of the input steam into the electrolyzer is adjusted. With lower S/C ratio (steam-to-carbon ratio), the produced syngas contains a higher amount of methane, which negatively affects methanol production. Conversely, as the S/C ratio increases, the water gas shift reaction is favored, and the conversion of CO_2 to CO decreases, resulting in lower methanol production for a given recycle rate. Considering the balance between methane production and CO_2 conversion, the S/C ratio is set at 4.98 to optimize the overall methanol production process. However, it should be noted that the overall impact of the S/C ratio on the MeOH production is quite small in the vicinity of the optimal value.

In the SyE plant, the purified syngas and steam are fed to the SOEC. The reactant conversion has been tuned to get the syngas module 2.05. The method to determine the amount of input steam is similar to the StCE plant, resulting in a S/C (referred to all carbon containing molecules in the syngas) of 2.3. In this configuration, the SOEC also serves as a methane reforming unit, as the feed has a 10.7 % content (dry basis) of CH_4 and C_2H_6, that contain about 20 % of the total carbon. Within the SOEC, assuming chemical equilibrium at the exit, all the C_2H_6 and 88 % of the CH_4 are converted, and only 1.4 % of the total carbon remains in the methane molecule.

2.5. Syngas compression

Within the StCE plant, the initially pure syngas is compressed to the necessary pressure of 30 bar before entering the CO_2 separation unit. This compression is achieved using a sequence of four intercooled compressors, with intercooling temperature of 40 °C.

In all plants, after cleaning and conditioning, the syngas undergoes compression to 90 bar, corresponding to the methanol synthesis reactor operating pressure, through an intercooled compressor. A six-stage compressor with intercoolers outlet temperature of 40 °C is employed. The resulting pressure ratio per stage is around 1.75, which yields a syngas outlet temperature from each compressor stage of 110–115 °C.

For the StE plant, at the H_2 enrichment step, the hydrogen stream is first compressed to 90 bar by a five-stage intercooled compressor without any aftercooler. The pressure ratio per stage in this compressor

is around 1.95. The compressed hydrogen at around 130 °C is then mixed with syngas

2.6. Methanol production and purification

The Methanol synthesis reactor is modeled using the kinetic model by Vanden Bussche and Froment [31] in Aspen Plus software. The syngas is blended with the recycled gas before being heated in a heat exchanger. It is then fed to the methanol synthesis reactor, where the reaction takes place in a fixed bed reactor filled with commercial copper, zinc oxide, and aluminum oxide catalyst pellets. The reactor is cooled by boiling water at 238 °C. After the reaction, the stream is cooled to 40 °C and separated from the light gases in a flash unit, with the gases being recycled back to the reactor. The reactor's tubes are 6 m long and 40 mm in diameter. The number of tubes in the reactor varies based on the Gas Hourly Space Velocity (GHSV, defined as the volume flow rate of the feed to the reactor (Nm^3/h) per the inner volume of reactor tubes (m^3)). In this study, the GHSV and recycle ratio (RR, defined as the recycle stream molar flow rate divided by the fresh syngas molar flow rate) are considered 5000 h^{-1} and 5, respectively [9,11]. It is worth mentioning that by increasing the RR and fixing inerts (CH_4 and N_2) concentration in the MeOH reactor inlet for each case, the methanol production slightly increases at the cost of a much larger MeOH reactor because of higher recycle flow. So, for all cases, RR is fixed as 5, leading to $CH_4 + N_2$ concentration in the recycled stream between 18 and 24.8 %.

The raw mixture containing methanol, water and other minor components, such as low boiling elements and ethanol, is sent to the purification section after throttling to 2 bar and cooling to 40 °C.

The methanol purification process involves two sequential distillation columns. The first column, the stabilizing column, removes most of the non-condensable gases and has 20 ideal trays. The second column, known as the concentration column, aims to increase the purity of the Methanol to 99.85 %wt and has 40 trays. The process ensures a minimum recovery of 99.5 %mol of Methanol.

2.7. Heat recovery and power generation

Power and Biomass-to-X plants have the potential to recover significant amounts of heat from various sources, such as hot syngas, flue gas, methanol synthesis, and hot SOEC products with a temperature range between 915 °C and 30 °C. Much of this recovered heat is required for preheating, methanol purification, and generating steam for gasification and SOEC units. Excess heat is recovered by raising steam and generate electric power in a steam cycle. Also, low-temperature heat can be recovered and delivered to a district heat network with a forward temperature of 80 °C and return temperature of 40 °C. Such heat can be provided by compressor intercoolers, Methanol condensers and ICE flue gas. If a district heating network is not available, the possibility of using dissipated heat to produce additional power via low-temperature Organic Rankine Cycle would be an option. However, because of the low efficiency and high Capex, this would be economically competitive only in scenarios with very high average electricity prices.

In all plants, an Internal Combustion Engine (ICE) is used to generate electricity from the purge gas from the methanol production and purification units. The performance of the ICE is evaluated using linearized equations as described in the works of Poluzzi et al. [9] and Zatti et al. [32]. ICEs yield an electric efficiency of between around 44.5 % and 46.5 % and a thermal efficiency of between 44.4 and 45.4 %. The exit gases from the ICE are at 400 °C and then cooled down to about 100 °C by reusing around 18.5–19.5 % of the fuel energy input.

The thermal integration for each configuration is dependent on the available heat, technical constraints, and thermal loads within the plant. The steam required in all plants is produced at 32 bar, which corresponds to an evaporation temperature of 236 °C, matching the MeOH reactor cooling temperature. The necessary energy for the evaporation process is supplied by cooling the MeOH reactor and syngas.

The saturated steam produced through the cooling of syngas and methanol reactors is combined and subsequently superheated in the lower temperature syngas cooler section. The superheated steam temperature varies between 485 °C and 350 °C, depending on the steam needs of the plant and the available heat for the superheating process. Such superheated steam is fed to the steam turbine to produce electrical power and steam at 6 bar for the gasifier and 3.2 bar for the SOEC.

Furthermore, a heat pump is used for upgrading the low-temperature heat source from concentration column condenser and providing the required heat for the column reboiler. A water loop at 1.5 bar is also designed to supply the necessary heat for the reboiler of the stabilizing column.

Next, to meet the energy demands of various processes, such as producing saturated water at 32 bar, stabilizing column reboiler, CO_2 capture and dryer unit, SOEC preheaters, and meanwhile to cool down the ICE flue gas, compressors intercoolers, MeOH condenser, and SOEC products, a pinch analysis of the HEN is utilized. The analysis and targeting are performed with a minimum temperature approach difference of 5 °C. The optimization task is carried out using Aspen Energy Analyzer. The resulting T-Q diagrams of the three plants are reported in supplementary material.

3. Result

3.1. Technical analysis

The results of the process simulations of each plant (Figs. 2–4) are presented in Tables 2–4. Each table includes the main streams properties, namely temperature, pressure, mass flow rate and mole fraction. It should be pointed out that the possibility of carbon formation is acknowledged, but it is improbable to occur as long as specific conditions such as moderate reactant conversion rates and high H/C ratios are maintained (Supplementary Material).

Table 5 also reports the main results obtained for the SOC.

To evaluate the performance of the assessed power and biomass-to-methanol plants, the following key performance indicators have been used.

The carbon efficiency (CE) measures how efficiently a process unit converts the carbon in the input biomass stream into the carbon in the output stream. It is calculated by dividing the carbon molar flow rate in methanol product ($F_{C,MeOH}$) by the carbon molar flow rate in the inlet biomass stream ($F_{C,biom}$):

$$CE = \frac{F_{C,MeOH}}{F_{C,biom}} \tag{1}$$

The global fuel efficiency ($\eta_{F,G}$) is defined as the ratio between the chemical energy of the product and the chemical energy input to the process including biomass resource and the produced syngas or hydrogen by SOEC (all based on LHV). It should be pointed out that the calculation of the syngas input's LHV is derived from the difference in LHV between the input to the SOEC and the resulting output stream from this device [33]:

$$\eta_{FG} = \frac{\dot{m}_{out} \bullet LHV_{out}}{\dot{m}_{in} \bullet LHV_{in}} \tag{2}$$

The equivalent fuel efficiency ($\eta_{F,eq}$) takes into account the equivalent biomass consumed due to the net electric power (P_{net}) consumed by the plant. To calculate this, a reference steam cycle with 35 % electric efficiency ($\eta_{el,ref}$) is assumed.

$$\eta_{F,eq} = \frac{\dot{m}_{MeOH} \bullet LHV_{MeOH}}{\dot{m}_{biomass} \bullet LHV_{biomass} + \frac{P_{net}}{\eta_{el,ref}}} \tag{3}$$

Finally, the overall energy efficiency is calculated based on Eq. (4), summing the net electric power either at the numerator (in case of net

Table 2
StE plant stream properties.

		1 Biomass Feed	2 Dried Biomass	3 O₂ to Gasifier	4 O₂ to ATR	5 Fluidizing Steam	6 Raw Syngas	7 Reformed Syngas	8 Steam to SOEC	9 Cleaned Syngas	10 H₂	11 O₂	12 Final Syngas	13 Recycle Flow	14 Syngas to Synthesis	15 MeOH to Purification	16 MeOH	17 Light Gas	18 Purge to ICE
Temperature	°C	25	80	25	25	200	870	915	200	52.66	132.95	30	115	47.02	58.42	41.28	64.58	40.00	37.76
Pressure	bar	1.01	1.01	30	30	5.90	3.80	3.60	3.20	3.20	92	3.20	91.90	92.00	91.90	2	1.01	86.36	1.35
Mass Flow	kg/s	10.27	6.64	1.87	0.60	3.46	11.93	12.53	6.72	8.94	0.75	5.97	9.38	35.42	44.79	9.05	7.02	35.74	0.55
Mole Flow	kmole/h			204	67	691	2047	2335	1342	1617	1342	671	2898	14,489	17,386	1170	790	14,622	157
Mole Fraction	H₂O					100	40.38	33.34	100	3.84				0.08	0.07	30.27	0.21	0.08	0.07
	H₂						20.11	30.26		43.69	100		70.72	73.52	73.05	0.24		73.52	64.36
	CO₂						17.31	15.98		23.05			12.86	3.52	5.07	1.22		3.52	12.04
	CO						14.64	19.07		27.51			15.35	1.04	3.42	0.01		1.04	0.92
	MeOH													0.41	0.34	68.03	99.77	0.41	2.87
	CH₄						4.46	0.39		0.56			0.31	5.61	4.73	0.14		5.61	5.77
	O2			100	100							100							
	C₂H₄						2.01												
	N₂						1.07	0.94		1.36			0.76	15.82	13.31	0.08		15.82	13.94
	H₂S						0.01	0.01											
	HCl						0.01	0.01											
	NH₃						0.001												
	ETOH															0.03	0.02		

Table 3
SyE plant stream properties.

		1 Biomass Feed	2 Dried Biomass	3 Fluidizing Steam	4 O₂ to Gasifier	5 Raw Syngas	6 Cleaned Syngas	7 Steam to SOEC	8 O₂	9 Syngas to Compression	10 Final Syngas	11 Recycle Flow	12 Syngas to Synthesis	13 MeOH to Purification	14 MeOH	15 Light Gas	16 Purge to ICE
Temperature	°C	25	80	200	25	870	69.67	200	30	850	115	46.86	58.15	41.76	64.58	40.00	38.25
Pressure	bar	1.01	1.01	5.90	30	3.80	3.20	3.50	3.07	3.07	92	92.00	92.00	2	1.01	86.43	1.35
Mass Flow	kg/s	10.27	6.64	2.66	1.87	11.93	7.98	9.22	5.38	11.83	8.50	36.18	44.67	8.23	7.09	36.44	0.41
Mole Flow	kmole/h	–	–	691	205	2047	1260	1841	604	3383	2717	13,586	16,303	10,101	797	13,684	117
Mole Fraction	H_2O	–	–	100	–	40.38	3.31	100	–	19.67	–	0.06	0.05	19.07	0.22	0.06	0.06
	H_2	–	–	–	–	20.11	32.69	–	–	55.30	68.85	68.51	68.57	0.30	–	68.51	60.31
	CO_2	–	–	–	–	17.31	28.10	–	–	5.96	7.42	1.75	2.70	0.74	–	1.75	7.87
	CO	–	–	–	–	14.64	23.77	–	–	18.10	22.55	0.81	4.43	0.01	–	0.81	0.73
	$MeOH$	–	–	–	–	–	–	–	–	–	–	–	0.38	79.45	99.76	0.46	3.82
	CH_4	–	–	–	–	4.46	7.20	–	–	0.29	0.37	0.46	6.39	0.25	–	7.59	8.58
	O_2	–	–	–	100	–	–	–	100	–	–	–	–	–	–	–	–
	C_2H_4	–	–	–	–	2.01	3.18	–	–	–	–	–	–	–	–	–	–
	N_2	–	–	–	–	1.07	1.74	–	–	0.65	0.81	20.81	17.48	0.14	–	20.81	18.60
	H_2S	–	–	–	–	0.01	–	–	–	–	–	–	–	–	–	–	–
	HCl	–	–	–	–	0.01	–	–	–	–	–	–	–	–	–	–	–
	NH_3	–	–	–	–	0.001	–	–	–	–	–	–	–	–	–	–	–
	$ETOH$	–	–	–	–	–	–	–	–	–	–	–	–	0.03	0.02	–	–

Table 4
StCE plant stream properties.

		1 Biomass Feed	2 Dried Biomass	3 O₂ to Gasifier	4 O₂ to ATR	5 Fluidizing Steam	6 Raw Syngas	7 Reformed Syngas	8 Cleaned Syngas	9 Compressed syngas	10 CO₂	11 Steam to SOEC	12 O₂	13 CO-H₂ to compressor	14 Low-CO₂ syngas to comp.	15 Final Syngas	16 Recycle Flow	17 Syngas to synthesis	18 MeOH to Purification	19 MeOH	20 Light Gas	21 Purge to ICE
Temperature	°C	25	80	25	25	200	870	915	52.66	40	40	440	30	30	30	115	46.92	58.16	41.81	64.58	40.00	38.27
Pressure	bar	1.01	1.01	30	30	5.90	3.80	3.60	3.20	30	30	3.20	3.20	3.17	30	92	92.00	92.00	2	1.01	86.45	1.35
Mass Flow	kg/s	10.27	6.64	1.87	0.60	2.66	11.93	12.53	8.94	8.94	4.33	10	5.97	4.03	4.30	8.33	36.44	44.77	8.08	7.12	36.69	0.40
Mole Flow	kmole/h	–	–	205	67	691	2047	2335	1617	1617	354	1998	672	1486	1201	2687	13,437	16,124	978	800	94	112
Mole Fraction	H_2O	–	–	–	–	100	40.38	33.34	3.84	3.84	–	100	–	–	–	–	0.06	0.05	16.05	0.22	0.06	0.05
	H_2	–	–	–	–	–	20.11	30.26	43.69	43.69	–	–	–	76.17	58.83	68.42	67.53	67.68	0.31	–	67.53	59.30
	CO_2	–	–	–	–	–	17.31	15.98	23.05	23.05	100	–	–	9.83	1.55	6.13	1.52	2.29	0.68	–	1.52	7.15
	CO	–	–	–	–	–	14.64	19.07	27.51	27.51	–	–	–	13.92	37.04	24.25	0.84	4.74	82.47	99.76	0.84	0.77
	$MeOH$	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	0.47	0.40	0.29	–	0.47	3.98
	CH_4	–	–	–	–	–	4.46	0.39	0.56	0.56	–	–	–	0.08	–	0.38	7.94	6.68	–	–	7.94	9.16
	O_2	–	–	100	100	–	–	–	–	–	–	–	100	–	0.76	–	–	–	–	–	–	–
	C_2H_4	–	–	–	–	–	2.01	0.94	1.36	1.36	–	–	–	–	–	0.82	21.63	18.16	0.16	–	21.6	19.5
	N_2	–	–	–	–	–	1.07	0.01	–	–	–	–	–	–	1.83	–	–	–	–	–	–	–
	H_2S	–	–	–	–	–	0.01	0.01	–	–	–	–	–	–	–	–	–	–	–	–	–	–
	HCl	–	–	–	–	–	0.01	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
	NH_3	–	–	–	–	–	0.001	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–
	$ETOH$	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	–	0.03	0.02	–	–

Table 5
Main results of the SOEC units model.

	StE	SyE	StCE
Cathode Inlet Mole flow (kmole/h)	2013	3101	2969
Cathode Inlet Mole Fraction (%)			
H_2O	83.33	60.72	74.93
H_2	16.66	13.21	10
CO_2	–	11.41	13.21
CO	–	9.65	1.82
CH_4	–	2.92	0.010
C_2H_4	–	1.29	–
N_2	–	0.7	–
Cathode Outlet Mole flow (kmole/h)	2013	3382	2966
Cathode Outlet Mole Fraction (%)			
H_2O	16.66	19.67	36.76
H_2	83.33	55.3	48.16
CO_2	–	5.96	6.21
CO	–	18.1	8.8
CH_4	–	0.29	0.05
C_2H_4	–	–	–
N_2	–	0.64	–
Current (kA)	71,955	64,810	72,029
Reversible Voltage (V)	0.90	0.94	0.95
Operational Voltage (V)	1.31	1.48	1.36
Reactant conversion (%)	80	58.49	51.34
Oxygen separated from cathode gas (kg/s)	5.96	5.37	5.97
Electricity consumption per O_2 separation (kWh/kg)	4.42	4.97	4.56

production) or at the denominator (in case of net consumption):

$$\eta_{tot} = \frac{\dot{m}_{MeOH} \bullet LHV_{MeOH}}{\dot{m}_{biomass} \bullet LHV_{Biomass} + P_{consumption} - P_{Production}} \quad (4)$$

Accordingly, Table 6 shows the key attributes of every Power and Biomass to Methanol plant suggested in this study as well as in the benchmark plant from Poluzzi et al. [11], based on low temperature electrolysis.

3.2. Economic analysis

The economic evaluation is conducted using the levelized cost methodology, which calculates the breakeven selling price of a product (M_{tot}) that covers all expenses (C_{tot}) over the lifetime of the plant. This approach takes into account various factors, including the total capital investment (TCI), the utility costs (C_{ut}), the feedstock expenses ($C_{feedstock}$), and fixed operation and maintenance costs ($C_{fixed\ O\&M}$). In the case of selling the excess low temperature heat to a district heating network, the revenue would be subtracted from the total cost ($DH_{revenue}$). Equation (5) illustrates the relationship between these variables, where $\dot{m}_{fuel}$ represents the nominal fuel production rate, and h_{eq} represents the equivalent annual operating hours. Also, the Capital Charge Factor (CCF) is used to annualize the capital investment accounting for the effects of investment depreciation and the interests during construction.

$$LCOF = \frac{C_{tot}}{M_{tot}} = \frac{TCI \bullet CCF + C_{fixedO\&M} + C_{feedstock} + C_{ut} - DH_{revenue}}{\dot{m}_{fuel} \bullet h_{eq}} \quad (5)$$

The methodology employed for estimating capital expenditures (Capex) and operational expenses (Opex), including utilities, maintenance and repairs, operating supplies, operating labor, laboratory costs, local taxes, insurance, and catalyst, and the heat transfer coefficient for heat exchangers area calculation is thoroughly described by Poluzzi et al. [9,11]. Furthermore, according to the literature on techno-economic studies investigating biomass conversion through gasification, a high plant capacity factor of 90 % has been assumed. High capacity factors are crucial to ensure the economic viability and competitiveness of processes with significant capex such as biomass gasification-based ones.

The key assumptions made in this study, which align with the prior works, are summarized in Table 7. Also, Table 8 reports information to estimate the cost of equipment. All monetary values reported in this paper are expressed in 2019 currency.

Table 9 provides the breakdown of the fixed capital investment (FCI) costs of the assessed plants. The FCI does not include the working capital. However, it is worth mentioning that Capex has significant uncertainty in light of the inflation rate starting from 2022.

Concerning the electrolyzer's FCI, whose voltage and gas composition differ in the assessed plants, the same specific cost referred to electrical capacity of 1000 €/kW [34] has been assumed for simplicity. Furthermore, a third of the investment is designated for the electrolyzer cells, which are projected to last 10 years in the steam-fed electrolyzer and 5 years in the electrolyzer fed with carbon-rich gases [35].

Table 6
Overall performance of the assessed plants.

		StE	SyE	StCE	LT electrolysis [11]
Inputs	Biomass input (MW)	100	100	100	100
	Electric Consumption (MW)				
	Electrolysis	94.95	96.29	98.18	129.04
	Multi-Stage Compressors	11.67	11.01	14.89	8.05
	MeOH Loop Recycle Compressor	0.92	0.84	0.85	0.90
	Heat Pump	1.92	1.79	1.75	–
	Dryer	0.65	0.65	0.65	0.65
	Water and Scrubber Pump	0.07	0.08	0.09	0.04
	O_2 Consumption (kg/s)	2.42	1.82	2.42	2.45
Outputs	MeOH Production (kg/s)	7.02	7.08	7.12	6.99
	O_2 Production (kg/s)	5.97	5.38	5.97	5.97
	District Heat Production (MW)	23.38	22.74	22.16	–
	Electric Production (MW)				
	ICE	4.46	3.53	3.49	4.44
	Steam Turbine	3.28	2.24	1.66	4.34
Performance Indicators	CO/CO_2 of the conditioned syngas	1.19	2.85	3.95	1.21
	MeOH output (MW)	139.66	140.90	141.65	139.07
	Total Electricity Consumption (MW)	110.20	110.67	116.41	138.67
	Total Electricity Production (MW)	7.75	5.78	5.15	8.78
	Net Electricity Output (MW)	−102.45	−104.89	−111.26	−129.89
	Carbon Efficiency (%)	91.06	91.86	92.35	90.46
	Global Fuel Efficiency (%)	73.42	74.19	73.55	73.35
	Equivalent fuel efficiency (%)	35.56	35.25	33.89	28.85
	Total efficiency (%)	68.98	68.76	67.04	61.94
	Total efficiency (%) (with DH)	80.53	79.86	77.53	–

Table 7
Variables and presumptions in the evaluation of LCOF.

Economic parameters	Value	Ref.
Discount rate, %	10	[11]
Lifetime, y	20	[11]
Capital Charge Factor, %	11.75	[11]
Availability, h/year	7884	[11]
Electrolyzer capacity factor, %	80	[11]
Variable Opex		
Biomass feedstock cost, €/t	45.72	[36]
Electricity price (2019 average in DK)[a], €/MWh	38.49	[37]
District Heat price, €/MWh	38.4	[38]
Fixed Opex		
Maintenance and repairs, % FCI	5	[9]
Operating supplies, % FCI	0.5	[9]
Operating labor, % Opex	10	[9]
Laboratory costs, % Opex	2.5	[9]
Local taxes, % FCI	1	[9]
Insurances, % FCI	1	[9]
Catalyst cost of MeOH reactor, €/kg	18.1	[9]
Catalyst lifetime, y	4	[9]

[a] The 2019 Danish electricity price has been selected to allow a more consistent comparison with the benchmark process based on low temperature electrolysis from [11]. Additionally, Denmark is taken as an example for the electric grid dominated by renewables, which may be representative of future electricity prices in Nordic countries.

4. Discussion

4.1. Technical performance

The SOEC of SyE case needs to separate lower O_2 flow rate (see Table 5), as the ATR is not present and no O_2 is added to syngas after the gasifier. Conversely, in StE and StC cases, O_2 which is added to the ATR (0.6 kg/s) has to be separated in the SOEC, which separates 0.6 kg/s of O_2 more than the SyE case. On the other hand, the flow rate of syngas flowing through the SOEC of the SyE case is higher, involving higher

heat generation in the SOEC and higher electricity consumption per unit of O_2 separated (4.97 kWh/kg$_{O_2}$ vs. 4.42–4.56 kWh/kg$_{O_2}$ of StE-StCE cases). As a result, the SOEC of the SyE case has similar electricity consumptions of the StE and StCE cells. StCE case shows the highest consumption due to a combination of higher O_2 separation and relatively high flow rate of H_2O-CO_2 gas to be heated through the cell. Overall, the SOEC electric power consumption differs by less than 4 % between the StE and the StCE cases and is overall quantitatively similar to the biomass energy input.

The other significant electric consumption in all plants is related to the multistage compressors. StCE case has a higher consumption compared to StE and SyE, due to the requirement for higher syngas pressure in CO_2 separation unit. Then, the CO_2 pressure is lost when solvent is regenerated, before mixing with the steam and entering the electrolyzer. The gross electricity consumption of StE, SyE and StCE plants is 110.2 MW, 110.67 MW and 116.41 MW, respectively.

Regarding the electricity generation, it is produced mostly from the ICE, followed by the steam turbine. Since the StE plant uses less high-temperature steam, the superheated steam produced by cooling the outlet syngas from the gasifier can be directed to the steam turbine with a higher temperature, resulting in around 3.28 MW of power generation. In comparison, the SyE and StCE plants produce 2.24 MW and 1.66 MW of power, respectively. Overall, the gross power production is in the 5–8 MW range, i.e. much less than electric consumption, leading to net electric consumption between 102 MW (StE case) and 111 MW (StCE case).

MeOH production is quite similar in all cases with small differences which are mainly related to the CO/CO_2 ratio in the MeOH synthesis section feed. The higher CO/CO_2 ratio of StCE case leads to slightly higher MeOH yield for the given fixed recirculation rate and therefore higher carbon efficiency and global fuel efficiency.

Overall, the combination of all mentioned results leads to similar carbon and fuel efficiencies in all cases. Equivalent and total efficiencies are also similar among the cases, with slightly better performance for the

Table 8
Data to estimate the cost of equipment.

Capital costs	Cost scaling parameter	Reference capacity	Scaling exponent	Reference purchase equipment delivered. M€	Lang factor	Ref.
Biomass-to-syngas island						
Feedstock handling	Biomass feed, MW$_{th}$	157.00	0.31	6.94	1.48	[24]
Belt dryer	Water evap., kg/s	0.34	0.28	2.49	1.48	[24]
Pressurized O_2 CFB gasifier	Dry biom., kg/s	17.80	0.75	49.38	1.42	[24]
Ceramic hot-gas filter	Syngas, kmol/s	1.47	0.67	8.91	1.48	[24]
Catalytic reformer	Syngas, kmol/s	2.04	0.67	28.55	1.42	[24]
Cleaning						
Sulfur removal (Liquid redox)	Synga, kmol/s	0.64	0.67	0.57	5.04	[23]
CO_2 separation	Separated CO_2. kg/h	46,600	0.67	16.69	1.40	[23]
Scrubber	Syngas at cleaning inlet, kmol/s	0.64	0.67	0.27	5.04	[39]
Activated carbon	Syngas at cleaning inlet, kmol/s	0.64	0.67	0.09	3.97	[39]
Waste water treatment	Waste water. m^3/h	22.56	0.67	0.45	5.04	[39]
Electrolyzer	Electricity Consumption. MW	1	1	1000	1.00	[34]
Compression						
Syngas compressor	Compressor power, MW$_{el}$	7.01	0.67	7.50	5.04	[39]
CO_2 compressor	Compressor power, MW$_{el}$	0.64	0.67	0.75	5.04	[23]
H_2 compressor	Compressor power, MW$_{el}$	0.64	0.67	0.75	5.04	[39]
Syngas-to-methanol island						
Methanol boiling water reactor	Syngas molar flow, kmol/s	2.20	0.67	1.72	4.28	[23]
Recycle compressor	Compressor power, MW$_{el}$	0.41	0.67	0.44	5.04	[23]
Stabilizing column	Raw Methanol. kmol/s	0.15	0.67	0.10	5.04	[23]
Concentration column	Raw Methanol. kmol/s	0.14	0.67	0.36	5.04	[23]
Heat Recovery System						
CHP internal combustion engine	Fuel input. kW$_{th}$	13,783	0.95	2.48	1.40	[32]
Heat Exchangers						
Economizer	Area, m^2	10,000	0.68	0.96	5.04	[40]
Evaperator	Area, m^2	5000	0.79	1.16	5.04	[40]
Superheater	Area, m^2	505	0.74	0.13	5.04	[40]
Shell & Tube	Area, m^2	500	0.60	0.18	5.04	[40]
Heat Pump	Compressor power, HP	$TotalInvestment = 9.65W^{0.62}$				[41]

Table 9
Breakdown of the fixed capital investment costs.

Components	Fixed capital investment M€		
	StE	SyE	StCE
Biomass-to-syngas island			
Feedstock handling	8.93	8.93	8.93
Belt dryer	7.15	7.15	7.15
Pressurized O_2 CFB gasifier	29.65	29.65	29.65
Ceramic hot-gas filter	6.99	6.99	6.99
Catalytic reformer	17.25	–	17.25
Total	69.97	52.72	69.97
Cleaning			
Sulfur removal (Liquid redox)	2.87	2.63	2.87
CO_2 separation	–	–	11.22
Scrubber	1.38	1.26	1.38
Activated carbon	0.36	0.33	0.36
Waste water treatment	2.14	2.02	1.97
Total	6.75	6.24	17.80
Electrolyzer	94.95	96.29	98.18
Compressors			
Syngas compressor	34.81	51.16	50.37
CO_2 compressor	–	–	13.21
H_2 compressor	15.91	–	–
Total	50.73	51.16	63.58
Syngas-to-methanol island			
Methanol reactor	12.47	11.94	12.38
Recycle compressor	3.81	3.59	3.60
Stabilizing column	0.84	0.77	0.75
Concentration column	3.09	2.86	2.79
Total	20.21	19.15	19.52
Heat Recovery System			
CHP internal combustion engine	2.50	2.03	1.99
Heat Exchangers			
Economizer	4.11	7.83	4.00
Evaporator	1.57	0.56	0.47
Superheater	4.73	1.52	2.13
Other heat exchangers	9.76	8.64	10.42
Heat Pump	1.19	1.14	1.13
Total	35.77	31.95	31.41
FCI	266.46	247.28	289.18

StE case ($\eta_{F,eq}$ = 35.56 % and η_{tot} = 68.98 %) and worse performance for the StCE case ($\eta_{F,eq}$ = 33.89 % and η_{tot} = 67.04 %).

When it comes to comparison of high-temperature-electrolysis based plants with low-temperature one (last column in Table 5), both the StE plant and the benchmark low temperature (LT) electrolysis case from Poluzzi et al. exhibit comparable performance regarding syngas composition and methanol production. The principal difference between the two lies in the SOEC electricity consumption, as the StE plant consumes 26.41 % less electricity compared to the LT electrolysis case. Indeed, the higher electricity consumption of LT electrolysis in the benchmark case renders it less competitive when compared to other plants concerning equivalent fuel efficiency and total energy efficiency (5–7 percent points difference for both indicators).

Based on the obtained results, the benefit of using high efficiency

high temperature electrolysis compared to benchmark low temperature electrolysis is clear. On the other hand, from the technical KPIs, it is not possible to identify a clearly best integration strategy of high temperature electrolysis systems into biomass to methanol plants. This will depend on economic considerations, on the maturity and reliability of the key components and possibly on the operational flexibility.

4.2. Economic performance

Regarding economic analysis, Table 9 provides an overview of the fixed capital investment costs for each plant. Specifically, regarding the Biomass to Syngas island, the StE and StCE plants have comparable investment costs. On the other hand, the investment cost for the SyE plant is 25.29 % lower compared to the others. This is primarily due to the absence of a catalytic reformer. As mentioned previously, in the SyE plant, CH_4 undergoes reformation within the SOEC. As for the cleaning section, the StCE plant features a higher cost due to the CO_2 separation unit (11.2 M€). As for the other cleaning components, all plants show similar values.

Regarding the FCI of the electrolyzer, because the electrical energy consumption is relatively similar across all plants, the FCI is comparable. However, since more than one-third of the investment is allocated to the electrolyzer cells, which have a projected lifespan of 10 years in the steam-fed electrolyzer and 5 years in the electrolyzer fed with carbon-rich gases, the main differences between electrolyzers costs are related to the O&M costs, which range from 1.74 M€/year for the StE to 5.3 5.4 M€/year for the SyE, and StCE plants. The electrolyzer share in the total investment cost is 36 % for the StE plant, 39 % for the SyE plants, and 34 % for the StCE plant. The other large cost share in all plants is related to the gas compression equipment, which accounts for around 19 %, 21 %, and 22 % of the total FCI for the StE, SyE, and StCE plants, respectively.

The fixed capital investment for the syngas to methanol island is comparable across all plants. The StE plant has a slightly higher methanol island cost, primarily due to the larger Methanol reactor and recycle compressor. The heat recovery system FCI cost share is 9 %, 9 %, and 7 % for the StE, SyE, and StCE, respectively.

In summary, the StCE plant has the highest fixed capital investment among the three cases, with costs exceeding those of the SyE and StE plants by 14.4 % and 7.8 %, respectively. This disparity can be attributed mainly to the inclusion of the CO_2 removal unit and CO_2 compressor in the StCE plant. On the other hand, the SyE plant exhibits the lowest FCI primarily due to lower costs of the biomass-to-syngas islands.

Table 10 presents the key outcomes of the economic analysis. It should be noted that the O&M costs exhibit a similar pattern to the TCI. When comparing the plants operating with SOEC, a notable observation is that SyE and StE plants have comparable LCOF. In contrast, the StCE case exhibits the highest LCOF due to its higher TCI and electricity consumption. It should be noted that by considering revenues achieved from DH network operating all the year in all cases, the LCOF would decrease by around 6 %.

Table 10
Key results of economic analysis.

Economic results	StE	SyE	StCE	Benchmark LT electrolysis[a] [11]
TCI, M€/y	36.81	34.16	39.95	39.33
O&M, M€/y	31.26	33.4	37.37	30.65
Purchased electricity cost, M€/y	31.09	31.83	33.76	28.45
Biomass cost, M€/y	13.32	13.32	13.32	13.32
Total costs, M€/y	105.4	105.83	117.7	111.76
District Heat revenues, M€/y	7.08	6.88	6.71	–
Methanol production, kt/y	199.2	201.0	202.1	177.3
LCOF (with DH revenues), €/t	529.02	526.49	582.43	–
LCOF (with DH revenues), €/GJ	26.59	26.46	29.27	–
LCOF (w/o DH revenues), €/t	564.55	560.74	615.62	630.28
LCOF (w/o DH revenues), €/GJ	28.38	28.18	30.94	31.67

[a] The benchmark case based on low temperature electrolysis from [11] considers flexible operation, where the electrolysis unit operates with a capacity factor of 80 %. Therefore, methanol production and purchased electricity costs derive from weighted average of the operating periods with and without electrolysis.

Comparing the SOEC-based cases with the benchmark LT electrolysis case [11], the TCI in the StE case is 6.4 % lower compared to the benchmark LT electrolysis case. Lower TCI have been obtained despite the higher cost of the electrolysis system (+16 %), due to additional equipment (WGS reactor for syngas conditioning, CO_2 separation unit and O_2 storage) required in the benchmark case. On the other hand, the assumption of replacing the cell every 10 years for the solid oxide steam-electrolysis cell lead to 2 % higher O&M cost compared to the LT electrolysis case.

It has to be noted that in the benchmark LT electrolysis case, the biomass to methanol plant was conceived to operate flexibly depending on the electricity price and worked in "baseline mode" (i.e. with electrolyzer off and no hydrogen addition to syngas) for 20 % of time and in "enhanced mode" (i.e. with electrolyzer on) for 80 % of the time (6308 h). Therefore, the purchased electricity price of the benchmark case is 12.5 % lower than the StE case and the amount of purchased electricity is 8.4 % less than that of StE case. On the other hand, when the plant works in baseline mode (i.e. as BtX plant), the MeOH production reduces from 6.99 kg/s to 3.29 kg/s, leading to lower yearly MeOH production than the StE plant. Taking into account all these effects, the levelized cost of fuel in the StE plant amounts to 28.38 €/GJ (without DH revenues), which is 10.38 % lower than that of the benchmark case.

Fig. 5 graphically shows the LCOF breakdown of the different plants. In the benchmark LT electrolysis plant, most of the cost share is attributed to the TCI, accounting for 35.1 %, followed by the fixed O&M cost, which makes up 27.4 %. In the high-temperature SOEC-based plants, the electricity cost takes the highest share of the total cost, averaging around 33 %. Overall, capital investment, O&M costs and purchased electricity contribute with a similar share (in the 30 % range each) to the LCOF.

To assess the systems performance under varying district heat and electricity prices, a sensitivity analysis is conducted. This analysis involves varying the electricity price between 0 and 100 €/MWh and the district heating price between 0 and 50 €/MWh to determine the effects on the plant LCOF. Fig. 6 depicts the relationship between the levelized cost of fuel and the average electricity and district heat prices. It should be mentioned that the electricity price constitutes a large portion of the plants total cost (Fig. 5), making the LCOF in the plants particularly sensitive to the electricity price. For instance, with a DH price of 50 €/MWh, a change in electricity price from 0 to 100 €/MWh increases the LCOF by 126 % and 122 % for SyE and StE plants, respectively.

Electricity price affects the LCOF much more than the heat selling prices. Additionally, it is worth noting that while the LCOF in the SyE plant is slightly lower than that of the StE plant at the reference point, this minor advantage becomes even less significant when electricity price exceeds a certain threshold relative to the DH price. For instance, when the DH price is assumed 38.4 €/MWh and the electricity price is 75 €/MWh, the LCOF for the StE plant starts to become more favorable compared to that of the SyE plant. Such a trend can also be seen in Fig. 7 where SOEC price increases from 500 €/kW to 1500 €/kW and StE plant depicts a little better performance compared to SyE case when the SOEC price is higher than 1100 €/kW.

5. Conclusions

This study provides a comprehensive analysis of three distinct configurations of a biomass-to-methanol plant, each incorporating a CFB O_2-blown biomass gasifier and a high temperature solid oxide electrolysis cell (SOEC). The configurations were differentiated based on whether the SOEC was fed with steam (StE), syngas (SyE), or a mixture of CO_2 and steam (StCE). The SOEC units have been sized to achieve high carbon efficiency, i.e. to retain the maximum amount of biogenic carbon in the produced methanol. From the results obtained, it is possible to list the following main conclusions:

- If SOEC system is sized properly, all the explored integration options can achieve high carbon efficiency, above 90 %.
- The energy efficiency (with both the "global fuel efficiency" and "equivalent efficiency" definitions used in the study) was also found to be similar in the different integration options, with differences below 2 percentage points. On the other hand, the advantage of integrating high temperature electrolysis systems with respect to low temperature electrolysis is significant, as it allows achieving about 7 percentage points higher efficiency.
- The capital investment cost of SOEC system (calculated with an assumed cost of 1000 €/kW) in the power and biomass to methanol plants is significant, with a share between 36 and 41 % of the total plant cost. The high Capex of the SOEC system also reflects on the fixed Opex, which are dominated by the cost of SOEC replacement. Reducing the SOEC Capex and increasing its lifetime would have a major impact on the levelized cost of fuel (LCOF).
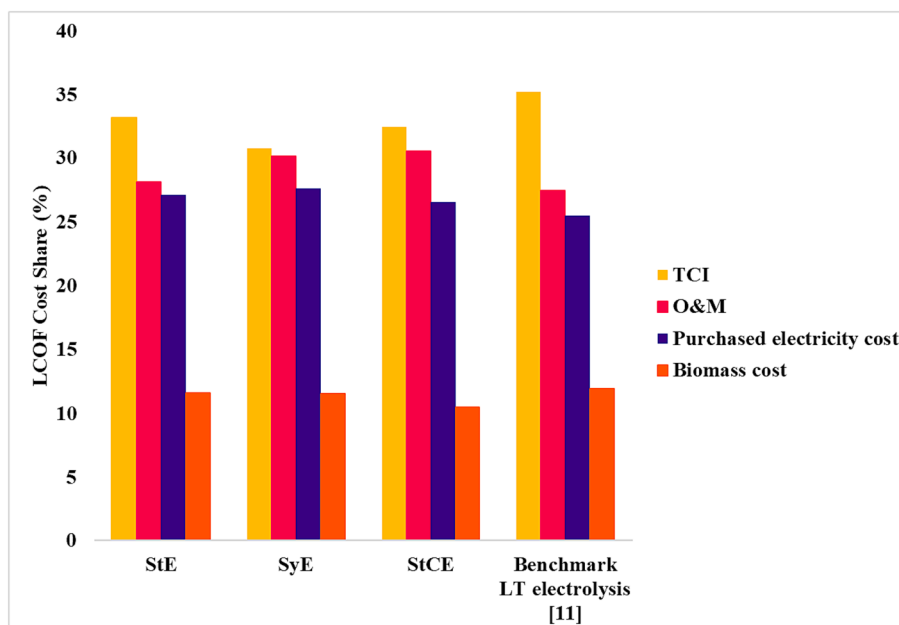

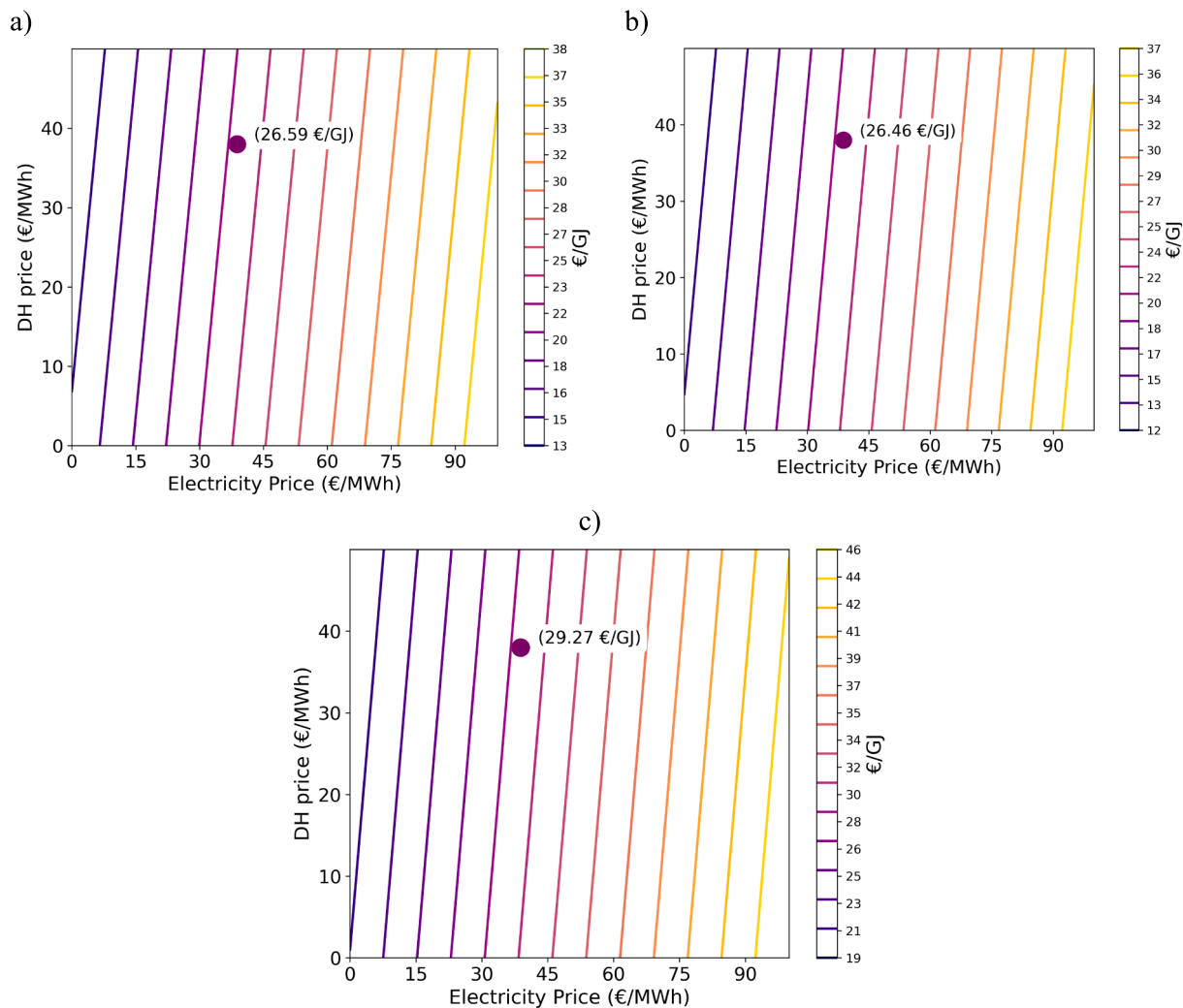**Fig. 5.** Contribution of each cost to the total cost of MeOH production.

Fig. 6. Sensitivity analysis of LCOF on average electricity and district heating prices, (a) StE plant, (b) SyE plant, and (c) StCE plant. The purple point shows the LCOF at the reference point where the electricity and district heating price are 38.49 and 38.4 €/MWh, respectively. (For interpretation of the references to colour in this figure legend, the reader is referred to the web version of this article.)
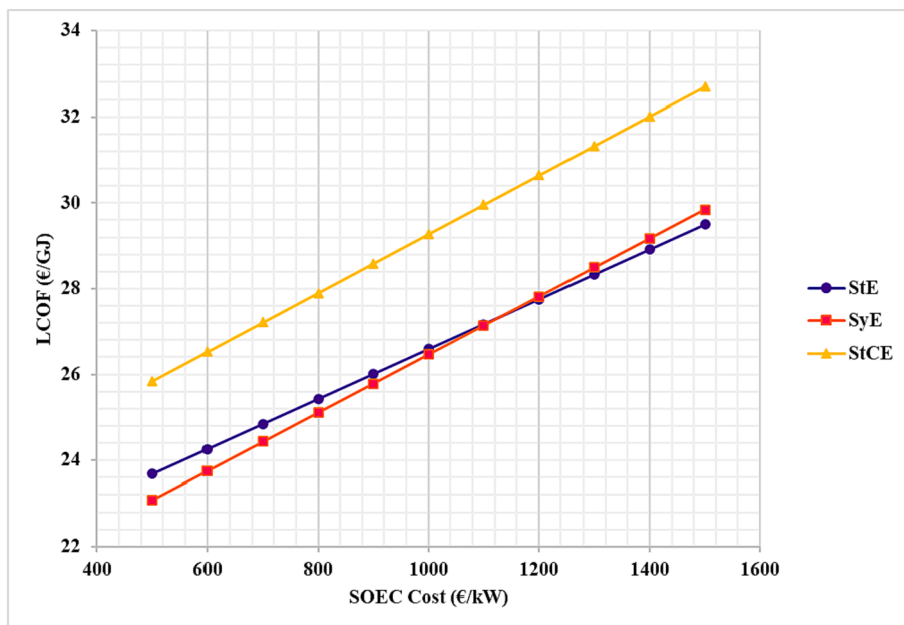

Fig. 7. Sensitivity analysis of LCOF on SOEC price.

- LCOF between 28.2 and 30.9 €/GJ have been found for the assessed plants. StE and SyE plants feature similar LCOF (28.2–28.4 €/GJ). Capital investment, O&M costs and purchased electricity contribute with a similar share (about 30 % each) to the LCOF. Higher LCOF have been obtained for the StCE case, mainly due to the higher consumption for electrolysis and consequently higher Capex, higher O&M and higher costs for purchased electricity. The possibility of recovering waste heat for district heating, allows reducing the LCOF to 26.6–29.3 €/GJ, i.e. 5.4–6.4 % less than without district heating availability. The calculated LCOF resulted 3.3–11.0 % lower than for the benchmark low temperature electrolysis case from literature, calculated with consistent assumptions.

Overall, the findings of this study provide valuable insights for the design and operation of carbon-efficient future biomethanol production plants. The choice of the most suitable SOEC integration strategy and system configuration will largely depend on the SOEC cost and lifetime and on the opportunities of flexible operation to take advantage of the variable electricity price.

CRediT authorship contribution statement

Fatemeh Rajaee: Conceptualization, Formal analysis, Investigation, Methodology, Software, Validation, Visualization, Writing – original draft, Writing – review & editing. **Giulio Guandalini:** Investigation, Methodology, Software, Validation. **Matteo C. Romano:** Conceptualization, Formal analysis, Investigation, Methodology, Project administration, Supervision, Validation, Writing – original draft, Writing – review & editing. **Jouni Ritvanen:** Conceptualization, Formal analysis, Investigation, Methodology, Project administration, Resources, Supervision, Writing – review & editing.

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.

Appendix A. Supplementary material

Supplementary data to this article can be found online at https://doi.org/10.1016/j.enconman.2023.118009.

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