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Methanol and carbon monoxide co-production *via* methane decomposition: techno-economic and environmental analysis

Aaditya Pandey, Farzeena Parayil Ayubkhan, Niharika Mylipilli, Amisha Thotakura and Navid Khallaghi *

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This study presents the development and optimisation of a co-production process for methanol and carbon monoxide *via* methane decomposition using Aspen Plus. The process model was designed to evaluate the system's energy efficiency, economic viability, and environmental impact. The overall energy efficiency of the process was calculated to be 89.4%, demonstrating its high performance in energy utilisation. The levelised cost of methanol production was determined to be 17.5 € per GJ, indicating competitive economic feasibility. Furthermore, a total life cycle CO_2 emission of 0.5 kg_{CO_2} $kg_{product}^{-1}$ was achieved, highlighting the process's potential for reduced environmental impact compared to conventional methods. The results suggest that methane decomposition for the co-production of methanol and carbon monoxide offers a promising pathway for sustainable chemical production, combining high energy efficiency with low carbon emissions.

1. Introduction

Methanol, primarily produced through the steam reforming of natural gas (NG), is a vital precursor for various chemicals and fuels, including formaldehyde, acetic acid, and dimethyl ether.[1] The increasing global demand for methanol (91 million tons in 2023[2]), driven by its energy and chemical synthesis applications, necessitates a thorough understanding of the underlying production processes, particularly those involving natural gas cracking and syngas generation. Methanol can be synthesised from methane *via* various pathways, with methane cracking being a promising method that offers the potential for lower carbon emissions compared to traditional reforming processes.[3] Methane cracking, or pyrolysis, involves the thermal decomposition of methane into hydrogen and solid carbon, thus eliminating carbon dioxide emissions associated with conventional methods of methanol production.

The economic feasibility of methane cracking technologies is being evaluated based on their ability to produce hydrogen at competitive prices. Research shows that methane pyrolysis can lower production costs considerably compared to conventional steam methane reforming (SMR) processes,[4] especially when factoring in the value of the solid carbon byproduct, which can be applied in multiple industrial sectors.[5,6] Pérez *et al.*[7] investigated the techno-economic feasibility of methane pyrolysis in a molten gallium bubble reactor under different heat supply scenarios: (1) carbon combustion with and without carbon capture and storage (CCS), (2) hydrogen combustion, (3) natural gas combustion with and without CCS, and (4) electricity. The levelised cost of hydrogen (LCOH) analysis showed that carbon combustion (2.94 € per kg_{H_2}) and electricity-based heating (3.16 € per kg_{H_2}) were cost-competitive with SMR without CCS (2.86 € per kg_{H_2}). In contrast, hydrogen combustion had the highest LCOH (4.03 € per kg_{H_2}) due to increased NG consumption for reactor heating. Regarding sustainability, carbon combustion with CCS and electricity were the most environmentally favourable options, offering lower CO_2 emissions and carbon taxes while maintaining economic viability. Sensitivity analysis highlighted that NG prices and carbon sales were key factors influencing the economic competitiveness of methane pyrolysis compared to conventional SMR. Kerscher *et al.*[8] and Riley *et al.*[9] conducted techno-economic assessments of methane pyrolysis using different reactor technologies: an electron beam plasma reactor and a catalytic fluidised bed reactor. Their analyses considered different energy supply scenarios in which the required heat or electricity was sourced from hydrogen, renewable energy, or natural gas. For electron beam plasma-driven methane pyrolysis, the lowest levelised cost of hydrogen was reported at 2.55 € per kg_{H_2}. However, this remained uncompetitive compared to conventional SMR with and without CCS, which ranged between 1.00–1.18 € per kg_{H_2}, though it performed better than water electrolysis (4.31 € per kg_{H_2}). The high LCOH of electrolysis was attributed to its significantly higher energy demand for splitting water (286 kJ per mol_{H_2}) compared to methane (37.5 kJ per mol_{H_2}). Additionally, the study identified the high capital cost of electron accelerators as a significant economic barrier, accounting for

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Centre for Energy Decarbonisation and Recovery, Cranfield University, Bedford, Bedfordshire, MK43 0AL, UK. E-mail: n.khallaghi@cranfield.ac.uk

over half of the total LCOH. However, as plasma technology advances and costs decline, the LCOH could drop below 1.5 € per kg_{H_2}. Riley *et al.*[9] estimated that methane pyrolysis in a catalytic fluidised bed reactor had an LCOH of 2.6–2.8 € per kg_{H_2}, assuming no financial benefit from carbon byproducts. Tabat *et al.*[10] introduced a mobile autothermal methane pyrolysis unit designed to overcome the challenges of limited hydrogen pipeline infrastructure while ensuring economic viability. The study assessed the system's efficiency and performance through energy and exergy analyses. The economic evaluation revealed a levelised cost of hydrogen (LCOH) between 1.1 € per kg and 1.3 € per kg, with a net present value ranging from 3.3 to 3.8 M€, depending on engineering, procurement, construction costs, and feedstock prices. The combination of a positive net present value, competitive LCOH, and a high methane conversion rate of 76.8% highlighted the profitability of the proposed system.

To the best of the authors' knowledge, there is a significant gap in the literature regarding comprehensive evaluations of methane decomposition from both techno-economic and environmental perspectives. This research aims to fill that gap by exploring the feasibility of co-producing methanol and carbon monoxide through methane thermal decomposition. The process is modelled using Aspen Plus, and its technical performance is assessed based on overall energy efficiency (OEE). The thermodynamic results are then used to evaluate the economic feasibility of the process. Lastly, the total life cycle CO_2 emissions (TLCCE) are calculated to assess its environmental impact.

2. Process description and model development

2.1. Process description

The process is modelled at steady state, based on a natural gas feed rate of 50 kg s^{-1} (methanol production of ~5700 t per day), the basis for mass and energy flow analysis throughout the system. The process commences by feeding methane at 3 bar and 450 °C into the decomposition reactor (Stream 2), where it undergoes thermal decomposition at 1000 °C to produce a hydrogen-rich gaseous mixture, solid carbon (C), and unconverted methane (Stream 5). After cooling to 573 °C, the product enters a separator to separate the solid carbon (Stream 18) from hydrogen and unconverted methane (Stream 7). This represents an idealised, high-temperature carbon removal step. In practice, however, such a unit would likely require further cooling (*e.g.*, to ~350 °C) before filtration or cyclone separation, due to material and thermal limitations. The Hydrogen-rich gas is subjected to the Pressure Swing Adsorption (PSA) Unit. The PSA unit helps yield the high-purity hydrogen required for methanol synthesis. After the PSA, the unconverted CH4 (Stream 11) is recycled back to the decomposition reactor to enhance the efficiency of H_2 production and minimise waste. On the other hand, pure CO_2 (captured CO_2 from any other industries) partially reacts with solid carbon to produce carbon monoxide. The reaction mechanism used to produce CO is known as the Boudouard reaction (R1).

$$C_{(s)} + CO_{2(g)} \rightarrow CO_{(g)}, \Delta H = 172 \text{ kJ mol}^{-1} \tag{R1}$$

The remaining pure CO_2 (Stream 21) is then mixed with the hydrogen from the PSA Stream and sent to the methanol reactor. The mixture is pressurised up to 43 bars *via* a multi-stage compressor before entering the methanol production reactor. The produced methanol is then entered into a distillation column for water separation, reaching 99% pure methanol (Fig. 1).

2.2. Model development

The Peng–Robinson equation of state is deployed to simulate the process in Aspen Plus software. The PSA unit is modelled with a simple separator in Aspen Plus. The pure CO_2 flow rate is calculated *via* Aspen (Design Specs), considering the solid carbon flow rate. Table 1 summarises the parameters used for other components.

Furthermore, assumptions are made to simulate the process presented in Table 2.

The mass balance and composition for the main streams of the process are summarised in Table 3.

3. Techno-economic and environmental assessment indicators

3.1. Techno-economic assessment indicators

The thermodynamic performance is evaluated by overall energy efficiency, considering the system's input/output thermal energy and available heat.

$$\text{OEE} = \frac{\dot{m}_{\text{MeOH}} \times \text{LHV}_{\text{MeOH}} + \dot{m}_{\text{CO}} \times \text{LHV}_{\text{CO}}}{\dot{m}_{\text{NG}} \times \text{LHV}_{\text{NG}} + W_{\text{tot}} + \ Q_{\text{tot}}} \tag{1}$$

where $\dot{m}_{\text{MeOH}}$, $\dot{m}_{\text{NG}}$ and $\dot{m}_{\text{CO}}$ represent the mass flow rate of methanol, natural gas and CO, respectively. Similarly, LHV_{MeOH}, LHV_{CO} and LHV_{NG} are respective lower heating values for methanol, CO and NG. In addition, Q_{tot} and W_{tot} correspond to the total required heat and power for the plant operation. To better assess the economics of the process, the levelised cost of methanol production (LCOM, € per GJ) is introduced as the crucial economic factors.

$$\text{LCOM € per GJ} = \frac{\text{TAC M€ per year}}{\dot{m}_{\text{MeOH}} \text{ kg per s} \times \text{LHV}_{\text{MeOH}} \times 3600 \times 7446 \text{ h per year}} \times 1000 \tag{2}$$

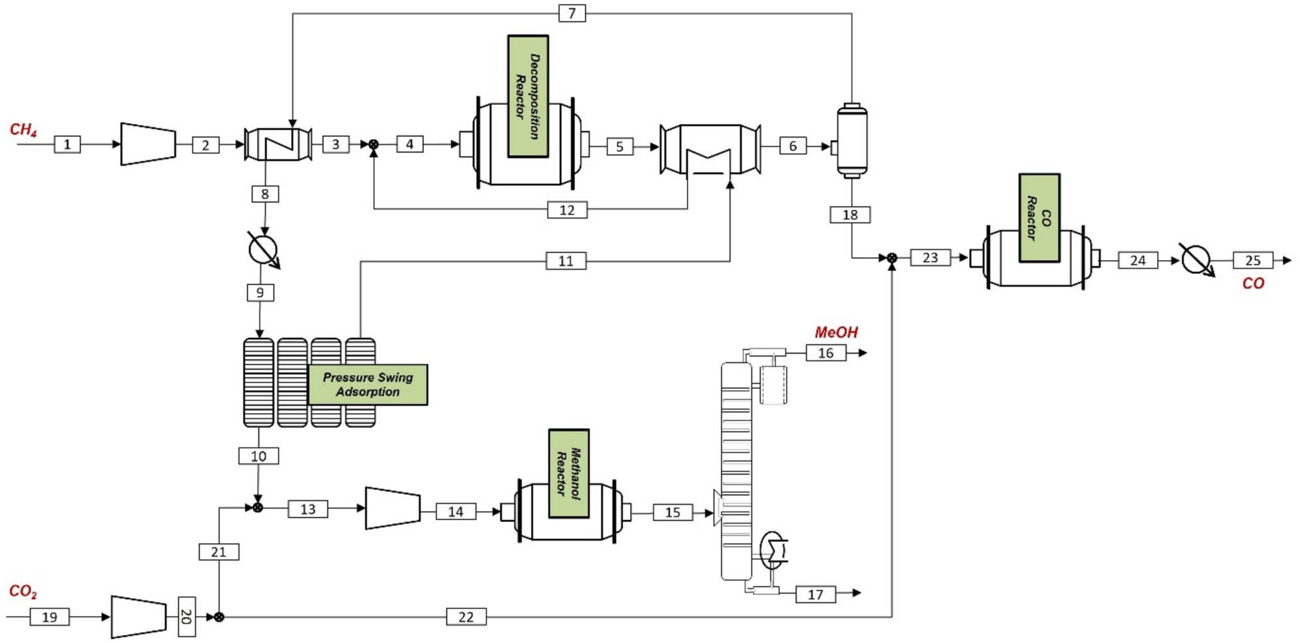

Fig. 1 Schematic of the MeOH and CO co-production plant.

Table 1 Aspen Plus components used in the process modelling

Unit	Aspen plus ID	Description
Cooler	Heater	
Heat exchanger(s)	HeatX	
Decomposition reactor	RGibbs	$T = 1000$ °C[11]
CO reactor	RStoic	$T = 700$ °C,[12] $C_{(s)}$ conversion = 1
Methanol reactor	RStoic	$T = 200$ °C,[13] H_2 conversion = 1 (ref. 14) 80% heat recovery
Compressor	Compr,MCompr	
PSA	Sep	95% hydrogen recovery[15]
Solid carbon separator	SSplit	
Distillation column	RadFrac	No. stages = 30, pressure = 1 bar,[16] Reboiler = kettle, condenser = total

Table 2 Assumptions to simulate the process configurations

Parameter	Value	Ref.
Heat exchangers: hot-inlet–cold-outlet temperature difference (°C)	50	17
Pressure drops-heat exchangers (%)	3	18
Pressure drops-reactors (%)	5	19
Isentropic efficiency of compressor (%)	85	20
Mechanical efficiency of compressors (%)	99.6	20

The total annualised cost (TAC) plays a key role in determining the LCOM calculation. This parameter is calculated based on the total plant cost (TPC), the fuel cost (C_{fuel}), the energy cost (C_{energy}), variable ($V_{\text{O\&M}}$) and fixed ($F_{\text{O\&M}}$) operating and maintenance costs.

$$\begin{aligned}\text{TAC M€ per year} = {} & \text{TPC} \times \text{ACCR} + V_{\text{O\&M}} + F_{\text{O\&M}} \\ & + C_{\text{fuel}} + C_{\text{energy}}\end{aligned} \tag{3}$$

The annualised capital charge ratio (ACCR) is a factor that is formulated in eqn (6), considering the project interest rate (r) and project lifetime (n).

$$\text{ACCR} = \frac{r(1+r)^{\,n}}{(1+r)^{\,n} - 1} \tag{4}$$

The equipment purchase cost (C_B) is estimated using the reference cost data from the literature and eqn (7), where C_A, Q_A and f represent the reference component cost, capacity and scaling factor, respectively.

$$C_B = C_A\left(\frac{Q_B}{Q_A}\right)^f \tag{5}$$

Table 3 Mass flow, temperature, pressure and composition of the main streams

Streams	1	2	3	5[a]	7	10	11	13	16	18	19	22	25
P (bar)	1	3	2.9	2.7	2.6	2.5	2.5	2.5	1	2.6	1	3	2.3
T (°C)	35	131.7	450	1000	983.3	35	35	64.9	198.9	983.3	35	129.8	250
m (kg s^{-1})	50	50	50	15.1	15.1	12.6	2.5	104.0	66.3	37.4	228.6	137.2	174.6
Composition (% mole)													
H_2O	—	—	—	—	—	—	—	—	—	—	—	—	—
H_2	—	—	—	97.5	97.5	100	—	75	—	—	—	—	—
C(s)	—	—	—	—	—	—	—	—	—	100	—	—	—
CH_4	100	100	100	2.5	2.5	—	100	—	—	—	—	—	—
CO	—	—	—	—	—	—	—	—	—	—	—	—	100
CO_2	—	—	—	—	—	—	—	23	—	—	100	100	—
CH_3OH	—	—	—	—	—	—	—	—	100	—	—	—	—

[a] Solid carbon produced in the methane cracking reactor is reported in the CISOLID substream of Stream 5. The values shown in this table for Stream 5 reflect only the mixed substream (gas and liquid phases).

Table 4 Scaling parameters for the component purchase cost[a]

Component	Scaling factor	C_A (M€)	Q_A	f	Ref.
Reactors CO_2 utilisation	Outlet flow rate (tonne per h)	12.5	42	0.65	16
Reactor decomp	Inlet volume flow (cum per h)	2.7	37 000	0.65	21
Distillation column	Methanol flow rate (tonne per h)	18.9	162	0.7	16
Compressor	Power (MW)	0.44	0.41	0.67	17
Heat exchanger	Heat duty (MW)	6.1	828	0.67	17
PSA unit	Flow rate (kmol h^{-1})	34.3	17 000	0.6	22

[a] "M" denotes million.

The summation of all the individual equipment purchase costs will result in the total equipment cost (TEC) illustrated in eqn (8). The scaling factor (f), reference component cost (C_A), and capacity factor (Q_A) of different plant components are presented in Table 4.

$$\mathrm{TEC} = \sum_{i}^{n} C_{\mathrm{B,i}} \tag{6}$$

The total direct purchase cost (TDPC) is the total of TEC and the total installation cost (TIC), as shown in eqn (9).

$$\mathrm{TDPC} = \mathrm{TEC} + \mathrm{TIC} \tag{7}$$

The TPC is calculated from the summation of TDPC, engineering procurement and construction cost C_{EPC}, the contingencies cost (C_{CO}) and the owner's cost (C_{OC}), as demonstrated in eqn (10). The C_{EPC} is equal to 15% of the TDPC and C_{CO} the and the C_{OC} are respectively 10% and 5% of the TDPC and C_{EPC} summation.

$$\mathrm{TPC} = \mathrm{TDPC} + C_{\mathrm{EPC}} + C_{\mathrm{CO}} + C_{\mathrm{OC}} \tag{8}$$

Assumptions for the calculation of the TAC are illustrated in Table 5.

3.2. Environmental assessment indicators

From the environmental perspective, LCA determines the environmental impact of the process during its lifetime. This

Table 5 Assumptions for the economic analysis

Parameter	Value	Ref.
Installation cost as a fraction of total purchase cost (%)	400	23
Variable operating cost (VOM) as a fraction of total capital cost (%)	2.0	24
Fixed operating cost (FOM) as a fraction of total capital cost (%)	1.0	24
Plant lifetime (T) (years)	30	25
Project interest rate (r) (%)	12	24
Capacity factor (CF) (%)	85	26
Fuel price[a] (€ per GJ)	5	17
Average CO market price (€ per t)	50	27
Average CO_2 market price (€ per t)	30	28

[a] The heat and electricity prices for the economic assessment are assumed to be 55% and 400% of the NG price, respectively.

study determined the total life cycle of CO_2 emissions per kg of product ($kg_{CO_2}\ kg_{product}^{-1}$), methanol and CO. The scope is established as cradle-to-gate, and the functional unit (FU) is defined as 1 kg of methanol produced. TLCCE is calculated using eqn (9), which ignores CO_2 released from equipment and infrastructure construction.[29] The CO_2 emission factors are illustrated in Table 6, which includes resource extraction, transport, *etc.*

$$\mathrm{TLCCE} = \frac{\sum_{i=1}^{n}\mathrm{CE}_i + \sum_{j=1}^{n}\mathrm{CE}_j}{\text{total no. of h per year} \times \text{product (kg)}} \quad (9)$$

where CE_i and CE_j are the CO_2 emissions of i_{th} sub-process in the supply chain and j_{th} sub-process in CO_2 compression, transport, and storage. It is assumed that natural gas, after extraction and sweetening in Alberta, is transported *via* a high-pressure pipeline to a liquified natural gas (LNG) export terminal (Goldboro LNG) located in Nova Scotia (4600 km east of Alberta). At the terminal, natural gas is liquified by cooling it to −162 °C, reducing its volume for shipping. LNG is assumed to be shipped to South Hook terminals in Wales (4800 km directly across the Atlantic). It is believed that the plant will be 50 km from the port, and therefore, the natural gas, after evaporation, is transported through a low-pressure pipeline to the plant.

3.3. Key assumptions and practical considerations

While the Aspen Plus simulations in this study assume pure methane and carbon dioxide as feedstocks to establish an idealised performance baseline, the TLCCE is based on natural gas, not pure CH_4. As shown in Table 6, the environmental impact accounts for upstream emissions from extraction, sweetening, liquefaction, and natural gas transport, thereby reflecting the realistic burden associated with gas purification. In practical applications, natural gas may contain various impurities such as hydrogen sulfide (H_2S), N_2, and heavier hydrocarbons (C_{2+}). These components can introduce operational challenges, including catalyst poisoning, coke formation, or altered reaction kinetics. For instance, H_2S can poison metal catalysts used in methane cracking and methanol synthesis, while C_{2+} compounds can degrade at high temperatures and contribute to fouling or carbon deposition in reactors.[11] Likewise, CO_2 feedstocks, particularly those derived from flue gas, fermentation, or syngas streams, may contain trace amounts of oxygen, sulfur compounds (SO_2, H_2S), NO_x, or volatile organic compounds. These can affect the Boudouard and methanol synthesis steps by degrading catalysts or introducing corrosive species.[32] Commercial processes address these issues through established purification technologies, including amine scrubbing, PSA, and membrane separation, which can deliver high-purity CO_2 (>99%). While these units increase capital and operational expenditure, their impact is generally moderate and would not drastically alter the process efficiency or environmental trends demonstrated in this study. Future work could incorporate impurity sensitivity analyses and model the integration of gas purification systems. It is worth mentioning that we assumed 100% hydrogen conversion in the methanol reactor and complete carbon conversion in the CO reactor to establish a theoretical performance baseline. This allowed us to evaluate methanol's best-case energy efficiency and levelised cost. However, lower conversion rates in either reactor would lead to reduced product yield per unit of feedstock, increased recycle or purge requirements, and greater energy consumption per unit of methanol produced. These effects would collectively lower overall energy efficiency and increase LCOM. In this study, a single distillation column was modelled to separate methanol and water. This simplification is justified by the assumption of ideal methanol synthesis, where no byproducts (*e.g.*, DME, higher alcohols, or unreacted gases) are formed. While real-world methanol plants typically employ multiple distillation columns to handle such byproducts, this study focuses on early-stage conceptual analysis. Similar simplifications have been used in previous process modelling work[33] to isolate the core performance of the system. In industrial implementation, additional separation units would be necessary to meet product purity requirements and manage non-condensable species. Moreover, a $CO_2:H_2$ molar ratio of ~1 : 3 was assumed for simplicity, in line with previous conceptual modelling studies

Table 6 CO_2 Emission data for LCA

Parameter	Value	Unit	Ref.
Methane			
Extraction and drying	0.0023	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
Sweetening	0.0059	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
Liquefaction	0.0055	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
Transport	0.0068	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
Evaporation	0.0024	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
NG distribution at high pressure to the consumer	0.00041	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
NG distribution at low pressure to the consumer	0.00018	$kg_{CO_2}\ MJ_{CH_4}^{-1}$	29
Electricity			
Natural gas combined cycle	425	$g_{CO_2}\ kWh^{-1}$	30
Heat			
Natural gas boilers	293	$g_{CO_2}\ kWh^{-1}$	31

on CO_2 hydrogenation to methanol,[34] where side reactions and recycle complexity are not explicitly considered.

4. Results and discussion

The technical performance of the plant is presented in Table 7. It reveals that the proposed system's methanol and CO production rate is 66.4 kg s^{-1} and 174.6, respectively, resulting in a thermal output of 3151.2 MW_{th}. The power and heat requirement for plant operation is 144.8 MW_{el} (11.6 MW_{el} for methane compression, 19.3 MW_{el} for captured CO_2 compression and 113.8 MW_{el} for the mixture compression before the methanol reactor). On the other hand, the heat requirement of the plant is 879.2 MW_{th}. It is worth mentioning that the heat requirement for the methanol distillation, CO production and methane cracking process is 287.3 MW_{th}, 600 MW_{th} and 317.3 MW_{th}, respectively, while 325.3 MW_{th} is recovered from the methanol reactor. Finally, the calculated overall energy efficiency from the abovementioned parameters is 89.4%, while the process CO_2 emission is −228.6 kg s^{-1}, representing the total captured CO_2 flow rate needed for CO and methanol production. It is worth mentioning that there is no CO_2 emission from the process since all the carbon in the methane is used for CO production.

Table 8 provides a detailed overview of the capital investment, operational expenditures, and revenue streams associated with the CO and methanol co-production plant. The economic breakdown highlights the key cost drivers and performance indicators, offering insight into the plant's financial feasibility. The total plant cost (TPC) is estimated at €1489.3 million, encompassing significant investments in critical equipment and installation. Installation expenses dominate the capital outlay, with the total installation cost (TIC) amounting to €900.9 million. Among the components, the CO reactor (€72.6 million) and methanol reactor (€51.8 million) represent the most critical cost centres, reflecting their pivotal roles in product synthesis. Other necessary equipment includes the methane cracking reactor (€10.7 million) and the PSA unit (€41.1 million), essential for feedstock conversion and gas separation. The largest share of operational expenditure is attributed to fuel costs (€335.1 million per year), underscoring the energy-intensive nature of the process. Utility costs for electricity (€79.9 million per year) and heat supply (€58.3 million per year) further contribute to the plant's high energy demand. Additionally, the purchase of CO_2 feedstock accounts for €183.8 million annually. Operational and maintenance costs, including variable (€14.9 million per year) and fixed (€29.8 million per year) components, are relatively modest but critical for maintaining consistent plant performance. Revenue from CO sales is projected at €234.2 million per year, providing a significant offset to the plant's operating costs. However, with total annualised expenses reaching €651.8 million, the financial sustainability of the plant is highly dependent on optimising feedstock and utility costs. The levelised cost of methanol (LCOM) is calculated at €17.5 GJ^{-1}, serving as a critical metric for benchmarking the plant's production cost against market prices and alternative production methods.

Table 7 Technical performance summary

Parameter	Unit	Value
NG flow rate	[kg s^{-1}]	50
LHV_{NG}	[MJ kg^{-1}]	50
Thermal input	[MW_{th}]	2500
Methanol production	[kg s^{-1}]	66.4
$LHV_{Methanol}$	[MJ kg^{-1}]	20.9
CO production	[kg s^{-1}]	174.6
LHV_{CO}	[MJ kg^{-1}]	10.1
Thermal output	[MW_{th}]	3151.2
Heat required	[MW_{th}]	879.2
Power required	[MW_{el}]	144.8
Direct CO_2 emission	[kg s^{-1}]	−228.6
Overall energy efficiency	[%]	89.4

Table 8 Economic performance summary of the proposed plant

Parameters	Unit	Value
PSA	[M€]	41.1
Heat exchangers	[M€]	1.9
Compressors	[M€]	22.3
Methanol reactor	[M€]	51.8
Distillation unit	[M€]	24.8
CO reactor	[M€]	72.6
Methane cracking reactor	[M€]	10.7
Total equipment cost (TEC)	[M€]	225.2
Total installation cost (TIC)	[M€]	900.9
Total direct plant cost (TDPC)	[M€]	1126.1
Total plant cost (TPC)	[M€]	1489.3
Annualised plant cost	[M€ per year]	184.2
Fuel cost	[M€ per year]	335.1
Heat cost	[M€ per year]	58.3
Electricity cost	[M€ per year]	79.9
CO_2 purchase cost	[M€ per year]	183.8
CO selling revenue	[M€ per year]	234.2
Variable O&M	[M€ per year]	14.9
Fixed O&M	[M€ per year]	29.8
Total annualised cost	[M€ per year]	651.8
LCOM	[€ per GJ]	17.5

Below is a sensitivity analysis of the levelised cost of Methanol (LCOM, € per GJ) to variations in fuel price, carbon dioxide market price, and interest rate. Fig. 2 shows a strong linear relationship between the fuel price and the LCOM. As the fuel price increases from 3 € per GJ to 10 € per GJ, the LCOM rises proportionally from approximately 12.4 € per GJ to 30.1 € per GJ. This result highlights that fuel cost, a key operational expense, is a dominant factor influencing methanol production costs. Such sensitivity is particularly critical for feedstock-dependent production pathways, where fluctuations in fuel markets can substantially impact economic performance.

Fig. 3 illustrates the impact of carbon dioxide market price (€ per t) on the LCOM. A linear increase in LCOM is observed, ranging from approximately 17.5 € per GJ at a CO_2 price of 30 € per t to 61.8 € per GJ at 300 € per t. This sensitivity underscores the influence of CO_2 pricing policies, particularly for processes involving carbon capture and utilisation (CCU). The steep slope of the trendline indicates that high CO_2 market prices can

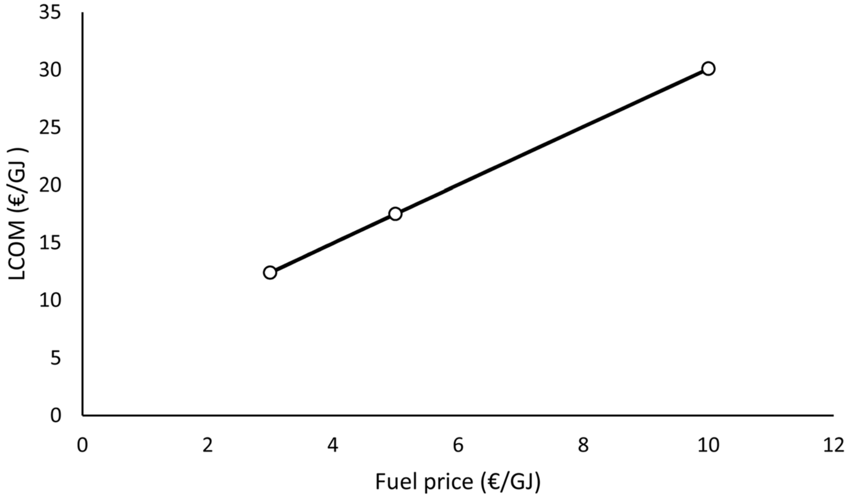

Fig. 2 Fuel price impact on the levelised cost of methanol production.

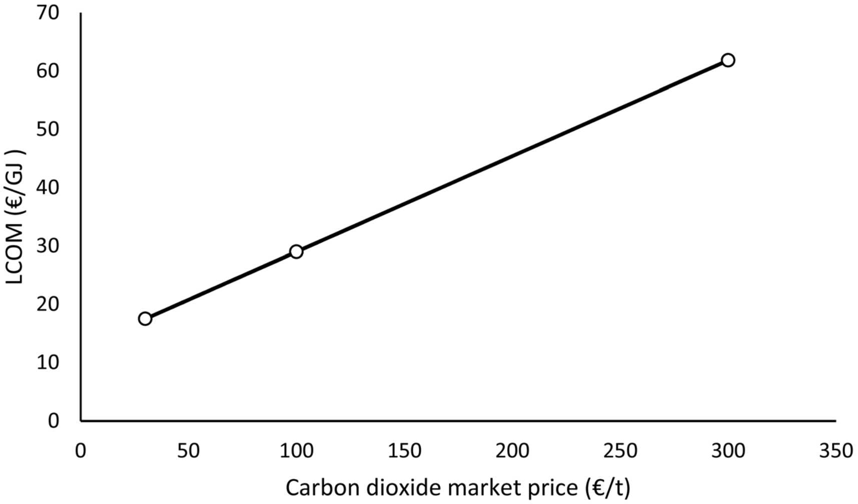

Fig. 3 Carbon dioxide market price on the levelised cost of methanol production.

significantly increase production costs, potentially affecting the competitiveness of methanol as a sustainable energy carrier.

Fig. 4 examines the sensitivity of LCOM to variations in interest rate, which applies to loans taken to finance the project. The results show a moderate, linear increase in LCOM as the interest rate rises from 3% to 15%, with LCOM increasing from approximately 13.8 € per GJ to 18.7 € per GJ. While less sensitive than fuel or CO_2 prices, this relationship reflects the influence of borrowing costs on overall production economics. For projects heavily reliant on external financing, higher interest rates can marginally increase LCOM, impacting financial viability. This effect is critical for investors and policymakers, especially in regions with varying lending rates.

The CO_2 emissions associated with the methanol and CO production were analysed and categorised into three key stages: (1) methane extraction, drying and sweetening; (2) methane transportation; and (3) plant energy consumption. The results (Fig. 5), expressed as $kg_{CO_2}\ kg_{product}{}^{-1}$), highlight the primary emission sources in the process chain. Methane extraction, drying, and sweetening contribute 0.09 $kg_{CO_2}\ kg_{product}{}^{-1}$. This stage involves upstream processing of natural gas, including removing impurities such as water, sulfur compounds, and CO_2, and accounts for emissions generated from energy-intensive operations and chemical treatments. Methane transportation results in 0.08 $kg_{CO_2}\ kg_{product}{}^{-1}$, encompassing the emissions associated with delivering methane to the production plant. These emissions are attributed to energy consumption during pipeline compression, liquefaction, or other logistics-related activities. Energy consumption within the production plant is the most significant contributor, with emissions reaching 0.3 $kg_{CO_2}\ kg_{product}{}^{-1}$. This reflects the reliance on carbon-intensive energy sources for process operations such as heating, compression, and reactor performance. It is worth mentioning that a TLCCE of 0.5 $kg_{CO_2}\ kg_{product}{}^{-1}$ does not reflect the complete LCA of the process, as emissions related to plant construction were not considered. The positive effect of

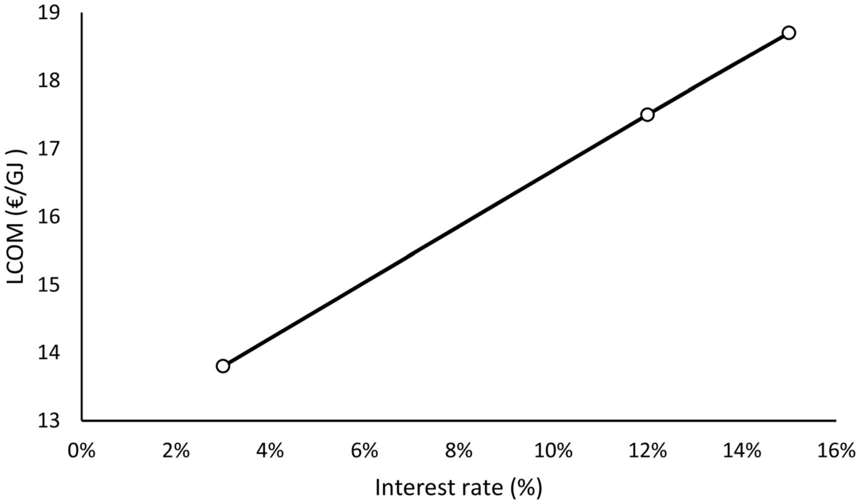

Fig. 4 Interest rate impact on levelised cost of methanol production.

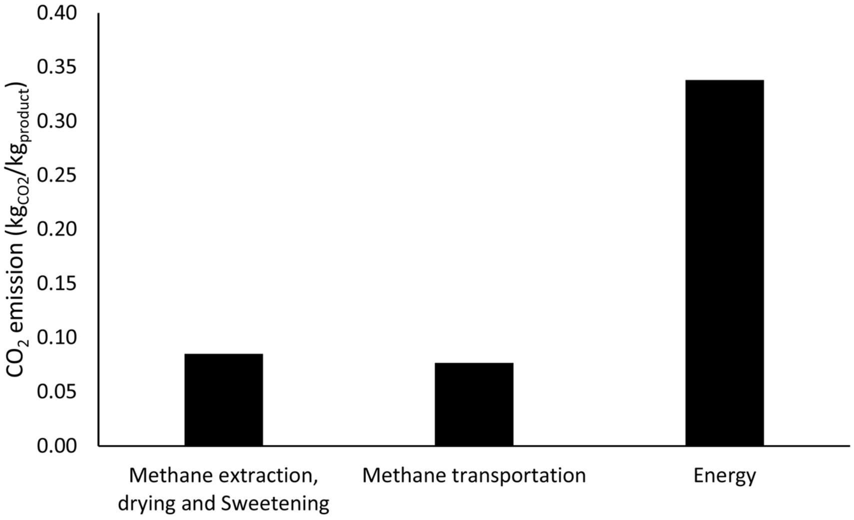

Fig. 5 CO_2 emission of the proposed process.

utilising captured CO_2 was not factored into the LCA calculation. By preventing CO_2 from being released into the atmosphere, this approach effectively turns a waste product into a valuable feedstock, supporting a circular carbon economy.

The economic and environmental implications of the evaluated plant, which operates at 17.5 € per $GJ_{methanol}$ (367 € per t_{MeOH}) and emits 0.5 kg_{CO_2} $kg_{product}^{-1}$, can be critically assessed against recent advancements in production technologies. A recent study by Peng *et al.*[35] evaluated the renewable methanol production from municipal solid waste and achieved 495–511 € per t_{MeOH} and a CO_2 emission of 0.8 kg_{CO_2} kg_{MeOH}^{-1}. Furthermore, methanol production cost with natural gas-based state-of-the-art technology was reported to be 268.5 € per t_{MeOH}.[36] Moreover, the TLCCE of methanol production through steam methane reforming and partial oxidation is reported to be 0.94 kg_{CO_2} $kg_{CH_3OH}^{-1}$ and 0.81 kg_{CO_2} $kg_{CH_3OH}^{-1}$, respectively.[37,38] It is found that the proposed cycle is competitive with the state-of-the-art methanol production plant.

It is worth mentioning that we assumed 100% hydrogen conversion in the methanol reactor and complete carbon conversion in the CO reactor to establish a theoretical performance baseline. This allowed us to evaluate methanol's best-case energy efficiency and levelised cost. However, Lower conversion rates in either reactor would lead to reduced product yield per unit of feedstock, increased recycle or purge requirements, and greater energy consumption per unit of methanol produced. These effects would collectively lower overall energy efficiency and increase LCOM. Based on the economic structure presented in this study, we estimate that the process would become economically uncompetitive if effective hydrogen conversion drops below approximately 75%, assuming no additional recycling or heat recovery measures are implemented. Below this threshold, the reduced methanol output and higher specific energy and capital costs would likely increase LCOM above 24 € per GJ, which exceeds typical methanol market prices.

5. Conclusion

This study developed and analysed a novel process for the co-production of methanol and carbon monoxide *via* methane decomposition, focusing on energy efficiency, economic feasibility, and environmental impact. The TLCCE was calculated to be 0.5 $\mathrm{kg_{CO_2}\ kg_{product}}^{-1}$, with methane extraction contributing 0.09 $\mathrm{kg_{CO_2}\ kg_{product}}^{-1}$, methane transportation accounting for 0.08 $\mathrm{kg_{CO_2}\ kg_{product}}^{-1}$, and the remaining emissions attributed to energy supply. The process demonstrated a high overall energy efficiency of 89.4%, supporting its potential for sustainable methanol and CO production. Economic analysis revealed that the total plant cost amounts to 1489.3 M€, with fuel costs being the most significant annual expense at 335.1 M€ per year. The total annualised cost was determined to be 651.8 M€, while revenue from CO sales reached 234.2 M€ per year. These findings indicate that methane decomposition-based methanol and CO co-production present a promising low-carbon alternative to conventional production pathways, offering competitive energy efficiency and reduced emissions. However, process economics remain sensitive to fuel prices and CO market value, necessitating further optimisation and potential integration with renewable energy sources to enhance long-term viability.

Data availability

All the original data generated in this study are included in the manuscript.

Conflicts of interest

There are no conflicts to declare.

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