

Switchable catalysis for methanol and synthetic natural gas synthesis from CO₂: A techno-economic investigation

Loukia-Pantzechroula Merkouri^{a,*}, Jayson Mathew^a, Jerin Jacob^a, Tomás Ramirez Reina^b, Melis S. Duyar^{a,*}

^a School of Chemistry and Chemical Engineering, University of Surrey, Guildford GU2 7XH, United Kingdom

^b Department of Inorganic Chemistry and Materials Sciences Institute, University of Seville-CSIC, 41092 Seville, Spain

ARTICLE INFO

Keywords:

Flared gas upgrading
Switchable catalysis
CO₂ utilisation
Techno-economic analysis
Methanol synthesis
Synthetic natural gas

ABSTRACT

The oil and gas sector produces a considerable volume of greenhouse gas emissions, mainly generated from flaring and venting natural gas. Herein, a techno-economic analysis has been performed of a switchable catalytic process to convert the CH₄ and CO₂ in flared/vented natural gas into syngas or methanol. Specifically, it was shown that depending on greenhouse gas composition, dry methane reforming (DRM), reverse water-gas shift (RWGS), and CO₂ methanation could be chosen to valorise emissions in an overall profitable and flexible operation scenario. The switchable process produced methanol and synthetic natural gas as its products, resulting in an annual income of €687m and annual operating expenses of €452m. The pre-tax profit was calculated at €234m, and at the end of the project, the net present value was calculated as €1.9b with a profitability index of 4.7€/€. The expected payback time of this process was ca. 4 years, and with a 35% internal rate of return (IRR). Most importantly, this process consumed 42.8m tonnes of CO₂ annually. The sensitivity analysis revealed that variations in operation time, green hydrogen price, and products' prices significantly impacted the profitability of the process. Overall, this techno-economic analysis demonstrated that switchable catalysis in greenhouse gas utilisation processes is profitable, and thus it could play an important role in achieving net zero emissions.

Nomenclature Symbol Explanation, Units

C	Cost (subscript denotes the source), €.
CAPEX	Capital expenses, €.
DCF	Discounted cash flow, -
IRR	Internal rate of return, %.
I _t	Cash inflows, €.
n _{Labour}	Number of employees, -
n _{project}	Project's lifetime, years.
NPV	Net present value, €.
O _t	Cash outflows, €.
PI	Profitability index, €/€.
PT	Payback time, years.
Q _p	Quantity of product produced (methanol or synthetic natural gas) tonne/year or MWh/year.
Q _u	Quantity of utilities used (subscript denotes the source) MWh for electricity and tonnes or kg for others.

R	Revenue (subscript denotes the source), €.
r _d	Discount rate parameter, %.
r _{workingcapital}	Working capital, %.

1. Introduction

In recent years, there has been a significant rise in greenhouse gas (GHG) emissions due to anthropogenic activities. Carbon dioxide (CO₂), the main GHG in the atmosphere that heavily contributes to global climate change, has dramatically increased in the atmosphere due to the combustion of fossil fuels for energy, transportation, and several industrial processes [1,2]. In addition to CO₂, there are other GHG that should also be abated, for instance, CH₄ whose global warming potential is 25 times higher than that of CO₂ within a 100-year time frame [3]. At the end of 2022, the concentrations of the CO₂ and CH₄ in the atmosphere were reported to be 419 ppm and 1923 ppb, respectively [4,5]. An appealing method to limit the CO₂ release includes carbon capture

* Corresponding authors.

E-mail addresses: l.merkouri@surrey.ac.uk (L.-P. Merkouri), m.duyar@surrey.ac.uk (M.S. Duyar).

<https://doi.org/10.1016/j.jcou.2023.102652>

Received 14 September 2023; Received in revised form 28 November 2023; Accepted 18 December 2023

Available online 26 December 2023

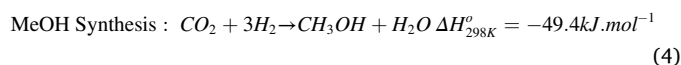
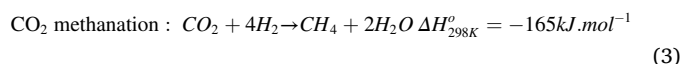
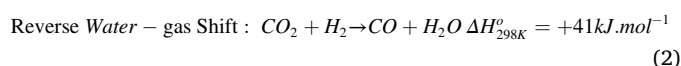
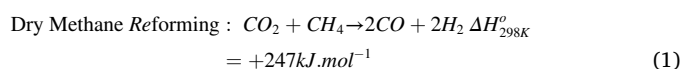
2212-9820/© 2023 The Authors. Published by Elsevier Ltd. This is an open access article under the CC BY license (<http://creativecommons.org/licenses/by/4.0/>).

and storage (CCS) and carbon capture and utilisation (CCU) [6,7]. Methane, on the other hand, is released through the venting of natural gas or it is combusted (flaring) in oil and gas refineries [8]. Approximately 140 billion cubic metres of natural gas streams are flared globally per year, resulting in nearly 300 million tonnes of CO₂ and 16 billion USD worth of gas discharged into the environment [8–10]. Flared gas mostly consists of hydrocarbons, such as methane [11]. There are two types of flared gas, associated and non-associated. The former refers to the flaring of gas that was originally dissolved in the oil in the underground formations and is then released from the upstream oil facilities. The latter refers to the gas flaring from refineries and petrochemical industries during routine operations and for safety purposes. Associated gas flaring accounts for 90% of the global gas flare, with Russia, Iraq, and Iran being the countries with the highest gas flaring volume. Its most popular uses include gas reinjection (i.e. enhanced oil recovery), liquefied natural gas, liquified petroleum gas, and compressed natural gas [11]. To date, various methods have been proposed to reduce and recover flaring gases, such as gas-to-liquid (GTL) production and electricity generation with gas turbines. In particular, flare gas conversion into liquid products is appealing due to their easier transportation from the stranded locations in which the majority of gas flaring occurs. As a result, it can technically be converted into syngas, hydrogen, hydrocarbons, methanol, and higher alcohols via a plethora of catalytic reactions [11,12].

A switchable catalyst is defined as one that is able to catalyse multiple reactions only when the reaction conditions are changed. It achieves high conversion and selectivity over a particular product and once the reaction atmosphere is altered, it exhibits high conversion and selectivity over another product. This flexibility in chemical production is important when dealing with supply and demand issues, for example, when pandemics and other crises occur. In addition, switchable catalysis fits well when there are seasonal discrepancies in the supply and demand of chemicals and fuels [13]. For instance, the daily natural gas demand in the US drops by 33% between summer and winter [14], prompting us to further explore the usage of switchable catalysis to manage these mismatches. When considering the CO₂ reduction reactions of dry reforming of methane (DRM), reverse water-gas shift (RWGS), and CO₂ methanation (Eqs. 1–3), it is possible to use a ‘switchable catalyst’ that is active for all three reactions under a distinct set of reaction conditions (temperature and co-reactant). Therefore, the reactor could operate the CO₂ methanation reaction, for example, during wintertime when the demand for natural gas is increased due to heating purposes. During summertime, the reactor temperature could increase, promoting the RWGS reaction and the formation of syngas which can be further upgraded or even combusted for power purposes. Furthermore, the price of green H₂ is currently as high as ca. USD 7.5 per kg, but it is expected to decrease in the future, motivating us to use either co-reactants, such as CH₄, in the near-term future in order to have a more profitable process [13,15]. This flexibility in chemicals and fuels synthesis is helpful in dealing with flare gas as its production is not continuous because it is affected by safety, regulatory, and technical reasons at any given time [11]. As a result, a switchable catalyst can be used for multiple reactions, depending on the availability of flare gas and its quantity produced, factors that have yet to be considered in other valorisation scenarios [11,12,16–19].

Some research studies have been conducted on switchable catalysts, which can potentially open up opportunities for major CO₂ emitters, such as refineries where CO₂ emissions are a major problem, to apply CCUS technologies [20–22]. Le Saché et al. [21] demonstrated that a Ru-promoted Ni-based catalyst remarkably enhanced the CO₂ conversion and selectivity and was able to switch between the CO₂ methanation and RWGS reactions in multiple cycles. Most recently, Merkouri et al. [20] conducted studies on versatile Ni-Ru catalysts that can catalyse DRM, RWGS, and CO₂ methanation, showing stable CO₂ conversion,

even when the operating conditions changed. Due to the fact that CO₂ has high stability, with its Gibbs free energy being at –394 kJ/mol, careful catalyst design needs to be employed to enhance the reaction kinetics [23]. Ni-based catalysts are promising candidates due to their high activity and low cost in the DRM [23–26], RWGS [27–30], and CO₂ methanation reactions [31–34]. Even though noble metal-based catalysts perform better, their high costs hamper their industrial application [35,36]. In more detail, CO₂ methanation (Eq. 3), which is useful in large-scale renewable energy storage through Power-to-Gas schemes, is an exothermic reaction that produces synthetic natural gas (SNG) [31, 32,37]. RWGS (Eq. 2) is an endothermic reaction and its main product, CO, is considered a versatile intermediate for a plethora of chemicals [29,30,38]. DRM (Eq. 1) is also an endothermic reaction during which syngas is produced with an H₂:CO ratio of 1 [23,36,39]. A switchable catalyst can be employed to switch between these three reactions because they are favourable at different reaction atmospheres. CO₂ methanation needs low temperatures and H₂ as a co-reactant, RWGS needs high temperatures and H₂ as a co-reactant, and DRM needs high temperatures and CH₄ as a co-reactant [13]. The products of RWGS and DRM can be further upgraded to produce higher hydrocarbons and/or alcohols. Among those, methanol is considered a promising chemical for it can be used as a transportation fuel and fuel-cell H₂ carrier, or to produce electricity, formaldehyde, and acetic acid [40,41]. The MeOH synthesis from syngas proceeds through three primary successive reactions: CO₂ hydrogenation, CO hydrogenation, and RWGS [37,42,43], and it is currently carried out at 50–100 bar and 220–240 °C by using a ternary Cu-Zn-Al oxide catalyst [44]. Consequently, the economic feasibility of these switchable catalysts can be tested in the flare gas/CO₂ emissions utilisation to produce on-demand chemicals, by simulating them as three different flow schemes to depict the three possible operation scenarios of the designed plant.



Several techno-economic studies have been conducted for either the production of MeOH or SNG. Relating to the MeOH production, some studies have shown that it is financially feasible to produce green MeOH, whereas some others have shown that it is not [45–50]. The high costs of MeOH production are associated with the costs of the water electrolyser for green H₂ generation, and the high reaction pressure needed to produce MeOH. In addition, the price of MeOH is a crucial parameter for the process profitability, which has varied a lot over the past years (392 \$/metric ton in Europe in June 2022) [51]. Concerning the SNG production, many studies have concluded that this process is economically feasible, especially when carbon tax is included, with the biggest cost contributor being the green H₂ production [52–57]. In order to deal with the high green H₂ costs, it has been proposed that SNG production should be adapted to long seasonal patterns, e.g. during the summer to allow maximum renewable energy production [56,58]. However, green H₂ costs are expected to decrease in the coming years [59]. It should be noted that the price of natural gas was 123\$/MWh in Europe in December 2022, but this price has fluctuated wildly over the past few years in this region [60]. The reported results of several techno-economic investigations of the DRM-based and RWGS-based processes in literature differ. However, all the studies agree that

maintaining a high reaction temperature is the biggest cost contributor [46,61–68]. Overall, it should be stated that the carbon tax is a vital incentive for both the existing and the new processes to become greener. The carbon tax has increased over the past years, reaching €100 a tonne in Europe at the end of February 2023 [69].

To date, there are no techno-economic investigations into the use of switchable catalysts to produce valuable chemicals and fuels on an industrial scale. Herein, we explore this scenario as a potential solution to valorising flared gas/CO₂ produce MeOH and SNG and investigate its carbon footprint and profitability. More specifically, we explore the case of a switchable catalytic reactor to upgrade the CO₂ in the flared gas to MeOH via syngas produced by DRM and RWGS reactions, and to SNG produced by CO₂ methanation reaction. Our techno-economic analysis evaluated the profitability of a plant located in Europe, which had a MeOH production rate of ca. 361.4k tonnes/year and a SNG production rate of ca. 404.1k tonnes/year. Overall, it was shown that it made sense from an economic perspective to use one reactor for the various CO₂ utilisation routes, paving the way for a more detailed exploration of switchable catalysis. In this way, chemicals and fuels can be produced on-demand, depending on the availability of flare gas without the need of shut-down or standby interruptions, taking into account seasonal needs and market fluctuations as well.

2. Methods

2.1. Process description

Fig. 1 shows the process of obtaining SNG and MeOH. The switchable aspect of the process was achieved by simulating the DRM, RWGS, and CO₂ methanation reactions separately. So, the overall process was split into three separate flow schemes in Aspen HYSYS. The DRM and RWGS flow schemes aimed at producing syngas, which was then utilised to produce methanol. The CO₂ methanation flow scheme was used to produce SNG. Those three different flow schemes showcased the three different scenarios in which the switchable catalyst could be utilised. The detailed process description can be found in the [Supporting Information](#).

In brief, the switchable reactor operated at 350 °C during CO₂ methanation and at 700 °C during RWGS and DRM reactions. The corresponding reactants entered the reactor and were converted into CH₄ and syngas, while the CO₂ conversion values were obtained experimentally [20]. Afterwards, the gaseous mixture was cooled down so as to remove the excess water formed during the CO₂ methanation and RWGS. The exhaust gases then entered a 6-stage compressor-intercooling step in order to increase their pressure. In the case of the CO₂ methanation, the outlet mixture after compression achieved the typical pipeline natural gas specification, and thus it was assumed that this mixture could enter the natural gas grid (see [Table S2](#)) [70–72]. In the cases of the RWGS and DRM, the outlet mixture entered the MeOH synthesis reactor, followed by a vapour/liquid separator for

the separation of H₂O and MeOH from the unreacted gases. These unreacted gases were mostly recycled back to the MeOH reactor, but a small purge stream was combusted to produce the required heat for the RWGS and DRM reactors. A distillation column was used for the separation of H₂O and MeOH.

2.2. Process simulation

The process was simulated on Aspen HYSYS V11, using the Soave-Redlich-Kwong (SRK) thermodynamic package. The SRK thermodynamic package was selected as the thermodynamic calculation method for this simulation because of its suitability for modelling gases and hydrocarbons produced at high temperatures and pressures (>10 bar). Moreover, the SRK made better predictions for the polar systems and it was a better thermodynamic model for the MeOH systems [19,40,73,74].

It was assumed that the switchable catalytic reactor and a downstream methanol synthesis process would be implemented in a natural gas production facility with some level of stranded hydrocarbon and hydrogen resources. Since those resources are too costly to purify, they could instead be used to reduce CO₂ to value-added products. The aim was to convert various sources of emissions (flared gas, vented natural gas, captured CO₂) into syngas or SNG. Hence, in this scenario, it was assumed that various levels of vented natural gas (rich in CO₂, but containing hydrocarbons), flared natural gas (rich in hydrocarbons) and concentrated CO₂ emissions (captured as part of standard amine scrubbing during natural gas production) were present on site, summarised in [Table 1](#). Information for the mass balances and conditions can be found in the SI ([Tables S4-S9](#)). The flow rates are given in the conditions specified in the table. It is worth mentioning that the total volumetric flow rates of the inlet streams of the switchable catalytic reactor were similar in all three CO₂ valorisation scenarios, i.e. approximately 294,500 m³/hr, meaning that the space velocity remained the same (see [Supporting Information](#)).

The overall process was simulated as three separate flow schemes to account for the three separate reactions catalysed by the switchable catalyst ([Fig. 1](#)). Both switchable catalyst and MeOH reactors were modelled as stoichiometric reactors. For both reactors, an energy stream was specified to ensure that the reactors operated isothermally, i.e. either at 350 °C or at 700 °C, depending on the reaction occurring in the switchable reactor, and at 240 °C in the MeOH reactor. The distillation columns were initially simulated as ‘shortcut’ columns. Once the product specification was reached, the ‘shortcut’ columns were changed into rigorous columns and were later optimised in terms of number of trays by using the temperature profile to ensure that they are redundant trays in the column (as described in the SI). The furnace was modelled as a Gibbs reactor and the components entering it were mainly CH₄, CO, and some CO₂. An additional air stream was specified to enable the combustion reaction to occur in the Gibbs reactor. More information about the process simulation can be seen in the [Supporting Information](#).

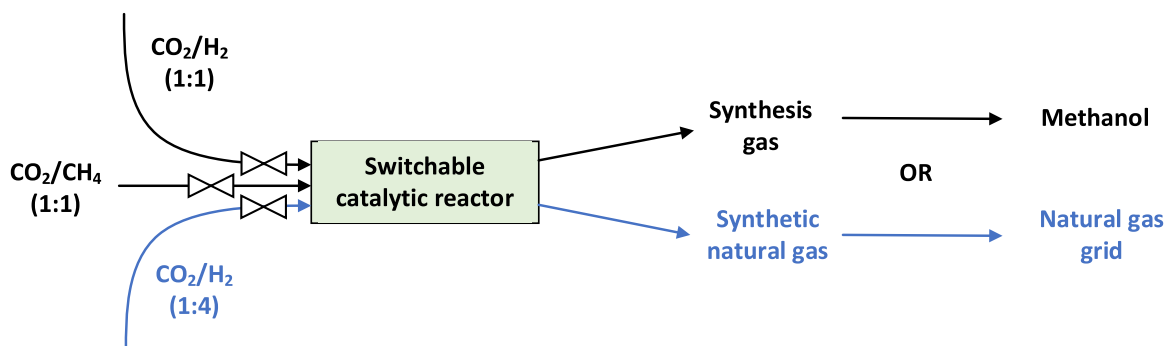


Fig. 1. : Illustration of the switchable catalytic process.

Table 1
Inlet specifications for the three flow schemes.

	DRM flow scheme	RWGS flow scheme	CO ₂ methanation flow scheme
CO ₂ flow rate (kmol/h)	5000	4500	1915
H ₂ flow rate (kmol/h)	-	5068	5650
CH ₄ flow rate (kmol/h)	5000	-	7656
Molar composition CO ₂ /H ₂ /CH ₄	0.5/0/0.5	0.47/0.53/0	0.13/0.37/0.5
Temperature (°C)	31.35	78.33	34.19
Pressure (bar)	3.05	2.78	3.05
Sources of inlet gas streams	Flared gas/vented natural gas	Captured CO ₂ /Green H ₂	Flared gas/Vented gas/Green H ₂

2.3. Economic model

In order to estimate the total capital costs, the equipment costs were calculated, using a combination of equipment costing equations and sizing procedures. The equations used to calculate the capital cost of each equipment and the detailed assumptions for the economic model can be found in the SI. The cost correlations for the compressors, pumps, reactors, distillation column, furnace, and separators were obtained from Towler & Sinnott, whereas the correlation for the heat exchangers was obtained from Seider *et al.* [75–77].

In the equipment purchase cost calculations, power and heat duty were the sizing parameters for the reciprocating compressors and heat exchangers, respectively. Volume was the sizing parameter for the reactors. Likewise, shell mass was needed to calculate the purchase cost for the distillation column and vapour/liquid separators. Those sizing parameters were calculated three times since the overall process was

Table 2
Input data used for technoeconomic analysis.

Data	Value	Reference
Price for MeOH, $P_{Methanol}$ (€/t)	370	[51]
Price for synthetic natural gas, $P_{synthetic\ natural\ gas}$ (€/MWh)	116	[56,80]
Price for electricity, $P_{electricity}$ (€/kWh)	0.1445	[81]
Price for cooling water, $P_{cooling\ water}$ (€/t)	0.0250	[61]
Price for switchable catalyst, $P_{switchable\ catalyst}$ (€/kg)	269	See Supporting Information
Price for methanol catalyst, $P_{methanol\ catalyst}$ (€/kg)	38	See Supporting Information
Price for catalyst charge/recharge, $P_{catalyst\ charge/recharge}$ (€/m ³)	2	[61]
Price for wastewater processing, $P_{wastewater\ processing}$ (€/m ³)	17	[61]
Price of hydrogen, $P_{hydrogen}$ (€/kg)	1.739	Calculated based on [78]
Price for maintenance, $P_{maintenance}$ (€/year)	5% of Annualised Capital Cost	[77]
Price of labour per employee, P_{labour} (€/person/year)	50,000	[82]
Price for insurance, $P_{insurance}$ (€/year)	1% of Annualised Capital Cost	[77]
Price for local taxes, $P_{local\ taxes}$ (€/year)	2% of Annualised Capital Cost	[77]
Price for royalties, $P_{royalties}$ (€/year)	1% of Annualised Capital Cost	[77]
Number of employees	10	[77]
Discount rate parameter, r_d (%)	4.75	[77]
Project lifespan (in years)	20	[77]
CAPEX (€)	397,921,699.59	See Supporting Information
Working capital, $r_{working\ capital}$ (%)	7.5% of total CAPEX	Assumed based on [77]
Chemical Engineering Plant Cost Index for 2010, CEPCI	550.8	[83]
Chemical Engineering Plant Cost Index for 2021, CEPCI	776.3	[83]

simulated in three separate flow schemes. For each equipment, the flow scheme which gave the highest value for its sizing parameter was selected to determine its purchase cost. That was because one plant was to be built, despite having three simulation schemes to model the part of the switchable catalysis. All the equipment was oversized so that it could function successfully when operating the plant for the production of either MeOH or SNG. The capital cost obtained from the aforementioned equations was for the year 2010, and thus the capital cost for 2021 was obtained using the Chemical Engineering Plant Cost Index (CEPCI), according to Eq. 5. To calculate the total investment, working capital was estimated to be a percentage of the fixed capital investment; 10% to 20% is the typical range [76]. Therefore, the working capital was presumed as the 7.5% of the fixed capital investment (Eq. 6). The analysis was carried out in euros for the year 2021.

The economic analysis was based on a 5,600-hour operation to produce MeOH, i.e. 2800 h of syngas produced through DRM plus 2800 h of syngas produced through RWGS, and 2800 h of operation to produce SNG. These operational hours were selected in order the switchable catalytic reactor to operate one reaction every four months. Therefore, the total operation time was 8400 h per year, and the remaining hours of the year (360 h) were attributed to the start-up, shut down, and maintenance procedures. This was the base case scenario. Heat recovery was achieved through combustion of the unreacted CH₄ and CO, using a furnace. The heat released was able to meet the majority of the heat demand of the switchable catalytic reactor when the DRM and RWGS reactions were in operation. This decreased operating costs and increased revenue. A sensitivity analysis was carried out to understand the effect of the operation time coupled with the green H₂ and products prices on the profitability of the designed process. The price of the green H₂ was assumed to be similar to that of the grey hydrogen in order to depict a future optimistic scenario, since its price is expected to drop within the next few years due to technological advances [78,79].

The economic model was based on the discounted cash flow method. This method included calculating the internal rate of return (IRR), payback time (PT), profitability index (PI), and net present value (NPV). The NPV and PI were calculated using Eqs. 7 and 8, respectively. The internal rate of return was calculated based on Eq. 9. Eqs. 10 and 11 were used to calculate the cash inflow and outflow. The project's lifespan was set at 20 years and the discount rate parameter was used to consider the time effect. The required loan for the investment was presumed to be the total capital cost, and the working capital was presumed to be 7.5%.

The revenue was generated from selling MeOH and SNG. Their total annual revenue was calculated by multiplying their prices by their annual production (using Eqs. 13 and 14). To calculate the revenue from the MeOH, its hourly mass flow rate obtained by the DRM flow scheme (leaving the top of the distillation column) was multiplied by 2800 h and the price of MeOH. Since the RWGS flow scheme also produced MeOH, the same procedure was followed in order to get the revenue from selling the MeOH when the switchable reactor operated the RWGS reaction. To calculate the revenue from the SNG, its lower heating value was multiplied by the mass flowrate of the CH₄ so as to obtain the total heat value per hour. The total hourly heat value was multiplied by 2800 h to calculate the total heat value of the stream per year and then, by the

price of SNG so as to measure its total revenue.

The cash outflows were calculated by grouping: $C_{Electricity}$, obtained from the price and quantity of electricity; $C_{Coolingwater}$, obtained from the price and quantity of cooling water; $C_{Switchablecatalyst}$, obtained from the price and quantity of the switchable catalyst; $C_{Methanolcatalyst}$, obtained from the price and quantity of the MeOH catalyst; $C_{Catalystcharge/recharge}$, obtained from the price for charging/recharging the catalyst and quantity of switchable and MeOH catalysts; $C_{wastewaterprocessing}$, obtained from the price and quantity of treating waste water; $C_{Maintenance}$, calculated as 5% of annualised CAPEX; C_{Labour} , obtained from the number of employees and their annual salaries; $C_{Insurance}$, calculated as 1% of the annualised CAPEX; $C_{Localtaxes}$, calculated as 2% of the annualised CAPEX; $C_{Royalties}$, calculated as 1% of the annualised CAPEX. Labour costs were obtained from using three operators per shift for 3 shifts per day, for a total of 10 operators. All those calculations were made using Eq. 15 to Eq. 26, respectively. The prices used for the operating costs were based on those of 2021. More details about the economic model used, the assumptions, and the calculations can be found in the SI. The input values for the TEA analysis are shown in Table 2.

$$\begin{aligned} \text{Capital cost year 2021} &= \text{Capital cost year 2010} \\ &\quad * \left(\frac{\text{CEPCI for year 2021 [776.3]}}{\text{CEPCI for year 2010 [550.8]}} \right) \end{aligned} \quad (5)$$

$$\text{Total investment} = \text{Total capital cost} + \text{Working capital} \quad (6)$$

$$NPV = \sum_{t=0}^n \frac{I_t - O_t}{(1 + r_d)^t} \quad (7)$$

$$PI = \frac{\sum_{t=0}^n \frac{I_t - O_t}{(1 + r_d)^t}}{C_{investment}} \quad (8)$$

$$IRR : NPV = \sum_{t=0}^n \frac{I_t - O_t}{(1 + r_d)^t} = 0 \quad (9)$$

$$I_t = R_{Methanol} + R_{Syntheticnaturalgas} \quad (10)$$

$$\begin{aligned} O_t &= C_{Electricity} + C_{Hydrogen} + C_{wastewaterprocessing} + C_{Maintenance} + C_{Labour} + C_{Royalties} \\ &\quad + C_{Localtaxes} + C_{Insurance} + C_{Switchablecatalyst} + C_{Methanolcatalyst} \\ &\quad + C_{Catalystcharge/recharge} \end{aligned} \quad (11)$$

$$C_{investment} = \frac{CAPEX * (100 + r_{workingcapital})}{100} \quad (12)$$

$$R_{Methanol} = Q_{Methanol} * P_{Methanol} \quad (13)$$

$$R_{Syntheticnaturalgas} = Q_{Syntheticnaturalgas} * P_{Syntheticnaturalgas} \quad (14)$$

$$C_{Electricity} = Q_{Electricity} * P_{Electricity} \quad (15)$$

$$C_{Hydrogen} = Q_{Hydrogen} * P_{Hydrogen} \quad (16)$$

$$C_{Coolingwater} = Q_{Coolingwater} * P_{Coolingwater} \quad (17)$$

$$C_{Switchablecatalyst} = Q_{Switchablecatalyst} * P_{Switchablecatalyst} \quad (18)$$

$$C_{Methanolcatalyst} = Q_{Methanolcatalyst} * P_{Methanolcatalyst} \quad (19)$$

$$C_{wastewaterprocessing} = Q_{wastewaterprocessing} * P_{wastewaterprocessing} \quad (20)$$

$$C_{Catalystcharge/recharge} = Q_{(Switchablecatalyst+Methanol)} * P_{Catalystcharge/recharge} \quad (21)$$

$$C_{Maintenance} = \frac{0.05 * CAPEX}{n_{project}} \quad (22)$$

$$C_{Labour} = n_{Labour} * P_{Labour} \quad (23)$$

$$C_{Insurance} = \frac{0.01 * CAPEX}{n_{project}} \quad (24)$$

$$C_{Localtaxes} = \frac{0.02 * CAPEX}{n_{project}} \quad (25)$$

$$C_{Royalties} = \frac{0.01 * CAPEX}{n_{project}} \quad (26)$$

3. Results and Discussion

3.1. CO₂ Consumption

The main aim of this work was the utilisation of CO₂, which meant that the designed process should consume more CO₂ than it generated. According to the simulation, the process chemistry consumed 5149 t/hr of CO₂ equivalent, resulting in an annual consumption of 43 million tonnes (see Supporting Information). However, in order to calculate the net CO₂ consumption, the carbon content of electricity needed to be considered. The electricity production in 2021 was reported to have had an average carbon footprint of 0.2123 kg CO₂ per kWh in the UK [84]. Therefore, 448k tonnes of CO₂ were released per year through the electricity used in this process. This meant that the overall annual consumption of CO₂ was 42.8m tonnes for 8400 h of plant operation time as opposed to the conventional ways of producing MeOH and natural gas, which are carbon positive technologies. Overall, the proposed process consumed more CO₂ than it emitted, which was the prime objective of this project. However, it is essential to note that the calculated value of the annual CO₂ equivalent consumption is only an approximation and that a more detailed life cycle assessment of the process is required to get a better estimate.

3.2. Techno-economic analysis

The annual income for this project was generated from producing 361.4m kg/hr of MeOH and 1.703m kW of SNG. The total income was €687m and the revenue breakdown can be seen in Fig. 2. SNG, which was produced in large quantities from the CO₂ methanation flow scheme, contributed more to the annual revenue, ca. 80% of the total revenue per year. Despite the fact that MeOH had a higher selling price

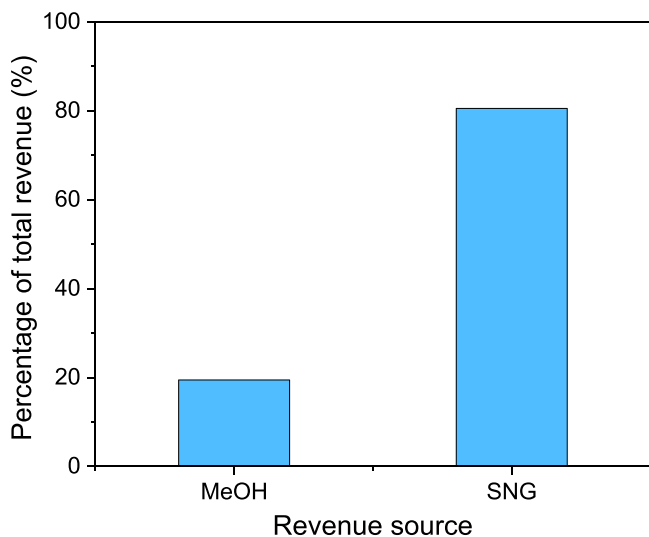


Fig. 2. : Breakdown of the total revenue from various revenue sources.

and was produced using both the DRM and RWGS flow schemes, its sales contributed only 20% to the total annual revenue. That was because the total MeOH produced was 722.8k tonnes per year and the SNG produced was 441.8m kg per year, but by multiplying the SNG amount by its heating value and 2800 h of production, 4.8m MWh were obtained. It is worth noting that this difference in production was due to the lower single pass conversion in the MeOH reactor compared to the SNG reactor. The DRM flow scheme produced more MeOH than the RWGS flow scheme, contributing six times more to the revenue obtained from the MeOH production, mainly attributed to the higher amounts of H₂ and, to a lesser extent, to the available CO₂ downstream the switchable catalytic reactor. Since H₂ was produced in DRM, but consumed in RWGS, its flow rate was 7198 kg/hr and 1738 kg/hr, respectively, in the stream after the switchable catalytic reactor. It should be noted that after the methanol synthesis reactor, the flow rate of H₂ was zero in both cases. The total annual income from the MeOH and SNG was calculated at €133m and €553m, respectively. If the switchable reactor had operated the DRM reaction for a longer period, it would have definitely increased the revenue as more syngas (and H₂) would have been produced via the DRM reaction than via the RWGS reaction, meaning a higher amount of MeOH.

The total capital cost was ca. €392m. In general, across the three flow schemes, the DRM flow scheme had the highest sizing parameters, which were used to calculate the purchase cost of each equipment. This was associated with the DRM reaction stoichiometry, and hence the higher amount of products downstream MeOH production was associated with higher capital costs than those of the production of SNG because additional expensive pieces of equipment, such as furnace, recycling loop, and distillation column, were required after multistage compression. The capital costing calculations showed that the compressor and furnace were the most expensive pieces of equipment, contributing significantly to the overall capital cost. The total fixed capital cost for both the compressor and the furnace were €163m with the compressors alone accounting for ca. 47% of the total capital costs. The main factor that affected the calculations for the capital and operating cost was the flowrate of the feed stream at the start of the process. Variations in flowrates had more impact on the capital costing of the compressor and furnace because a significant increase in compressor power and furnace duty was observed when processing a considerable amount of feed, resulting in a substantial increase in the overall capital cost due to the need for bigger equipment. The advantage of the studied process was that it utilised a furnace to produce a hot flue gas stream, which was in turn used for heat integration. Before being combusted, the

stream included mainly CO, CO₂, and CH₄ (in the case of DRM only) after the recycle loop of the MeOH synthesis reactor, helping to both achieve heat integration and prevent the release of highly toxic CO into the atmosphere. In general, even though furnaces significantly increase capital costs, they reduce operational expenses by lowering electricity requirements since the flue gas is hot enough to heat cold streams. The capital cost could have been reduced more if the MeOH synthesis had been able to be operated at lower pressures with the aid of a more effective catalyst [37] as fewer compressors would have been required.

Fig. 3 shows the operating expenses in the base case scenario, calculated as €452m in total with the largest contributors being electricity, cost of H₂, and maintenance at €305m, €106m, and €19m per year, respectively. Even though H₂ cost was significant and was expected to have had a higher contribution, the operating costs were dominated by the cost of electricity. This was attributed to the high amounts of electricity needed to operate the compressors and the reactors isothermally (mainly during DRM), in addition to avoiding using H₂ for a third of a year due to the DRM flow scheme. The pre-tax profit was calculated as €234m per year when taking the annual sales income and operating expenses into consideration.

Fig. 4 shows the comparison of the operating costs between the DRM, RWGS, and CO₂ methanation flow schemes. The operating costs were at their highest point when the switchable reactor was operating the DRM, followed by the RWGS, and the CO₂ methanation reactions. The total electricity requirement was the main reason why the DRM flow scheme had the highest operating costs, mainly due to the power required to operate the switchable reactor isothermally, in accordance with previous literature findings [62,63]. The required power for the compressors and the electric heater utilised for preheating the syngas for the MeOH synthesis also contributed to the increased electricity requirement for the DRM flow scheme. Likewise, the electricity costs were the biggest contributor to both the RWGS and the CO₂ methanation flow schemes, followed by the H₂ costs. CO₂ methanation utilised more H₂ compared to that of the RWGS flow scheme, but the operating costs for the RWGS flow scheme were €21m higher than those of the CO₂ methanation flow scheme due to the increased electricity costs associated with maintaining the temperature of the RWGS reactor. It should be noted that the low costs of SNG production are not only associated with the lower reaction temperature, but also with the presumed price of the green H₂ at its lowest possible cost in order to depict a near-future scenario. As regards the costs of the catalysts, the total switchable catalyst cost was estimated as €1.9m per year and the total MeOH catalyst cost as €1.8m per year.

A company selects its discount rate on the cost of capital, which comprises the cost of equity and the cost of debt incurred. Therefore, a project of this scale would have come under the 10-year AA grade corporate bond, with coupon rates between 0.75% and 10% [60]. Hence, a coupon rate of 4.75% was selected (Bloomberg price index for a 10-year AA bond) [80]. By using a discount rate of 4.75%, the cumulative NPV at the end of the project's lifespan was calculated as €1.9b, and the profitability index (PI) as 4.7€/€. From an investor's preceptive, the value calculated for the profitability index showed that for every €1 invested, the investor would receive an additional €4.7. Since investors seek projects with a PI greater than 1€/€, the findings of this work clearly demonstrate that a flared gas-to-MeOH/SNG process is financially attractive, while the flexibility offered by the switchable catalytic reactor allows for increasingly sustainable CO₂ and H₂ sources to be used through the lifetime of the plant.

Fig. 5 shows the cumulative cash flow, which was used to calculate a payback period of ca. 4 years with a discount rate of 4.75%, meaning that the proposed process would have required 4 years to become profitable. Those economic factors were calculated on the basis that all three reactions would have operated evenly across the course of a year, i. e. 4 months for each reaction. The SNG production was the top contributor to the profitability of the process as it enabled higher income from the sales and lower operating expenses, which also allowed for positive cumulative profits after the third year of operation. The results

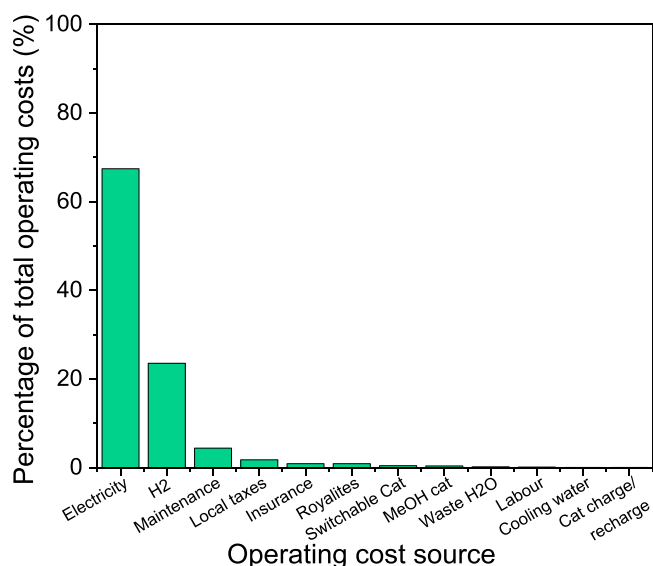


Fig. 3. Percentage breakdown of the total costs per cost source.

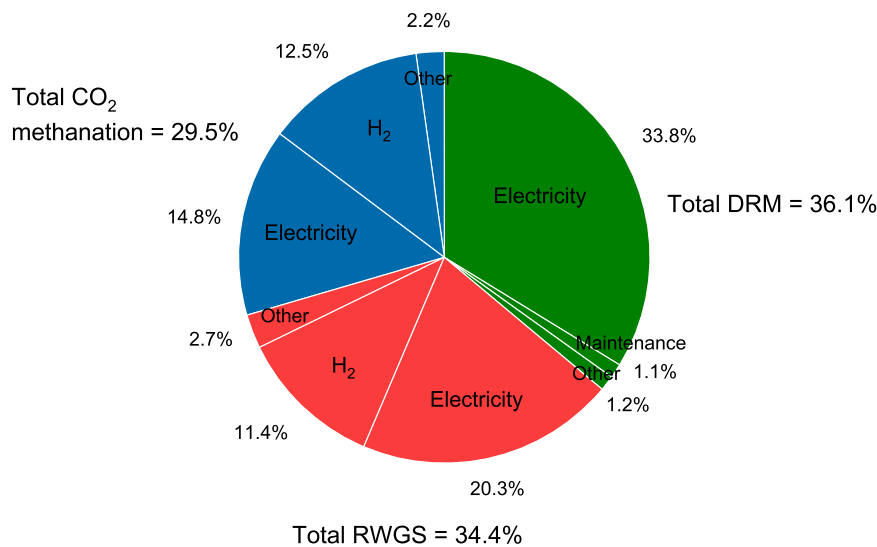


Fig. 4. : Comparison of operating costs across the different flow schemes.

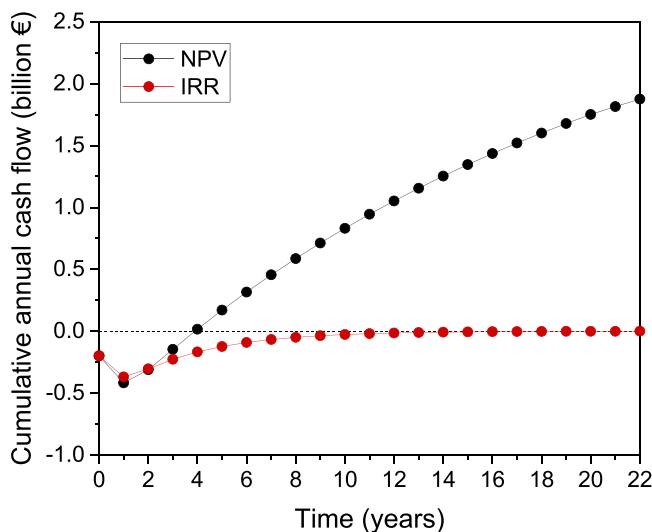


Fig. 5. : Project lifespan with an internal rate of return (IRR) and project net present value (NPV).

demonstrated that the flexibility offered by the switchable catalyst could help manufacturers switch between reactions in line with the seasonal demands for the product, thus ensuring that profits were realised for the overall process.

The internal rate of return (IRR) is defined as the discount rate that a project's NPV requires in order to become zero at the end of its lifespan and it indicates the maximum interest that could be paid to reach the break-even point. In this way, the comparison of projects is feasible since

the IRR is independent of a project's scale. According to the World Bank Group, a project is considered appealing when the NPV has a positive value, the IRR is higher than 10%, and the payback period is less than 10 years [61]. The IRR was calculated at 35%, which was a good indicator of the project's profitability. The results of this paper strongly illustrate how the proposed process would be profitable with the utilisation of switchable catalysis. The project is therefore appealing, taking account of its profitability and its remarkable reduction in CO₂ emissions. A summary of the key results is presented in Table 3.

3.3. Sensitivity analysis

A sensitivity analysis was conducted to gain a better understanding of the impact of different parameters on the profitability of the process. Hence, the selection of the parameters was based on maximising the revenue by examining how the operation time and the prices of green H₂ and products impacted the profitability of the process.

3.3.1. The effect of switching between DRM, RWGS, and methanation

As previously discussed, this process was designed to be flexible in terms of chemical synthesis production in line with supply and demand. Hence, three scenarios were analysed to investigate the impact of operation time on the profitability of the overall process. Scenario A was at work when the plant only produced MeOH via the DRM and RWGS flow schemes. Scenario B was activated when the plant only produced SNG via the CO₂ methanation flow scheme. Scenario C was put into action when the plant operated all three flow schemes, but with different operation times. In Scenario C1, the DRM and CO₂ methanation flow schemes were operated for 8 and 4 months, respectively. In Scenario C2, the RWGS and CO₂ methanation flow schemes were operated for 10 and 2 months, respectively. In Scenario C3, the DRM and CO₂ methanation flow schemes were operated for 10 and 2 months, respectively. In Scenario C4, the DRM flow scheme was operated for the entire year. In the

Table 3
Summary of key economic results.

Economic parameter	Value
Capital costs (€)	392m
Annual operating costs (€)	452m
Annual pre-tax profit (€)	234m
PI (€/€)	4.7
IRR (%)	35
NPV (€)	1.9b
Break-even point (years)	4
Carbon footprint (kg CO ₂ equivalent)	-42.8m

Table 4
Duration (in hours) of each reaction for all scenarios tested in sensitivity analysis.

	Base case	A	B	C1	C2	C3	C4
CO ₂ methanation	2800	-	8400	2800	1400	1400	-
RWGS	2800	4200	-	-	7000	-	-
DRM	2800	4200	-	5600	-	7000	8400

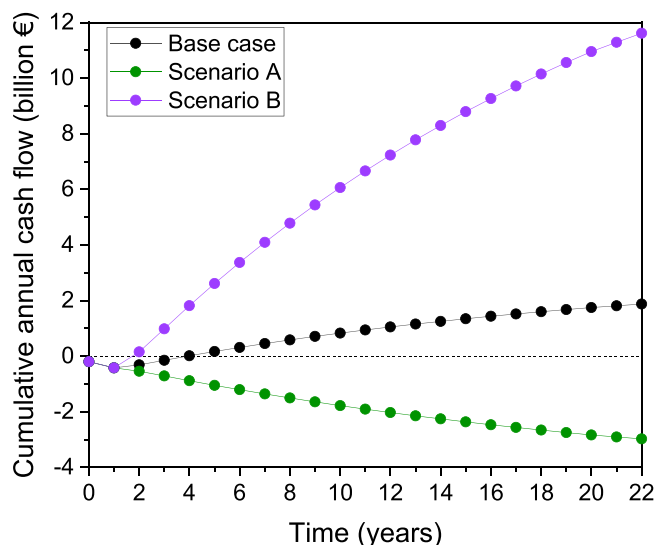


Fig. 6. Cumulative cash flow comparison between base case and scenarios A and B.

base case scenario, all three flow schemes were operated equally, i.e. 4 months of operation for each flow scheme in the course of a year. Table 4 shows the duration of each flow scheme in hours for all the aforementioned scenarios.

Fig. 6 illustrates the comparison of cumulative cash flows between A, B, and the base case scenarios. In the case of scenario A, it became apparent that the plant did not turn any profit when it only operated the DRM and RWGS reactions to produce MeOH. The cumulative cash flow for that scenario kept running negative over the years because the overall expenses outweighed the income generated from selling MeOH; the revenue from selling MeOH was €200m and the total operating costs were €474m, reaching a negative NPV of -€3b. The DRM flow scheme contributed to the considerably high operating costs, mainly due to the significant electricity requirements. More MeOH was produced when the switchable reactor operated the DRM instead of the RWGS reaction and the amounts of the MeOH produced from the RWGS flow scheme were not adequate enough for the overall process to turn a profit. Therefore, the plant should not solely operate on the MeOH production mode because it would be operating at a loss. As for scenario B, it proved profitable when it only operated the CO₂ methanation flow scheme for a

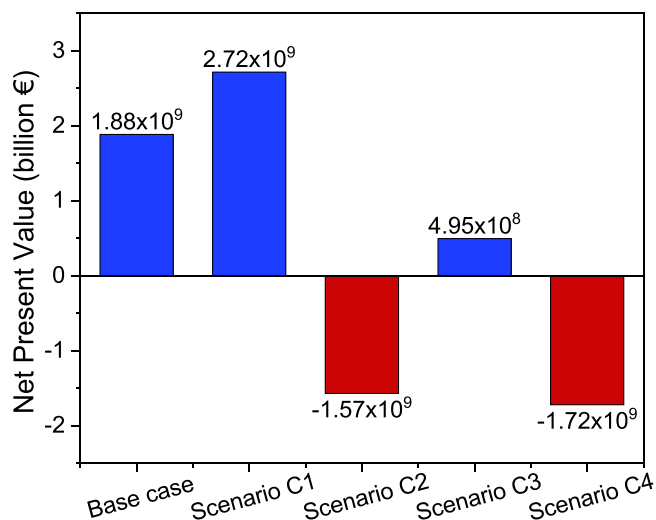


Fig. 7. Effect of varying the operating time on the NPV of the process (for scenario C).

whole year. NPV was calculated as €18b and the payback period was ca. 2 years, whereas the payback period for the base case scenario was 4 years. The fact that Scenario A was unprofitable and Scenario B profitable was the result of the lower capital costs due to the avoidance of a multistage compression and distillation column, as well as the result of the lower electricity costs associated with the switchable reactor operation, i.e. the annual electricity costs were €149m, €90m, and €66m for the DRM, RWGS, and CO₂ methanation reactions, respectively, in the base case scenario.

Fig. 7 depicts how the NPV for the overall process fluctuates, depending on how long each flow scheme operated over the course of a year. When each of these scenarios was compared with the base case one, it became clear that the operating scenario C1 was more profitable than the base case scenario; their difference in NPV was €841m, C1 payback period was 1 year shorter than the base case one, and its IRR was 45%. The reason why C1 was more profitable than the base case scenario was that despite the fact that DRM and RWGS had similar annual operating costs, i.e. €150m and €143m, respectively, the amount of MeOH produced in the DRM flow scheme was about six times larger than that of the RWGS, resulting in higher sales income. Even though 70% of the year was dedicated to the MeOH production, most of the revenue was still generated from the SNG production. Operating scenario C3 was not as profitable as the base case one for it had a considerably lower NPV and a longer payback period of 9 years. In scenario C3, the revenue mostly derived from the MeOH production, meaning that it took 10 months of operating the DRM flow scheme for the MeOH production to generate a larger revenue than that of the SNG production. Even though the electricity costs during DRM were significant, that configuration was favourable because no H₂ was required during the DRM. PI and IRR were calculated at 1.24€/€ and 15%, respectively, barely demonstrating a profitable process. Furthermore, it became obvious that it was not profitable to operate the RWGS flow scheme for most of the year (scenario C2) and the DRM flow scheme for a whole year (scenario C4) because the NPVs were calculated as -€1.6b and -€1.7b, respectively. As aforementioned, the amount of the MeOH produced through the RWGS was considerably small in comparison with its high operating expenses, which meant that the plant failed to reach the break-even point. Considerable electricity requirements for the operation of the DRM flow scheme for a whole year caused a significant increase in operating costs, rendering scenario C4 unprofitable. For scenarios C2 and C4 to be profitable, the MeOH prices had to increase so as to boost the overall revenue, reaching above 1900€/tonne to achieve the minimum payback time of 10 years and IRR of 13% for scenario C2, and above 600€/tonne for scenario C4. These values are even higher than the highest price of MeOH ever recorded to date, making the proposed scenarios unrealistic when it comes to real application in the short-to-medium term [45,48,51].

In general, in order to boost the profitability of the MeOH production, more heat integration should have been used to reduce the operating costs. For instance, the excess heat from the furnace, which had already been used to preheat the feed entering the switchable reactor, could have been further utilised to preheat the syngas mixture before entering the MeOH reactor. This would have resulted in a substantial reduction in electricity requirements for the DRM and RWGS flow schemes and lower operating costs for the electric heater. Instead of using electricity for the switchable reactor, other heating methods, such as green H₂ and natural gas combustion with carbon capture, could also be considered [78]. Additionally, natural gas prices and demand have fluctuated recently due to the COVID-19 pandemic, the war in Ukraine, and the ensuing cost of living and energy crises, especially in Europe, where the gas spot price has reached an all-time high [85]. The annual demand for natural gas in the UK in 2020 was ca. 812k GWh [86]. 14305 GWh could be produced annually by the CO₂ methanation flow scheme if it operated for 8400 h per year. This meant that this innovative strategy could meet the 1.8% of the UK natural gas demand. However, this process utilised 14% of the available H₂ in the UK to meet the 1.8%

of the UK natural gas demand, since the UK produced 700k tonnes of H₂ in 2020 in total, and this process consumed 96,000 tonnes of H₂ per year [87], further manifesting that there is a crying need for green H₂ production on a larger scale.

3.3.2. Effect of hydrogen

All the previous results represent the economics when using green low-cost H₂. Calculations were therefore carried out once again by using the price of green H₂ in 2016 (12\$/kg) and in 2019 (7.7\$/kg) to evaluate the economic feasibility of this process [78,88]. NPV was calculated as -€2.9b and -€1.49m, respectively. The process in its base case scenario failed to reach the break-even point due to a massive increase in annual operating costs, and thus it was not economically feasible to use green H₂ at its current price. The price of H₂ had to be below 2.23 €/kg for the process to break even, as Fig. 8 shows. This price is above the blue H₂ cost at present, which is 1.1–1.9€/kg [78,88], and within the price range projections of green hydrogen for 2050 [15,88]. Although green H₂ is too expensive to make this process feasible, renewable H₂ production costs are expected to decline in the future due to technological advances in both renewable electricity generation and electrolysis, enabling a significant reduction in operating costs [59]. In only three years from 2016 to 2019, there was a 36% drop in its price, confirming its downward trend. It should also be mentioned that grey H₂ price depends on the price of natural gas, which has fluctuated a lot over the past couple of years [60]. Additionally, this flexible process can switch between the three reactions, meaning that the CO₂ methanation and RWGS reactions can run during the summer months when the production of renewable electricity peaks and green H₂ is more widely available.

3.3.3. Effect of product prices

A final sensitivity analysis was carried out to determine the impact of the products' prices on the profitability of this project. When calculating the NPV for varying natural gas prices, the MeOH price was stable at its original value, i.e. 370 €/t. When calculating the NPV for varying MeOH prices, the natural gas price was stable at its original value, i.e. 116.08 €/MWh. These particular price ranges of products were selected to depict a wide range of possible price changes, with the lowest value being lower than ever recorded over the past years and the highest value being higher than ever recorded over the past years [51,60]. The results are presented in Fig. 9. It was shown that the prices of natural gas had a bigger effect on the project's profitability than MeOH. Indeed, a positive NPV was obtained regardless of the MeOH price (even when it was set to 0), indicating that the process was profitable due to the natural gas production. This has already been highlighted in Fig. 2 and particularly

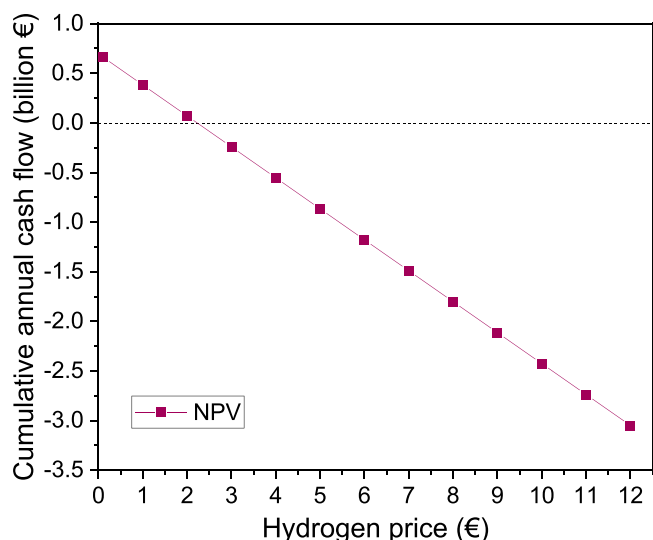


Fig. 8. NPV values (€) vs hydrogen price (€/kg) for the base case scenario.

in Fig. 7 when the different operation scenarios were considered. Natural gas prices had a huge impact on the NPV values, with the NPV becoming zero when the natural gas price was approximately 75 €/MWh. This confirmed that the process profitability was dependent on the CO₂ methanation mode.

4. Discussion

To date, CO₂ utilisation technologies have been viewed as a mean to convert flared gas into valuable chemicals, while using flared gas for electricity production has also been investigated [17,18]. However, most of the techno-economic studies surrounding the flared gas/CO₂ utilisation processes have been economically unfeasible because of the high capital investments and operating costs, which mainly pertain to electricity and catalysts, and the scant government subsidies.

The results of this work demonstrate an interesting switchable chemical synthesis strategy for utilising flare/vent gas streams and producing on-demand chemicals and fuels. It is shown that SNG production is profitable if green H₂ reaches cost parity with grey hydrogen; however, the process is profitable with the current grey and blue H₂ cost. In addition, it was found that DRM and RWGS schemes are not profitable in the short term, but there can be ways to future-proof our investment and monetise emissions. For example, the switchable reactor can be set up for SNG production and then, it can gradually be shifted to supply low emission syngas to other processes, especially when the demand for SNG is low, e.g. during summer. It should also be noted that taxes or penalties imposed on venting CO₂ and hydrocarbons to the atmosphere have not been considered in this work. This cost is avoided when the gases are used through our carbon negative process, making it even more attractive to utilise these emissions rather than release them. Hence, our analysis does not include savings/revenue from the carbon tax, making our profit calculations conservative. In this regard, even methanol can be more appealing, offsetting its production costs to a certain extent, because the conventional carbon-positive production of methanol can even become unprofitable in the future due to carbon tax increase. As mentioned earlier, being on the increase in recent years, the carbon tax reached €100/tonne in Europe at the end of February 2023 [69] and is expected to increase even more in the near future due to current environmental regulations, potentially reaching €200/tonne by 2050, according to the International Energy Agency [89]. By multiplying only the CO₂ inlet flow rates (Table 1) of the three flow schemes by their corresponding operation times, the molecular weight of CO₂, and the price of carbon tax in February 2023 in Europe, the avoided tax based on CO₂ emissions alone is calculated to be €143.6m. As a result, the carbon tax can be more than €287.2m in 2050, without even considering the contribution of hydrocarbons that will produce additional CO₂ when combusted.

Moreover, this work illustrates the importance of exploring greater synergies between CO₂ utilisation and downstream process treatment to decarbonise various industry sectors. Hence, more efforts should go into the oil and gas sectors to utilise waste streams so as to produce sustainable chemicals and fuels, especially when considering how much CO₂ and hydrocarbons are lost through oil and gas operations per year. While the process that can recover these waste streams is profitable, it can benefit from further financial assistance from government subsidies and incentives. The switchable process can further be optimised by ensuring that it has the capabilities to adapt to other sources of emissions from refineries and other heavy-carbon industrial processes, such as steel-making, cement-manufacturing, etc. In addition, this process can be improved in terms of the heat and power integration as well as the activity and durability of switchable catalysts. In any case, this study demonstrates how advances in catalysis affect the economic viability of alternative chemical and fuel production methods owing to the fact that when the widely unprofitable DRM is combined with CO₂ methanation, the overall process becomes profitable.

In general, these findings reveal the adverse effect of supply and

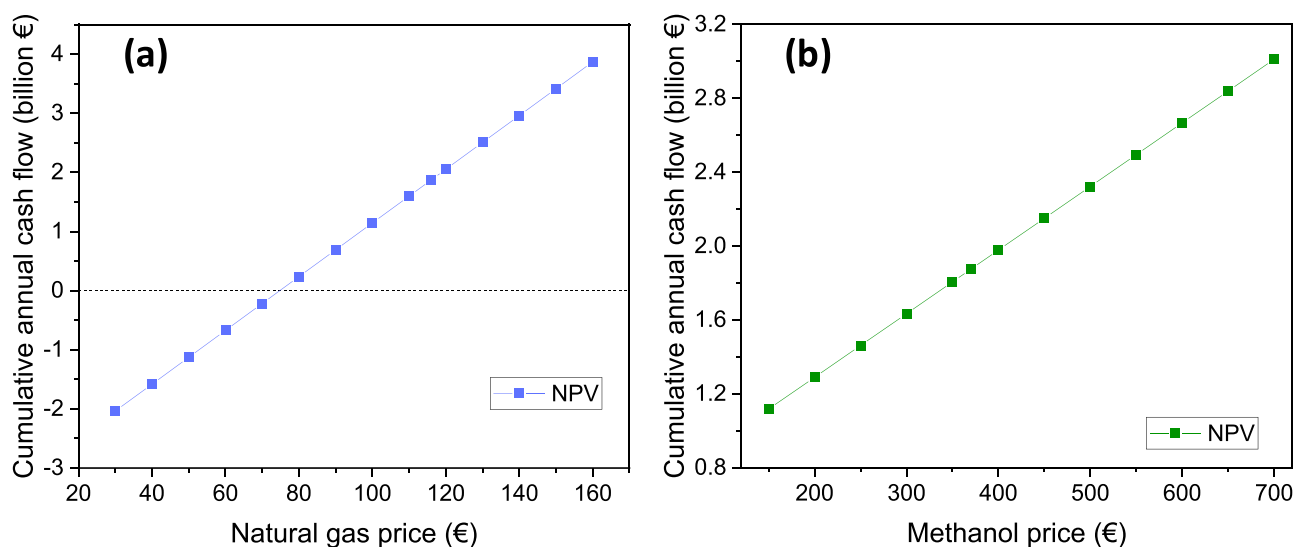


Fig. 9. NPV values (€) vs (a) natural gas price (€/MWh) and (b) methanol price (€/t) for the base case scenario.

demand as well as the uncertainties of the global market on the profitability of industrial projects, showcasing that the use of switchable catalysis can be useful in scenarios of this type. For example, the demand for natural gas for heating purposes by residential and commercial consumers increases the overall demand for natural gas during the winter [14,90]. Therefore, rather than operating the DRM and RWGS flow schemes for MeOH production, a plant could focus on operating only the CO₂ methanation reaction to produce SNG during these periods to meet the demands. Similarly, profitable switches in operation can be made by keeping an eye on market prices and aiming for the most profitable ones at any given time. In addition, with this study, we show that petrochemical and oil and gas companies can begin to gradually utilise their emissions and streamline their processes around the use of CO₂-derived syngas so as to re-envision chemical synthesis and consider using a switchable reactor that utilises the waste CO₂. The utilisation of such technologies can help countries become less reliant on fuel imports, which can only be beneficial. Finally, it is vital that the prices of renewable H₂ decrease so as to generate positive cumulative annual cash flows.

5. Conclusions

The work presented in this paper examined the techno-economic feasibility of switchable catalysis for chemical synthesis flexibility. The use of a switchable catalyst is a novel approach towards the reduction of CO₂ emissions since a catalyst can be used not only for an individual CO₂ utilisation scheme, but also for multiple switchable operation scenarios that allow flexible CO₂ recycling. The techno-economic analysis demonstrated that this approach is currently economically feasible because of the production of SNG. The production of MeOH is not profitable, but there are other benefits deriving from the flexibility of chemical synthesis in adapting to the changing demands and the availability of feedstocks. The annual revenue for this process was €687m, generated from the MeOH and SNG sales. The total capital investment required for this project was €392m and the yearly operating costs were €452m. While there were various operating expenses, the major ones were the costs of electricity and H₂ at €305m and €106m, respectively. The pre-tax profit was calculated as €234m per year. By using a discount rate of 4.75%, the NPV was calculated as €1.9b at the end of the 20-year-lifespan of the project with a PI of 4.7€/€. The process had a payback time of nearly 4 years and its IRR was calculated at 35%. When all the sources of the emissions in the process were taken into account, the yearly CO₂ consumption was estimated at 42.8m tonnes of

CO₂ equivalent. According to the sensitivity analysis, the production of MeOH/SNG was extremely sensitive to operation time, and H₂ and natural gas prices. In this work, we illustrated that the overall process became profitable when the DRM and CO₂ methanation flows schemes operated together or when only the CO₂ methanation was operated for an entire year. The use of green H₂ at its current price is not economically viable for this process, unlike blue H₂, whose implementation makes the process economically feasible. Significant government support is therefore essential to promote the usage of renewable H₂. Overall, the utilisation of switchable catalysis to produce MeOH and SNG from flared natural gas manifests itself as a unique approach towards the minimisation of GHG emissions and the move towards a circular economy. Nonetheless, new synergies between flared gas/CO₂ utilisation and upgrading should be further investigated in order to decarbonise strategic chemicals and fuels, such as formic acid, acetic acid, and dimethyl ether.

CRediT authorship contribution statement

Loukia-Pantzechroula Merkouri: Conceptualization, Formal analysis, Investigation, Methodology, Validation, Visualization, Writing – original draft. **Jayson Mathew:** Formal analysis, Investigation, Methodology, Visualization, Writing – original draft. **Jerin Jacob:** Formal analysis, Investigation, Methodology, Visualization, Writing – original draft. **Tomás Ramirez Reina:** Conceptualization, Funding acquisition, Resources, Supervision, Writing – review & editing. **Melis S. Duyar:** Conceptualization, Funding acquisition, 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.

Acknowledgements

Financial support for this work was provided by the School of Chemistry and Chemical Engineering (Chemical and Process

Engineering) and the Doctoral College of the University of Surrey.

Appendix A. Supporting information

Supplementary data associated with this article can be found in the online version at [doi:10.1016/j.jcou.2023.102652](https://doi.org/10.1016/j.jcou.2023.102652).

References

- [1] IEA, CO₂ Emissions in 2022, IEA. (2023). (<https://www.iea.org/reports/co2-emissions-in-2022>) (accessed March 15, 2023).
- [2] W.M. Budzianowski, Value-added carbon management technologies for low CO₂ intensive carbon-based energy vectors, *Energy* 41 (2012) 280–297, <https://doi.org/10.1016/j.energy.2012.03.008>.
- [3] O. Boucher, P. Friedlingstein, B. Collins, K.P. Shine, The indirect global warming potential and global temperature change potential due to methane oxidation, *Environ. Res. Lett.* 4 (2009), <https://doi.org/10.1088/1748-9326/4/4/044007>.
- [4] X. Lan, K.W. Thoning, E.J. Dlugokencky, Trends in atmospheric methane, NOAA Glob. Monit. Lab. (2023). (https://gml.noaa.gov/ccgg/trends_ch4/). accessed March 30, 2023.
- [5] P. Tans, R. Keeling, K.W. Thoning, Trends in atmospheric carbon dioxide, NOAA Glob. Monit. Lab (2023), <https://doi.org/10.15138/9NOH-ZH07>.
- [6] J.C.M. Pires, F.G. Martins, M.C.M. Alvim-Ferraz, M. Simões, Recent developments on carbon capture and storage: an overview, *Chem. Eng. Res. Des.* 89 (2011) 1446–1460, <https://doi.org/10.1016/j.cherd.2011.01.028>.
- [7] A. Al-Mamoori, A. Krishnamurthy, A.A. Rowanagi, F. Rezaei, Carbon capture and utilization update, *Energy Technol.* 5 (2017) 834–849, <https://doi.org/10.1002/ente.201600747>.
- [8] C.D. Elvidge, M.D. Bazilian, M. Zhizhin, T. Ghosh, K. Baugh, F.C. Hsu, The potential role of natural gas flaring in meeting greenhouse gas mitigation targets, *Energy Strateg. Rev.* 20 (2018) 156–162, <https://doi.org/10.1016/j.esr.2017.12.012>.
- [9] J. Aizarani, Natural gas flaring volume worldwide from 2014 to 2021, Statista. (2023). (<https://www.statista.com/statistics/653611/natural-gas-flaring-at-oil-production-sites-worldwide/>) (accessed March 30, 2023).
- [10] G.G.F.R.P. (GGFR), Global Gas Flaring Data, World Bank. (2022). (<https://www.worldbank.org/en/programs/gasflaringreduction/global-flaring-data>) (accessed March 30, 2023).
- [11] R. Mansoor, M. Tahir, Recent developments in natural gas flaring reduction and reformation to energy-efficient fuels: a review, *Energy Fuels* 35 (2021) 3675–3714, <https://doi.org/10.1021/acs.energyfuels.0c04269>.
- [12] M.R. Rahimpour, S.M. Jokar, Feasibility of flare gas reformation to practical energy in Farashband gas refinery: no gas flaring, *J. Hazard. Mater.* 209–210 (2012) 204–217, <https://doi.org/10.1016/j.jhazmat.2012.01.017>.
- [13] L.-P. Merkouri, A.I. Paksoy, T.R. Reina, M.S. Duyar, The need for flexible chemical synthesis and how dual-function materials can pave the way, *ACS Catal.* 13 (2023) 7230–7242, <https://doi.org/10.1021/acscatal.3c00880>.
- [14] O. Comstock, U.S. natural gas consumption has both winter and summer peaks, U. S. Energy Inf. Adm. (2020). (<https://www.eia.gov/todayinenergy/detail.php?id=42815>) (accessed October 4, 2023).
- [15] The Future of Hydrogen, IEA. (2019). (<https://www.iea.org/reports/the-future-of-hydrogen>) (accessed October 2, 2023).
- [16] M. Zolfaghari, V. Pirouzfard, H. Sakhaeinia, Technical characterization and economic evaluation of recovery of flare gas in various gas-processing plants, *Energy* 124 (2017) 481–491, <https://doi.org/10.1016/j.energy.2017.02.084>.
- [17] A. Khalili-Garakani, M. Iravaninia, M. Nezhadfar, A review on the potentials of flare gas recovery applications in Iran, *J. Clean. Prod.* 279 (2021), 123345, <https://doi.org/10.1016/j.jclepro.2020.123345>.
- [18] P. Iora, P. Bombarda, S.L. Gómez Aláez, C. Invernizzi, T. Rajabloo, P. Silva, Flare gas reduction through electricity production, *Energy Sources, Part A Recover. Util. Environ. Eff.* 38 (2016) 3116–3124, <https://doi.org/10.1080/15567036.2015.1129471>.
- [19] C. Jensen, M.S. Duyar, Thermodynamic analysis of dry reforming of methane for valorization of landfill gas and natural gas, *Energy Technol.* 9 (2021) 1–12, <https://doi.org/10.1002/ente.202100106>.
- [20] L.-P. Merkouri, E. le Sache, L. Pastor-Pérez, M.S. Duyar, T.R. Reina, Versatile Ni-Ru catalysts for gas phase CO₂ conversion: bringing closer dry reforming, reverse water gas shift and methanation to enable end-products flexibility, *Fuel* 315 (2022), 123097, <https://doi.org/10.1016/j.fuel.2021.123097>.
- [21] E. le Saché, L. Pastor-Pérez, B.J. Haycock, J.J. Villora-Picó, A. Sepúlveda-Escribano, T.R. Reina, Switchable catalysts for chemical CO₂ recycling: a step forward in the methanation and reverse water-gas shift reactions, *ACS Sustain. Chem. Eng.* 8 (2020) 4614–4622, <https://doi.org/10.1021/acscuschemeng.0c00551>.
- [22] Q. Zhang, J.J. Villora-Pico, M. Joyce, A. Sepúlveda-Escribano, M.S. Duyar, T. R. Reina, Ni-Phosphide catalysts as versatile systems for gas-phase CO₂ conversion: Impact of the support and evidences of structure-sensitivity, *Fuel* 323 (2022), 124301, <https://doi.org/10.1016/j.fuel.2022.124301>.
- [23] E. le Saché, T.R. Reina, Analysis of Dry Reforming as direct route for gas phase CO₂ conversion. The past, the present and future of catalytic DRM technologies, *Prog. Energy Combust. Sci.* 89 (2022), <https://doi.org/10.1016/j.peccs.2021.100970>.
- [24] B. Abdullah, N.A. Abd Ghani, D.V.N. Vo, Recent advances in dry reforming of methane over Ni-based catalysts, *J. Clean. Prod.* 162 (2017) 170–185, <https://doi.org/10.1016/j.jclepro.2017.05.176>.
- [25] S. Wang, G.Q. (Max) Lu, G.J. Millar, Carbon dioxide reforming of methane to produce synthesis gas over metal-supported catalysts: state of the art shaobin, *Energy Fuels* 10 (1996) 896–904, <https://doi.org/10.1021/ef950227t>.
- [26] O.A. Berekidou, M.A. Goula, Biogas reforming for syngas production over nickel supported on ceria-alumina catalysts, *Catal. Today* 195 (2012) 93–100, <https://doi.org/10.1016/j.cattod.2012.07.006>.
- [27] M.D. Porosoff, J.G. Chen, Trends in the catalytic reduction of CO₂ by hydrogen over supported monometallic and bimetallic catalysts, *J. Catal.* 301 (2013) 30–37, <https://doi.org/10.1016/j.jcat.2013.01.022>.
- [28] L. Yang, L. Pastor-Pérez, S. Gu, A. Sepúlveda-Escribano, T.R. Reina, Highly efficient Ni/CeO₂-Al₂O₃ catalysts for CO₂ upgrading via reverse water-gas shift: Effect of selected transition metal promoters, *Appl. Catal. B Environ.* 232 (2018) 464–471, <https://doi.org/10.1016/j.apcatb.2018.03.091>.
- [29] W. Wang, S. Wang, X. Ma, J. Gong, Recent advances in catalytic hydrogenation of carbon dioxide, *Chem. Soc. Rev.* 40 (2011) 3703–3727, <https://doi.org/10.1039/c1cs15008a>.
- [30] Y.A. Daza, J.N. Kuhn, CO₂ conversion by reverse water gas shift catalysis: comparison of catalysts, mechanisms and their consequences for CO₂ conversion to liquid fuels, *RSC Adv.* 6 (2016) 49675–49691, <https://doi.org/10.1039/c6ra05414e>.
- [31] P. Frontera, A. Macario, M. Ferraro, P.L. Antonucci, Supported catalysts for CO₂ methanation: a review, *Catalysts* 7 (2017) 59, <https://doi.org/10.3390/catal7020059>.
- [32] J. Gao, Q. Liu, F. Gu, B. Liu, Z. Zhong, F. Su, Recent advances in methanation catalysts for the production of synthetic natural gas, *RSC Adv.* 5 (2015) 22759–22776, <https://doi.org/10.1039/c4ra16114a>.
- [33] F. Ocampo, B. Louis, L. Kiwi-Minsker, A.C. Roger, Effect of Ce/Zr composition and noble metal promotion on nickel based CeZr_{1-x}O₂ catalysts for carbon dioxide methanation, *Appl. Catal. A Gen.* 392 (2011) 36–44, <https://doi.org/10.1016/j.apcata.2010.10.025>.
- [34] S. Rönisch, J. Schneider, S. Matthischke, M. Schlüter, M. Götz, J. Lefebvre, P. Prabhakaran, S. Bajohr, Review on methanation - from fundamentals to current projects, *Fuel* 166 (2016) 276–296, <https://doi.org/10.1016/j.fuel.2015.10.111>.
- [35] M. Usman, W.M.A. Wan Daud, H.F. Abbas, Dry reforming of methane: Influence of process parameters - a review, *Renew. Sustain. Energy Rev.* 45 (2015) 710–744, <https://doi.org/10.1016/j.rser.2015.02.026>.
- [36] D. Pakhare, J. Spivey, A review of dry (CO₂) reforming of methane over noble metal catalysts, *Chem. Soc. Rev.* 43 (2014) 7813–7837, <https://doi.org/10.1039/c3cs60395d>.
- [37] X. Jiang, X. Nie, X. Guo, C. Song, J.G. Chen, Recent advances in carbon dioxide hydrogenation to methanol via heterogeneous catalysis, *Chem. Rev.* 120 (2020) 7984–8034, <https://doi.org/10.1021/acs.chemrev.9b00723>.
- [38] G. Torres-Sempere, L. Pastor-Pérez, J.A. Odriozola, J. Yu, F.J. Duran-Olivencia, L. F. Bobadilla, T.R. Reina, Recent advances on gas-phase CO₂ conversion: catalysis design and chemical processes to close the carbon cycle, *Curr. Opin. Green. Sustain. Chem.* 36 (2022), <https://doi.org/10.1016/j.cogsc.2022.100647>.
- [39] C. Jensen, M.S. Duyar, Thermodynamic analysis of dry reforming of methane for valorization of landfill gas and natural gas, *Energy Technol.* 9 (2021), 2100106, <https://doi.org/10.1002/ente.202100106>.
- [40] G. Leonzio, Methanol synthesis: optimal solution for a better efficiency of the process, *Processes* 6 (2018), <https://doi.org/10.3390/pr6030020>.
- [41] J.L. Martín-Espejo, J. Gandara-Loe, J.A. Odriozola, T.R. Reina, L. Pastor-Pérez, Sustainable routes for acetic acid production: Traditional processes vs a low-carbon, biogas-based strategy, *Sci. Total Environ.* 840 (2022), <https://doi.org/10.1016/j.scitotenv.2022.156663>.
- [42] M. Bowker, Methanol synthesis from CO₂ hydrogenation, *ChemCatChem* 11 (2019) 4238–4246, <https://doi.org/10.1002/cctc.201900401>.
- [43] M.D. Porosoff, B. Yan, J.G. Chen, Catalytic reduction of CO₂ by H₂ for synthesis of CO, methanol and hydrocarbons: Challenges and opportunities, *Energy Environ. Sci.* 9 (2016) 62–73, <https://doi.org/10.1039/c5ee02657a>.
- [44] J.C.J. Bart, R.P.A. Sneed, Copper-zinc oxide-alumina methanol catalysts revisited, *Catal. Today* 2 (1987) 1–124, [https://doi.org/10.1016/0920-5861\(87\)80001-9](https://doi.org/10.1016/0920-5861(87)80001-9).
- [45] D. Bellotti, M. Rivarolo, L. Magistri, A.F. Massardo, Feasibility study of methanol production plant from hydrogen and captured carbon dioxide, *J. CO₂ Util.* 21 (2017) 132–138, <https://doi.org/10.1016/j.jcou.2017.07.001>.
- [46] A.S. Alsuhaibani, S. Afzal, M. Challiwal, N.O. Elbashir, M.M. El-Halwagi, The impact of the development of catalyst and reaction system of the methanol synthesis stage on the overall profitability of the entire plant: a techno-economic study, *Catal. Today* 343 (2020) 191–198, <https://doi.org/10.1016/j.cattod.2019.03.070>.
- [47] J. Nyári, M. Magdeldin, M. Larmi, M. Järvinen, A. Santasalo-Aarnio, Techno-economic barriers of an industrial-scale methanol CCU-plant, *J. CO₂ Util.* 39 (2020), 101166, <https://doi.org/10.1016/j.jcou.2020.101166>.
- [48] C. Zhang, K.W. Jun, R. Gao, G. Kwak, H.G. Park, Carbon dioxide utilization in a gas-to-methanol process combined with CO₂/Steam-mixed reforming: Techno-economic analysis, *Fuel* 190 (2017) 303–311, <https://doi.org/10.1016/j.fuel.2016.11.008>.
- [49] K. Harris, R.G. Grim, Z. Huang, L. Tao, A comparative techno-economic analysis of renewable methanol synthesis from biomass and CO₂: Opportunities and barriers to commercialization, *Appl. Energy* 303 (2021), 117637, <https://doi.org/10.1016/j.apenergy.2021.117637>.

- [50] H. Nieminen, A. Laari, T. Koironen, CO₂ hydrogenation to methanol by a liquid-phase process with alcoholic solvents: a techno-economic analysis, *Processes* 7 (2019) 1–24, <https://doi.org/10.3390/pr7070405>.
- [51] L. Fernández, Monthly methanol spot prices worldwide from January 2020 to June 2022, by region, Statista. (2022). (<https://www.statista.com/statistics/1323381/monthly-methanol-spot-prices-worldwide-by-region/>) (accessed March 14, 2023).
- [52] G. Iaquaniello, S. Setini, A. Salladini, M. De Falco, CO₂ valorization through direct methanation of flue gas and renewable hydrogen: a technical and economic assessment, *Int. J. Hydrog. Energy* 43 (2018) 17069–17081, <https://doi.org/10.1016/j.ijhydene.2018.07.099>.
- [53] F.M. Baena-Moreno, Z. Zhang, X.P. Zhang, T.R. Reina, Profitability analysis of a novel configuration to synergize biogas upgrading and Power-to-Gas, *Energy Convers. Manag.* 224 (2020), 113369, <https://doi.org/10.1016/j.enconman.2020.113369>.
- [54] B. Lee, H. Lee, S. Kang, H. Lim, Stochastic techno-economic analysis of power-to-gas technology for synthetic natural gas production based on renewable H₂ cost and CO₂ tax credit, *J. Energy Storage* 24 (2019), 100791, <https://doi.org/10.1016/j.est.2019.100791>.
- [55] F.M. Baena-Moreno, N. Cid-Castillo, H. Arellano-García, T.R. Reina, Towards emission free steel manufacturing – exploring the advantages of a CO₂ methanation unit to minimize CO₂ emissions, *Sci. Total Environ.* 781 (2021), 146776, <https://doi.org/10.1016/j.scitotenv.2021.146776>.
- [56] J. Guilerá, J. Ramon Morante, T. Andreu, Economic viability of SNG production from power and CO₂, *Energy Convers. Manag.* 162 (2018) 218–224, <https://doi.org/10.1016/j.enconman.2018.02.037>.
- [57] M. Thema, F. Bauer, M. Sterner, Power-to-Gas: electrolysis and methanation status review, *Renew. Sustain. Energy Rev.* 112 (2019) 775–787, <https://doi.org/10.1016/j.rser.2019.06.030>.
- [58] M. Götz, J. Lefebvre, F. Mörs, A. McDaniel Koch, F. Graf, S. Bajohr, R. Reimert, T. Kolb, Renewable power-to-gas: a technological and economic review, *Renew. Energy* 85 (2016) 1371–1390, <https://doi.org/10.1016/j.renene.2015.07.066>.
- [59] G. Gahleitner, Hydrogen from renewable electricity: an international review of power-to-gas pilot plants for stationary applications, *Int. J. Hydrog. Energy* 38 (2013) 2039–2061, <https://doi.org/10.1016/j.ijhydene.2012.12.010>.
- [60] A. O'Neill, Monthly prices for natural gas in the United States and Europe from January 2014 to December 2022, Statista (2023). (<https://www.statista.com/statistics/673333/monthly-prices-for-natural-gas-in-the-united-states-and-europe/>). accessed March 14, 2023.
- [61] L.-P. Merkouri, H. Ahmet, T. Ramirez Reina, M.S. Duyar, The direct synthesis of dimethyl ether (DME) from landfill gas: a techno-economic investigation, *Fuel* 319 (2022), 123741, <https://doi.org/10.1016/j.fuel.2022.123741>.
- [62] F.M. Baena-Moreno, D. Sebastia-Saez, Q. Wang, T.R. Reina, Is the production of biofuels and bio-chemicals always profitable? Co-production of biomethane and urea from biogas as case study, *Energy Convers. Manag.* 220 (2020), 113058, <https://doi.org/10.1016/j.enconman.2020.113058>.
- [63] F. Vidal-Barrero, F.M. Baena-Moreno, C. Preciado-Cárdenas, Á. Villanueva-Perales, T.R. Reina, Hydrogen production from landfill biogas: Profitability analysis of a real case study, *Fuel* 324A (2022), 124438, <https://doi.org/10.1016/j.fuel.2022.124438>.
- [64] K. Mondal, S. Sasmal, S. Badgandi, D.R. Chowdhury, V. Nair, Dry reforming of methane to syngas: a potential alternative process for value added chemicals—a techno-economic perspective, *Environ. Sci. Pollut. Res.* 23 (2016) 22267–22273, <https://doi.org/10.1007/s11356-016-6310-4>.
- [65] J.L. Ong, A.C.M. Loy, S.Y. Teng, B.S. How, Future paradigm of 3D printed Ni-based metal organic framework catalysts for dry methane reforming: techno-economic and environmental analyses, *ACS Omega* 7 (2022) 15369–15384, <https://doi.org/10.1021/acsomega.1c06873>.
- [66] T. Hos, M. Herskowitz, Techno-economic analysis of biogas conversion to liquid hydrocarbon fuels through production of lean-hydrogen syngas, *ACS Eng. Au.* 2 (2022) 450–460, <https://doi.org/10.1021/acseengineeringau.2c00019>.
- [67] E. Portillo, J. Gandara-Loe, T.R. Reina, L. Pastor-Pérez, Is the RWGS a viable route for CO₂ conversion to added value products? A techno-economic study to understand the optimal RWGS conditions, *Sci. Total Environ.* 857 (2023), 159394, <https://doi.org/10.1016/j.scitotenv.2022.159394>.
- [68] E. Rezaei, S. Dzuryk, Techno-economic comparison of reverse water gas shift reaction to steam and dry methane reforming reactions for syngas production, *Chem. Eng. Res. Des.* 144 (2019) 354–369, <https://doi.org/10.1016/j.cherd.2019.02.005>.
- [69] C. Hodgson, D. Sheppard, EU carbon price tops €100 a tonne for first time, *Financ. Times*. (2023). (<https://www.ft.com/content/7a0dd553-fa5b-4a58-81d1-e500f8ce3d2a>) (accessed March 14, 2023).
- [70] R. Chauvy, L. Dubois, P. Lybaert, D. Thomas, G. De Weireld, Production of synthetic natural gas from industrial carbon dioxide, *Appl. Energy* 260 (2020), 114249, <https://doi.org/10.1016/j.apenergy.2019.114249>.
- [71] S. Fendt, A. Buttler, M. Gaderer, H. Spliethoff, Comparison of synthetic natural gas production pathways for the storage of renewable energy, *WIREs Energy Environ.* 5 (2016) 327–350, <https://doi.org/10.1002/wene.189>.
- [72] W.A. Poe, S. Mokhtab, Introduction to natural gas processing plants, in: *Model. Control. Optim. Nat. Gas Process. Plants*, Elsevier, 2017, pp. 1–72. (<https://doi.org/10.1016/b978-0-12-802961-9.00001-2>).
- [73] P. Haro, F. Trippe, R. Stahl, E. Henrich, Bio-syngas to gasoline and olefins via DME - a comprehensive techno-economic assessment, *Appl. Energy* 108 (2013) 54–65, <https://doi.org/10.1016/j.apenergy.2013.03.015>.
- [74] F. Trippe, M. Fröhling, F. Schultmann, R. Stahl, E. Henrich, A. Dalai, Comprehensive techno-economic assessment of dimethyl ether (DME) synthesis and Fischer-Tropsch synthesis as alternative process steps within biomass-to-liquid production, *Fuel Process. Technol.* 106 (2013) 577–586, <https://doi.org/10.1016/j.fuproc.2012.09.029>.
- [75] W.D. Seider, J.D. Seader, D.R. Lewin, S. Widagdo, *Product and Process Design Principles: Synthesis, Analysis and Evaluation*, 3rd ed., Wiley, 2009.
- [76] G. Towler, R. Sinnott, Capital cost estimating, in: *Chem. Eng. Des.*, 2nd ed., 2013: pp. 307–354. (<https://doi.org/10.1016/b978-0-12-821179-3.00007-8>).
- [77] R.K. Sinnott, Coulson & Richardson's Chemical Engineering Design, 4th ed., Elsevier, Oxford, 2005. (https://www.academia.edu/37875648/Coulson_Richardsons_Chemical_Engineering_Vol_6_Chemical_Engineering_Design_4th_Edition).
- [78] R.M. Bown, M. Joyce, Q. Zhang, T.R. Reina, M.S. Duyar, Identifying commercial opportunities for the reverse water gas shift reaction, *Energy Technol.* 9 (2021), 2100554, <https://doi.org/10.1002/ente.202100554>.
- [79] Global Hydrogen Review 2021, IEA. (2021). (<https://doi.org/10.1787/39351842-en>).
- [80] Trading Economics, EU Natural Gas - TTF Gas, *Trading Econ.* (2023). (<https://tradingeconomics.com/commodity/eu-natural-gas>) (accessed April 7, 2023).
- [81] EC, Electricity price statistics, Eurostat - Stat. Explain. (2016) 1–10. (https://ec.europa.eu/eurostat/statistics-explained/index.php?title=Electricity_price_statistics) (accessed April 7, 2023).
- [82] G. Towler, R. Sinnott, Estimating Revenues and Production Costs, in: *Chem. Eng. Des.*, 2nd ed., 2013: pp. 355–387. (<https://doi.org/10.1016/b978-0-08-096659-5.00008-0>).
- [83] C. Maxwell, Cost indices, *Towering Ski*. 1 (n.d.). (<https://toweringskills.com/financial-analysis/cost-indices/>) (accessed April 7, 2023).
- [84] Department for Energy Security and Net Zero and Department for Business Energy & Industrial Strategy, Greenhouse gas reporting: conversion factors 2021, Gov. UK (2021) 1. (<https://www.gov.uk/government/publications/greenhouse-gas-reporting-conversion-factors-2021>) (accessed April 7, 2023).
- [85] B. Zakeri, K. Paulavets, L. Barreto-Gomez, L.G. Echeverri, S. Pachauri, B. Boza-Kiss, C. Zimm, J. Rogelj, F. Creutzig, D. Ürges-Vorsatz, D.G. Victor, M.D. Bazilian, S. Fritz, D. Gielen, D.L. McCollum, L. Srivastava, J.D. Hunt, S. Pouda, Pandemic, War, and Global Energy Transitions, *Energies* 15 (2022) 6114, <https://doi.org/10.3390/en15176114>.
- [86] J. Aizarani, Demand of natural gas in the United Kingdom (UK) from 2010 to 2021, Statista. (2023). (<https://www.statista.com/statistics/550946/natural-gas-demand-uk/>) (accessed April 7, 2023).
- [87] UK hydrogen strategy, Government UK. 85 (2021) 17. (<https://doi.org/10.1002/cind.859.6.x>).
- [88] IEA, Global average levelised cost of hydrogen production by energy source and technology, 2019 and 2050, IEA. (2020). (<https://www.iea.org/data-and-statistics/charts/global-average-levelised-cost-of-hydrogen-production-by-energy-source-and-technology-2019-and-2050>) (accessed June 16, 2023).
- [89] IEA, Global Energy and Climate Model - Macro drivers, IEA. (2022). (<https://www.iea.org/reports/global-energy-and-climate-model/macro-drivers>) (accessed June 16, 2023).
- [90] Supply, transformation and consumption of gas - monthly data, Eurostat. (2022). (https://ec.europa.eu/eurostat/databrowser/view/NRG_CB_GASM_custo_m_4164035/default/line?lang=en) (accessed October 4, 2023).