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Techno-economic Analysis of Alternatives for Carbon Capture and Utilization and Green Ammonia Production from a Cement Plant Flue Gas

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ABSTRACT

The manufacturing industry is the second largest emitter of CO_2, with the cement industry being one of the main contributors (7-8 % of the global emissions). Carbon capture and utilization (CCU) technologies are promising decarbonization solutions for the cement industry, addressing both fossil fuel-related (40 %) and process-derived emissions (60 %). Within a cement plant, producing synthetic natural gas (SNG) from captured CO_2 is particularly suitable, as it is sufficient to fully replace solid fuels in the rotary kiln. On the other hand, the use of zero-carbon fuels, such as green ammonia, is also recognized as a promising approach for decarbonization. In this work, a superstructure was developed to explore alternative routes for producing SNG and green ammonia from CO_2 and N_2 in cement plant flue gas, respectively. The routes were modelled in Aspen Plus® V14, and their economic viability was assessed. Currently, the most promising route, at a cost of 109 €/tonne of flue gas, involves CO_2 capture (90% efficiency) and storage, with the N2-rich stream emitted. However, all the routes remain uncompetitive compared with the full emission of the flue gas (22 €/tonne). Routes for green ammonia production from N_2 in flue gas were compared with those from atmospheric air but were found to be uncompetitive. Nevertheless, with future developments, the economic feasibility of integrating pathways for synthesizing high-value chemical products into the cement industry is expected to increase.

Keywords: Cement industry, SNG production, Green ammonia production, Aspen Plus, Techno-economic analysis

INTRODUCTION

Climate change continues to pose a pressing global challenge, driving extensive research and policy initiatives worldwide. The planet's temperature has been exponentially increasing over the past decades, reaching the value of 1.28 °C above the pre-industrial average in 2024 [1]. This translates to a new record high of 37.8 Gt of CO_2 emitted [2], which takes responsibility for global warming, as it is the main greenhouse gas (GHG). Relocating electricity and heat consumption to their respective sectors, the industry sector becomes the most impactful one for GHG emissions, accounting for 39% of global CO_2 emissions. The cement industry alone is responsible for around 7-8% of GHG emissions, as CO2 is a direct by-product of its production process, which explains the high CO_2 content in its flue gas. Carbon capture, utilisation and storage (CCUS) is considered a promising strategy for reducing these emissions. The literature on the treatment of cement plant flue gas focuses primarily on carbon capture and storage technologies [3]. However, studies have reported the possibility of producing high-value chemicals, such as formic acid, dimethyl ether and synthetic natural gas (SNG), despite having obstacles in terms of high energy and raw materials consumption [4]. Among these carbon utilization strategies, the conversion of the captured CO_2 into SNG through the Sabatier process, with the addition of green hydrogen and in the presence of metal catalysts, is particularly suitable for the cement industry, as the

produced SNG is enough to fully replace the solid fuels fed into the rotary kiln and precalciner of a cement plant [5].

Since the cement plant flue gas is composed of not only CO_2, but also N_2 and some minor compounds, the obtention of a high-purity CH_4 stream usually requires pre-treatment of the flue gas to remove NO_x, SO_x, HCl and dust particles and an additional capture unit for N_2 separation from CO_2 [6]. The best available technologies for pre-treatment include wet scrubbing with hydrated lime and selective catalytic reduction (SCR) for SOx and NOx removal, respectively, while the most mature carbon capture technology is absorption with monoethanolamine (MEA) [7-9].

Traditionally, ammonia (NH_3) is synthesized through the Haber-Bosch (H-B) process, which relies on fuel gasification, leading to large CO_2 emissions. Power-to-ammonia technologies offer a sustainable alternative, requiring nitrogen (N_2), usually obtained from an air separation unit (ASU), and green hydrogen from water electrolysis, which are both highly energy-intensive [10]. Because common cement plant flue gas has a high N_2 content, utilizing this component in an ammonia production process offers a promising alternative to bypass the ASU and reduce energy consumption. This way, transforming N_2 into NH_3, a versatile compound with fertilizer and energy applications, highlights the potential of generating valuable products from the cement plant flue gas.

Despite their potential, CCU and green ammonia production face economic and technical challenges, mainly due to the high cost of green hydrogen. Existing pilot plants study these technologies separately, missing potential synergies between the processes. This work introduces a novel approach by using industrial flue gas as an N_2 source, thereby eliminating the need for an ASU.

METHODS

The flue gas of a cement plant was considered to go through a set of processes defined in a superstructure to capture and either store or convert CO_2 into SNG and N_2 into ammonia. These processes were modelled using Aspen Plus® v14 and the techno-economic viability of each route of the superstructure was evaluated. The heat exchangers of the processes considered were designed using Aspen Exchanger Design & Rating (EDR)® [11].

Cement Plant Flue Gas

The typical composition of a cement plant flue gas, which was used as a basis for this work, was retrieved from the work of Schakel et al. [6]. However, only 10 % of the total mass flowrate of the cement plant flue gas was considered, to avoid ammonia market saturation.

Superstructure

The most mature technologies were integrated to assess the techno-economic feasibility of converting a typical cement plant's flue gas into SNG and green ammonia, reaching the superstructure shown in Fig. 1.

Before feeding the production processes, the flue gas must undergo pre-treatment to remove NO_x, SO_x, HCl, and dust particles. After this process, four options were considered: the separation of CO_2 from N_2 with 90 % efficiency ('MEA absorber (90 %)') or with approximately 100 % efficiency ('MEA absorber (~100 %)'), no separation of these components or direct emission of the flue gas. The 'MEA absorber (~100 %)' process was investigated to obtain a CO_2-free outlet, which can be sent directly to the H-B process. In the MEA absorption units, the stream is separated into a CO_2-rich and a N_2-rich stream. The CO_2-rich stream can either be sent to the 'Sabatier' unit to generate SNG or to storage, while the N_2-rich stream can be valorized into green ammonia in the H-B process or emitted. If the high-purity N_2 stream comes from a capturing unit with 90 % efficiency, a methanation step is required beforehand, since the catalysts used for ammonia production are poisoned by oxygen-containing compounds [12]. The 'MEA absorber (~100 %)' option was therefore considered to avoid the need for this additional methanation step. Regarding the option of no CO_2/N_2 separation, the pre-treated flue gas stream is subjected to methanation for converting the CO_2 to methane and sent to the Haber-Bosch process to generate green ammonia. The ammonia is then separated from the methane through distillation. All these routes were compared with the full flue gas emission and with the production of green ammonia through the hydrogenation of atmospheric air. Tab. 1 presents the nomenclature of the different superstructure paths.

Table 1: Nomenclature of the superstructure routes. MEA stands for monoethanolamine, and ASU stands for an air separation unit.

Separation	CO_2 stream	N_2 stream	Route
MEA absorber (90 %)	Sabatier	Haber-Bosch 1	$A90^{S}_{HB}$
		Emit	$A90^{S}_{E}$
	Storage	Haber-Bosch 1	$A90^{CCS}_{HB}$
		Emit	$A90^{CCS}_{E}$
MEA absorber (~100 %)	Sabatier	Haber-Bosch 2	$A100^{S}_{HB}$
		Emit	$A100^{S}_{E}$
	Storage	Haber-Bosch 2	$A100^{CCS}_{HB}$
		Emit	$A100^{CCS}_{E}$
No separation	Sabatier + Haber-Bosch 3		$A0^{S}_{HB}$
	Emit		E
ASU	-	Haber-Bosch 2	ASU_{HB}

The route nomenclature follows the rule: X^{i}_{j}. X represents the separation options of the pre-treated flue gas/air stream, i represents the options of the CO_2-rich stream generated after absorption and j represents the

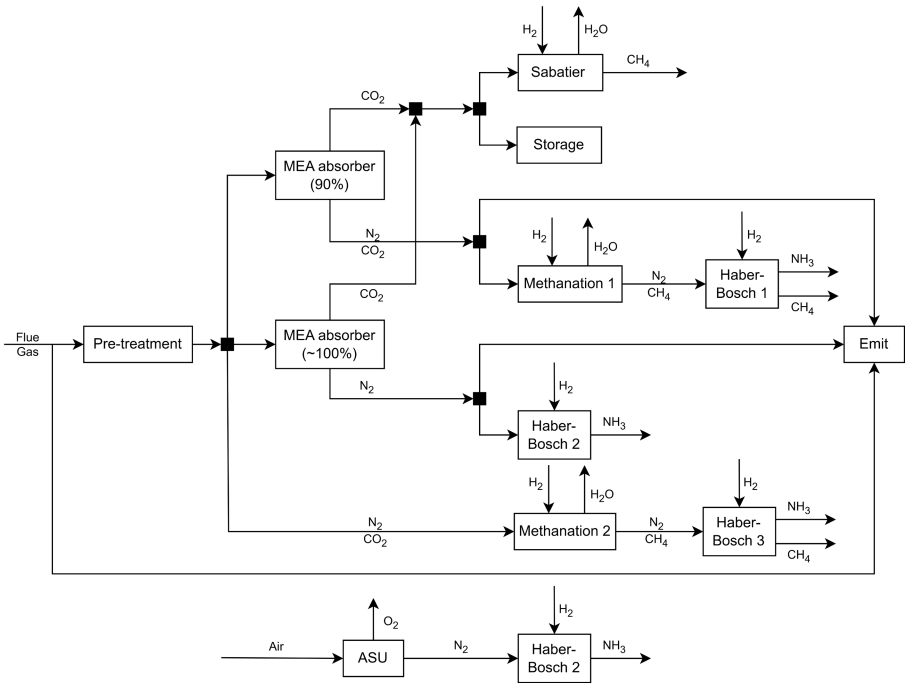

Figure 1. Superstructure for synthetic natual gas and green ammonia production from a cement plant flue gas.

options of the N_2-rich stream generated after absorption.

Pre-treatment

The pre-treatment process aims to remove O_2, NO_x, SO_2, HCl, and dust particles from the initial flue gas. For the removal of SO_2 and HCl, a 10 %wt. $CaCO_3$ aqueous solution was added to a scrubber, which was simulated in Aspen Plus® V14 by using a *RadFrac* model and implementing the equilibrium reactions described by Duque et al. [13]. The dust particles were also assumed to be removed by the scrubber. On the other hand, rate-controlled reactions were implemented to model the NO_x elimination with NH_3 in the SCR using a *RPlug* block, as suggested by Sweeney and Liu [14]. The O_2 elimination was modelled using a *RStoic* model, with its reaction with CH_4 following the work of Lu et al. [15]. The diameter of the scrubber was calculated by Eq. (1)

$$D_{scrubber} = \sqrt{\frac{4S}{\pi}} = \sqrt{\frac{4Q_{v,gas}}{\pi v_{gas}}} \quad (1)$$

, where $Q_{v,\,gas}$ is the gas volumetric flowrate and v_{gas} is the gas velocity inside the scrubber, which was considered 1 m/s [16]. The scrubber's height (L) was attained using Eq. (2) [16].

$$\frac{L}{D} = 3 \quad (2)$$

Regarding the SCR, a tubular flow reactor must be used, which has a usual residence time (τ) of 0.8 s [17]. This, combined with the gas flowrate going through the reactor, enables the calculation of the volume occupied by the gas (V_{gas}). The catalyst volume (V_{cat}), which can often be neglected for being considerably smaller than V_{gas}, can be calculated using the mass of catalyst inside the reactor (m_{cat}) and its density (ρ_{cat}), leading to Eq. (3).

$$V_{reactor} = V_{gas} + V_{cat} = Q_{v,gas} \bullet \tau + m_{cat} \bullet \rho_{cat} \quad (3)$$

Applying Eq. (2) to the formula of the volume of a tubular reactor, it is possible to calculate directly the diameter of the reactor (D), reaching Eq. (4) [16].

$$D = \sqrt{\frac{4\,V_{reactor}}{\pi L}} \quad (4)$$

Regarding the removal of O_2, Lu et al. [15] also suggests a tubular flow reactor, which can be designed using the previous expressions.

MEA absorption

The goal of the MEA absorption process is to separate N_2 from CO_2 in the treated flue gas. For this, two MEA absorption systems were simulated: one with 90 % carbon capture efficiency and the other with an almost complete CO_2 capture. The two simulations differ only on the reboiler duty needed for stripping the CO_2 from the solvent, to regenerate it. An AspenTech example [19] was adapted with the chemistry and reaction models retrieved from the work of Madeddu et al. [9] to build a rate-based model. The main equipment consists of two packed columns, which were simulated as RadFrac models with Sulzer Mellapak 250Y packing: an absorber, where the flue gas enters and contacts with the MEA solvent, which absorbs the CO_2, and a stripper, where the CO_2-rich solvent is separated from the CO_2 and pumped

back into the absorber. The design of both the columns followed the method of Madeddu et al. [9].

Methanation

The Sabatier process can have two objectives: to solely produce CH_4 or to prevent the poisoning of the downstream Haber-Bosch system by converting CO_2 into CH_4. For the simulation, the reaction of Eq. (5) was implemented with the kinetics retrieved from the study of Koschany et al. [20] with a $NiAl(O)_x$ catalyst.

$$CO_2 + 4H_2 \rightleftarrows CH_4 + 2H_2O \quad (5)$$

Before the reactors, the stream is compressed to 7 bar, not only because the reaction is favoured by higher pressure, but also because the natural gas distribution system operates at this pressure, and the product will be at the required conditions for integration. To achieve this, a three-stage compression system with interstage cooling is implemented. After the stream is compressed and mixed with H_2 in a 1:4 mole ratio, it is heated to 290-320 °C and goes through 3-4 Sabatier reactors, simulated by RPlug models, with condensation between each stage. As in the work of Marnate et al. [21], multitubular fixed-bed reactors were chosen for methanation. The dimensions of each tube were fixed at 2 m in height and 1 in. in diameter. Using Eq. (3), the required reactor volume can be calculated, assuming a catalyst density of 2300 kg/m^3 [22]. The mass of catalyst was adjusted to maintain the gas flowrate at 120 NL/h/g_{cat} [20]. Dividing the required reactor volume by the volume of each tube yields the required number of tubes.

Haber-Bosch

The H-B process aims to produce NH_3 from N_2. An AspenTech example model [19] was adapted, and reaction of Eq. (6) was implemented with the kinetics studied by Tripodi et al. [23] in iron-based catalysts.

$$N_2 + 3H_2 \rightleftarrows 2NH_3 \quad (6)$$

The inlet stream is mixed with H_2 at a 1:3 molar ratio and then compressed to 150 bar to favour the reaction [21], using four compression stages with intermediate cooling. The stream is heated to 400-540 °C before entering each reactor, depending on the optimal temperature found. After passing through 3 reactors with intermediate cooling, the product is condensed at -35 °C, and liquid NH_3 is separated from the unreacted gases that must be recycled back into the first reactor. The design of the Haber-Bosch reactors involved a sensitivity analysis varying the mass of catalyst and checking its influence on the conversion of each reactor: when the maximum conversion was reached, the value for the mass of catalyst was found. This way, employing Eqs. (3) and (4), it was possible to compute the dimensions of each reactor, using 2900 kg/m^3 [24] as ρ_{cat}.

Economic analysis

The cost of each piece of equipment for each other process was primarily estimated using the Aspen Process Economic Analyzer® (APEA) [25]. The equipment costs were then updated to 2025 using the corresponding CEPCI value [26]. The capital expenditure (CAPEX) and the operational expenditure (OPEX) were calculated using the factorial method, as described by Timmerhaus et al. [27]. The CAPEX was annualized using a 0.098 factor, assuming a plant lifespan of 20 years and a discount rate of 7.5%. The OPEX was calculated based on the values in Tab. 2. Adding the OPEX to the annualized capital cost enables the economic analysis of every process and, consequently, every superstructure route.

Table 2: Main values for the calculation of the OPEX and of the revenues [28-34].

Parameter	Value	Unit
H_2 price	5.1	€/kg
Electricity cost	0.06	€/kWh
Cooling water	0.01	€/m^3
Refrigerant	2.8	€/GJ
Steam	0.04	€/kg
NH_3 value	521	€/t
CH_4 value	1806	€/t
O_2 value	290	€/t
Storage cost	30	€/t
Emission cost	70	€/t

Although the Air Separation Unit was not modelled using Aspen Plus® v14, the associated costs were also evaluated. In this case, Eq. (7) [35] was used for the estimation of the equipment cost (EC), being $\dot{m}_{O_2}$ the mass flowrate of O_2 entering the system, in tonnes per day.

$$EC\ (M€_{2014}) = 22 \bullet \left(\frac{\dot{m}_{O_2}[t/d]}{432}\right)^{0.6} \quad (7)$$

Electric and steam consumptions of 230 kWh/t_{O2} and 16.2 kWh/t_{O2} were considered, respectively [35]. The value of the mass flowrate of O_2 used was calculated considering the mass flowrate of N_2 that enters the 'Haber-Bosch 2' process and the mole fractions of N_2 (~79 %) and O_2 (~21 %) in the air.

Since not all superstructure routes produce the same products, these were assumed to be sold to enable an economic comparison of the routes. In this sense, the revenue was calculated by multiplying the generated flowrate by the respective unitary values. In routes that do not generate any product, such as $A90_E^{CCS}$, $A100_E^S$ or $A0$, no revenue was calculated, but the emissions/storage fees were applied. Adding the costs, and subtracting the revenue, leads to the overall annual cost of each route. This value is then divided by the initial mass flowrate of the cement plant flue gas (42.8 t/h) to assess the cost of treating it per metric tonne.

RESULTS AND DISCUSSION

The economic results are detailed and discussed in this section for each individual process considered in the superstructure and for each route of the superstructure.

Pre-treatment

The dimensions of the scrubber, SCR and O_2 removal reactor are presented in Tab. 3.

Table 3: Design of the main equipment in the pre-treatment section.

Equipment	Height/Length (m)	Diameter (m)
Scrubber	13.0	4.3
SCR reactor	5.8	1.9
O_2 removal reactor	2.4	0.8

After determining the equipment dimensions, the total estimated cost was calculated at approximately 1.56 M€. The simulation enabled the calculation of utility and raw material consumption costs, which totaled 7.9 M€/y, indicating a profit, since steam is generated in this process and can be used in downstream processes.

MEA absorption

Subsequently to the pre-treatment process, the resulting stream can go through MEA absorption that can either capture 90 % or approximately 100 % of the CO_2 present. The dimensions of the absorber and stripper columns are similar for both capturing efficiencies and are presented in Tab. 4.

Table 4: Design of the main equipment in the capturing section.

Equipment	Height (m)	Diameter (m)
Absorber	26.9	3.4
Stripper	12.0	2.6

The equipment costs for both processes were 3.4 M€, while the total expenses for raw materials and utilities were 9.7 M€/y for the 'MEA absorption (90 %)' process and 76.6 M€/y for the 'MEA absorption (~100 %)' process. The large difference in operating costs reflects different steam consumption needs: the last process requires a higher heat duty to regenerate the MEA solvent because it absorbs more CO_2.

Methanation

The Sabatier processes have the objective of solely producing CH_4 ('Sabatier') or of eliminating the CO_2 content of the stream ('Methanation 1' and 'Methanation 2'), readying the stream for a subsequent H-B process. The design of these processes is presented in Tab. 5.

Table 5: Number of tubes (N_{tubes}) and mass of catalyst ($m_{catalyst}$) required for each reactor (R) of the Sabatier (S), Methanation 1 (M1) and Methanation 2 (M2) processes.

	N_{tubes}			$m_{catalyst}$ (kg)		
	S	M1	M2	S	M1	M2
R1	1855	1346	3151	240	185	451
R2	665	1300	1860	86	162	240
R3	460	1370	1768	56	161	215
R4	-	-	1867	-	-	214

The estimated equipment costs are 2.0, 3.7, and 5.2 M€ for the 'Sabatier', 'Methanation 1', and 'Methanation 2' processes, respectively, as they progressively handle higher flow rates and increase the dimensions of the respective equipment. Regarding the consumption of utilities and raw materials, the respective values are 86.2, 10.6, and 96.5 M€/y for the processes. The values differ due to the different needs for green H2 in each process: the inlet stream of the 'Methanation 1' process contains approximately 10 % CO2 relative to the other methanation processes, leading to the lowest green hydrogen consumption and thus the lowest operating costs.

Haber-Bosch

The design of the reactors of the Haber-Bosch process is presented in Tab. 6.

Table 6: Reactor (R) design and mass of catalyst ($m_{catalyst}$) required for each reactor (R) of the Haber-Bosch (HB) processes.

	Diameter (m) x Length (m)			$m_{catalyst}$ (kg)		
HB	1	2	3	1	2	3
R1	1.0x3.1	1.0x3.0	1.4x4.2	2	2	13
R2	1.3x3.9	1.2x3.7	1.8x5.4	10	9	34
R3	1.6x4.8	1.5x4.6	2.6x7.9	24	21	120

The estimated equipment costs for producing NH_3 through the H-B process are 11.7, 11.6, and 18.5 M€ for the 'Haber-Bosch 1', 'Haber-Bosch 2', and 'Haber-Bosch 3' processes, respectively. The 'Haber-Bosch 3' has the highest equipment costs because it handles the largest amount of inert gas (CH_4). This translates into the largest circulating flowrates and, consequently, the largest equipment dimensions.

As for the consumption of utilities and raw materials, all three processes have a value of approximately 211 M€/y, as their green H_2 consumption is similar because their respective inlet streams contain equal amounts of N_2.

Air Separation Unit

The estimated equipment cost of the ASU is 17.3 M€, while the estimated annual cost of electricity and steam is 0.9 M€.

Superstructure

Tab. 7 summarizes the results of the economic assessment for each possible superstructure route, each designated by its respective nomenclature (see Tab. 1).

It is evident that OPEX has a stronger influence on production costs than CAPEX, which is expected, since annualized CAPEX is 1-10% of OPEX. The high OPEX of the processes is mainly related to their high energy and green H_2 consumption.

Regarding route revenues, those that include the H-B process to produce green ammonia yielded higher revenues, due to the high N_2 content in the flue gas stream, which is converted into green ammonia. However, the routes that consider the emission of the CO_2-lean stream after the separation unit ($A90_E^S$, $A90_E^{CCS}$, $A100_E^S$ and $A100_E^{CCS}$), present fewer total costs than the ones that include a H-B system. This means that, despite having the highest revenues, the Haber-Bosch process is currently not economically viable, due to its large equipment costs and high green H_2 consumption.

When comparing the routes producing green ammonia from the cement plant flue gas with the benchmark technology to produce green ammonia (ASU_{HB}), the latter presents lower total costs, which means that it is currently still uncompetitive to utilize the N_2 from the cement plant flue gas instead of using an ASU.

Tab. 7 also reflects the lower costs of the routes including a CO_2 capturing unit with 90 % efficiency when compared with the ones comprising ~100 % MEA absorption systems. For example, route $A90_E^{CCS}$ is approximately 70 % less expensive than $A100_E^{CCS}$. This is mainly related to the high medium-pressure (MP) steam consumption required in the MEA absorption unit to capture approximately 100 % of the CO_2 emissions.

Among the routes that produce both green ammonia and SNG, route $A100_{HB}^S$ has the highest revenue value due to producing the largest amount of methane and ammonia. However, it also presents the highest total cost, since it not only requires the high green H_2 consumption of both the Haber-Bosch and the Sabatier processes, but it also contains the 'MEA absorption (~100 %)' system, which has a high utility demand. Route $A90_{HB}^S$ appears to be a more promising option for producing these two products, since it has lower energy requirements in the CO2-capture unit, leading to significantly lower OPEX. On the other hand, the route $A0$ exhibits the lowest OPEX value of the three, as it does not contain a CO_2 capturing unit. For this same reason, it needs to handle the total flue gas flowrate, which requires larger equipment dimensions, therefore it presents the highest CAPEX. The $A0$ route is the one which presents the lowest revenues, due to the need of higher purge flowrates in the process to accommodate the high flowrate of CH_4 in the H-B process. The total net cost of treating the flue gas in this route is similar to the $A90_{HB}^S$ route, indicating that the $A90_{HB}^S$ and $A0$ are the most promising routes for the simultaneous production of SNG and green ammonia from a cement plant's flue gas. However, these two routes present total costs approximately 40 times higher than the emission of the cement plant flue gas. This low economic attractiveness of the simultaneous production of e-methane and green ammonia from the cement plant flue gas is mainly related to the high costs of green hydrogen.

Although not producing methane or ammonia, route $A90_E^{CCS}$ is currently the less expensive path to treat the cement plant flue gas. Nonetheless, it is worth noticing the relatively low total cost of path $A90_E^S$, which includes methanation of the CO_2-rich outlet of the 'MEA absorption (90 %)' process and emission of the CO_2-lean outlet stream, as it presents the second lowest total costs while producing SNG through the Sabatier process. However, under current conditions, no route is economically competitive with the cement plant's total emissions from flue gas. The greatest contributor to the high cost of this route is the cost of green H_2, which is expected to decrease in the coming years [36]. By analysing the economic feasibility with the green H_2 price evolution, it is possible to conclude that the $A90_E^S$ route becomes the most profitable route when the green H_2 price reaches 1.81 €/kg and that its break even H_2 price is 1.51 €/kg of H_2.

Conclusions

Given the cement industry's elevated GHG emissions, this work proposes applying SNG and green ammonia production processes to cement plant flue gas. The most mature technologies were investigated and assembled into a superstructure comprising different routes, which were modelled using Aspen Plus®. Firstly, a pre-treatment was employed to remove unwanted components from the flue gas. Then, the possibility of separating N_2 from CO_2 was studied by applying a MEA absorption system with either 90 % or ~100 % capturing efficiency.

The obtained N_2-rich stream could be valorized to NH_3 via the Haber-Bosch process or emitted, while the CO2-rich stream could be sent to storage or converted to CH_4. Comparing the MEA units studied, as the complete CO_2 removal is approached, the system requires higher reboiler duties to strip the CO_2 from the solvent due to a substantial increase in MP steam consumption. Thus, the 90 % capture configuration appears to be more energetically and economically advantageous. The ~100% separation unit has the advantage of not requiring a subsequent methanation step to treat the N^2-rich stream before it enters the Haber-Bosch process. However, the cost increase of applying a ~100 % separation unit is greater than the cost of adding an extra

Table 7: Economic analysis of the superstructure routes.

Route	$A90^{S}_{HB}$	$A90^{S}_{E}$	$A90^{CCS}_{E}$	$A90^{CCS}_{HB}$	$A100^{S}_{HB}$	$A100^{S}_{E}$	$A100^{CCS}_{E}$	$A100^{CCS}_{HB}$	$A0$	ASU	E
EC (M€)	22.4	6.9	4.9	20.4	18.6	7.0	4.9	16.5	25.2	28.9	-
CAPEX (M€/y)	13.1	4.1	3.0	12.0	11.0	4.1	3.0	9.7	14.8	17.3	-
OPEX (M€/y)	477.7	152.0	31.0	356.7	587.5	280.2	122.9	430.1	468.9	332.7	-
CH_4 sales (M€/y)	66.5	66.5	-	-	73.8	73.8	-	-	62.4	-	-
NH_3 sales (M€/y)	111.1	-	-	111.1	113.6	-	-	113.6	101.9	113.6	-
O_2 sales (M€/y)	-	-	-	-	-	-	-	-	-	16.3	-
Total sales (M€/y)	177.6	66.5	-	111.1	187.4	73.8	-	113.6	164.3	129.3	-
Emissions (M€/y)	-	0.7	0.7	-	-	0.0	0.0	-	-	-	7.5
Storage (M€/y)	-	-	3.0	3.0	-	-	3.3	3.3	-	-	-
Total costs (M€/y)	313.2	90.3	37.6	260.5	411.1	210.6	129.1	329.5	319.5	220.7	7.5
€/t of treated FG	910	263	109	757	1195	612	375	958	928	-	22

methanation system.

The routes considering carbon capture with 90 % efficiency, with the emission of the N_2-rich stream and either the methanation or storage of the CO_2-rich stream are the most economically promising alternatives to treat the flue gas stream, with respective total costs of 263 and 109 €/t of flue gas. However, they are currently uncompetitive due to flue gas emissions (22 €/t).

Nevertheless, it is possible that H2-requiring routes become economically feasible as green hydrogen costs are expected to decline in the coming years. This trend indicates great potential to integrate pathways for synthesizing high-value chemical products into the cement industry, contributing to GHG mitigation and the transition to a more sustainable industrial sector.

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