C2PAT – Carbon to Product Austria

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Abstract: In this paper, a power-to-liquid pilot plant including carbon capture, hydrogen production by electrolysis and conversion in a reverse water-gas shift and Fischer Tropsch synthesis is simulated in ASPEN Plus®, investigating the changes in recycle streams and catalyst properties such as CO conversion and CH_4 selectivity. An investment and operating cost calculation is also performed to show the differences of economic considerations. In the end, a proposal for a PtL plant is made, considering the best efficiencies, product quantities and cost calculations. A process design with full gaseous recycle stream from Fischer Tropsch synthesis back to reverse water-gas shift reactor provides the best key performance indicators.

Keywords: Power to Liquid, reverse water-gas shift (rWGS), Fischer Tropsch Synthesis, Syncrude Production

1 Introduction

1.1 Motivation for the project "C2PAT – Carbon to Product Austria"

The effects of global warming are undisputed and are currently evident in the increase in catastrophic events, also in Austria. Due to the increasing cement production in the current construction boom, this industrial sector contributes a considerable share to carbon dioxide emissions into the atmosphere. In order to be able to achieve the national and international climate targets by 2030 and 2040, the "Carbon to Product Austria (C2PAT)" project is a first milestone towards climate neutrality in the cement industry. The use of carbon capture and utilization (CCU) is intended to reduce carbon dioxide emissions and counteract climate change. In the C2PAT project, nationally and internationally known major companies from a wide range of industrial sectors, such as Lafarge Zementwerke GmbH, Verbund AG, OMV AG and Borealis AG, are uniting to implement a power-to-liquid (PtL) plant in Austria.

The use of CO_2 from cement plant off-gases is very suitable for CCU due to its high concentration of about 14 vol.-%. The main component in the cement manufacturing process is limestone, which is burned to clinker in a rotary kiln at nearly 1,450 °C. About one third of the carbon dioxide emissions come from the firing of substitute fuels (plastic flakes, tire lint, and others) for heat input into the system, the rest results from the process-related CO_2 generated during the burning of the limestone mix in the rotary kiln.

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1.2 Project demonstration and implementation

Basically, a "Carbon Capture and Utilization" pilot plant is to be implemented in the C2PAT project at the Mannersdorf am Leithagebirge cement plant (Lower Austria). Carbon dioxide is to be captured from the cement plant exhaust gas and used as a feedstock for the production of renewable based plastics.

Carbon dioxide will be separated from cement plant off-gas in an amine-based scrubber and concentrated. Green hydrogen will be produced locally at the Mannersdorf site using an electrolysis system powered by electricity from a newly constructed PV park and renewable electricity from the grid. In a "reverse water-gas shift" reactor (rWGS) and a downstream Fischer-Tropsch reactor, carbon dioxide will react overall with green hydrogen to produce the product "syncrude". This syncrude will consist of a wide range of mainly normal paraffin-rich hydrocarbons and is separated into the fractions naphtha, middle distillate and wax. Further processing will take place at the Schwechat (Austria) and Burghausen (Germany) refineries, where the fractions will be processed in steam crackers to ethylene and propylene, and the end products polyethylene and polypropylene are produced in a polymerization step. This closed-loop system, from the burned substitute fuel via a carbon capture and utilization plant, to the production of renewable based plastics, is unique and has not been implemented before.

At the cement plant in Mannersdorf am Leithagebirge, the capture of CO_2 from the cement plant off-gas with a subsequent PtL pilot plant with an annual intake capacity of 10,000 tons of CO_2 is to be built. During the operation of the pilot plant, critical areas of the process route, long-term behavior and the stability or dynamics of the processes will be investigated and further developed. The pilot plant is a preliminary project in which the results of the investigations will be used to improve the equipment used, the optimized connection to the cement plant, the use of electrolysis and the chemical reaction processes to produce plastics. This project is a prerequisite for the implementation of the PtL process on a large industrial scale at the Mannersdorf cement plant, where the annually emitted CO_2 quantity of 700,000 tons is to be converted into renewable based products.
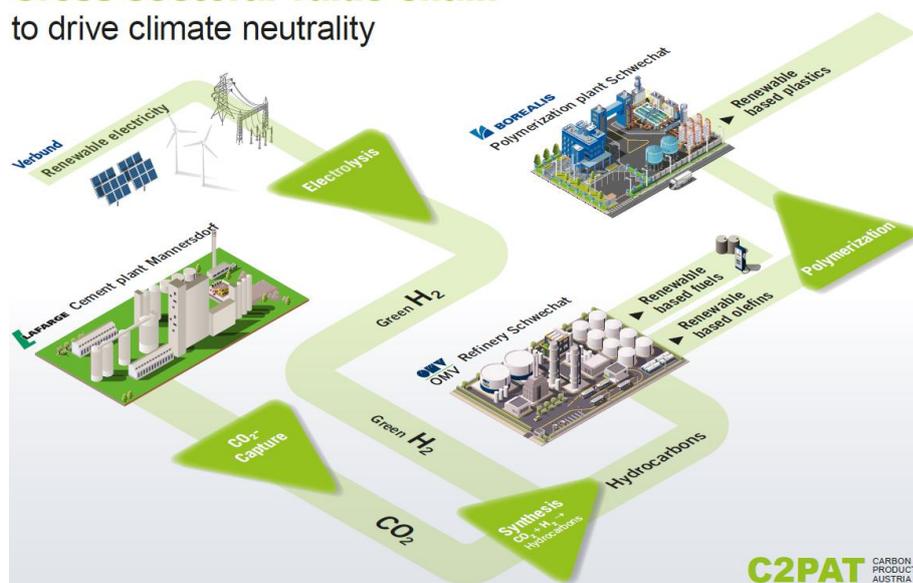
Figure 1: Schematic diagram of the new planned power-to-liquid plant in Lafarge's cement plant Mannersdorf am Leithagebirge in Austria (source: www.lafarge.com)

1.3   Process description

The block flow diagram shows the structure of a PtL process, which was also used in this analysis. It involves solvent-based CO_2 capture, followed by the reverse water-gas shift reaction fed with pure CO_2 and green hydrogen. The production and separation of hydrocarbons in a Fischer Tropsch synthesis is downstream. Highly concentrated carbon dioxide (14 vol.-%) from cement plant off-gas is used as feedstock for the solvent-based CO_2 capture. Separation of CO_2 and nitrogen using a MEA solvent is a proven process and is demonstrated in pilot [1] and full-scale plants [2]. Research on advanced solvents is expected to reduce the high energy requirements of the desorption process. Carbon dioxide and water are very stable molecules that require high energy to convert to carbon monoxide and hydrogen. Since a ratio of $CO:H_2 = 1:2,1$ must be fed to the Fischer Tropsch synthesis, carbon dioxide must be converted to syngas in an additional process step. The hydrogen is produced from renewable energy sources to earn the designation "green hydrogen". Production in alkali (AEL) and proton exchange membrane (PEM) electrolyzes is well established, while the level of development of high temperature electrolysis is low [3].

The conversion of CO_2 and H_2 in an rWGS reactor is also still subject of research. Unde has studied this heterogeneous catalytic reaction using nickel- and alumina catalysts [4]. Adelung et al. [5], König et al. [6] and Unde [4] analyzed the operation conditions for a rWGS reactor under special boundary conditions. Temperature and pressure are adjusted depending on the outcome of the desired products, in Adelung et al. and König et al. the production of kerosine is the goal of the PtL plant. To achieve high yield of desired hydrocarbons for kerosene (C9-C16), the chain growth probability of the catalyst is assumed to be 0.839 and 0.85 [5,6]. When Fischer Tropsch syncrude is used as a feedstock for plastics production, the desired product distribution differs from kerosene or diesel production. Performing such low temperature Fischer Tropsch synthesis enables the use of syncrude for many product pathways. By operating with high chain growth probability, the highly paraffinic product can be processed into e-fuels with hydrocrackers on the one hand, and on the other hand, it can be integrated into existing refinery equipment (e.g., steam cracker) as a renewable based feedstock for the production of renewable based plastics.
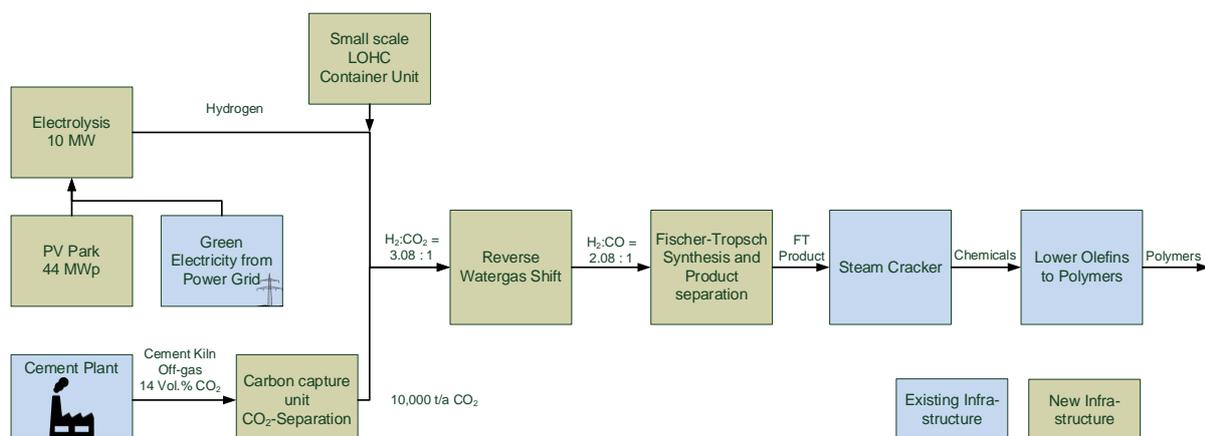
Figure 2: Block-flow-diagram to produce polypropylene and polyethylene from a cement plant off-gas

1.4 Aim of this work

As mentioned earlier, Adelung et al. and König et al. used the same process flow structure of a power-to-liquid process, but the application of operation conditions for the rWGS reactor differs [5,6]. The integration of a rWGS reactor into a pilot plant is new, so the investigation of real operation conditions has not yet been performed. The introduction of recirculating streams around the Fischer Tropsch synthesis and back to the rWGS reactor may affect the size of both reactors and additional piping. This must also be considered when calculating investment costs. In order to compare the entire value chain, the PtL plant, consisting of carbon capture, rWGS with Fischer Tropsch synthesis, and product separation, is simulated in ASPEN Plus®. Since CO conversion and CH_4 selectivity vary over a wide range (20-80 % and 6-12 %, respectively), the impact of these changes must be analyzed in a sensitivity study [7]. High CH_4 selectivity significantly reduces the product yield. The effects of variations in the recycle streams of the PtL routes needs to be compared, to define the best case for Fischer Tropsch synthesis resulting in the highest renewable based syncrude yield. In addition to typical key performance indicators (KPIs) such as PtL-efficiency, chemical and carbon conversion, a cost calculation of the pilot plant is performed to provide realistic process comparison of syncrude production.

2 Process description

In the following chapter, the carbon capture, electrolysis and synthesis technologies used are explained and the main simulation assumptions are given.

2.1 Amine scrubber unit

The carbon dioxide removal process is the first step of the value chain. The exhaust gas from the cement plant, which consists of about 14 vol.-% CO_2, is fed to an amine scrubbing system. The exhaust gas contains traces of impurities such as SO_2, NO_x, which originate from fuel combustion for heat supply and clinker production in the rotary kiln. Therefore, a pre-cleaning system must be installed to remove these unwanted impurities to a level that meets the amine solution manufacturer's specification. These pre-cleaning systems are not considered in the simulation. A clean exhaust gas is assumed, consisting of CO_2, N_2, O_2, steam and dust. In a pre-washer, the exhaust gas is cooled to an outlet temperature of 40 °C with a sodium hydroxide solution, which binds dust and cleans the gas stream of residual sulfur oxides. Sulfur oxides lead to the degradation of the amine, and the replacement in form of makeup is unfortunately higher [8]. In the absorber tower, CO_2 is dissolved in the amine solvent, while in the desorber tower, CO_2 is released and then steam is condensed to obtain CO_2 pure gas (>95 wt.-% CO_2). The desorption process requires a heat input of 3.8 MJ/kg CO_2 (e.g. steam from cooling system of the Fischer Tropsch synthesis) supplied by a reboiler [9]. The lean absorbent is recycled through a heat exchanger to close the amine scrubber unit circuit [10].

2.2 Hydrogen production

In the Fischer Tropsch synthesis, hydrogen and carbon monoxide are the reaction partners for the production of hydrocarbons. The definition green hydrogen means that the electrical energy comes from renewable energy production. In Austria today, these carbon dioxide

neutral production plants consider electricity generation for example from wind, photovoltaic or hydroelectric power generation. Two types of low temperature electrolysis (alkali- and proton exchange membrane electrolysis) are commercially available on the market. The high temperature electrolysis is still under development, which makes it difficult to implement in a pilot plant. The water is taken from a well and pretreated in a water treatment plant. The by-product oxygen is not used in this consideration, but could be part of the business plan calculation for large production volumes.

The main advantages and the choice of PEM electrolysis are the higher applied current of each cell to develop a lower overall specific power consumptions and high-pressure stream outlet (30-40 bar (g)) without compression. Dynamic operation (start/stop time) is also an advantage of PEM electrolysis used in combination with a cement plant because of unplanned and short stops due to cement production problems [11].

A low temperature PEM electrolysis with an operating temperature of 75 °C and an outlet pressure of 30 bar (g) is integrated in the simulation. The overall specific energy consumption is 4.7 kWh/Nm³ H_2 [3]. The injected water has to be treated first; a detailed water treatment plant is included in the simulation. A simple assumption is to double the water flow used for the electrolysis for operation cost calculation. Also, the electrolysis requires cooling water for constant hydrogen production. For heat integration, a constant cooling amount of 17 % of the electric power demand is assumed [12].

2.3   LOHC release unit

For demonstration and development purposes, a LOHC (liquid-organic-hydrogen-carrier) release unit should be installed and integrated as an option for hydrogen supply. Local hydrogen production will not be feasible for a full-scale PtL plant, because the local grid infrastructure is not well developed to operate a 700 MW electrolysis plant. Therefore, LOHC release studies and process route implementation research open the possibility for hydrogen import and use, avoiding local production near the cement plant. Due to the small share of released hydrogen from the LOHC unit in the total hydrogen consumption, the LOHC unit is not included in the simulation.

2.4   Syngas production

The separated carbon dioxide and green hydrogen must be converted into synthesis gas in an rWGS reactor. Therefore, Adelung et al. and König et al. found their optimal operation conditions of 825 °C and 5 bar and 900 °C and 25 bar, respectively. The rWGS reaction is not well established, so a new optimization of rWGS conditions is performed in this work. Due to thermodynamic equilibrium, CO_2 and H_2 favor conversion to syngas at high temperatures and low pressure. Moreover, energy in form of heat must be supplied to drive the endothermic heterogenous catalytic reaction.

$$CO_2 + H_2 \leftrightarrow CO + H_2O \qquad\qquad \Delta H_r^0 = +41 \; kJ/mol \qquad\qquad\qquad \text{Equation 1}$$
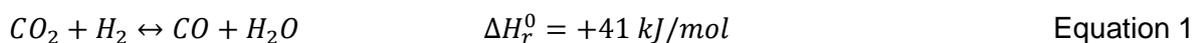
Methane and higher order gaseous hydrocarbons can affect the reaction as a recycle stream from the product separation unit back to the rWGS reactor is installed. Steam (Equation 2) and dry (Equation 3) reforming can also occur, and these reactions are highly endothermic, which can result in higher heat requirements compared to the rWGS-only reaction. Theoretically, this

process design allows for a higher total carbon conversion since enrichment of inert gas with methane is avoided in the Fischer Tropsch synthesis [13].

$$CH_4 + H_2O \leftrightarrow CO + 3H_2 \qquad \Delta H_r^0 = +205 \; kJ/mol \qquad \text{Equation 2}$$

$$CH_4 + CO_2 \leftrightarrow 2CO + 2H_2 \qquad \Delta H_r^0 = +247 \; kJ/mol \qquad \text{Equation 3}$$
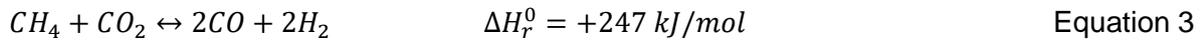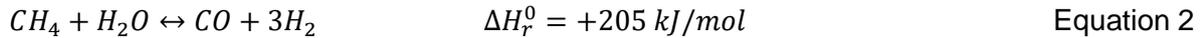
It is very important to consider all reactions that may occur in the rWGS reactor. If the reactor is operated under unfavorable conditions, undesirable reactions such as the Sabatier and methanation reactions, as well as carbon formation according to the Bosch reaction and Boudouard equilibrium, can deactivate the catalyst in the rWGS reactor. Due to reactions 2 and 3, the operation conditions are also pressure dependent. The rWGS reaction proceeds as a constant-volume reaction and is pressure-independent, shifting the equilibrium of all other reactions according to the Le-Chatelier principle. In general, higher pressures lead to higher conversion to undesired by-products.

The production of additional methane from CO_2 and H_2 lowers the CO selectivity. Figure 3 shows the thermodynamic equilibrium composition of the previously mentioned reactions. Higher Temperature and lower pressure result in better CO selectivity. The inlet gas stream has an H_2:CO_2 ratio of about 3:1, corresponding to the required stochiometric reaction conditions of the series of rWGS and Fischer Tropsch synthesis. Another important aspect is the shift and avoidance of carbon formation resulting from the excess of hydrogen in the rWGS reactor.
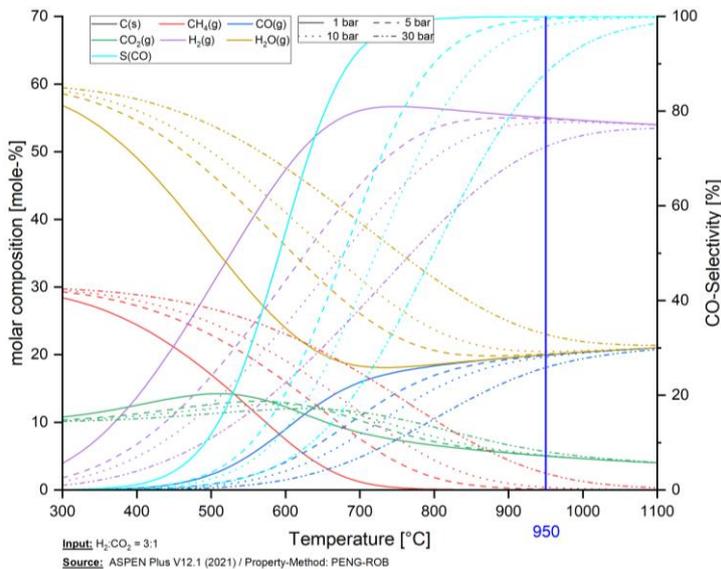
Figure 3: Thermodynamic equilibrium of the rWGS reaction, as a function of temperature and pressure

The pressure dependence can be clearly seen in Figure 3. Higher pressures generally lead to lower CO selectivity. This means that low pressure favors the desired syngas production. The PEM electrolysis has a hydrogen pressure output of about 30 bar (g), while the Fischer Tropsch synthesis usually operates at 25 bar (g) [11,14]. High pressure variations within the synthesis process chain require undesirable additional compression stages, and thus additional energy consumption. Therefore, the pressure of the whole process chain is set to 10 bar (g). The second observation is the increase of CO selectivity with higher temperature. Consequently, the operation temperature is set to 950°C. The reactor must be designed for

high temperatures. The selection of materials is similar to steam reformers, and the investment costs are considerable.

Nowadays, two techniques of rWGS reactors are developed and tested in laboratory scale. The main difference is the heat supply for the endothermic reaction. In addition to an autothermal reactor (fired by H_2), an electrified rWGS reactor has been developed and is used in this simulation. Electrically driven reactors are currently being developed [15,16].

2.5 Fischer Tropsch synthesis

The choice of a Fischer Tropsch synthesis (FTS) opens a wide range of product processing options. Integration into an existing refinery is possible, and the replacement of fossil crude oil contributes positively to the avoidance of greenhouse gases. The categorization of Fischer Tropsch technology depends on desired products and the operating conditions based thereon. The production of hydrocarbons from synthesis gas is strongly exothermic and is based on the following main chemical equation for alkanes as the main component [17].

$CO + 2H_2 \rightarrow$ -CH_2- $+ H_2O$ $\qquad \Delta H_r = $ -140 to -160 kJ/mol_{CO} $\qquad$ Equation 4
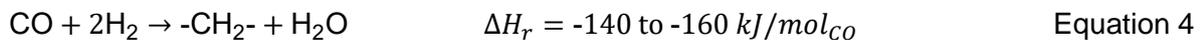
In general, the product composition of a Fischer Tropsch synthesis is based on the Anderson-Schulz-Flory (ASF) distribution. Depending on the chain growth probability α (usually about 0.9-0.95 for industrial scale FTS), the formation of desired hydrocarbon chain length is described [18]. The variation of temperature, pressure, catalysts, and H_2:CO ratio significantly affects the chain growth probability. Moreover, two main techniques are distinguished, Fischer Tropsch synthesis at low (220-230 °C) and high (320-340 °C) temperature. The advantage of a low-temperature FTS (LTFT) is the predominant production of long chain paraffinic hydrocarbons, while in a high-temperature FTS (HTFT) the main products are low-chain olefins. By-products such as oxygenates, aromatics and naphthene occur in both operating modes. In LTFT, Co- or Fe-catalysts are usually used. Due to its thermal stability, the implementation of Fe-catalyst has shown good performance in HTFT [17].

In this simulation, it is assumed that the syncrude will be integrated into an existing refinery. Therefore, a very simple pilot plant should be designed for syncrude production and transport by train or truck to the nearby refinery. The Fischer Tropsch synthesis operates with a carbon chain growth probability of 0.92, a temperature of 220 °C and an operating pressure of 25 bar (g). Product separation into naphtha (C5-C10), middle distillate (C11-C22) and waxes (C22+) is realized by using two flash units. The wax phase is taken off directly in the FTS reactor as a liquid phase. The gaseous product stream is cooled down to 100 °C that the middle distillate and partly steam condensates and is withdrawn as a liquid. Before entering the second separation stage, the gas stream is depressurized to 11 bar (g) and cooled to 20 °C. The separated liquid phase is naphtha and water, and the gaseous phase (including tailgas and LNG C1-C4) is a recyclable stream and is sent to a splitter. The splitter ratio is part of process optimization in this work. The products naphtha, middle distillate and wax are stored in storage tanks and prepared for further transportation.

2.6 Investment and operational cost estimation

For a meaningful comparison of the Fischer Tropsch routes, the capital and operational cost calculation also plays an important role. Increasing the recycle stream around the Fischer

Tropsch synthesis leads to higher equipment demand (in particular, the piping costs for a second recycle connection are higher) and a smaller size of the rWGS reactor. Therefore, the investment cost calculation can provide an economic statement about the construction of the plant.

The investment cost calculation is performed using Aspen Process Economic Analyzer V12® (APEA) with the latest database from Q1/2019. The current project is in the design phase, with an initial draft inside battery limits (ISBL) cost estimate using cost estimate class 5, i.e., calculated results may vary from -50/+100 % [19]. The inclusion of outside battery limits (OSBL) costs is not included due to uncertainties in the local infrastructure. For further and more detailed calculation, OSBL costs can vary from 10 to 100 % of ISBL costs, according to local conditions [20]. Chilton's method is used to account for detailed design changes in the investment cost calculation. The piping factor plays an important role in changing split scenarios, with the factor being higher when two recycle streams need to be installed. This is due to the complexity of the additional piping.

Calculating the capital cost for the main reactors, rWGS and Fischer Tropsch reactor, requires a detailed design study or literature data for evaluation. The structure of an rWGS reactor is similar to that of a steam methane reformer (SMR). For small-scale SMRs (0.15-15 MW), estimates of investment costs in the range of 3,000-5,000 USD/kW are available [21]. As mentioned in the rWGS chapter, in this simulation comparison, an electrified rWGS reactor is implemented in the simulation, where the size of the reactor becomes smaller according to Wisman et al. [16] Therefore, due to the smaller size and similarity with SMR, the assumption for the e-rWGS is one third lower than the literature values, namely 2,000 €/kW.

For the Fischer Tropsch reactor design, a gas hourly space velocity of 2,592 h^{-1} is reported for C_{5+} selectivity [22]. With this assumption, the number of tubes (diameter of 2.5 cm, length of 5 m) is calculated and APEA estimates a tube bundle reactor. Due to the same gas flow from split ratio 0 to 0.5, the cost estimate of 180,000 € is considered in both calculations.

High temperature (>950 °C) materials are used in the process, but are not available in APEA database, so changes to temperatures of 550 °C must be made for this comparison. Table 1 shows the assumptions for the main equipment used for the cost calculation.

The operation cost calculation is divided into the main and driver cost factor of electricity, followed by water, operations & maintenance (O&M) and personnel. The return on investment (ROI) in 12.5 years, the avoidance of CO_2 costs (actual cost of 80 €/t$_{CO2}$) and the newly introduced tax in Austria (30 €/t$_{CO2}$ stated in Austrian law "ÖkoStRefG 2022 Teil I") due to the lower carbon dioxide emissions are also considered in operation cost calculation [23]. Personnel costs include an operations manager and a maintenance team with five shifts.

Table 1: Assumptions on equipment and utility costs for the calculation of investment and operating costs

Equipment	Cost estimation		Utility category	Cost estimation	
PEM Electrolysis [3]	€/kW	1,400	Electricity costs [24]	€/MWh	80
e-rWGS reactor [21]	€/kW	2,000	Water cost [25]	€/m³	1.85
Fischer Tropsch reactor	€	180,000	Operational and maintenance cost [26]	% of CAPEX	4
Piping factor–recycle 0	%	35	Personnel costs	% of CAPEX	1.5
Piping factor–recycle 0.5	%	50	Return on Investment	%	8
			CO_2 price incl. tax	€/t_{CO2}	110

3 Simulation description

The PtL pilot plant is simulated in ASPEN Plus® V12.1. In general, the structure of the plant is divided into the amine scrubber unit (including the pre-wash unit), the electrolysis followed by the rWGS reactor and Fischer Tropsch synthesis including product separation and recycle stream. The general flowsheet is constructed in two different sections, using the ELECNRTL property method for the amine scrubber unit and the PENG-ROB (Peng Robinson) property method for the synthesis part. The following flow diagram shows the process flow from the cement plant off-gas to the desired product syncrude, which is separated into three fractions, namely naphtha (C5-C10), middle distillate (C11-C22) and wax (>C22). The effects of catalyst performance is simulated in a Fischer Tropsch conversion reactor, where the paraffine reaction and additional the methane reaction (to reach the methane selectivity) are implemented. A wide range of catalyst performance is reported in the literature. The data from Adelung et al. with a CO conversion of 40% and CH_4 selectivity of 16% are chosen as a starting point for this simulation [5]. A second operating point is chosen from the data of Ostadi et al. with a CO conversion of 64 % and a CH_4 selectivity of 8 % [7]. These two assumptions are also cross varied so that these data are used for a case study with four variants. The second simulation to be investigated is the variation of the split ratio in the product separation unit. The objective of this work is to calculate key performance indicators (KPIs) in dependence of catalyst and split ratio. The KPIs include power-to-liquid and carbon efficiencies, as well as specific energy consumption, product quantity, and cost calculation of each process variation.
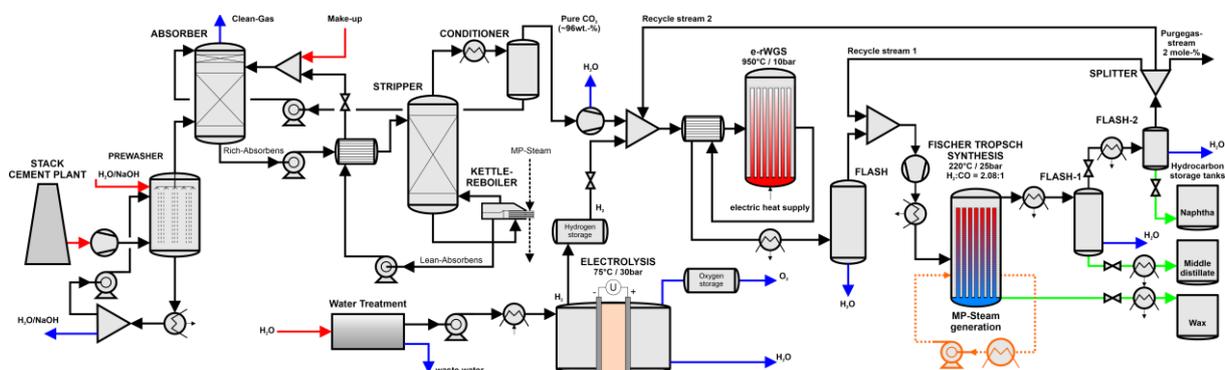
Figure 4: Simplified flowsheet of the ASPEN Plus simulation from cement plant off-gas, carbon capture unit, PEM electrolysis and synthesis with product separation to syncrude product storage

4 Results

For the correct set-up of a Fischer Tropsch process, the most important influencing factors must be evaluated and discussed in order to achieve high KPIs and product quantities. Figure 5 shows the KPIs and product quantity (A: PtL efficiency, B: Chemical conversion, C: Product quantity, D: Carbon conversion incl. carbon capture unit, E: Carbon conversion without carbon capture unit, F: Specific energy consumption). The KPIs are calculated as described in König et al. [6]

The first observation is the variation of the catalyst performance. Four scenarios with different CO conversion and CH_4 selectivity are simulated. As shown in the diagrams of Fig. 5, the impact of increasing split ratio from 0 to 0.5 leads to constant PtL efficiency, chemical and carbon conversion. Afterwards a decrease of all KPIs is visible, best seen in the product quantity. A full recycle stream (and 2 mole-% purge gas stream) results in high losses of desired product, high Fischer Tropsch recycle streams and an undesired high purge gas stream. This problem already occurs downstream of the rWGS reactor. A single pass of CO_2 and H_2 at a ratio of 1:3.08 through the rWGS-reactor produces syngas, maximal up to the thermodynamic equilibrium. Thus, the unconverted CO_2 continues to enter the Fischer Tropsch synthesis loop, where CO_2 is treated as inert gas, along with methane and higher order hydrocarbons. Due to the enrichment of these inert gases, the purge gas stream, set with 2 mole-% of the total gas stream, also draws off useful carbon dioxide and methane, resulting in poor KPIs and product quantities.

The figures represent that best KPIs and product quantity are achieved with a split ratio between 0 and 0.5. Higher split ratio leads to lower efficiencies, lower product quantity and higher specific energy consumption. No implementation of a recycle stream is due to very bad results unacceptable.
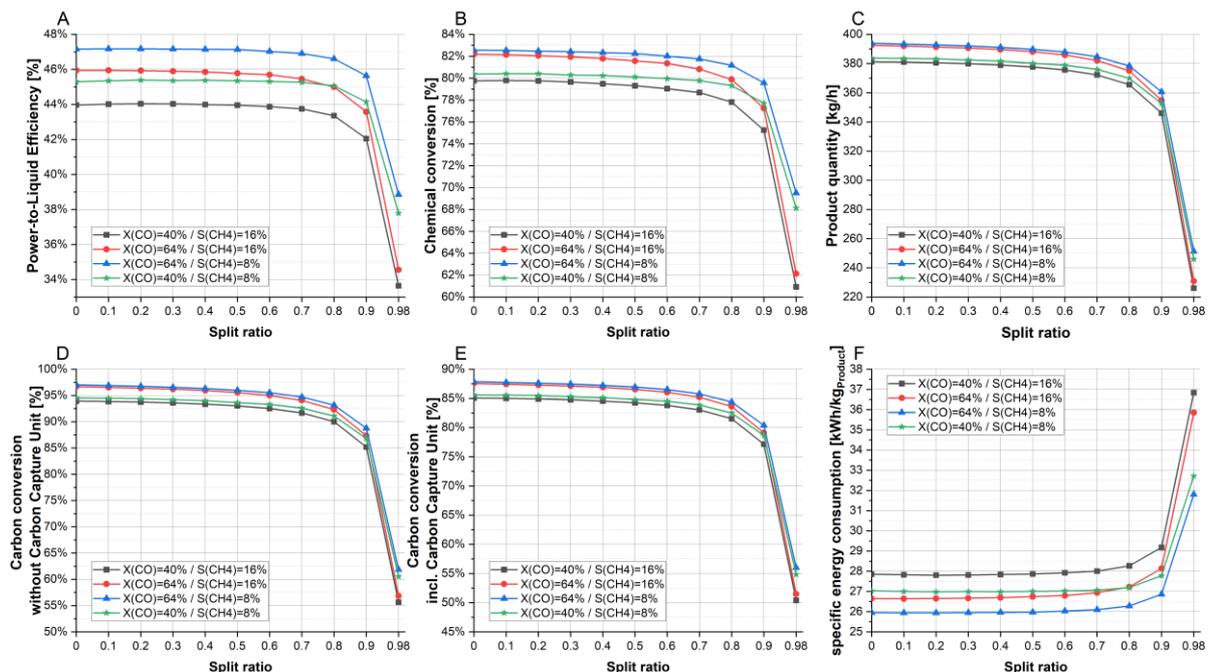
Figure 5: Schematic representation of efficiencies and product quantity as a function of catalyst properties and split ratio.

For a real and meaningful comparison, the piping costs for the synthesis part, calculations of investment and operating costs including return on investment (ROI) have to be performed. The investment costs are calculated for a split ratio of 0 and 0.5, which shows the difference between one and an additional second recycle stream. In the detailed Chilton calculation, the piping factor is increased by 15 percentage points from 35 % to 50 %. The 0.5 split ratio scenario, with a total volume of 46.7 MEUR, is 0.3 MEUR higher than the 0 split ratio, which has an investment volume of 46.4 MEUR. Operating costs including ROI remain constant for both split ratios at 13 MEUR per year. As already mentioned, the product volume decreases slightly with a higher split ratio.

5   Conclusion

In summary, a large number of factors play a role in the structuring of a power-to-liquid plant. In addition to the proper selection of a suitable catalyst, the interconnection of piping and equipment also leads to different results. The best application is to use catalysts with high CO conversion and low CH_4 selectivity, avoiding the production of useless gaseous byproducts such as methane and higher order lower hydrocarbons. KPIs and product quantity decrease with higher split ratio. Therefore, a value between 0 and 0.5 should be aimed for. Due to the slightly higher product quantity, lower investment and the same operating costs with a split ratio of 0, the recommendation for a new power-to-liquid plant to be built is: the PtL pilot plant should consist of an amine scrubbing unit for CO_2 capture, a PEM electrolysis for hydrogen production and an e-rWGS reactor with downstream Fischer-Tropsch synthesis and product separation, with unreacted gas being recycled to the rWGS reactor.

6   References

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