Final report

1.1 Project details

Project title	C3U, Cryogenic Carbon Capture and Use
Project identification (program abbrev. and file)	**64017-0026**
Name of the programme which has funded the project	**EUDP 2017-I**
Project managing company/institution (name and address)	Department of Energy Technology, Pontoppidanstræde 111, 9220 Aalborg Øst
Project partners	Aalborg Portland, Aalborg Energi Holding, EMD, Hydrogen Valley, Brigham Young University
CVR (central business register)	29102384
Date for submission	

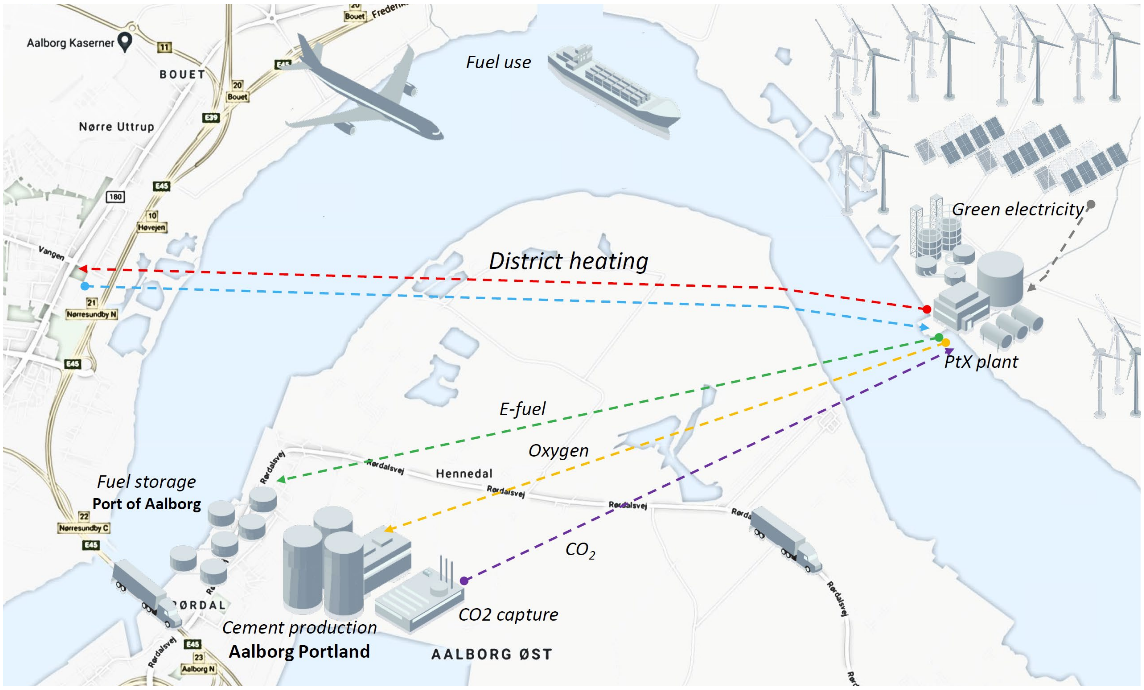
1.2 Short description of project objective and results

The main scope of this project was to establish techno-economic scenarios for grid balancing by cryogenic carbon capture coupled with synthesis of transportation fuels. As the first step, site specific data was collected to assess the process integration potential (capture and fuel synthesis) with the existing operation of the cement and CHP plants. The carbon capture technology was based on a recent break-through in cryogenic capture technology that has proven superior to any other capture technology in terms of energy efficiency and cost projections. The project validated the capture technology for the Aalborg Portland site and estimated capital cost for the installation. Experimental studies were made of CO_2 conversion into methanol by catalytic reaction with hydrogen. Topologies for the integration and further processing of the captured CO_2 to e-methanol was investigated in terms of cost and energy efficiency. The economic analyses were used to assess the feasibility of a demonstration plant located near Aalborg Portland. Overall, a feasible business case can be established provided a number of conditions are met including reduced grid tariffs and a willingness to pay a premium for e-methanol over fossil methanol.

Formålet med projektet var at gennemføre tekno-økonomiske studier af net-balancering ved hjælp af kryogen CO_2 fangst koblet med syntese af transportbrændstof. Som første skridt blev der indsamlet data om røggasmængder og sammensætning for at kunne undersøge integrationen af CO_2 fangst. I projektet undersøges kryogen CO2 fangst, da det har vist et potentiale for billigere og mere energieffektiv fangst af CO_2 fra røggasser. I projektet er CO_2 fangst fra cementfremstilling valideret på anlæg i USA og der er gennemført detaljerede procesberegninger og estimat på CAPEX for et anlæg designet specifikt til Aalborg Portland. I projektet er der gennemført eksperimentelle studier af metanolfremstilling fra CO_2 og brint ved forskellige procesbetingelser. Der er gennemført detaljerede simuleringer ved hjælp af EnergyPRO, som kobler det samlede CCU-anlæg til energisystemet i Aalborg og optimerer driften ud fra el spotpriser, varmeforbrug etc. Konklusionen på denne analyse er at det er muligt at opnå en favorabel business case under en række forudsætninger her under at der kan opnås reduceret el-tarif og at der kan findes aftagere til grøn metanol som vil betale en merpris i forhold til fossil metanol.

1.3 Executive summary

The overall vision for the integration of the C3U carbon capture and utilization plant in the local area around Aalborg Portland, Port of Aalborg and the Green Test Center at the North shore of Limfjorden (the Nordjyllandsværket site) is illustrated in the graphics below. The whole concept is a Giga-Scale green symbiosis that use the unique infrastructure available around Aalborg in terms of district heat and transmission grid connection points.



Aalborg Portland cement factory is among the largest CO_2 point sources in Denmark with an annual emission of 2,2-2,3 mio. tons. About 60% of the CO_2 originates from the calcination process and is unavoidable whereas the remaining CO_2 can be net zero emission through fuel switch and process modifications. A cryogenic carbon capture plant was designed to capture CO_2 from grey and white cement production, respectively. Depending on the required purity of the CO_2 and the fraction of CO_2 captured, the total cost of the plant is estimated at between USD 150 mio. and USD 220 mio. The carbon capture plant consumes about 5-6% of the energy of the total CCU plant, with the electrolyzer consuming the large majority.

Since green carbon-based fuels are required, in particular for aviation, there are good reasons to use the captured biogenic CO_2 but possibly also unavoidable CO_2 to synthesize green electro-fuels using renewable energy. One significant regulatory challenge in this regard is the fact that Aalborg Portland is not credited the CO_2 reduction if the CO_2 is used to produce green fuels that are to be considered CO_2-neutral.

In the C3U analysis, a fuel factory with nearly 1200 MW of electrolysis capacity was considered which is capable of converting up to 1,5 mio. tons of CO2/year into green e-methanol. Methanol can be used directly or further converted to aviation fuel.

A detailed model was set up in EnergyPRO including the district heating system in Aalborg and historical hourly electricity spot market prices based on which plant operation was optimized. The results were used in the techno-economic analysis. The analysis showed that the cost of electricity is the single most important factor influencing the plant economy. The cost of electricity consists of two components, the spot market electricity cost and the grid tariffs. Using historical spot market data from 2018 (high cost year) and 2015 (low cost year), the annual profit was found to vary by as much as a factor of two. A future electricity spot market price was also synthesized based on the 2018 data by amplifying the variations around the mean by a factor of 1,5. This increased the profit by 30% even though the number of operating hours per year decreased slightly. By reducing the transmission grid tariff from 80 DKK/MWh to 30 DKK/MWh, the annual profit increased by about 50%. The grid tariffs alone amount to 64 M€/year in the reference scenario using 2018 grid tariffs. Potential profits from grid balancing services was not considered and is regarded as a potentially very significant economic upside.

In all scenarios, the plant operates between 50% and 62% of the time and de facto becomes a very substantial grid balancing plant. As a consequence, only about 50% of the captured CO_2 is converted into fuel with a typical annual methanol production between 500.000 and 850.000 tons. The fuel factory can be oversized by as much as a factor of two to fully utilize the CO_2 although this will result in a significant amount of waste heat not being used for district heating.

The ability to sell the waste heat from the electrolyzer and the fuel synthesis plant is very important for the overall economy as it contributes 33 M€ to the annual income. With no heat sales, the net present value of the plant is reduced by 60%. Adding a seasonal heat storage, heat sales increases slightly above the reference case. The current district heating system in Aalborg is not able to absorb all the waste heat from the C3U factory and will constrain the number of annual operating hours unless heat ejection is allowed. Allowing the plant to eject heat, the income improves significantly as more low-cost electricity can be used. If significantly more efficient electrolyzers become available this will greatly influence plant economy and the reduced waste heat production will increase the ability to absorb this in the district heating system. In the analysis we used EU KPI data for alkaline electrolyzers in 2020 (50 kWh/kg H2).

A future scenario was investigated using the 2018 electricity spot prices with price variations amplified by a factor of 1,5, the fuel factory oversized by 50%, grid tariffs reduced to 30 DKK/MWh and the methanol selling price set at 400 €/ton. This resulted in a net present value (20 years horizon) of 332 M€ corresponding to and IRR of about 9%. Based on this analysis important conclusions can be drawn:

1. It is possible to reach near fossil price parity with the e-methanol provided grid tariffs are reduced significantly to acknowledge the very substantial balancing potential of 1200 MW.
2. It is a prerequisite that heat can be sold to the district heating grid. A possible oxygen sale will further improve the case.
3. A larger electricity price spread is required compared to historical spot market prices from 2018 allowing the plant to purchase sufficient low-cost electricity.

In conclusion, the analyses have shown great potential in the CCU plant with respect to grid balancing and production of green fuels at reasonable cost if the required regulatory conditions are established. Based on the findings of the C3U project, the partners formed a consortium to proceed with more detailed analyses of the technical, economic and environmental aspects of the concepts.

1.4 Project objectives

The objective of C3U was to explore an innovative Carbon Capture and Utilization (CCU) concept that potentially offers gigawatt-scale balancing power to the grid through energy efficient cryogenic carbon capture (CCC) technology coupled with electro-fuel production. Processing of the captured carbon dioxide to high value transportation fuels with hydrogen from water electrolysis will provide long term energy storage and secure green fuels for heavy duty vehicles, ships and aviation.

The main scope was to establish well documented and quantified techno-economic scenarios for grid balancing by cryogenic carbon capture coupled with synthesis of transportation fuels. The project leveraged a recent break-through in the cryogenic carbon capture technology that has proven superior to any other capture technology in terms of energy efficiency and cost. In addition, the cryogenic capture technology offers an inherent low-cost ability to provide energy storage services to the grid. Integration of the cryogenic process with transportation fuel production based on the captured CO_2 and hydrogen from electrolysis will add very significant balancing capacity and contribute with an important value stream to the system.

The scope of the project was addressed through the following activities:

- Detailed mapping of the plant data and CO_2 sources at the two sites in Northern Jutland including future development scenarios.
- Further development and optimization of the cryogenic capture technology towards the specific Danish energy systems and the considered plants
- Theoretical and experimental investigation of synthetic fuel production routes from the captured CO_2. Future development scenarios.
- Detailed studies of the balancing potential of the investigated plant concepts and techno-economic assessments. Feasibility study of large-scale demonstration.

Overall, the project has met the planned objectives but with a few changes in scope along the way. As the project progressed, it was decided to focus on Aalborg Portland as the CO_2 source since the plans to convert the local power station, Nordjyllandsværket, into biomass firing was abandoned. With this change in plans, Nordjyllandsværket will not be in operation after 2028 and hence not a relevant CO_2 source.

More time than expected was spent on the commissioning and debugging of the laboratory mini-plant to demonstrate the synthesis of different fuels from CO_2 and hydrogen partly due to technical issues with the process plant and partly due to the ongoing renovation of the laboratory building that delayed access to the facility.

There complexity of the European Emission Trading System (ETS), and the fact that there are ongoing discussions about the regulation of CO_2 utilization, made it impossible to reach the planned GO/NO-GO decision on a demonstration project. There was a clear need for further and deeper investigations of the complete business case including the perception of CCS and CCU by the customers for green cement and for the ETS discussions concerning CCU to reach a conclusion.

1.5 Project results and dissemination of results

This section presents the main project results in terms of the Cryogenic Carbon Capture process, the synthetic fuel production and the techno-economic analysis.

1.5.1 Cryogenic Carbon Capture

Sustainable Energy Solutions (SES) demonstrated the cryogenic carbon capture™ (CCC) process in the field at two different cement plants, removing over 90% of the CO_2 in a stream of CO_2 that is greater than 99% pure. The CCC process removes CO_2 from flue gas by cooling and drying the gas, desublimating the CO_2, separating the solid CO_2 from the light gas components, pressurizing the CO_2 stream, and warming all streams back to near ambient or initial temperatures. This relatively simple process requires only electrical power and flue gas. The flue gas passes through the system to produce a CO_2-depleted light gas at ambient conditions and a pressurized, liquid stream. This process produces high-purity CO_2 with capture percentages of 99+% at approximately half the cost and energy consumption as comparable amine processes. Nearly all the sensible energy needed for the process comes through heat integration of the inlet and outlet streams. However, the CO_2 enters the process at ambient conditions as a gas mixture and leaves the process as a pressurized, separated liquid. The energy and cost associated with pressurization are minimal since they occur in a condensed phase. Most of the process energy drives the phase change and separation, with some additional energy associated with heat losses and process inefficiencies. Nearly all this energy (approx. 90%) provides refrigerants. The sensible and especially the phase-change and compression energy required to generate these refrigerants provides them with a very high energy density. This high energy density enables compact energy storage.

1.5.1.1 Process Description

Figure 1 provides a simplified process flow diagram. The process cools flue gas in several steps, first to approximately cooling water temperature by direct heat exchanger, then to

nearly 0 °C, then to near the frost point of CO_2, which is approximately -100 °C for these flue gases. The figure represents all of these cooling steps as a single cooling heat exchanger. In fact, it occurs in three major steps, all of which minimize pressure drop and some of which utilize a process patented by SES that simultaneously cools and dries the gas. At cold, dry flue gas has a moisture content well below its dew (frost) point at -100 °C (typically in the single digit ppt range). The flue gas then enters a desublimating heat exchanger where it further cools as CO_2 desublimates to form solids. The desublimating heat exchanger operates as a counter-current, direct-contact heat exchanger that contacts the flue gas with a cryogenic liquid. CO_2 forms solid particles in the liquid, which exit the exchanger as a slurry. The capture efficiency of the process depends on the coldest contact liquid temperature. The slurry leaves the desublimating heat exchanger and enters a screw press, which separates the solids and liquids. The contacting liquid passes through a heat exchanger that returns its temperature to the set point (-147 °C in this case) as it recycles back to the desublimating heat exchanger. The screw press or a slurry pump, or both, pressurize the solid stream to about 8 bar. It passes from the screw press directly into a melter where its temperature increases to about -54 °C, at which condition the CO_2 is a liquid. Some of the contact fluid inevitably remains with the particles as the slurry passes through the screw press. This residual contact liquid separates from the CO_2 stream in either a flash drum or a distillation column. The flash drum produces about 99.4% pure CO_2 while the distillation column produces beverage-grade (99.999%) CO_2. This simulation uses a distillation column. The distillation column also separates out residual light gases (typically O_2 and N_2).
Most of the pollutants (NO_x, SOx, Hg, PMxx, etc.) accumulate in the melter or the distillation column reboiler. These are separated as an additional small stream from these locations. A more detailed process flow diagram and explanation is available but can only be shared under a non-disclosure agreement.

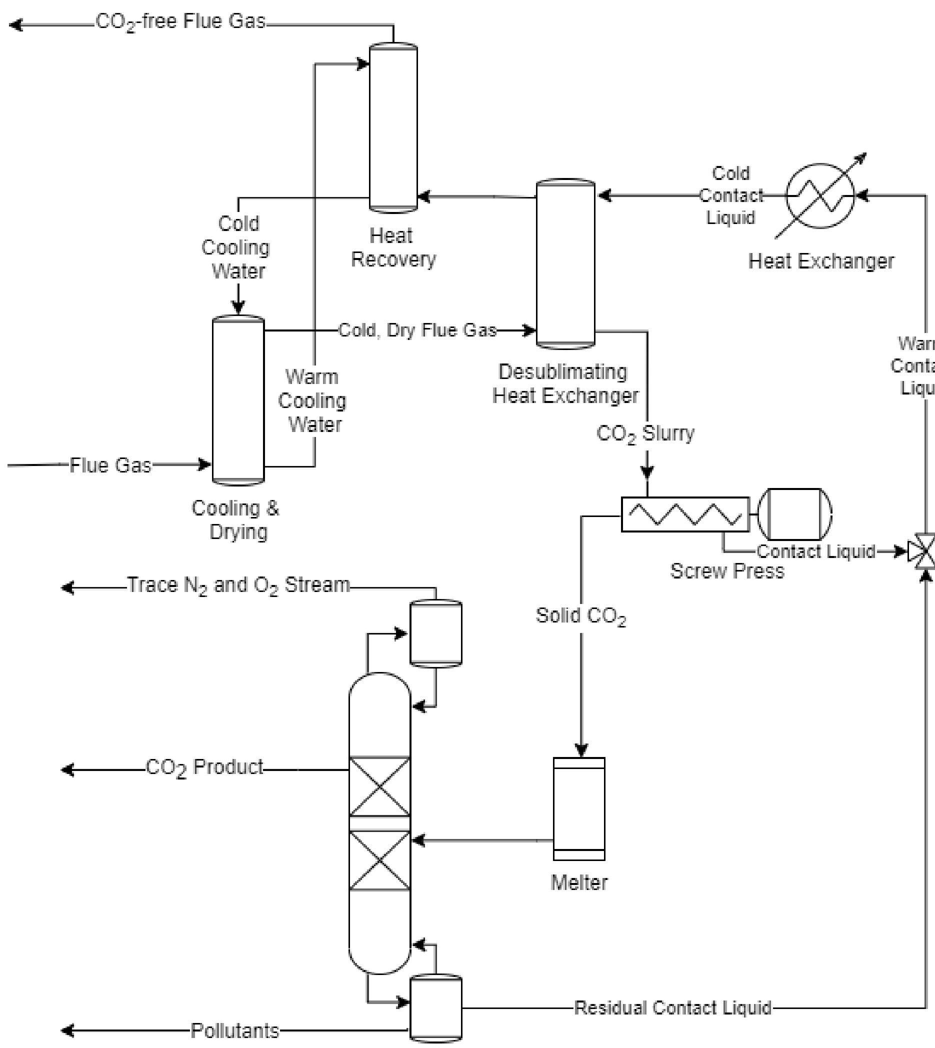


Figure 1 A simplified process flow diagram for the CCC process.

1.5.1.2 Cooling from LPG

Shaft work or electricity represents slightly over 90% of the total energy demand for the CCC process. Nearly all of this could be supplied by vaporizing LNG. There is a version of the CCC process that stores and releases energy (CCC-ES) in large quantities and on short time-scales. It is described in more detail at the end of this report, but it involves using natural gas (NG) as a refrigerant in the LNG. Alternatively, if there exists an LNG import hub, the CCC-ES process can use the cooling from vaporizing the LNG to drive the carbon capture process. Depending on how it is implemented, the CCC-ES process decreases the overall energy demand when using LNG as refrigerant by up to 80-90%.

1.5.1.3 Gray Cement

The following sections summarize the performance of the CCC process in terms of heat and material balances and costs, in that order. The simulations also depend on details such as temperature approaches (typically about 5 K), equipment efficiency (varies), and thermodynamic models (Peng Robison with interaction parameters). Most of these represent conservative estimates and have experimental or operational validation. The overall process cost and energy demand could be lower with more aggressive assumptions.

1.5.1.3.1 Capture Rates, Purities, and Heat and Material Balances

Table *1* summarizes the inlet and outlet conditions, respectively, for the gray cement case. The inlet conditions came from Aalborg Portland and represent a saturated gas flow. A small liquid water component in this stream ensures the stream is saturated. Four outlet streams appear in the table, namely the treated flue gas, two streams that describe the captured CO_2, and one of the pollutant streams. Several additional streams in the process include a substantial water outlet stream and several minor blow-down streams from several pieces of equipment. These additional streams contain most of the water and some of the remaining pollutants. Several additional chemical species primarily recirculate in the process rather than entering or exiting it. These include refrigerants, drying agents and the contact liquids, most of which are also not in the table. The last two columns in *Table 1* highlight the capture amounts for several key species. The columns depend on the amount of the species that remains in the flue gas (fraction captured) and the amount in the capture streams (fraction in outlet), respectively. The capture stream is the CO_2 stream for CO_2 and the pollutant out stream for everything else. The pollutant out stream in the table is the largest, but not the only, exiting stream that contains pollutants.

This simulation ran under conditions that capture about 99.2% of the CO_2. The process can capture arbitrarily high amounts, including up to and exceeding 100% of the CO_2 that did not enter into the system with the combustion air. The conditions in the table represent typical operating conditions for the CCC-ECL process when it has operated in the field at cement plants. Similarly, the CO_2 and flue gas purity depend on the potential CO_2 use. For example, this simulation estimates a CO_2 purity of 99.999+% (roundoff in the table makes the stream 100% CO_2) in the large CO_2 stream that contains 99% of the captured CO_2 and a CO_2 purity of 98% in the smaller stream. The large stream is beverage grade by almost any standard in that it contains minimal O_2, N_2, or other gases. The smaller stream is not beverage grade. Both streams probably exceed the CO_2 purity specifications for most industrial applications or for sequestration. Some of the capture highlights of these simulations include:

- 99+% CO_2 capture
- 99.999+% CO_2 purity
- 96% SO_2 capture (SO_3, if present, is captured at higher rates)
- 99.9+% NO_2 capture (NO, if present, is captured at lower rates)
- 0% CO capture
- 60% HCl capture
- 99.9+% NH_3 capture

The CCC process design focuses on CO_2 capture. In the process of capturing CO_2, it captures several other pollutants. In general, the process captures CO_2 and everything heavier than CO_2 quite efficiently. CO and NO volatility exceed CO_2 volatility and these species are not effectively captured by the process unless they react to form CO_2 and NO_2. HCl volatility slightly exceeds that of CO_2 and CCC partly captures it. The process captures most other pollutants, including pollutants not included in the table (hydrocarbons, Hg, particulate, ozone, etc.).

Table 1 Summary of primary flue gas flows for the gray cement flue gas case.

Description	Units	Flue Gas In	Flue Gas Out	CO2 Out	CO2 Out	Pollutants	Capture Fraction	Fraction in Outlet
Phase		Vapor Phase	Vapor Phase	Liquid Phase				
Temperature	C	70.0000	31.1520	-54.2014	-87.6709	-53.1500		
Pressure	bar	1.0000	1.0025	150.0000	150.0000	6.9102		
Average MW		29.7368	28.0067	43.7213	44.0097	37.3668		
Mole Flows	kmol/sec	8.8885	6.4316	0.0181	1.5111	0.0000		
O2	kmol/sec	0.5410	0.5400	0.0002	0.0000	0.0000	0.001966	0.000295
N2	kmol/sec	5.6401	5.6382	0.0002	0.0000	0.0000	0.000336	3.81E-05
CO2	kmol/sec	1.5458	0.0127	0.0177	1.5110	0.0000	0.991776	0.988983
SO2	kmol/sec	0.0001	0.0000	0.0000	0.0000	0.0000	0.957401	2.71E-17
NO2	kmol/sec	0.0013	0.0000	0.0000	0.0000	0.0000	0.999609	0
CO	kmol/sec	0.0003	0.0003	0.0000	0.0000	0.0000	0	0
HCL	kmol/sec	0.0001	0.0000	0.0000	0.0000	0.0000	0.595209	0.092194
NH3	kmol/sec	0.0003	0.0000	0.0000	0.0000	0.0000	1	0
HG	kmol/sec	0.0000	0.0000	0.0000	0.0000	0.0000	0.999181	0
Mole Fractions								
O2		0.0609	0.0840	0.0085	0.0000	0.0000		
N2		0.6345	0.8766	0.0116	0.0000	0.0000		
CO2		0.1739	0.0020	0.9798	1.0000	0.0000		
H2O		0.1304	0.0374	0.0000	0.0000	0.0000		

1.5.1.3.2 Equipment Costs and Weights and Utilities

The total installed capital cost for the process at this scale comes to about $128M USD. The most significant portion of this cost are for turbomachinery associated with compression-expansion refrigeration, and for process pressure rise. The next most expensive equipment involves heat exchange equipment in the form 5 direct-contact spray towers or packed columns and one multi-stream heat exchanger. Three of the spray towers and the multi-stream heat exchanger require stainless steel construction because of their operating temperatures. The remainder of the equipment includes distillation columns, a slurry pump, a screw-press filter, and several flash drums, pumps, flow splitters and combiners, etc. The total installed costs should be about $128M USD, these costs exceed those for the white cement case discussed later primarily because of the high flow rate and secondarily because of the higher capture fraction and CO_2 purity.

The utilities for this process include only electricity and depend on compressor efficiency and heat exchange approach temperature. These simulations estimate a total CO_2 capture energy (electricity) demand of about 1.02 MJ_e/kg of CO_2 captured corresponding to 283 kWh/ton. This is 5-6% of the energy consumption of the complete CCU process to generate methanol.

1.5.1.4 White Cement

The following sections summarize the performance of the CCC process in terms of heat and material balances and costs, in that order for the flue gas stream originating from white cement production. In this simulation there is no CO2 distillation column hence the CO2 is less pure than for the grey cement. This is only to show the difference between the two options, with and without distillation.

1.5.1.4.1 Capture Rates, Purities, and Heat and Material Balances

Table *2* summarizes the inlet and outlet conditions, respectively, for the white cement case. The inlet conditions came from Aalborg Cement and represent a saturated gas flow. A small liquid water component in this stream ensures the stream is saturated.
The same three outlet streams as for the grey cement case appear in the table, namely the treated flue gas, the captured CO_2, and the one of the primary pollutant streams. The last two columns in Table 3 highlight the capture amounts for several key species.
This simulation ran under conditions that capture about 96% of the CO_2. Similarly, the CO_2 and flue gas purity depend on the potential CO_2 use. As an alternative to the above for the grey cement, this simulation estimates 99.9+% CO_2, which probably exceeds the purity needed for sequestration or almost any industrial application but which is not beverage grade. Some of the capture highlights of these simulations include:

- 96% CO_2 capture
- 99.9+% CO_2 purity
- 95% H_2O capture (closer to 100%, but 5% is added back to flue gas)
- 93% SO_2 capture (SO_3, if present, is captured at higher rates)
- 99.9+% NO_2 capture (NO, if present, is captured at lower rates)
- 0% CO capture
- 52% HCl capture
- 99.9+% NH_3 capture

Table 2 Summary of outlet flows for the white cement flue gas case.

Description	Units	Flue Gas In	Flue Gas Out	CO2 Out	Pollutant Out	Fraction Captured	Fraction in Outlet
Phase		Vapor Phase	Vapor Phase	Liquid Phase	Liquid Phase		
Temperature	C	70.00	30.45	-87.65	-52.20		
Pressure	bar	1.01	1.02	150	0.99		
Average MW		27.30	28.21	44.02	45.63		
Mole Flows	kmol/sec	4.3531	2.2279	0.6742	2.3335		
O2	kmol/sec	0.1981	0.1978	0.0001	0.0002		
N2	kmol/sec	1.9225	1.9222	0.0001	0.0002		
CO2	kmol/sec	0.7074	0.0275	0.6738	0.0060	0.96	0.96
H2O	kmol/sec	1.5236	0.0800	0.0000	0.0357	0.95	0.02
SO2	kmol/sec	0.0004	0.0000	0.0000	0.0004	0.93	0.93
SO3	kmol/sec	0.0000	0.0000	0.0000	0.0000		
NO	kmol/sec	0.0000	0.0000	0.0000	0.0000		
NO2	kmol/sec	0.0007	0.0000	0.0000	0.0006	1.00	0.84
CO	kmol/sec	0.0002	0.0002	0.0000	0.0000	0.00	0.00
HCL	kmol/sec	0.0000	0.0000	0.0000	0.0000	0.52	0.15
NH3	kmol/sec	0.0002	0.0000	0.0000	0.0000	1.00	0.00
Mole Fractions							
O2	frac	0.0455	0.0888	0.0001	0.0001		
N2	frac	0.4416	0.8628	0.0001	0.0001		
CO2	frac	0.1625	0.0124	0.9994	0.0026		
H2O	frac	0.3500	0.0359	0.0000	0.0153		

1.5.1.4.2 Equipment Costs and Weights and Utilities

The total installed capital cost for the process at this scale comes to about $68M USD. Again, the most significant portions of this cost are for turbomachinery associated with compression-expansion refrigeration, and for process pressure rise.

The utilities for this process include only electricity and depend on compressor efficiency and heat exchange approach temperature. These simulations estimate a total CO_2 capture energy (electricity) demand slightly less than 1.0 MJ_e/kg of CO_2 captured.

1.5.1.5 Using LNG as a Refrigerant – CCC-ES

This project demonstrates energy storage by means of generating extra refrigerant for the CCC process during times of excess/inexpensive energy availability and restoring that energy to the grid by using the stored refrigerant to decrease the CCC parasitic loss during high demand/cost time periods. This strategy allows the process to store substantial energy when energy costs are low because of excess renewable generation or any other reason and restore that energy when costs are high, typically because of high energy demand compared to supply. For a thermal power plant, the incremental capital expense increases for this process compared with operating a non-energy storing CCC process primarily because of the cryogenic storage tank. For a cement plant that operates continuously independently of electricity production from renewables there is an additional capital cost for larger LNG compressors making the energy storage less economically attractive compared with installations on thermal power plants.

The process efficiency is very high since the refrigerant must be generated at some point and this process just shifts its generation to a time when power demand is low. The primary efficiency decrease comes from loss of refrigerant during storage, which is well under 1%/day. While the focus of this project was on energy storage, this process also stores NG itself, and parallel work indicates the economic value of the natural gas storage exceeds that of the energy storage.

1.5.1.6 Flow Diagrams for a Thermal Power Plant equipped with CCC-ES:

The process flow diagrams (Figure 3-Figure 5) illustrate the CCC-ES process when it is neither storing nor recovering energy, followed by the conditions during energy storage and energy recovery. These diagrams differ primarily in the state of the storage tank, the state of the refrigerant NG source to the plant, and the state of the refrigerant NG output from the plant, as summarized in Table 5. In this table and in the figures, the NG refers only to NG used as a refrigerant. This refrigerant NG differs from NG consumed in combined-cycle and similar power plants as fuel.

Table 3 Summary of main process changes based on energy transfer mode.

Energy Storage State	Normal	Storing	Releasing
Storage Tank	no net flow	filling	draining
Refrigerant NG to plant	no net flow	net inflow	no net flow
Refrigerant NG from plant	no net flow	no net flow	net flow to pipeline or auxiliary turbine
Net Power Production	Normal	Low	High

Figure 3 illustrates the CCC-ES process during normal (neither storing nor recovering energy) operation. The following description steps through most of the process during normal operation. The ambient-pressure flue gas (orange line) exits the cement kiln and passes through a series of innovative coolers and dryers, eventually cooling to just above the CO_2 frost point, which is typically about −100 °C. This gas enters the desublimating heat exchanger, which removes CO_2 by further cooling the gas to −120 °C for 90% capture or −132 °C for 99% capture. The exact temperatures depend on the initial gas composition. The cold and CO_2-depleted flue gas leaves the desublimator (purple line) and warms back to nominally ambient temperature by cooling the incoming stream.

The CO_2 separates from the flue gas in the desublimating heat exchanger (black lines) as a slurry. The slurry passes through a slurry pump to increase the pressure to 8-10 bar and then cools to slightly below the coldest temperature the flue gas attains (typically −136 °C). The slurry passes through a solid-liquid separation vessel. The liquid phase of the slurry passes from this separation and recycles back to the desublimating heat exchanger.
The solid CO_2 leaves the solid-liquid separator (red line) and warms to its melting point (about −56 °C). This liquid CO_2 warms further in the multi-stream heat exchanger to somewhat below ambient temperature and then passes into a CO_2 polisher. The CO_2 polisher technology depends on the CO_2 purity specification and consists of a two-stage flash drum for 99.4% CO_2 or a distillation column for 99.999% pure CO_2. The CO_2 stream then passes through a pump that raises the pressure to typically 125-150 bar and finally exits the process as a high-pressure, ambient temperature liquid.

The process uses up to two primary refrigeration loops, one of which (green lines) serves as a first stage of the lowest-temperature refrigeration loop (blue lines). The CCC-ES process uses NG for the lowest temperature loop. This NG passes through a compressor and then cools and condenses in the main multi-stream heat exchanger. The liquid pressure then decreases and subsequently cools the slurry from the desublimating heat exchanger as it vaporizes. The NG vapor then warms back to ambient pressure in the multistream heat exchanger and recycles back to the compressor. The dashed blue lines to and from the NG pipeline and LNG storage tank transport nominally nothing in this operation mode.
The power plant produces electricity, including mostly net production for the grid but some electricity needed for the CCC process, primarily to operate the turbines. In the case of the cement plant electricity is consumed from the grid. The level of LNG in the tank does not change when operating in "normal" mode.

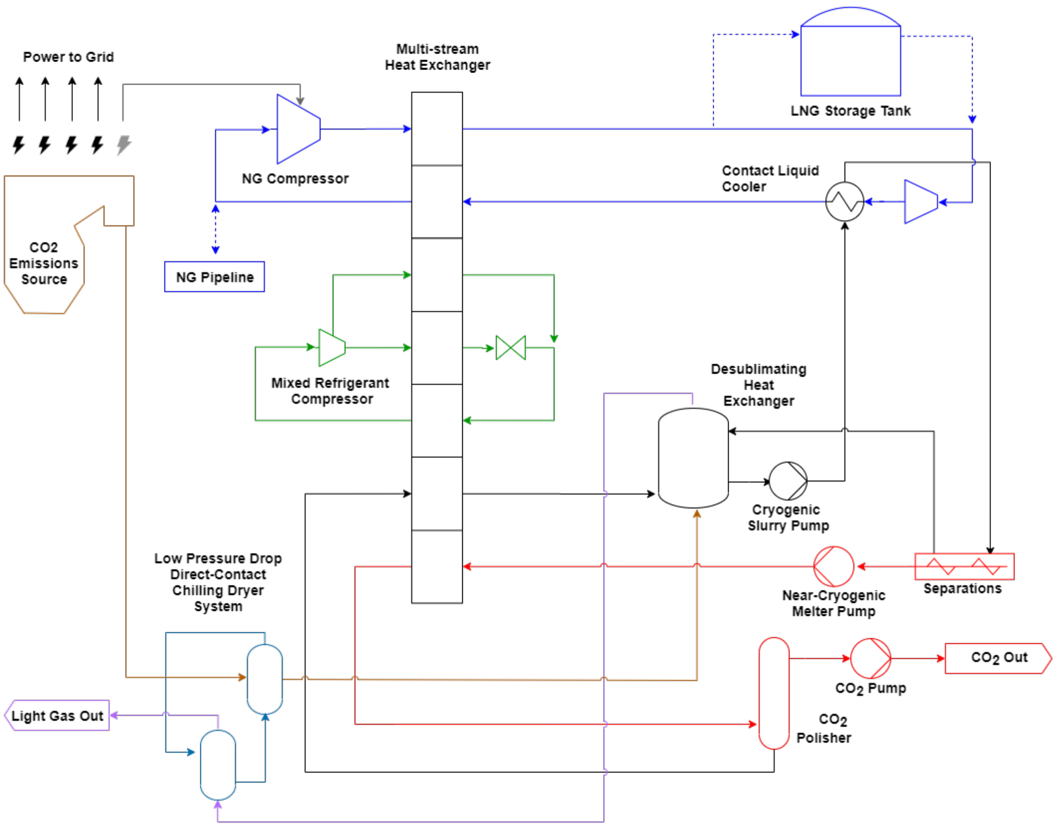


Figure 2. Process flow diagram for normal conditions (neither storing nor releasing energy).

The process changes during energy storage and recovery apply almost exclusively to the blue NG lines and the power to grid lines, and the subsequent discussions involves only these. The remainder of the process does not change.

The process stores energy from the grid when power generation costs are low, such as when there are excess renewable energy supplies or during low-demand periods (Figure 4). During energy storage, less power goes to the grid and more power generates refrigerant. The excess refrigerant, in the form of LNG, fills an LNG storage tank while the amount needed to operate the CCC process continues through the process as normal. The NG needed for this storage comes from a NG pipeline. The rate of energy storage can vary substantially but it averages to 5-10% of the power plant capacity. The flow rate of NG into the multi-stream heat exchanger no longer balances with that coming back from the CCC process, causing a potential imbalance in the heat exchanger that can be corrected with SES's dynamic heat exchanger technology. These dynamic heat exchangers maintain constant temperature profiles in the heat exchangers even as the flows temporarily imbalance. The LNG refrigeration process operates independently of the CCC process in that the changes in the LNG process that enable energy storage do not affect the CCC portion of the process.

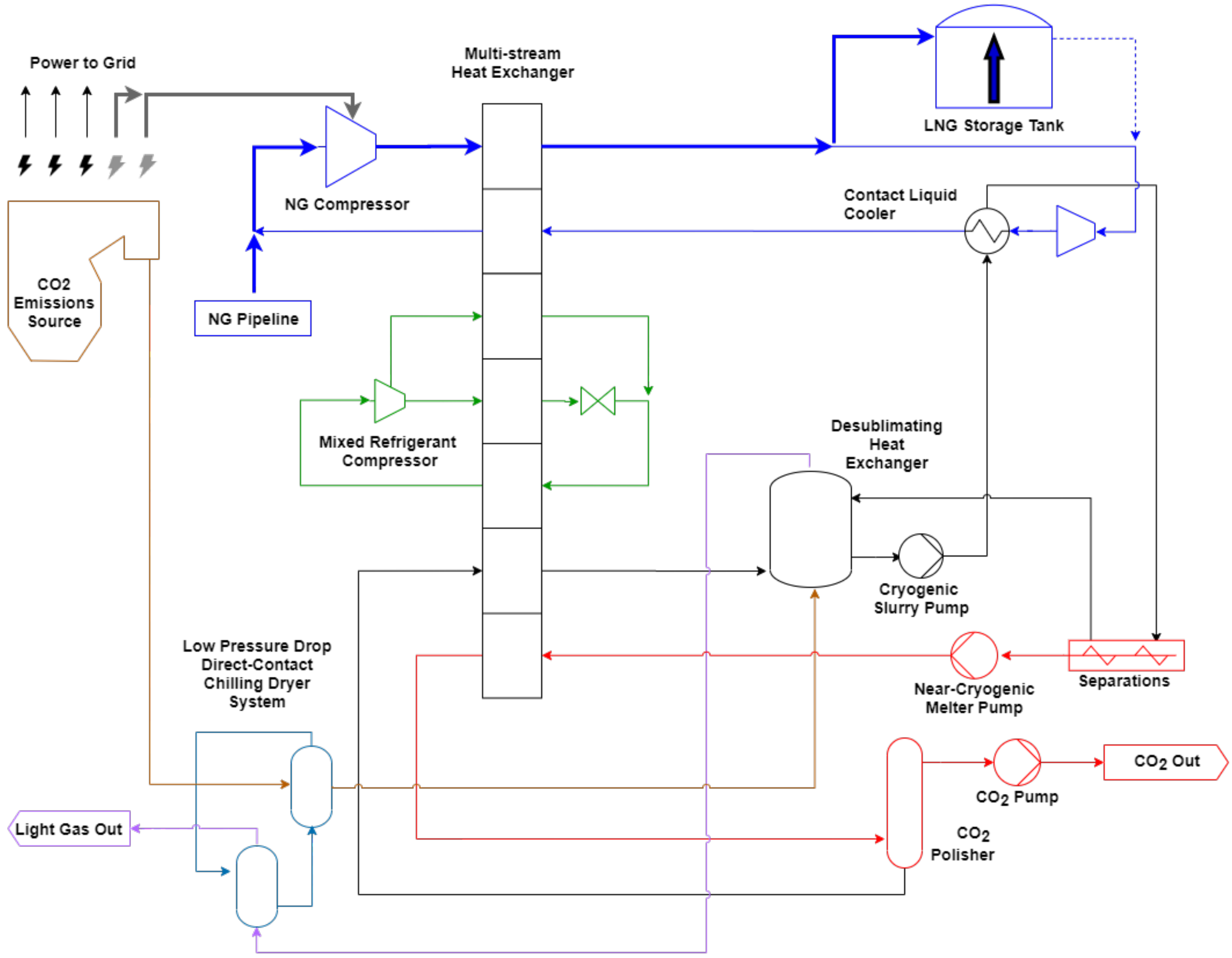


Figure 3. Process flow diagram for energy storing conditions.

Figure 5 illustrates how the CCC-ES process releases energy back to the grid by using the stored LNG to decrease the parasitic load associated with refrigerant generation, redirecting the saved energy back to the grid. The dashed lines indicate that a small amount of NG still passes through the compressor. This small amount keeps the turbomachinery rotating, which greatly improves its dynamic response. However, the great majority of the LNG comes from the storage tank, provides refrigeration for the CCC process, and then passes back into the NG pipeline as a pressurized but ambient-temperature vapor.

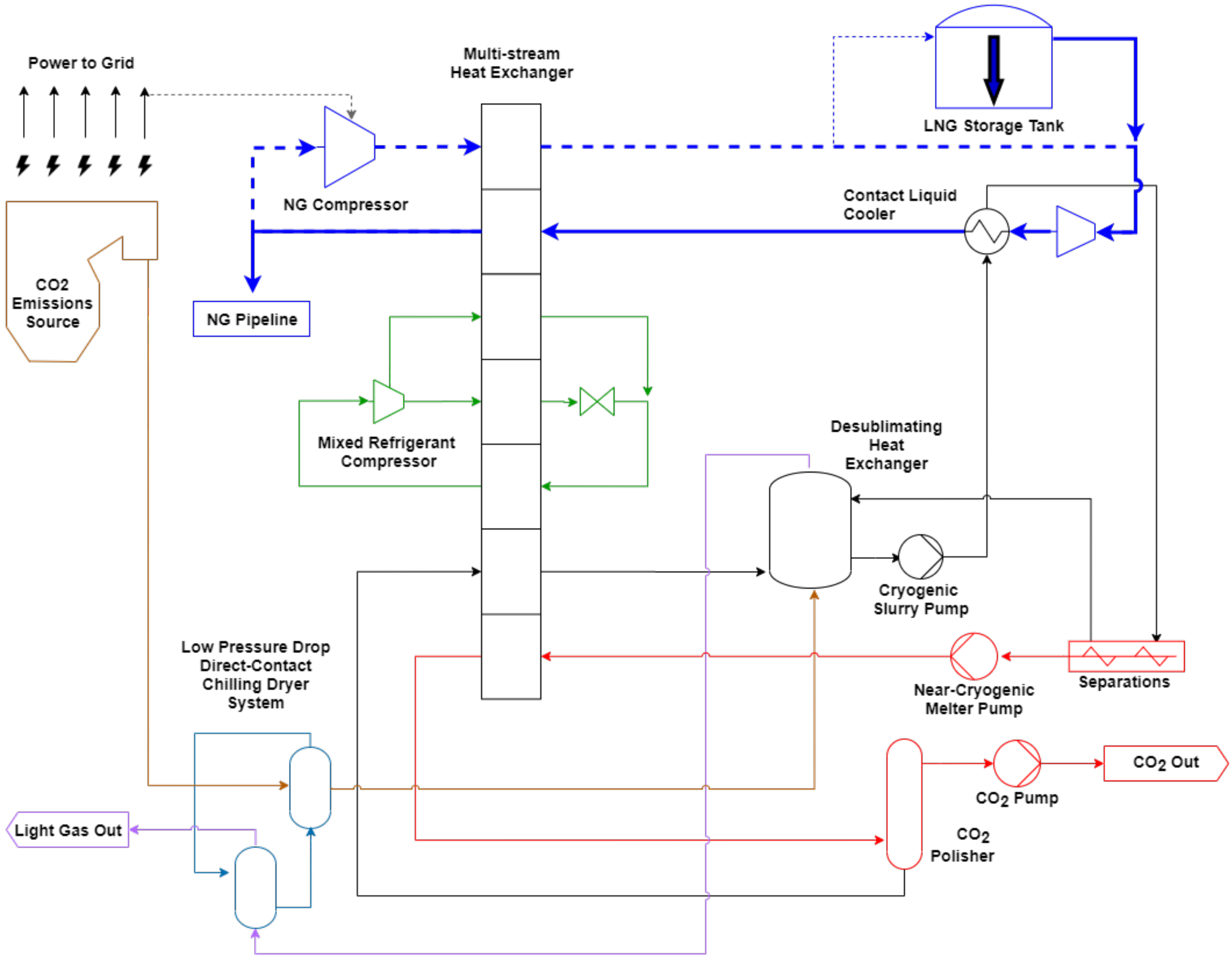


Figure 4. Process flow diagram for energy releasing conditions.

1.5.2 Transportation fuel from CO2

The third work package was a theoretical and experimental feasibility study of using the captured CO_2 as a carbon source in transportation fuel synthesis. Overall, the time taken to get the laboratory scale fuel synthesis plant operational was longer than expected which influenced the range of processes that could be simulated. It was decided to focused on methanol synthesis since it is relatively simple and could be completed within the available project time frame.

1.5.2.1 Laboratory scale fuel synthesis

The lab-scale experimental test was conducted in the established Mini-Plant testing platform for the process of methanol production from CO_2 and H_2. The following objectives were defined:

- To investigate the catalyst performance (activity, selectivity and stability) for methanol production from CO_2 and H_2 under different operating conditions (pressure, temperature and gas hourly space velocity (GHSV))
- To investigate the impurities in the crude methanol product.
- To investigate possible operation issues for the recycle-mode.

The testing results can form the basis for validating the performance predicted from ASPEN simulations that can later be used to extrapolate findings to a complete methanol plant.

1.5.2.2 Experimental setup

The Mini-plant test platform used for the catalyst testing is shown in Figure 6.

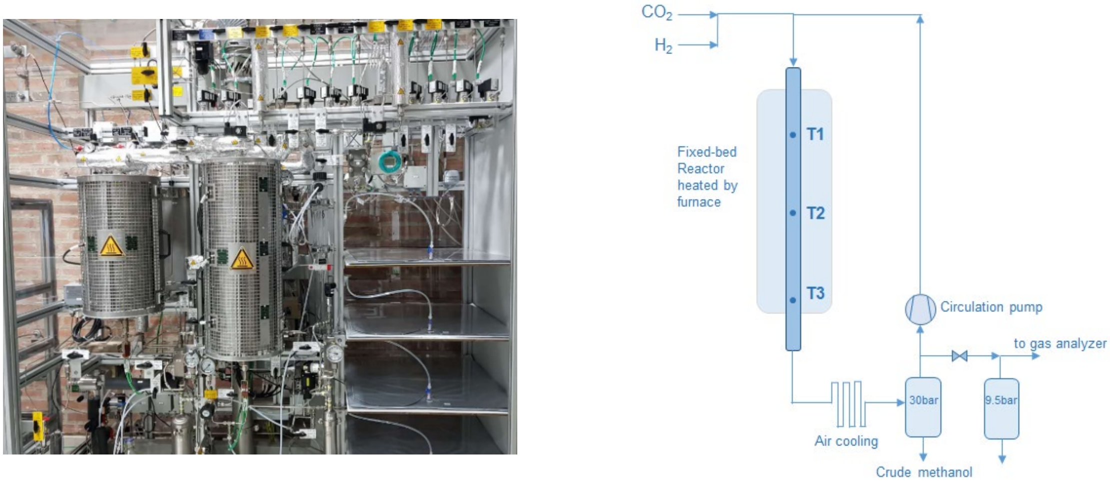


Figure 5. Photo of Mini-Plant (left) and Process scheme (right) of the Mini-Plant.

The reactor parameters and operating conditions are shown in Table 4.

Table 4 Testing parameters

Catalyst	
Commercial Catalyst	Topsøe MK-121
Pellet size, mm	0.425-0.85
Reactor	
Tube diameter, mm	9
Height of catalyst bed	231
Operating Conditions	
Absolute Pressure, bar	30
Inlet temperature, ℃	220
Furnace temperature, ℃	c.a. 250
GHSV, h^{-1}	4100, 15000−50000

1.5.2.3 Single-pass mode (Thermodynamic equilibrium conditions)

Due to the low GHSV value (4100 h^{-1}) under the one-pass mode in the testing, the thermodynamic equilibrium was supposed to be achieved in the methanol reactor. The testing results were compared with the simulation results by ASPEN PLUS (shown in Figure 7). The overall trend is well captured but there is some variation in absolute values with the experimental conversion exceeding the predicted. The reason for the difference is most likely a combination of uncertainty in the kinetic data and the measurement uncertainties related to the gas analysis.

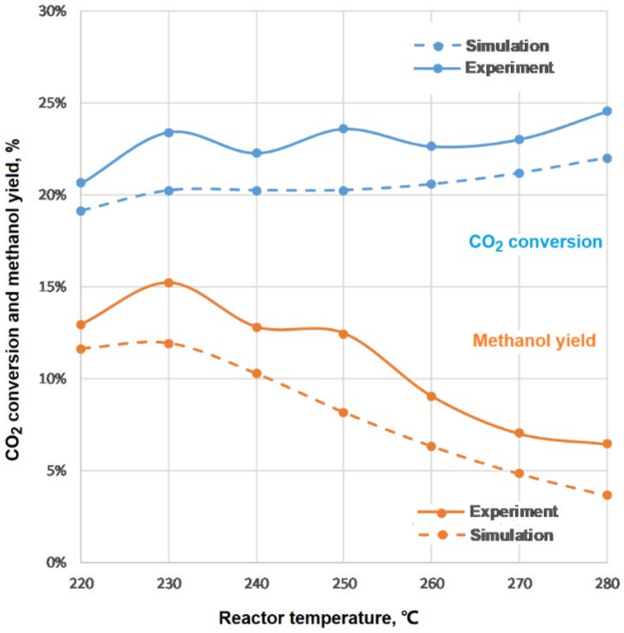


Figure 6 CO2 conversion and methanol yield at different reactor temperatures.

1.5.2.4 Recycle mode

The methanol synthesis process under recycle mode was conducted which is close to the condition of a conventional methanol plant. The GHSV value of 15000 h^{-1} was used for the **baseline case**. The reactor temperatures were monitored by the thermocouples (T1, T2 and T3) along the reactor, where T1 (220℃) was close to the inlet of the catalyst bed, T2 is in the middle and T3 is near the outlet. With the increase of the GHSV value the temperatures (T2 and T3) in the catalyst bed moved, which is shown in Figure 3. The temperature T2 decreased with GHSV value, the trend is different from the simulation result, which could be attributed to that the height of the catalyst bed in the reactor tube is different from the calculated value (using a bulk density of 1.3 g/ml), for example, the temperature trend 40mm away from T2 (named position adjusted, shown in Figure 8) agrees well with the experimental result. The temperature of T3 achieved maximum value at around **GHSV=30000 h^{-1}**, which reveals that T3 is close to the hot spot and all the catalyst in the reactor is involved in the methanol synthesis reactions under this GHSV value. Additionally, the operation interface for the baseline test was shown in Figure 9.

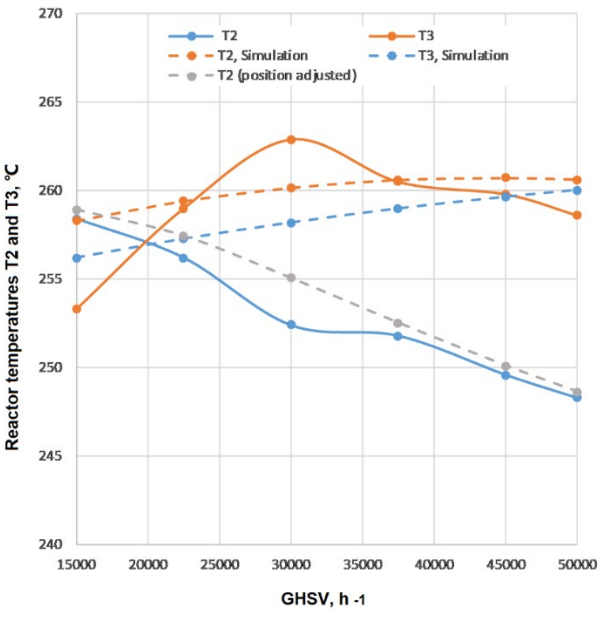


Figure 7 Reactor temperatures at different GHSV values.

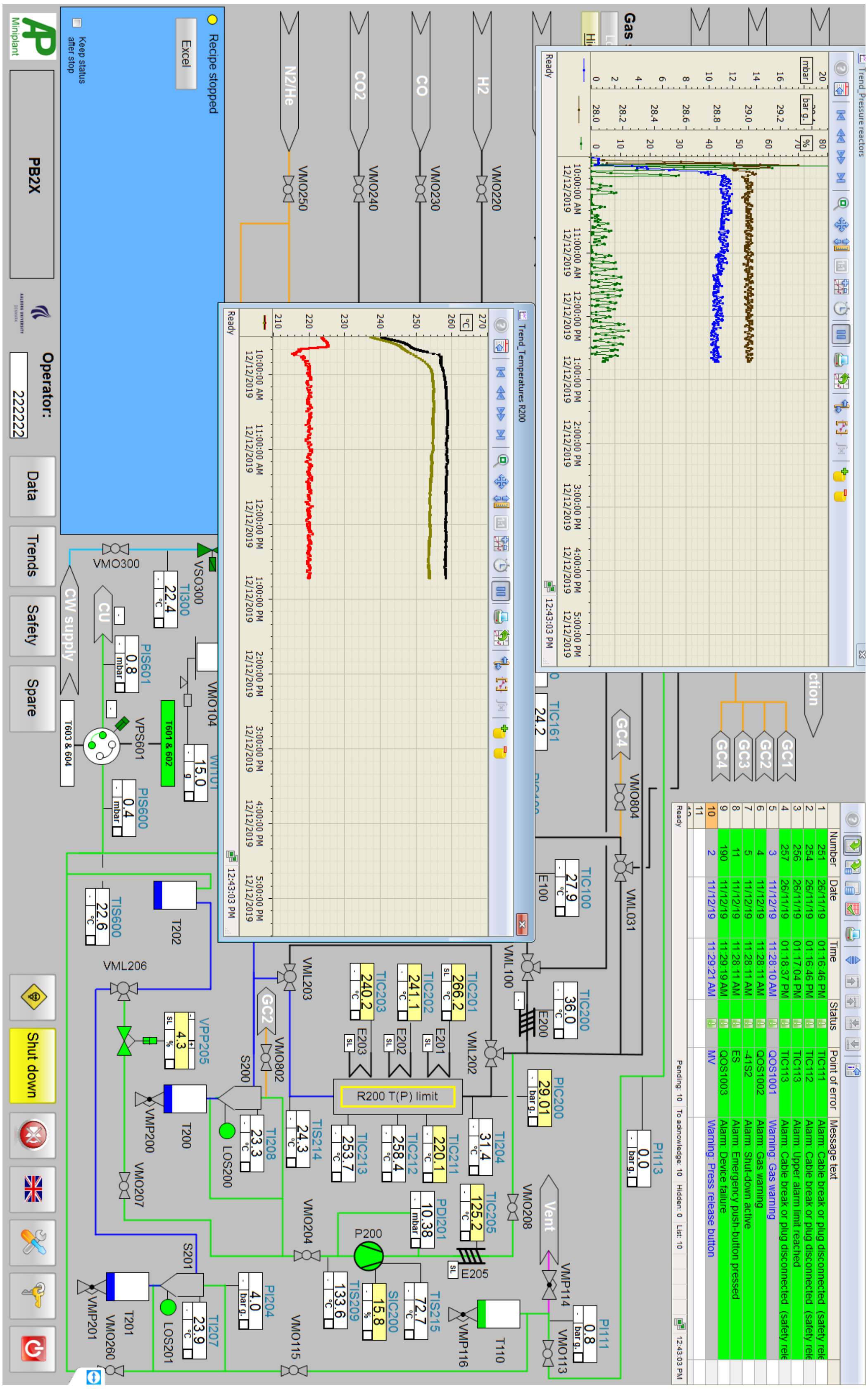


Figure 8 Operation interface for the baseline case (P=30bar, T1=220℃, GHSV=15000 h-1)

1.5.2.5 Influence of inert gas (N_2)

Inert gas may exist in the feed gas such as nitrogen and methane, which can accumulate in the recycle process of methanol synthesis and influence the reactor performance. Therefore, the inert gas should be considered for the reactor and process design. The feed gases with N_2 content of 1−5% were considered for the experimental and simulation study. The gas compositions of the recycle stream under different N_2 contents were obtained and shown in Figure 10, which are close to simulation results (except the trend of the CO content). The results showed obvious N_2 existing (0−35% in this study) in the recycle stream, which results in a higher volume flow rate of the recycle stream and compressor power (shown in Figure 11(a)). Additionally, the change of methanol production rate under different N_2 contents is not obvious with around 8 g/h (shown in Figure 11(b)). Notably, a recycle ratio of 99% (1% of the product gas goes to the exhaust) was assumed in the present study, this ratio could be higher for a methanol plant, e.g., 99.9%, which may result in a higher N2 level in the recycle stream.

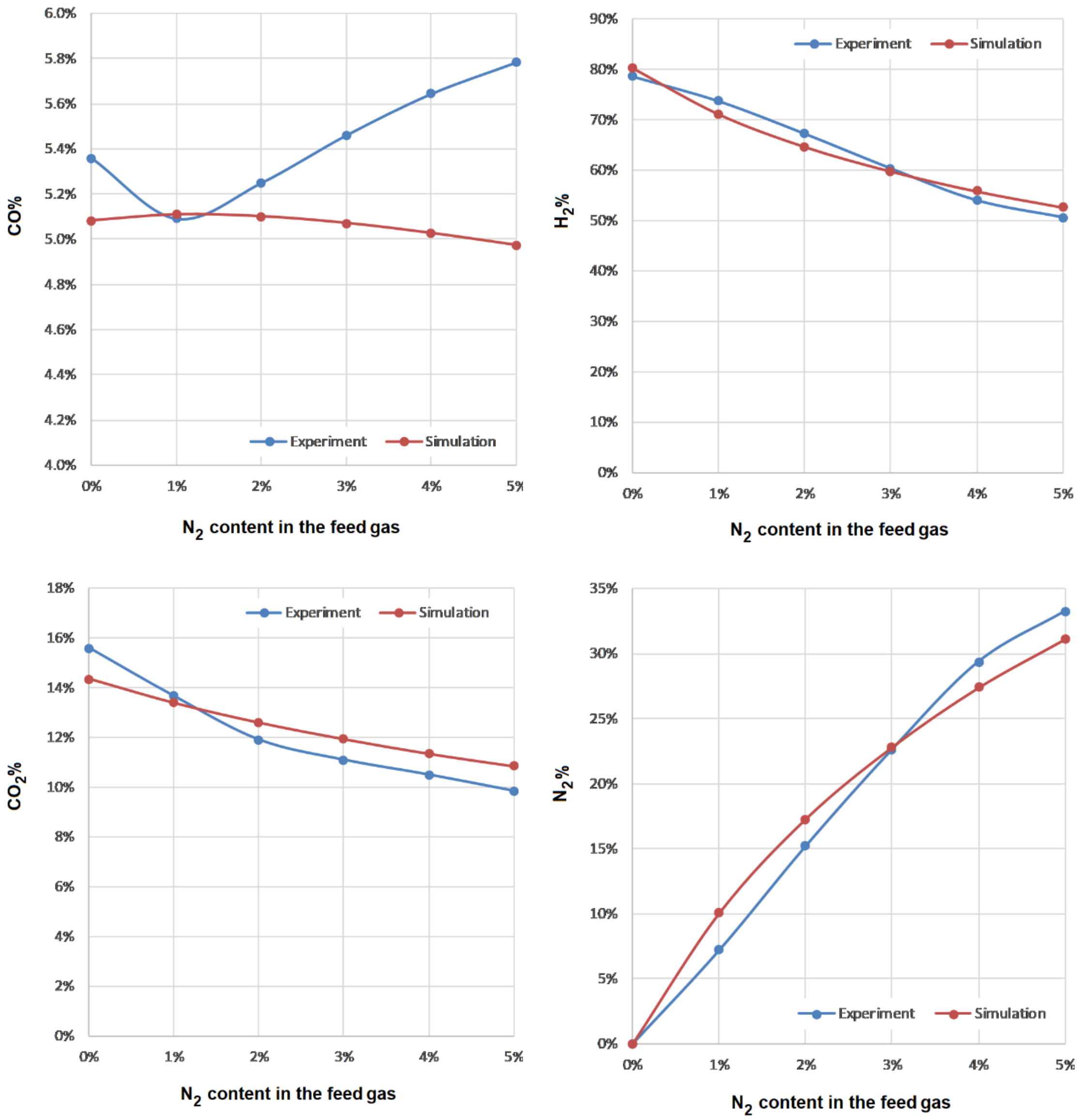


Figure 9 Gas composition in of the recycle stream with different N2 content in the feed.

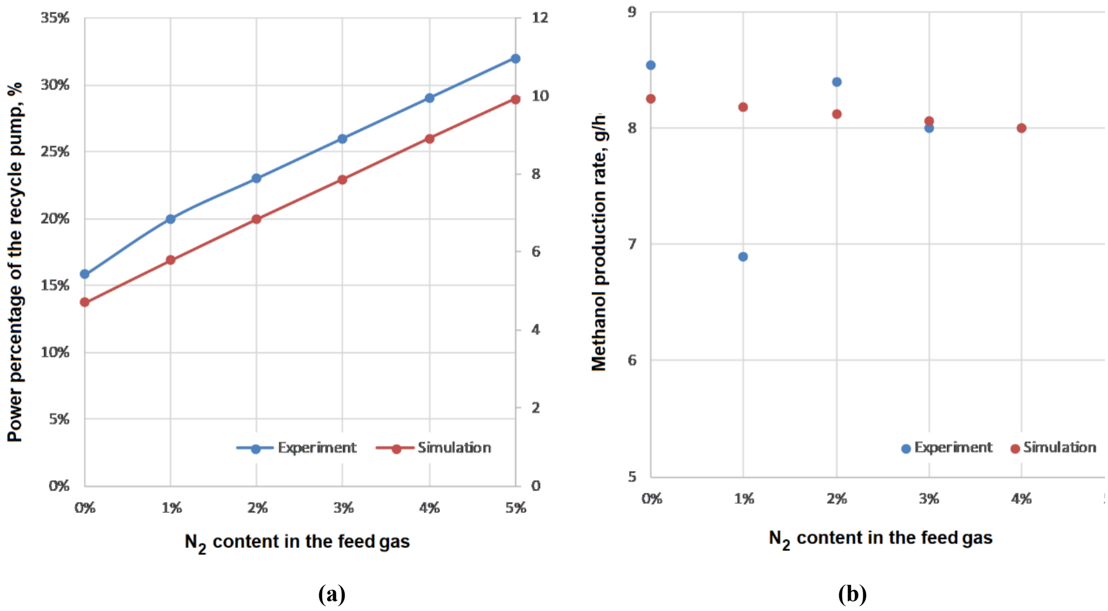


Figure 10 (a) Volume flow rate and compressor power of the recycle stream and (b) methanol production rate with different N2 content in the feed gas.

1.5.2.6 Influence of pressure (50 bar)

The condition of a higher operating pressure (50 bar) has also been investigated. The experimental results show a higher T2 temperature and lower power of the recycle pump, which indicates a higher hot-spot temperature and lower recycle ratio compared with the results at 30 bars. The comparison between the two operating pressures is shown in Table 5.

Table 5 Comparison of the testing results under 30 bar and 50 bars.

Pressure, bar	T1(inlet) , ℃	T2, ℃	T3, ℃	Power output percentage of the recycle pump, %
30	220	258.4	253.3	15.8%
50	220	263.9	253.3	11%

The composition of the non-condensable gas, the recycle stream, has been analyzed, which are close to the simulation results by Aspen Plus (shown in Table 6). Higher content of H_2 and lower content of CO and CO_2 were found compared with those under 30 bars.

Table 6 Noncondensable gas composition in the recycle stream.

	CO mol%	CO_2 mol%	H_2 mol%
Experimental result	2.8%	13.9%	83.3%
Simulation result	3.4%	14.1%	82.5%

1.5.2.7 Summary

- The test demonstrated the feasibility of converting CO2 and H2 directly into methanol using a standard methanol synthesis.
- The experimental study of lab-scale methanol synthesis has been conducted in the Mini-plant. Both one-pass and recycle mode were investigated. The results agree well with the simulations by Aspen Plus.
- The catalyst in the reactor was fully used under the testing conditions of around GHSV=30000 h^{-1}, P=30bar, T1=220℃.
- The accumulation of the inert N_2 gas was found with N_2 content up to 35mol% in the recycle stream in this study, which increases the power requirement of the recycle pump. The influence of N_2 accumulation on the methanol production rate is not obvious.
- The experimental result of the high pressure (50bar) condition shows a slightly higher T2 temperature and lower power consumption of the recycle pump.

1.5.3 Balancing potential and scenarios.

The background of this chapter is, that in a future with increasing amounts of fluctuating RES from wind and sun, the price structure of the energy sector may change significantly. Electricity prices may vary a lot more over time than we see today, and balancing elements might be an extremely valuable part of the grid system. Other elements that may have great influence on tomorrows energy system are the implementation of RED II, and the electricity grid tariff structure. New definition of renewable energy sources and legislative demands for e.g. use of certain types of fuel may change the economic condition completely. This chapter presents different scenarios set up to determine the effects of different external influences on the system. This information will be used to analyze the robustness of the system and determine in which ranges it will be economically feasible. The elements to be changed are:

- The tariffs paid to the TSO and DSO
- The selling price of methanol
- The spot price on electricity
- The price profile of electricity
- The fuel factory capacity
- The liquefier capacity

All of these are changed from the baseline scenario which will be presented in the following section.

1.5.3.1 Baseline scenario:

The baseline scenario is used as a starting point for the analysis. The values used in this scenario are set to resemble the values of a C3U plant used in 2018, the specific values of the system can be found in the Appendix. The price of methanol is fixed at 600 €/ton, which is higher than the present market price for fossil methanol. This is done under the assumption that the ability to produce and sell sustainably produced methanol will be favored by a future price premium.

A graphical representation of the model that was implemented in energyPRO by EMD is shown in Figure 12. This model includes details of the Aalborg district heating system with consumption and production data as well as historical electricity spot market prices.

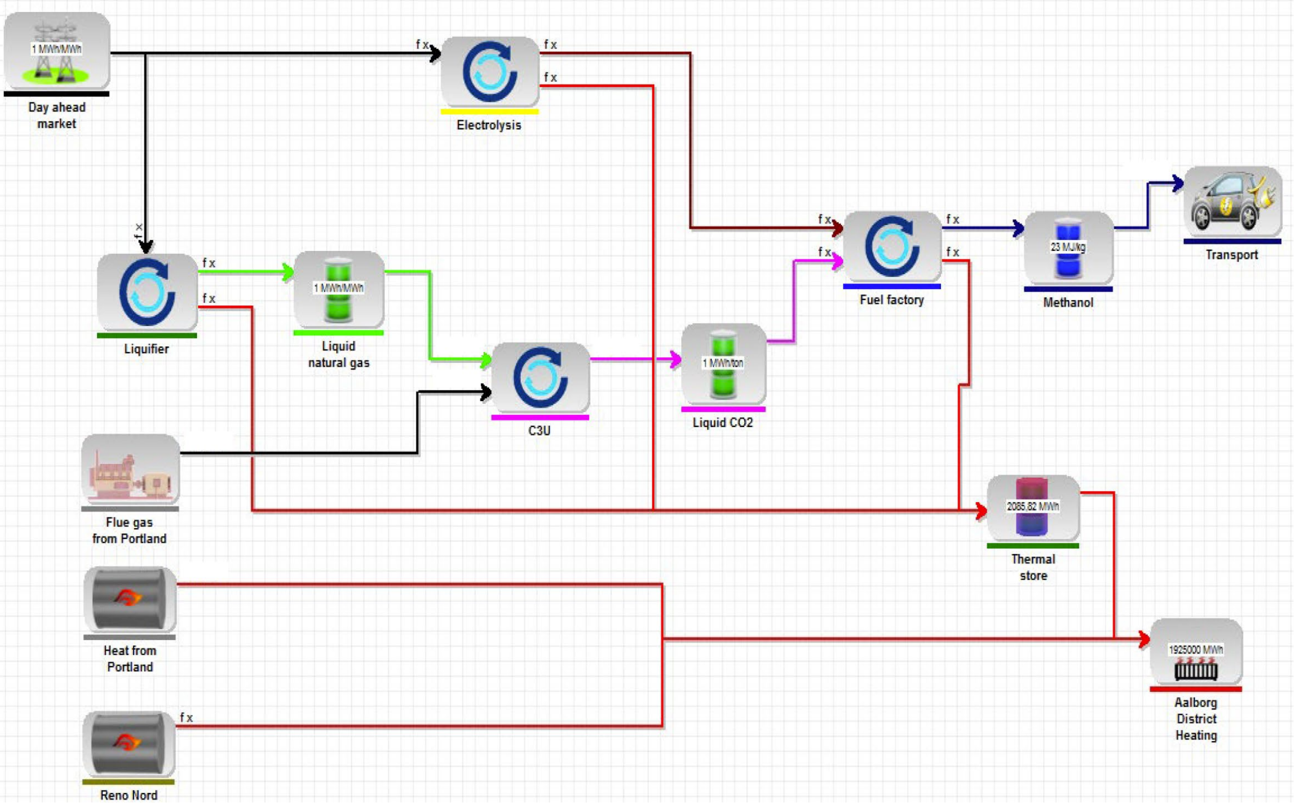


Figure 11. Graphical representation of the C3U system, made in energyPRO.

From Figure 12 it is seen that the methanol production price is dependent on the electricity price, the availability of CO2 and the ability to sell or repel the excess heat. The varying price of electricity in the 2018 price profile has a high influence on the methanol production price, thus energyPRO optimizes the C3U operation and methanol production to achieve the most economical operation. The LNG storage (i.e. the CCC-ES process) and the liquid CO2 storage gives the system an ability to produce at low electricity prices and shut down at high prices, creating a limited flexibility. The operation income, net present value, methanol production and captured CO2 will be used as the basis of comparison. These results of the baseline reference scenario are presented in the table below:

Scenario:	Operation profit [M EUR/year]	Net present value [M EUR]	Methanol production [ton/year]	CO2 captured [ton/year]
Baseline Reference	90,5	464	585.147	798.950

The operation profit includes the income from selling methanol and heat subtracted the expenses from producing methanol. The specifics of these are presented in the table below.

Expenses:	Electricity purchase	Grid tariffs	Maintenance cost	Water purchase
	214 M€	64 M€	13 M€	1,7 M€
Income:	**Methanol and heat**			
	384 M€			

It is observed from this table that the electricity related expenses (electricity purchase and grid tariffs) is around 95 % of the total operation expenses why it is assessed as being the most interesting expense to change in this sensitivity analyses. Using these results as a reference the subsequent sections will analyze and discuss the various changes previously presented.

1.5.3.2 Methanol price

The methanol price has a high influence on the economic feasibility of the system since it creates the main income in all scenarios. The price of 600 €/ton methanol is high compared to the average 2018 price at 335€/ton[1][2], but was evaluated to be fitting for a scenario where electro fuels are used to replace regular fossil-based fuels in competition with other green fuels. In addition to this, the future price of methanol is extremely hard to predict, and the analysis would therefore benefit from testing at a high and a low price. Testing the effects of this price gives a better overview of the resilience for the C3U plant giving an insight as to what the acceptable price range is, for the business case of the C3U plant to be feasible. The results of the scenario run with a selling price of 400 €/ton methanol is shown in the table below.

Scenario	Operation profit [M EUR/year]	Net present value [M EUR]	Methanol production [ton/year]	CO2 captured [ton/year]
Low price (400 EUR/ton)	17,2	-450	185.918	253.850

The results from this scenario shows that when the selling price of methanol is decreased the yearly operation hours and income of the system is also lowered. This also influences the net present value to be negative since the total income over 20 years, does not exceed the expenses and the investment price of the C3U facility. These results generally show that the system is very sensitive towards changes in the methanol price: by changing it to two thirds of the original price the operation income is lowered to a fifth the original income. When assessing the system, it would be highly recommended to consider this effect.

[1] https://www.methanol.org/methanol-price-supply-demand/

[2] Average taken of Rotterdam FOB spot prices over 2018 with an exchange rate of 0,833 € to USD

1.5.3.3 Electricity price and profile

The electricity spot price is the largest part of the operation expenditure. In the reference scenario it accounts for 73% of the total operation expenditure. Given that the production hours are highly dependent on the ability to create a certain profit for the system, changing the electricity prices and the profile of the electricity prices is therefore deemed to have a high influence on production hours as well the operation expenses. The reference scenarios use the price profile of 2018 which is higher than the last 15-year average. The C3U facility is therefore tested with 2015 prices, which had the lowest average price in this 15-year period. The average prices of the 15-year period with the total average price is shown in Figure 13.

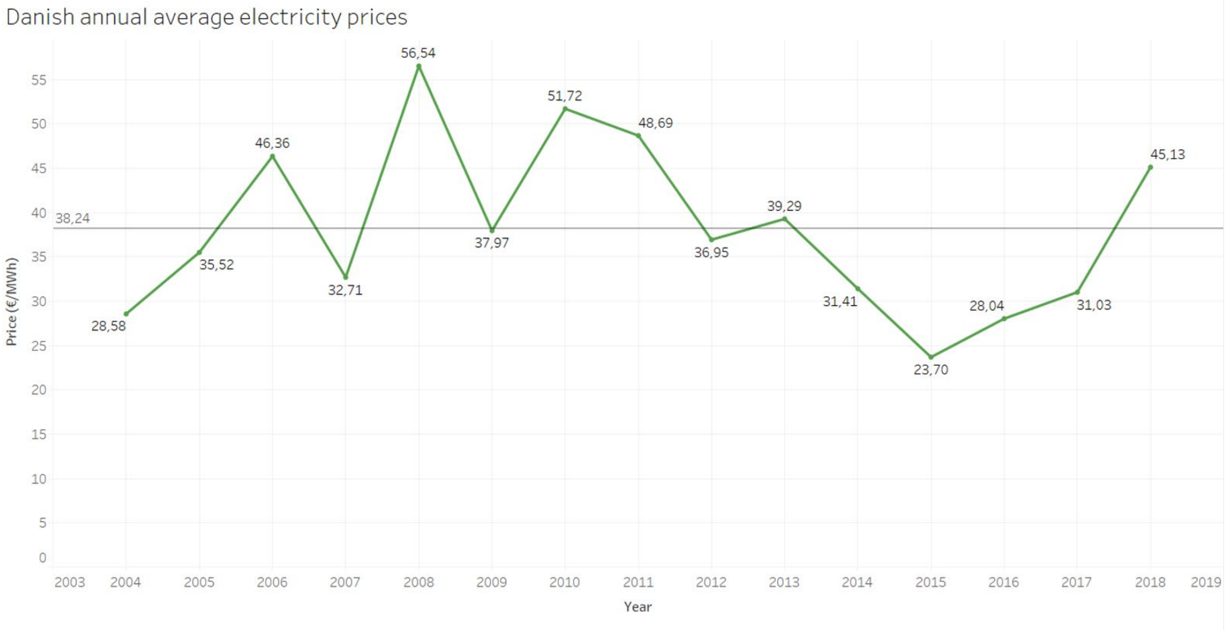


Figure 12 - Average yearly prices from 2004 to 2018, total average shown with a line. Data from Nordpool.com

The electricity price is expected to be lowered with a higher amount of RES in the system and future cost reduction of wind and solar energy, adding to the argument of testing the system with lower prices i.e. the spot prices from 2015. A higher amount of RES in the system is assumed to create a more fluctuating electricity production. Energinet.dk suggested that a factory acting as a balancing unit to counteract this high fluctuating for the electricity system could receive a price reduction on the system tariffs. We use a reduction from 80 DKK pr. MWh to 30 DKK pr. MWh[3] which is considered realistic and required. In order to test the effects of having a higher amount of RES in the system, the 2018 spot prices are tested with an amplitude of 1.5 times the original price profile, i.e. larger price spread. Imagining the fuel factory as a balancing unit, the system is also tested with a reduction in the price of the tariffs. The results of these three scenarios are presented in the following section:

1.5.3.4 Results of electricity price and profile

The results of altering the spot price, increasing the amplitude of the price profile, and the results of reducing the system tariffs from 80 DKK to 30 DKK pr. MWh is shown in the table below.

Scenarios	Operation profit [M EUR/year]	Net present value [M EUR]	Methanol production [ton/year]	CO2 captured [ton/year]
Spot prices from 2015	196,8	1.789	648.965	886.087
1.5X amplitude on 2018	116,9	793	541.549	739.423
Lowered tariffs	131,8	979	625.814	854.477

[3] This tariff reduction is not set but is considered realistic and necessary for PtX plants to be established.

1.5.3.5 2015 spot prices

The results from using the 2015 spot prices shows an increase in both operational income, methanol production and NPV. This indicates that the operational expenses are lower in this scenario, increasing the amount of production hours available for the factory. Where the baseline reference scenario was able to operate for 55 % of the year, the changed spot prices enables the system to run 62% of the year, meaning an increase of 7% of the total run time.

The relatively large increase in operational income, and with that NPV, can be explained by the general lowered operation cost. Where the baseline used 366 €/ton in electricity expenses, the 2015 spot price scenario only uses 217 €/ton. All in all, this scenario shows that with a decrease in operational cost, specifically electricity cost, the number of operation hours is increased and the cost of production per ton is lowered.

1.5.3.6 Profile amplitude changed

The changed amplitude of the price profile of 2018 shows an increased yearly operational income, but a decreased CO2 collection and methanol production. This is explained by the new profile creating less hours of operation, while also creating a lower average price of the hours where it is able to create a profit from operating. This effect is shown in in Figure 14 and Figure 15, which shows the distribution of price hours for the baseline scenario and the scenario using an amplified spot price profile.

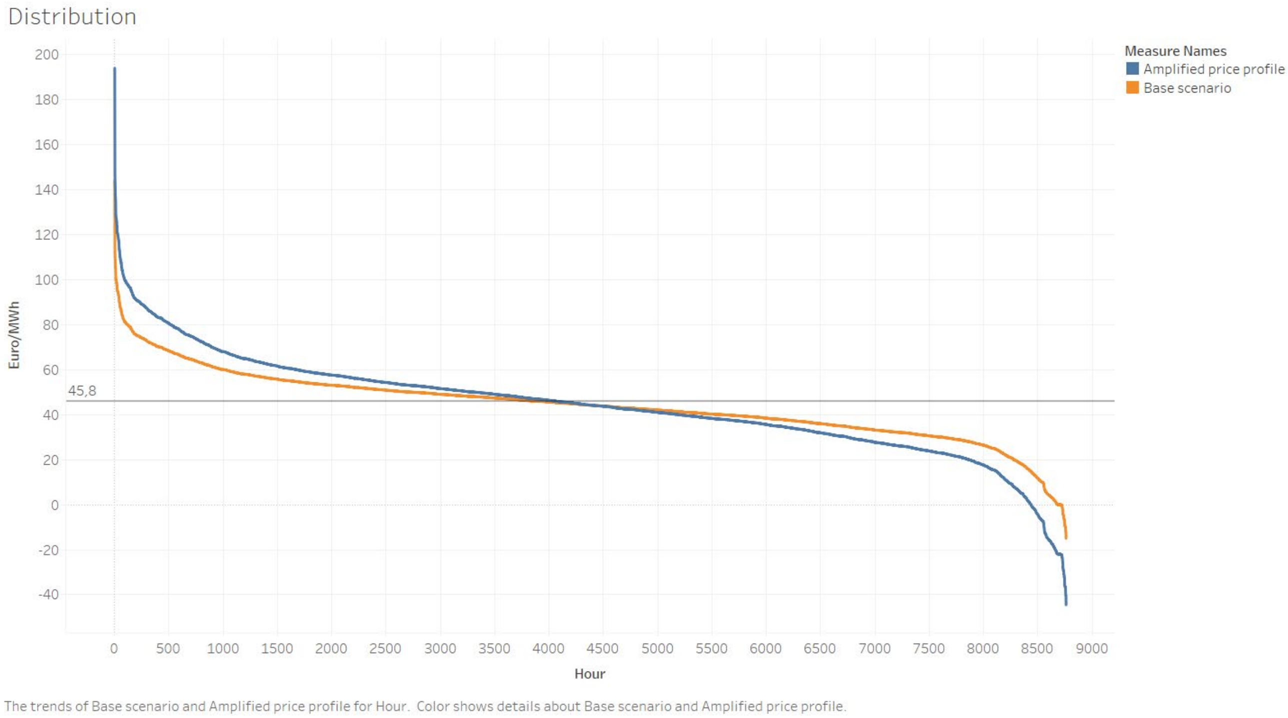


Figure 13, Distribution of prices to various hours over a year for respectively the baseline spot price (orange) and the amplified 2018 spot price (blue).

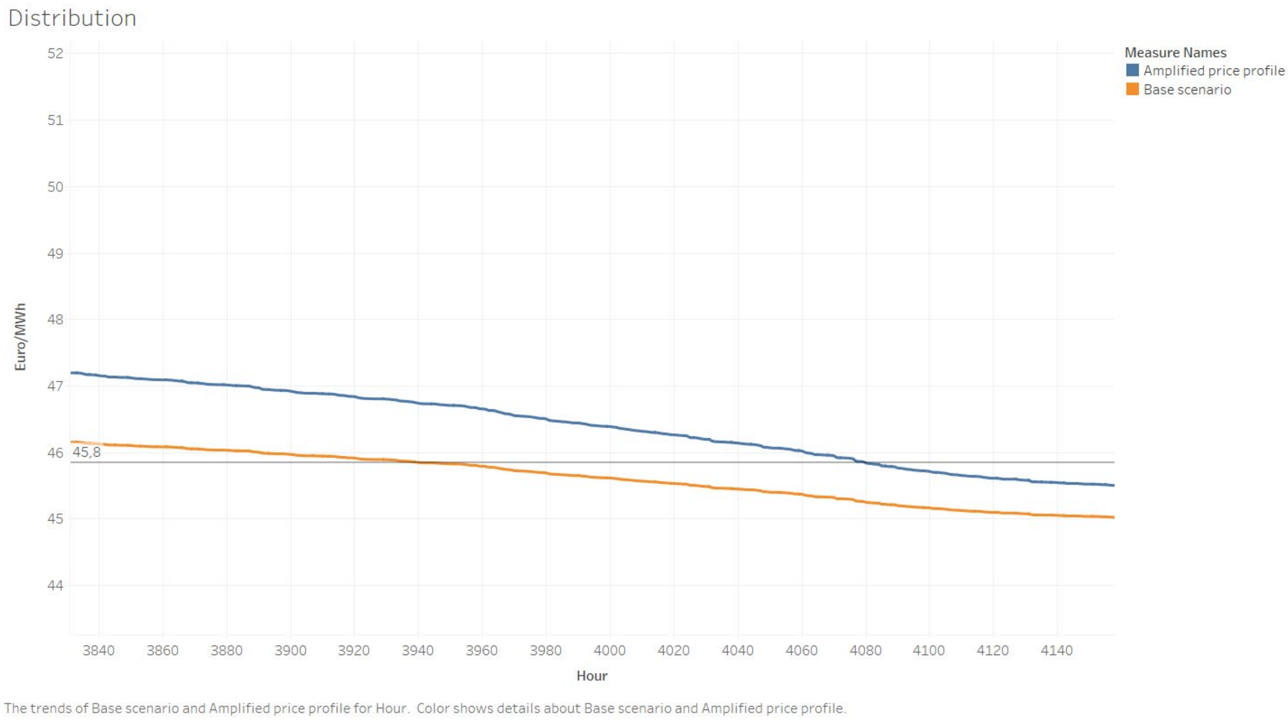


Figure 14, Cut of graph shown in figure 3.

Knowing that the baseline scenario has the ability to operate 55% of the time, corresponding to the 4818 hours with the lowest electricity price, it can be observed that this approximately corresponds to electricity price just below 46 €/MWh for the yellow curve. It is also observed from the blue curve, indicating the amplified spot price, has a lower number of hours where it is profitable to run the fuel factory. In this scenario, the plant runs 51% of the time, being 4% less than the reference scenario. The total electricity price pr. Ton methanol produced is 305 €, which is 62 € lower than that of the reference scenario, explaining the higher yearly operation income and with that the higher NPV after 20 years.

1.5.3.7 Low tariffs

The scenario using reduced tariff shows improved operational income, amount of methanol production and carbon captured. Here we observed the same effect as that of changing the spot prices to be lower than those of 2018. The decreased total electricity price (consisting of the spot prices plus the tariffs paid to the TSO and DSO), compared to the reference scenario enables more production hours for the plant. The production hours are increased to be 59% of the time during the year, instead of being only 55% in the reference scenario. With regards to electricity, the total production price including the tariffs is 420 €/ton, which is 56 €/ton lower than that of the reference scenario, therefore increasing the operation income and with that the net present value of the plant.

1.5.3.8 Conclusion

The electricity cost pr. ton, production hours for the different scenario are presented in the table below.

Scenario	**Reference**	**2015 spot prices**	**Changed amplitude**	**Reduced tariffs**
Electricity spot price pr. ton	366,22 €	217,44 €	304,98 €	378,91 €
Total electricity price pr. ton (with tariffs)	476,19 €	327,40 €	414,94 €	420,18 €
Production hours (% of total hours in a year)	55,2 %	61,7%	51,1 %	59 %
Operational profit (M EUR/Year)	90,5	196,8	116,9	131,8

The largest increase in operational income was created by lowering the spot price. The same effect was created by reducing the tariffs, however this change was not as significant as changed spot prices, because the effect on the total price pr. ton methanol is low. The changed amplitude reduced the total electricity price pr. ton, thereby increasing the operational income while also decreasing the number of hours where production would be profitable.

1.5.3.9 Heat utilization

The processes happening in the system produces excess heat. This heat is sold off in the reference scenario at a price of 20 €/MWh and having an excess heat production of 1.658 GWh it creates an operational income of 33 M€ annually. This heat is sold off to the district heating system of Aalborg, and in the reference scenario it accounts for 86% of the total heat demand. The system is dependent on being able to repel the excess heat to the district heating system, meaning that if there is no demand for the excess heat, the liquefier and the fuel factory will shut down, and not start up until the heat demand allows for it again. The heat demand is fluctuating every hour based on the values for 2018, having a highest demand 597,8 MW and the lowest at 79,1 MW. The limited heat demand, both yearly and hourly, could potentially limit the possible production hours for the system and with that decrease the methanol production.

In order to see the effects of being independent on the hourly price a seasonal storage is added to the system creating the possibility of storing the heat produced by the system to be used in times with a higher heat demand. To illustrate the effect of being independent of the yearly heat demand, a scenario is created where the heat can be rejected from the system without necessarily being sold to the district heat system. The heat will however be used in the system at times of demand. In order to see the effects on the business case of being able to sell the heat, a third test is conducted where the heat is not sold but the system is still dependent on having to supply a heat demand. The results of these three scenarios is presented in the following section.

1.5.3.10 Results of different heat utilization scenarios

The results from adding a seasonal storage, making the system independent of having to supply heat, and having no income from the heat are shown in the table below.

Scenarios	Operation profit [M EUR/year]	Net present value [M EUR]	Methanol production [ton/year]	CO2 captured [ton/year]
Seasonal storage	94,1	509	601.606	821.423
Excess heat rejection	107,6	677	846.274	1.155.490
No heat sale	67,8	181	696.990	951.660

1.5.3.10.1 Discussion

From the scenarios using different heat utilisation we can see that all but the scenario with no heat sale has a higher NPV than the refence scenario. In the "No heat sale" scenario, the methanol production is not limited to times where there is a heat demand in the distribution system, but at the same time it only produces methanol when it is profitable. Compared to the scenario including the seasonal storage the methanol production is considerable higher, but because the seasonal storage has an extra income of 34 M€/year, it exceeds the operational income and with that the total NPV.

The scenario that is able to reject the heat when there is no demand and still able to sell the heat when there is, can be considered a "best of both worlds scenario" Here the methanol production is at its highest and the yearly operational cost is 17 M€ higher than that of the reference scenario. The increased methanol production compared to the scenario with no heat sale, can be explained by the added income from the heat sale: this allows for the system to run in times where the electricity price might have been too high before, because the total sum of the system would still be positive i.e. give a profit.

1.5.3.11 Variable amplitude with respectively added liquefier capacity and fuel factory capacity.

The main operational cost for the system is the electricity purchase. The C3U system consists of two electricity consuming elements, the liquefier and the fuel factory. By increasing one of these two units, it would allow the plant to run more in times with low electricity prices thereby decreasing the overall operational cost but would at the same time increase the total investment cost of the system. In order to see the effects of this, the following scenarios increases the liquefier and the fuel factory capacity with 50%. The investment and operation cost of the increased liquefier and fuel factory are presented in the table below, along with the same costs of the reference scenario:

Investments [M EUR]	Baseline	150% fuel factory	150% Liquefier
Liquefier	73	73	95
C3U	65	65	65
Electrolyser	466	606	466
Fuel factory	60	78	60
Total Investment	664	822	686

Maintenance costs [EUR/hour of operation]	Baseline	150% fuel factory	150% Liquefier
Liquefier	310	310	403
C3U	280	280	280
Electrolyser	1950	2535	1950
Fuel factory	250	325	250

It is worth noting that despite being increased with 150%, the investment cost of both the fuel factory capacity and the liquefier is only being increased to 130%. This is done under the assumption that the last 50% increase will not be as expensive as the original 100%. The same applies for the maintenance cost.

As mentioned in the section on electricity price and the profile section, a future system highly relying on RES primarily from wind turbines and solar cells, could result in a more fluctuating electricity production, and with that a more fluctuating electricity price. Since this could prove most beneficial in scenarios with a higher adjusted production, this changed spot price profile is tested with both an increased liquefier and an increased electrolysis plant. The results are presented in the following section.

1.5.3.12 Results of added liquefier capacity and added fuel factory capacity

The results for using respectively an increased liquefier capacity and an increased fuel factory capacity at different spot price amplitudes are presented in the table below.

Scenarios	Operation profit [M EUR/year]	Net present value [M EUR]	Methanol production [ton/year]	CO2 captured [ton/year]
Fuel factory, Amp factor 1,0	121	687	623.616	851.476
Fuel factory, Amp factor 1,5	159	1.162	599.754	818.895
Liquefier, Amp factor 1,0	91	452	584.724	798.373
Liquefier, Amp factor 1,5	118	785	539.874	737.136

1.5.3.13 Result discussion

At both amplitudes it is seen that the fuel factory has the largest increase in operation income, resulting in a higher NPV. This can be explained by the fuel factory having a higher electricity usage and therefore benefit more from being able to produce at times with low electricity prices. The liquefier already has some flexibility in its operation, having a storage connected to it corresponding to one day's capacity. Adding an extra production capacity does create a better NPV and operation income compared to the reference scenario, but it is not nearly as high as the scenario using an increased fuel factory capacity.

1.5.3.14 Future scenario

Based on the conclusions from the various scenarios investigated a future scenario was designed to determine whether or not it is feasible to reach near fossil price parity. The scenario uses a fuel factory oversized by 150% to better utilize periods of low electricity cost, it assumes tariffs are reduced to 30 DKK/MWh when connected at the transmission system level and the price spread was amplified by a factor of 1,5 as mentioned under Section 4.3. Based on these assumptions a methanol selling price of 400 €/ton was set and the operational economy investigated as in previous scenarios.

As seen from the last row in Table 10, the annual income is comparable with the reference but the investment is higher due to the oversizing resulting in lower NPV. It is also evident that the plant gets relatively few operation hours and consequently produce relatively little methanol. This indicates that the carbon capture capacity can be better utilized by increasing the fuel factory size by 2-300% instead of only 150% as was used in this analysis.

1.5.3.15 Conclusions

The results from the different scenarios are presented in table 10.

Table 7 Summary of simulation cases

Scenarios	Income [M€/year]	NPV [M EUR]	MeOH production [ton/year]	CO2 captured [ton/year]
Baseline scenarios				
Reference	90,5	464	585.147	798.950
Reference incl. Seasonal storage	94,1	509	601.606	821.423
Reference, excess heat rejection	107,6	677	846.274	1.155.490
Reference, no heat sale	67,8	181	696.990	951.660
Reference, Amp factor 1,5	116,9	793	541.549	739.423
Price sensitivity				
Low tariffs (30 DKK/MWh)	131,8	979	625.814	854.477
Low Methanol price (400 EUR/ton)	17,2	-450	185.918	253.850
Spot prices from 2015	196,8	1.789	648.965	886.087
Flexibility analysis: 150% Fuel factory capacity				
Fuel factory, Amp factor 1,0	121	687	623.616	851.476
Fuel factory, Amp factor 1,1	129	780	617.793	843.525
Fuel factory, Amp factor 1,3	143	966	608.829	831.286
Fuel factory, Amp factor 1,5	159	1.162	599.754	818.895
Flexibility analysis: 150% Liquifier capacity				
Liquifier, Amp factor 1,0	91	452	584.724	798.373
Liquifier, Amp factor 1,1	97	518	575.257	785.447
Liquifier, Amp factor 1,3	107	651	556.964	760.470
Liquifier, Amp factor 1,5	118	785	539.874	737.136
Future scenario: 150% Fuel factory cap., Low tariffs, Amp factor 1.5, methanol price 400 €/ton				
Future scenario	93	332	436.530	596.031

The comparison between the net present values and the methanol production of the various scenarios are presented in respectively figure 16 and figure 17.

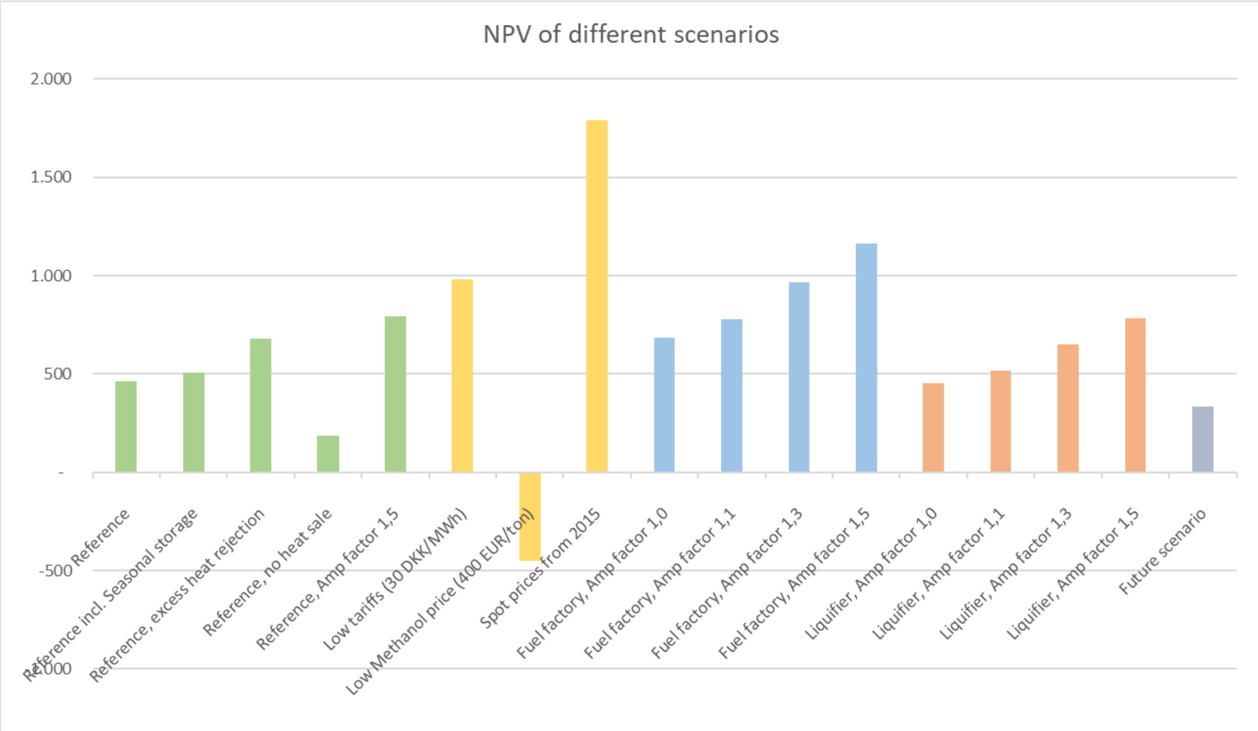


Figure 15 Net present values from the different scenarios. The color division corresponds to that presented in table xx.

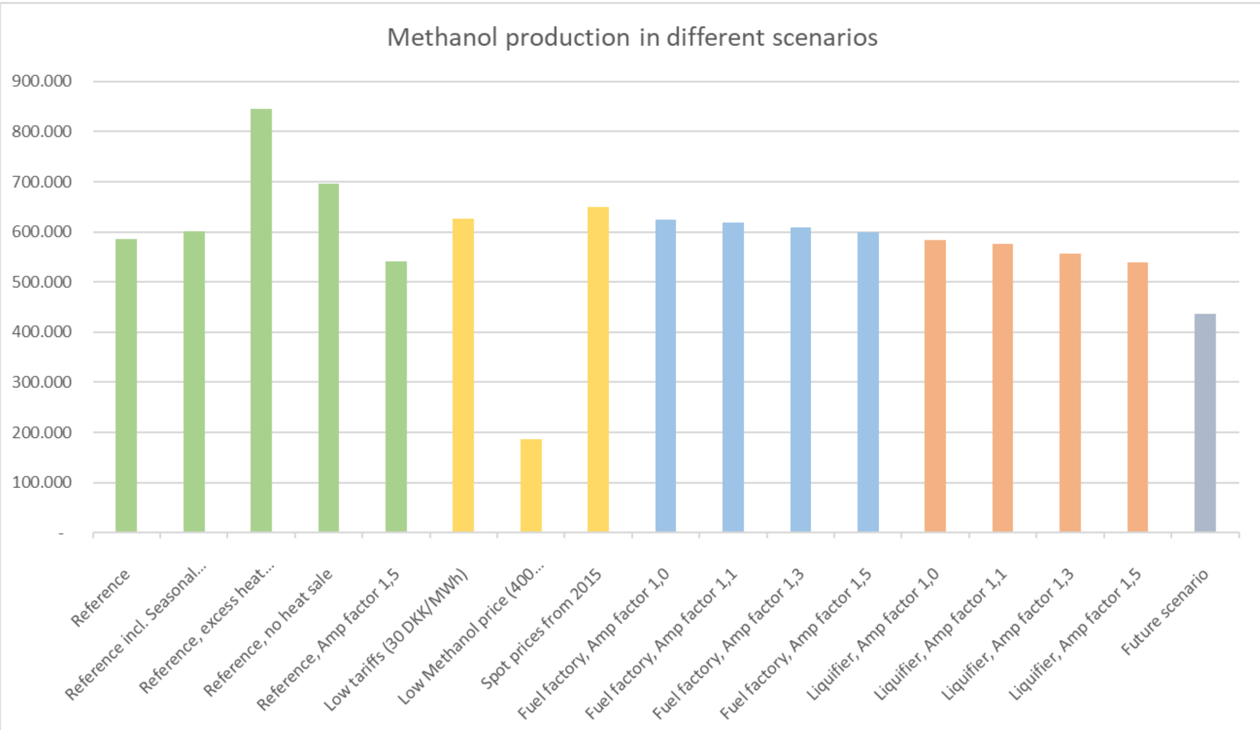


Figure 16 Methanol production from the different scenarios. The color division corresponds to that presented in table x.x.

From figure 16 and 17 it is seen that the electricity price is one of the highest influencing factors on the profitability of producing methanol. When comparing the values of operation income and methanol production from the reference scenario with the scenario using electricity prices from 2015, we can see that approximately 106,3 M€ pr. year corresponding to an increase of 117 %, while the methanol production is increased with only 63 818-ton pr. year, corresponding to an increase of 11%. This suggest that the increase is not created by the revenue caused by the extra production, but by the reduced cost of production.

The scenarios concerning heat rejection and sale, showed that the systems where the methanol production is dependent on the ability to sell heat to the district heating grid, had a generally lower production than those independent off the district heating system. Independence from the district heating system can increase the production of methanol up to 45%, as seen in the scenario: Reference, excess heat rejection. It is also seen that the income from selling heat is quite high: the scenario with no income from heat sale produces 19% more methanol pr. year, but only have 75 % the income of the reference scenario. Adding a seasonal storage to the system enables a higher flexibility of selling the excess heat. This created a 3% increase of methanol production and a 5% increase in the operational income.

Changing the capacity of the fuel factory and the liquefier enable the system to run with higher capacities and with that better avoid unfavorable electricity prices. Changing the fuel factory capacity has the largest impact on the operational income, it increases with 34% while the production only increases with 7%. Showing that the value of utilizing low electricity prices is very high.

The amplitude changes were conducted to simulate a system with highly fluctuating electricity prices. Here the results from the scenarios compared to the reference scenario shows that more hours with lower electricity prices can increase the operation income 29% while also decreasing the methanol production with 7%. Increasing the amplitude while also increasing the size of the fuel factory has the same effect as increasing the amplitude on the reference scenario. Any future research could be aimed at investigating whether changing the amplitude would keep on increasing the operation income or if there is a breaking point where the income will start to decrease.

The largest effect seen on the system, was decreasing the methanol selling price from 600 €/ton to 400 €/ton. Decreasing the price with 33%, decreases the operational income to 19 % of its original value and decreases the methanol production to 32 % its original value. The price reduction decreases the number of hours where the cost of producing methanol is lower than the selling price, why the net present value becomes negative. The price of 400 €/ton is closer to the current price of fossil-based methanol than the price at 600 €/ton which is used in all other scenarios. To investigate if fossil price parity can be achieved, a future scenario was defined with a 150% oversized fuel factory, reduced tariffs and an amplification of price variations to 150% of the 2018 level. The conclusion from this analysis was that near fossil parity is feasible with the assumed conditions and the fuel factory should be oversized by up to 300% possible by storing the hydrogen and maintaining the size of the fuel synthesis plant. The feasibility of the project will also benefit greatly from utilizing the oxygen from the electrolyzer which constitute a potentially very substantial source of income.

1.5.4 Dissemination of project results

The early results were presented in 2018 at a technical meeting in IDA supplemented by the article "Kan vi fryse CO og omdanne det til brændstof?". At the Nordic Clean Energy Week 2019, the project gave presentation on CO2 utilization as well as on Cryogenic Carbon Capture. The project background and preliminary findings were also presented in the article "Med CO2-fangst kan vi få vindmøllestrøm ned i tanken på lastbiler, busser og fly" published in the magazine Biopress in January 2019.

1.6 Utilization of project results

The project was a pioneer on large-scale carbon capture and utilization in Denmark and brought an increased focus on the possibilities offered by this technology both among the project partners and in the broader context through presentations at meetings and workshops.

The C3U project helped facilitate that Aalborg Portland decided to engage in the GreenCem project supported by EUDP as the main applicant. This is a major step towards realizing a CCUS project in Aalborg with Aalborg Portland as the main CO2 source.

At EMD the project continued the EnergyPRO modelling from the HyBalance project and extended the model to a complete CCU plant integrated in Aalborg's energy infrastructure with the district heat grid and heat suppliers. This modelling provided very useful input for the business case considerations and highlighted the key influencing parameters. This work probed the interest in further developing the simulation capabilities of EnergyPRO into the CCU/PtX field including more advanced features that consider the grid regulation services offered by the technology.

At Aalborg University the project initiated the studies of fuel synthesis from CO2 and kicked off the strategic decision to make electro-fuels a dedicated research program. It also formed the basis for the decision to engage in the Power2Met project as key partner hosting the industry-scale demonstration plant.

In terms of future perspectives, there is still a strong ambition for a large-scale CCUS demonstration project in Aalborg possibly including also a future CO_2 collection network that connects Reno Nord (municipal solid waste incineration plant), Randers Tegl (large brick production facility), and possibly Rockwool. From this hub biogenic CO_2 can be fully or partly utilized whereas the reminder mostly fossil CO_2 is stored. Altogether, the potential CO_2 volume handled will exceed 3.500.000 tons/year.

There are many similar projects in the making in the European Union and there will be strong competition for the government/EU funding that is a prerequisite for any CCUS project. Aalborg has advantages in terms of infrastructure that makes us believe that it has a good chance in the international competition.

1.7 Project conclusion and perspective

In conclusion, the C3U project showed that there is a large potential for carbon capture and use in Aalborg and given the right regulatory conditions a feasible business case can be established which is fundamental to attract the massive private investments required.

The project also found that full utilization of the 2,2 mio tons CO2 emitted by Aalborg Portland will generate more waste heat than the district heating grid in Aalborg can absorb today (even when Nordjyllandsværket has been decommissioned). The ability to use the waste heat is important to the overall efficiency as well as the business case. Development of more efficient electrolyzers and production of higher temperature waste heat can change this.

Looking into a future scenario of larger price spread on the electricity cost, a CCU plant – or at least the electrolyzer – will probably only operate around 5000 hours annually. The final business case optimization will depend significantly on the development of the grid regulation services market that can shift the balance towards fewer or more operating hours. Paying the full grid tariffs is not considered a viable option, it will have to be off-set by regulation service contracts or other payments for flexible operation otherwise the cost of the produced fuel will exceed what the market is willing to pay. The quantities of produced fuel from a large-scale CCU plant will require a large market which today can only be shipping or aviation.

Altogether, Aalborg is considered among the best suited locations for a large-scale CCU(S) demonstration project due to the available infrastructure and the fact that there are no other alternatives to achieve deep CO_2 reduction from cement production.

Appendix:

This section includes all the technical values used to model the Baseline scenario. All of these values will be kept through the rest of the sensitivity analysis, unless otherwise specified.

Fuel values:

- Heating value of methanol: 5,05 kWh/l or 6,39 kWh/kg
- Heating value of hydrogen: 3,55 kWh/l or 39,4 kWh/kg
- Heating value of natural gas 11 kWh/Nm3

Electrolyser

	Electric cons. (MW)	Water cons. (l)	Hydrogen prod. (MW)	Heat prod. (MW)
Full load	1167	1 L/Nm^3 H_2	919	248
Minimum load	10 % hot standby or off			

Fuel factory

	CO2 cons. (ton/hour)	Hydro-gen cons. (MW)	Metha-nol prod. (MW)	Heat prod. (MW)	Electric cons. (MW)	Water prod. (ton/hour)
Full load	170	919	795,6	0	35	69,5 (MeOH)
Mini-mum load	Without hy-drogen stor-age assumed to be 10 %					

Liquefier

	Electric cons. (MW)	Cooling delivered (MW)	Heat prod. (MW)
Full load	48,75 (alt. Sce-nario 62)		
Minimum load			

C3U plant

	Electric cons. (MW)	Cooling con-sumption (MW)	Flue gas cons. (ton CO2/h)	Liquid CO2 prod. (ton/h)
Full load	Baseload 24 MW (11 alt. Scenario)		170	170

Storage:

- Capacity of liquid natural gas should vary, starting at a capacity corresponding to one day use, but should be expanded if seen as a limiting factor.
- Same procedure should be taken when determining the capacity of the liquid CO2 storage.

Variable operating costs:

- Heat selling price: 20 €/MWH
- System tariff for Energinet: 80 kr/MWh
- Potentially a gas net tariff should be added.
- Selling price of methanol: 600 €/ton
- Water use costs: 10 kr/m3 (numbers from NEAS/Centrica)
- Variable maintenance for each plant equal to 2% CAPEX (Liquefier, Electrolyser and fuel plant)

Investment cost:

CAPEX for each plant pr. Installed capacity.

- Liquefier: 500 – 1000 $/ton LNG/year (insecure, but estimated to be quite high)
- Electrolyser: 0,4 mio €/MW
- Cryogenic Carbon Capture: 80 €/ton CO_2/year
- Fuel plant: 100 €/ton MeOH/year