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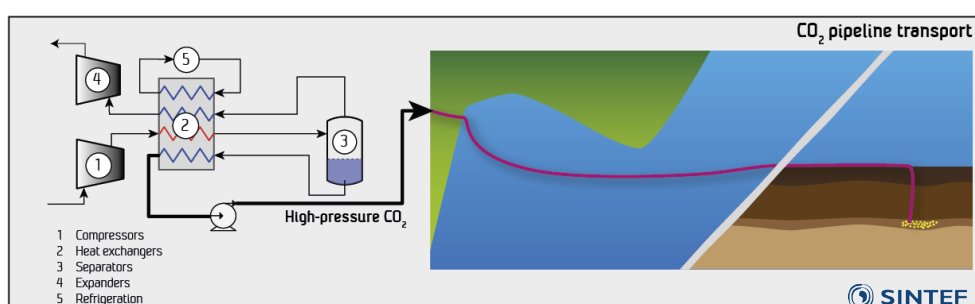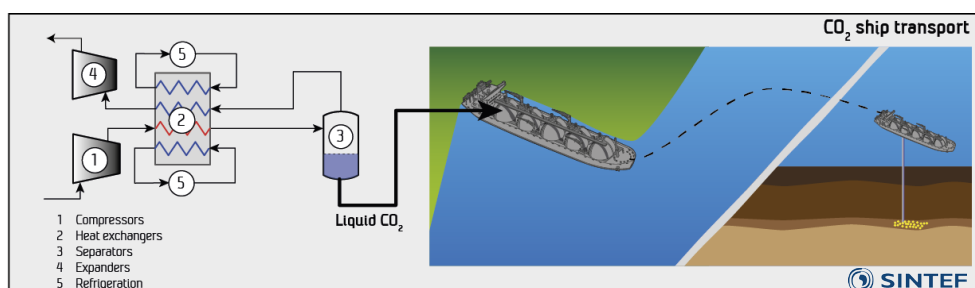## Low Temperature CO2 separation

Benchmarking and recommendations for further work

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ABSTRACT

Low Temperature CO_2 separation

A benchmarking of low-temperature CO_2 capture against solvent-based technologies was conducted, based on process performance calculations and cost analysis. Main results are:

- Low-temperature capture technology offers a clear potential for cost reduction and efficiency increase for pre-combustion capture from coal (IGCC).
- For post-combustion capture from coal, a hybrid concept (polymeric membrane followed by low temperature capture) is an interesting alternative to MEA-capture, including retrofit.
- Production of high purity hydrogen from gasified coal was investigated. The concept, a combination of H_2-selective membrane and low temperature capture, has potential to be refined into a self-sustained process for H_2 production that also produces the power required for CO_2 capture and H_2 liquefaction.

With low temperature technology CO_2 is captured in the liquid phase. This gives flexibility over time for choice of transport solution, compared to the reference solvent-based technologies. The results obtained are promising enough to justify further experimental and theoretical investigations of low temperature CO_2 capture technology.

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1 Executive summary – Low Temperature CO$_2$ Separation

In order to implement Carbon Capture and Storage (CCS) as a technology to mitigate climate change, challenges must be faced regarding energy efficiency and cost, as well as infrastructure roll out and financing. There may also be HSE concerns related to current solvent-based CO$_2$ capture technologies. In this context, low temperature capture technology, alone or in combination with other technologies, has been found to provide a promising alternative for power production, and may also be useful for industrial applications.

This project report presents a benchmarking of low-temperature CO$_2$ capture against reference solvent-based technologies. The benchmarking is based on process performance calculations and cost analysis. SINTEF expertise on low-temperature processing has been applied in the conducted investigations. This expertise has been built up over years through research in fields such as natural gas liquefaction, hydrogen liquefaction and use of CO$_2$ as a refrigerant.

Main project results are:

- Low-temperature capture technology offers a clear potential for cost reduction and efficiency increase for pre-combustion capture from coal. This is the application that has been investigated most in detail, with efficiency and cost calculations that cover processing for both pipeline and ship transport.

- For post-combustion applications, including CO$_2$ capture from industries such as cement and iron/steel, a hybrid concept (low temperature capture in combination with a polymeric membrane) was found to be an interesting alternative to the reference MEA-based technology, even though the cost advantage is smaller than for the pre-combustion case.

- Production of high purity hydrogen from gasified coal, using a combination of H$_2$-selective membranes and low temperature capture was investigated. The concept should be possible to refine into a self-sustained process that maximises H$_2$ production while producing the power required for CO$_2$ capture and H$_2$ liquefaction.

- The fact that CO$_2$ is captured in liquid phase with low temperature technology, gives flexibility over time for choice of transport solution compared to the reference solvent-based technologies that deliver captured CO$_2$ in gaseous phase.

Low-temperature capture of CO$_2$ from IGCC (pre-combustion capture)

The low-temperature capture technology has been most deeply investigated for the Integrated Gasification Combined Cycle (IGCC), which is a coal-based power plant where pre-combustion capture of CO$_2$ can be applied. It was found that for pipeline transport, compared to capture with the physical solvent Selexol, the power consumption for CO$_2$ capture was reduced with ~30%, and the cost of avoiding CO$_2$ emissions was reduced with ~35% (in €/t$_{CO2,avoided}$) compared to Selexol. Also for ship transport, the power consumption with low-temperature capture was reduced with ~30% compared to the best liquefaction technology applied after Selexol capture. In this case, the cost of CO$_2$ avoided was reduced with 40%.
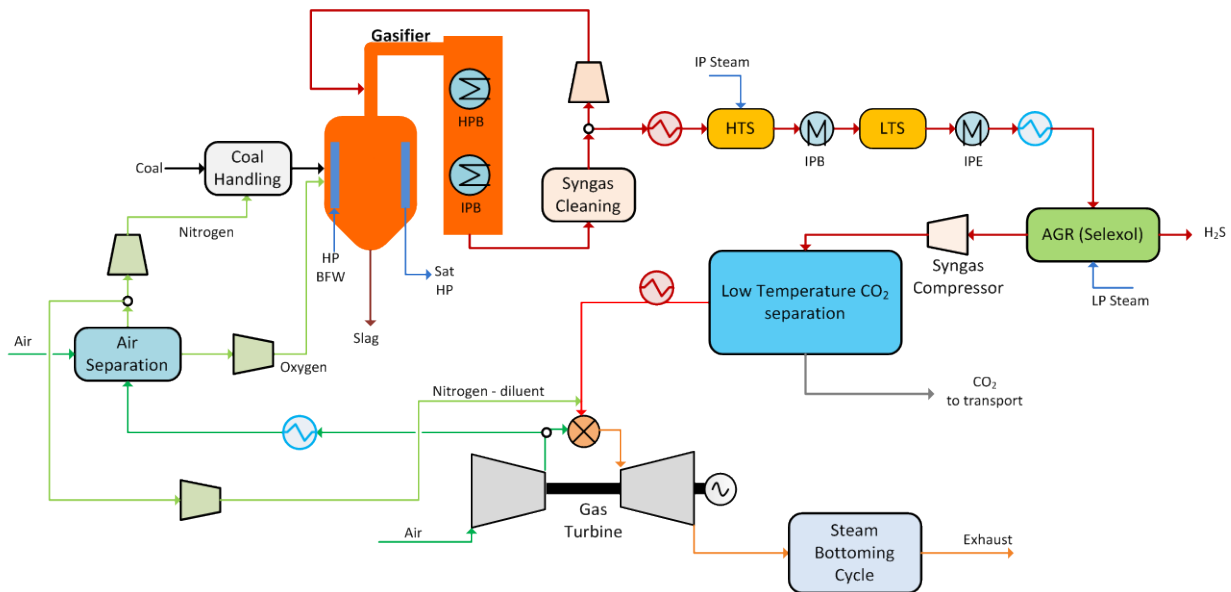
Figure 1. IGCC process with low-temperature CO₂ capture.

Post-combustion CO₂ capture from coal-fired steam power plants

An innovative hybrid concept was proposed for post-combustion CO_2 capture from coal fired power plants. The concept consists of two steps, first CO_2 enrichment of the exhaust through the use of a polymeric membrane, followed by a low-temperature capture process that separates CO_2 from the enriched exhaust gas. With the hybrid concept, the cost of CO_2 avoided was reduced with 9% compared to the reference post combustion process using MEA. The hybrid concept could be a retrofit option for steam power plants, since it requires no steam for solvent regeneration, but it could also be relevant to consider for newly built power plants. The hybrid concept should also be of interest to investigate for CO_2-emitting industries such as cement and iron/steel in future projects.

The hybrid concept with membrane and low temperature separation should also be investigated for natural gas fired power plants, to verify if any benefit could be obtained within this application. In particular, it could be interesting to investigate natural-gas fired combined cycles with Exhaust Gas recirculation (EGR).
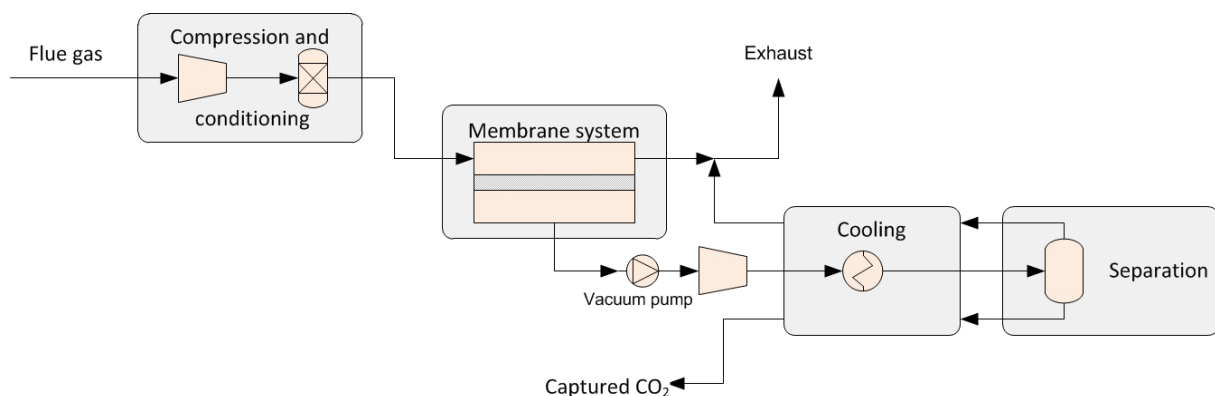
Figure 2. CO₂ capture hybrid process from CO₂-containing flue gas, using first a membrane system for CO₂ enrichment, thereafter low-temperature separation of CO₂.

High-purity H₂ production from coal with low-temperature CO₂ capture

Based on the same coal gasifier as in the IGCC study, a concept was proposed for production of high-purity hydrogen, which is an application of high interest for centralized production of H_2 with CCS. Hydrogen is separated from a shifted and desulfurized syngas through the use of H_2-selective Pd-membranes. Thereafter, low-temperature separation of CO_2 from the hydrogen-depleted syngas is conducted. The remaining low-heating value syngas, depleted of CO_2 and H_2, is fed to a gas turbine combined cycle that generates the steam required for gasification and water-gas shift, as well as a surplus of electricity after the power requirements of CO_2 capture and H_2 liquefaction are satisfied. This concept should be of interest to investigate further and compare to other technologies, and also to investigate for hydrogen production with CO_2 capture based on natural gas in future projects. It should be possible to develop the concept into a stand-alone process that delivers required steam and electric power, without any connection to an electricity grid.

Low-temperature capture and CO₂ infrastructure

It has become clear during the project execution that one of the key benefits of the low temperature CO_2 capture technology, compared to the reference technologies, is that CO_2 is captured in the liquid phase, while the reference technologies deliver captured CO_2 in the gaseous phase. Capturing CO_2 in the liquid phase provides a possibility for significant cost reductions for ship transport, compared to other liquefaction technologies. Furthermore, if the low-temperature process initially is designed and built for delivering CO_2 for ship transport, it is rather simple to modify the process later on to deliver CO_2 for pipeline transport. This means that transport and storage of captured CO_2 is possible in coastal areas already prior to a pipeline infrastructure is put into operation. This flexibility can enable postponing costly CO_2 infrastructure investments around e.g. the North Sea until a sufficient number of CO_2 suppliers are in place. Obviously, the low-temperature capture process can also be designed and built to deliver CO_2 for pipeline transport also from the beginning.

Figure 3. Transport flexibility when using low temperature CO₂ separation technology. Process redesign between the two transport options should be straightforward.

Low-temperature capture technology basics and future development

The idea behind low temperature separation is in principle fairly simple. CO_2 molecules are, relatively speaking, heavy and condense at higher temperature than lighter gas compounds such as e.g. hydrogen, nitrogen and oxygen. CO_2 can therewith, under the right conditions, be condensed as a liquid, while the other components from a CO_2-rich gas mixture mainly remain in the gaseous phase. For the condensation to take place, the gas mixture must be compressed and cooled down to a temperature close to -56°C.

No major obstacles have been identified for the low temperature capture technology. The process can be designed and built with known process components: heat exchangers, compressors, expanders, phase separators and a refrigeration circuit, using e.g. ethane and propane as refrigerants. Neither advanced materials nor large volumes of liquid solvents are required. The key challenge lies in the high precision that is required in process design and operation, but such high precision is known from other more complex processes, such as natural gas liquefaction, hydrogen liquefaction, air separation units and refrigeration technology. Altogether, it appears that, as the market for CO_2 capture evolves, it should be possible to progress towards implementation of the technology at full scale relatively quickly.

There are, however, still some knowledge gaps remaining, that need to be filled through lab and pilot scale investigations. The first knowledge gap that needs to be filled is the fact that current thermodynamic models, that describe the separation of liquid CO_2 from gaseous hydrogen, rely on a few experimental data points that are rather old. In order to increase the confidence in the results obtained in this report, measurement of CO_2/H_2 vapor/liquid equilibrium data should be conducted at relevant pressures and temperatures, and the results should be applied to update the process simulations. Further, non-equilibrium conditions that will occur in a real plant should be investigated more in detail.

2 Introduction

2.1 The Low Temperature CO$_2$ Separation project – objectives and contents

This technical report summarizes the findings from the "Low Temperature CO$_2$ Separation" project. The project is a pre-study under the CLIMIT-Demo programme (Gassnova Project Number 223123) and has been carried out from January till November 2013.

In the project application phase, the objective was defined to be "proving the potential of the low-temperature concept for CO$_2$ removal in the IGCC and establish the applicability window of the concept". At project startup, the scope of the project was further increased to cover two more concepts. Hence, the project has investigated low-temperature CO$_2$ separation processes applied to

- CO$_2$ capture from syngas in an IGCC process (pre-combustion capture)
- A hybrid concept for post-combustion capture, where bulk CO$_2$ separation through polymeric membranes is followed by low-temperature separation of CO$_2$
- CO$_2$ capture from hydrogen production through coal gasification (same gasifier as in the IGCC case). Here, high-purity hydrogen is separated through a palladium or palladium alloy membrane. The remaining CO$_2$-enriched stream is purified by low-temperature CO$_2$ separation.

It should be noted that for gas compositions with a high CO$_2$ content, such as in oxy-fuel applications, low-temperature CO$_2$ removal by cooling and partial condensation is already assumed to be state-of-the-art technology for CO$_2$ separation and capture (Berstad et al. 2013a). No oxy-fuel concepts were therefore included in this project.

It should also be noted that some work on low-temperature CO$_2$ separation already has been presented in the literature for pre-combustion applications (Berstad et al., 2013b; Mori and Forsyth, 2013)

In the present project, the IGCC application has been investigated in more depth, and two different process layouts with CO$_2$ capture for pipeline transport are presented. Furthermore, since liquefying CO$_2$ is the fundamental idea behind low-temperature separation, capture of liquid CO$_2$ that is ready for ship transport has also been investigated for IGCC. It should be noted that capturing liquid CO$_2$ for ship transport is equally applicable to the post-combustion capture and hydrogen production concepts, as well as any other process where low temperature CO$_2$ capture could be envisaged.

After the technical concepts were investigated, economic key performance indicators were determined, in order to enable a techno-economic benchmarking of the low-temperature capture technology against reference technologies (capture with Selexol for IGCC and capture with MEA for post-combustion capture). No investment cost or other economic results were determined for the H$_2$ production concept, since this is still in an early phase of investigation.

2.2 The need for different CO$_2$ capture technologies

The chart in Figure 4 gives an overview of CO$_2$ capture conditions (concentration and partial pressure at the capture point) for a wide range of power generation cycles and industrial processes. As can be observed, the driving force, in terms of CO$_2$ partial pressure, varies substantially with as much as four orders of magnitude, between different CO$_2$ sources. An important interpretation of this observation is that that different conditions for CO$_2$ containing streams, which occur in power generation and industries, will require different capture technologies for techno-economically optimal performance.
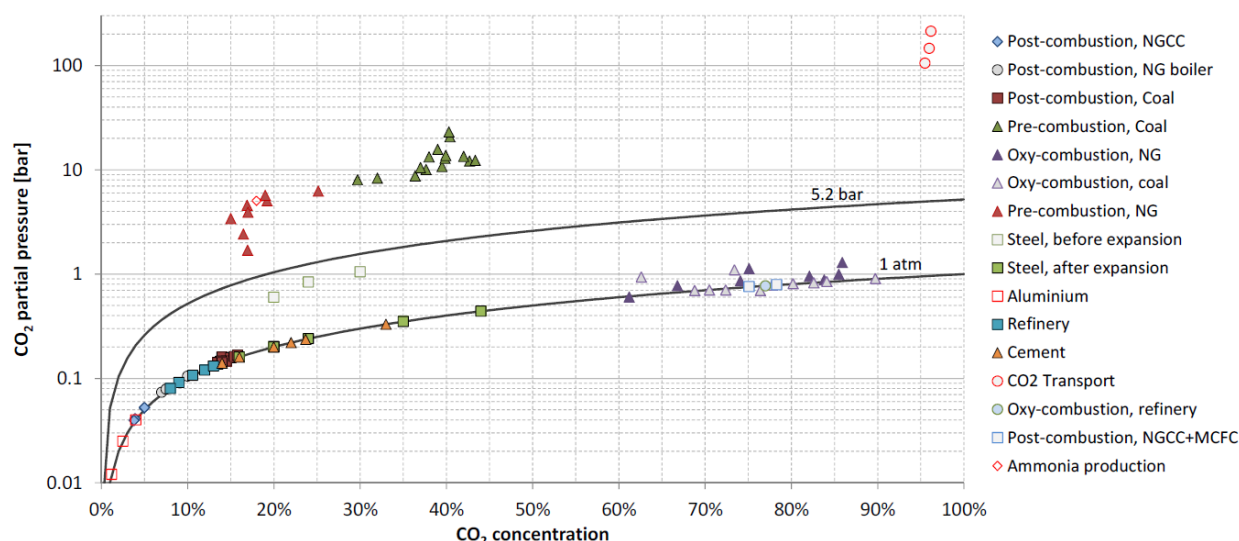
Figure 4. Capture conditions overview chart (Berstad et al., 2013a).

The techno-economically optimal technology for a certain CO_2-containing stream depends on several factors. In addition to the CO_2 concentration and partial pressure illustrated in Figure 4, parameters such as CO_2 diluents (e.g. N_2, O_2, H_2, CO), required CO_2 product purity, required CO_2 capture rate, energy cost will have an impact.

When investigating the applicability of low-temperature CO_2 capture from a CO_2 source, the CO_2 concentration cannot be too low, as this will give too low capture rate or far too high energy consumption. However, if a gas with low CO_2 concentration first can be subject to CO_2 enrichment through the use of another technology, low-temperature separation can yield satisfactory capture rate at an acceptable power input. An example of this is the membrane-hybrid concept for post-combustion capture from coal (section 3.2 in this report).

It should be noted that CO_2 separation by liquefaction is considered as state-of-the-art for oxy-combustion capture. The high CO_2 concentration of the product gas makes this choice obvious. Through previous work at SINTEF Energy Research, and in this project, the area of applicability of the low-temperature capture technology has been expanded to cover also pre-combustion capture from coal and post-combustion capture from coal. As can be seen in Figure 4, the CO_2-concentration is higher at capture conditions for coal for both pre-combustion and post-combustion capture routes, compared to natural gas. This made it reasonable to investigate capture from coal prior to natural gas. If positive results were obtained for coal, it would be relevant to proceed to investigate also natural-gas based pre-and post-combustion capture technologies.

2.3   Key performance indicators for benchmarking of power processes

Two types of power processes are investigated in this report, the Integrated Gasification Combined Cycle (IGCC) and the Advanced Supercritical (ASC) pulverized coal fired plant. Capturing CO_2 in the IGCC is done before combustion, i.e. it is a pre-combustion type of power plant with CO_2 capture. Capture of CO_2 from the ASC is done from the exhaust gas after the combustion process, i.e. it is a post-combustion capture process. In addition to these power processes, a concept for H_2 production from coal is presented, but this concept is at a too early stage of development to be benchmarked against a reference case.

In order to be able to conduct a benchmarking of different CO_2 capture alternatives and describe the increase in cost compared to power plants without CO_2 capture, five key performance indicators are employed:
- Net process efficiency

- CO_2 capture rate
- Net Present Value (NPV)
- Levelized Cost of Electricity (LCOE)
- Cost of CO_2 avoided.

The definitions of these are given in below.

2.3.1 Process efficiency

The net efficiency η_{th} of any process producing an amount of net power P_{net} from combustion of an amount of fuel m fuel with a lower heating value Q_{LHV} is

$$\eta_{th} = \frac{P_{net}}{Q_{LHV}}$$

2.3.2 CO_2 capture rate (CCR)

The CO_2 capture rate (CCR) is defined as the CO_2 captured divided by the total CO_2 available in a specific CO_2 source, e.g. a syngas or an exhaust gas from a power plant.

$$CCR = \frac{CO_{2\ captured}}{CO_{2\ available}}$$

2.3.3 Net Present Value (NPV)

Net Present Value (NPV) of project costs is the sum of the discounted cash flows during the project duration. The NPV is commonly used to benchmark projects and make investment decisions. One should note that usually, since there is no revenue supposed in the project, this NPV is the same as a classical Discounted Cash Flow (DCF). The NPV is calculated for each investigated power process assuming a real discount rate of 8%[1], 7,500 operating hours per year and an economic lifetime of 25 years (Anantharaman et al., 2011).

$$\text{Net Present Value of costs } = \text{ Investment} + \sum_{\text{project duration}} \frac{\text{OPEX}}{(1+ \text{discount rate})^k}$$

2.3.4 Levelized Cost of Electricity (LCOE)

The levelized cost of electricity (in €/MWh) is here used as a key performance indicator to measure the unitary cost of the electricity production of a plant with and without CO_2 capture. The electricity production cost approximates the average discounted electricity price over the project duration, which would be required as income to match the net present value of capital and operating costs for the project. It is equal to the annual costs divided by the annual net electricity production, as shown below.

$$\text{Levelized cost of electricity} = \frac{\text{Annualized investment} + \text{Annual OPEX}}{\text{Annual gross electricity production} - \text{Plant auto-consumption}}$$

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[1] This real discount rate of 8 % corresponds to a nominal discount rate around 10% if an inflation rate of 2% is considered.

2.3.5 CO$_2$ avoided cost

The CO$_2$ avoided cost (in €/tCO$_2$) is obtained by comparing the levelized cost of the plant with and without the CO$_2$ capture as shown below (Rubin et al., 2013).

$$CO_2 \text{ avoided cost} = \frac{(LCOE)_{CCS} - (LCOE)_{ref}}{(t_{CO2}/MWh)_{ref} - (t_{CO2}/MWh)_{CCS}}$$

Where:
- $(LCOE)_{CCS}$ is the levelized cost of electricity of the plant with CCS (€/MWh)
- $(LCOE)_{ref}$ is the levelized cost of electricity of the reference plant without CCS (€/MWh)
- $(t_{CO2}/MWh)_{CCS}$ is the CO$_2$ emission rate to the atmosphere of the plant with CCS (t_{CO2}/MWh)
- $(t_{CO2}/MWh)_{ref}$ is the CO$_2$ emission rate to the atmosphere of the reference plant without CCS (t_{CO2}/MWh)

3 Investigated power processes with low-temperature CO_2 capture

3.1 CO_2 capture from the Integrated Gasification Combined Cycle (IGCC)

3.1.1 IGCC power plants and process overview

One of the three main capture routes identified for reducing CO_2 emissions from fossil fuel power generation is the pre-combustion capture route. With this capture route, the energy contents of the fuel are transferred from a carbon-containing fossil fuel to hydrogen, while the carbon contents of the fossil fuel are converted to CO_2 and captured. For solid fuels (such as bituminous coal or lignite), the fuel conversion is based on syngas production through gasification in an Integrated Gasification Combined Cycle (IGCC).

The IGCC is of interest as an efficient means of producing power from coal also without CO_2 capture. Gasification of coal enables the production of a CO/H_2 syngas that can be burnt in a highly efficient gas turbine combined cycle for power production. This means that IGCC provides a possibility for coal-based power production with higher thermal efficiency than for pulverized fuel (PF) combustion where power is generated in a steam cycle. For instance, the European Benchmarking Task Force (Franco et al., 2011), calculated the efficiency of the ASC PF bituminous coal plant to be 45.5%, whereas the efficiency for the IGCC case was calculated to be 46.9%, based on present technologies.

In the 1990's, a few IGCC power plants were built to demonstrate the technology, notably in Buggenum (NL), Puertollano (ES), and in Wabash River, Tampa and Pinon Pine (USA), with a design point power output ranging from 95 to 320 MW (Olsson, 1999). Thereafter no major projects were planned for several years, until recently. The Edwardsport IGCC plant (618 MW) begun commercial operation in June 2013 in Indiana, USA. Fuel is bituminous coal, and no CO_2 capture is implemented. A lignite-fuelled 582 MW IGCC plant (Kemper County) is under construction in Mississippi, USA, and is estimated to be put into operation in the fourth quarter of 2014. The Kemper County plant will capture at least 65% of the generated CO_2, thus reducing the CO_2 emissions level to that of an Natural Gas Combined Cycle (NGCC) without CO_2 capture. The captured CO_2 will be transported by pipeline (61 miles, corresponding to ~98 km) for enhanced oil recovery (EOR).

Adding CO_2 capture to the IGCC process is in principle a straightforward, feasible process, and the capture unit will be considerably smaller and more cost-efficient than for post-combustion capture. Once the syngas (consisting mainly of H_2 and CO) has been produced in the gasifier, steam is added and the syngas is shifted to a H_2/CO_2-rich mixture. Thereafter the CO_2 can be separated from the H_2-rich fuel and the separated CO_2 is directed to transport and storage, while the H_2-rich fuel is burnt in the gas turbine. Also, since coal in general contains sulfur, H_2S must be removed from the syngas. This could be done either before or after the water-gas shift reactors. This sulfur-removal is typically carried out in a selective Selexol or amine process where H_2S is absorbed and removed from the flue gas. The inlet temperature into the Selexol unit should be 40°C according to Franco et al (2011), meaning that it is suitable to carry out the sulfur removal downstream of the water-gas shift reactors to avoid the energy penalty associated with cooling down the syngas only to reheat it again before the water-gas shift. A good overview of previous work on pre-combustion decarbonisation for coal with conventional technologies is given by Kunze and Spliethoff (2010). The IGCC process flowsheet with CO_2 capture using selexol is illustrated in Figure 5.
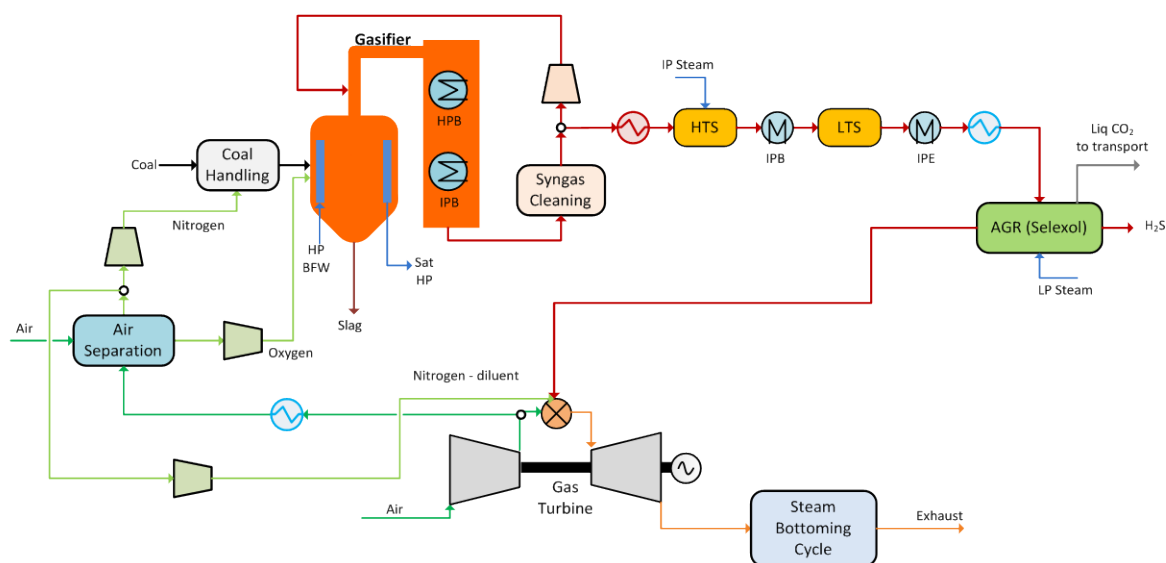
Figure 5. Process flowsheet, IGCC with CO₂ capture using Selexol for CO₂ and H₂S removal.

The conventional technology for performing the CO_2/H_2 separation is in a pressurized absorption-desorption process using a physical solvent. However, as initially observed in the DECARBit project and thereafter described by Berstad et al. (2013a,b) the composition of the shifted, pressurized syngas is well suited for CO_2 removal at low temperature. The process flowsheet of the IGCC with low-temperature CO_2 capture is shown in Figure 6. It should be noted that in practice the AGR (Selexol) process unit for this process will be significantly smaller than for the process illustrated in Figure 5.
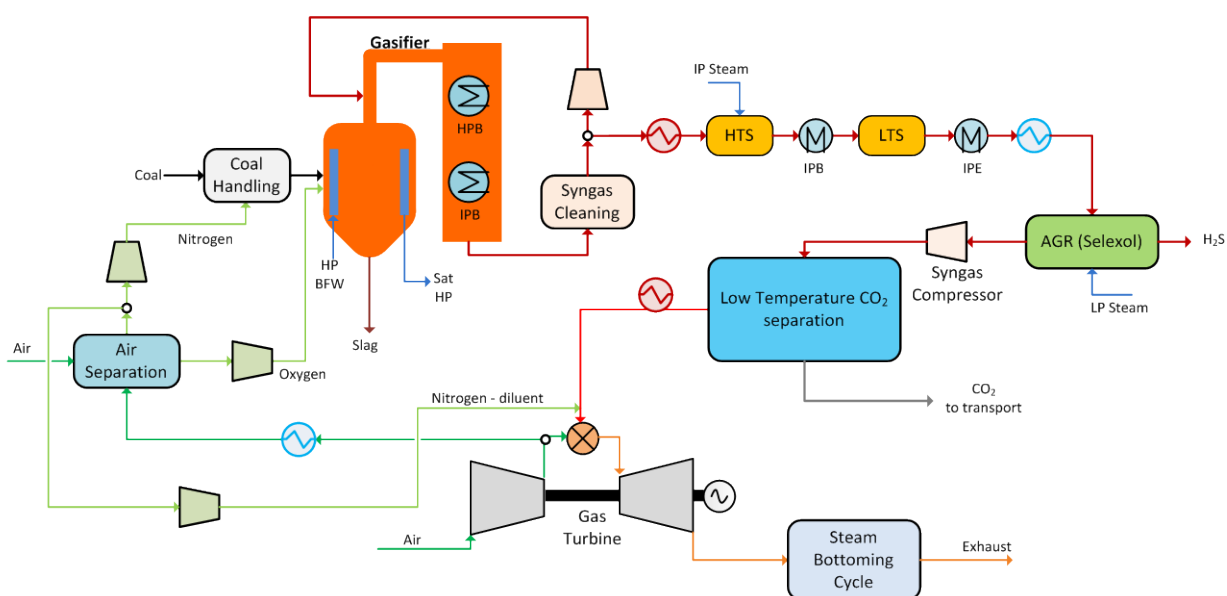
Figure 6. Process flowsheet, IGCC with low-temperature CO₂ capture, using Selexol for H₂S removal.

The composition of the desulfurized, shifted and dried syngas is given in Table 1.

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Table 1. Composition of shifted syngas after desulfurization and drying

Component	Hydrogen	CO_2	CO	N_2	Ar
Mol fraction	0.5375	0.3804	0.0160	0.0571	0.0090

The gasifier technology employed in the EBTF, and also in this project, is the Shell-type entrained flow dry-feed gasifier. Gasification pressure is set to 44 bar and the temperature to 1550°C. Fuel to the gasifier (bituminous coal) is fed with a pneumatic lock-hopper system using nitrogen with a small fraction of argon as the driving gas (this is the waste gas from the cryogenic ASU). The gas turbine of the IGCC is an F-class gas turbine and the HRSG/steam turbine bottoming cycle is a triple pressure configuration with reheat. CO_2 concentration in the exhaust is 7.75% and specific CO_2 emissions are 734 g CO_2/kWh net. It should be noted that since the CO_2 concentration is much lower in the exhaust for the IGCC, the post combustion capture solution would be much more energy demanding here than for an equivalent steam power plant such as that described in section 3.2.1

For the CO_2 transport product no specification has been defined apart from a transport pressure of 110 bar. The purity should however be as high as possible, thus minimizing the amount of H_2 dissolved in the captured CO_2.

3.1.2 Low-temperature capture from IGCC – baseline design

The process flowsheet for the baseline low-temperature separation process designed for CO_2 capture from an IGCC is shown in Figure 7. The syngas feed is assumed to undergo pre-treatment prior to cooling and partial condensation. After desulphurisation (H_2S removal) the syngas feed is assumed to be dehydrated to 0.1–1 ppm H_2O by molecular sieve adsorption, and the dry syngas is subsequently entering the inter-cooled compressor train where the pressure is increased from about 35 bar to 115 bar in two stages. The pressurized syngas feed then enters a heat exchanger network in which it is cooled to a final separation temperature of about -56°C.
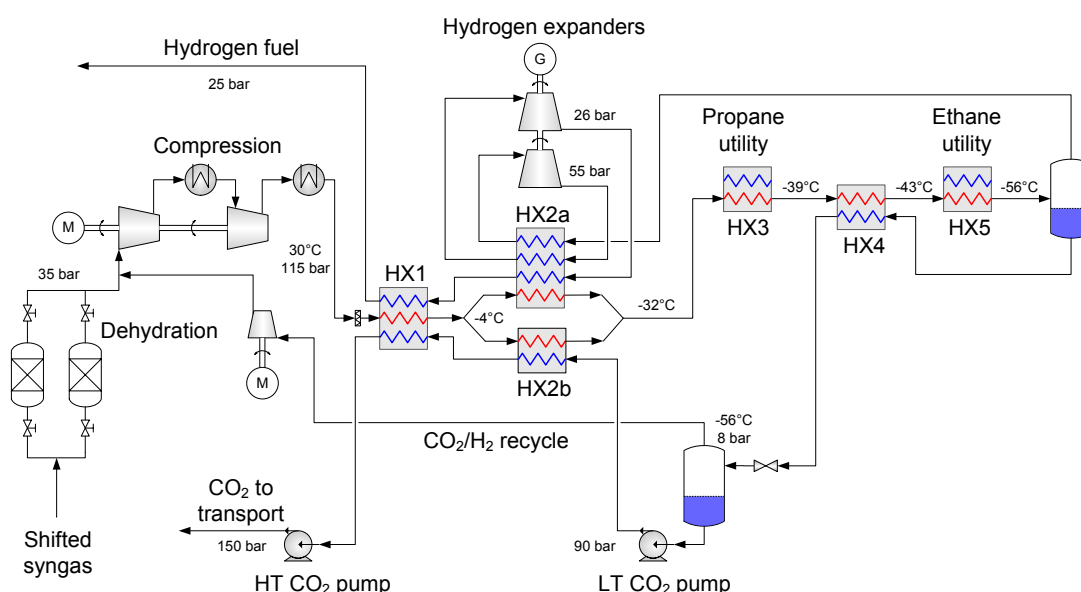
Figure 7. Baseline process flow diagram for the low-temperature separation process for CO_2 capture from shifted syngas in e.g. an IGCC process.

Based on the compressor, expander and heat exchanger efficiencies assumed in the simulation model, the energy performance of the baseline process before heat exchanger optimization is summarized in Table 2. 85% CO_2 capture rate was obtained in the simulation and as can be observed, syngas compression accounts for the main portion of the power requirement, about 22.4 MW. Second main power consumer is the refrigeration cycle, which consumes an estimated 5.7 MW, while water pumping, recycle compression and CO_2 pumping give minor contributions to the overall power consumption. The power generated by expanders is substantial; 9.4 MW, and the recovery of this power is crucial for obtaining high efficiency.

Table 2. Baseline design energy performance before heat exchanger optimization

Syngas compressors [MW$_e$]	Expanders [MW$_e$]	Auxilliary refrigeration [MW$_e$]	Cooling water pumping [MW$_e$]	Recycle compression [MW$_e$]	CO_2 pumping [MW$_e$]	Net power consumption [MW$_e$]	Specific energy consumption [kJ$_e$/kgCO$_2$ capt.]
22.392	9.445	5.691	0.209	1.123	1.446	22.122	275.6

In order to further verify the simulation results of the base case, geometric heat exchangers models were used for the largest and most complex heat exchanger (HX2a). After several iterations a feasible HX2a design was found and the energy performance after heat exchanger optimization can be seen in Table 3.

Table 3. Baseline design energy performance after heat exchanger optimization

Syngas compressors [MW$_e$]	Expanders [MW$_e$]	Auxilliary refrigeration [MW$_e$]	Cooling water pumping [MW$_e$]	Recycle compression [MW$_e$]	CO_2 pumping [MW$_e$]	Net power consumption [MW$_e$]	Specific energy consumption [kJ$_e$/kgCO$_2$ capt.]
22.392	9.575	5.427	0.205	1.123	1.457	22.067	274.9

Net electric efficiency of the IGCC with this optimized low-temperature separation process is 38.9% and net power output is 372.0 MW$_e$.

3.1.3 Low-temperature capture from IGCC – design for improved operability

The pressure-based energy contained in the high-pressure fuel stream leaving the main separator must be utilised in order to obtain a satisfactory overall efficiency for the baseline process. In the base case, the expansion of this stream takes place at sub-ambient temperature and provides cooling duty for the feed pre-cooling as well as power for driving compressors. By moving the expansion to higher temperatures a simpler process design can be derived from the viewpoint of heat exchanger design. A process flow diagram with so-called high-temperature expanders is shown in Figure 8, and as can be observed, the expander inlet temperatures are maximised by heat exchange against hot compressor outlet streams. The latter streams, however, have a temperature of 90–100°C only, limiting the expander inlet temperature.
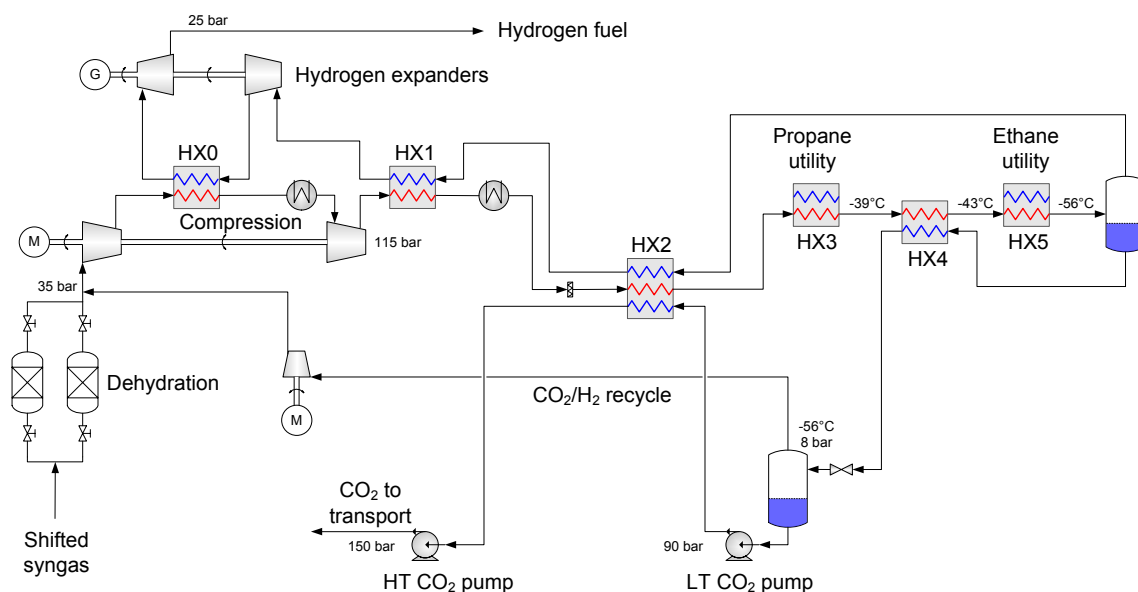
Figure 8. Process flow diagram for the low-temperature syngas separation process with high-temperature expanders.

With the expander inlet temperatures at about 70°C, obtained in the process model shown in Figure 8, the generated expander power output is higher than for the base case – a 36% increase from 9.45 MW to 12.9 MW. Furthermore, the power consumption in the auxiliary refrigeration cycle increased considerably due to lower degree of heat integration at sub-ambient temperature. This component increases by 87%, from 5.69 MW to 10.6 MW. Assuming 90% power recovery for expander shaft power, the net power consumption increases by about 7%.

Table 4. Energy design performance for the improved operability concept.

Compressors [MW$_e$]	Expanders [MW$_e$]	Auxilliary refrigeration [MW$_e$]	Cooling water pumping [MW$_e$]	Recycle compression [MW$_e$]	CO$_2$ pumping [MW$_e$]	Net power consumption [MW$_e$]	Specific energy consumption [kJ$_e$/kg CO$_2$ capt.]
22.216	12.875	10.631	0.215	1.112	1.404	23.990	299.3

The net IGCC efficiency is 38.7%, i.e. it is slightly reduced with the improved operability concept. More comprehensive process integration may contribute to further efficiency improvements. If the fuel streams were integrated with higher temperatures available in the main power cycle the power output from each expander would be further increased.

3.1.4 Low-temperature capture from IGCC – internal CO$_2$ liquefaction for ship transport

Most capture technologies deliver CO$_2$ in gaseous phase, leading to high energy requirements for CO$_2$ liquefaction for ship transport. Different options for large-scale liquefaction of gaseous CO$_2$ for ship transport are described by Aspelund et al (2004) and Aspelund and Jordal (2007), where CO$_2$ can, depending on process be compressed to somewhere between ~30 and 95 bar before liquefaction at 6-7 bar. As can be observed in the process flow diagrams in this report, the low-temperature capture technology generates CO$_2$ in the liquid phase at 6–10 bar after the second phase separator, potentially reducing the energy requirement considerably, thus, reducing the number of processing steps for generating liquid CO$_2$.

The layout of a low-temperature syngas separation process delivering liquid CO$_2$ for ship transport can be seen in Figure 9. The main difference from the baseline design is the internal heat exchange of the captured liquid-phase CO$_2$ stream. Since the LCO$_2$ for ship transport is required to have a temperature of at least -

50°C or lower and a pressure level somewhat above the CO_2 triple point pressure (above 5.2 bar), the state of the captured CO_2 stream is well adjusted to ship transport after the purification unit. Unlike for the baseline design where CO_2 is subsequently pressurised by pumping and heated by internal heat exchange, the liquid-phase CO_2 is now transport-ready.
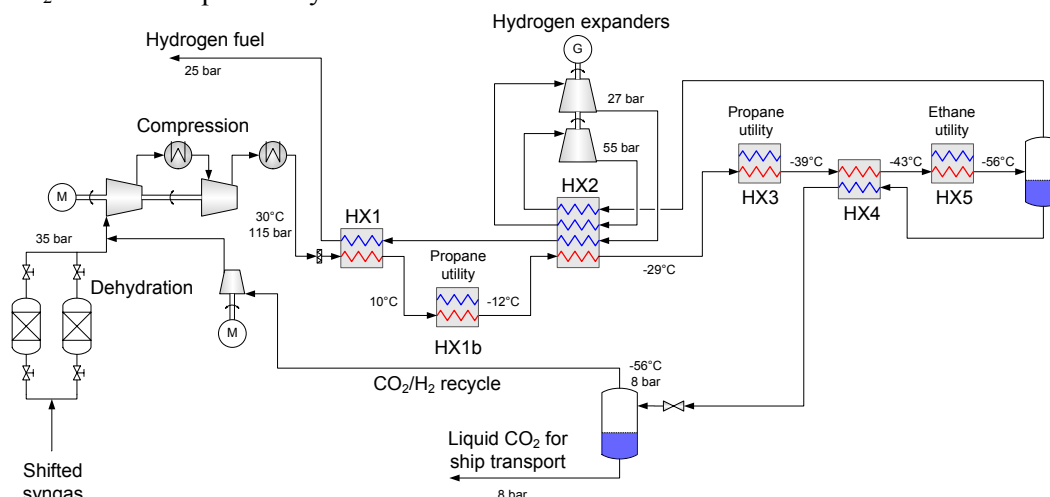
Figure 9. Process flow diagram for low-temperature syngas separation process for ship transport of liquid CO_2 (LCO_2).

As the LCO_2 stream will have to be kept at low temperature, it is not available for utilisation for syngas cooling, as was the case in HX1 and HX2b in the baseline design. The reduced internal heat exchange thus has to be compensated for by increased utility cooling load, and as the required additional cooling duty arises at rather high temperatures, an additional propane cooling level is considered to be appropriate. This is indicated by the additional heat exchanger HX1b between HX1 and HX2 (HX2a is the baseline design). From HX3 to HX5 there are no differences between the designs. The same applies upstream in the syngas pre-compressor train as well as for the recycle loop from the CO_2 purification separator. It is therefore the area between, and including, HX1 and HX2 where the process changes are visible, as well as the additional auxiliary refrigeration level. The energy figures for this process scheme are summarised in Table 5.

Table 5. Process performance for direct liquefaction of CO_2 for ship transport.

Compressors [MW$_e$]	Expanders [MW$_e$]	Auxilliary refrigeration [MW$_e$]	Cooling water pumping [MW$_e$]	Recycle compression [MW$_e$]	CO_2 pumping [MW$_e$]	Net power consumption [MW$_e$]	Specific energy consumption [kJ$_e$/kg CO_2 capt.]
22.138	8.783	9.986	0.290	1.114	n/a	25.702	320.7

The capture process for liquid CO_2 for ship transport is more energy-intensive than the baseline design for pipe CO_2. The net process efficiency is 38.5 % and the net power output is 368.4 MWe.

Compared to the pipeline case delivering compressed CO_2 at 150 bar, the specific energy requirement increases from 276 kJ/kg$_{CO2}$ to 321 kJ/kg$_{CO2}$. To a certain extent, this is an expected result when considering the exergy content (Table 6) of LCO_2 and high-pressure pipe CO_2. From the numbers in Table 6, for pure CO_2 for simplicity, the exergy level for ship CO_2 (-56°C, 8 bar) is about 5% higher than for pipeline CO_2 (15°C, 110 bar). The net power consumption for the ship CO_2 case, however, is 16% higher than the baseline design, and the former process therefore has a lower efficiency than the latter. It is anticipated that the efficiency of the CO_2 capture case for ship transport can be improved through further process optimization.

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Table 6. Thermomechanical exergy level for pipe and liquid CO₂.

	T [°C]	P [bar]	h [kJ/kg]	s [kJ/kgK]	e [kJ/kg]
Ambient	15	1.013	497.38	2.708	0
Pipe	15	110	228.92	1.0738	202.2
Ship	-56	8	81.204	0.52561	212.4

As the LCO₂ process could be deployed in the early stages before a pipeline infrastructure is built, process designs allowing for later modification to pipeline CO₂ transport should be of high interest. For now, however, the designs have been made for LCO₂ and pipe-CO₂, without considering the modification options. An optimal LCO₂ process designed with the possibility to modify it for delivering pipe-CO₂ will probably differ somewhat from the presented design, and efficiency may have to be reduced slightly to allow for flexibility in CO₂ capture process design over the CO₂ chain lifetime.

3.1.5 Alternative processes for liquefaction of CO₂ captured from IGCC

In order to give a more complete picture of CO₂ liquefaction as such and provide a benchmark for the low-temperature capture process with internal liquefaction, two other liquefaction processes were considered:

- The liquefaction process using ammonia liquefaction cycle, shown in Figure 10(a), has been proven by Alabdulkarem et al. to be the most energy and cost-efficient solution for liquefaction of gaseous CO₂ (Alabdulkarem et al., 2012). This process cannot be applied to CO₂ captured with low-temperature technology, since CO₂ captured with low-temperature technology never is available the in gaseous phase. Hence, this liquefaction technology can therefore only be considered after CO₂ capture with Selexol within the framework of the technologies investigated in this report.

- The liquefaction process by expansion of supercritical CO₂, shown in Figure 10(b), has been considered especially in the ZEP report (European Technology Platform for Zero Emission Fossil Fuel Power Plants (ZEP), 2011) due to the flexibility of the system when converting it from a shipping to a pipeline transport system. The COCATE project has however showed that when starting from a CO₂ stream at ambient conditions, this liquefaction system is 18% more expensive than the ammonia cycle liquefaction (Bureau-Cauchois et al., 2011). Although expansion liquefaction of supercritical CO₂ can be applied after the low-temperature CO₂ capture process, it does not make sense, since it actually means that the CO₂ is liquefied twice, both inside the capture process and after the capture process. This alternative was however investigated in this project, to provide a reference energy consumption (35.2 MW$_e$) as a comparison for the expansion liquefaction option when capturing CO₂ with Selexol (43.0 MW$_e$).
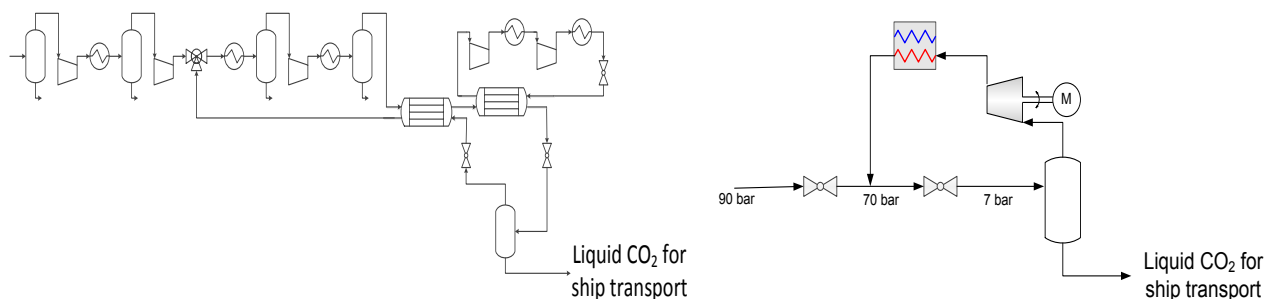
Figure 10. (a) Liquefaction by ammonia cycle (Roussanaly et al., 2013) (b) liquefaction by expansion.

The calculated energy consumption for each liquefaction option for each possible combination of CO₂ capture and liquefaction for ship transport, are given in Table 7.

Table 7. CO$_2$ capture, compression and liquefaction power consumption for possible combinations of capture and liquefaction processes.

Capture process	Ammonia cycle (Figure 10a)	Expansion liquefaction (Figure 10 b)	Low Temperature internal liquefaction (Figure 9)
Low temperature capture		35.2 MWe	25.7 MWe
Selexol capture	35.2 MWe	43.0 MWe	

The low-temperature separation with internal liquefaction is, with an electricity consumption of 25.7 MWe, by far the most energy-efficient option for supplying CO$_2$ for ship transport. The ammonia cycle is indeed more energy efficient than the expansion liquefaction technology, but still it exceeds the energy consumption of the low temperature internal liquefaction significantly (with 37%). In addition, CO$_2$ capture with internal liquefaction inherently provides a possible future simple adaptation to a pipeline transport network if/when such a network becomes available at the CO$_2$ capture site. It must be emphasized in this context that, although the investigation is made for CO$_2$ capture from the IGCC, the flexibility in terms of transport options is general for capture from all CO$_2$ point sources. Provided that the CO$_2$ concentration is sufficiently high, it is expected that the advantage in terms of low energy penalty for CO$_2$ liquefaction is valid also for CO$_2$ point sources other than the IGCC.

3.1.6 Low temperature capture of CO$_2$ from IGCC – summary of energy performance

So far, section 3.1 in this report has described low-temperature CO$_2$ separation processes as an alternative to CO$_2$ capture with Selexol in the IGCC power process. The low-temperature separation technology employs only electric power for CO$_2$ capture, and does not require any process integration to obtain steam or other process heat from the IGCC process. As a summary, and in order to visualize what happens in the processes, the internal power consumption and power generation of the low-temperature CO$_2$ capture technology is shown in Figure 11.
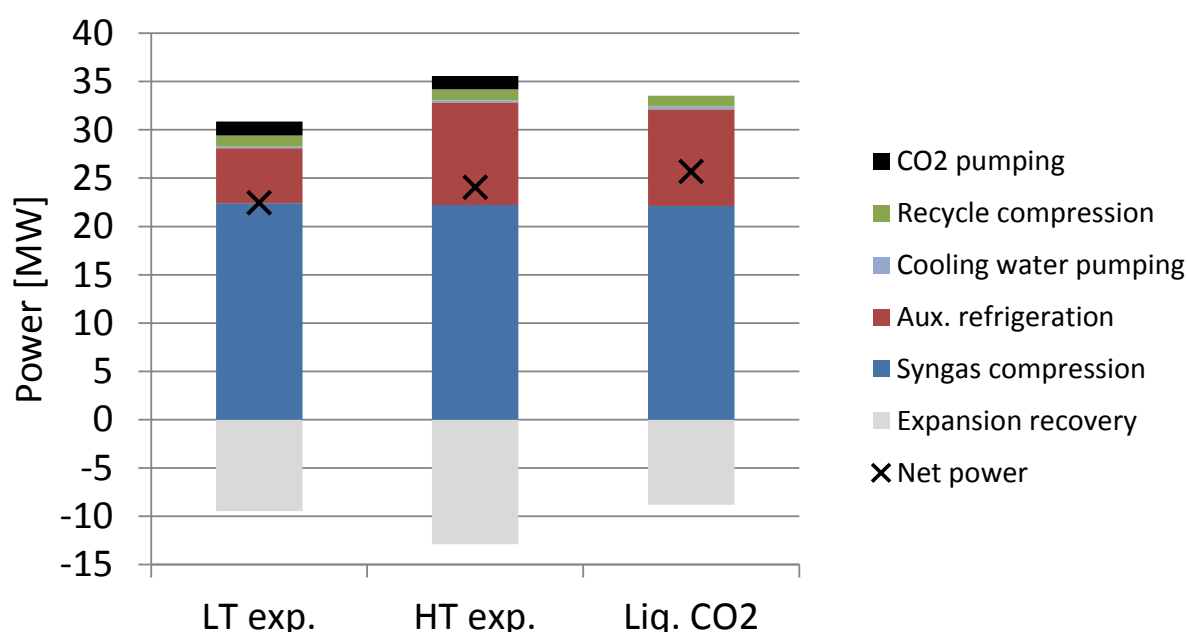

Figure 11. Decomposed power figures for the baseline case (LT exp.), improved operability case (HT exp.) and internal CO$_2$ liquefaction for ship transport (Liq CO2).

The main observations that can be made based on Figure 11 are:
- The syngas compression power does not change for the three cases, as this is governed by the targeted CCR and thus separation pressure
- Auxiliary refrigeration duty for the 'HT exp.' and 'Liq. CO_2' cases (shown in Figure 8 and Figure 9 respectively) are considerably higher than for the 'LT exp.' Case (shown in Figure 7). This is due to lower internal cooling from cold product streams
- Running the process in 'HT exp.' mode makes the process generate more power by expansion of H_2 rich fuel, but also consume more energy due to lower internal heat recuperation
- Energy requirements related to recycle compression, cooling water pumping and CO_2 pumping are considerably lower than for the syngas compression and auxiliary refrigeration for all cases.

3.2 Hybrid membrane/low-temperature CO_2 capture process for post combustion capture

3.2.1 The ASC reference plant without and with post-combustion capture

Coal-fired steam power plants are by far the dominating technology for generating electricity from fossil fuels, with high availability and relatively low cost of electricity. Power is generated in a Rankine cycle with water/steam as working fluid in a power process that has undergone evolutionary development over decades. Materials development set the limitations on the steam operating parameters in terms of pressure and temperature. The steam data for the Advanced Supercritical Coal (ASC) plant with pulverized bituminous coal that were selected for the EBTF reference plants were 300 bar and 600°C at the boiler exit (270 bar a at the HP turbine inlet), with single steam reheat to 610°C at 60 bar a. The CO_2 concentration at the exhaust stack is 13.7% and the specific CO_2 emissions are 763 g CO_2/kWh net, without CO_2 capture. These exhaust data are employed as the starting point for the investigations described in this project report.

3.2.2 Membrane separation of CO_2 from exhaust gas

Gas separation membranes are considered among one of the promising emerging capture technologies for post-combustion capture and this process has been studied extensively, from material innovation related to material development aspects to the process design, integration and assessment (Bounaceur et al., 2006; Favre, 2011; He et al., 2009; Merkel et al., 2010). Membrane technology has a few advantages over amine based solvent systems in that the process has no steam requirement and is thus easy to retrofit into existing power plants, and it has no environmental impacts caused by solvent losses and degradation. The disadvantage of this technology is limitations in membrane selectivity and permeability, multiple stage processes and significant compression work are required to attain high CO_2 capture ratios and product purity while ensuring sufficient driving force. While there is no clear consensus in the literature in comparing performance of amine based solvent systems and membrane systems, earlier work has shown that membrane processes are more energy intensive than amine-based post-combustion capture (Zhang et al., 2013).

3.2.3 Hybrid concept overview

Due to the low triple point of pure CO_2 (216.58 K, 5.185 bar), separating CO_2 from a gas mixture with the liquefaction method, especially for high CO_2 partial pressure gases, is quite attractive and close to commercialization in the oxy-combustion carbon capture process (Berstad et al., 2013a). Interestingly enough, the permeate from the 1st stage of a polymeric membrane separation process can reach relatively high CO_2 concentration (around 70%) which is similar to that of oxy-combustion exhaust. This permeate is here proposed to be used as a feed gas to a low temperature separation process. The purpose of this approach is to investigate whether a membrane process combined with a low temperature method has potential to

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reduce the capture energy penalty when both technologies are combined to match their operating windows. The principle of such an integration can be seen in Figure 12.
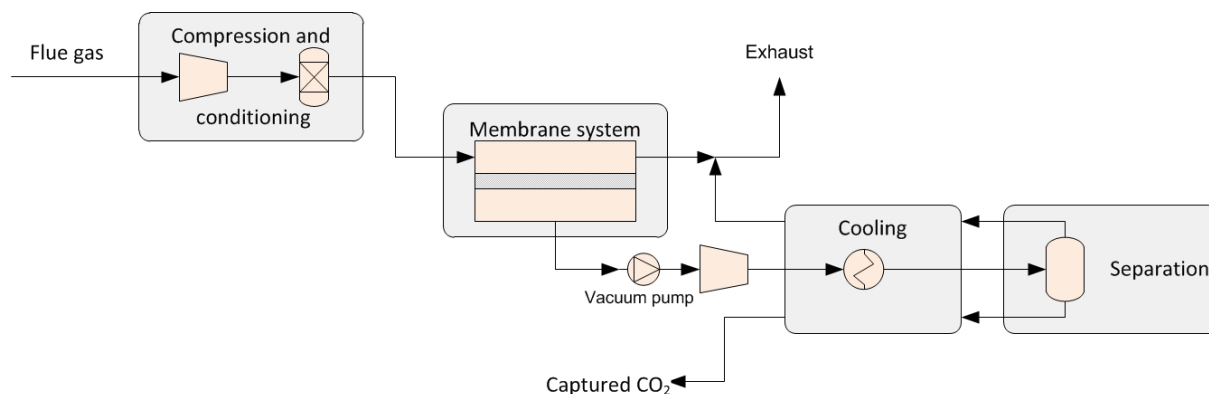

Figure 12. Process block diagram for membrane low temperature hybrid process for post-combustion capture.

In the hybrid process, the CO_2 concentration in the flue gas is increased using a single stage membrane process. This gas is then compressed and CO_2 is separated using a low temperature separation process. The single stage membrane process is designed to capture 90% of the CO_2 from the exhaust gas. The purity of the "CO_2 rich" permeate stream depends on the membrane properties, driving force and membrane area. The relevant membrane properties, selectivity and permeance are fixed to 80 (CO_2/N_2) and 5 $Nm^3/m^2 \cdot bar \cdot hr$ respectively. The driving force is the CO_2 partial pressure difference across the membrane. This is manipulated by pressurizing the membrane feed and/or operating the permeate side under vacuum. The energy penalty of a membrane process increases with increased purity of the permeate stream. The CO_2 rich permeate stream is fed to a low temperature capture unit where the CO_2 capture rate (CCR) and energy penalty depend on the CO_2 concentration of the feed. The higher the CO_2 concentration of the permeate, the higher the attainable CCR and lower the capture penalty in the low temperature process. The CO_2 purity of the permeate from the membrane process determines the overall capture process energy penalty and attainable CCR. To identify the permeate purity at which the overall energy penalty is minimum at an acceptable CCR is at the crux of this work.

3.2.4 Capture unit inlet gas composition and boundary conditions

Flue gas from a coal fired power plant with stream properties given in Table 8 is used as the feed to the hybrid capture unit. The permeate from the membrane unit was set to different purities: 52 – 75% and fed to the low temperature capture unit.

Table 8. Stream properties for coal fired power plant exhaust gas.

Temperature	85	°C
Pressure	1.02	bar
Mass flow	781	kg/s
Composition		
CO_2	13.7	mol%
N_2	72.9	mol%
O_2	3.7	mol%
H_2O	9.7	mol%

3.2.5 Process performance

After passing through the membrane, the CO_2-enriched flue gas is subject to a similar kind of processing as the syngas processing described earlier in this report. The gas is pressurized in a multi-stage intercooled compression process, and thereafter cooled to -56°C using external cooling utilities and cooling available by converting the pressure exergy in the flue gas stream. Liquid CO_2 is condensed at these conditions and pumped to a transport pressure of 110 bar. A simplified process flow diagram is shown in Figure 13.
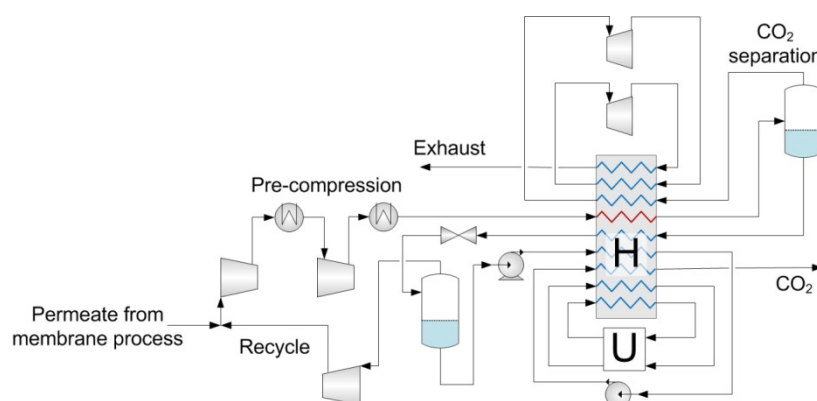
Figure 13. Simplified process flow diagram for low-temperature capture from CO_2-enriched exhaust.

As mentioned earlier, the energy penalty for capture and CCR depend on the CO_2 concentration in the permeate gas from the membrane process. The energy penalty for the low temperature process is evaluated at different possible values for CCR for the various CO_2 concentrations in the permeate. The results are plotted in Figure 14. The results show that:

- The specific energy penalty for the low temperature process decreases as the CO_2 concentration in the permeate increases.
- There is an optimum CCR, from an energy penalty perspective, for each CO_2 concentration in the permeate.
- There is a sharp increase in the specific energy penalty as the CCR approaches its highest attainable value.
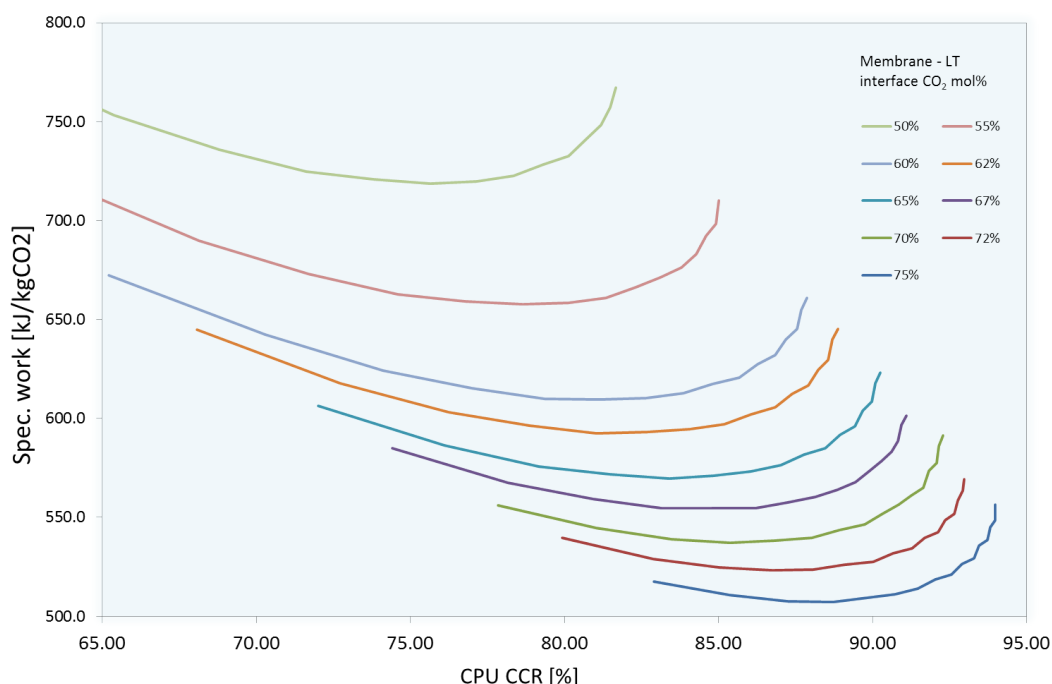
Figure 14. Specific energy penalty for the low temperature capture process as a function of the CCR for different permeate CO$_2$ compositions.

The total power consumption (or efficiency penalty) depends on the trade-off between the energy penalty in the membrane process for producing a permeate with high CO$_2$ concentration, and the energy penalty for separating this CO$_2$ in the low-temperature process. Further, the achievable CO$_2$ capture ratio also depends on the CO$_2$ concentration in the feed to the low-temperature process. Figure 15 shows how the system efficiency and the CO$_2$ capture ratio vary with the different permeate CO$_2$ concentrations. It can be seen that an increase in the capture ratio results in decreasing the overall system efficiency. This relationship is approximately linear from 50-85% capture ratio, after which the efficiency drops significantly for a small increase in capture ratio. Hence, rather than selecting the typically assumed CCR of 90% that is used by e.g. the EBTF, it makes sense when studying in the membrane/low-temperature hybrid concept for coal-fired power plants to select a CCR of 85%. Therefore, this CCR was chosen to evaluate the economics of the hybrid process with the permeate CO$_2$ concentration set to 67%. With the selected operating point, the net power output is 549.5 MWe, and the overall process efficiency with CO$_2$ capture is 33.4%.
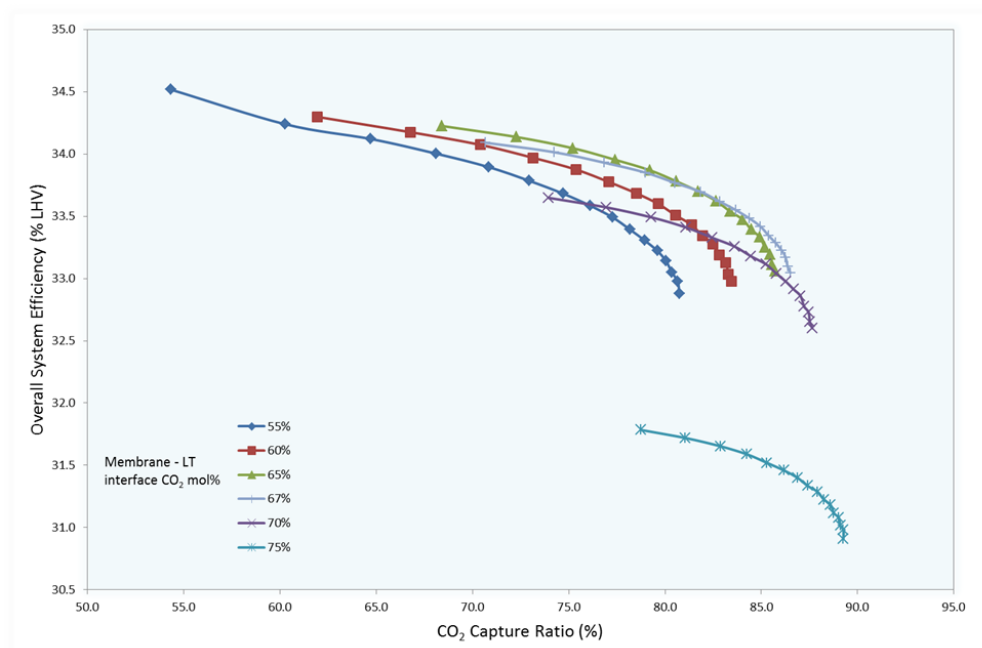
Figure 15. Overall post combustion capture efficiency and achievable CO$_2$ capture ratio as a function of membrane permeate CO$_2$ concentration.

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4 Benchmarking of power processes with CO_2 capture for pipeline transport of CO_2

This chapter describes the benchmarking of the IGCC and ASC power processes with CO_2 capture for pipeline transport. In the cases with low temperature CO_2 capture, this means that CO_2 is pumped up till the dense phase at a pressure of 110 bar, whereas for the reference cases with CO_2 capture using solvents, gaseous CO_2 is first compressed and thereafter pumped to 110 bar. A benchmarking of different concepts for liquefaction of CO_2 for ship transport is presented in chapter 5.

4.1 Efficiency and CO_2 capture rate

For delivery of CO_2 at pipeline transport conditions, the performance data of the IGCC and ASC plants and their reference cases are given in Table 9 below

Table 9. Power output, net electrical efficiency and CO_2 capture rate for investigated low-temperature capture concepts for pipeline transport (in bold face) and their reference cases.

	Net power output [MWe]	Net efficiency [%]	CO_2 capture rate [%]
IGCC, no CO_2 capture	391.5	46.9	---
IGCC, CO_2 capture with selexol (reference technology)	362.5	37.9	81.9
IGCC, low-temperature CO_2 capture for pipeline transport	**372.0**	**38.9**	**81.4**
ASC, no CO_2 capture	754.0	45.5	---
ASC, CO_2 capture with MEA (reference technology)	549.2	33.5	89.0
ASC with hybrid membrane-low temperature CO_2 capture for pipeline transport	**549.5**	**33.4**	**85.0**

It can be seen that for the IGCC case, there is a clear efficiency improvement of 1 %-point when applying low-temperature CO_2 capture compared to Selexol. This efficiency improvement is not possible to obtain with the hybrid membrane/low temperature concept for post-combustion capture from the ASC plant, which is in parity with the reference technology.

However, it should be noticed that the hybrid concept for post combustion capture has a clear advantage in terms of simplicity for process integration, since no heat integration is required between the power plant and the capture process. This indicates an advantage for retrofit options and also that the technology could be of interest to evaluate also for industrial sources with high concentration of CO_2 in the exhaust, such as cement and iron/steel.

More detailed results from the power output and process efficiency calculations can be found in appendix B.

4.2 Net present value

In order to be able to calculate the NPV for the power processes with low temperature capture, and compare them against the reference cases from the European Benchmarking Task Force (EBTF), CAPEX and OPEX were first calculated for the processes using the cost evaluation methodology described in Appendix C. Costs are evaluated using 2008 as cost basis year to make the calculations directly comparable to the results from the EBTF.

Table 10. CAPEX and OPEX and NPV for the investigated low-temperature capture cases for pipeline transport, cost basis 2008.

	CAPEX [M€$_{2008}$]	OPEX [M€$_{2008}$/y]	NPV [M€$_{2008}$]
IGCC, no CO_2 capture	926	95.1	1855
IGCC, CO_2 capture with selexol (reference technology)	1133.5	112.5	2230
IGCC, low-temperature CO_2 capture for pipeline transport	**1086.5**	**107.2**	**2135**
ASC, no CO_2 capture	1770	169	3552
ASC, CO_2 capture with MEA (reference technology)	2012	198	4099
ASC with hybrid membrane-low temperature CO_2 capture for pipeline transport	**2006.0**	**179**	**3891**

The CAPEX, OPEX and NPV of the different coal-based power production concepts with or without capture are shown Table 10. However it is important to consider the NPV as an intermediate result, since different power plants have different power output (Table 9). It is the cost of electricity and CO_2 avoided cost that are directly comparable. It should also be noted for the ASC plant that the MEA reference technology cost is assessed for a newly built plant, whereas the hybrid membrane/low-temperature technology cost is applicable both for a newly built power plant and for a retrofit plant of the same size.

4.3 Levelized Cost of Electricity (LCOE) and CO_2 avoided cost

The definitions of these two key performance indicators are given in sections 2.3.4 and 2.3.5.

4.3.1 Cost summary table

In order to compare the low temperature capture to the reference technologies, the cost evaluation was done on the same basis as the EBTF, and results are therefore directly comparable to the EBTF cases. CO_2 avoidance cost calculations include compression/pumping up till 110 bar. As can be seen in Table 11, the LCOE and CO_2 avoided cost are lower for the low-temperature capture technologies, both for the IGCC case and the ASC case.

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Table 11: LCOE and CO₂ avoided cost of the IGCC and ASC power plants with CO₂ capture for pipeline transport.

	LCOE [$€_{2008}$/MWh]	CO₂ avoided cost [$€_{2008}$/$t_{CO2\ avoided}$]
IGCC, no CO₂ capture	64.6	---
IGCC, CO₂ capture with selexol (reference technology)	86.2	33.3
IGCC, low-temperature CO₂ capture for pipeline transport	**78.7**	**21.7**
ASC, no CO₂ capture	59.6	---
ASC, CO₂ capture with MEA (reference technology)	94.4	52.9
ASC with hybrid membrane-low temperature CO₂ capture for pipeline transport	**88.7**	**47.9**

4.3.2 Analysis of results for the IGCC plant

The cost of electricity for the IGCC cases is visualized in Figure 16. The cost of electricity with low temperature CO₂ capture is only 18% more expensive than the plant without capture. It is further 10% cheaper than the reference plant with Selexol CO₂ capture. When looking at the cost of capturing CO₂, the low temperature concept, with a cost of 21.7 $€_{2008}$/$t_{CO2,avoided}$, is 35% more cost-efficient than the Selexol process as shown in Figure 17.
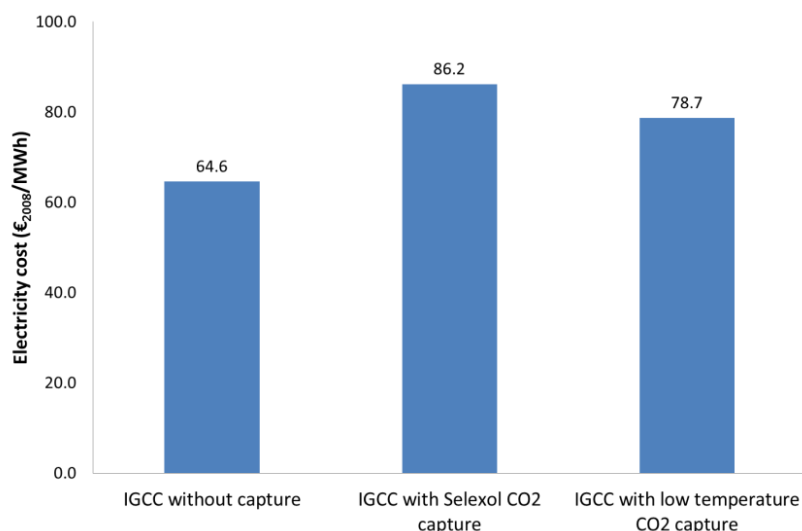

Figure 16: Electricity costs of the IGCC case, the IGCC case with selexol CO₂ capture and the IGCC case with low temperature CO₂ capture
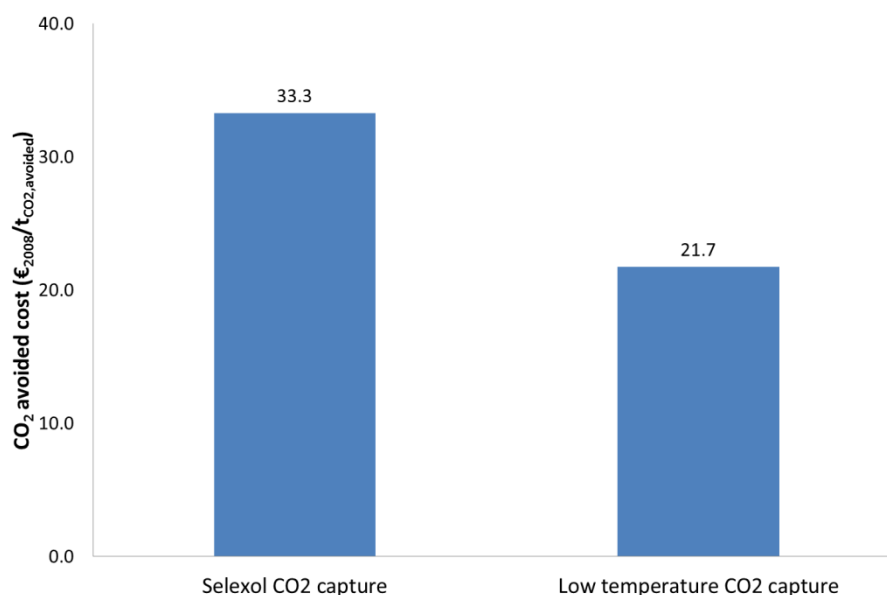
Figure 17: CO$_2$ avoided cost of the Selexol and low temperature capture technologies for the IGCC.

Sensitivity analyses were run on the capture process investment cost and operating and maintenance cost (CAPEX and OPEX in Table 10), to illustrate their impact on the IGCC cost of electricity as shown in the two uppermost bars in Figure 18. To facilitate the interpretation of the results, the electricity cost of the IGCC with Selexol capture (86.2 €$_{2008}$/MWh, refer to Table 11) is added as a red line. It can be seen that even with an uncertainty of 50% on capture investment and O&M cost, the resulting cost of electricity of the IGCC with low-temperature technology is well below that of the IGCC with Selexol capture.

Furthermore, sensitivity analyses were made of fuel price, operating and maintenance cost, utilization rate, project duration and discount rate as shown in Figure 18. Here, no direct comparison is made in the diagram with the reference IGCC with Selexol, since these five parameters are not directly related to the capture technology, but to the overall operating conditions of the power plant, regardless of capture technology employed. These results show that the IGCC with low temperature capture exhibit the same sensitivity trends as the IGCC with Selexol capture (Anantharaman et al, 2011). This means that the uncertainties connected to fuel cost, project duration, discount rate and plant utilisation rate are far more important for plant electricity cost than the CAPEX and OPEX for the capture unit in itself. In this context it should be noted that the CAPEX of the low-temperature CO$_2$ capture system is 5.9% of the total calculated IGCC plant CAPEX and the OPEX of the low-temperature CO$_2$ capture system is 2.1% of the total calculated IGCC plant OPEX.

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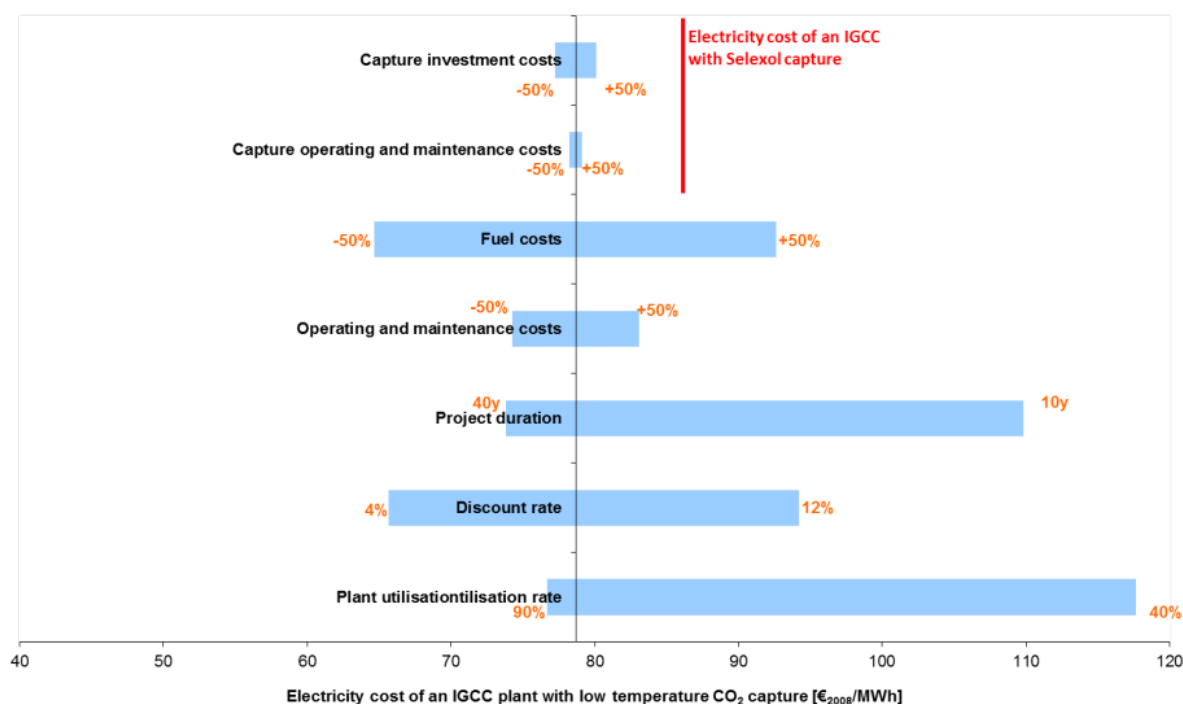
Figure 18: Sensitivity analyses on COE of the IGCC plant with low temperature capture for pipeline export

4.3.3 Analysis of results for the ASC plant

The cost of electricity for the different ASC cases is visualized in Figure 19. The cost of electricity with hybrid membrane/low temperature CO_2 capture is 33% more expensive than the plant without capture. It is further 6% cheaper than the reference plant with 30%wt MEA CO_2 capture. For the cost of capturing CO_2, the hybrid/low temperature concept, with a cost of 47.9 $€_{2008}/t_{CO2,avoided}$, is 9% more cost-efficient than the MEA process as shown in Figure 20.

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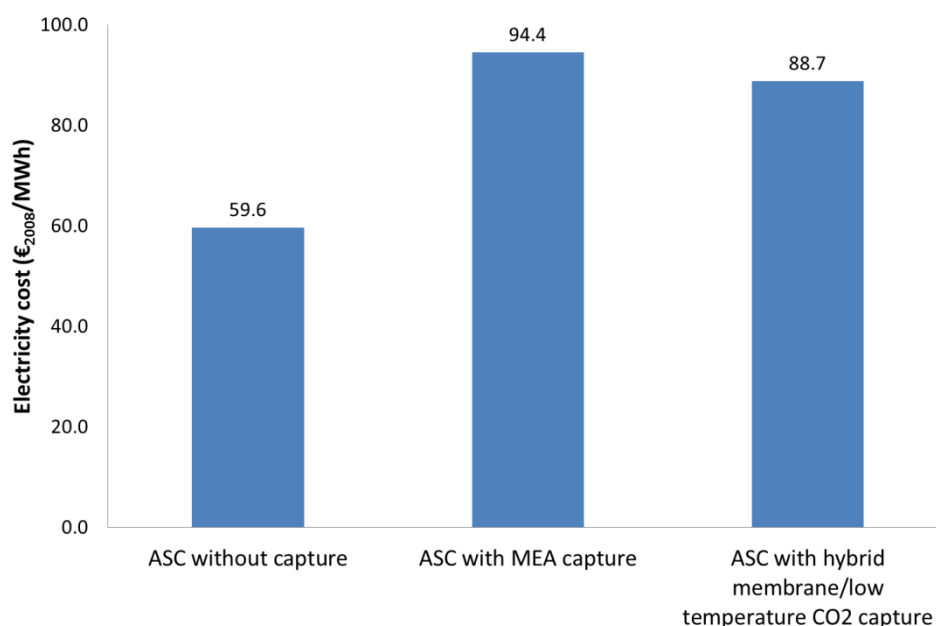
Figure 19: Electricity costs of the ASC case, the ASC case with MEA CO$_2$ capture and the ASC case with hybrid membrane/low temperature CO$_2$ capture.
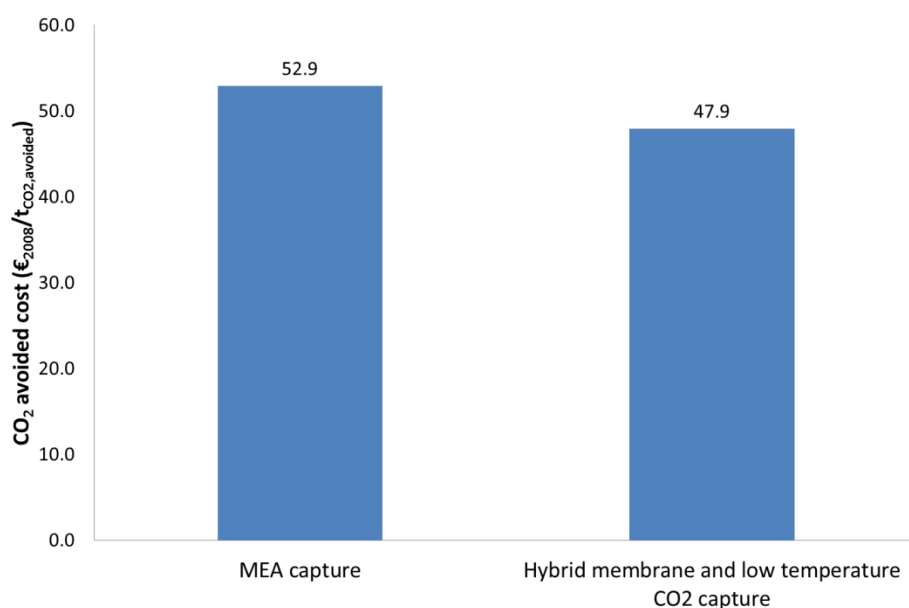
Figure 20: CO$_2$ avoided costs of the MEA and hybrid membrane/low temperature concepts for the ASC plant.

As for the IGCC case, sensitivity analyses were run on the capture CAPEX and OPEX. The results are shown in the three uppermost bars in Figure 21. The "capture investment cost" shows the variation for the overall hybrid membrane/low-temperature system. In addition, the estimated investment cost of the membrane only was varied. The reason for this sub-system variation is that it is often assumed that membranes are costly and add significantly to investment cost. However, as can be seen in Figure 21, the investment cost of the membrane itself is negligible, compared to other parts of the hybrid system, such as compressors. The red bar that shows the cost of electricity of an ASC plant with MEA capture (94.4 €$_{2008}$/MWh, refer to Table 11) is included in order to facilitate the interpretation of the results. As for the

IGCC case in Figure 18, it can be seen that even with an increase in investment cost and O&M cost of 50%, the resulting cost of electricity is lower than for the reference case with 30%wt MEA capture, although the margin is smaller than in the comparison between low-temperature capture and Selexol in the IGCC study.

Also, as for the IGCC case, variations were made of fuel price, operating and maintenance cost, utilization rate, project duration and discount rate to investigate impact on cost of electricity, as shown in Figure 21. It can also be seen in this figure that the cost of electricity is far more sensitive to variations in these parameters than in the capture process investment and O&M cost. As for the IGCC, the sensitivity analysis shows that regarding fuel cost, project duration, discount rate and plant utilization rate can be far more important for the economy of a power plant with CO_2 capture over the plant lifetime. The sensitivity analysis exhibits the same trends as the ASC with 30%wt MEA capture, described by Anantharaman et al. (2011). In this context it should be noted that the CAPEX of the hybrid membrane/low-temperature CO_2 capture system is 11.7% of the total calculated IGCC plant CAPEX, and the OPEX of the low-temperature CO_2 capture system is 5.4% of the total calculated IGCC plant OPEX.
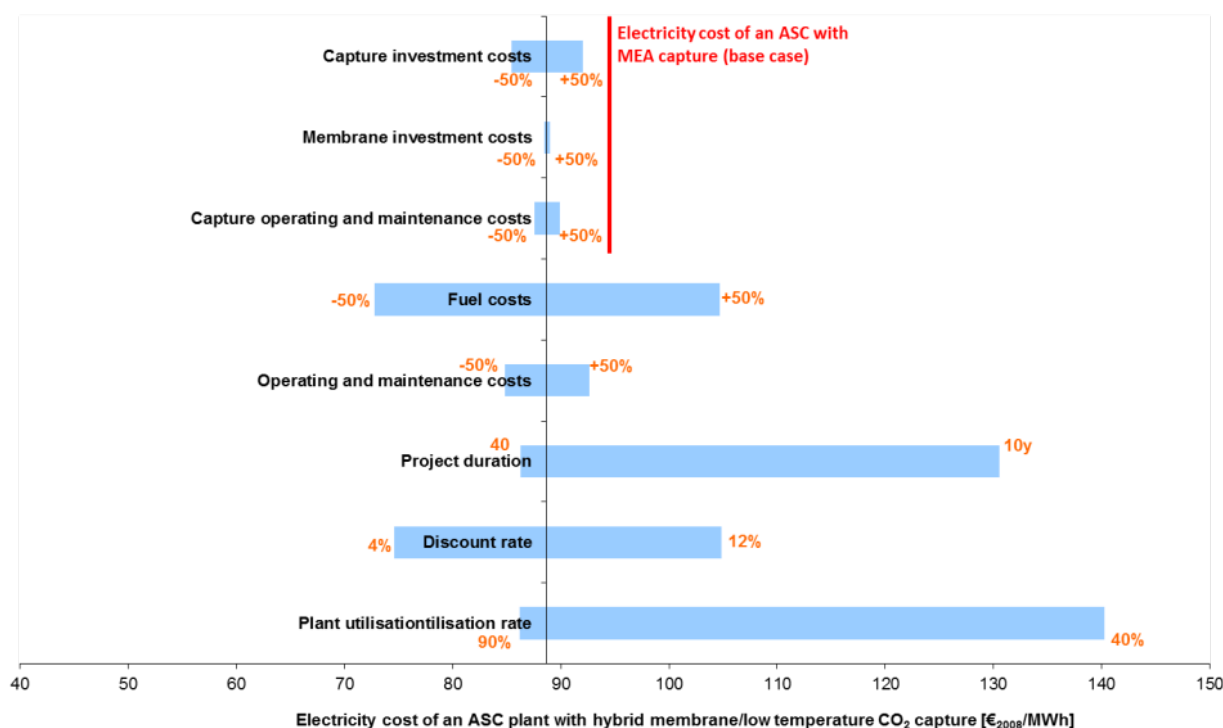

Figure 21: Sensitivity analyses on costs of the ASC plant with hybrid membrane/low temperature capture.

5 Benchmarking of IGCC processes with capture for ship transport of CO$_2$

As explained in section 3.1.5, altogether four different versions of CO$_2$ liquefaction were investigated for the IGCC power plant with CO$_2$ capture. This chapter presents a benchmarking of these four processes. A comparison between ship transport and pipeline transport is beyond the scope of this project, since the main focus has been on the low-temperature capture technology itself, not on value chain analysis.

5.1 Efficiency and CO$_2$ capture rate

Four different options for CO$_2$ liquefaction were investigated, as described in section 3.1.5; two for capture with Selexol and two for low-temperature capture. The reason for this investigation is that since CO$_2$ is captured in the liquid phase in the low-temperature process, it is of interest to investigate if this gives a benefit compared to alternative concepts for ship transport. As can be seen in Table 12, it was found that the liquefaction concept with the highest efficiency is the low-temperature process with internal liquefaction (refer to Figure 9 for process flowsheet).

Table 12. Power output, net electrical efficiency and CO$_2$ capture rate for investigated cases with CO$_2$ capture for ship transport (all cases are capture from IGCC).

	Net power output [MWe]	Net efficiency [%]	CO$_2$ capture rate [%]
IGCC, no CO$_2$ capture	391.5	46.9%	---
Selexol capture with expansion liquefaction	349.4	36.6	81.9
Selexol capture with ammonia cycle liquefaction	357.2	37.4	81.9
Low temperature capture with expansion liquefaction	358.9	37.5	81.4
Low temperature capture with internal liquefaction	368.4	38.5	81.4

It is emphasized that since the CO$_2$ is delivered in different states (dense phase or liquid) a direct comparison of results between Table 9 and Table 12 cannot be made.

5.2 Net present value

In order to be able to calculate the NPV for the power processes with low temperature capture, and compare them against the reference cases from the European Benchmarking Task Force (EBTF), CAPEX and OPEX were first calculated for the processes using the cost evaluation methodology described in Appendix C. Cost basis 2008 is applied to make the calculations directly comparable to the results from the EBTF.

Table 13. Power output, net electrical efficiency and CO$_2$ capture rate for investigated cases with CO$_2$ capture for ship transport (all cases are capture from IGCC).

	CAPEX [M€$_{2008}$]	OPEX [M€$_{2008}$/y]	NPV [M€$_{2008}$]
IGCC, no CO$_2$ capture	926	95.1	1855
Selexol capture with expansion liquefaction	1145	114	2260
Selexol capture with ammonia cycle liquefaction	1155	114	2274
Low temperature capture with expansion liquefaction	1098	109	2165
Low temperature capture with internal liquefaction	1081	107	2178

5.3 Levelized Cost of Electricity (LCOE) and CO$_2$ avoided cost

5.3.1 Cost summary table

Based on the electricity output and the plant costs, the cost of electricity (LCOE) of the IGCC power plant with CO$_2$ capture processes including liquefaction are shown in Table 14.

Table 14: LCOE and CO$_2$ avoided cost of the IGCC power plant with CO$_2$ capture for ship transport.

	LCOE [€$_{2008}$/MWh]	CO$_2$ avoided cost [€$_{2008}$/t$_{CO2\ avoided}$]
IGCC, no CO$_2$ capture	64.6	---
Selexol capture with expansion liquefaction	90.7	40.3
Selexol capture with ammonia cycle liquefaction	89.2	38.0
Low temperature capture with expansion liquefaction	82.7	28.0
Low temperature capture with internal liquefaction	79.3	22.7

5.3.2 Analysis of results for CO$_2$ liquefaction

The evaluation shows that:

- For the low temperature process toward shipping, the process with internal liquefaction seems to be much more cost-efficient than the liquefaction by expansion.
- The internal liquefaction of the low temperature concept is also more cost-efficient than Selexol capture in combination with ammonia liquefaction[2]. This is especially interesting, as the liquefaction with the ammonia cycle in earlier reports has been considered to be one of the most energy and cost-efficient processes to liquefy CO$_2$.
- In the case of a capture process and conditioning oriented towards ship transport, the low temperature process, with a cost of 22.7 €$_{2008}$/t$_{CO2,avoided}$, is 40% more cost-efficient than the Selexol

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[2] Based on the selexol process.

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process (38 €$_{2008}$/t$_{CO2,avoided}$). This is due to the already existing cost advantage of the process in the case of a pipeline export and the extremely cost efficient low temperature internal liquefaction for shipping transport.

- The electricity (LCOE) and CO$_2$ avoided costs for two alternative capture technologies, Selexol and low temperature is higher with ship transport than for pipeline transport. However, it is important to not compare these costs directly as the cost for the export systems are not included. Except for small capacities and distances, the shipping export system can be expected to be cheaper than the pipeline export, which will make the shipping chain more interesting for certain conditions.

Altogether, there appears to be a significant potential for low temperature capture technology designed for ship transport. It is noted by the authors that the investigations done so far are for an IGCC power plant, and that the probability of one or several IGCC plants with CO$_2$ capture being built in a coastal area is currently not overwhelming, However, the benefits with liquefaction through low-temperature CO$_2$ capture should be quite similar for CO$_2$ capture from other sources where the low-temperature capture technology is of interest i.e. the CO$_2$ concentration from the source is sufficiently high or can be increased with membranes.

The results for CO$_2$ avoided are visualized in Figure 22 for the different CO$_2$ liquefaction concepts. The blue bars show the CO$_2$ avoided cost when capturing CO$_2$ for pipeline transport (refer to Figure 17), and the red field at the top of the bars show the additional cost with liquefying the CO$_2$.
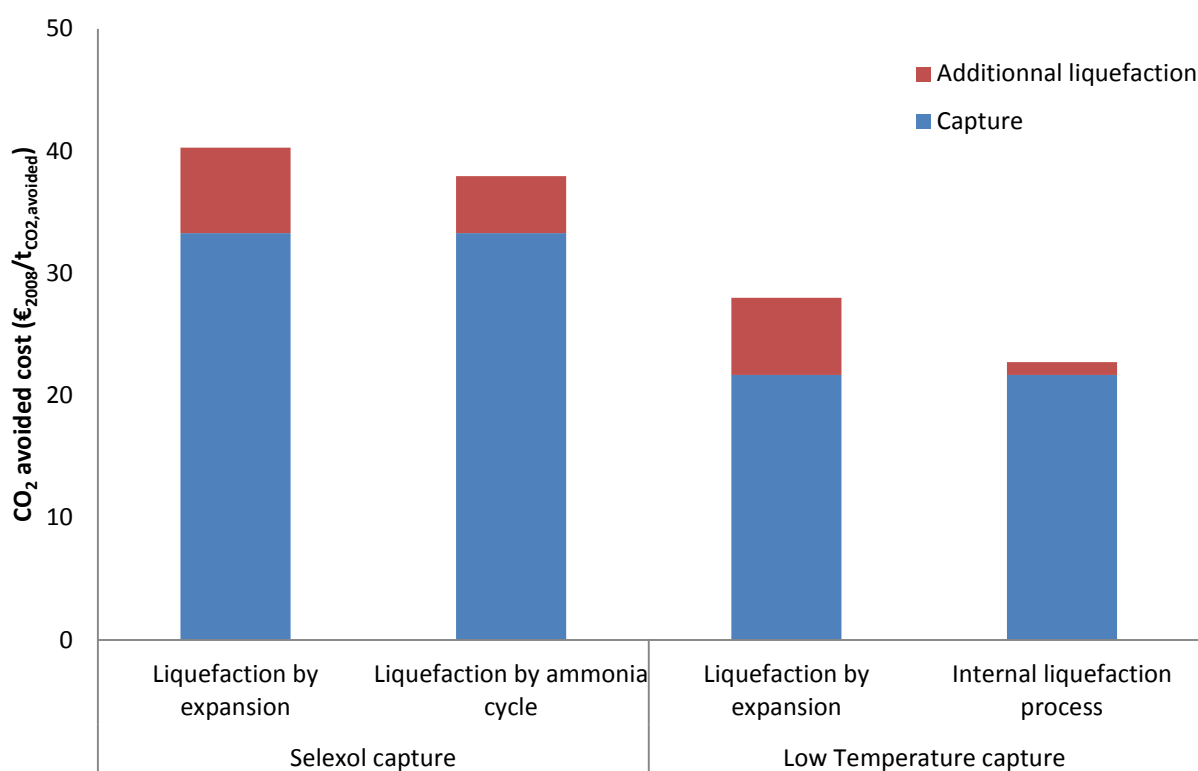
Figure 22: CO$_2$ avoided costs for IGCC with Selexol and low temperature capture for pipeline (blue bars) and ship transport (additional liquefaction cost in red bars).

Sensitivity analyses were run on capture process CAPEX and OPEX. The results are shown in the two uppermost bars in Figure 23. To facilitate the interpretation of the results, the red line gives a reference for the alternative with CO$_2$ capture using Selexol, followed by ammonia cycle liquefaction (electricity cost 89.2

$€_{2008}$/MWh in Table 14). It can be seen that even with a 50% increase in CAPEX or OPEX, the electricity cost for CO_2 capture with internal liquefaction for ship transport clearly results in a lower cost of electricity.

As in Figure 18 and Figure 21 a sensitivity analysis was made for fuel price, IGCC plant operating and maintenance cost, utilization rate, project duration, and discount rate and their impact on cost of electricity. The results are shown in Figure 23, and it can as in the other two sensitivity analyses be seen that relative variation of these parameters have a far more significant impact on the overall cost of electricity of the IGCC plant with CO_2 liquefaction. The EBTF cost analysis contains no investigation of CO_2 liquefaction, but as can be seen when comparing Figure 18 and Figure 23, the results obtained for the variation of these parameters are quite similar. Still, it should be noted that it is not correct to do a direct comparison of the ship and pipeline transport costs given in this report – a correct direct comparison must take the overall CO_2 chain into account, all the way to the CO_2 injection. A study of the value chain is however far beyond the scope of this project report.
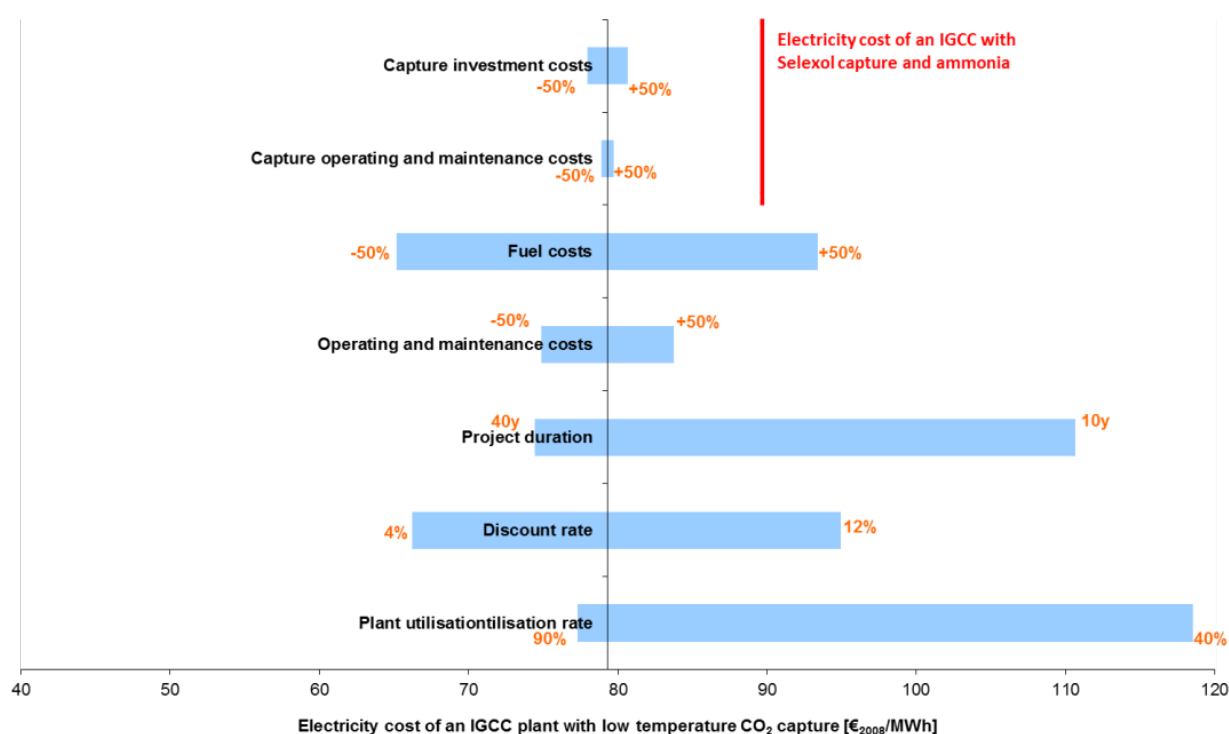

Figure 23: Sensitivity analyses on costs of the IGCC plant with low temperature capture (internal liquefaction) for ship transport.

6 Hydrogen production from coal gasification

6.1 Concept outline

The use of hydrogen as an energy carrier for e.g. automotive transport is anticipated to increase in the future, in Japan, Europe and many other countries, and hydrogen may also find additional users for instance in industries as a substitute for fossil fuels. In the IEA 2 degree scenario – High H_2 (IEA, 2012), a significant increase in the use of hydrogen is projected, with an increase from current annual use of 6 EJ to nearly 30 EJ in 2050.

It has therefore been judged to be of interest to evaluate a novel concept for production of high-purity hydrogen from coal in combination with low-temperature CO_2 capture. The concept is based on the EBTF gasification island presented by Franco et al. (2011). The same unit as was employed to generate the syngas for the low-temperature separation cases in the IGCC studies presented above. It has become clear as the hydrogen production concept has evolved that the pressures for steam generation (cooling of gasification product and shift reactors) could be selected differently in this concept than for an IGCC, where the purpose is to generate maximum power.

In the proposed concept pure hydrogen is separated from a shifted syngas through a H_2-selective Pd or Pd-alloy membrane, and the CO_2-rich retentate is sent to dehydration and subsequently to the low-temperature CO_2 separation unit. One tube in the membrane separation module is shown in Figure 24 – in practice the membrane unit for an industrial scale process would consist of several thousand tubes of several meters length and a diameter of a few centimetres. Since high-purity hydrogen is required for e.g. fuel cells or hydrogen liquefaction, no sweep-gas can be applied to reduce the permeate-side partial hydrogen pressure in the tubes. It is not practical or even possible to separate 100% of the hydrogen through the membrane, meaning that there will always be some combustible gases in the retentate. Moreover, there will be a few percentages of N_2 and Ar, due to the feeding of coal to the gasifier. Altogether, the CO_2-enriched retentate in Figure 24 must be further treated before the CO_2 is sufficiently pure for transport and storage. Due to the presence of N_2 and Ar, oxy-combustion of the stream is not possible since it would yield a CO_2-rich stream of insufficient purity. Hence, low-temperature separation should be the preferable CO_2 capture option. The retentate has a higher CO_2 concentration compared to the IGCC cases presented in section 3.1 above, meaning that the required work to separate CO_2 from volatile components is reduced.

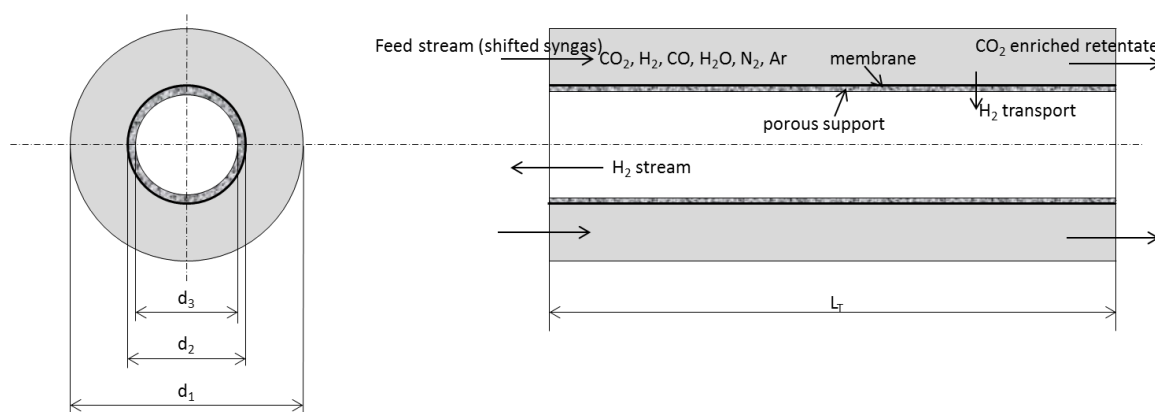**Figure 24. Principle of a H_2 separation unit using a H_2-selective Pd-membrane.**

A possible self-contained concept for high-purity hydrogen production from coal is illustrated in Figure 25 After gasification, cleaning, cooling and sulfur removal, the CO/H_2-rich syngas is sent to a water-gas shift (WGS) reactor. Since the H_2 membrane cannot operate satisfactorily at high CO-concentrations, the

concentration of this gas component must be reduced to 3%, which requires substantial amounts of steam in the WGS reactor. After separation of H_2 through the membrane, the remaining CO_2-rich stream is sent to the low-temperature separation process, which is very similar to the baseline process illustrated in Figure 7. The main difference is that since the CO_2 concentration is higher for the retentate than for the shifted syngas, CO_2 can be condensed and separated at a lower pressure. The H_2-rich fluid that is obtained in the gaseous phase from the low-temperature separation process is envisaged to be employed as low heating-value (LHV) gas turbine fuel in a gas turbine of suitable size. The water gas shift reactor, as well as the gasification process will require significant amounts of steam. A heat recovery steam generator is added to recover heat from the gas turbine exhaust and produce the required amounts of steam, and possibly also additional power. Also the heat available at 400°C in the CO_2-rich retentate upstream of the low-temperature separation process is possible to employ for steam generation. Furthermore, cooling of the stream leaving the gasifer and the water-gas shift reactors can be employed for generating steam from saturated, pressurized water.
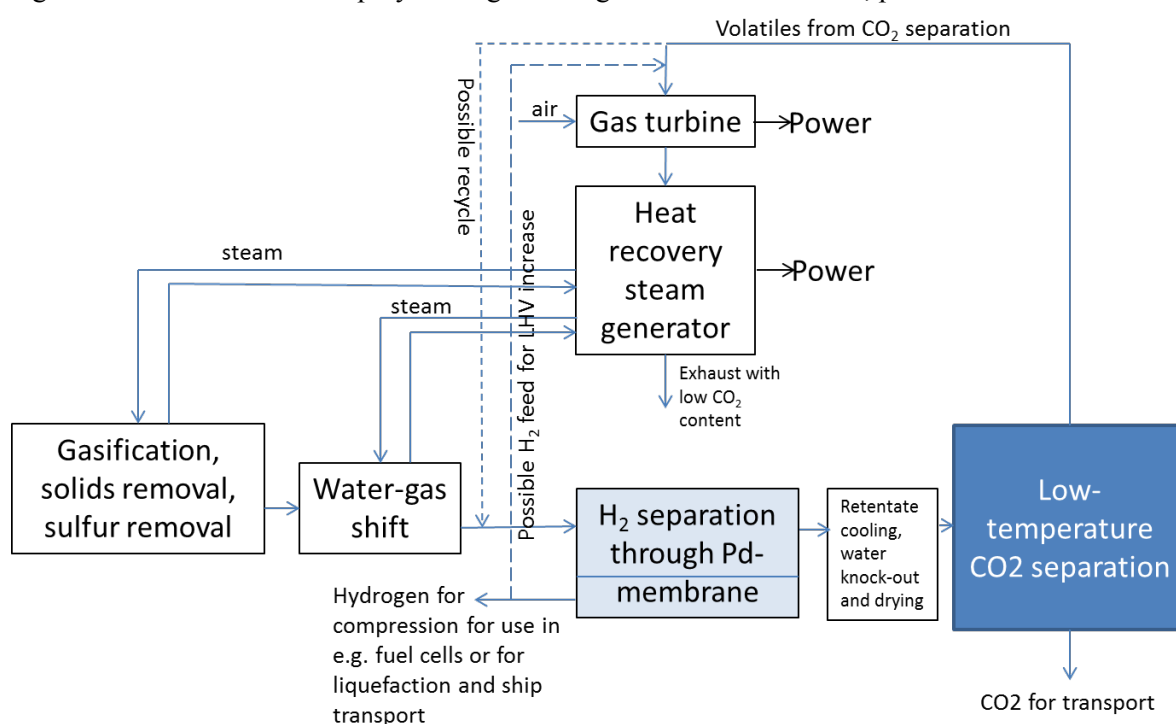
Figure 25. Conceptual process block diagram for H_2 production through H_2-separating membranes in combination with low-temperature CO_2 separation.

6.2 Performance data

For the first overall process simulation reported here, the intention is to close the heat and mass balances. It was assumed that 75% of the hydrogen available after the water-gas shift reactor is separated through the Pd membrane. The ability of the membrane unit to separate this amount of hydrogen will depend on variables like membrane tube length, membrane thickness, flow rate through the membrane tubes et cetera. The requirements that must be put on the membrane and membrane unit, to be able to separate this amount of hydrogen, is a topic for further investigations.

The retentate composition after drying can be seen in Table 15.

Table 15. Retentate composition after drying and prior to low-temperature separation of CO₂

Component	Hydrogen	CO₂	CO	N₂	Ar	CH₄
Mol fraction	0.2155	0.5980	0.0755	0.0962	0.0146	0.0002

A gas turbine model was set up in Aspen HYSYS to reflect a typical gas turbine that has the ability to burn syngas and/or low-heating value fuels. The choice was made to set up a model that is similar to the Siemens SGT5-2000E (former V94.2, which was employed in the Buggenum IGCC plant). Data employed as basis for the gas turbine model were retreived from the 2006 Turbomachinery Handbook. However, the total available fuel energy in the retentate after low temperature separation is roughly half of the fuel required for one SGT5-2000E, which means that one gas turbine unit would be suitable for roughly two gasifiers and twice the envisaged H_2 production in the investigated concept. A summary of the performance of the concept is given in Table 16 below. It can be seen that more than half of the coal thermal energy is transferred to the H_2 product, and that the resulting concept has an electric efficiency of 11.2%.

Table 16. Performance data for concept with H₂ production from coal.

Coal flow rate	155.1	t/h
Coal LHV	25.17	MJ/kg
Thermal energy of fuel	**1084**	**MWth**
Thermal energy for coal drying	**9**	**MWth**
H₂ product	4.69	kg/s
LHV H₂	119.96	MJ/kg
Thermal energy of H₂ product	**563**	**MWth**
H₂ production thermal efficiency	**51.9**	**%**
Gas turbine output	100.8	MWe
Steam turbine output	90.5	MWe
Gross power output	**191.3**	**MWe**
ASU power consumption	27.7	MWe
O₂ compression	13.2	MWe
N2 to gasifier compression	5.8	MWe
Selexol H₂S removal	0.4	MWe
CO₂ capture	16.35	MWe
Power Island aux.	2.6	MWe
Coal Handling	1.6	MWe
Other	0.6	MWe
Total ancillary power consumption	**68.2**	**MWe**
Net power output	**123.0**	**MWe**
Electric efficiency	**11.2**	**%**
Overall first law efficiency	**62.7**	**%**
CO₂ captured	85.2	kg/s
CO₂ emitted	21.4	kg/s
CCR	**79.9**	**%**

Further work on this concept can go in several directions, it is clear that one of the things that must be done is to establish a reference case for H_2 production with CO_2 capture using most mature reference technologies (Pressure-Swing Adsorption and Selexol). Experiences should be possible to draw from the EU FP6 project Dynamis, which was coordinated by SINTEF Energy Research. However, the prospects of this concept seem to be very interesting.

6.3 Hydrogen liquefaction energy consumption

The very pure H_2 that can be obtained through Pd membrane separation is a good starting point for hydrogen liquefaction. The hydrogen that is available at 1 bar and 400°C would then first have to be cooled before it can be compressed and liquefied. Cooling the separated hydrogen down to 30°C corresponds to recovering 25.6 MWth that can be employed in the steam bottoming cycle. Roughly it can be estimated that this would yield an additional power production of 10 MWe, i.e. the total power production would be around 133 MWe.

Hydrogen liquefaction has been studied recently in the EU FP7 FCHJU research project IDEALHY. State-of-the art technology for H_2 liquefaction was established, and also future more advanced technologies were investigated. Based on knowledge about existing H_2 liquefaction plants, reported in the IDEALHY project, result in power consumption for hydrogen liquefaction of 228 MWe for the 4.69 kg/s of hydrogen that is produced. With advanced, future technology, developed in IDEALHY, calculations give a power consumption of 144 MWe for 4.69 kg/s of hydrogen. Altogether, this shows that with conventional technology, even when taking into account the additional power generated from heat recovered through hydrogen cooling, there is an electric power deficit of ~95 MWe. When considering advanced hydrogen liquefaction technology, there is a power deficit of only ~11 MWe, which is quite close to a self-sustained hydrogen production and liquefaction process with CO_2 capture.

Altogether it can be concluded that many pieces must match to make the concept a viable hydrogen production technology with CO_2 capture. It is clear that for a self-sustained process, i.e. a process that produces the power it consumes, more hydrogen can be produced if a more advanced liquefaction technology is applied. With current liquefaction technology, the hydrogen separation rate must be reduced, so that less hydrogen is liquefied and more hydrogen is burnt in the gas turbine. This will also lead to a higher power consumption in the low-temperature separation unit, since the CO_2 concentration of the stream entering the unit will be reduced.

7 Recommendations for future work

This report shows that there is a significant potential for low-temperature CO_2 separation for coal-based applications. In particular the concept with CO_2 separation from the IGCC process has been investigated in-depth. It has also been found that there is an interesting possibility for capturing CO_2 in a state ready for ship transport, both in terms of economics and in terms of flexibility for conversion to pipeline transport later on. A sensitivity analysis on the results for the low-temperature capture process for pipeline transport is provided in section 7.1.

The low-temperature separation concept is not yet fully developed into a commercial technology, as described in this chapter. Additional research is required for technology development and demonstration as described in section 7.2.

7.1 Sensitivity analysis – parameters influencing capture process efficiency

In order to find the main uncertainties in the IGCC application and find the directions for further work on the low-temperature process, a sensitivity analysis was conducted with respect to the specific power consumption for the IGCC syngas separation unit. The sensitivity analysis was conducted early in the project, to also give directions for the research described in this report, and is therefore based on an earlier version of the low-temperature separation process and not the reference process illustrated in Figure 7. This means that the specific power consumption is slightly higher in the reference point (281 kJ/kg) than it is for the improved process (275 kJ/kg). Nevertheless, the response to the varied parameters should be very similar.

A graphical display of results from the sensitivity analysis is shown in Figure 26. As can be observed, the units varied along the horizontal axis are diverse and cannot be directly compared with respect to scale. Hence, the slope of each line plotted in the diagram does not give an unequivocal impression of its sensitivity relative to any other parameter. Due to several extreme differences in slope, however, the investigated parameters can still be divided into different categories of high and low sensitivity.
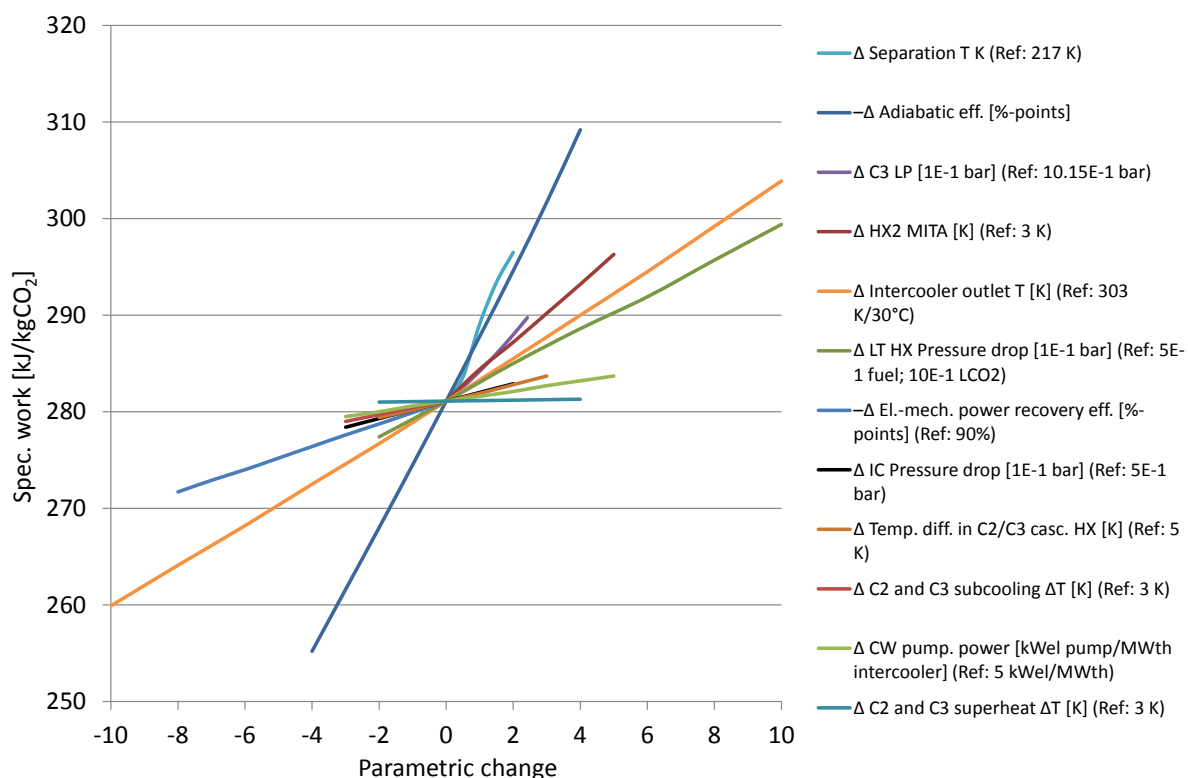
Figure 26. Low temperature process specific work sensitivity for various for process parameters

The two most sensitive parameters pointed out in the analysis are the adiabatic efficiencies for compressors and expanders, as well as the minimum allowed syngas separation temperature. This finding strengthens the already established view of the high importance of verifying efficiencies of turbomachinery components and conducting thermodynamic simulations and experiments to find the CO_2 freeze-out temperature for the syngas compositions under consideration. It is emphasized that the assumed turbomachinery efficiencies are believed to be conservative in the conducted simulations.

The pressure drops in heat exchangers are rather sensitive parameters. The same applies to the intercooler outlet temperature, which is governed by mainly ambient temperature and ambient cooling fluid – water or air. The two other parameters also considered to be rather sensitive are the minimum temperature approach in heat exchangers and the lowest allowed pressure level in the propane refrigeration cycle. The reference values for these parameters, however, have been chosen conservatively. When going in-depth on heat exchanger design in the project, the minimum temperature approaches in the detailed HX2a designs (made in Aspen MULE) turned out to be considerably lower than what has been assumed in the process models (3°C). Also, the pressure drops in the different streams of HX2a were found to be below 1 bar.

Among less important parameters are those concerning superheating, sub-cooling and cascade temperature differences in the propane–ethane auxiliary refrigeration cycles. Another group of insensitive parameters concern inter-coolers, as neither pressure losses nor cooling water pumping power have significant influence on the specific power consumption. The power recovery efficiency from fuel expanders, assumed to be 90% in the baseline design, is another parameter with relatively low sensitivity to the overall efficiency

7.2 Additional research required for low temperature capture technology development

7.2.1 Vapour/liquid equilibrium of H_2/CO_2 mixtures

In this project, the Aspen HYSYS process simulator has been applied with a cubic equation of state (EOS), namely Peng-Robinson. This type of EOS are fast and robust, but the process simulation results may deviate from the actual conditions when there are large differences in molecular weight in the fluid components such as for CO_2/H_2 mixtures. This means that the process simulations for the syngas separation case and the retentate processing case have been conducted as well as possible, but not with full confidence in the thermodynamic calculations. The results obtained with Peng-Robinson EOS are in accordance with older experimental data (Spano et al, 1968), but do not match with the more recent experimental data from Tsang and Streett (1981).
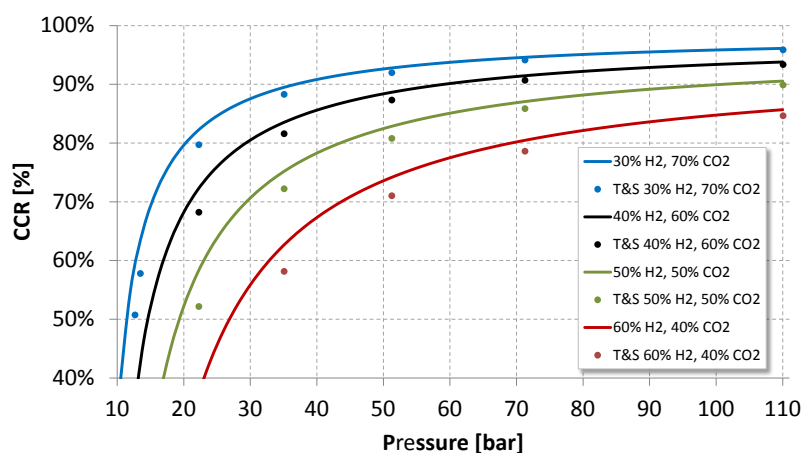

Figure 27. Results with Peng-Robinson for CO_2–H_2 phase separation, compared with experimental results from Tsang and Streett (1981).

Use of an appropriate mixing rule with all necessary binary interaction parameters regressed against either high-quality experimental values or values calculated by a reference equation of state can give a high accuracy cubic EOS for a region of interest. A very accurate reference EOS for pure CO_2 exists (Span and Wagner, 1996) and an accurate pure H_2 reference EOS is described by Leachman et al. (2009). Gas/liquid mixtures of CO_2 and H_2 can be described with the GERG 2008 EOS (Kunz and Wagner, 2012), but current limited availability of high-quality VLE experimental data reduces the accuracy of this model. Ongoing experimental and modelling work on CO_2/H_2 mixtures primarily has focus on compositions and conditions relevant for CO_2 transport (Sanchez-Vicente et al., 2013), which differ from those relevant for low-temperature capture. It can be seen in Figure 27 that the largest deviations between Peng-Robinson and experimental data from Tsang and Streett (1981) is at low pressures, whereas the main part of the CO_2 capture is obtained through the separation at high pressure. Hence, it can be presumed that the CO_2 capture rate is fairly well predicted with Peng-Robinson for the base case predictions, unless new experimental data differ significantly from Tsang and Street (1981) and Spano et al, (1968). More noticeable deviations from calculated results with Peng-Robinson may be found for hydrogen recovery and CO_2 purity. Still, it is expected that both hydrogen recovery and CO_2 purity will be high, also if novel experimental data were produced and taken into account in the simulations.

Experimental data for CO_2/H_2 VLE could be possible to obtain by experiments in e.g. the CO2MIX test rig, owned by SINTEF Energy research in Trondheim. The rig is designed to operate at a pressure range of 1 – 200 bar and a temperature range of -60 – 150 °C. Temperature accuracy and stability should be less than 10 mK, while pressure measurement stability should be less than 0.1 % (Stang et al., 2013).

In the post-combustion application of low-temperature separation, the difference in molecular weight between the species (N_2/O_2/CO_2) is smaller, and confidence is therefore sufficiently high in the thermodynamic results obtained with Aspen HYSYS.

7.2.2 Process simulations with improved thermodynamic and process unit models

It is not sufficient only to establish high-accuracy VLE thermodynamic data for CO_2/H_2 mixtures at relevant pressures, temperatures and composition mixtures, as described in section 7.2.1. These data must also be applied in improved thermodynamic models that are applicable in process simulations, i.e. thermodynamic models must be established that are robust, reliable and have a higher accuracy that what is presumed to be obtainable with the current HYSYS models. In this work, it should be possible to recur also to detailed, computationally demanding thermodynamic models.

Once the improved thermodynamic models have been established, the process simulations for the IGCC cases and the H_2 production concept should be repeated with improved accuracy.

In addition, process simulations for other relevant capture scenarios should be performed, focusing on e.g. H_2 production with CO_2 capture from reforming of natural gas. Also in these process simulations detailed heat exchanger modelling good models for compressors and expanders must be included, to ensure the best possible accuracy.

Although confidence is high in the thermodynamics for CO_2/N_2/O_2 mixtures, it is still of interest to perform improved process analysis of the hybrid post-combustion capture concept described in section 3.2. In addition to the possibility to provide a retrofit option for coal power plants, it should also be of interest to investigate CO_2 capture from cement and iron/steel production.

Post-combustion capture from natural gas has a low CO_2 concentration and is perhaps not the most obvious process for investigating low-temperature CO_2 separation. Nevertheless, it could be of interest to investigate the hybrid membrane-low temperature capture process and establish the requirements on CO_2 membrane selectivity and permeability to make low-temperature separation an alternative. In particular, it could be of interest to investigate the NGCC with exhaust gas recirculation, which brings up the CO_2 concentration from ~ 4% to ~ 8%.

7.2.3 Experimental setup for main phase separator and heat exchanger

VLE measurements will give the equilibrium data for H_2/CO_2 mixtures that can be obtained after a long residence time in a large volume separator. In practice, however, gas separators must be designed that are of a reasonable scale, i.e. gives sufficiently high vapour/liquid separation at a sufficiently low process unit cost.
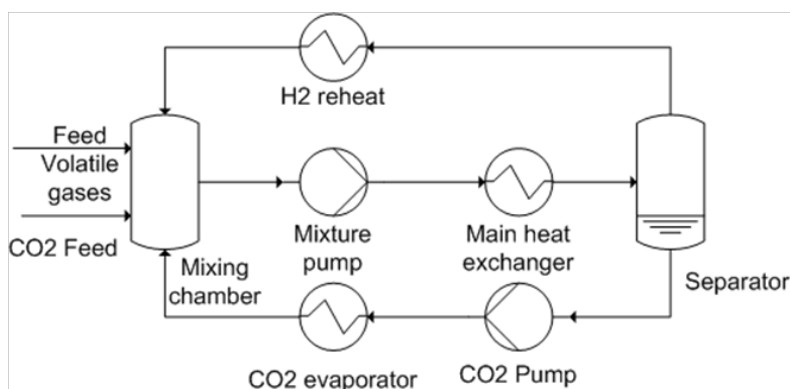
Figure 28. Sketch of experimental setup principle for investigating heat exchanger and phase separator behaviour.

Hence, in order to establish knowledge about the necessary residence time and give guidelines for the design of the gas-liquid separator, an experimental setup shown in principle in Figure 28 is proposed. This experimental setup can enable studies of the kinetics of gas-liquid separation, i.e. of how fast and well CO_2 (condensed in the "main heat exchanger") is separated from hydrogen.

It is also of interest to measure how the presence of hydrogen affects the CO_2 freezing point (so-called freezing point depression). The principle to investigate is the same as when salt (e.g. NaCl) is mixed with water to reduce the freezing point of water. If hydrogen can reduce the CO_2 freezing point in the "main heat exchanger", this could enable the walls at the heat exchanger outlet to operate at e.g. -60°C (i.e. below the CO_2 freezing point), without CO_2 forming a solid phase on the walls. The possibility to operate with heat exchanger walls below -56°C would enable design of a more compact and cheaper heat exchanger.

The suggested experimental setup would also enable investigations of heat exchanger and phase separator behaviour at part load operation, start-up and load changes. For instance, it is of relevance to investigate if/how much the heat exchanger outlet temperature must be increased during part load operation to avoid CO_2 freeze-out. It may also be possible to investigate heat exchanger dynamics and change of heat transfer coefficient for condensation of CO_2 from a CO_2/H_2 mixture.

If experimental work in a rig of the kind that is sketched in Figure 28 is successful, the acquired knowledge will be possible to apply in a subsequent process development step, which could be to design a pilot-scale experimental facility that could be of relevance to test at e.g. Test Centre Mongstad.

7.2.4 Investment cost and value chain analysis

Investment costs determined in this project for the IGCC application of low temperature separation are very favourable. It could be of interest to re-evaluate these costs with help of external cost assessment expertise. It should also be of interest to do a cost optimization and evaluate trade-offs between simplified process configurations that may lead to reduced efficiency and/or CO_2 capture rate on one hand, and reduced investment cost on the other. Due to the interesting perspectives for flexible build-up of a CCS infrastructure in e.g. the North Sea area, the full value chain perspective for low-temperature capture of CO_2 should be assessed. Such value chain assessment could take into account e.g. initial ship transport of CO_2 at an early stage of CCS deployment, followed by subsequent conversion to pipeline transport, and if and when such transition is attractive. It could also be of interest to investigate concepts that have been suggested in the NORDICCS project, where infrastructure has been proposed for shipping of liquid CO_2 to an offshore hub, for subsequent pipeline transport to a storage site.

8 Concluding remarks

This project report has described process simulations, economic analysis and benchmarking of the low temperature CO_2 capture technology against reference solvent-based technologies.

The IGCC process with low temperature capture of CO_2 was found to be more efficient and have a significantly lower cost of electricity than the reference technology (capture with Selexol) for both CO_2 for ship transport and for pipeline transport, provided that the capture rate is at around 82%, rather than the more commonly assumed 90% originating from the characteristics for other capture technologies.

For a coal-based steam power plant (ASC process) with post combustion capture, there was no decisive advantage for a hybrid membrane/low temperature concept in terms of efficiency, and only a slight advantage terms of cost of electricity when applying a hybrid post-combustion capture process. This hybrid process consists of a polymeric membrane that increases the CO_2 concentration of the exhaust gas, followed by the low-temperature capture process. In this context, however, it must be observed that the hybrid membrane/low temperature process does not require any process steam to capture CO_2, only electricity. Hence, this hybrid concept is relevant as a retrofit option for steam power plants and should also be of interest to investigate for CO_2-emitting industries such as cement and iron/steel, where access to process steam may be difficult. Furthermore, the hybrid concept with membrane and low temperature separation should be if interest to investigate for natural gas fired power plants, to verify if any benefit could be obtained with this application. In particular, it could be interesting to investigate natural-gas fired combined cycles with Exhaust Gas recirculation (EGR).

In addition, a first investigation was made of a hybrid concept for high-purity H_2 production from gasified coal. The concept utilises a H_2-selective Pd membrane in combination with low-temperature separation of CO_2. The heat and mass balance was closed, with a net power output, generated from an LHV gas turbine. Based on this first investigation, and with the on-going development of Pd-membranes in mind, the concept appears possible to develop into a viable option for hydrogen production with CO_2 capture. When including the required energy consumption for hydrogen liquefaction it is concluded that the concept has potential to be refined into a self-sustained process for H_2 production that also produces the power required for CO_2 capture and H_2 liquefaction. As for the post-combustion capture hybrid, the hydrogen production concept should be of interest to investigate also for natural gas, thus generating H_2 with CO_2 capture from natural gas reforming.

It has become clear during the project execution that one of the key benefits of the low temperature CO_2 capture technology, compared to the reference technologies, is that CO_2 is captured in the liquid phase, while the reference technologies deliver captured CO_2 in the gaseous phase. Capturing CO_2 in the liquid phase provides a possibility for significant cost reductions for ship transport, compared to other liquefaction technologies. Furthermore, if the low-temperature process initially is designed and built for delivering CO_2 for ship transport, it is rather simple to modify the process later on to deliver CO_2 for pipeline transport. This means that transport and storage of captured CO_2 is possible in coastal areas already prior to a pipeline infrastructure is put into operation. This flexibility can enable postponing costly CO_2 infrastructure investments around e.g. the North Sea until a sufficient number of CO_2 suppliers are in place. Obviously, the low-temperature capture process can also be designed and built to deliver CO_2 for pipeline transport also from the beginning.

Altogether, it is concluded that the results obtained are highly interesting, and promising enough to justify further experimental and theoretical investigations of the low temperature CO_2 capture technology with the ultimate objective to realize full-scale low-cost CO_2 capture that allows for flexibility in the choice of transport means (ship or pipeline).

9 List of Acronyms

ASC	Advanced Supercritical pulverized fuel power plant
CCR	CO_2 capture ratio
CCS	CO_2 Capture and Storage
EBTF	European Benchmarking Task Force (Anantharaman et al., 2011, Franco et al., 2011)
EOS	Equation of State
IGCC	Integrated Gasification Combined Cycle
LCOE	Levelized cost of electricity
LNG	Liquefied Natural Gas
NGCC	Natural Gas Combined Cycle
NPV	Net Present Value
VLE	Vapor-Liquid Equilibrium

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A Auxiliary refrigeration

The auxiliary refrigeration cycles used for all simulations of the low-temperature capture separation processes are conventional vapor-compression refrigeration cycles with propane and ethane as refrigerants. This is mature technology with a very long history of industrial application in for instance LNG production, and therefore involves low technological risk. Even though the refrigerants are flammable, safety can be maintained by taking appropriate measures is process design.

As can be observed in Figure 29, the propane and ethane cycles have a cascade connection in which the propane cycle pumps the heat rejected from the ethane cycle and rejects this to the ambient.

With the current configuration and parameters the coefficient of performance (COP) – the ratio between cooling duty and power input – calculate to:
- 2.11 for refrigeration at -42°C (231 K)
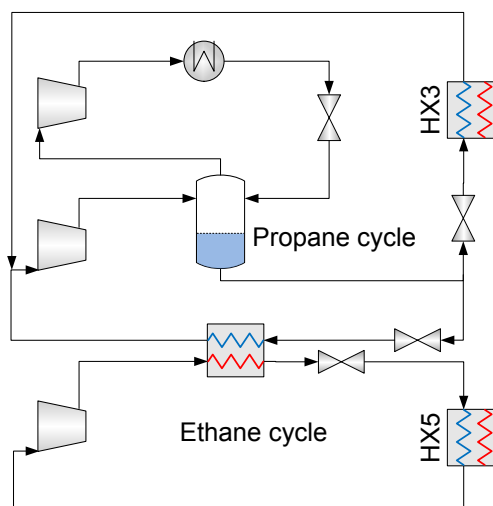- 1.48 for refrigeration at -59°C (214 K)

Figure 29. Auxiliary refrigeration unit with two-stage propane vapour compression cycle in cascade configuration with single-stage ethane vapour compression cycle.

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B Performance data summary, all IGCC cases with CO_2 capture

Gasifier					G1			
Capture process		Selexol	LT	LT improved operability	Selexol+exp. liq.	Selexol + NH3 liq.	LT + exp liq.	LT, internal liq.
Coal flowrate	tonnes/h	136.6	136.7	136.7	136.6	136.6	136.7	136.7
Coal lower heating value (LHV)	MJ/kg	25.2	25.2	25.2	25.2	25.2	25.2	25.2
Thermal Energy of fuel (LHV)	MWth	955.1	955.8	955.8	955.1	955.1	955.8	955.8
Thermal Energy for coal drying	MWth	8.1	8.1	8.1	8.1	8.1	8.1	8.1
Gas turbine output	MWe	283.1	284.3	284.3	283.1	283.1	284.3	284.3
Steam turbine output	MWe	171.6	172.2	172.2	171.6	171.6	172.2	172.2
Air expander	MWe	5.8	5.8	5.8	5.8	5.8	5.8	5.8
GT fuel expander		—	—	—	—	—	—	—
Gross electric power output	MWe	460.5	462.3	462.3	460.5	460.5	462.3	462.3
ASU power consumption	MWe	13.2	13.3	13.3	13.2	13.2	13.3	13.3
O_2 compression	MWe	11.6	11.6	11.6	11.6	11.6	11.6	11.6
N_2 to gasifier compression	MWe	5.1	5.1	5.1	5.1	5.1	5.1	5.1
N_2 to GT compression	MWe	27.4	26.8	26.8	27.4	27.4	26.8	26.8
CO_2 capture	MWe	11.0	22.1	24.0	24.1	16.3	35.2	25.7
Selexol – AGR	MWe	included above	0.3	0.3	included above	included above	0.3	0.3
CO_2 compression	MWe	18.9	—	—	18.9	18.9	---	---
Other	MWe	10.8	11.1	11.1	10.8	10.8	11.1	11.1
Total ancillary power	MWe	98.0	90.3	92.2	111.1	103.3	103.4	93.9
Net electric power output	MWe	362.5	372.0	370.1	349.4	357.2	358.9	368.4
Net electric efficiency	**%**	**37.9**	**38.9**	**38.7**	**36.6**	**37.4**	**37.5**	**38.5**
Overall CO_2 capture rate	%	81.9	81.4	81.4	81.9	81.9	81.4	81.4
Specific CO_2 emissions	kg/MWh	175.1	177.5	178.5	172.3	168.8	179.4	179.2

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C Cost evaluation methodology

C.1 General parameters

C.1.1 Type of costs estimates

In this work, costs corresponding to an "NOAK" (Nth Of A Kind) plant to be built sometime in the future when the technology is mature are calculated. Such estimates reflect the expected benefits of technological learning, but may or may not adequately account for the increased costs that typically occur in the early stages of commercialization (Metz et al., 2005). The project estimates presented in this report are in the feasibility order of magnitude estimates (i.e. ± 40 % of accuracy) as shown in Table 17.

Table 17. Project estimates degree of accuracy (U.S. Department of Energy, 2011)

Type	Purposes	Accuracy
Planning/Feasibility order of Magnitude Estimate (Proposal)	1. Scoping Studies. 2. Preliminary budget estimates of Total Project Cost 3. Support Key Decision 0.	± 40 %
Budget/Conceptual Design Estimate (Equipment Factored)	1. Ensure project feasibility. 2. Develop reliable project cost estimate. 3. Establish baseline project definitions, schedules, and costs. 4. Support Key Decision 1. (Design 10% to 15% Complete)	± 30 %
Title I Estimate	1. Verify that Title I design details still remain within the project funding. 2. Support Key Decision 2. (Design 25% to 35% complete)	± 20 %
Title II or Definitive Estimate (Detailed)	1. Estimate construction costs as accurately as possible, prior to the commencement of competitive bidding and construction activities. 2. Support Key Decision 3. (Design 60% to 100% Complete)	-5% to +15%
Construction Estimate	1. Estimate is based on bid information. (Design 100% complete)	-5% to 10%

C.1.2 Cost Year and currency

For calculations of CAPEX and OPEX, the directions given by the European Benchmarking Task force (EBTF, joint effort from EU projects DECARBIT, CAESAR and CESAR) were applied. All costs estimates are reported in **Euros 2008** which correspond to the reference costs year in the EBTF report (Anantharaman et al., 2011). The reason for applying the EBTF and applying 2008 as the base year is that this makes the results from the present assessment directly comparable to other assessments of CO_2 capture technologies within SINTEF and from other research organisations that also apply the EBTF. In addition, an evaluation of

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the electricity production cost in €$_{2012}$/Mwh is also provided[3]. Investment costs of the capture plant are updated according to the Chemical Engineering Plant Cost Index (CEPCI) (Chemical Engineering, 2011) while investment of the power plant are updated according to the EPCCI without nuclear[4] (IHS, 2013). The utilities costs are corrected considering a yearly inflation of 2% (Trading Economics, 2011). The CEPCI, the EPCCI without nuclear and the EUR/USD (Eurostat, 2010) exchange rate from 2001 to 2012 are presented in Figure 30.

	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012
CEPCI	394	396	402	444	468	500	525	575	522	551	586	593
EPCCI without nuclear	110	118	121	127	133	154.5	173	180	164	169.5	167.5	173
€/$	0.90	0.95	1.13	1.24	1.24	1.26	1.37	1.47	1.39	1.33	1.39	1.29
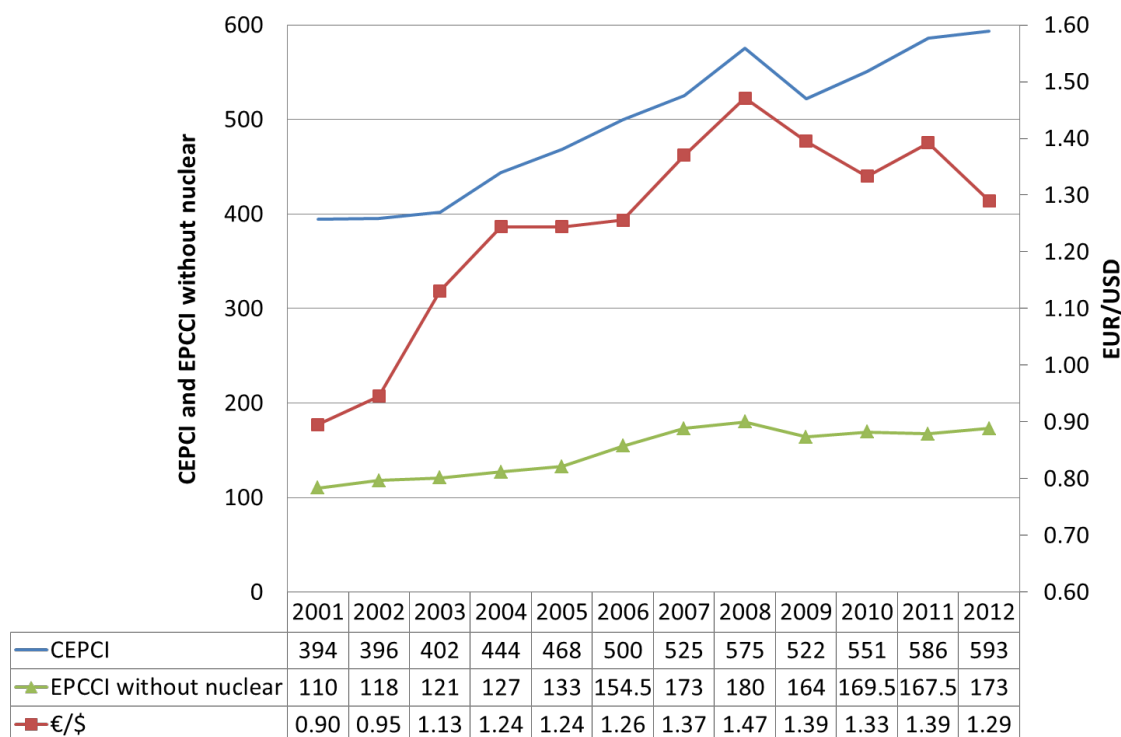
Figure 30. CEPCI, EPCCI without nuclear and Exchange rate EUR/USD Evolution

C.2 Investment costs

Investment costs are given in 2008 prices or reported using the CEPCI Index (Chemical Engineering, 2011). However, in the cash flow profile, the investment costs are reported as an overnight cost occurring at the end of the construction. Investment cost is distributed over the 4-year construction time as shown in Table 18, and then reported in year 0 of the power plant operation, while including the interest rate for the cost over the construction time.

Table 18. Annual allocation of costs for plan construction (Anantharaman et al., 2011).

Year	1	2	3	4
Cost share per year (%)	20	30	30	20

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[3] The CEPCI, EPCCI and exchange rate are not fully available for the current year (2013), the cost were therefore updated in 2012 for which all indicators are available.

[4] The EPCCI track and forecast the costs associated with the construction of a portfolio of power generation plants in Europe, and is as such an indicator of the market price of the power plants.

Different investment costs estimation methods are used: a common method for the CO_2 capture process and a specific one for the power plant.

C.2.1 CO_2 capture process

A factor estimation method is used to estimate the investment costs of a CO_2 capture plant were the estimated equipment costs are multiplied with direct (includes piping/valves, civil work, instrumentation, electrical installations, insulation, paintings, steel structures, erections and outside battery limits) and indirect (includes yard improvement, service facilities, engineering/consultancy cost, building, miscellaneous, owner's costs and contingencies) cost factors to obtain the investment costs. Equipment costs of equipment in the appropriate material[5] are estimated using Aspen Process Economic Analyzer® (Aspen Tech, 2010), based on results from the process simulations in Hysys® that were presented in deliverable D.A. The investment cost for given equipment is then calculated by multiplying the equipment cost with the appropriate direct and indirect cost factors (Table 19). The total investment cost in €$_{2008}$ is then determined by summarizing the estimated investment cost for all components within defined system boundaries. For consistency, the EBTF average direct and indirect cost factors of the IGCC plant with CO_2 capture are used to evaluate the investment cost of the CO_2 capture process in this work (Table 19).

Table 19. Direct and indirect cost factor as function of the equipment cost (Anantharaman et al., 2011).

Direct Cost Factor	1.77
Indirect Cost Factor	1.31

Since they are not available in the Aspen Economic Analyzer database, Plate and Fin Heat Exchangers (PFHE), syngas compressors and syngas expander costs are modelled based on literature figures and using the heat exchange area/power consumption/power production obtained through component sizing. It should be noted that cost data for PFHE (Table 20) are scarce in the open literature, and that the obtained data are rather old[6]. Based on the available costs, the subsequent scaling for the proper process capacity is performed using the cost power law (see under) and installation factors. Where C_0 is the equipment or erected cost of unit with a capacity S_0, while C_1 is the equipment cost of the unit with a capacity S_1. The reference cost and exponential coefficient "n" of the special equipment in the capture process are given in Table 20. The direct and direct costs are then obtained using the cost factor methodology presented above.

$$C_1 = C_0 \cdot \left(\frac{S_1}{S_0}\right)^n$$

Table 20. Reference cost and exponential coefficient of special equipment of the CO_2 capture process

Plant component	Reference parameter	Reference cost	Type of cost	Scale factor	Year
Syngas compressor	Compression work	11.1 M\$	Investment	0.85	2001

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[5] The CO_2 capture process equipment is modelled in Stainless steel due to the low temperature applications, while sea water coolers are in titanium and brazed plate and fin heat exchangers are in aluminium.

[6] For the HX2A in the IGCC case, this lead to a equipment cost and direct cost of respectively 462 and 820 k€$_{2012}$. From vendors contact, the same heat exchanger was estimated to cost between 400 and 450k€ (equipment cost), indicating that the value available in the literature and used for costing of heat exchanger in the report is within a realistic range.

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(Hamelinck and Faaij, 2002)					
Syngas expander[7] (Hamelinck and Faaij, 2002)	Expansion work	11.1 M$	Investment	0.85	2001
Brazed plate & fin Heat Exchanger (Lunsford, 1996)	Cold side area	8 $/ft^3	Investment	1	1996

C.2.2 Power plant process

Due to their specificity, the power plant equipment is modelled based on the EBTF report (Anantharaman et al., 2011). As previously, based on the available costs, the subsequent scaling for the proper process capacity is performed using the cost power law for equipment costs and installation factors. The reference cost and exponential coefficient "n" of the special equipment in the power plant and capture process are given in Table 21 to Table 23.

Table 21. Reference cost and exponential coefficient of the IGCC power plant equipments (Anantharaman et al., 2011)

Plant component	Reference gross electric power output [MWe]	Reference equipment cost [M€]	Reference direct cost [M€]	Scale factor	Year
Coal handling	457.2	24	53.9	0.67	2008
Gasifier	457.2	108	180	0.67	2008
WGS reactor	457.2	13.2	21.1	0.67	2008
Ash handling	457.2	7.8	17.4	0.67	2008
Acid gas removal	457.2	12	20.7	0.67	2008
Gas cleaning	457.2	4.3	6.9	0.67	2008
Claus burner	457.2	8	12.8	0.67	2008
Gas turbine	457.2	51	93.3	1	2008
Steam turbine	457.2	32	52	0.67	2008
Heat recovery steam generator	457.2	15.5	34.1	0.67	2008
Low temperature heat recovery	457.2	5.3	10.9	0.67	2008
Cooling	457.2	15	39	0.67	2008
Water treatment	457.2	13.2	21	0.67	2008
Air separation unit	457.2	45.5	72.8	0.67	2008

Table 22. Reference cost and exponential coefficient of the ASC power plant (Anantharaman et al., 2011)

Plant	Reference net power output [MWe]	Investment [M€]	Scale factor	Year
ASC power plant	754.3	1456	-	2008

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[7] Considered to have equivalent cost to a syngas compressor due to their similarity.

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Table 23. Reference cost and exponential coefficient of membrane system equipments (Zhai and Rubin, 2013)

Equipment	Direct cost	Scale factor	Year
Membrane reactor	50 $/m^2	1	2011
Vaccum pump	1000 $/kW	1	2011

C.3 Operating costs

The operating costs are split into fixed and variable operating costs.

- Fixed operating costs

The fixed operating cost depends on the investment cost and covers maintenance, insurance, and labour costs. The annual fixed operating cost is set to 3.6%[8] of total direct costs (equipment and direct costs[9]) of process units (Anantharaman et al., 2011). The fixed operating costs of the ASC power plant are based on the cost of the reference plant without capture (Anantharaman et al., 2011) as shown in Table 25.

- Variable operating costs

The variable operating cost, being a function of the amount of electricity produced, covers consumption of utilities: coal, ash disposal, make up water, process water and sea water cooling. The electricity consumed is considered to be auto-consumption in the power plant. The annual variable operating costs are estimated using the utilities consumptions given by the simulation results and utility costs given in Table 24. The utilities costs which are not in the proper year prices are corrected considering a yearly inflation of 2% (Trading Economics, 2011).

Table 24. Utility costs

Utilities	Reference costs	Cost Units	Reference year
Coal (Anantharaman et al., 2011)	3	€/GJ	2008
Ash disposal (Anantharaman et al., 2011)	0	€/t	2008
Make-up water (Anantharaman et al., 2011)	0.35	€/m^3	2008
Process water (Anantharaman et al., 2011)	6	€/m^3	2008
Sea water cooling (Fimbres Weihs and Wiley, 2012)	0.03[10]	€/m^3	2008

---

[8] To be consistent with the EBTF evaluation, this value corresponds to the average value for the IGCC case with capture in the EBTF report.

[9] Includes piping/valves, civil work, instrumentation, electrical installations, insulation, paintings, steel structures, erections and outside battery limits.

[10] Corresponding to 0.225 NOK_{2009}/m^3 converted to $\$_{2012}$ using the $NOK_{2009}/€_{2009}$ exchange rate of 0.114 and an annual inflation rate of 2%.

As the IGCC process is not fully designed in our simulation, the utilities consumption of the IGCC parts are evaluated considering the reference IGCC without capture (Anantharaman et al., 2011), shown in Table 25, and scale up based on the thermal energy of fuel required (955.8 MW_{th}) in the IGCC case with low temperature CO_2 capture.

As a principal purpose of this study is to calculate the cost of capturing CO_2 with the low-temperature capture process and compare this technology against reference technologies, no CO_2 tax is considered. It is assumed that sulphur can be sold for a price of 80€/$t_{sulphur}$.

Table 25. Consumables and costs of the reference IGCC power plant without capture (Anantharaman et al., 2011).

Utilities	Consumable [t/y]	Cost [€$_{2008}$/y]
Make up cooling water	3,300,000	1,155,00
Process water	420,000	2,520,000
Selexol sulphur removal	20	100,000
Catalysts	-	180,000
Ash	125,000	0
Miscellaneous	-	1,800,000
Sulphur credit	4,375	-350,000
Fuel cost	-	67,200,000

The operating costs of the ASC power plant (without capture) are evaluated using the EBTF report (Anantharaman et al., 2011) as shown in Table 26.

Table 26. Consumables and costs of the reference ASC power plant without capture (Anantharaman et al., 2011)

Operating costs	Cost [M€$_{2008}$/y]
Fixed operating and maintenance cost	27
Fuel	133
Variable operating costs	9.2

D Low Temperature syngas separation and CO$_2$ capture from IGCC

By David Berstad, Petter Nekså, Rahul Anantharaman, Geir Skaugen (SINTEF Energy Research)

Presented by David Berstad during the 7th Trondheim Conference on CO$_2$ capture, Transport and Storage (TCCS7), Trondheim, June5-6, 2013.

Presentation available on https://conference.preseria.com/?session=g2yf0bvbslc35oe&stamp=204#

E Assessment of Hybrid Membrane – Low Temperature Process for Post-Combustion Capture

By X. P. Zhang, T. Gundersen (NTNU) R. Anantharaman, D. Berstad (SINTEF Energy Research)
Presented by Rahul Anantharaman during the 7th Trondheim Conference on CO_2 capture, Transport and Storage (TCCS7), Trondheim, June5-6, 2013.
Presentation available on https://conference.preseria.com/?session=14qqen6k211bsje&stamp=342#

F Integrated Low-Temperature CO$_2$ Capture from IGCC Power Plant by Partial Condensation and Separation of Syngas

By David Berstad, Petter Nekså, Rahul Anantharaman, Geir Skaugen (SINTEF Energy Research)
Presented by David Berstad during the 16th Conference on Process Integration, Modelling and Optimisation for Energy Saving and Pollution Reduction (PRES'13), Rhodes, Greece, 2013.
Presentation on subsequent pages

Integrated Low-Temperature CO$_2$ Capture from IGCC Power Plant by Partial Condensation and Separation of Syngas

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16th Conference on Process Integration, Modelling and Optimisation for Energy Saving and Pollution Reduction

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Outline

- Pre-combustion CO_2 capture from IGCC
  - Focus on the capture unit

- Cooling of syngas to condense and separate CO_2 from combustibles
  - Alternative to physical and chemical solvents

- Explain how Pinch and Extended Pinch analysis has been used to design an efficient process
  - Subambient process design

- Briefly on work performed beyond the scope of the conference paper

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Subambient process integration

Important points mentioned by Prof. Truls Gundersen yesterday (right before the blackout):

- Compressors and expanders are very central units in Pinch analysis of subambient process

- Complicates Pinch analysis and may add several degrees of freedom

- Non-fixed path between supply and target state

- Relationship between pressure, temperature and phase

- Process streams may act as utility streams

- Due to high exergy content, tight heat integration of subambient streams is crucial for high energy efficiency

Introduction to IGCC CO_2 capture conditions

- IGCC syngas has relatively high CO_2 concentration and partial pressure
- Condensation and separation of CO_2 is an alternative to baseline technology (physical and chemical solvents)
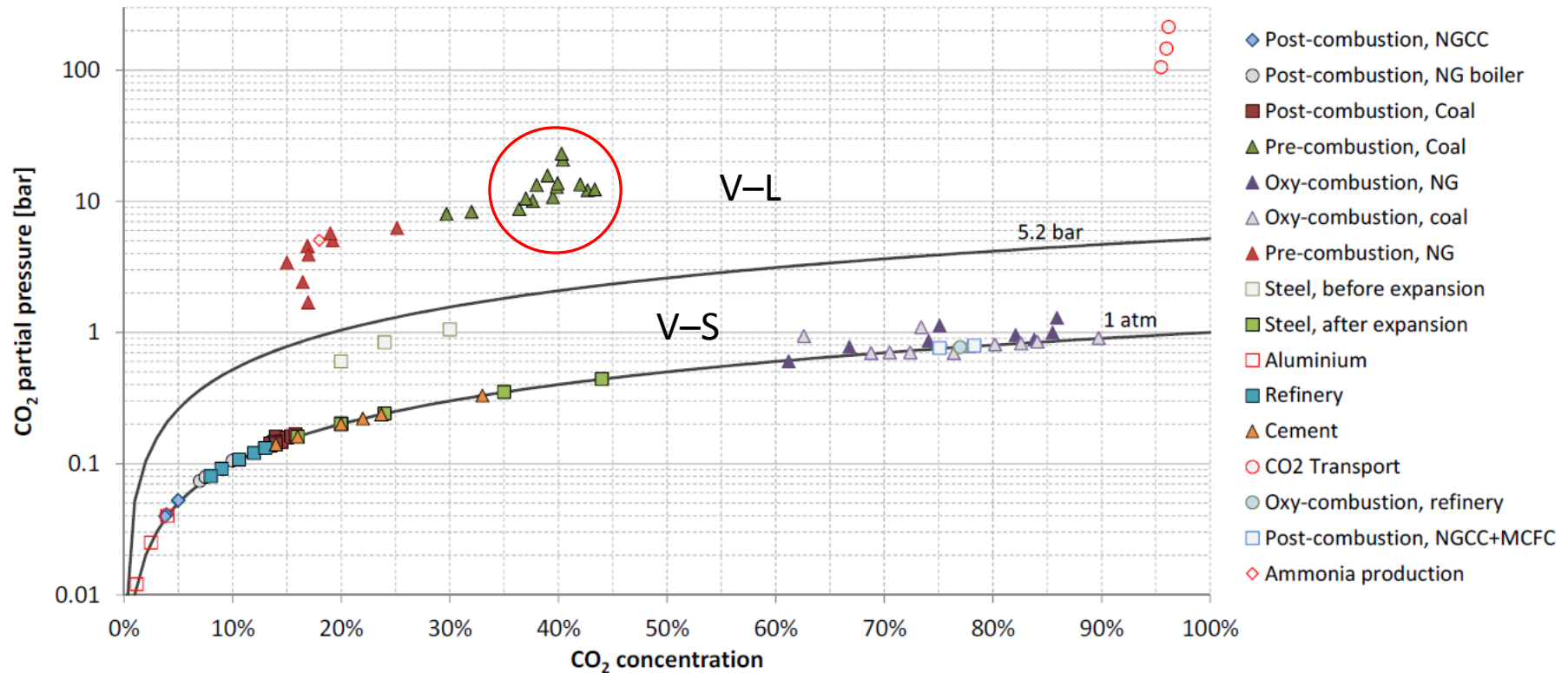
Figure from: David Berstad, Rahul Anantharaman, Petter Nekså. Low-temperature CO_2 capture technologies – Applications and potential (2013). International Journal of Refrigeration 36(5), 1403–1416.

CO$_2$ capture ratio based on V–L equilibrium

H$_2$ + CO$_2$

-53.15°C (220 K)

CO$_2$
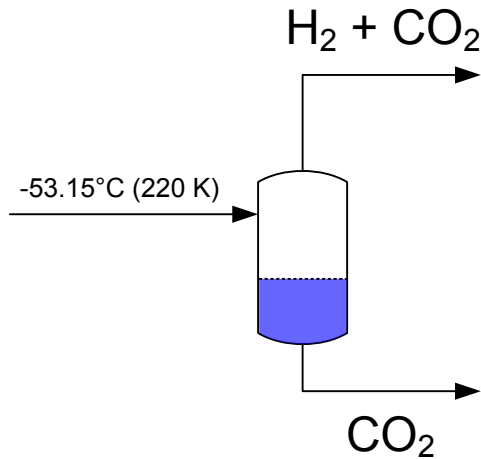
Composition for this case

Hydrogen	53.75%
CO$_2$	38.04%
CO	1.60%
N$_2$	5.71%
Ar	0.90%

Legend:
- Peng–Robinson: 42% CO2, 58% H2
- Peng–Robinson: 40% CO2, 60% H2
- Peng–Robinson: 38% CO2, 62% H2
- Spano et al.: 40% CO2, 60% H2
- Tsang & Streett: 40% CO2, 60% H2

Y-axis: CCR [%]
X-axis: Separation pressure [bar]
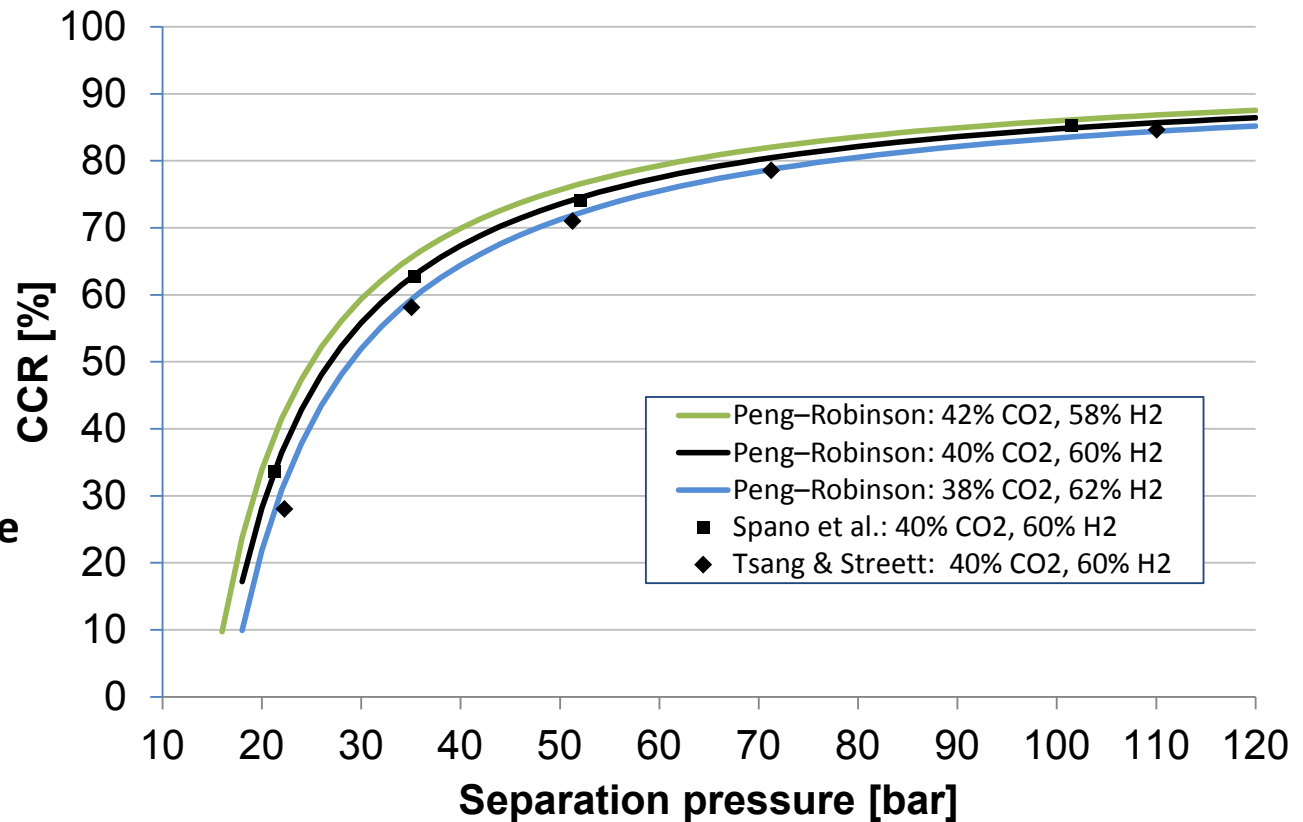
Flowrate: 114 kg/s
Available at 35 bar, 30°C

Figure from: David Berstad, Rahul Anantharaman, Petter Nekså. Low-temperature CCS from an IGCC power plant and comparison with physical solvents (2012). Energy Procedia 37, 2204–2211.

Process structure with respect to the design hierarchy

'Onion diagram' from: Linnhoff B., Townsend D.W., Boland D., Hewitt G.F., Thomas B.E.A., Guy A.R., Marsland R.H., 1982, A User Guide on Process Integration for the Efficient Use of Energy, IChemE, Rugby, UK.
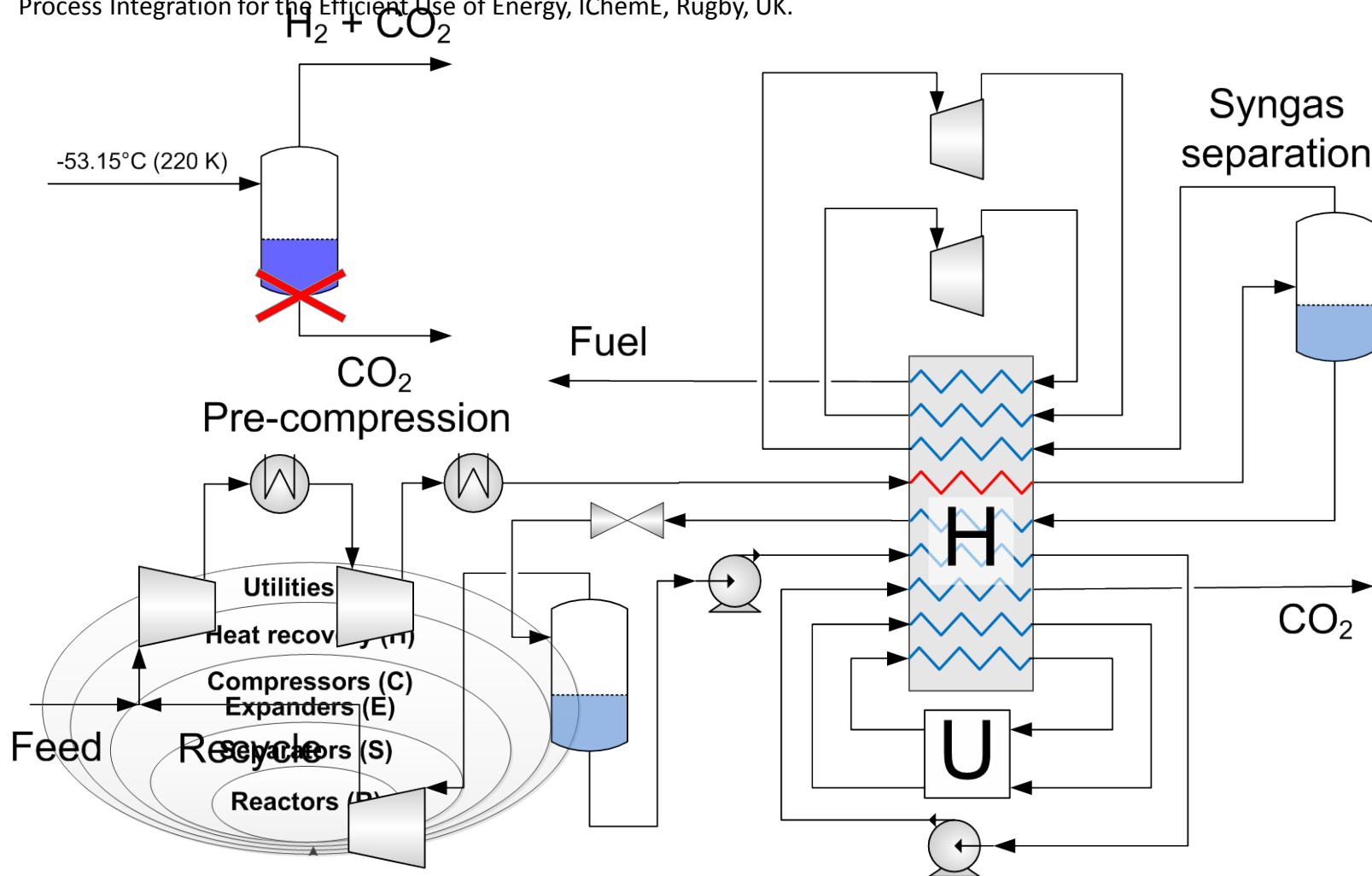
Process assumptions made before Pinch analysis

- 115 bar pre-compression (85% CO_2 capture ratio)
- -56°C separation temperature
- Purification flash at ca. 7 bar, -56°C
- -42°C pre-flash temperature
- Two gas-product expanders
- 25 bar fuel pressure
- Two CO_2 pumps (95 bar, 150 bar)

Syngas separation

Fuel

Pre-compression

H

U

Feed   Recycle

CO_2
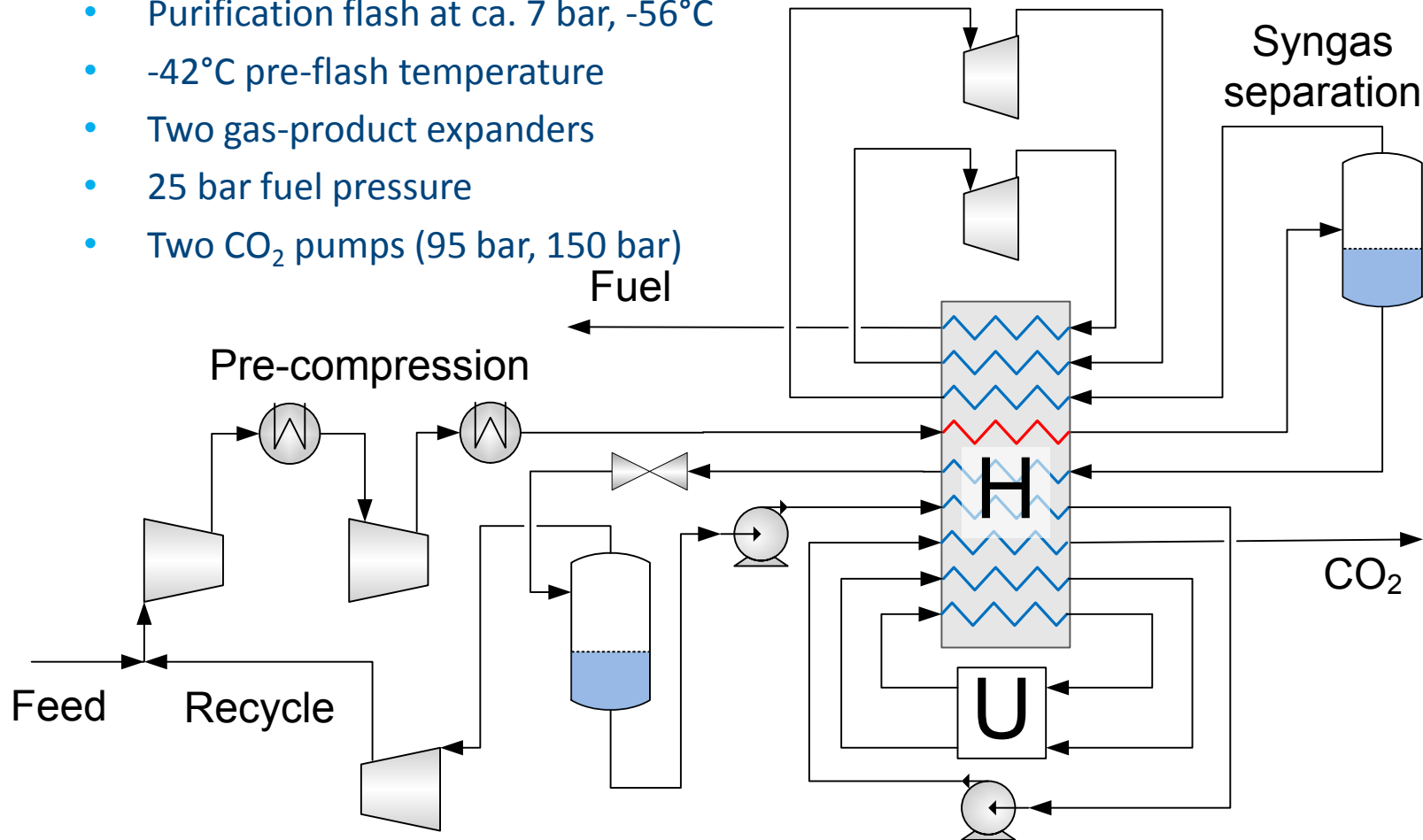
Pinch analysis of process streams
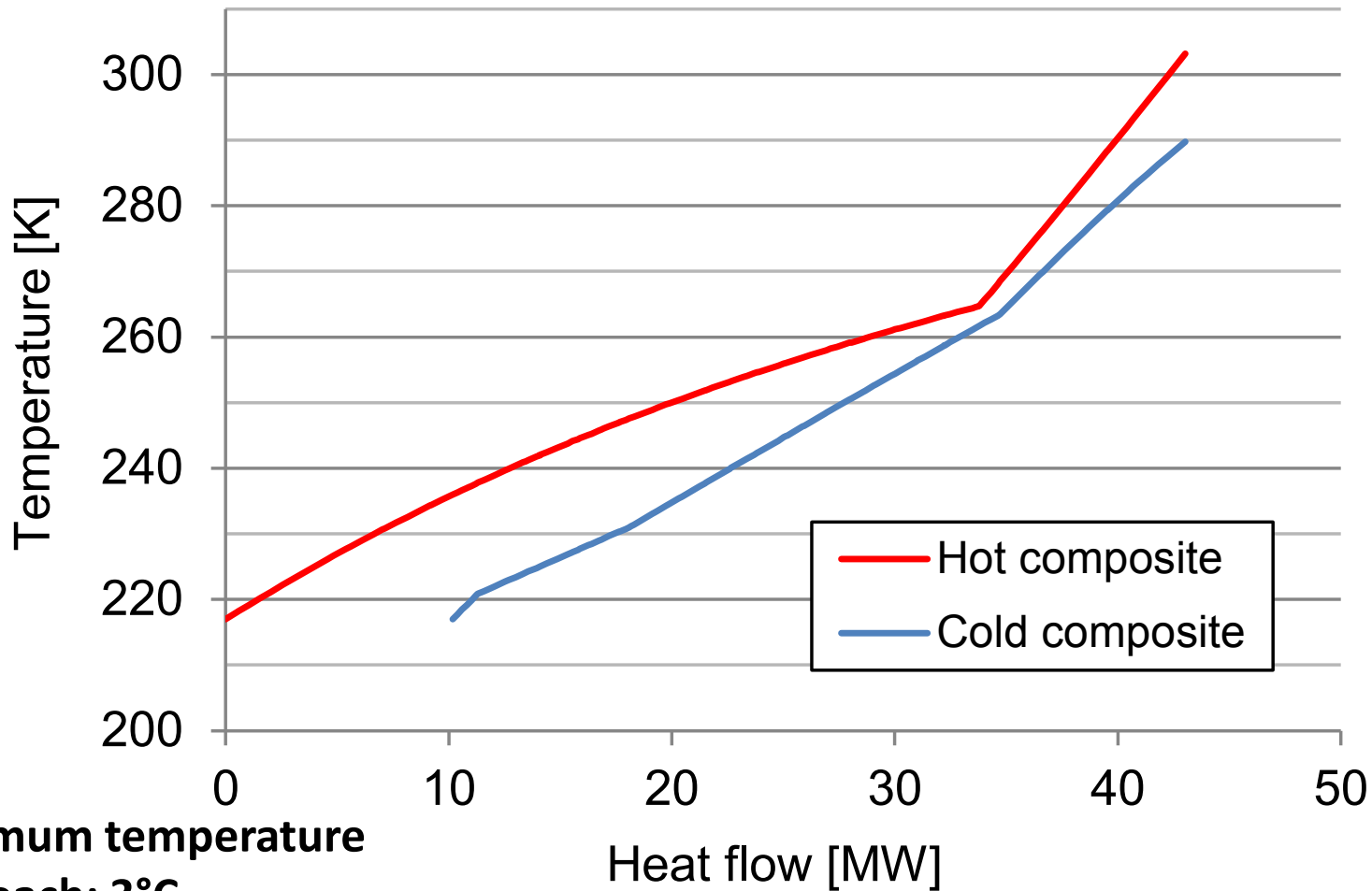
Minimum temperature approach: 3°C

Utilities selection: propane/ethane vapour compression cascade cycle
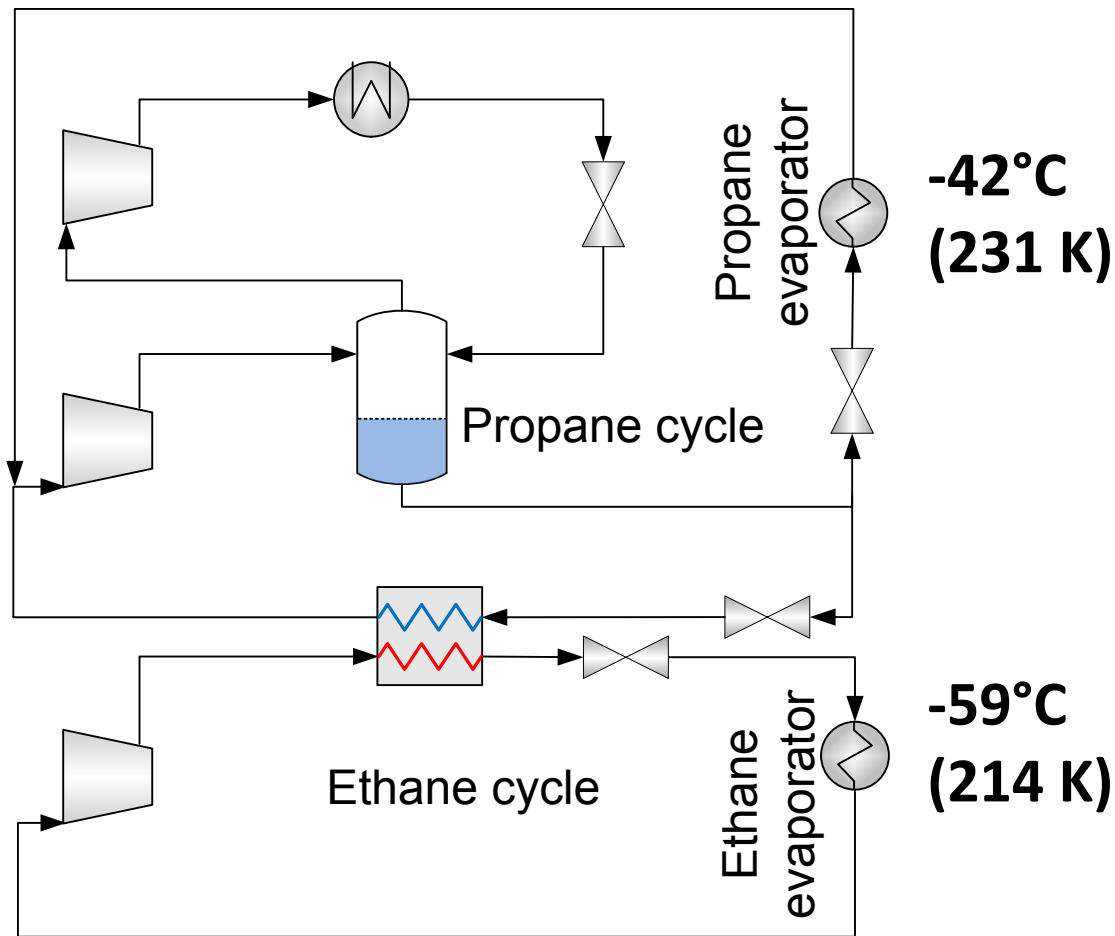
Pinch analysis and energy targeting with utilities

Minimum temperature approach: 3°C
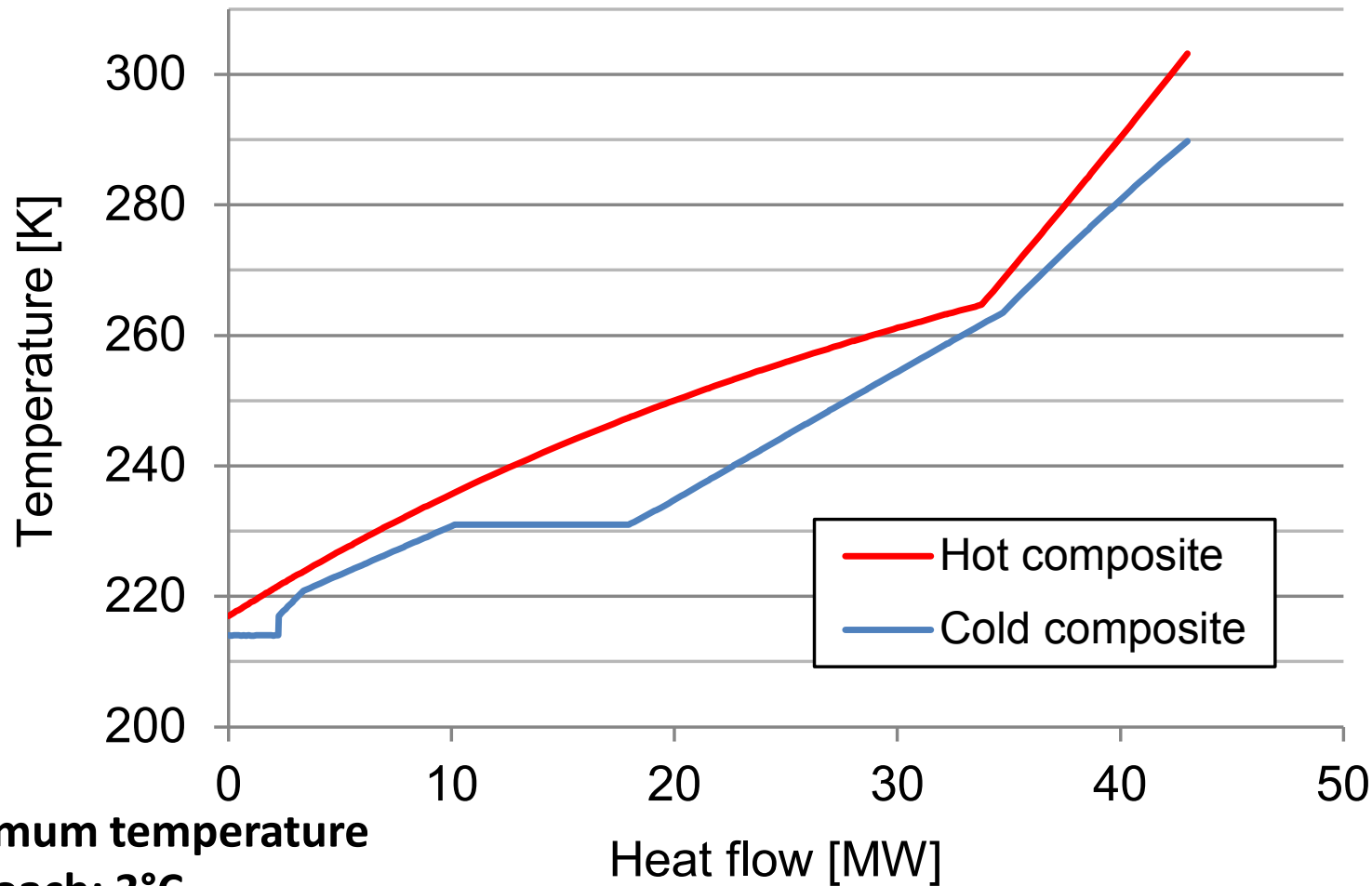
Pinch analysis and energy targeting with utilities
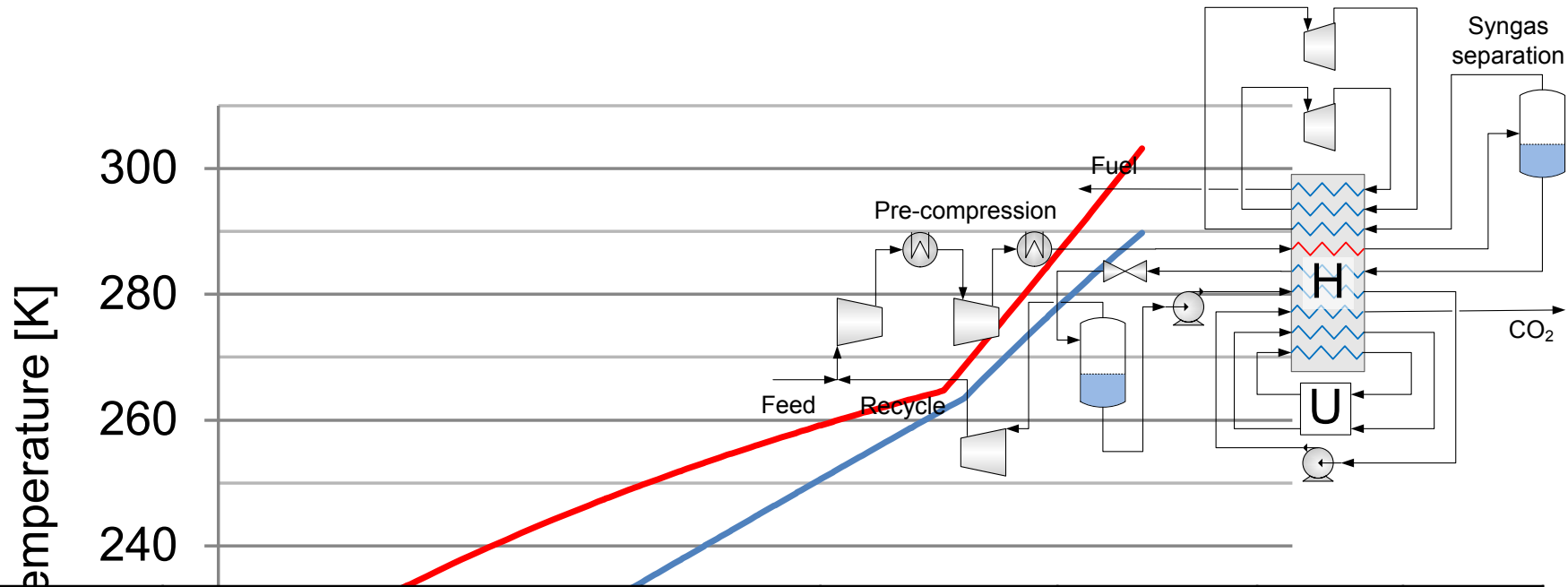
	Syngas compression	Recycle compression	CO_2 pumping	Auxiliary refrigeration	Fuel expansion	CW pumps	Net power
Energy targeting [MW]	22.1	1.12	1.41	5.29	9.17	0.21	21.9*
* 90% shaft power recovery assumed							

Heat exchanger network design

- Plate-fin heat exchangers selected

- Two- and multi-stream heat exchangers

- Widely used in gas liquefaction and separation, e.g. cryogenic air separation, LNG, hydrogen liquefiers, helium liquefiers

- Design pressure > 100 bar and temperature < -269°C (4 K) possible

Stream 1 (outlet)
Stream 4 (outlet)
Stream 2 (inlet)
Stream 3 (outlet)
Stream 5 (outlet)
Stream 1 (inlet)
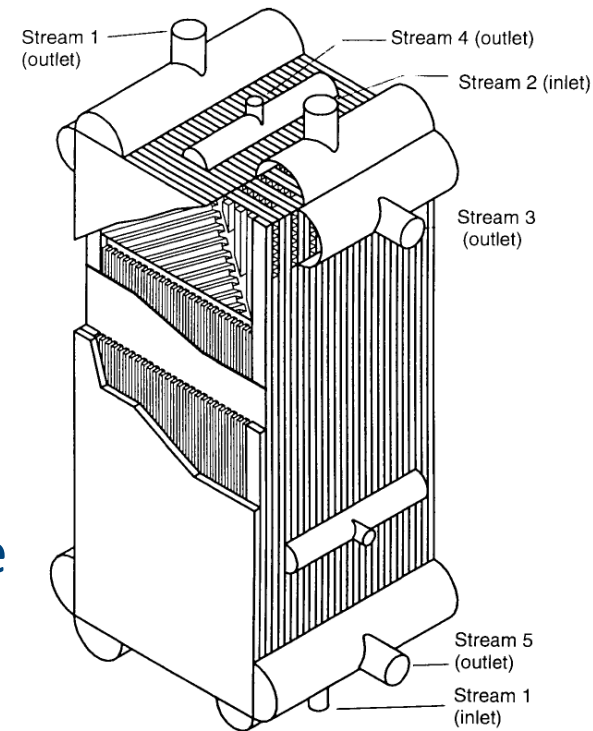
Source: ALPEMA

The Standards of The Brazed Aluminium Plate-Fin Heat Exchanger Manufacturers' Association (www.alpema.org)

Process layout
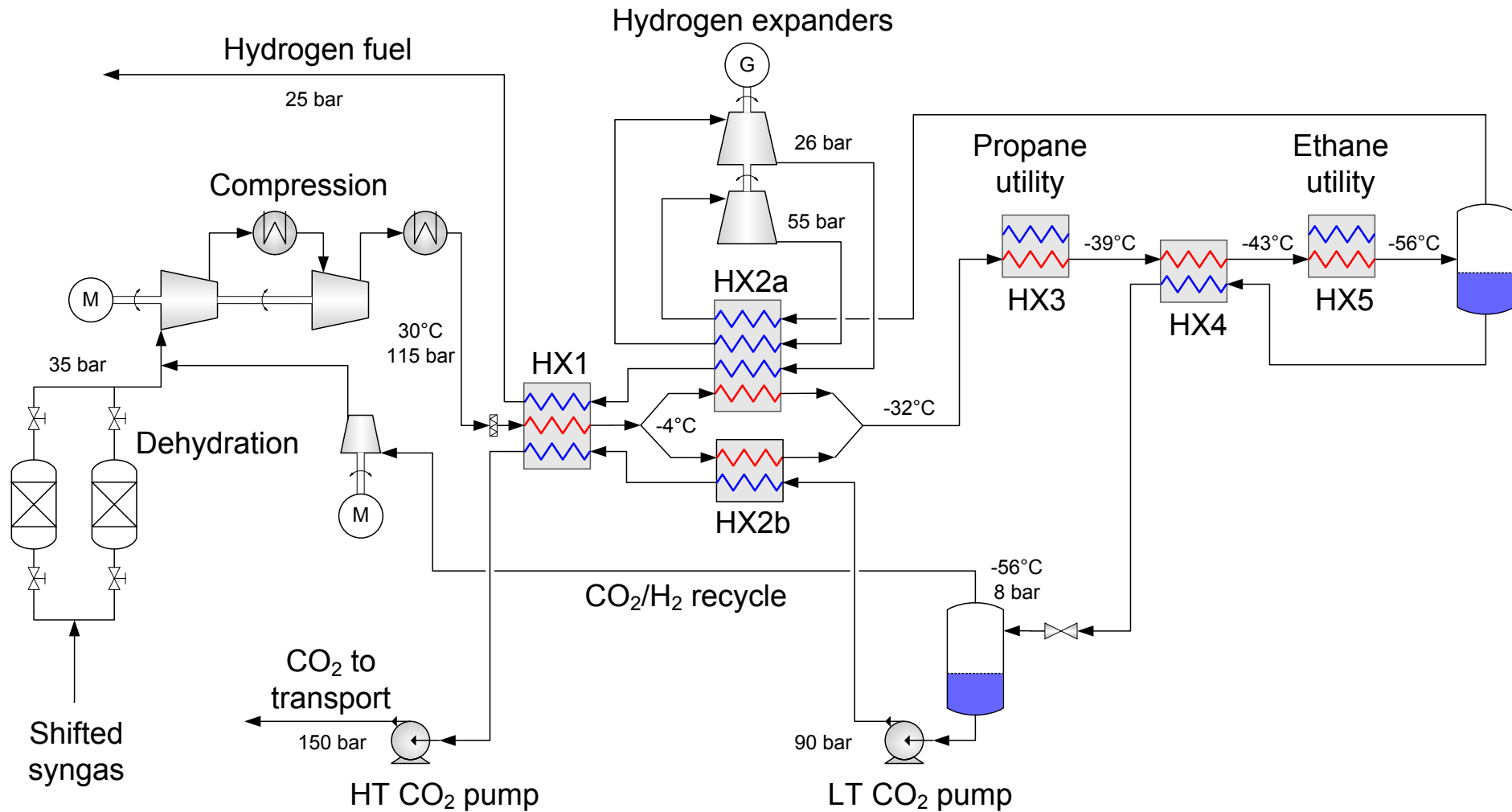
Process layout
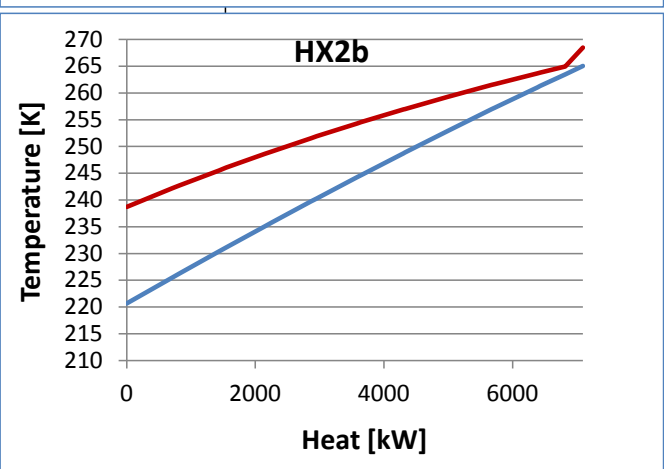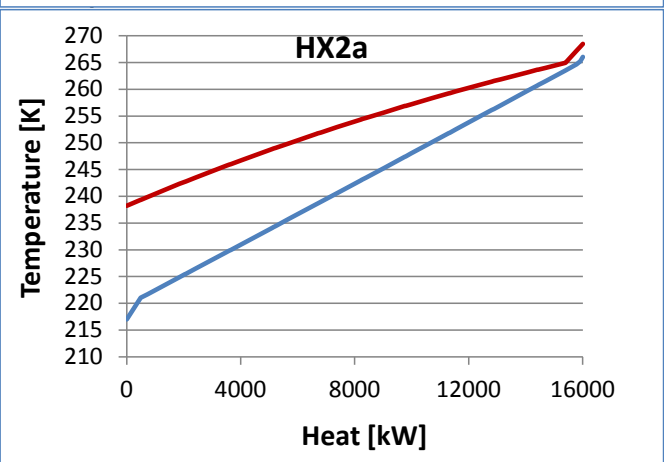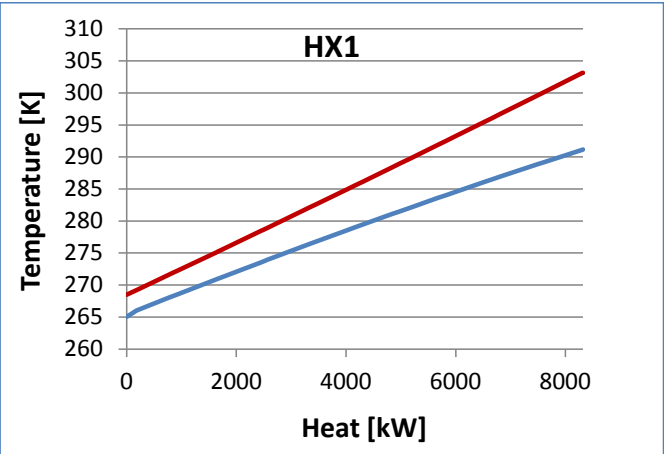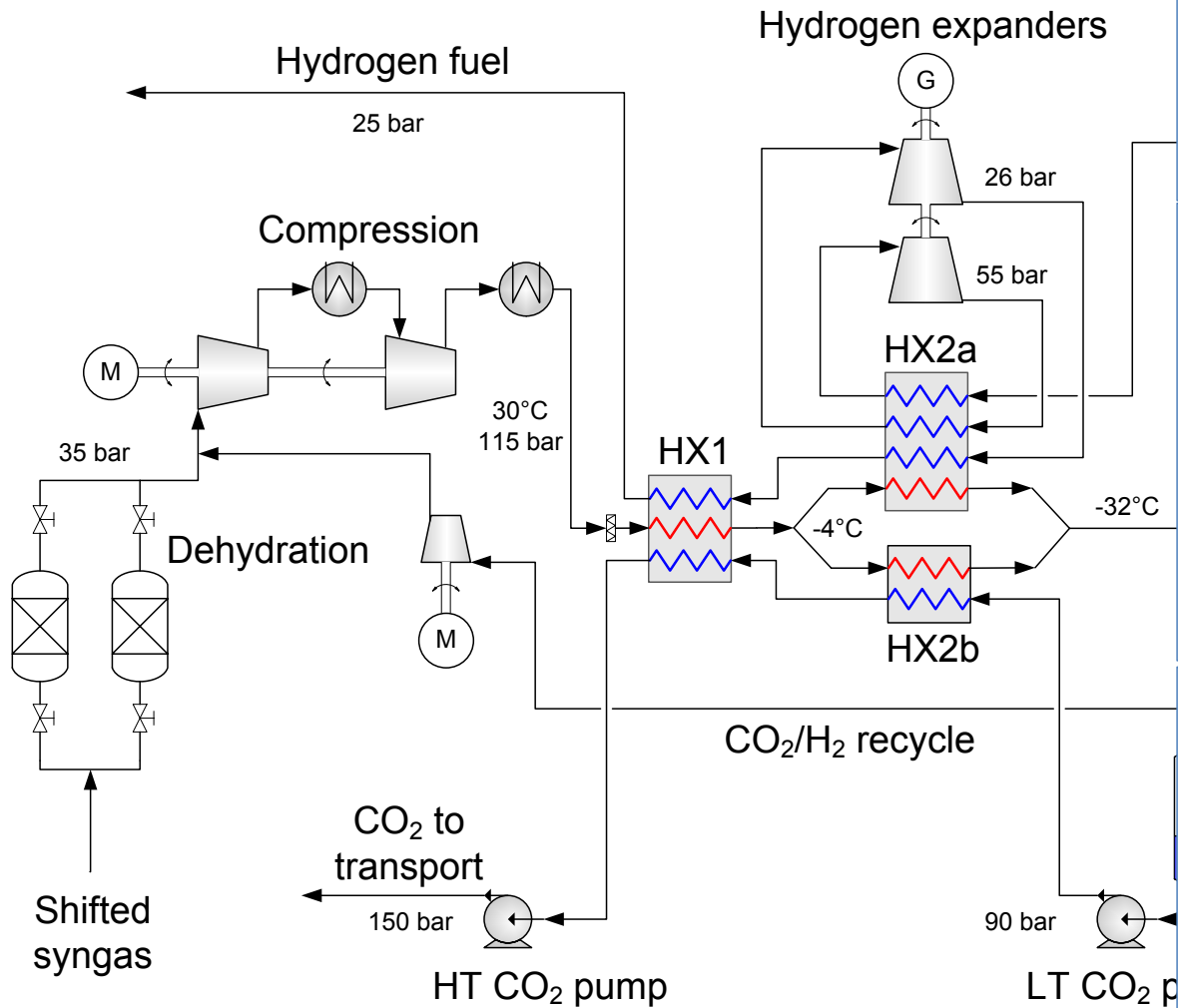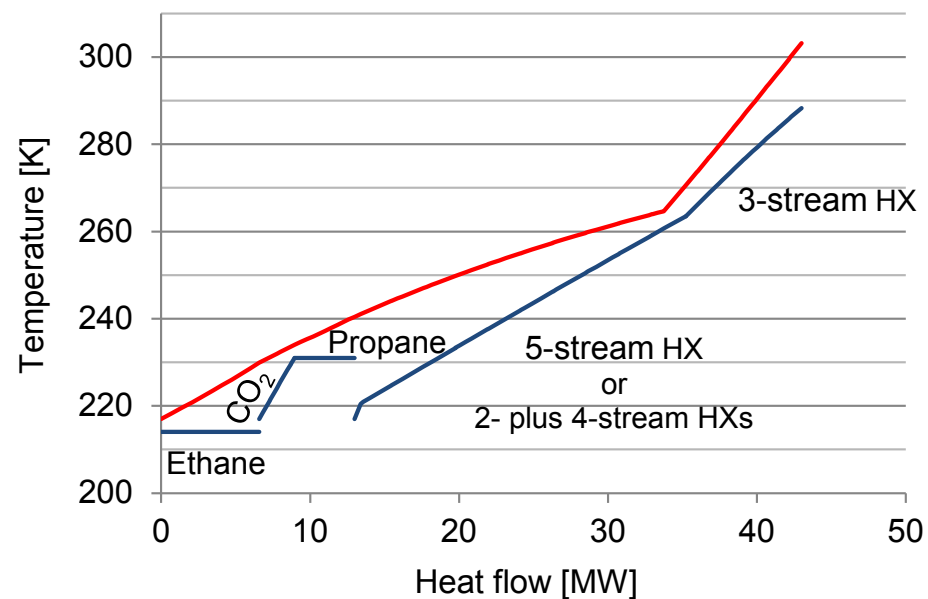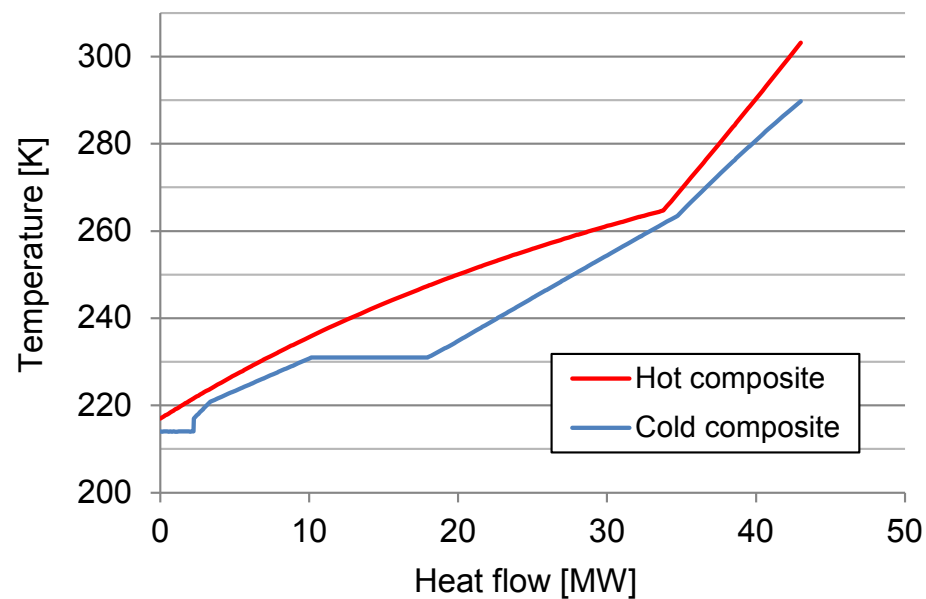
	Syngas compression	Recycle compression	CO_2 pumping	Auxiliary refrigeration	Fuel expansion	CW pumps	Net power
Energy targeting [MW]	22.1	1.12	1.41	5.29	9.17	0.21	21.9*
Proposed design [MW]	22.1	1.12	1.40	6.40	9.25	0.22	22.9*
* 90% shaft power recovery assumed							

Beyond Pinch analysis and heat exchanger network design

Simulation results for HX2a – the most complex heat exchanger

- 2 heat exchanger blocks in parallel
- Layer sequence: ACABAD/46
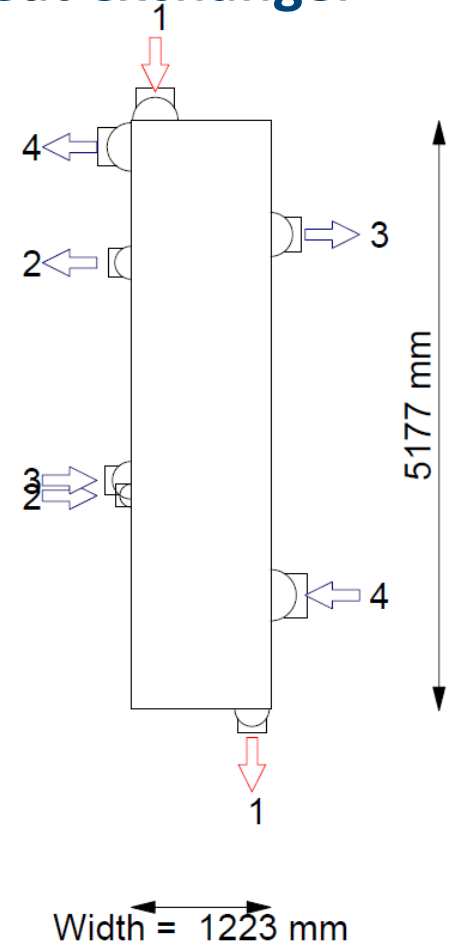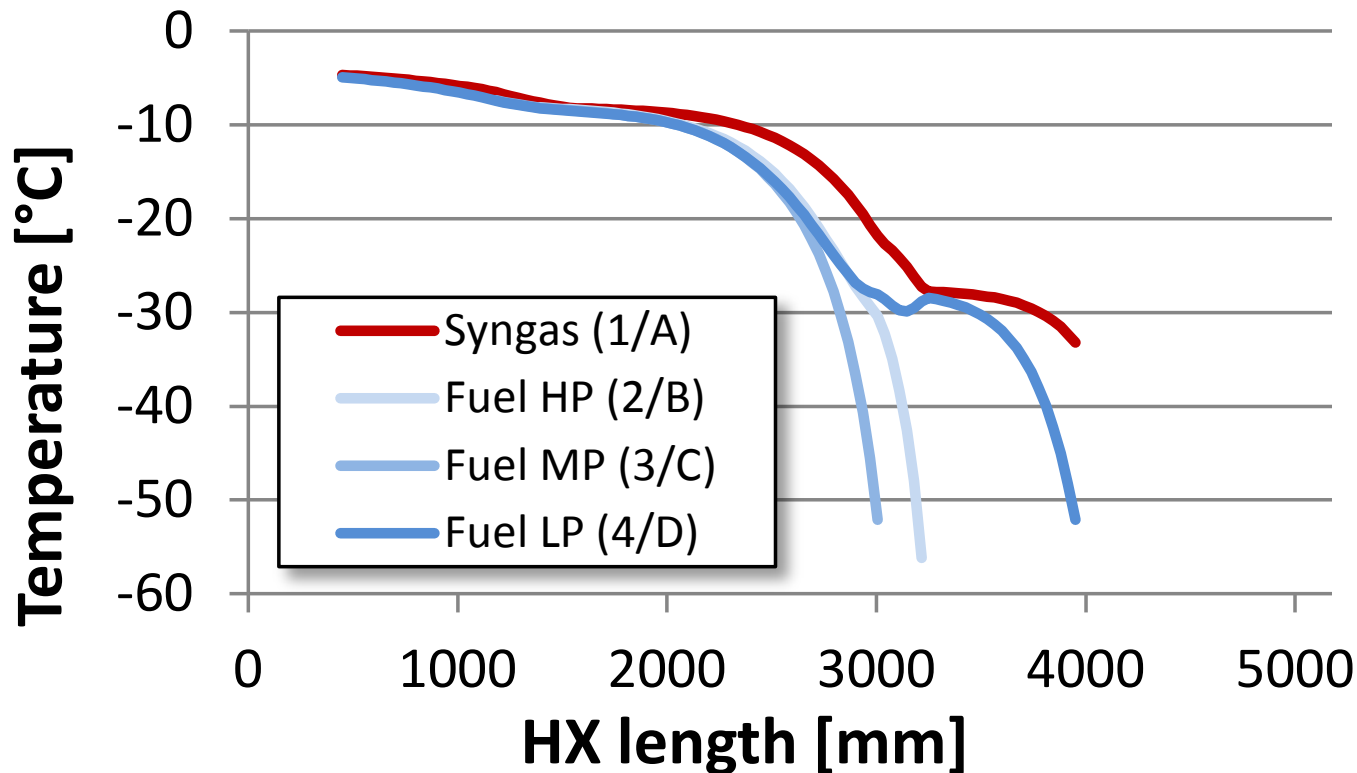
Summary and conclusions

- Pinch and extended Pinch analysis for subambient processes have been used heuristically to obtaining an energy efficient CO_2 capture process for IGCC synthesis gas

- Energy target based on extended Pinch analysis was 21.9 MW while actual process design with heat exchanger network design resulted in 22.9 MW (+4.6%)

- The low-temperature separation is energy efficient and highly competitive with physical solvents
  - 0.8–1.2 %-points improvement in IGCC net electric efficiency vs. baseline (Selexol) for 85% capture-unit CCR (Berstad et al., 2013)

- As shown when switching from overall process design to detailed plate-fin heat exchanger design, the rigorously defined minimum temperature approach (3°C) no longer constrains actual process design

- Further and ongoing work comprises: heat exchanger design, compressor and expander performance verification, off-design operation and flexibility.

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Acknowledgements

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The authors also acknowledge Gassnova SF for funding of this research – Project number (CLIMIT) 223123.

Rotating equipment

- Syngas compressors (2 units)
  - Type: centrifugal turbo-compressors
  - Suction volumetric flow: 15,000 and 7,800 m^3/h
  - Molecular weight of gas: 20.9
  - Efficiency: 82–85% isentropic efficiency assumed
  - Load: 11–12 MW per machine
- Fuel expanders (2 units)
  - Type: centrifugal turbo-expanders
  - Inlet volumetric flow: 2,700 and 5,700 m^3/h
  - Molecular weight of gas: 8.9
  - Efficiency: 85–88% isentropic efficiency assumed
  - 4.5–5.0 MW per machine
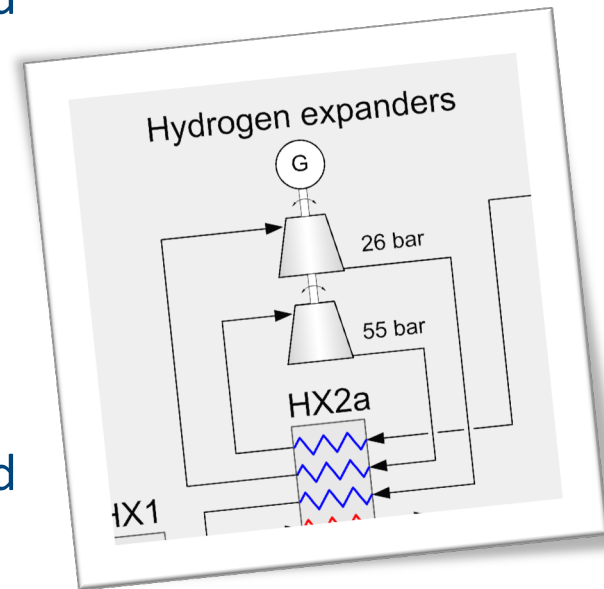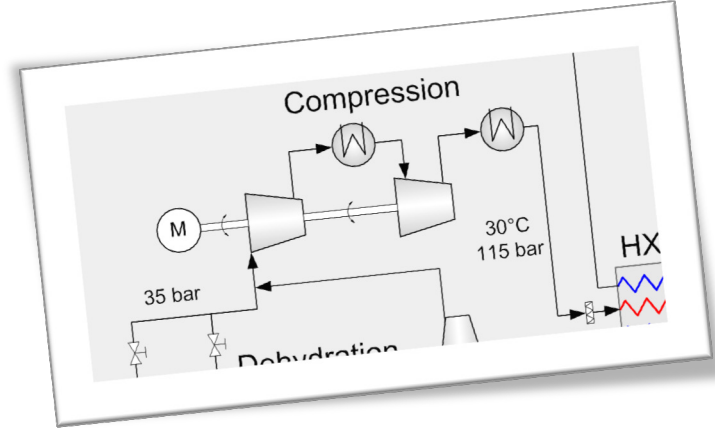
Critical pressure

- $P_{critical}$ > 600 bar for 50% CO_2, 50% H_2
- $P_{critical}$ > 1200 bar for 40% CO_2, 60% H_2
- $P \ll P_{critical}$ → Cubical EOS are okay to use

CO$_2$ freeze-out

Binary H$_2$–CO$_2$:

- For -53°C melting point pressure ~ 350 bar

- For 110 bar: freezing point likely between -56 and -57°C

CO$_2$ freeze-out calculations for actual syngas composition with embedded HYSYS utility:

- -58.49°C

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VLE

- Current H_2–CO_2 EOS interaction parameter: 0.1202

- If changed to 0.104:
  - CCR: Drops by about 0.1 %-point
  - CO_2 purity: Drops by about 0.01 mol%-point

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Paper can be found using the DOI-link: http://www.aidic.it/cet/13/35/059.pdf