# CHALMERS



# Integration of a Carbon Capture process in a chemical industry

## Case study of a steam cracking plant

*Master's Thesis within the Innovative and Sustainable Chemical Engineering programme*

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*Division of Heat and Power Technology*
CHALMERS UNIVERSITY OF TECHNOLOGY
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MASTER'S THESIS

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Cover:
Schematic figure of a carbon capture plant with a transportation pipeline to a storage well at the bottom of the ocean.

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ABSTRACT

One of the largest environmental challenges of modern time is to reduce the emissions of green house gases (GHG) in general, and particularly the emissions of $CO_2$. One large source of GHG emissions is process industry. Carbon capture and storage (CCS) technologies are currently being developed to separate the $CO_2$ from flue gases generated from combustion. Borealis site in Stenungsund is involved in an initiative that aims to reduce the $CO_2$ emissions in the Skagerrak region with CCS technology. A post combustion carbon capture process with mono-ethanolamine (MEA) absorption is evaluated to separate $CO_2$ from flue gases generated in Borealis' cracker plant.

The post combustion process was simulated in Aspen Plus to find the optimal operating conditions. The post combustion process is very energy demanding. To reduce the energy demand process integration possibilities with the cracker plant were investigated with help of pinch analysis. Utilizing excess heat from the carbon capture process showed substantial heat savings potentials. It was also shown that the heat saved could be used in the post combustion process. 40 MW heat representing 64 % of the carbon capture process heating demand could be saved by process integration and optimization.

A study by Tel-Tek investigated the implementation of the post combustion process without process integration. In order to meet the heating demand of the process a steam boiler was proposed to supply the heat. This thesis shows that investment of a steam boiler can be avoided thanks to process integration. Compared to the study made by Tel-Tek, the investment costs for the capture plant were reduced from 114.5 M€ to 66.9 M€ and the operating costs were reduced from 24 M€/yr to 15.9 M€/yr. The resulting $CO_2$ net capture cost was reduced from 80 €/ton $CO_2$ to 55 €/ton $CO_2$.

It should be pointed out that the results are based on theoretical savings. Costs for retrofits should be studied in detail to assess the actual savings. Transportation and storage costs should also be considered in possible future studies.

Key words: Process integration, pinch analysis, post combustion process, CCS, process simulation, $CO_2$ capture

Integration av en koldioxidavskiljningsprocess i en kemisk industri
Fallstudie av en ångkrackeranläggning
Examensarbete inom masterprogrammet *Innovative and Sustainable Chemical Engineering*

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SAMMANFATTNING

En av vår tids stora miljöutmaningar är att minska utsläppen av växthusgaser i allmänhet, och $CO_2$ i synnerhet. Många av de största utsläppskällorna av växthusgaser finns inom processindustrin. Teknologier för koldioxidinfångning och lagring (CCS) håller för närvarande på att utvecklas för att avskilja $CO_2$ från förbränningsgaser. Borealis kracker i Stenungsund är en del av ett initiativ där målet är att minska utsläppen av $CO_2$ i området runt Skagerrak med CCS teknologi. Den studerade $CO_2$ avskiljningsprocessen är med absorption där mono-etanolamin (MEA) använts som absorptionsmedel.

För att bestämma driftförhållanden i avskiljningsprocessen har simuleringar gjorts i Aspen Plus. Processen är energikrävande och för att minska energibehovet har möjligheter till processintegrering med krackern studerats med hjälp av pinchanalys. Genom att nyttiggöra överskottsvärme i avskiljningsprocessen kunde betydande mängder energi sparas i krackern. Det visade sig att den sparade energin kunde tillföras i avskiljningsprocessen. 40 MW värme vilket motsvarar 64 % av energibehovet i avskiljningsprocessen kunde sparas tack vare processintegrering och processoptimering.

En studie av Tel-Tek har gjorts där implementering av denna avskiljningsprocess studerats utan processintegrering. För att tillfredsställa energibehovet i processen hade man föreslagit investering i en ångpanna. Detta examensarbete visar på att en investering av en ångpanna kan undvikas tack vare processintegrering. Investeringskostnaderna för ingångningsanläggningen minskade från 114.5 M€ till 66.9 M€ och driftkostnaderna från 24 M€/år till 15.9 M€/år. Detta resulterade i en lägre nettoinfångningskostnad av $CO_2$, från 80 €/ton $CO_2$ till 55 €/ton $CO_2$.

Det skall poängteras att resultaten är baserade på den teoretiska mängd energi som kan sparas. Ombyggnadskostnader bör studeras i detalj för att fastställa de faktiska energibesparingarna. I eventuella framtida studier bör hänsyn också tas till transport- och lagringskostnader av $CO_2$.


Nyckelord: Processintegrering, pinchanalys, avskiljningsprocess, processimulering, koldioxidinfångning

# Contents

# Preface

This master thesis was carried out in cooperation with the Division of Heat and Power Technology at Chalmers University of Technology and Borealis AB in Stenungsund. The study investigates a simulated and optimized $CO_2$ capture process integrated with the cracker plant.

I would like first to thank my supervisors Roman Hackl and Jonas Sjöblom at the Division of Heat and Power Technology, Chalmers, and Jonny Andersson and Reine Spetz at Borealis, for their excellent and valuable help. I would also like to thank Professor Simon Harvey and Professor Thore Berntsson at the Division of Heat and Power Technology, Chalmers, for this great opportunity. A special thanks to Stefan Heyne and Rickard Fornell for their time and helpful contribution to the simulations. I would also like to thank all other employees who have contributed with information to this thesis. Finally, I like to thank my family for the support during the thesis.

Gothenburg, November 2010

Ahmed Sherif

# Notations

**Abbreviations:**

| | |
|---|---|
| CAPEX | Capital Expenditures |
| CC | Composite Curve |
| CCS | Carbon Capture and Storage |
| ETS | Emissions Trading System |
| GCC | Grand Composite Curve |
| GHG | Green House Gases |
| HDPE | High Density Polyethylene |
| HP | High Pressure |
| IP | Intermediate Pressure |
| LDPE | Low Density Polyethylene |
| LLDPE | Linear Low Density Polyethylene |
| LP | Low Pressure |
| LPG | Liquefied Petroleum Gas |
| MEA | Mono-EthanolAmine |
| MP | Medium Pressure |
| MVR | Mechanical Vapour Recompression |
| OPEX | Operational Expenditures |
| PE | Polyethylene |

**Roman upper case letters**

| | |
|---|---|
| $C_k$ | Cost of Equipment k, [€] |
| $C_x$ | Cost of Equipment x, [€] |
| $C$ | Concentration, [mol/m$^3$] |
| $Cp$ | Heat Capacity, [kJ/kg,K] |
| $E_k$ | Equipment of Size k, [m$^2$] or [m$^3$] or [kW] |
| $E_x$ | Equipment of Size x, [m$^2$] or [m$^3$] or [kW] |
| € | Euros, [-] |
| $F$ | Mass Flow Rate, [kg/s] |
| $N$ | Number of Sections in Column, [sections] |
| $P$ | Pressure, [Pa] |
| $Q$ | Duty, [kW] |
| $R$ | Ideal Gas Constant, [kJ/kg,K] |
| $T$ | Temperature, [K] |
| $\Delta T_{min}$ | Minimum Temperature Difference, [K] |
| $\dot{V}$ | Volumetric Flow Rate, [m$^3$/s] |

**Roman lower case letters**

| | |
|---|---|
| $i$ | Interest Rate, [1/yr] |
| $n$ | Number of Years, [years] |
| $n_c$ | Cost Capacity Exponent, [-] |
| $\dot{n}_{CO2}$ | CO$_2$ Molar Flow Rate, [mol/s] |

# 1      Introduction

This thesis was made in collaboration with Borealis AB. Borealis AB is the Swedish entity of Borealis AG, which is a manufacturer of chemicals and plastics with production facilities in Europe, Brazil and the US. As a large producer of polyolefins Borealis supplies plastics to infrastructure, automotive and advanced packaging markets in Europe, North and South America, Middle East and Asia. (Borealis AB, 2010A)

Borealis in Stenungsund is Sweden's only producer of polyethylene (PE). Three types of PE are produced.

- Low Density PE (LDPE) – branched PE chains produced in a high pressure process
- Linear Low Density PE (LLDPE) – significantly linear PE chains with short branches produced in a low pressure catalytic process
- High Density PE (HDPE) – linear PE chains with no branches produced in a low pressure catalytic process

PE produced in Stenungsund is mostly sold to customers for wire, cable and pipe manufacturing. The old high-pressure plant has been replaced by a new unit (LD5), which was taken in operation in 2010. Compared to the old, not only it has a higher production capacity, but also the LD5 is more energy efficient due to advanced heat recovery and other energy efficiency measures. Table 1 shows the production rate of PE. (Borealis AB, 2010B)

*Table 1. Production of PE in 2007. *Design capacity. (Borealis AB, 2007).*

| Product | [ktonnes/year] |
|---|---|
| High Pressure PE (LDPE): Old | 154 |
| High Pressure PE (LDPE): LD5 | 350* |
| Low Pressure PE (HDPE) | 162 |
| PE3 Borstar (HDPE) | 215 |

Ethylene for the PE plants is produced in a steam cracker plant from among others ethane, propane, butane and naphtha. The flexibility of the cracker furnaces makes it possible to use various feedstocks as the residence time in the furnaces can be varied to obtain the wanted conversion of the feed to products. Table 2 shows the feedstock consumption and production rate. (Borealis AB, 2010B)

*Table 2. Cracker plant production rate in 2007. Note: fuel gas is a mixture of methane and hydrogen. ETBE is ethyl-tert-butylether. (Borealis AB, 2007)*

| Product | [ktonnes/year] |
|---|---|
| Ethylene | 622 |
| Propylene | 200 |
| Others (incl fuel gas) | 488 |
| ETBE | 28 |

Gas product cooling from the furnaces produces steam, which is used in the cracker process. Steam is also produced in steam boilers to meet the heating demand of the cracker process. The current hot utility consumption is 94 MW and the cold utility consumption is 120 MW (Hedström, 2008). Moreover, the cracker in Stenungsund is one of Sweden's largest emitters of fossil $CO_2$, 1.2 % of the Swedish overall $CO_2$ emissions (Hedström, 2008). The annual $CO_2$ emissions from the cracker furnaces is about 560 000 tonnes, which is approximately 83 % of the $CO_2$ emitted from the cracker plant (Borealis AB, 2007).

Emission trading system (ETS) and other mitigation initiatives
One of the largest environmental challenges of modern time is to reduce the emissions of green house gases (GHG) in the atmosphere. To meet these challenges 192 parties have signed and ratified the Kyoto Protocol to date (UNFCCC, 2010), with the purpose to prevent levels of GHG in the atmosphere that can affect the climate negatively and damage the environment. The Protocol was signed in Kyoto in 1997 and took effect in 2005 aiming to reduce the total GHG emission in developed countries during 2008 – 2012 with 5 % of the GHG emitted year 1990 (UNFCCC, 2010). However, within the European Union (EU) there has been an agreement to make that reduction to 8 %. To reach this target the ETS is an essential tool, a system that was taken into force 2005 (Energimyndigheten, 2009). To work towards the GHG emission goal set by the Swedish parliament the cracker plant became partially included in the ETS between 2005 and 2007 (Energimyndigheten, 2009). Since 2008, all parts of the plant are a part of the trading system including cracking furnaces and flares (Borealis AB, 2007).

Carbon capture and storage (CCS) technology in the industrial sector is investigated to reduce the emissions in the EU. CCS technology aims to separate $CO_2$ from flue gases and transport the separated carbon dioxide in pipelines for storage e.g. in depleted oil fields. Figure 1 shows an example of how CCS can be implemented in industry.
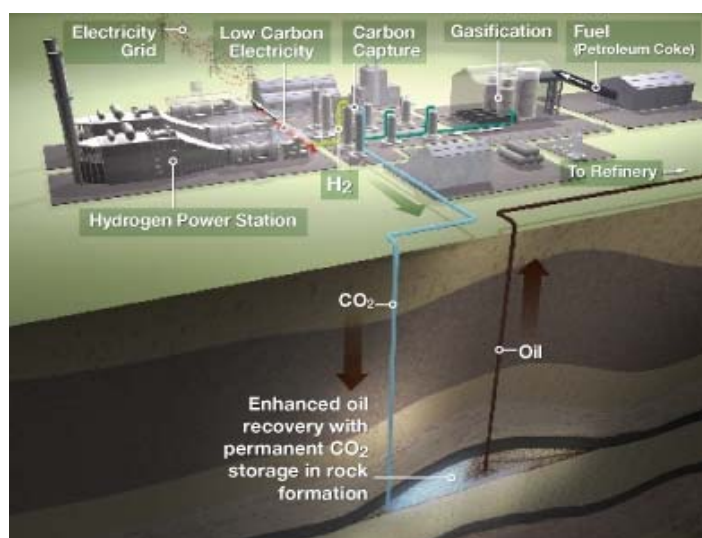


*Figure 1. An example of CCS technology. In this example the $CO_2$ is produced from a gasification process, but CCS can be used for any combustion process using a fuel that contains carbon. The $CO_2$ can be stored in different places. In this example the $CO_2$ is stored in an oil field, which also enhances the oil recovery.*

Skagerrak III

Recently, Borealis joined an initiative where the aim is to reduce $CO_2$ emissions in the Skagerrak region. The idea is to explore the potential of implementing CCS technology to industries located in West Sweden, East Norway and North Denmark (see Figure 2). Eventually, the captured $CO_2$ from each region can be transported in a joint high-pressure pipeline system and stored in depleted oil and gas reservoirs or other geological formations at the bottom of the ocean. These reservoirs and geological formations are preferably located in the Skagerrak area, but the North Sea is also of interest even though it is located further than the Skagerrak area. However, the geological formation around the potential reservoirs has to be studied carefully.
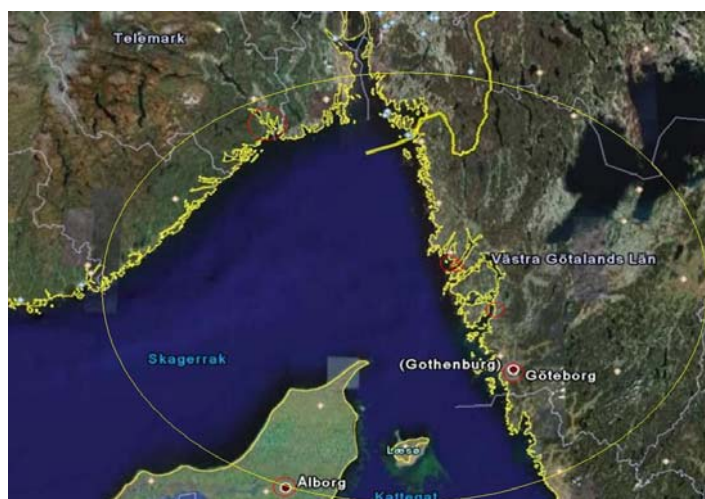


*Figure 2. The CO₂ emissions in this region are studied in the Skagerrak initiative.*

The $CO_2$ emissions from this region are about 13 million tons per year so there is a significant potential for reduction of the $CO_2$ emissions in this area. It is estimated that 10 million tons per year can be captured and stored. If the Skagerrak area could be used, the compressed $CO_2$ does not have to be transported long distances (within a radius of 100 km), which is an advantage from an economical point of view. (Eldrup, 2008A)

Carbon capture technologies

Currently, there exist three carbon capture technologies, presented below. These are pre-combustion, oxy-fuel, and post-combustion.

- Pre-combustion is a process that treats the fuel before the combustion. An example is gasification of carbon rich materials. Gasification is an endothermic process that converts a fuel to synthesis gas (CO and $H_2$) in lean oxygen conditions. Other products formed are mainly $CO_2$ and $CH_4$. The synthesis gas is then used as a fuel in a combustion process. The gasification is followed by a $CO_2$ shift to convert the CO and water to $CO_2$ and more $H_2$. Then the $CO_2$ is separated in a membrane from $H_2$ and the resulting pure $H_2$ stream can be burned in the combustion process.
- Oxy-fuel combustion is a process where the fuel is burned in pure oxygen instead of air. This reduces the volume of the flue gas since nitrogen is not present and therefore generates a flue gas with a high concentration of $CO_2$. Although, there

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are many advantages of this process the production of pure oxygen is energy consuming and therefore cost intensive.

- <u>Post-combustion</u> process is the third alternative in which the $CO_2$ is separated after the combustion. Technically, this is less complicated to implement in an existing plant than the previous mentioned technologies. The flue gases are directed from the stack into the capture process, where the $CO_2$ is absorbed from the flue gases in a solvent, usually aqueous amine and then desorbed from the solvent to obtain a pure $CO_2$ stream.

In this project a post-combustion process was analyzed since it is the only reasonable option for an existing plant to date. The most developed post-combustion process for $CO_2$ capture is by amine absorption, where the solvent is mono-ethanolamine (MEA) in an aqueous solution, usually 30 % MEA by weight (Eldrup, 2008A). Not only is it the most developed, but it also has potential of high $CO_2$ removal capacity. However, there is an important economical aspect. Due to the high heat of absorption generated, the endothermic $CO_2$ desorption reaction in the desorber is a very energy demanding step (Eldrup, 2008A).

# 2   Objective

The objective of this thesis is to investigate process integration possibilities for a post combustion carbon capture process with amine absorption and the cracker process at Borealis. Previously, a study was conducted by Tel-Tek in cooperation with Borealis AB, where the implementation of an amine carbon capture process was investigated. In that study no process integration was considered. Therefore all the heat supply to the carbon capture process was assumed to be supplied by a new boiler. However, integration of the two processes could reduce the total utility consumption substantially, as excess heat from the cracking process can be supplied to the capture process and vice versa, thereby decreasing the demand of external heat from a boiler. The objective is therefore also to determine if it is possible to supply enough heat from the cracker process to the carbon capture process, or vice versa, so that a boiler can be avoided.

Aspen Plus® is used as a tool to simulate the carbon capture process, to determine the energy demand. Information from a previous pinch analysis study of the cracker plant conducted at the Division of Heat and Power Technology were used in order to examine the integration possibilities between the cracker and carbon capture process. Different heat integration opportunities are examined, with respect to total external utility demand. The most promising configuration is analyzed with respect to investment and operating costs. The resulting specific carbon capture cost is then compared with results from the previously conducted conceptual Tel-Tek study of a non-integrated carbon capture process.

# 3 Borealis' Process Descriptions

The background describes Borealis' cracker process, the HPPE process, the utility systems and the carbon capture process. The descriptions in section 3.1 – 3.3 is from previous theses made at Borealis (Hackl, 2009) and (Hedström, 2008).

## 3.1 Borealis' cracker process
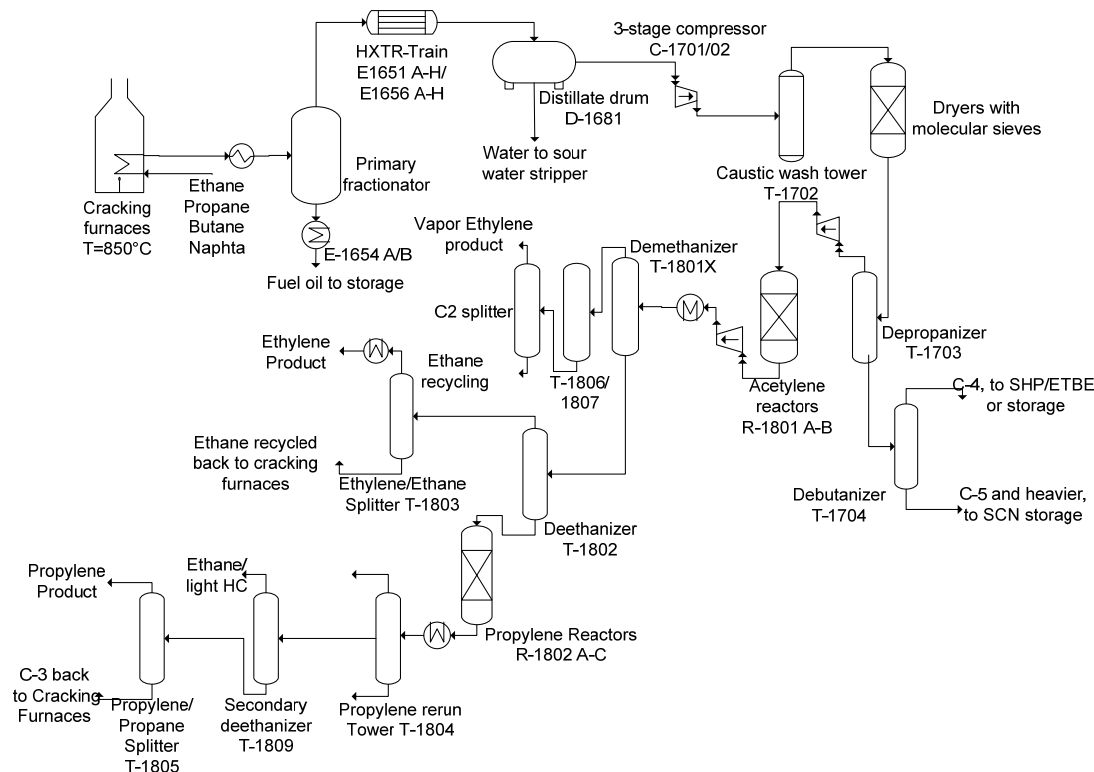


*Figure 3. The figure shows a schematic flowsheet of the cracker process*

The feedstock to the cracker plant is ethane, propane, naphtha and butane. The main products are ethylene and propylene and examples of by-products are ethylene-tert-butyl-ether (ETBE) and steam cracked naphtha (SCN).

### 3.1.1 Cracking units

The cracker plant consists of nine cracking furnaces, fractioning units and separation units. The feedstock enters the convection section of the cracking furnaces where they are preheated with the hot flue gases from the combustion of fuel gas (methane and hydrogen). After the preheating, the feedstock is mixed with steam and further heated right below the cracking temperature. The cracking into smaller molecules takes place in the radiation section of the furnaces. The exposure time in the radiation section, i.e. reaction time is very short. The 850 °C hot gases are heat exchanged with boiler feed water to produce high-pressure (HP) steam. To prevent further cracking and coke formation the stream is quenched. Quenching implies injecting liquid oil that cools the cracked gases when evaporated. The cracked hydrocarbons have to be cooled before the fractioning unit. To enable separation between the lighter and heavier fractions

produced after the cracker furnaces, it is necessary to cool the gases. By injecting quench oil the gases are cooled to 200 °C, before entering the primary fractioning section.

### 3.1.2 Primary fractioning

The cracked gas goes to the primary fractioning where a separation between light and heavy fractions takes place. The liquefied heavy fractions are stored and sold to customers. The light fractions (and steam) are cooled in a heat exchanger train. The cooled stream consists of a light fraction vapour, a liquid phase and water. The phases are separated in a drum in which the light fraction liquid phase is sent to a distillate stripper. The water is sent to a sour water stripper as it contains volatile hydrocarbons and phenols. The light fraction gas phase is sent to the light fractioning section.

### 3.1.3 Light fractioning

The light fractioning system consist of four parts; compression, caustic washing tower, propane and butane columns.

The light fraction gas phase from the drum is compressed in three stages and cooled after each stage. A fraction of the hydrocarbons in the compressed gas is liquefied and separated from the hydrocarbon vapour phase.

The hydrocarbon vapour phase is fed to the caustic washing tower, where the acidic gases are removed by sodium hydroxide. This step is also necessary to prevent disturbances in the catalytic reactor and to meet ethylene specifications. The gas is dried before the depropanizer to prevent freezing and hydrates in later steps. In the depropanizer, the hydrocarbon gas is separated from $C_4$ and heavier fractions (bottoms). $C_4$ is then separated from heavier fractions in the debutanizer and stored.

### 3.1.4 Ethylene recovery

The ethylene recovery consists of several compression, reaction and separation steps. The distillate from the depropanizer is compressed and fed to two acetylene reactors to remove acetylene (interstage coolers). From the reactor effluent, methane is separated before enters the deethanizer, where ethane and ethylene are separated from propane and propylene. The product, ethylene, is finally separated from ethane in the ethane/ethylene splitter.

The bottom product from the deethanizer is fed to the propylene reactor (hydrogenation of methyl-acetylene and propadiene). Thereafter, the reactor effluent is fed to the propylene rerun tower to separate propylene and propane from heavier fractions. The propylene and propane are fed to the secondary deethanizer before separated in the propane/propylene splitter to give the second product, propylene.

## 3.2 The HPPE plant (LD5)

The LD5 is the high-pressure plant that is going to phase out the old HPPE plant. The polymerization reaction is strongly exothermic and takes place at relatively high temperatures. By cooling the reaction large amounts of steam is produced. The steam is fed to the utility system and can be used in other processes. The utility system is described in the following section.

## 3.3    Utility system

The utility system consists of a system for energy supply and a cooling system.

### 3.3.1    Energy supply

<u>Steam system</u>

Steam is used as an energy source to supply heat and power to the cracker process and to the PE processes. As mentioned in section 3.1.1 HP steam is produced in the cracking furnaces when the 850 °C product gases are heat exchanged with feed water. Additional HP steam needed in the process is produced in three steam boilers. Currently, the steam boilers operate at minimum load since steam from the LD5 process delivers substantial amounts of heat to the cracker process, enough to satisfy the heating demand. Figure 4 shows the steam utility system.
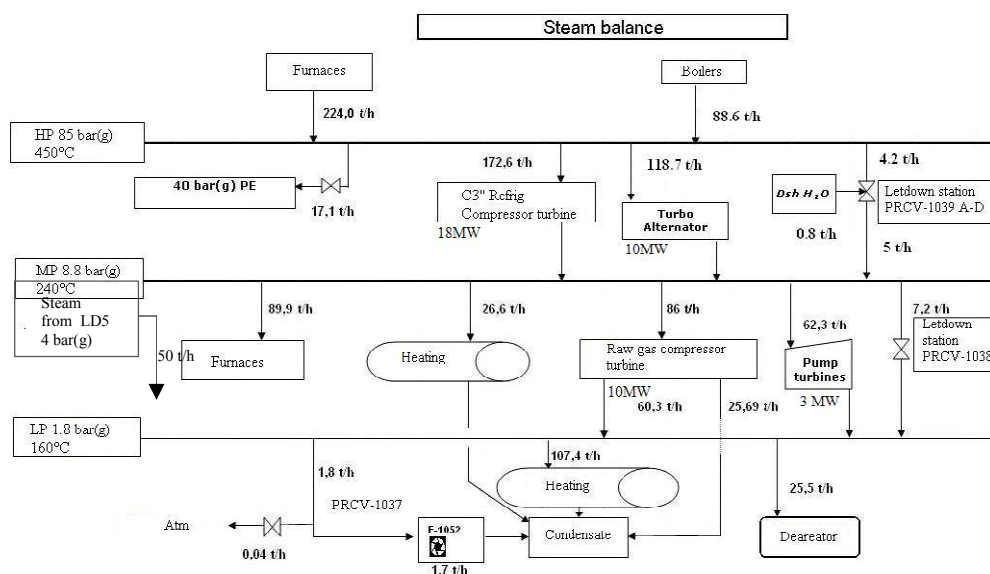


*Figure 4. The figure shows a scheme of the steam system. Intermediate pressure (IP) steam is not shown in the figure. Both the intermediate pressure steam and the LD5 steam are let down to the low-pressure header. The steam production and consumption are not up to date, see Table 3.*

Table 3 shows the steam levels that are represented in Figure 4. The steam production in the LD5 plant is predicted, as it has not yet reached full production capacity.

*Table 3. The table presents steam temperature and current production for the different pressure levels.*

| Steam | Pressure [bar(g)] | Temperature [°C] | Production [ton/h] |
|---|---|---|---|
| High pressure (HP) | 80 | 450-520 | 290 |
| Medium pressure (MP) | 8,8 | 185 | 279 |
| Intermediate pressure (IP) | 2,7 | 141 (saturated) | 20 |
| Low pressure (LP) | 1,8 | 131,4 (saturated) | 146 |

| LD5 steam | ~ 4 | ~150 (saturated) | ~ 50 |
|-----------|-----|------------------|------|

The steam use at each pressure level is presented below.

- HP steam: is not used for process heating. A majority of the steam is expanded in the turbine CT-1951 to produce shaft work for the compressor in the propylene refrigeration system. The steam is also used to drive a turbo alternator that produces electricity. Small amounts of the steam goes to the PE process (expanded to 38 – 40 bar(g)) and through the letdown station to the MP header.

- MP steam: is used in the cracking furnaces after the feed preheater. It is also used in the raw gas compressor turbine CT-1701 and pump turbines to produce shaft work for the compressors and pumps respectively. The steam is also used for process heating e.g. to preheat (E-1609 X) and vaporize (E-1609 Y) the liquefied petroleum gas (LPG) and as heat in the reboilers E-2218 N, E-2218 X and E-2225 N. A small amount goes through the let down station to the LP header.

- IP steam: is produced in the primary fractionating heat recovery section. The steam is mostly used in the debutanizer (T-1704) reboiler and in the distillate stripper (T-1701) reboiler.

- LP steam: is used as the heat source in heat exchangers, tracing and stripping in the deaerators. The steam is generated from backpressure turbines. In addition, steam from the LD5 goes to this header via a valve that reduces the pressure of the steam to 1.8 bar(g).

Electricity system

The electricity consumption for the cracker plant is 340 GWh/year. A large part of the consumption goes to the two stage compressor in the light fractioning section.

Fuel

The cracker plant consumes about 4.3 TWh fuel/year. The fuel gas has a composition similar to natural gas and is sulphur free. The fuel gas is a by-product from the production of ethylene and is mainly consumed in the cracker furnaces, steam boilers and the hot oil furnace.

### 3.3.2 Cooling system

Water system

Borealis cracker plant uses cooling water from Askeröfjorden. The water inlet temperature varies from 5 °C to 25 °C depending on the time of year. Inlet cooling water has a capacity of 12 000 m³/hour. To prevent fouling and mussels in the system the cooling water is mixed with sodium hypochlorite. Before the cooling water is pumped back to the fjord it has to be treated. The returning water is divided into four categories depending on contamination risk.

1. Water that in case of a leak can be contaminated by gas goes to a tank for degasification and thereafter directly to Askeröfjorden.
2. Water that in case of a leak can be contaminated by gas and oil goes to a separation tank. Thereafter, the water goes to the cooling water channel in the water treatment.
3. Water that in case of a leak can be contaminated by oil goes to the cooling water

channel via the oil separator.

4. The water from heat exchangers in the tank area goes to the tank for degasification.

The treated water from categories 2-4 is mixed and pumped back to the fjord.

<u>Refrigeration system</u>

The refrigeration system consists of ethylene and propylene at three temperature levels each. The temperature levels for the ethylene system are -62 °C, -84 °C and -100 °C. The temperature levels for the propylene system are 9 °C, -21 °C and -40 °C. The refrigerants are mainly used to cool streams in the ethylene recovery and the light fractioning section. Not only are the refrigerants used as a cooling medium, but also as a heating medium e.g. propylene condensation supplies heat to the demethanizer and ethane/ethylene splitter reboilers. Additionally, the refrigeration system is used to preheat the ethane and propane feedstock.

## 3.4    Hot and cold utility demands

The current hot utility demand of the cracker plant is represented by $Q_{H1}$ and the current cold utility demand is represented by $Q_{C1}$. An example of a hot and cold utility is steam and cooling water respectively. Implementation of a carbon capture process results in an increased hot and cold utility consumption and to minimize the consumption process integration becomes essential. This is illustrated with following figures. Figure 5 below gives an overview of the hot and cold utility demand in the cracker and a future carbon capture process that is not integrated.
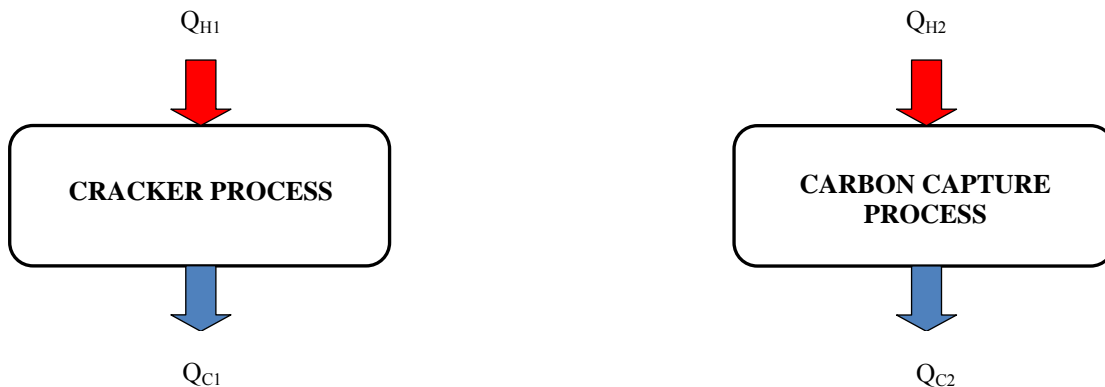


*Figure 5. Schematic figure to show the cracker hot and cold utility and the unknown hot and cold utility of the carbon capture process.*

In a case where a carbon capture process is taken into operation the total utility demand will increase, since the process requires heating and cooling.

*Total hot utility demand = $Q_{H1}$ + $Q_{H2}$*

*Total cold utility demand = $Q_{C1}$ + $Q_{C2}$*

To minimize the utility usage related to the carbon capture process, not only does internal heat exchanging become essential, but also heat exchanging with the cracker process. Figure 6 illustrates a heat exchange possibility between the processes, where e.g excess heat from the carbon capture process can be used in the cracker process.
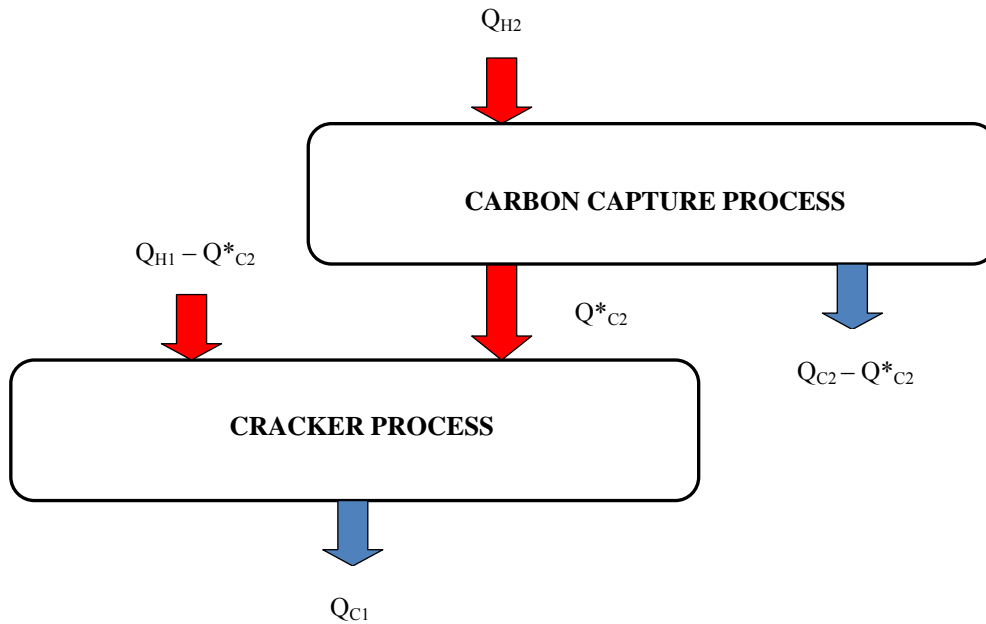
*Figure 6. An example of how process integration can be executed.*

Utilizing excess heat from the carbon capture process can reduce the total hot and cold utility demands. The combined utility demands therefore become,

*Hot utility demand = $(Q_{H1} - Q*_{C2}) + Q_{H2}$*

*Cold utility demand = $Q_{C1} + (Q_{C2} - Q*_{C2})$*

From an energy saving point of view, maximizing $Q*_{C2}$ is the most optimal solution of process integration. As the hot utility demand (e.g. steam) in the cracker process decreases with $Q*_{C2}$ it might be possible to supply $Q*_{C2}$ to the carbon capture process, if the saved hot utility is at an appropriate temperature. If this integration would be possible, the hot utility demand in the cracker process decreases with $Q*_{C2}$. If the heat $Q*_{C2}$ is at an appropriately high temperature it might be possible to supply $Q*_{C2}$ to the carbon capture process. Figure 6 is an example of integration between the cracker and carbon capture process. It can be possible to supply excess heat from the cracker process to the carbon capture process. Figure 6 is only one example of integration possibilities. Excess heat from the cracker process might be utilized in the carbon capture process as well. The integration possibilities are examined with pinch analysis, which is described in more detail in section 5.3.

# 4 Carbon Capture Process

## 4.1 Background

Post-combustion $CO_2$ capture that utilizes MEA as the absorbent is well developed and commercially available since it is generally relatively simple to implement this technology for retrofitting existing processes. In addition to the new ground area needed for the capture plant, piping work between the stack and the capture plant is necessary to redirect the flue gases into the capture plant. Although, the latter is the only process modification, it is important to minimize the distance between the existing process and the capture process. Large volumetric flue gas flows require large pipe diameters, which become costly when routed over long distance.

### 4.1.1 Mono-ethanolamine (MEA)

MEA is an organic compound with the molecular formula $C_2H_7NO$. The molecular structure is $H_2N$-$CH_2$-$CH_2$-$OH$. It is formed when ethylene oxide reacts with aqueous ammonia.

The reason behind this choice of solvent is the ability of forming strong chemical bonds to $CO_2$, which allows low partial pressures of $CO_2$ in the flue gas. The physically absorbed $CO_2$ reacts according to the reactions R1 and R2. The chemical absorption reactions that take place in the liquid phase are presented below R refers to $HO$-$CH_2$-$CH_2$. (Hassan, 2005).

$$CO_2 + 2\ RNH_2 \leftrightarrow RNH_3^+ + RNHCOO^- \qquad (R1)$$

$$CO_2 + RNHCOO^- + 2\ H_2O \leftrightarrow RNH_3^+ + 2HCO_3^- \qquad (R2)$$
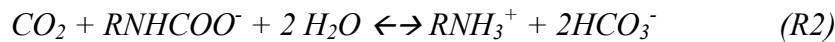
The forward reaction in R1 is the exothermic absorption reaction and the backward reaction is the endothermic solvent regeneration reaction that takes place in the desorber. The forward reaction is favoured by high pressures and low temperature, and the backward reaction is favoured by low pressures and high temperatures. The absorption reaction R2 is neglegible if the gas/liquid exposure time is short. Reaction R1 is a total of two reactions, R3 and R4. (Hassan, 2005)

$$CO_2 + RNH_2 \leftrightarrow H^+ + RNHCOO^- \qquad (R3)$$

$$H^+ + RNH_2 \leftrightarrow RNH_3^+ \qquad (R4)$$

Reaction R3 is a first order reaction with respect to carbon dioxide and MEA i.e. a second order reaction. Since reaction R4 is an ionic instanteneous reaction, R3 is the rate determining step. (Hassan, 2005)

## 4.2 MEA absorption plant proposed by Nexant

The process description is based on a carbon capture plant proposed by Nexant (Choi, 2004) and the process is described in detail by Tel-Tek (Haugen, 2009). The flue gases are generated in a gas fired power plant. The inlet flow rate of $CO_2$ is 70 ton/h, where 60 ton $CO_2$/h is captured. The flue gases enter the carbon capture plant at 80 °C and 1.02 bar at a flow rate of 1128 ton/h. Additional flue gas data and composition are presented in Table 4.

*Table 4. Data for the flue gases from the gas fired power plant with gas turbines.*

| Composition | Mole fraction |
|---|---|
| $N_2$ | 0,7528 |
| $CO_2$ | 0,0398 |
| $O_2$ | 0,124 |
| $H_2O$ | 0,0834 |
| **Other** | **Value** |
| Molar flowrate [kmol/h] | 39832 |
| Mass flowrate [ton/h] | 1128 |
| Molecular weight [g/mol] | 28,3 |

The process can be divided into three parts; flue gas pretreatment, $CO_2$ absorption (absorber) and MEA regeneration (desorber) and finally $CO_2$ compression. In the first step the flue gases are treated to reduce the $NO_x$ and $SO_x$ content to the restricted levels (presented in section 4.2.1). The reason is to minimize the consumption of MEA by reactions with $NO_x$ and $SO_x$. However, the $NO_x$ concentration in the flue gases from Borealis cracker furnaces already fulfils the concentration limit. In addition, the flue gases are free from $SO_x$ since the fuel gas combusted in the cracker furnaces does not contain any sulphur. Consequently, flue gas desulfurization and further $NO_x$ treatment can be excluded in this thesis. The MEA absorption process is shown in Figure 7 (Pfaff, 2010). A more detailed flowsheet is presented in Appendix 3.
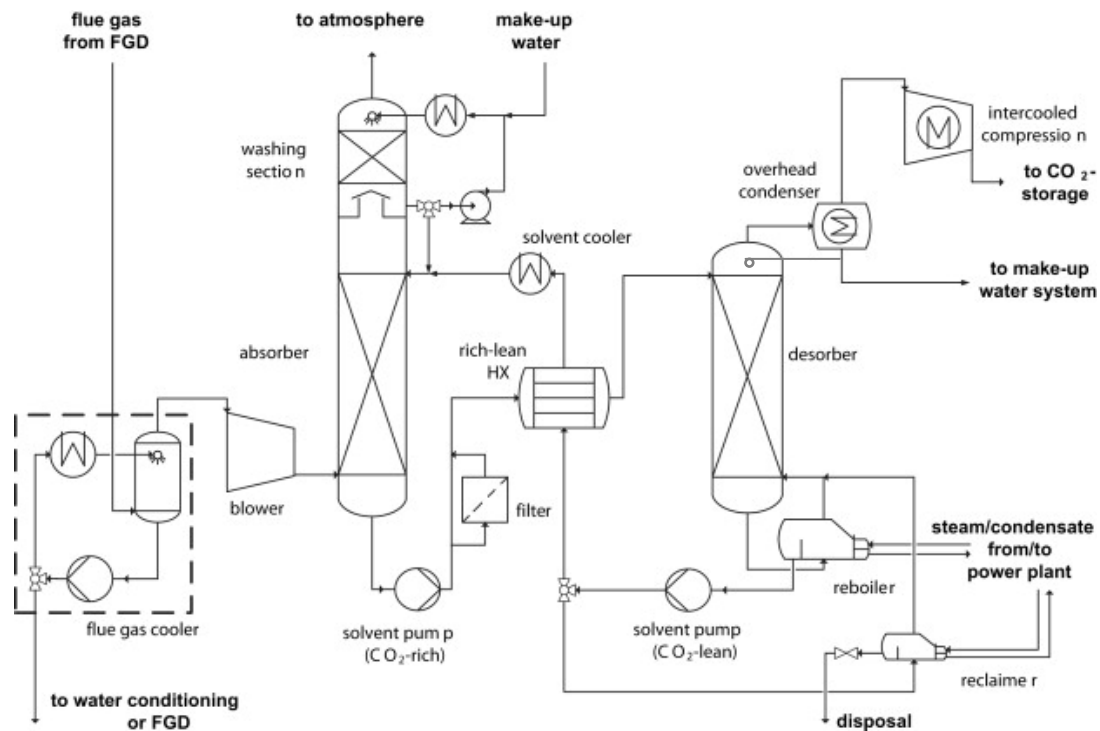


*Figure 7. Schematic flowsheet of the Nexant carbon capture process. The flue gases enter the system from the cracker furnaces. The desorber also consists of a washing section (above the feed inlet), which is not shown in the figure.*

## Cooling and compression

The flue gases are cooled to 47 °C prior to the absorption column with water in a direct contact cooler (DCC) to enhance the $CO_2$ capture efficiency. The volumetric flue gas flow rate increases after the DCC, since water evaporates in contact with the hot, unsaturated flue gases.

Thereafter, a blower increases the pressure of the flue gas in order to compensate for the pressure drop in the column. The blower power consumption is approximately 3 MW to increase the flue gas pressure by 0.09 bar. Reducing the flue gas volumetric flow rate before the blower is essential to reduce the power consumption.

## Absorption

The absorption column consists of two packed beds. One is used to absorb the $CO_2$ from the flue gases (lower bed) and the other to wash the flue gases from solvent (upper bed). The column operates at atmospheric pressure and the partial pressure of $CO_2$ is low, which motivates the choice of MEA. The flue gases are fed at the bottom of the column and the aqueous MEA solution ($CO_2$-lean solution) is fed between the sections. As the flue gases flow upwards through the lower packed bed, countercurrent to the aqueous MEA solution, the $CO_2$ is absorbed into the aqueous MEA solution. After the $CO_2$ absorption the gases enter the upper packed bed. Due to the exothermic reaction the temperature of the solution increases, resulting in evaporation of the solvent. Consequently, the treated flue gases have to be cleaned from evaporated MEA in the upper packed bed before being released to the atmosphere. This water wash section prevents high emissions of MEA by injecting water above the upper packed bed that flows down in contact to the flue gases. The top product is flue gases cleaned from $CO_2$ and the bottom product is a $CO_2$-rich solution that consists of MEA, $H_2O$ and absorbed $CO_2$ respectively. A $CO_2$-rich solution is a solution with high concentration of $CO_2$ compared to the $CO_2$-lean solution.

## Regeneration of MEA

The rich solution is then pumped and preheated before it enters the desorber where the solvent is regenerated. The $CO_2$ is desorbed from the rich solution by supplying energy i.e. steam. As in the absorber, the desorber also consists of two packed beds; MEA regeneration (lower bed) and water wash (upper bed). The feed enters between the sections and flows down in the column as vapour stripps off the $CO_2$ from the rich solution. This vapour is generated from the $CO_2$-lean solution (contains a small fraction of $CO_2$) at 123 °C in the reboiler. The regeneration is the most energy-consuming step in the process as about 79 MW (4740 kJ/kg $CO_2$ captured) of heat is required in the reboiler. The bottom product (lean solution) is recirculated to the absorber via a number of cooling steps. Also, it is necessary to supply water and MEA to cover up the losses in the process. The top product, a $CO_2$-$H_2O$ mixture that contains 98 mol% $CO_2$, goes to the $CO_2$ compression. The reflux to the column from the partial condenser is condensed water, which is used in the desorber water wash to avoid MEA in the top product. The condenser duty is approximately 41 MW (2450 kJ/kg $CO_2$ captured).

The desorber column is the most energy consuming unit, compared to others. The reboiler duty shows that it requires 79 MW of hot utility, i.e. steam, to evaporate the

lean solution at 123 °C. The difference between the reboiler and condenser duty is mainly due to the endothermic character of the desorption reaction.

CO_2 compression to liquefaction prior to transportation

To ease transportation of $CO_2$ to the storage location, compression of the gas mixture is necessary. The gas mixture (98 mol% $CO_2$) is compressed towards the critical point of pure $CO_2$ i.e. that temperature and pressure where pure carbon dioxide is in a liquid state. The critical point is at 31.1 °C and 73.83 bar, whereas the mixture that is transported is at 25 °C and 75 bar. Figure 8 shows the phase diagram of pure carbon dioxide.
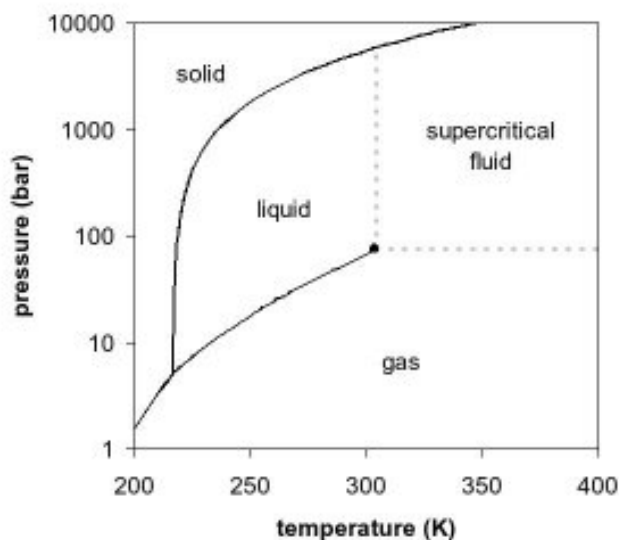


*Figure 8. Phase diagram of pure carbon dioxide. The critical point is marked in the figure. The pressure scale is logarithmic.*

The 98 mol% $CO_2$-$H_2O$ stream is compressed in three stages, with cooling between each stage, to a pressure at approximately 75 bar. Between the compression stages water is condensed and separated. Before the last compression stage there is also a dryer that removes most of the remaining moist (not present in Figure 7). As a result, the multistage effluent is > 99 mol% pure liquid with respect to $CO_2$ (Chakravarti, 2001). This stream is then pumped to 150 bar and transported for storage.

## 4.2.1 Disadvantages and limitations of the MEA absorption process

Even though this process is the most developed compared to other $CO_2$ capture processes, there are a number of challenges that has to be studied in detail to identify the consequences and the solutions. The significant challenges are:

- Treating flue gases from $NO_x$ and $SO_x$. If the flue gases contain $NO_x$ and $SO_x$ they will react with MEA resulting in an undesirably high consumption of the solvent. Therefore, the flue gases have to be pre-treated to reach the concentration limits of $NO_x$ and $SO_x$; 100 ppm $NO_x$ (maximum 10 ppm $NO_2$) and 5 ppm $SO_x$ respectively. In addition, flue gas may contain dust that can cause blockage problems; hence the concentration should not exceed 10 mg dust/$Nm^3$ flue gas. The $NO_x$ content in the flue gases generated in the cracker furnaces is 37 ppm (Borealis AB, 2007). Combustion of the fuel gas (mixture of $CH_4$ and $H_2$) in the

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furnaces does not generate significant amounts of $SO_x$ and dust and can be considered negligible in the flue gases (Borealis AB, 2007).

- Reduce the energy cost and the external energy supply (e.g. steam) of the process and especially in the solvent regeneration step (reboiler). It is difficult to reduce the energy demand since the desorption reaction is endothermic, which means that there is a minimum energy demand for the reaction. One possible solution to reduce the energy cost and the external energy supply is through process integration, which is studied in this thesis.

- Even though MEA is recycled in the process, small quantities are emitted from the absorber to the atmosphere. In the atmosphere, MEA reacts and forms a number of degradation products (e.g. aldehydes and amides) that might be toxic and carcinogenic (Shao, 2009). Therefore, it is essential to evaluate the environmental and human impacts caused by degradation products (Veltman, 2010). In this thesis the MEA concentration in the flue gases leaving the carbon capture plant is 5 ppm.

- Process water has to be treated from MEA and its degradation products. The separation of such species is handled in the reclaimer, which is the source of the main amine waste. This waste is classified as hazardous waste (Shao, 2009).

- The reaction between $CO_2$ and MEA generates corrosive solutions. The corrosion rate depends on several factors such as the total MEA concentration, the $CO_2$ loading (defined as mole $CO_2$ per mole MEA) and solution temperature. The corrosion rate increases with solution temperature i.e. a high temperature region give rise to a high corrosion rate and vice versa. Relating this to the process, the corrosion rate in the absorber is low compared to the corrosion rate in the desorber. In addition, a high temperature in combination with a high $CO_2$ loading leads to additional corrosion. To further reduce the corrosion rate the total MEA concentration should not exceed 30 % and the $CO_2$ loading in the lean solution (mole $CO_2$ per mole MEA) should not exceed 0.5 (Chakravarti, 2001).

It is important to keep in mind that the waste and emission quantities of MEA and its degradation products depend on the flue gas composition. Solutions to the problems mentioned above are not to full extent established. Considering the species in the flue gases these can be reduced below the limits by conventional $NO_x$ and $SO_x$ treatments. This may require retrofits of the existing treatment plants to achieve this. Considering the corrosion, other materials can be used in the process where corrosion might be severe (e.g. stainless steel) to reduce corrosion, even though it cannot be completely avoided. MEA is the most corrosive amine, so other amines should be studied (Shao, 2009). Finally, it is necessary to improve separation techniques to decrease the quantity of amine waste to air and water to reduce the environmental and human impacts.

### 4.2.2 Summary of Tel-Tek study – carbon capture plant coupled to Borealis cracker plant

The Tel-Tek study made at Borealis in Stenungsund investigated the possibility of implementing a carbon capture process with MEA absorption in the cracker plant. The carbon capture process is based on the process proposed by Nexant (Choi, 2004) (Appendix 3). No process integration was studied and, therefore a natural gas fired steam boiler was proposed to supply heat to the carbon capture process. In addition, two steam turbines were proposed to drive the blower and the compressors in the $CO_2$

compression with shaft power. HP steam is produced in the steam boiler and expanded in the turbines to deliver heat and shaft power to the carbon capture process. This can be related to Figure 5 in section 3.4. The capital expenditures (CAPEX) and operational expenditures (OPEX) are determined after scaling of the MEA absorption process proposed by Nexant. The process is scaled with respect to flue gas flow before the absorber, and with respect to $CO_2$ flow downstream the absorber (Eldrup, 2010). The work flowsheet for the Tel-Tek study is shown in Figure 9.

```
┌─────────────────────┐        ┌─────────────────────┐
│  Steam boiler and   │        │ Nexant process with │
│   steam turbines    │        │   plant specific    │
│                     │        │        data         │
└─────────────────────┘        └─────────────────────┘
           │                              │
           ▼                              ▼
┌──────────────────────────────────────────────────────┐
│ Scaling of the process to design it for the cracker   │
│     flue gases and to meet the energy requirements    │
└──────────────────────────────────────────────────────┘
                          │
                          ▼
               ┌─────────────────────┐
               │    CAPEX, OPEX      │
               └─────────────────────┘
```
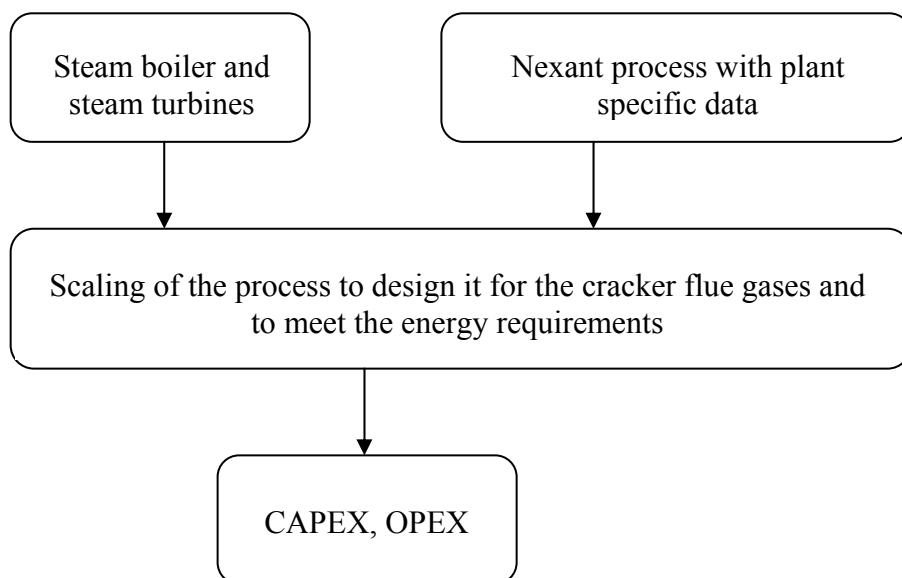
*Figure 9. Tel-Tek study work flowsheet.*

The total capital investment (CAPEX) became 114.5 M€ and total operational costs (OPEX) were 24 M€/year. The electricity and fuel consumption and other detailed costs related to investment and operating costs are shown in Appendix 2 (Eldrup, 2008B). The costs are converted to Euros 2010 using data in Appendix 8. The net capture cost was 80 €/ton $CO_2$ and the calculation can be seen more in detail in section 7.1.1. This cost was also converted to Euros 2010 using data in Appendix 8.

### 4.2.3    Comparison between the Tel-Tek study and this thesis

In the Tel-Tek study, the flue gases generated in the cracker furnaces and the flue gases generated in the invested steam boiler goes to the carbon capture plant. Thus, a higher flue gas flow rate is treated and larger quantity of $CO_2$ is captured in the Tel-Tek study compared to this thesis, 70 ton/h and 53.5 ton/h respectively (Eldrup, 2008B). The work flowsheet for this study is presented in Figure 10 below.
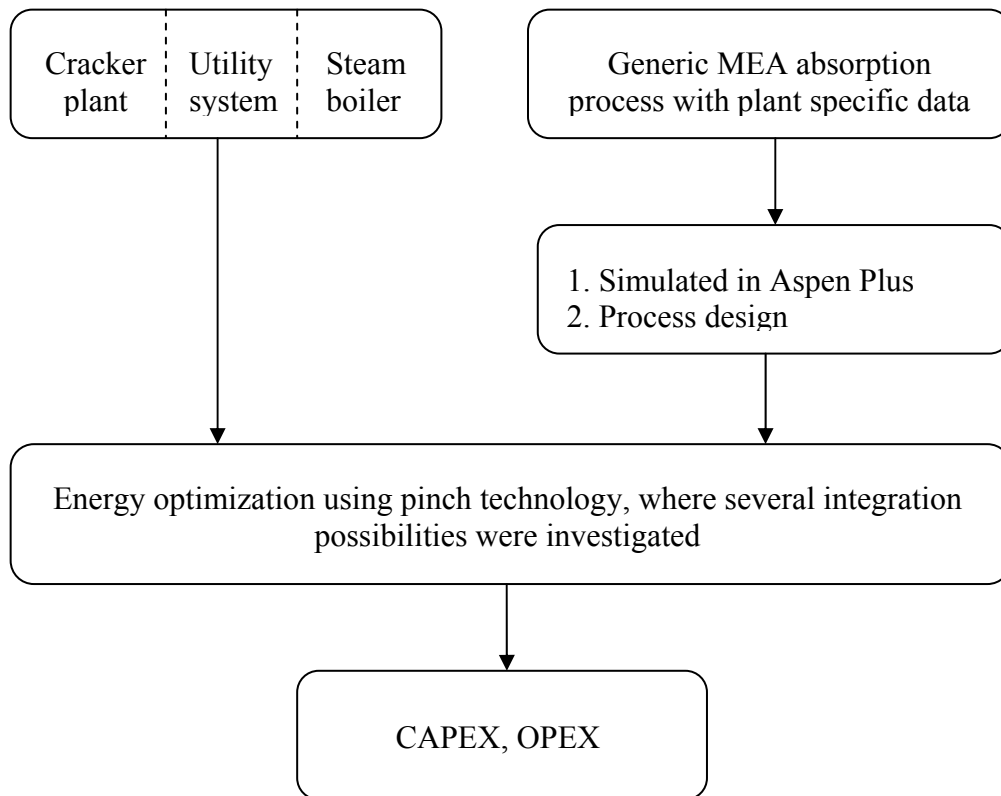
*Figure 10. Work flowsheet of this thesis.*

As mentioned previously, process integration is studied in this thesis. This can be related to Figure 6 in section 3.4. The aim of the integration is to supply enough heat to the carbon capture plant and avoid an additional steam boiler used in the Tel-Tek study. The procedure to calculate CAPEX and OPEX in this thesis is based on Aspen Plus simulations, instead of scaling. The simulation was developed from the Nexant proposed MEA absorption process to find operating conditions designed for the cracker flue gases. Thereafter, the process was energy optimized with process integration and the CAPEX and OPEX were calculated and compared to the results from the Tel-Tek study.

# 5    Methodology

## 5.1    Overview

The methodology was carried out using the simulation tool Aspen Plus® and pinch technology. The former is used to determine the energy demand and equipment sizes in the process and the latter is used to establish the possibilities of process integration. The study was carried out using the four cases presented below (all simulated in Aspen Plus®).

**Base Case**
- Simulation of the process used in the Tel-Tek study (section 4.2).
- Used to verify the Aspen Plus simulation model.

**Case 1**
- Simulation carried out using data from flue gases generated in Borealis' cracker furnaces. Equipment data are the same as the base case.

**Case 2**
- Investigates the possibility of condensing the water in the cracker flue gases.

**Case 3**
- Based on case 1 and 2, but it is optimized with respect to energy and economy by changing absorber and desorber dimensions.

**Case 4**
- Case 4 investigates the possibilities of heat pump implementation in the desorber for Case 3.

*Figure 11. The studied and simulated cases.*

To establish the integration possibilities a pinch analysis was performed for Case 3 and 4. These cases are further described in section 5.2.2.

## 5.2    Aspen Plus® simulations

To determine the energy demands in the capture process a simulation in Aspen Plus® was conducted. Detailed information of how the simulations were conducted is presented in Appendix 1.

### 5.2.1    Assumptions

Following assumptions are made to reduce the complexity of the simulations.

- The flue gases do not contain any $NO_x$ and $SO_x$. The main species entering the carbon capture process are $N_2$, $H_2O$, $O_2$ and $CO_2$.
- No reaction occurs between MEA and oxygen in the flue gas.
- Corrosion is not considered in the simulations.
- The dryer in the $CO_2$ compression is not included.

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- Amine losses are assumed to be to atmosphere only i.e. no by-product formation from the amine (reclaimer is not simulated).

## 5.2.2 Process simulations

The base case process has then been stepwise modified to come up with a process that suits the flue gases from the cracker process at Borealis. Moreover, improvements have been suggested and studied to optimize the process in terms of energy. An overview of the simulation steps is presented below.

Verification of the carbon capture process proposed by Nexant

The base case process has been simulated in Aspen Hysys, from which stream and equipment data was available. These data were used to validate the Aspen Plus simulation model.
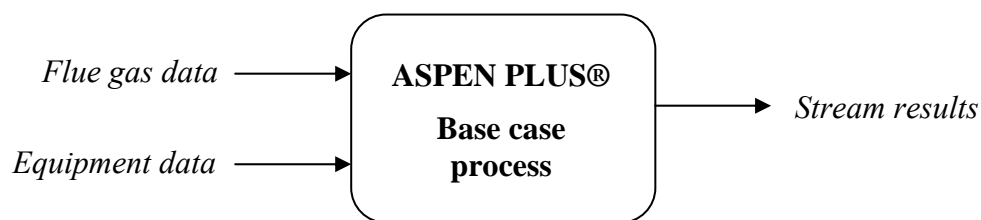


*Figure 12. In the process reconstruction case both input and output data were known. Flue gas data are assumed to be from a gas fired power plant.*

The same process was simulated in Aspen Plus® and the results of both studies are compared to determine if there were large deviations between Aspen Plus® and Aspen Hysys. As all data were available it was easy to compare results from both simulation tools. The base case process was also developed in Aspen Plus® to act as a basis for the cases below. The inlet flue gas temperature was assumed to be 80 °C.

Case 1 – Borealis flue gas data

After successfully completing the process verification simulation in Aspen Plus®, the next step was to simulate the process adjusted to include specific Borealis flue gas data. The first case uses the cracker flue gas data as input. The equipment data were the same as for the process reconstruction case.
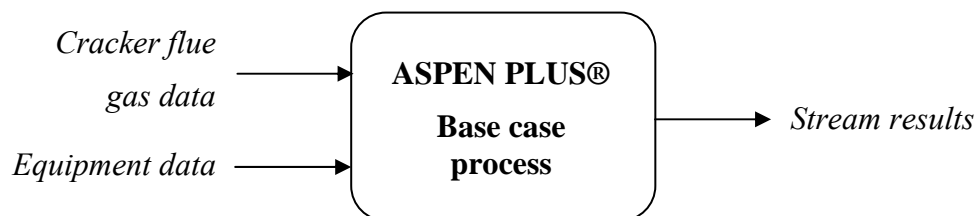


*Figure 13. In this case the cracker flue gas data were used, together with equipment data from the conventional process.*

Equipment size and specifications in the process remained unchanged and taken from the process presented in section 4.2. The flowsheet constructed in this case is developed and improved in the remaining cases. The inlet flue gas temperature is 144 °C.

<u>Case 2 – Flue gas water condensation</u>

Case 2 is based on the previous case. The cracker flue gases contain 21.1 mol% $H_2O$, compared to 8.34 mol% in the base case. This case is considered to show that it is possible to condense and separate the water from the flue gases. The water could be used as process water since Borealis is reaching its maximum authorized consumption level of process water (fresh water) consumption. To utilize the water, the flue gas cooling section in the carbon capture process has been modified in this case.

<u>Case 3 – Process design</u>

This case is based on case 1 and case 2. These cases have been combined and the absorber and desorber are designed in this case by varying column dimensions to fit for the cracker flue gases.

*Cracker flue*
*gas data* → **ASPEN PLUS®**
**Base case process**
**optimized** → *Equipment and*
*stream results*
*Equipment data* →
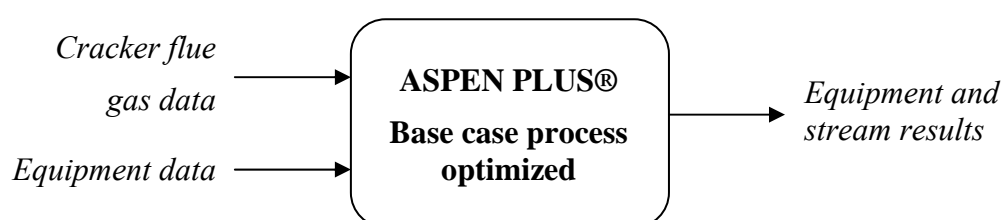
*Figure 14. The equipment data input and output were unknown.*

Two variables were varied to find suitable absorber and desorber dimensions. The absorber dimension varies depending of the number sections of packed bed. The number of sections, on the other hand, depends on the ability of the solvent (lean solution) to absorb the $CO_2$ in the flue gases. Thus, the second variable is the $CO_2$ loading in the lean solution (mole $CO_2$ per mole MEA), named α.

<u>Case 4 – Mechanical Vapour Recompression (MVR)</u>

A MVR heat pump could be implemented to save energy in such a way that heat at low temperature is lifted and supplied at a higher temperature. The use of a MVR in combination with the desorber was investigated to reduce the external hot utility supply to the reboiler. This might be possible if the top vapour stream from the desorber is compressed to lift the temperature above the reboiler temperature. The vapour stream can then be heat exchanged in the reboiler. The remaining amount of energy is supplied by steam.

## 5.2.3 Process integration

The process integration between the carbon capture process and other processes at Borealis was conducted according to the illustration and description below.
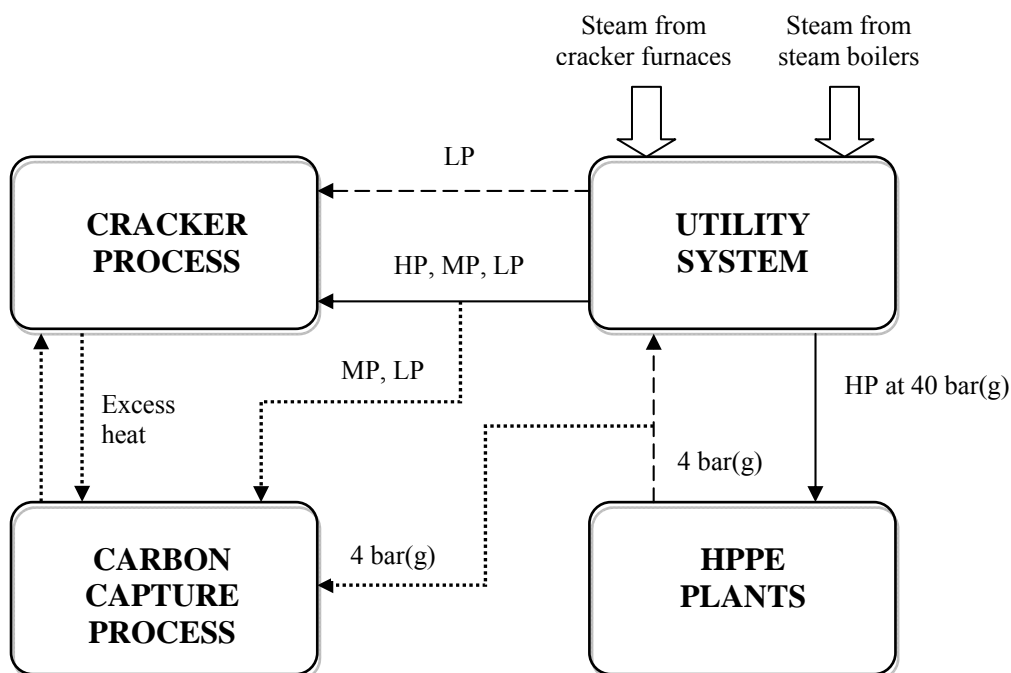
*Figure 15. The figure shows the systems involved in the process integration and the energy flow in between them. The HP steam is not used as heat in the process, rather to produce work required in the cracker process.*

The *solid lines* shows the LP and MP steam from the utility system currently used to satisfy the heat demand in the cracker process. Also, as can be seen in Figure 4 steam from the LD5 process at 4 bar(g) is currently expanded to the LP header and used in the cracker process. The *dashed lines* represent this steam. To satisfy the energy demands in the carbon capture process a number of options are investigated, shown by the *dotted lines*. First, the integration possibilities between the cracker and the carbon capture process will be investigated utilizing excess heat from the processes as mentioned in section 3.4. Secondly, integration possibilities that can utilize the LD5 steam (at 4 bar(g)) are likely to be assessed. Finally, steam generation in the steam boilers will be studied as a supplementary heat source for a potential integration solution.

## 5.3    Pinch analysis

The carbon capture process with MEA is an energy demanding process. Therefore, efficient use of energy is essential to reduce the operating costs. Pinch analysis is a tool used to determine the minimum external hot and cold utility in a process (heaters and coolers respectively). To begin with, an overview of all process streams must be compiled to be able to select those streams that are either heated or cooled and evaporated or condensed. Stream data of importance are start and target temperatures and heating/cooling demand. Streams with a heat surplus (needs cooling) are hot streams, whereas streams with a heat deficit (needs heating) are cold streams. An example of a hot stream is the gas product from the cracker at ~850 °C that has to be cooled to 200°C. The cracker gas is first cooled by heat exchanging with saturated

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boiler feed water that evaporates. Then, the cracker gases are quenched i.e. oil is injected to cool the gases. The oil evaporates in the system and condenses in downstream units in the process.

Considering above example, any hot stream at a high temperature can be used as an energy source for a cold stream at a lower temperature. From a pinch analysis it is possible to identify the maximum amount of heat that can be exchanged between hot and cold streams internally in a process. The additional heat needed (to heat cold streams) in a process that cannot be satisfied by internal heat exchange must be supplied from an external heat source (e.g. steam). Thus, the minimum hot utility demand is defined. The amount of heat that has to be removed from the hot streams (cooling hot streams) in the process that cannot be satisfied by internal heat exchange must be removed using an external cooling medium (e.g. cooling water). Thus, the minimum cold utility is defined. The temperature at which there is a heating demand above and a cooling demand below is defined as the pinch temperature. This will be illustrated and described later in this section. To obtain the minimum external hot and cold utility three "golden" rules should be respected.

1. Do not heat with external heaters below the pinch (there is a heat surplus below the pinch)
2. Do not cool with external coolers above the pinch (there is a heat deficit above the pinch)
3. Do not transfer heat through the pinch

If any of these rules are broken a so called "pinch violation" is committed. The consequences of the pinch violations are.

1. Increased cooling demand (not taken advantage of the heat surplus below the pinch)
2. Increased heating demand (not taken advantage of the heat deficit below pinch)
3. Increased heating and cooling demand

To perform a pinch analysis following steps are usually performed.

- Stream system definition and data collection
- Define hot and cold streams i.e. start and target temperatures, mass flow and heat capacity
- Set an appropriate minimum temperature difference, $\Delta T_{min}$, for heat exchange
- Construct composite curves (CC) to determine the minimum external hot and cold utility (and the maximum internal heat exchange)
- Construct a grand composite curve (GCC) for the background process
- Construct a GCC for the foreground process
- Make a background/foreground analysis to establish the integration possibilities for the carbon capture process and the cracker.
- Calculate the new minimum heating and cooling demands after process integration
- Design a heat exchanger network for maximum energy recovery

Composite curves (CC)

To establish the maximum internal heat exchange between the hot and cold streams in the process CC are constructed. Given start and target temperature of a hot or cold stream together with known flow rate (F) and heat capacity (Cp) the heat load (Q) of that stream can be calculated. To construct the hot composite curve, the hot streams are divided in temperature ranges. In this temperature range the slope and width of the curve is given from the $\sum F \cdot Cp$ and the $\sum Q$ respectively. In other words, there is at least one stream that influences the slope and width of the curve in a certain temperature range. In a similar way the cold composite curve is constructed. An example of how to construct a cold composite curve using two streams is shown below.

An example of a CC is illustrated in Figure 5.
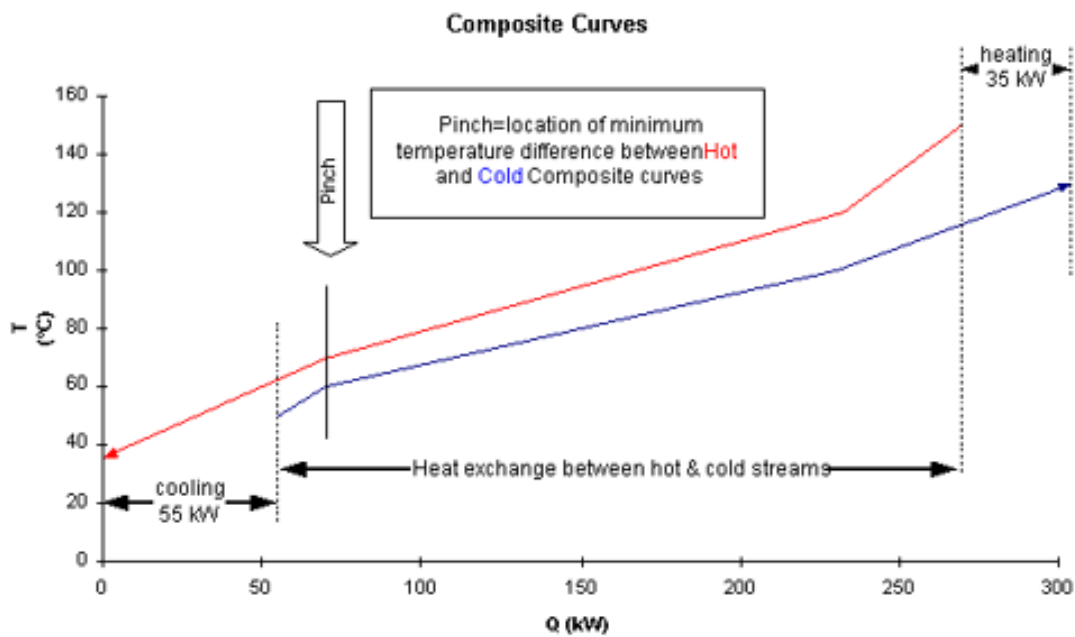


*Figure 16. The CC shows the minimum external heating and cooling demand of the process, 35 kW and 55 kW respectively. Considering energy consumption, a perfectly constructed heat exchanger network would result in these minimum heating and cooling demands. However, this might not be the most economical solution.*

The maximum internal heat exchange can be observed from the overlapping curves. The limitation of this overlap is the $\Delta T_{min}$. The maximum overlap is therefore when the cold CC is shifted to the left below the hot CC until $\Delta T_{min}$ is reached (the pinch). Therefore, a smaller $\Delta T_{min}$ implies more internal heat exchange compared to a larger $\Delta T_{min}$ that results in increased hot and cold utility. On the other hand, the driving force will be smaller and heat exchangers become larger i.e. the investment costs increases.

Grand composite curve (GCC)

The GCC is used to study the net heat content of a process at a certain temperature range. It is used to identify at what temperatures hot and cold utility are desired in the process. It is also possible to identify a pinch point (if the system is not unpinched) at the point where the heat content is zero (i.e. Q = 0 kW). As mentioned there is a heat deficit above the pinch and a heat surplus below. In a GCC the pinch point (L in

Figure 17) separates these parts. Therefore, a hot utility e.g. steam has to be supplied above the pinch as illustrated in the GCC. In similar way cold utilities e.g. cooling water or refrigeration system has to remove heat from the process below the pinch. Note that Figure 17 does not give a realistic example when it comes to LP steam generation or the highest steam level. LP steam generation below the pinch at 100 °C requires vacuum conditions and the highest steam level requires saturated steam at approximately 130 bar to get condensation at 330 °C, which both are not common in the process industry.
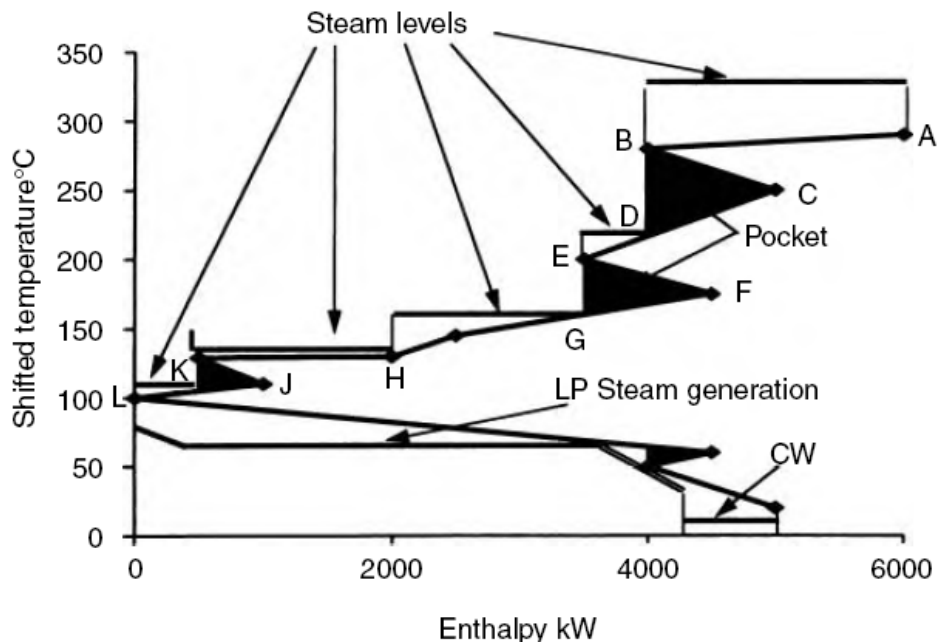


*Figure 17. The GCC represented together with steam levels above the pinch (L), LP steam generation and cooling water levels below the pinch. The steam levels correspond to the minimum hot utility and the LP steam generation and cooling water represent the minimum cold utility.*

The GCC makes it possible to tell whether there is a heat deficit or surplus in a temperature range from the slope of the curve. A positive slope is an indication of a heat deficit in the process and a negative slope indicates a heat surplus. Therefore, pockets can be formed as shown in the figure (e.g. B-C-D). This shows that heat exchange in the process is possible. In addition, condensing steam is represented as a horizontal line.

Finally, the concept of interval temperatures is used in the GCC. For every hot stream ½ $\Delta T_{min}$ is subtracted and for every cold stream ½ $\Delta T_{min}$ is added. Thus, in the GCC the minimum temperature difference will be zero in interval temperatures. A GCC is unique for a specified value of $\Delta T_{min}$.

Background/Foreground (BF) analysis

A BF analysis is conducted to find solutions for process integration. Firstly, a GCC for the background process (in this case: carbon capture process) is constructed.

Secondly, a GCC for the foreground process (in this case: selected streams from the cracker process) is constructed. Finally, the GCC: s are plotted in the same diagram.

For further reading in pinch analysis the reader is referred to Kemp Ian C. "Pinch analysis and process integration".

<u>ProPi</u>

A brief introduction to pinch analysis has now been presented. The pinch analysis was carried out in ProPi, an add-in application in Microsoft Excel developed by Chalmers Industriteknik AB. Input stream data required are

- Start and target temperatures
- Required heating/cooling demand of each stream
- Minimum temperature difference
- Value of the heat transfer coefficient ($kW/m^2K$)

With these inputs, CC and GCC can be constructed in the program. Additional outputs are pinch temperature, minimum hot and cold utility and the quantity of pinch violations. Also, this program is used to make a background/foreground analysis.

# 5.4    Cost evaluation

The cost evaluation is made to make a comparison between the Tel-Tek study and this thesis and to determine if there is profitability with process and energy optimization. The investment and operational costs for the Tel-Tek study have been evaluated in detail and are shown in Appendix 2 (Eldrup, 2008B).

Considering the calculation of the investment cost in this thesis, the sixth-tenths rule has been used. This equation has been used to calculate the investment cost for equipments that make up to a large percentage of the total investment cost given from the Tel-Tek study (see Appendix 2).

$$C_x = C_k \cdot \left[\frac{E_x}{E_k}\right]^{n_c}$$

Where

$C_x$ = cost of equipment item of size $E_x$

$C_k$ = known cost of equipment item of size $E_k$

$n_c$ = cost capacity exponent = 0.6 – 0.7

This equation has been used to calculate the investment cost of following equipments in this thesis: flue gas cooler, blower, absorber, rich-lean heat exchanger, desorber, desorber condenser and reboiler, and finally the compressors in the $CO_2$ compression. The remaining equipments in this thesis are assumed to cost the same as in the Tel-Tek study, since its costs are small compared to the equipments mentioned above. Also, due to the simplified simulated process information about e.g. filters and storage tanks is not available (see Appendix 2). The cost capacity exponent is chosen to 0.7.

Considering the variable operating costs, up to date prices of electricity and fuel have been used (Spetz, 2010). Other variable and fixed costs (except operator cost) have been taken from the Tel-Tek study. The operating costs for the Tel-Tek study are also presented in Appendix 2.

# 6 Results and discussions

The presentation of the results and discussions is based on the base case process (presented in section 4.2). The base case process is further developed in consecutive steps to find the process condition that suits the flue gases generated in the cracker furnaces at Borealis. The results in each case are presented and discussed to motivate the following step.

## 6.1 Base case verification

The flowsheet for the base case verification is shown in the Figure 18 below. The stream results related to the flowsheet are shown in Appendix 4.
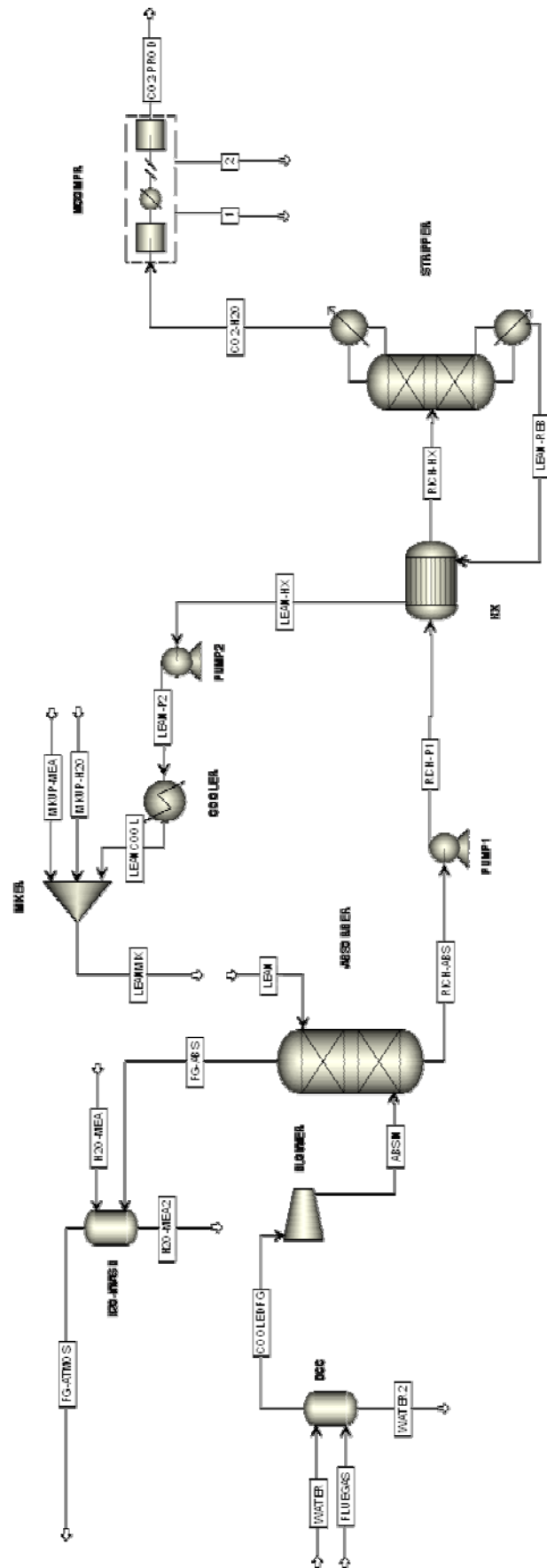
*Figure 18. The flowsheet constructed in Aspen Plus®.*

The flowsheet, Figure 18, represents the MEA carbon capture process proposed by Nexant, shown in section 4.2. The stream LEANMIX and LEAN have the same mass flow rate and composition with respect to water, MEA and $CO_2$. To enhance flowsheet numerical convergence these streams are separated. The simulation results were considered satisfactory and showed good agreement (see deviations in Appendix 4) for the process up until the desorber. The $CO_2$ compression results obtained from the Aspen Plus simulation model showed, however, a high deviation from the Aspen Hysys model. A plausible explanation is the different simulation approach between the simulation tools. More detailed information in regard to deviation and simulation approach is presented in Appendix 4. Though, the constructed model was considered valid to use as a basis for further simulations in the study. For further reading of differences between Aspen Hysys and Aspen Plus the reader is referred to (Øi, 2007).

## 6.2    Case 1 – Borealis flue gas data

In this case the flue gas input data are changed to those generated in the cracker furnaces. A comparison between the flue gases generated in the cracker furnaces (Borealis) and the flue gases generated in the gas fired power plant (Nexant) is shown in Table 5.

*Table 5. Borealis flue gas data in comparison with the flue gas data from the carbon capture process proposed by Nexant.*

|  | *Borealis* | *Nexant* |
|---|---|---|
| **Composition** | **Mole fraction** | **Mole fraction** |
| $N_2$ | 0,719 | 0,7528 |
| $CO_2$ | 0,05 | 0,0398 |
| $O_2$ | 0,02 | 0,124 |
| $H_2O$ | 0,211 | 0,0834 |
| **Other** | **Value** | **Value** |
| Inlet temperature [°C] | 144 | 80 |
| Inlet pressure [bar] | 1,02 | 1,02 |
| Molar flowrate [kmol/h] | 28544 | 39832 |
| Mass flowrate [ton/h] | 765 | 1128 |
| Molecular weight [g/mol] | 26,8 | 28,3 |

The main differences in the flue gas composition between the cases are:

- Higher water fraction in case 1 compared to the base case due to the hydrogen in the fuel gas used in the furnaces.
- Higher oxygen fraction in base case compared to case 1.
- The mass flow rate of flue gases generated in the cracker furnaces is smaller than those generated in the natural gas fired power plant.

The simulated flowsheet and stream data of the process are shown in Appendix 5. To enhance convergence in the simulations the rich-lean heat exchanger (HX in Figure 18) is decomposed and represented by a separate heater on the rich solution stream (62.9 MW). The cooling of the lean solution stream from 123 °C to 38 °C is represented by one cooler only. In other words, the cooling of the lean solution in the rich-lean heat exchanger (HX in Figure 18) and in the downstream lean cooler (COOLER in Figure 18) is modelled as one cooler (76.2 MW). This is also

advantageous in the pinch analysis since streams with heating and cooling demands are separated into cold (needs heating) and a hot streams (needs cooling). Therefore, this will not affect the results related to this thesis. Simulation data and results are shown in Table 6.

*Table 6. Simulation results of relevance.*

| Equipment | | Value |
|---|---|---|
| Absorber | | |
| | Number of absorption sections | 8 |
| | Number of washing sections | 3 |
| | Lean mass flow [ton/h] | 908 |
| | Alfa [mol CO2/mol MEA] | 0,186 |
| Desorber | | |
| | Number of desorption sections | 19 |
| | Number of washing sections | 3 |
| | Condenser duty [MW] | 32,1 |
| | Reboiler duty [MW] | 65,9 |
| Flue gas cooler | | |
| | Duty [MW] | 16 |
| DCC | | |
| | Duty [MW] | 48 |
| Blower | | |
| | Work [MW] | 2,1 |
| Rich heater | | |
| | Duty [MW] | 62,9 |
| Lean cooler | | |
| | Duty [MW] | 76,2 |
| $CO_2$ compression | | |
| | Number of compression stages | 3 |
| | Number of cooling stages | 3 |
| | Total work [MW] | 4,7 |
| | Total cooling duty [MW] | 8,2 |

These results will be used as a reference for the upcoming cases to determine if improvements are made along the way.

### 6.2.1 Discussion of case 1 results

The reboiler duty is 65.9 MW and the condenser duty is 32.1 MW. According to specifications 53.5 ton $CO_2$ per hour must be captured, which corresponds to an energy demand of 4440 kJ/kg $CO_2$ in the reboiler. Corresponding value for the condenser is 2160 kJ/kg $CO_2$.

The flue gases from the cracker furnaces contain 21.1 mol% water compared to 8.34 mol% in the base case. As the water fraction is higher in this case it is possible to condense a significant amount of water from the flue gas. Condensing the water reduces the flue gas volumetric flow rate and consequently the work needed for the blower. In addition, the condensed water can be used as process or cooling water. To determine these possibilities, this is studied in case 2.

## 6.3     Case 2 – Flue gas water condensation

Based on the outcomes in the previous case, this case investigates the configuration of unit operations before the absorber. There are two configurations as shown in Figure 19. In configuration A, the flue gases from the cracker furnaces at 144 °C are cooled in the flue gas cooler (external cooler). The flue gases are further cooled in the direct contact cooler (DCC) to 47 °C with process water. The top stream is cooled flue gas that is compressed in a blower before fed to the absorber. The bottom stream consists of water, mainly condensed water from the flue gases. In configuration B the flue gases are cooled in the flue gas cooler to 47 °C. The water in the flue gases is condensed during the cooling. The vapour and liquid phase is then separated in a splitter (SEP). The top stream is fed to the blower before it goes to the absorber.
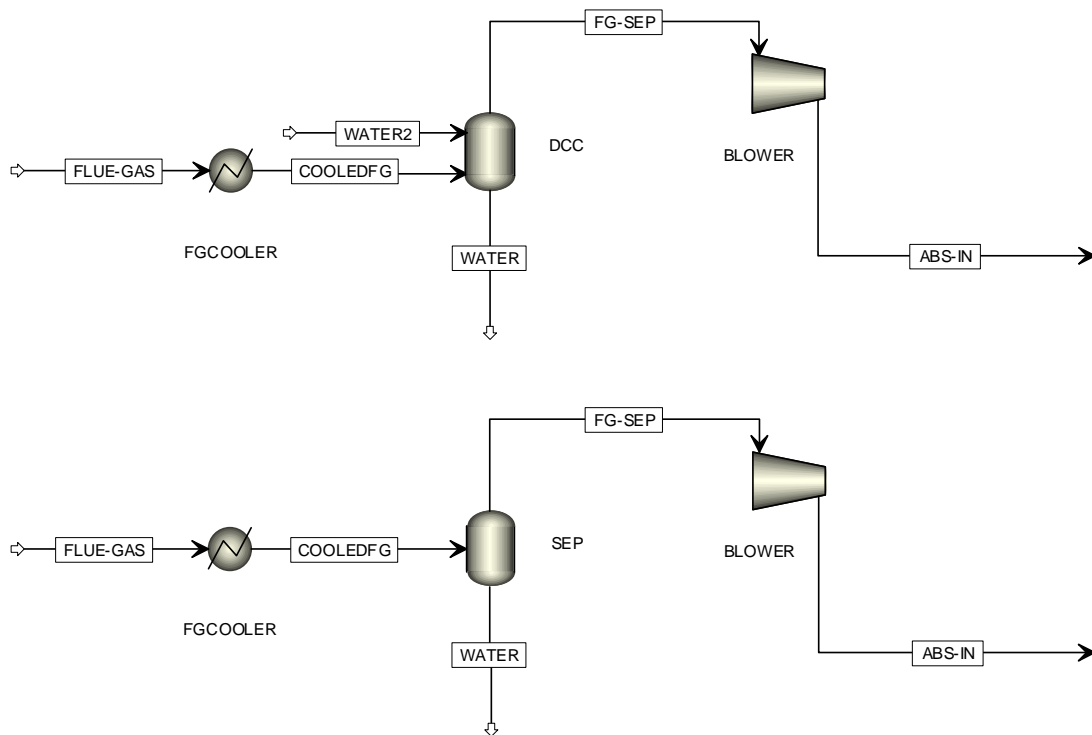


*Figure 19. Above (a): The configuration used in case 1 (configuration A). Below (b): Alternative setup studied in case 2 (configuration B). SEP represents a separator that separates the flue gases and the condensed water.*

The purpose of the flue gas cooler is to decrease the temperature of the flue gas going to the DCC in configuration A to 80 °C (the same flue gas inlet temperature as in the generic case). The differences between the configurations are shown in Table 7.

*Table 7. Stream differences between the configurations.*

| Stream | Config. A - Temp. [°C] | Config. B - Temp. [°C] |
|---|---|---|
| Flue gas | 144 | 144 |
| Cooled flue gas | 80 | 47 |
| Flue gas to blower | 47 | 47 |
| Flue gas to absorber | 57 | 57 |
| **Stream** | **Mass flow [ton/h]** | **Mass flow [ton/h]** |
| Water 2 | 12,77 | 0 |
| Water | 73,62 | 60,85 |
| Cooled flue gas | 764,5 | 764,5 |
| Flue gas to blower | 703,6 | 703,6 |

The water supplied to the DCC cools the flue gases and ends up in the bottom stream (73.62 – 12.77 = 60.85 ton/h). The blower in both configurations consumes 2.1 MW of electricity and increases the flue gas temperature to 57 °C.

### 6.3.1 Discussion of case 2 results

The configurations generate the same results of the stream fed to the blower. The reason is that the flue gases are cooled to 47 °C in both cases before the blower. Considering configuration A, the injected water is evaporated at first since the flue gas is unsaturated. However, as the flue gases are cooled and reach the dew point temperature at 61.6 °C the water condenses. Thereafter, the cooling occurs along a saturation temperature down to 47 °C (shown in the Appendix 6). As a result, both configurations generate the same stream results. However, there are advantages and disadvantages with each configuration. These are presented in Table 8 below.

*Table 8. Advantages and disadvantages with each configuration.*

| | Configuration A | Configuration B |
|---|---|---|
| **Advantages** | Can remove particles from the flue gas. Particles e.g. dust can cause blocking problems in pipes and process equipments. <br><br> More efficient heat exchange. <br><br> Lower pressure drop, which results in a lower operating cost. | No process water consumption. Only cooling water is consumed and it does not get contaminated, which results in a lower operating cost. |
| **Disadvantages** | Unnecessary process water consumption. The contaminated water has to be treated before further use, which results in a higher operating cost. | Cannot remove particles to the same extent as configuration A. <br><br> Less efficient heat exchange, which results in a large heat exchanger area <br><br> Higher pressure drop, which results in a higher operating cost. |

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Flue gas cooling according to configuration A implies lower pressure drop compared to configuration B. As a result, more work has to be supplied to the flue gas fans in the furnace chimneys, but the pressure drop over the flue gas coolers is assumed to be the same in both configurations. The amount of water that is condensed from the flue gases, 60.85 ton/h, make up to approximately 20 % of the total process water (fresh water) consumption at the cracker plant in both configurations (Andersson, 2010). Fresh water is a limited resource especially in Stenungsund, making it important to reduce its consumption (Andersson, 2010). Thus, configuration B is preferred even though configuration A offers more efficient heat exchange and lower pressure drop.

As the absorption reaction is exothermic, decreased temperature of ABS-IN enhances the absorption i.e. the solvent flow rate is decreased. A temperature below 57 °C requires further cooling of the flue gases (increased flue gas cooler duty), but this reduces the flue gas volume because more water would condense. This will also reduce the blower work consumption. However, it is important to remember that the absorber inlet temperature affects the temperature of the flue gases released to the atmosphere. The outlet temperature should be high enough to give the wanted buoyant force of the flue gases. In the conventional process described in section 4.2 the temperature of the vented flue gases is 51 °C. Therefore, in this thesis the target temperature of the flue gases released to the atmosphere is around 56 °C. Although, this is a limitation in this thesis, it could be interesting to investigate further cooling to regenerate process water and heating of the flue gas stream after the blower to achieve the wanted stack temperature.

## 6.4 Case 3 – Process design

Case 3 is based on case 1 and case 2. The purpose is to determine suitable dimensions and operating conditions for the absorber and desorber. As mentioned in section 5.2.2, the absorber and desorber were evaluated using two variables; α and number of packed bed sections (N). The number of sections is proportional to the packing volume. The diameter and packing height per section are fixed at 11.4 m and 1.5 m, as used in the generic process. Figure 20 and 21 show the results from the simulations. Stream results and data for Figure 20 and 21 are tabulated in Appendix 7.
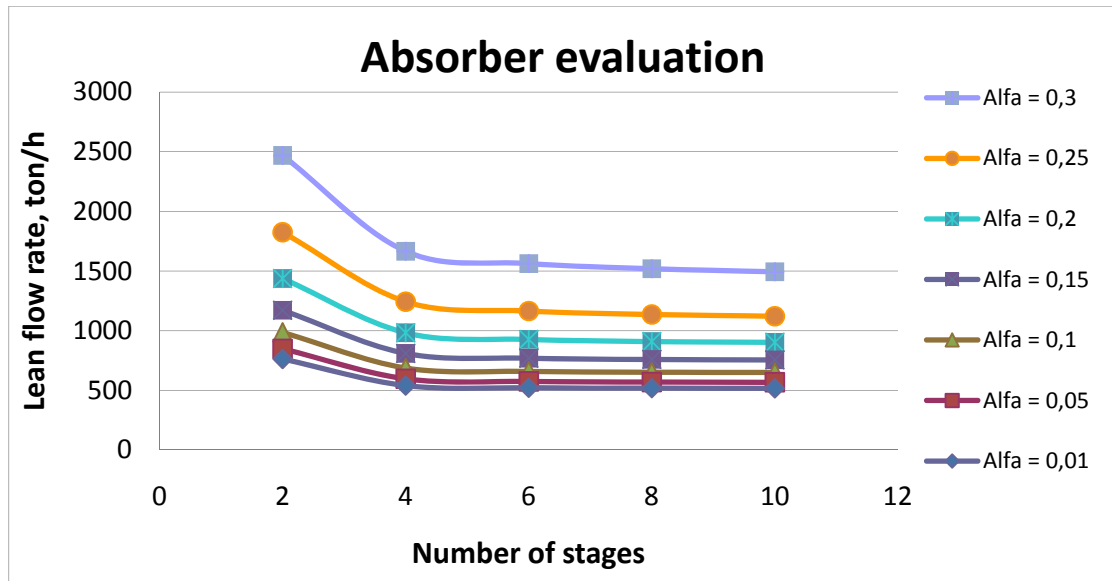
*Figure 20. The lean flow rate plotted against the number of sections at different α-values.*

The required lean flow rate to absorb 85 % $CO_2$ from the flue gases is given by the y-axis values. The lean flow rate decreases at first until a more or less constant value is reached. In other words, the lean flow rate required to absorb 85 % $CO_2$ does not decrease after N = 4 for all values of α. The graphs also show a clear variation of the lean flow rate with α. A low α value results in a low value of the lean flow rate and vice versa. In other words, the lean solution flow rate is to some extent influenced by the number of sections i.e. packing volume, but mostly dependent on the $CO_2$ loading in the lean solution, α. A low $CO_2$ loading favors the absorption as the driving force is larger compared to a lean solution with high $CO_2$ loading (explained more in detail in Appendix 7).

Figure 20 is related to Figure 21 through the value of α and to choose an appropriate value of α Figure 21 is evaluated. Figure 21 is constructed with the chosen value, N = 4, in the absorber and shown below.
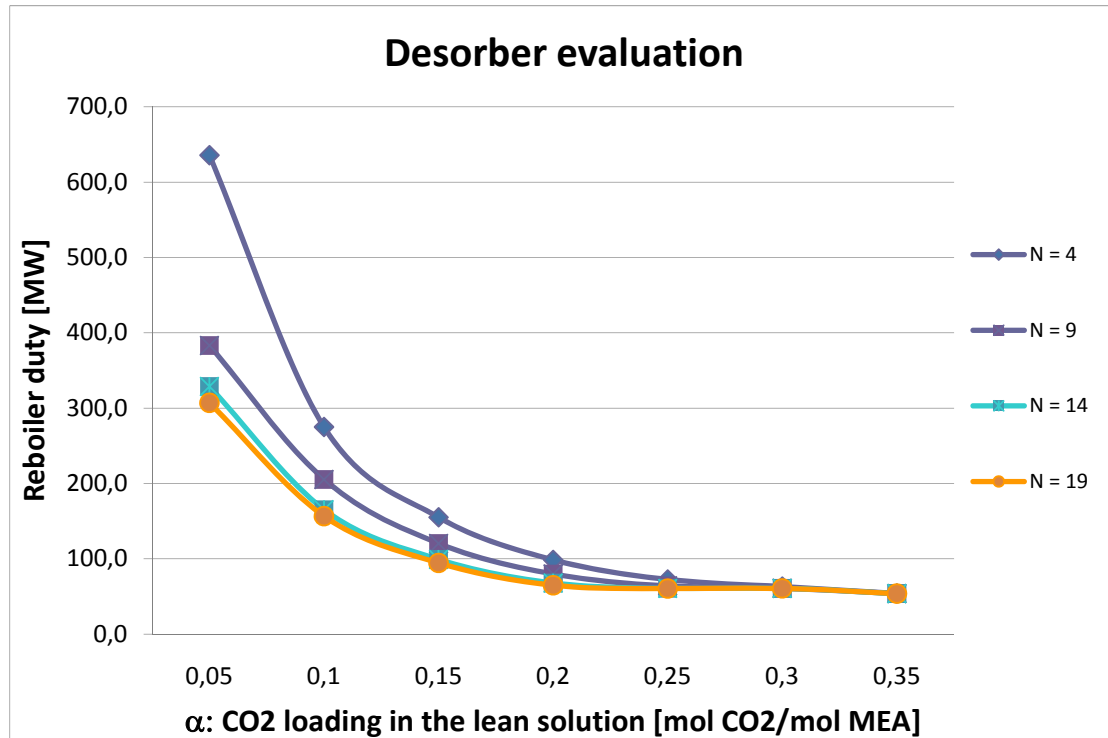
*Figure 21. The reboiler duty was plotted against α at different number of sections in the desorber (N).*

The energy supplied to the reboiler required to regenerate the solvent is given by the y-axis values. As can be seen from Figure 21 the reboiler duty is high at low α for all values of N. The reboiler duty decreases as α increases and reaches a point where there is no dependence of N. The absorbed $CO_2$ in the absorber should be separated in the desorber to avoid accumulation. The absorbed amount of $CO_2$ is 53.5 ton/h, which is 85 % of the ingoing 63 ton/h. Keep in mind that a low α in the lean solution means that there is a low flow of $CO_2$ compared to a high α. A lean solution with a low α that absorbs 53.5 ton $CO_2$/h will therefore result in a lower $CO_2$ flow in the rich solution compared to a lean solution with high α. The rich solution is fed to the desorber and consequently, the separation is more difficult in the former mentioned scenario compared to the latter. A difficult separation will increase the reflux ratio in the column, which means more liquid is recirculated back to the column. The liquid will then end up in the reboiler where it is evaporated. Hence, the reboiler duty increases with low α and decrases with high α, as can be seen in Figure 21. Suitable conditions in the desorber are further evaluated below for α – values 0.15, 0.2 and 0.25.

Operating conditions in the columns

Figure 20 and 21 assesses the absorber and desorber to choose appropriate values of α and N. The most reasonable value of α is between 0.15 and 0.25. To determine what value of α and N should be used, Figure 23 is constructed.
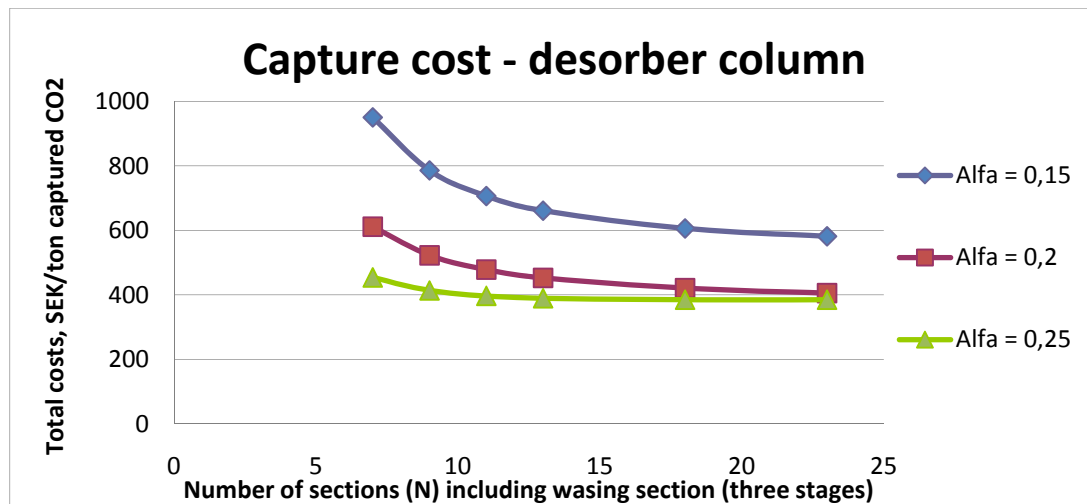
*Figure 23. The total costs (investment and operational) to capture one ton CO₂ varied with the number of section at different values of α.*

Figure 23 evaluates the total cost of the desorber unit for capturing one ton of $CO_2$. The total costs include the investment cost of a desorber column (shell and packings), and the operating cost is the cost of steam supplied to the reboiler, 36 €/MWh (Spetz, 2010). The investment cost is annualized and divided by the captured $CO_2$ quantity per year. The operating cost is also divided by the captured $CO_2$ quantity per year. As shown in Figure 23, the total cost decreases at first because the profit made by steam savings is larger than the cost of investing in a larger column. Thereafter, the cost reaches a minimum at a certain value of N (not reached in the graphs). Beyond that value of N the cost will increase due to the fact that the investment cost increases faster compared to the profit made from steam savings.

Even though the cost is lower for α = 0.25 the chosen value of α is 0.2. As can be seen in Figure 20, higher α results in a higher flow rate of the lean solution. The costs difference between α – values is considered too small to motivate larger flows in the process. Larger flows in the process would imply investment of larger equipment. The resulting values of N in the columns and the value of α become:

$N_{absorber} = 4$

$N_{desorber} = 22$

$α = 0.2$

The stream and equipment results (from case 3 simulation) of relevance are presented in Table 9.

*Table 9. The table shows equipment results from case 3 simulations, and the difference from case 1 simulations. For more detailed data of these cases, see Appendix 5 and Appendix 7.*

| Equipment | | Value Case 3 | Value Case 1 |
|---|---|---|---|
| Absorber | | | |
| | Number of absorption sections | 4 | 8 |
| | Number of washing sections | 3 | 3 |
| | Lean mass flow [ton/h] | 996 | 908 |
| | Alfa [mol CO2/mol MEA] | 0,2 | 0,186 |
| Desorber | | | |
| | Number of desorption sections | 19 | 19 |
| | Number of washing sections | 3 | 3 |
| | Condenser duty [MW] | 26,5 | 32,1 |
| | Reboiler duty [MW] | 64,9 | 65,9 |
| Flue gas cooler | | | |
| | Duty [MW] | 64 | 16 |
| DCC | | | |
| | Duty [MW] | 0 | 48 |
| Blower | | | |
| | Work [MW] | 2,1 | 2,1 |
| Rich heater | | | |
| | Duty [MW] | 60,7 | 62,9 |
| Lean cooler | | | |
| | Duty [MW] | 84,1 | 76,2 |
| CO2 compression | | | |
| | Number of compression stages | 3 | 3 |
| | Number of cooling stages | 3 | 3 |
| | Total work [MW] | 4,7 | 4,7 |
| | Total cooling duty [MW] | 8,2 | 8,2 |

Stream results in the simulation are presented in Appendix 7.

### 6.4.1 Pinch analysis of the carbon capture process

In order to construct composite curves and a grand composite curve, individual minimum temperature differences are chosen for each stream. The stream data sheet with all information related to the pinch analysis is presented in Appendix 8. The results from the pinch analysis for the carbon capture process in case 3 are presented below.
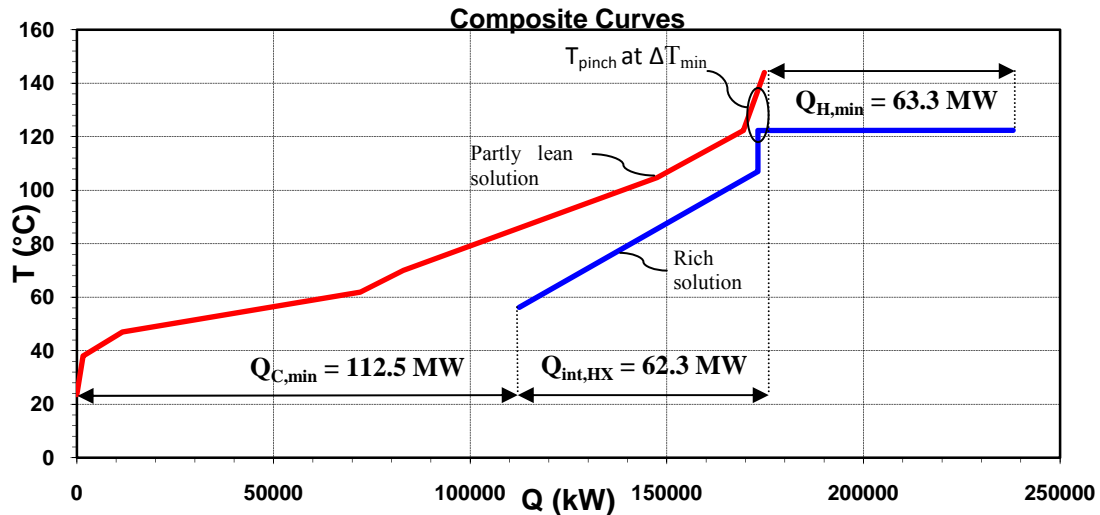
*Figure 24. Composite curve of the carbon capture process.*

The minimum hot utility is 63.3 MW and the minimum cold utility is 112.5 MW. The maximum internal heat exchange is where the red and blue curve overlaps, 62.3 MW. The maximum internal heat exchange is to a large extent between the rich and lean solution. The GCC for the carbon capture process is represented in the figure below. Note that the temperatures on the y-axis are given in interval temperatures.
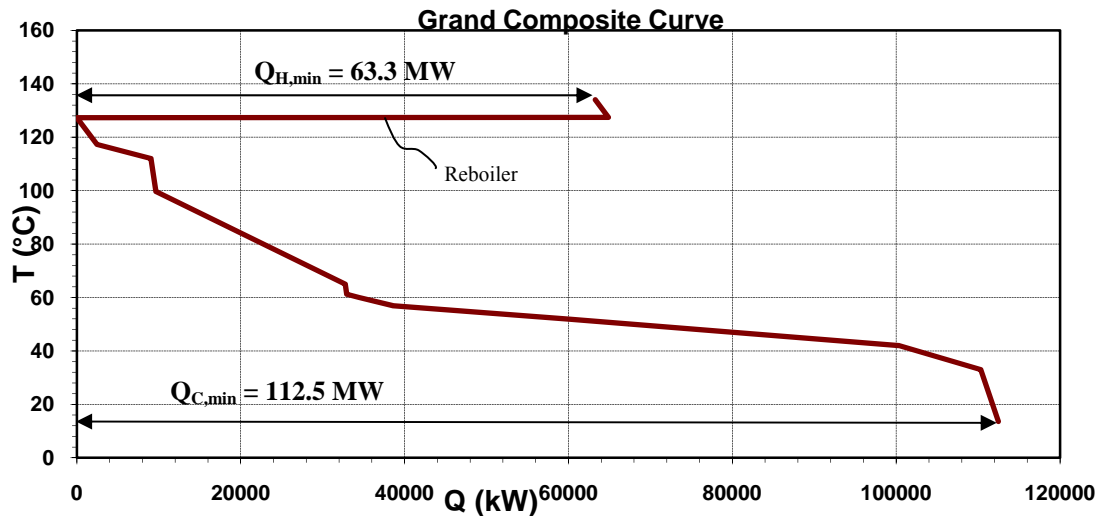


*Figure 25. The GCC of the carbon capture process.*

As mentioned in section 5.3 the GCC shows the temperature level of which the hot and cold utilities (steam and cooling water respectively) have to be supplied. It also shows that the pinch temperature is 127.3 °C. The flat section of the curve represents the reboiler i.e. the liquid in the reboiler boils at a constant temperature. However, the condenser is not represented as a flat section. Due to the fact that the vapour stream from the top stage in the desorber is a mixture of $CO_2$ and $H_2O$, condensation does not occur at a constant temperature. Instead, the condensing $CO_2$-$H_2O$ stream goes from 104 °C to 23.6 °C to get a 98 % $CO_2$ mole purity. The water condensation begins at the dew point temperature at 101°C, shown in Figure 26 below.
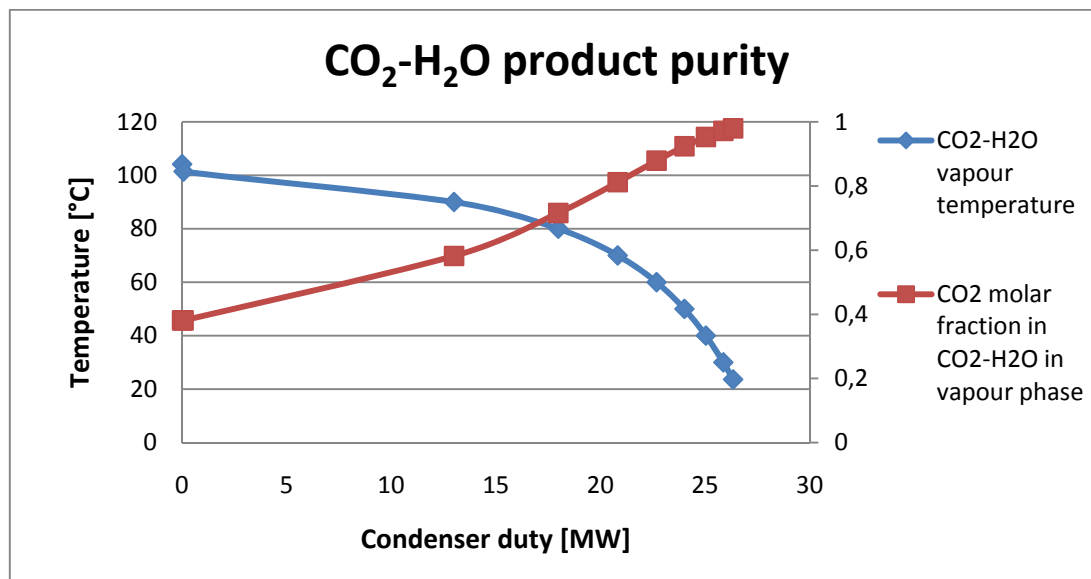
*Figure 26. Condensation line of the $CO_2$ and water vapour stream and its $CO_2$ purity variation with temperature.*

The next step is to determine the integration possibilities with streams in the cracker plant. From the shape of the GCC in Figure 25, it can be seen that there is 34 MW of excess heat at a temperature above 60 °C. This might open for integration possibilities with streams in the cracker process that can be heated with the excess heat at these temperatures. This is investigated in the background/foreground analysis.

Background/Foreground Analysis

All streams with a heating or cooling demand in the cracker process are given from a previous pinch study of the cracker process, presented in Appendix 8 (Hedström, 2008). From these streams, a number of streams that fulfils following criteria are chosen:

- The start and target temperatures of the streams should be lower than the pinch temperature of the carbon capture process. If this criterion is fulfilled, these streams can potentially receive excess heat from the carbon capture process.
- The streams should currently be heated with a hot utility that has a higher temperature than the reboiler temperature. The hot utility can be used in the carbon capture process instead of heating streams at low temperatures.
- The heating demand of the streams should be relatively high compared to those in the carbon capture process. Few streams with large duties reduce the number of heat exchangers needed for heat recovery compared to a lot of streams with small duties.
- Physically possible i.e. sufficient space for piping and possible new heat exchangers.

The streams that fulfil the criteria are presented in Table 10. The individual minimum temperature difference is 8 K for liquids and 5 K for evaporation/condensation (Spetz, 2010).

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*Table 10. Streams chosen from the cracker plant.*

| Stream no. | Stream name | Tstart [°C] | Ttarget [°C] | Duty [kW] | Currently used utility | ΔT$_{min}$ |
|---|---|---|---|---|---|---|
| 1 | LPG feed heater/vaporizer | 4 | 73 | 2377 | MP steam | 8 |
| 2 | LPG feed heater/vaporizer | 73 | 83 | 3875 | MP steam | 5 |
| 3a | Preheating demin. water | 44 | 128 | 17640 | LP steam | 8 |
| 3b | Preheating demin. water | 44 | 118 | 15540 | LP steam | 8 |
| 4 | Reboiler T-1809 | 52 | 53 | 3090 | LP steam | 5 |
| 5 | Reboiler T-1805 | 55 | 56 | 21230 | LP steam | 5 |

The streams in Table 10 are represented in the GCC below (blue line) together with the GCC from the carbon capture process (red line). The red line is the background process and the blue line is the foreground process. Figure 27(a) shows the integration using stream 3a in Table 10, and Figure 27(b) shows integration using stream 3b. The other streams in the table remain the same in both figures.

As can be seen in Figure 27, the foreground process curve intersects the background process curve. At the intersection the heat exchange between a hot and cold stream is at the defined minimum temperature difference i.e. the hot stream temperature is ΔT$_{min}$ °C higher than the cold stream temperature. Consequently, due to ΔT$_{min}$ restrictions heat exchange is only possible between the intersections.
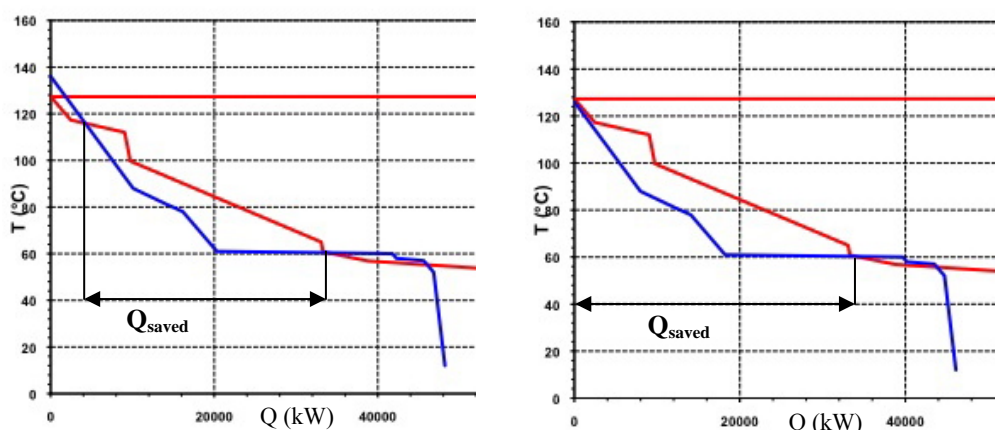


*Figure 27. (a) shows the starting point towards finding the best possible integration (b) a stream has been modified and results in better integration compared to (a).*

The integration in Figure 27(a) shows that it is possible to save 29.7 MW of steam. However, by modifying stream 3a to 3b in Table 10 it is possible to integrate 33.7 MW, as shown in Figure 27(b). The improvement indicates that it is more beneficial to change from stream 3a to 3b. Excess heat from the cracker process is then used to preheat the demineralised water from 44 °C to 118 °C and 1.8 bar(g) steam is used to heat it from 118 °C to 128 °C.

Removing stream 1 and 2 from the foreground process, further integration improvements can be made. The streams used in the foreground process are 3b, 4 and 5 and the result is presented in Figure 28.
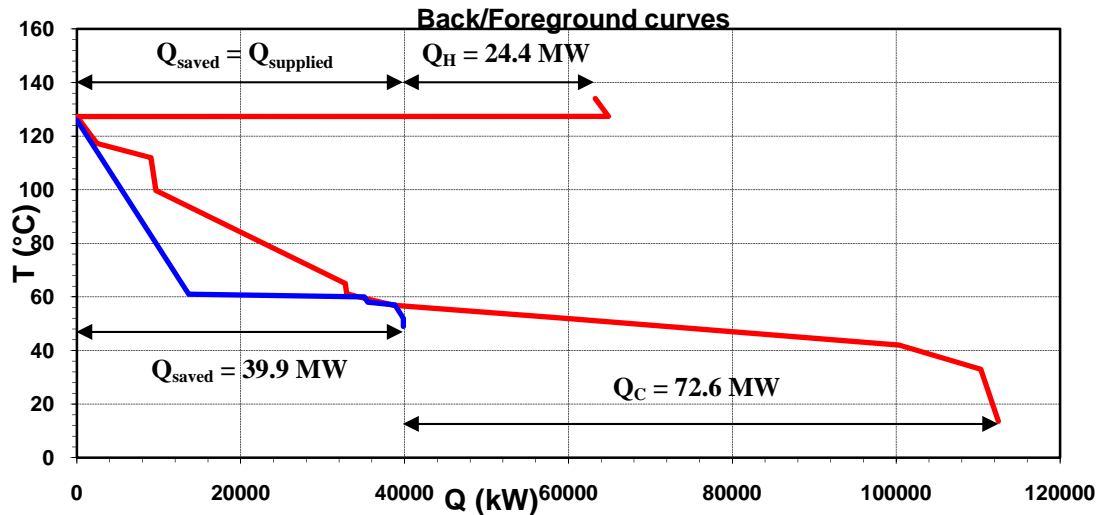
*Figure 28. The background/foreground analysis for the best integration case.*

Heat supply from the carbon capture process to the reboilers at T-1805 and T-1809 and to heat the demineralised water can save 39.9 MW of LP steam in the cracker plant. This integration is the most beneficial concerning energy saving in the cracker. In addition, a fewer number of streams are used in the integration compared to the integration in Figure 27. This is advantageous considering investment costs and process retrofits.

Utilizing excess heat from the carbon capture process can release the steam that is currently used to heat the chosen streams in the cracker process. Figure 28 shows that it is possible to release 39.9 MW, which can be supplied to the carbon capture process since the steam temperature is higher than the reboiler temperature. The remaining heat needed in the reboiler, $Q_H$, is 24.4 MW (= 63.3 MW – 39.9 MW). This heat can be supplied from the steam boilers as these are currently operating at minimum load. Also, thanks to the steam savings, it can be possible to supply steam to the reboiler from the LD5 process. As a result, an additional steam boiler can be avoided with the integration shown in Figure 28. Not only does this integration save steam, but also cooling water. The cooling duty, $Q_C$, for the carbon capture process decreases by 39.9 MW, from 112.5 MW down to 72.6 MW.

Limitations

Although the results seems promising from an energy saving point of view, it is worth to mention that this is the theoretical amount that can be saved. This means that a constructed heat exchanger network has to recover 39.9 MW of heat from the carbon capture process. In fact, several heat exchangers may perhaps be too small to motivate an investment of the heat exchangers, which results in a decreased heat recovery. Additionally, it might not be possible to heat exchange e.g. the flue gases with a column reboiler located in the cracker process. Therefore, it could be necessary to use indirect heat exchanging i.e. transfer heat from the flue gases to another medium that in turn is heat exchanged with the reboiler. If that is the case, the minimum temperature difference has to be increased. However, this thesis only investigates the theoretical amount of energy that can be saved and these practical issues are not considered.

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## 6.5    Case 4 – Mechanical vapour recompression (MVR)

Case 4 is based on case 3 to investigate a solution that can reduce the supplied external hot utility (i.e. steam) to the desorber reboiler by introducing mechanical vapour recompression (MVR). Figure 29 shows the MVR configuration around the desorber. The $CO_2$-$H_2O$ stream in the figure is the vapour stream from the desorber top stage. The composition of the $CO_2$-$H_2O$ vapour stream going in to the compressor is, on a molar basis, 38 % $CO_2$ and 62 % $H_2O$ (the same as in case 3). The compression increases the temperature of the stream from 104 °C to 260 °C and can therefore be heat exchanged with the desorber reboiler (reboiler temperature is 123 °C). During the heat exchange with the reboiler, water in the $CO_2$-$H_2O$ stream condenses. This stream is further cooled in the condenser to 23.6 °C prior to the multistage compressors. The $CO_2$-$H_2O$ vapour is separated from the condensed water in a splitter. The top stream consists of > 99 mol% $CO_2$ fed to the multistage compressor. The bottom stream, which consists of condensed water, is recirculated to the desorber water wash section. Stream and equipment data related to Figure 29, which are of interest, are presented in Table 11.
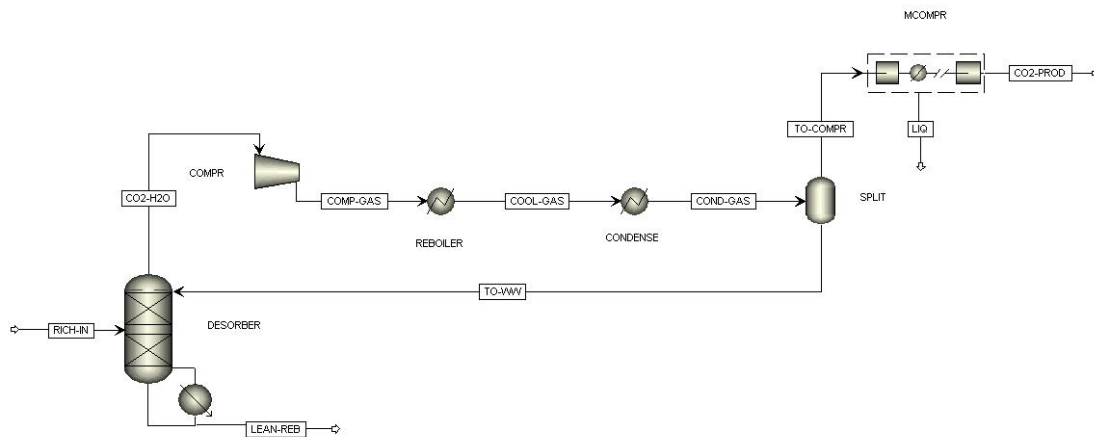


*Figure 29. MVR introduced in the desorber.*

*Table 11. The results from the MVR are shown in the table.*

| Equipment | Specification | Value |
|---|---|---|
| Compressor | COMPR | |
| | Work [MW] | 5,3 |
| | Outlet temperature [°C] | 260 |
| | Inlet pressure [bar] | 1,9 |
| | Outlet pressure [bar] | 6,8 |
| Reboiler | | |
| | Supplied energy from MVR [MW] | 15,5 |
| | Supplied external heat [MW] | 49 |
| | Outlet temperature [°C] | 133 |
| | Outlet vapor fraction | 0,68 |
| Condenser | | |
| | Duty [MW] | 16,8 |
| | Outlet temperature [°C] | 23,6 |
| | Outlet molar CO2 purity (vapour phase) | > 0,99 |
| CO2 compression | MCOMPR | |
| | Number of compression stages | 2 |
| | Number of cooling stages | 2 |
| | Total work [MW] | 2,9 |
| | Total cooling duty [MW] | 6,3 |

The energy balance over the MVR is:

*Supplied work to the compressor = 5.3 MW*

*Supplied energy to the reboiler = 15.5 MW*

*Duty of condenser (in MVR flowsheet) = 16.8 MW*

*Heat removed from the CO2-H2O stream = 16.8 + 15.5 – 5.3 = 27 MW*

The heat removed from the $CO_2$-$H_2O$ stream without MVR (i.e. case 3) is 26.5 MW. The difference arises due to the mechanical efficiency in the compressor after the desorber (set to 0.98).

### 6.5.1 Pinch analysis of process with MVR configuration
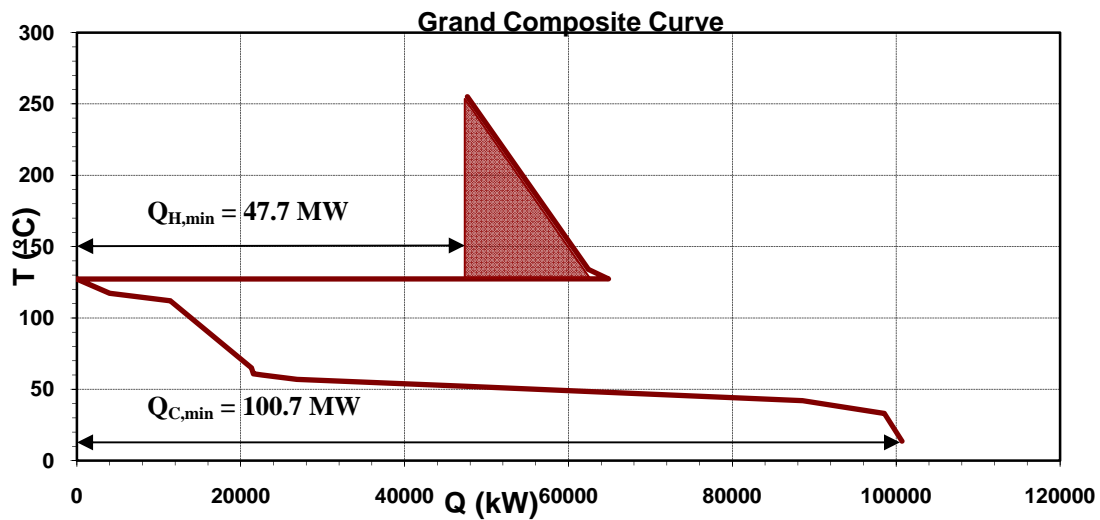


*Figure 30. The figure shows the GCC with MVR.*

Stream data used in the pinch analysis are shown in Appendix 10. The GCC for the carbon capture process with MVR has changed in such a way that earlier available heat from the condenser at low temperature is now supplied to the reboiler (pocket). Therefore, the GCC shows that the minimum hot utility decreases to 47.7 MW. Consequently, the minimum cold utility decreases to a value of 100.7 MW. The background/foreground analysis will change as the shape of the GCC is changed. To compare whether the integration possibilities are changed compared the one presented in case 3 (Figure 28), a background/foreground analysis was made.
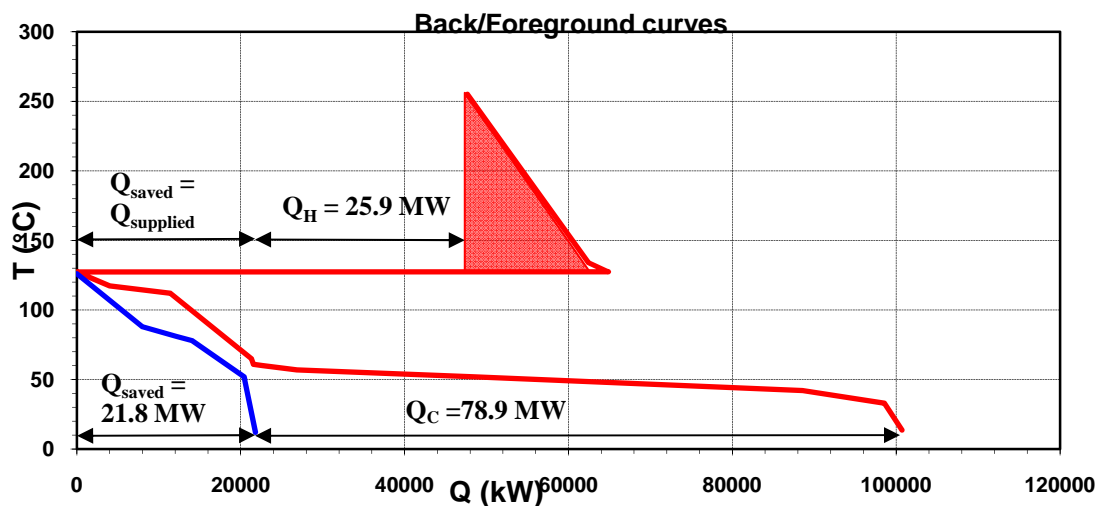


*Figure 31. The figure shows the background/foreground curves with MVR for the best integration solution.*

The background/foreground analysis shows best integration possibilities with stream 1, 2 and 3b in Table 10. The integration shows that it is possible to save 21.8 MW of steam in the cracker process i.e. the steam to the reboiler can be supplied from the LD5 process. The minimum cooling demand becomes 78.9 MW. If the saved steam is supplied to the reboiler the hot utility demand, $Q_H$, becomes 25.9 M kW (= 47.7 MW

– 21.8 MW), which can be supplied by increasing the load in the steam boilers. As in case 3 this is the theoretical amount of steam savings.

## 6.6    Comparison between case 3 and 4

Relevant results to compare the carbon capture process without MVR (case 3) and with MVR (case 4).

*Table 12. The differences in the $CO_2$ compression between case 3 and case 4.*

| Equipment | | Value Case 3 | Value Case 4 |
|---|---|---|---|
| Compressor | COMPR | None | |
| | Inlet pressure [bar] | - | 1,9 |
| | Outlet pressure [bar] | - | 6,8 |
| | Work [MW] | - | 5,3 |
| | Inlet mass flow [ton/h] | - | 90 |
| CO2 compression | MCOMPR | | |
| | Number of compression stages | 3 | 2 |
| | Number of cooling stages | 3 | 2 |
| | Inlet pressure [bar] | 1,7 | 6,6 |
| | Outlet pressure [bar] | 75,5 | 75,5 |
| | Inlet CO2 purity [mol%] | 0,98 | > 0,99 |
| | Total work [MW] | 4,7 | 2,9 |
| | Total cooling duty [MW] | 8,2 | 6,3 |
| | Mass flow to first stage [ton/h] | 54 | 53.5 |

The total work required in the $CO_2$ compression is lower with MVR compared to case 3 without MVR, since the ingoing pressure is at 6.8 bar. However, compression of the $CO_2$-$H_2O$ stream to 75.5 bar is more work consuming with MVR compared to case 3 due to the compressor after the desorber. The total compressor work becomes 8.1 MW and 4.7 MW for case 4 and case 3 respectively. It can be concluded that the compressor after the desorber is large in comparison with the other compressors as it stands for 65 % of the total work consumption due to the higher inlet mass flow. Therefore, it is also expected that the investment costs for the MVR configuration is higher than a 3-stage compressor train. The difference in mass flow to the first compressor in the $CO_2$ compression and the total cooling duty arise as the inlet $CO_2$ purity is higher i.e. more water has condensed in case 4 compared to case 3.

Moreover, Table 13 shows the energy balance for case 3 and 4.

*Table 13. Results from the background/foreground analysis for case 3 and 4.*

| | Case 3 | Case 4 |
|---|---|---|
| Hot utility demand [MW] | 63,3 | 47,7 |
| Cold utility demand [MW] | 72,6 | 78,9 |
| Heat integrated [MW] | 39,9 | 21,8 |
| Heat from MVR [MW] | 0 | 15,5 |
| Heat from steam boilers [MW] | 24,4 | 25,9 |

The integration according to case 3 is more advantageous since the heat supply from the steam boilers is lower than in case 4. As mentioned previously, excess heat (below the pinch) in the carbon capture process is lifted and supplied at a higher temperature (above the pinch) to the reboiler (case 4). As the amount of excess heat in carbon capture process is reduced, there is less excess heat that can be utilized for the cracker process. The lost heat of integration between the cracker and carbon capture process with MVR is larger than the heat supplied from the MVR, which results in a higher steam boiler duty in case 4. This means that less fuel is combusted in the steam boilers in case 3, which reduce the operational cost.

The COP of the MVR heat pump is 2.9 (heat supplied to the reboiler/work supplied to the compressor). There are a number of reasons, coupled to one another. The compressed vapour consists of $CO_2$ and water vapour. The mixture condenses according to Figure 26 i.e. not at a constant temperature. Therefore, compared to a pure component stream that condenses at constant temperature, a higher temperature lift is necessary to supply the same amount of heat, which requires higher pressure increase in the compressor. In addition, since the inlet vapour stream to the compressor is a mixture of $CO_2$ and water vapour only the water condenses in the temperature regions of the MVR (outlet vapour fraction after heat exchange with the reboiler is 0.68). As the stream is not completely condensed the heat supplied to the reboiler is reduced. Hence, the COP does not reach typical values for common MVR configurations (COP is approximately 12 for this temperature lift for a pure component).

Finally, case 3 is used in the economical evaluation. The flowsheet and stream data are presented in Appendix 7.

# 7    Economic evaluation

Detailed costs (investment and operating) of all process equipment required in the carbon capture process used in this thesis for the cost evaluation are shown in Appendix 8. The costs used in this thesis are updated to year 2010.

Total capital investment (CAPEX)

The cost evaluation is made for case 3 as it offers the best integration possibilities. In the cost evaluation, heat exchangers for the process integration with the cracker process are *not* included. In the cost evaluation the carbon capture process is divided into four blocks; absorber, desorber, compression and piping. All equipments in the carbon capture process that are included in the absorber, desorber and compression are shown in Appendix 2. The capital investment for each block is shown in Figure 32.
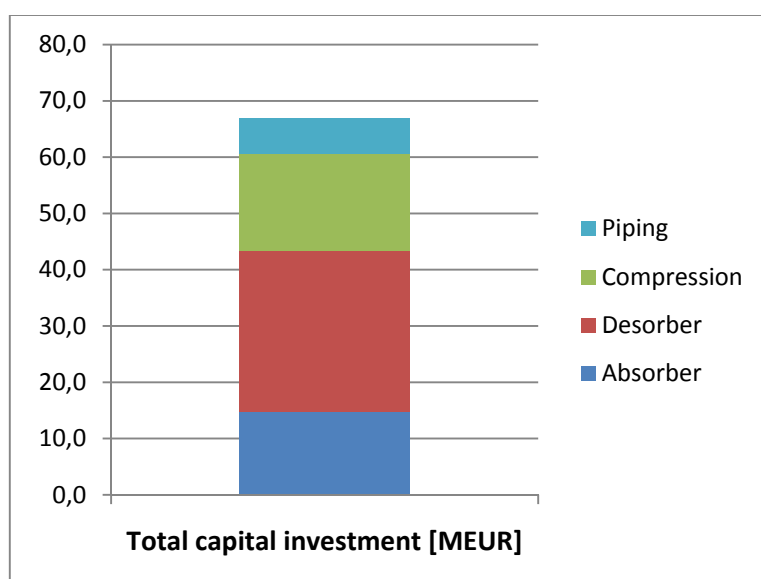


*Figure 32. Total capital investment (CAPEX) in million Euros.*

The total capital investment sums up to a cost of 66.9 M€. The largest investment costs within the blocks are:

- Absorber: flue gas cooler, blower and absorber column.
- Desorber: rich-lean heat exchanger, desorber column, condenser and reboiler.
- Compression: compressors in the $CO_2$ compression.
- Piping from cracker furnaces chimneys to the carbon capture process

Total operating costs (OPEX)

The carbon capture process has a power (electricity) demand of 11 MW. The heating demand is 63.3 MW, of which 39.9 MW is supplied from the process integration. The remaining heat demand is 24.4 MW and can be supplied from the existing steam boilers via the steam system presented in section 3.3.1 and Figure 3. Following steps are the calculation basis for the total operating costs.

- 30.5 MW of fuel gas is burnt in the steam boilers to produce 24.4 MW of heat. The fuel cost is 31 €/MWh.

- HP steam is produced in the boilers and expanded to MP steam through the turbo alternator. 24.4 MW of heat is delivered at MP level from the turbo alternator.
- The turbo alternator also produces electricity and has available capacity to produce electricity. The electricity produced is 2.1 MW and can be used in the carbon capture process.
- The remaining electricity needed in the carbon capture process is 8.9 MW, which is purchased from the grid. The price of the electricity is 57 €/MWh.

The total operating costs consist of variable operating costs (e.g. fuel and electricity) and fixed operating costs (e.g. labour and maintenance), which are presented in detail in Appendix 8. The total operating cost sums up to 15.9 M€/year, where the variable costs make up to 81 % (12.8 M€/year). Figure 33 gives an overview of the variable operating costs.
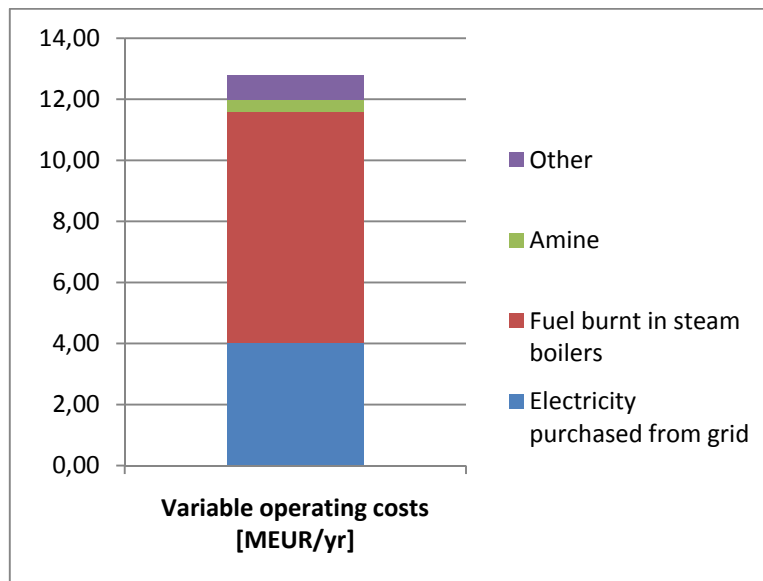


*Figure 33. The figure shows the total operating costs (OPEX) in million Euros per year.*

The electricity and fuel costs cover 32 % and 59 % of the variable operating costs respectively.

## 7.1 Comparison with Tel-Tek study

To determine if it is possible to decrease the capture cost (€/ton captured $CO_2$) with process optimization and particularly with process integration, the CAPEX and OPEX for the integrated case are compared with results from the study without process integration. Detailed costs and cost factors used in the Tel-Tek study for the comparison are shown in Appendix 2. The costs used in the Tel-Tek study are from 2006. To determine the cost for 2010 from the costs for 2006, the CEPCI (Chemical Plant Cost Index) has been used (Chemical Engineering).

*CEPCI, 2006 = 499.6 (mean value of all values in 2006)*

*CEPCI, 2010 = 542.3 (mean value of values from January to April 2010)*

Total capital investment (CAPEX)

The difference in the total capital investment is shown in the Figure 34.
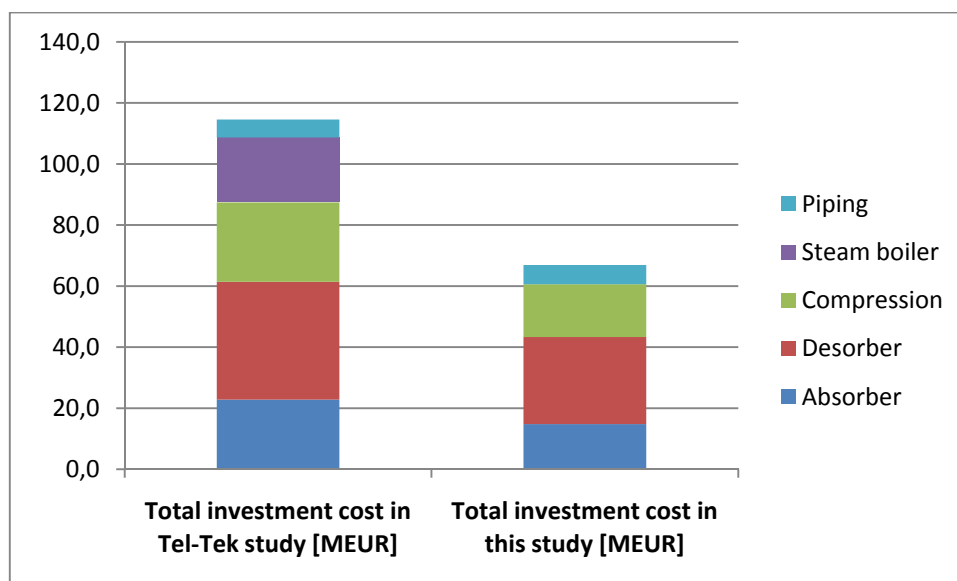


*Figure 34. The figure compares the CAPEX in both studies.*

The total capital cost in the Tel-Tek study is 114.5 M€, which is 47.6 M€ more than in this thesis. The main difference is that with process integration the steam boiler proposed in the Tel-Tek study can be avoided (see section 4.3). Furthermore, the turbine that drives the blower in the Tel-Tek study is included in the absorber block and the turbine that drives the compressors in the Tel-Tek study in the $CO_2$ compression is included in the compression block.

The cost differences between the absorber blocks mainly arise from:

- Absorber column evaluation results in a lower packed bed volume, but also since the flue gas flow rate into the absorber is lower in this thesis compared to the Tel-Tek study (which includes flue gases from the proposed steam boiler). The packed bed volume is reduced from 2847 m$^3$ to 1071 m$^3$.

The differences between the desorber blocks arise mainly from:

- Lower flue gas flow rate and lower captured $CO_2$ quantities results in a lower lean solution flow rate in this thesis compared to the Tel-Tek study. As a result, the rich-lean heat exchanger has a lower area, which results in a lower cost.
- Lower captured quantities of $CO_2$ (53.5 ton/h) results in a lower heat demand (proportional to area) in the reboiler in this thesis compared to the Tel-Tek study (70 ton/h). Decreased areas results in a lower cost.

The differences between the compression blocks arise mainly from:

- Lower quantity of $CO_2$ captured results in a lower desorbed amount of $CO_2$ that is fed to the compression, which decreases the work duty of the compressors. As a result the costs decrease.

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Total operating costs (OPEX)

The difference in total operating costs is shown in Figure 35.
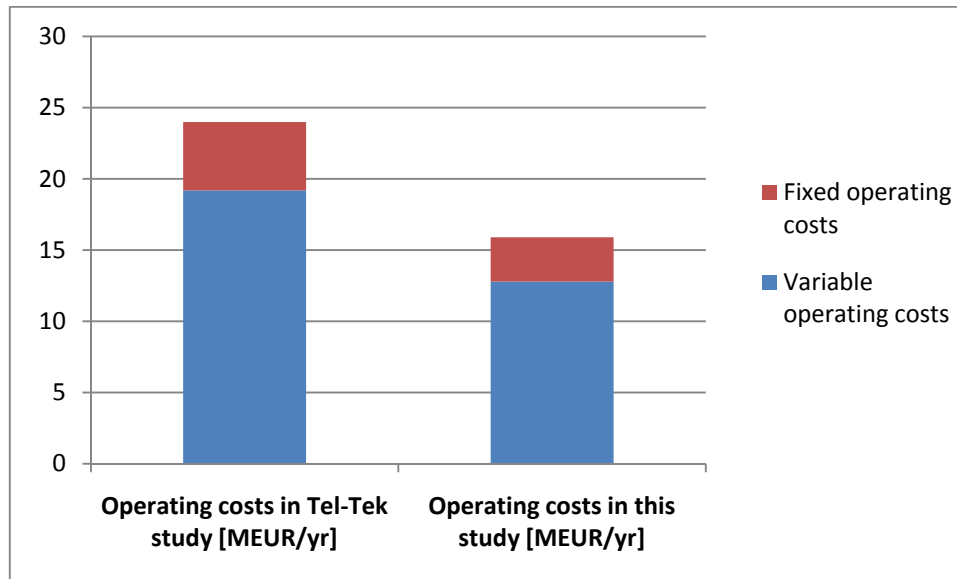


*Figure 35. The figure shows the OPEX in both studies.*

The total operating cost for the Tel-Tek study is 24.0 M€/year. The total operating cost for this thesis is 15.9 M€/year. The difference in fixed operating costs is given from lower labour (less salaries) and lower maintenance cost. Lower maintenance cost is a result from the lower total capital cost (4 % of total capital cost). The differences in variable costs are discussed with Figure 36.
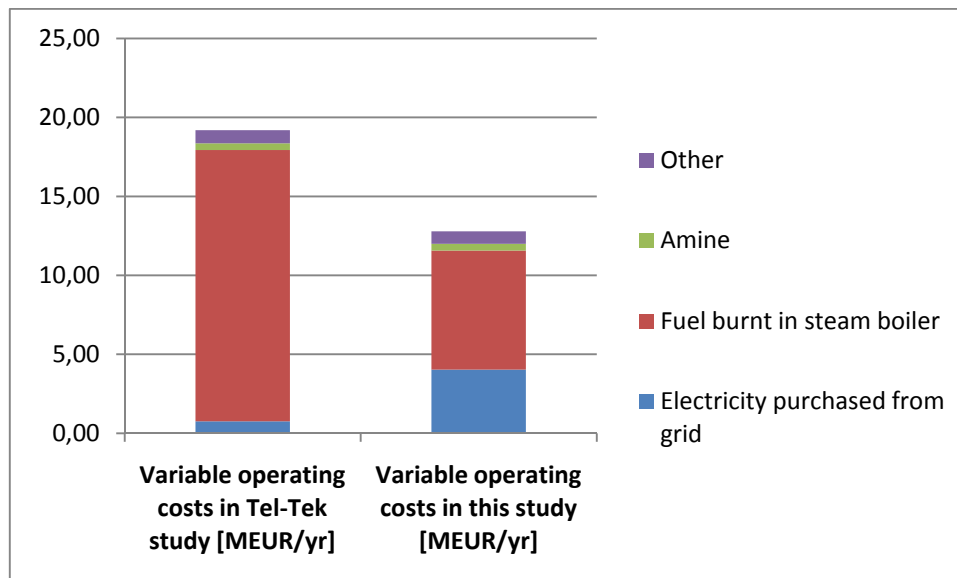


*Figure 36. The figure shows the differences in variable costs. The variable costs in the Tel-Tek are from 2006. The amine and other costs in this thesis are from 2006 as well. The electricity and fuel costs in this thesis are from 2010.*

The electricity and fuel costs in this thesis make up to 91 % of the total variable costs (90 % in Tel-Tek study). Consumption and prices related to electricity and fuel in Figure 36 are shown in Table 13.

*Table 13. The costs for fuel and electricity used in both studies.*

Operating plant time = 8000 [h/year]

|  | Tel-Tek study | This thesis |
|---|---|---|
| Costs for year | 2006 | 2010 |
| Electricity price [€/MWh] | 43,2 | 56,6 |
| Fuel price [€/MWh] | 21,2 | 30,9 |
| Electricity consumption from grid [MW] | 2,2 | 8,9 |
| Fuel consumption [MW] | 85 – 95 | 30,5 |

The amine costs and other costs are shown in Appendix 8. The differences in variable costs are to some extent a result of the costs for electricity and fuel, which are presented in Table 13.

As can be seen in Figure 36, the variable costs in the Tel-Tek study is 19.2 M€/year compared to 12.8 M€/year in this thesis. Thanks to the process integration, a lower amount of steam can be supplied to the carbon capture process from the existing steam boilers in the cracker plant in this thesis compared to the previous. This implies that a lower amount of fuel is burnt in the steam boilers, which gives lower fuel costs. The steam generated in the boilers is also HP steam, which is expanded through the turbo alternator that produces electricity (see Figure 3). As less steam is produced in the boilers, less power will be generated and more has to be bought from a power plant compared to the Tel-Tek study.

Considering the amine costs, these are based on the losses in the process. The losses are amine emissions to air, and amine that forms by-products (outside the scope of this thesis). There is no information in the Tel-Tek study of how much amine is transformed into by-products, or how much of the amine cost arise due to these losses. As a result, the amine cost is considered the same. The prices that cover the other costs are also assumed to be the same. Since amine and other costs are small compared to the fuel and electricity costs these have a minor impact on the results.

### 7.1.1 Cost of net captured $CO_2$

The $CO_2$ capture cost is based on the total investment and operating costs and given in €O per ton captured $CO_2$. This is defined as:

$$CO_2 \ capture \ cost = \frac{Annualized \ tot. investment \ cost + Tot. operating \ costs}{Yearly \ captured \ CO_2}$$

The annualized investment cost (€/year) is calculated from:

$$Annualized \ tot. investment \ cost = Tot. investment \ cost \cdot \left[\frac{i \cdot (i+1)^n}{(i+1)^{n-1}}\right]$$

Where *i* is the interest rate and *n* is the number of years. For *i = 0.075* and *n = 25 (the annuity factor is 0.09)* the annualized investment costs can be calculated.

The purpose of the carbon capture plant is to capture 85 % of the flue gases generated in the cracker furnaces. In the Tel-Tek study heat and power to the carbon capture process was supplied from the natural gas fired steam boiler. The combustion of natural gas also generates flue gases containing $CO_2$, which partly is vented to the atmosphere (15 %) and partly captured in the carbon capture plant (85 %). This implies that 85 % absorption of $CO_2$ from the cracker furnaces results in additional $CO_2$ emissions from the natural gas fired steam boiler. The *net captured $CO_2$* is therefore:

*Net captured $CO_2$ = $CO_2$ emissions without $CO_2$ capture (before) – Total $CO_2$ emissions with $CO_2$ capture (after)*

The net captured $CO_2$ per year in the Tel-Tek study is 413 600 tonnes and in this thesis 386 400 tonnes (8000 operating hours per year). The net captured $CO_2$ is shown in more detail in Appendix 11. The net $CO_2$ capture cost for the studies is shown in Figure 37.
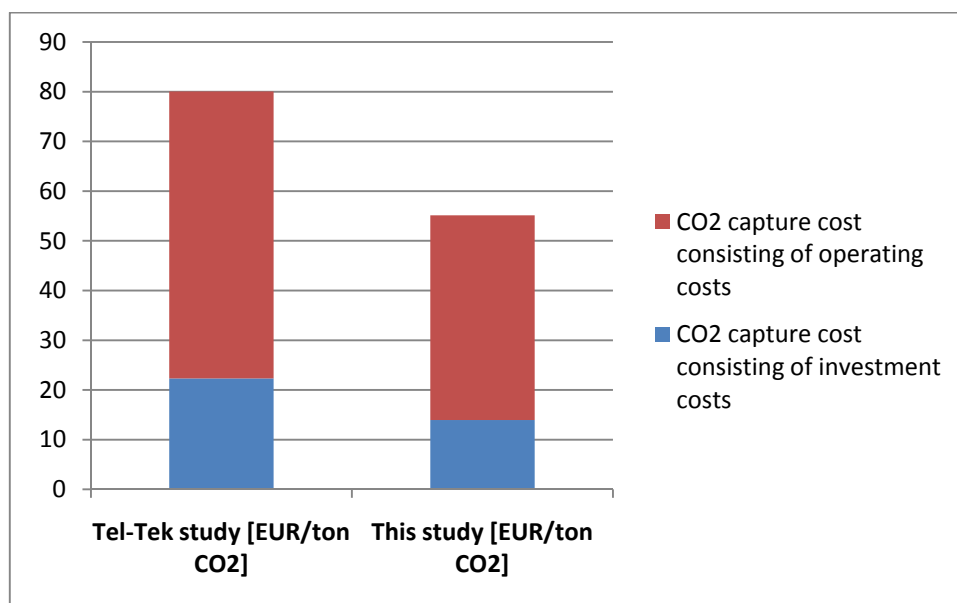


*Figure 37. Comparison of the $CO_2$ capture costs between both studies.*

The $CO_2$ capture cost for the Tel-Tek study is 80 €/ton $CO_2$, compared to 55 €/ton $CO_2$ in this thesis. The operating costs in this thesis make up to 75 % (72 % in Tel-Tek study) of the $CO_2$ capture cost and that shows the importance of energy optimizing the process.

Although this thesis shows promising ways of reducing the $CO_2$ capture cost, from Borealis point of view the conclusion is that the reduction is not to a sufficient extent. The $CO_2$ net capture cost is 3 – 4 times higher compared to today's cost of $CO_2$ emissions for Borealis at 15 €/ton $CO_2$ and almost twice as high to the predicted emission cost in the upcoming trading period, 30 €/ton $CO_2$. However, in the longer perspective, the price for $CO_2$ emissions is probably going to increase substantially (Axelsson, 2010) and CCS can then even show to be profitable.

# 8　Conclusions

Simulation of the carbon capture process makes it possible to construct a process that is specifically designed for the cracker flue gases. The packed bed volume in the absorber column was reduced with 62 % and the costs by 47 % compared to the Tel-Tek study (where the carbon capture process was scaled from a generic case). Due to the fact that the column handles large volumetric flue gas flows a high packing volume is necessary, and for that reason the absorber column has the highest investment costs. Consequently, it becomes important to optimize the column to reduce its investment costs. Evaluation of the desorber column, which is the most energy consuming unit in the process, results in a reboiler duty of 64.9 MW and a condenser duty of 26.5 MW. These duties correspond to 4370 kJ/kg $CO_2$ and 1780 kJ/kg $CO_2$ respectively.

Furthermore, 39.9 MW of excess heat in the carbon capture process can be utilized and release steam that is currently used in the cracker process. The released steam can be supplied to the carbon capture process and the remaining 24.4 MW can be supplied from the existing steam boilers. Thus, investment in an additional steam boiler can be avoided.

The possibility of utilizing heat from the stripper condenser to heat the reboiler by integration of a MVR was investigated. The $CO_2$–$H_2O$ vapour stream from the top stage in the desorber was compressed i.e. the stream temperature was lifted above the reboiler temperature. The compressed vapour stream heat supply was 15.5 MW when heat exchanged with the reboiler. However, this solution was not the optimal integration solution in this thesis as additional heat to the reboiler from the steam boilers became 25.9 MW. Less excess heat could be recovered due to the fact that the heat was supplied to the reboiler and no longer available for the cracker process. The fact that MVR is not advantageous in this thesis is also because the value of COP at 2.9 is low compared to more common values for MVR configurations. On the other hand, this shows theoretical integration potential within the carbon capture process that can reduce the hot utility demand.

Moreover, the cracker flue gases contains 21.1 mol% water, which in the process proposed in this thesis, showed that it was possible to condense substantial amounts of water that make up 20 % of the total fresh (process) water consumption at Borealis.

The process proposed in this thesis reduces the CAPEX with 42 % (from 114.5 M€ to 66.9 M€) and the OPEX by 34 % (from 24 M€/yr to 15.9 M€/yr). The economic evaluation of the integrated process also shows a large reduction in $CO_2$ net capture costs. The net capture cost is decreased with 31 %, to 55 €/ton $CO_2$ compared to the Tel-Tek study. The largest contribution to the $CO_2$ capture cost was from the operating costs, which indicates the importance of reducing the operating costs e.g. by process integration. Before comparing the $CO_2$ capture cost with today's costs of $CO_2$ emissions it is important to remember that transportation and storage costs are not included.

# 9    Future work

The results are based on theoretical energy savings. Retrofit studies should be made to establish the heat exchanger network between the integrated streams and the actual energy savings. Costs for retrofits should be included in the carbon capture process to reduce the contingency of the cost evaluation results. Also, it might be worth to consider a heating water system that removes energy from the carbon capture plant and delivers energy to the cracker. In such a way, long pipes and inappropriate heat exchanging (e.g. flue gases with a reboiler) might be avoided.

The amine by-products, amine waste and amine emissions to air should be studied to determine the impacts on the process and environment. Also, reactions and kinetics should be studied to find process conditions that might be able to avoid the mentioned amine losses. Corrosion problems are also worth to consider as the $MEA-CO_2$ mixture can form acidic solutions. This can have an effect on process material choices.

Moreover, other absorbents and carbon capture processes can be investigated e.g. the chilled-ammonia is a process, which is less energy demanding than the MEA absorption process.

The flowsheets constructed in Aspen Plus do not include all important parts of the process e.g. amine waste treatment as this thesis was focused on energy demands. In future studies a more detailed flowsheet should be simulated to investigate techniques that can solve the drawbacks with the MEA absorption process mentioned in the report.

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**Veltman** Karin, Singh Bhwana (2010): *Human and environmental impact assessment of post combustion $CO_2$ capture focusing on emissions from amine-based scrubbing solvents to air.* Environ. Sci. Technol. vol.44, p.1496 – 1502.

# 11    Appendices

## Appendix 1 – Aspen Plus® simulations

**Property methods**

A property method is essential to model thermodynamic and transport properties of the species in the system. In Aspen Plus® there are special property method packages to chose from, designed for certain systems. In this thesis a MEA property insert package called *kmea* was used to develop the flowsheet. This package is suitable for processes involving MEA and it is based on the existing ELECNRTL property method in Aspen Plus®. ELECNRTL property method is used for non-ideal ionic interactions in liquid solutions. Species in gas phase such as nitrogen and carbon dioxide obey Henry's law. Since nitrogen is not included in the *kmea* insert package it is added as a component in Aspen Plus®. In addition, hydrogen sulfide is a component that is included in the *kmea* insert package, but since the flue gases from the cracker plant do not contain sulphur formation of hydrogen sulfide is not possible and this component is ignored.

The multistage compressors are simulated using the PENG-ROBINSON property method since this unit operation handles gases at high pressures.

**Unit operation models**

The stream names and equipments in this section are referred to the flowsheet in section 6.1, Figure 18.

The process equipments are simulated using a number of unit operation models. These models are presented below to give an introduction of how Aspen Plus® calculates outputs from specified inputs.

*Table 6. The table shows the how the different process units were simulated.*

| Process Unit | Unit Operation Model | Equations |
|---|---|---|
| Flue gas cooler | HEATER | Energy and mass balance |
| Rich Heater | HEATER | Energy and mass balance |
| Lean Cooler | HEATER | Energy and mass balance |
| Rich-Lean heat exchanger | HeatX | Energy and mass balance |

The HEATER unit operation model can calculate the outlet conditions and heat duty from known inlet conditions and two out of three outlet specifications; temperature, pressure or vapour fraction. Energy and mass balances are then used to calculate the output. The HeatX unit operation model requires an input specification such as an outlet temperature of the cold stream outlet temperature.

Two process units are modelled as a FLASH2 unit operation model, the DCC and the water wash in the absorber.

*Table 7. The table shows the how the different process units were simulated.*

| Process Unit | Unit Operation Model | Equations |
|---|---|---|
| DCC | FLASH2 | VLE, energy and mass balance |
| Water wash | FLASH2 | VLE, energy and mass balance |

The FLASH2 unit operation model adds equations for vapor-liquid-equilibrium to determine the composition and outlet conditions. In the FLASH2 model it is not possible to specify the outlet temperature, outlet pressure and the heat duty.

The unit operation models for the blower and the 3-stage compressors are presented in the table below.

*Table 8. The table shows the how the different process units were simulated.*

| Process Unit | Unit Operation Model | Equations |
|---|---|---|
| Blower | COMPR | Energy and mass balance |
| 3-stage compressors | MCOMPR | Energy and mass balance, VLE |

The blower is modeled as an isentropic one-stage compressor using the COMPR unit operation model is used. An energy and mass balance is used to calculate the outlet conditions. It is necessary to specify whether it is an isentropic or polytropic compressor. In addition, the discharge pressure or pressure increase has to be specified. It is also possible to specify the isentropic and mechanical efficiency, which are set as default to 0.7 and 1.0 respectively. The MCOMPR unit operation model is basically the same as COMPR. However, since one compressor stage includes a cooler (HEATER) and eventually a liquid draw out stage (FLASH2) several equations are added.

*Table 9. The table shows the how the different process units were simulated.*

| Process Unit | Unit Operation Model | Equations |
|---|---|---|
| Absorber | RADFRAC | Rate-based non-equilibrium |
| Desorber | RADFRAC | Rate-based non-equilibrium |

The absorber and desorber are modeled as RADFRAC columns. The latest version of Aspen Plus® has a built in *RateFrac* model in the RADFRAC unit operation model. *RateFrac* is a rate-based non-equilibrium model used to simulate fractionation columns. The model simulates the actual packed column with kinetics, mass transfer and heat transfer rates taken into account i.e. the simulations are not based on an equilibrium stage representation and tray efficiencies are not included either. As a result, the predictability is better compared to an equilibrium model (Øi, 2007).

Moreover, the column dimensions and operating conditions need to be specified in Aspen Plus®. It is necessary to specify where the feed and product streams are connected to the column. In addition, the number of stages and the operating pressure of the column have to be specified. Other inputs are the existence of a condenser or a reboiler, column packed bed type and geometry.

The pumps in the process are simulated with a PUMP unit operation model, where energy and mass balances together with a specified outlet pressure generates an output.

*Table 10. The table shows the how the different process units were simulated.*

| Process Unit | Unit Operation Model | Equations |
|---|---|---|
| Rich pump | PUMP | Energy balance |
| Lean pump | PUMP | Energy balance |

*Design specifications*

In order to achieve the wanted specifications in the flowsheet e.g. 85 % $CO_2$ absorption from the flue gases, variables have to be defined to adjust the $CO_2$ absorption rate. Another example is design specification of the $CO_2$ purity of the top product going from the desorber to the compression that is adjusted by a number of variables. In Aspen Plus® it is possible to introduce a design specification that can be fulfilled by defining a variable. These are shown in Table 11.

*Table 11. The flowsheet design specifications to achieve wanted $CO_2$ flow and MEA concentration in the outgoing flue gases, together with its variables.*

| Design Specifications | Stream | Specification | Variable | Stream |
|---|---|---|---|---|
| $CO_2$ mass flow | FG-ABS | CO2out = 0,15 • CO2in | Mass flow | LEAN |
| MEA mass fraction | FG-ATMOS | MEAout/5e-6 = 1 | Mass flow | WATER |

Table 11 shows the design specifications related to the absorber. The first design specification implies that 15 % of the total amount of $CO_2$ fed to the absorber should exit the first packed bed i.e. 85 % absorbed $CO_2$. This is achieved by varying the mass flow of the solvent (stream name: LEAN). The other design specification is that the outlet concentration of vapour MEA in the flue gases vented to the atmosphere should not exceed 5 ppm. This target is reached by varying the mass flow of water fed to the water wash section.

As the design specifications involving the absorber were accomplished the next step was to reach purity and mass flow targets of the desorber top product. These are shown in Table 12.

*Table 12. The desorber design specifications with corresponding variables.*

| Design Specifications | Stream | Specification | Variable | Equipment |
|---|---|---|---|---|
| $CO_2$ mole purity | CO2-H2O | 0,98 | Reflux ratio | DESORBER |
| Mass flow | CO2-H2O | 53,96 [ton/h] | Bottoms to feed ratio | DESORBER |

The first desorber design specification is to have a distillate stream that consists of 98 % $CO_2$ on a molar basis. The reflux ratio is varied to regulate the purity of that stream. The second desorber design specification is to reach a distillate mass flow to prevent accumulation of $CO_2$ in the system by regulating the bottoms to feed ratio. The amount of $CO_2$ that goes out as a top product from the desorber should be the same amount as that absorbed in the absorber.

Finally, there are two additional design specifications cover up the losses of MEA and water in the process. In the process there are losses of MEA with e.g. absorber flue gases and losses of water with the desorber top product. Therefore, make-up streams of MEA and water are used to cover up these losses. The make-up streams are added to the solvent stream going into the absorber.

*Table 13. The design specifications for the lean solution.*

| Design Specifications | Stream | Specification | Variable | Equipment |
|---|---|---|---|---|
| Mass flow | LEANMIX | LEANMIX/LEAN = 1 | MEA mass flow | MKUP-MEA |
| MEA mass fraction | LEANMIX | 0,30 | Water mass flow | MKUP-H2O |

There are two criteria's of the solvent stream. Firstly, it should contain 30 % MEA by weight. Secondly, the mass flow should be that given in the first design spec in Table 11 (LEAN mass flow). To achieve these critera's the mass flow of the stream is adjusted by varying the mass flow of the MEA make-up stream. However, to avoid too high mass fractions of MEA (> 0.30) the mass flow of the make-up water is regulated.

**Reactions**

The reactions presented in section 4.1.1 are divided into elementary steps in the *kmea* insert package. The reactions are presented below.

$$MEA^+ + H_2O \Leftrightarrow MEA + H_3O^+$$

$$2\ H_2O \Leftrightarrow H_3O^+ + OH^-$$

$$HCO_3^- + H_2O \Leftrightarrow H_3O^+ + CO_3^{2-}$$

$$MEACOO^- + H_2O \Leftrightarrow MEA + HCO_3^-$$

$$CO_2 + OH^- \rightarrow HCO_3^-$$

$$HCO_3^- \rightarrow CO_2 + OH^-$$
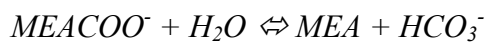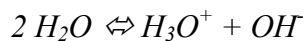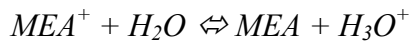
Reactions R5 to R8 are equilibrium reactions while R9 and R10 are kinetic reactions. The *kmea* insert package has a built in expression to calculate equilibrium constants for the reactions (Arrhenius equation). The rate of reaction for the last two reactions is also calculated in this package using activation energies and rate constants.

# Appendix 2 – Economic tables for the Tel-Tek study

Equipment investment cost

Equipment costs for the carbon capture plant in Norwegian Crowns (NOK), 2006. The cost for the steam turbines that drives the blower and the compressors in the $CO_2$ compression is included in the cost "flue gas blower" and "$CO_2$ product compressor" in the table. The cost of these steam turbines are 1000 NOK/kW installed.

| Utstyrs-nr. | Utstyrsnavn | Antall strenger | Beregnet antall utstyrs-enheter pr streng | Installert reserve pr streng | Antall enheter totalt | Mengde pr utstyrs-enhet (dim. Enhet) | Dimen-sjonerende enhet | Materiale | Utstyrs-kostnad kNOK/stk |
|---|---|---|---|---|---|---|---|---|---|
| C104 | Seal Pot | 1 | 1 | 0 | 1 | 543 | m3 | CS | 1 141 |
| ABSORBER | | | | | | | | | |
| C101 | Flue gass absorber | 1 | 1 | 0 | 1 | 2 847 | m3 | CS | 26 773 |
| E105 | Wash Water Cooler | 1 | 1 | 0 | 1 | 136 | m2 | CS | 1 278 |
| G103A/B | Wash Water Circ Pump | 1 | 1 | 1 | 2 | 57 | kW | CS | 219 |
| G105A/B | Flue Gas Humifidier Pump | 1 | 1 | 1 | 2 | 101 | kW | CS | 309 |
| K101 | Flue Gas Blower | 1 | 1 | 0 | 1 | 3 027 | kW | CS | 10 311 |
| C105 | Feed Cooler | 1 | 1 | 0 | 1 | 241 | m2 | CS | 9 665 |
| STRIPPER | | | | | | | | | |
| C102 | Amine Stripper | 1 | 1 | 0 | 1 | 736 | m3 | CS | 11 957 |
| C103 | Stripper Overhead Receiver | 1 | 1 | 0 | 1 | 15 | m3 | CS | 502 |
| C106 | Condensate Flash Drum | 1 | 1 | 0 | 1 | 21 | m3 | CS | 276 |
| D101 | MEA Storage Tank | 1 | 1 | 0 | 1 | 1 339 | m3 | CS | 1 087 |
| D102 | MEA Sump | 1 | 1 | 0 | 1 | 64 | m3 | CS | 302 |
| D103 | Soda Ash pack | 1 | 1 | 0 | 1 | 4 580 | kg | CS | 1 021 |
| E101 | Lean Rich solution exch | 1 | 5 | 0 | 5 | 2 477 | m2 | CS | 1 441 |
| E102 | Lean solution Cooler | 1 | 1 | 0 | 1 | 270 | m2 | TI | 2 159 |
| E103 | Stripper Condenser | 1 | 2 | 0 | 2 | 524 | m2 | TI | 3 801 |
| E104 | Stripper reboiler | 1 | 5 | 0 | 5 | 706 | m2 | CS | 1 089 |
| E107 | Flash steam Condenser | 1 | 1 | 0 | 1 | 160 | m2 | CS | 977 |
| E108 | Condesate Cooler | 1 | 1 | 0 | 1 | 299 | m2 | TI | 2 336 |
| G101A/B | Rich Solution Pump | 1 | 1 | 1 | 2 | 347 | kW | CS | 1 121 |
| G102A/B | Lean Solution Pump | 1 | 1 | 1 | 2 | 408 | kW | CS | 1 468 |
| G104A/B | Regeneartion Reflux Pump | 1 | 1 | 1 | 2 | 13 | kW | CS | 110 |
| G106 | Amine Fill Pump | 1 | 1 | 0 | 1 | 18 | kW | CS | 127 |
| G107 | Amine Sump Pump | 1 | 1 | 0 | 1 | 13 | kW | CS | 110 |
| G108 | Soda Ash Feed Pump | 1 | 1 | 0 | 1 | 2 | kW | CS | 43 |
| G109A/B | Condensate return Pump | 1 | 1 | 1 | 2 | 66 | kW | CS | 234 |
| 0 | Cooling water pump | 1 | 2 | 1 | 3 | 563 | kW | CS | 1 557 |
| C108 | Carbon Drum | 1 | 1 | 0 | 1 | 17 | m3 | CS | 689 |
| T101 | Pre filter | 1 | 1 | 0 | 1 | 152 | kg | CS | 77 |
| T102 | Post filter | 1 | 1 | 0 | 1 | 152 | kg | CS | 77 |
| | | | | | | | | | |
| E106 | Reclaimer | 1 | 1 | 0 | 1 | 209 | m2 | CS | 416 |
| KOMPRESJON | | | | | | | | | |
| E301 | 1 st stg Aftercooler | 1 | 1 | 0 | 1 | 247 | m2 | TI | 1 938 |
| E302 | 2nd Stg Aftercooler | 1 | 1 | 0 | 1 | 136 | m2 | TI | 1 170 |
| E303 | 3 rd Stg Aftercooler | 1 | 1 | 0 | 1 | 69 | m2 | TI | 659 |
| G301 A/B | Condensate pump | 1 | 1 | 1 | 2 | 2 | kW | CS | 50 |
| G302 | CO2 product pump | 1 | 1 | 0 | 1 | 604 | kW | CS | 2 582 |
| K301 | CO2 Product compressor | 1 | 1 | 0 | 1 | 7 772 | kW | CS | 46 600 |
| 0 | Dehydration unit | 1 | 1 | 0 | 1 | 70 | kg/h | CS | 3 540 |
| C301 | 1st stg CO2 comp suction drum | 1 | 1 | 0 | 1 | 11 | m3 | CS | 373 |
| C302 | 2nd stg CO2 comp suction drum | 1 | 1 | 0 | 1 | 5 | m3 | CS | 211 |
| C303 | 3rd stg CO2 comp suction drum | 1 | 1 | 0 | 1 | 2 | m3 | CS | 138 |
| C304 | 4rd stg CO2 comp suction drum | 1 | 1 | 0 | 1 | 15 | m3 | CS | 543 |
| ENERGIVERK | | | | | | | | | |
| 0 | Steamgenerator | 1 | 1 | 0 | 1 | 104 | t/h | CS | 34 692 |
| 0 | Demin. water plant | 1 | 1 | 0 | 1 | 32 | t/h | CS | 3 005 |
| 0 | Deaerator | 1 | 1 | 0 | 1 | 21 | m3 | CS | 532 |
| 0 | BFW pumps | 1 | 1 | 1 | 2 | 604 | kW | CS | 6 027 |
| SOX Scrubber | | | | | | | | | |
| 0 | Sox Scrubber | 1 | 1 | 0 | 1 | 822 787 | Nm3/h | 0,0 | 0 |

Total capital investment (CAPEX)

Costs for the carbon capture plant in Norwegian Crowns (NOK), 2006.

| Prosjekt | Borealis Stenungsund | risstigning | | 0 % | | | | |
|---|---|---|---|---|---|---|---|---|
| Byggestart | mai.10 | Rente | | 0 % | | | | |
| Anl.periode | | 18 v. prosess | | ? | | | | |
| Estimator | Nils Henrik Eldrup | Dato | | 01.10.2008 | | | | |

| Investerings-analyse | Absorber | Absorber | Stripper | Kompresjon | Energiverk | Kanaler | Samlet |
|---|---|---|---|---|---|---|---|
| | kNOK | kNOK | kNOK | kNOK | kNOK | kNOK | kNOK |
| Utstyrskostnader | 50 223 | 0 | 52 942 | 57 852 | 50 282 | 29 082 | 240 381 |
| Montasjekostnader | 3 766 | 0 | 7 361 | 4 273 | 3 406 | 0 | 18 806 |
| Rør inkl. montasje | 18 759 | 0 | 36 670 | 21 286 | 16 966 | 0 | 93 681 |
| Elektro (utstyr & montasje) | 10 246 | 0 | 20 028 | 11 626 | 9 266 | 0 | 51 166 |
| Instrument (utstyr & montasje) | 12 430 | 0 | 24 298 | 14 104 | 11 242 | 0 | 62 075 |
| Grunnarbeider | 3 449 | 0 | 6 741 | 3 913 | 3 119 | 1 997 | 19 219 |
| Stål & betongarbeider. | 10 642 | 0 | 20 803 | 12 075 | 9 625 | 0 | 53 145 |
| Isolasjon av rør og utstyr | 1 953 | 0 | 3 819 | 2 217 | 1 767 | 0 | 9 755 |
| *Fysiske kostnader* | 111 468 | 0 | 172 663 | 127 346 | 105 672 | 31 079 | 548 228 |
| | | | | | | | |
| Prosj. prosess | 2 994 | 0 | 5 853 | 3 398 | 2 708 | 1 734 | 16 687 |
| Prosj. mekanisk | 1 439 | 0 | 2 812 | 1 632 | 1 301 | 833 | 8 017 |
| Prosj. rør | 5 594 | 0 | 10 934 | 6 347 | 5 059 | 0 | 27 934 |
| Prosj. elektro | 2 682 | 0 | 5 243 | 3 043 | 2 426 | 0 | 13 393 |
| Prosj. instrument | 3 850 | 0 | 7 525 | 4 368 | 3 482 | 0 | 19 224 |
| Prosj. grunnarbeider | 565 | 0 | 1 105 | 642 | 511 | 327 | 3 151 |
| Prosj.stål og betong | 1 645 | 0 | 3 215 | 1 866 | 1 487 | 0 | 8 212 |
| Prosj. isolasjon | 317 | 0 | 620 | 360 | 287 | 0 | 1 584 |
| *Prosjektering* | 19 085 | 0 | 37 307 | 21 655 | 17 261 | 2 894 | 98 203 |
| | | | | | | | |
| | | | | | | | |
| Anskaffelse | 2 150 | 0 | 4 202 | 2 439 | 1 944 | 311 | 11 046 |
| Prosjektstyring | 954 | 0 | 1 865 | 1 083 | 863 | 270 | 5 036 |
| Anleggs ledelse | 4 844 | 0 | 9 469 | 5 496 | 4 381 | 1 351 | 25 540 |
| Prosjektledelse | 4 709 | 0 | 9 204 | 5 343 | 4 258 | 675 | 24 189 |
| *Sum adm* | 12 656 | 0 | 24 740 | 14 361 | 11 446 | 2 607 | 65 811 |
| | | | | | | | |
| Uttesting | 1 959 | 0 | 3 830 | 2 223 | 1 772 | 0 | 9 784 |
| | | | | | | | |
| **Sum anleggs kostnader** | 145 169 | 0 | 238 541 | 165 585 | 136 151 | 36 580 | 722 026 |
| | | | | | | | |
| Forventet tillegg | 26 162 | 0 | 51 140 | 29 685 | 23 661 | 7 316 | 137 964 |
| | | | | | | | |
| **Totale kostnader 2006** | 171 331 | 0 | 289 681 | 195 270 | 159 812 | 43 896 | 859 990 |
| | | | | | | | |
| **Lokaliseringskostnader** | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | |
| **Lokale kostnader** | 171 331 | 0 | 289 681 | 195 270 | 159 812 | 43 896 | 859 990 |

Utility and chemicals consumption for the carbon capture plant

**Forbrukstall/time**

| | | Absorber | Sox Scrubber | Stripper | Kompressor | Energi-verk | Kanaler | Totalt |
|---|---|---|---|---|---|---|---|---|
| Elektrisk kraft | kW | 111 | 0 | 631 | 796 | 656 | 0 | 2 194 |
| Ekstra el-kraft/kull | kW | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Tapt el-kraft kull | kW | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| | | | | | | | | |
| Naturgass | Sm3/h | 275 | 0 | 7 571 | 860 | 0 | 0 | 8 706 |
| | | 0 | 0 | 0 | 0 | 0 | 0 | |
| | | 0 | 0 | 0 | 0 | 0 | 0 | |
| Friskt vann | m3/h | 31 | 0 | 0 | 0 | 0 | 0 | 31 |
| Kjølevann | M3/h | 358 | 0 | 7 052 | 638 | 0 | 0 | 8 048 |
| Amin erstatning | kg/h | 30 | 0 | 0 | 0 | 0 | 0 | 30 |
| NA2CO3 | kg/h | 92 | 0 | 0 | 0 | 0 | 0 | 92 |
| Aktivt kull | kg/h | 4 | 0 | 0 | 0 | 0 | 0 | 4 |
| Korrosjonsinhibitor | kg/h | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Destruksjon | kg/h | 83 | 0 | 0 | 0 | 0 | 0 | 83 |
| | | | | | | | | |
| *Investering* | kNOK | 171 331 | 0 | 289 681 | 195 270 | 159 812 | 43 896 | 859 990 |

Total operating costs (OPEX)

Costs for the carbon capture plant in Norwegian Crowns (NOK), 2006.

**Driftskostnader/år**

| *Antall driftstimer pr år* | 8000 | | | | | | | | |
|---|---|---|---|---|---|---|---|---|---|
| | Enhetspris (NOK/enhet) | Enhet | Absorber | SOX Scrubber | Stripper | Kompressor | Energiverk | Kanaler | Totalt |
| | | | kNOK/år | kNOK/år | kNOK/år | kNOK/år | kNOK/år | kNOK/år | kNOK/år |
| Elektrisk kraft | 0,35 | kW | 310 | 0 | 1 766 | 2 230 | 1 838 | 0 | 6 143 |
| Naturgass | 2 | Sm3 | 4 398 | 0 | 121 140 | 13 757 | 0 | 0 | 139 295 |
| Energi kostnader | kNOK | | 4 707 | 0 | 122 906 | 15 987 | 1 838 | 0 | 145 438 |
| | | | | | | | | | |
| Friskt vann | 0,1 | m3 | 25 | 0 | 0 | 0 | 0 | 0 | 25 |
| Kjølevann | 0,01 | m3 | 29 | 0 | 564 | 51 | 0 | 0 | 644 |
| Amin erstatning | 14,4 | kg | 3 435 | 0 | 0 | 0 | 0 | 0 | 3 435 |
| NA2CO3 | 4,64 | kg | 3 427 | 0 | 0 | 0 | 0 | 0 | 3 427 |
| Aktivt kull | 44 | kg | 1 275 | 0 | 0 | 0 | 0 | 0 | 1 275 |
| Korrosjonsinhibitor | 15 | kg | 30 | 0 | 0 | 0 | 0 | 0 | 30 |
| Destruksjon | 2 | kg | 1 336 | 0 | 0 | 0 | 0 | 0 | 1 336 |
| Andre kostnader | kNOK | | 9 557 | 0 | 564 | 51 | 0 | 0 | 10 172 |
| | | | | | | | | | |
| Variable kostnader | kNOK | | 14 264 | 0 | 123 470 | 16 038 | 1 838 | 0 | 155 610 |
| | | | | | | | | | |
| Skiftbemanning | 400 | hours | 821 | 0 | 1 388 | 936 | 766 | 0 | 4 121 |
| Funksjonær | 400 | hours | 137 | 0 | 231 | 156 | 128 | 0 | 687 |
| Vedlikehold | 4 % | | 6 853 | 0 | 11 587 | 7 811 | 6 392 | 1 756 | 34 400 |
| Faste kostnader | kNOK | | 7 811 | 0 | 13 207 | 8 902 | 7 286 | 1 756 | 38 962 |
| | | | | | | | | | |
| Totale kostnader | kNOK | | 22 075 | 0 | 136 677 | 24 941 | 9 124 | 1 756 | 194 572 |
| Prosentfordeling | | | 11 % | 0 % | 70 % | 13 % | 5 % | 1 % | 100 % |
| | | | | | | | | | |
| Totale kostnader | mNOK | | 22 | 0 | 137 | 25 | 9 | 2 | 195 |
| Investering | mNOK | | 171 | 0 | 290 | 195 | 160 | 44 | 860 |

# Appendix 3 – Carbon capture process proposed by Nexant



Figur A2. *Prosessdata – PDF-diagrammer.*

TA-2528/2009.

Innfanget CO2 produkt

MEA tap: 2,61E-6 kg/s

Returkondensat fra CO2 kmopresjon

Vannavløp

Massestrøm H2O: 0,91 kg/s
MEA tap: 0,000011 kg/s

Dampforbruk: 2,20 kg/kg CO2

Damp, 4 bar a

Varme: -42 521 kW

Dampforbruk: 36,71 kg/s
Varme: 78 340 kW

Molfraksjon CO2: 2,11 %
Molfraksjon H2O: 86,58 %
Molfraksjon MEA: 11,31 %

Varme: 23 077 kW

Effekt: 2 590 W

Effekt: 4 kW

Effekt: 0,5 kW

Effekt: 40 kW

CO2 ladning: 0,186 mol CO2/mol MEA

Molfraksjon CO2: 2,05 %
Molfraksjon H2O: 86,92 %
Molfraksjon MEA: 11,03 %

Mager løsning til absorpsjon

MEA løsning justeringstilsats

Vann til justeringsløsningens vanninnhold

Massestrøm H2O: 6,64 kg/s
Massestrøm MEA: 0,34 kg/s
Inklusive fersk MEA: 32E-3 kg/s

Molfraksjon CO2: 4,89 %
Molfraksjon H2O: 84,17 %
Molfraksjon MEA: 10,94 %

Rik løsning fra desorpsjon

CO2 ladning: 0,447 mol CO2/mol MEA

CO2-fangst eksempel 1 - Desorberdel

Demonstrasjons-eksempel

| | 1,013 | Trykk [bar a] |
| | 41 | Temperatur [°C] |
| | 44.858 | Molarstrøm [kmol/h] |
| | 1 244 | Massestrøm [t/h] |

| P-8 | P-3 | H-4 | H-3 | P-4 | X-2 | T-2 | H-6 | H-5 | H-9 | P-10 | V-10 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| (Justerings-pumpe for løsning) | Pumpe for rik løsning | Mager løsning kjøler | Varmeveksler rik til mager løsning | Pumpe for mager løsning | Kaustisk soda pakkeenhet | CO2 Desorp-sjons-kolonne | Descorpsjons-koker | Desorber kondensator | Reclaimer | Reflukspumpe | Desorber reflukstank |

Figur A3. Prosessdata – PDF-diagrammer.

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TA-2528/2009.

**Figur A4.** *Prosessdata – PDF-diagrammer.*

# Appendix 4 – Base case verification stream data results from Aspen Plus® simulations

| | COOLEDFG | FG-ABS | FG-ATMOS | FLUEGAS | H2O-MEA |
|---|---|---|---|---|---|
| | | | | | |
| Temperature C | 47 | 57 | 55,8 | 80 | 35 |
| Pressure   bar | 1,02 | 1,02 | 1,013 | 1,023 | 1,02 |
| Vapor Frac | 1 | 1 | 1 | 1 | 0 |
| Mole Flow  kmol/hr | 40796,734 | 41278,609 | 41251,989 | 39832 | 1530,031 |
| Mass Flow  tonne/hr | 1145,042 | 1118,619 | 1118,027 | 1127,662 | 35,057 |
| Volume Flow cum/hr | 1,06E+06 | 1,11E+06 | 1,11E+06 | 1,14E+06 | 35,655 |
| Enthalpy   MMkcal/hr | -390,161 | -367,142 | -365,959 | -325,197 | -109,319 |
| Mass Flow  tonne/hr | | | | | |
| H2O | 77,227 | 1,10E+02 | 110,003 | 59,847 | 25,245 |
| MEA | | 0,238 | 0,108 | | 2,043 |
| CO2 | 69,769 | 10,465 | 10,431 | 69,769 | |
| N2 | 839,999 | 839,991 | 839,991 | 839,999 | |
| O2 | 158,047 | 1,57E+02 | 157,494 | 158,047 | 0,018 |
| HCO3- | | | | | 0,074 |
| MEACOO- | | | | | 4,646 |
| MEA+ | | | | | 2,97 |
| CO3-2 | | | | | 0,06 |
| H3O+ | | | | | |
| OH- | | | | | |
| Mass Frac | | | | | |
| H2O | 0,067 | 0,099 | 0,098 | 0,053 | 0,72 |
| MEA | | | | | 0,058 |
| CO2 | 0,061 | 9,00E-03 | 0,009 | 0,062 | |
| N2 | 0,734 | 7,51E-01 | 0,751 | 0,745 | |
| O2 | 0,138 | 0,141 | 0,141 | 0,14 | 0,001 |
| HCO3- | | | | | 0,002 |
| MEACOO- | | | | | 0,133 |
| MEA+ | | | | | 0,085 |
| CO3-2 | | | | | 0,002 |
| H3O+ | | | | | |
| OH- | | | | | |
| Mole Flow  kmol/hr | | | | | |
| H2O | 4286,734 | 6,13E+03 | 6106,098 | 3321,989 | 1401,3 |
| MEA | | 3,90E+00 | 1,771 | | 33,451 |
| CO2 | 1585,313 | 2,38E+02 | 237,005 | 1585,314 | |
| N2 | 29985,529 | 29985,241 | 29985,241 | 29985,53 | 0,01 |
| O2 | 4939,157 | 4921,883 | 4921,874 | 4939,168 | 0,553 |
| HCO3- | | | | | 1,214 |
| MEACOO- | | | | | 44,634 |
| MEA+ | | | | | 47,861 |
| CO3-2 | | | | | 1,005 |
| H3O+ | | | | | |
| OH- | | | | | 0,002 |
| Mole Frac | | | | | |
| H2O | 0,105 | 1,48E-01 | 0,148 | 0,083 | 0,916 |
| MEA | | | | | 0,022 |
| CO2 | 0,039 | 0,006 | 0,006 | 0,04 | |
| N2 | 0,735 | 0,726 | 0,727 | 0,753 | |
| O2 | 0,121 | 0,119 | 0,119 | 0,124 | |
| HCO3- | | | | | 0,001 |
| MEACOO- | | | | | 0,029 |
| MEA+ | | | | | 0,031 |
| CO3-2 | | | | | 0,001 |
| H3O+ | | | | | |
| OH- | | | | | |

|  | 1 | 2 | ABSIN | CO2-H2O | CO2-PROD |
|---|---|---|---|---|---|
|  |  |  |  |  |  |
| Temperature C |  |  | 56,6 | 9,8 | 25 |
| Pressure   bar |  |  | 1,102 | 1,7 | 75 |
| Vapor Frac |  |  | 1 | 1 | 0 |
| Mole Flow   kmol/hr | 0 | 0 | 40796,734 | 1385,507 | 1385,438 |
| Mass Flow   tonne/hr | 0 | 0 | 1145,042 | 60,5 | 60,5 |
| Volume Flow cum/hr | 0 | 0 | 1,02E+06 | 19171,345 | 85,208 |
| Enthalpy   MMkcal/hr |  |  | -387,345 | -128,388 | -131,523 |
| Mass Flow   tonne/hr |  |  |  |  |  |
| H2O | 0 | 0 | 77,227 | 1,83E-01 | 0,18 |
| MEA | 0 | 0 | 0 | 0 | 0 |
| CO2 | 0 | 0 | 69,769 | 59,756 | 59,753 |
| N2 | 0 | 0 | 839,999 | 0,008 | 0,008 |
| O2 | 0 | 0 | 158,047 | 0,553 | 0,553 |
| HCO3- | 0 | 0 |  |  | 0,004 |
| MEACOO- | 0 | 0 |  |  |  |
| MEA+ | 0 | 0 |  |  |  |
| CO3-2 | 0 | 0 |  |  |  |
| H3O+ | 0 | 0 |  | 0,00E+00 | 0,001 |
| OH- | 0 | 0 |  |  |  |
| Mass Frac |  |  |  |  |  |
| H2O |  |  | 0,067 | 0,003 | 0,003 |
| MEA |  |  |  |  |  |
| CO2 |  |  | 0,061 | 0,988 | 0,988 |
| N2 |  |  | 0,734 |  |  |
| O2 |  |  | 0,138 | 0,009 | 0,009 |
| HCO3- |  |  |  |  |  |
| MEACOO- |  |  |  |  |  |
| MEA+ |  |  |  |  |  |
| CO3-2 |  |  |  | 0,00E+00 |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow   kmol/hr |  |  |  |  |  |
| H2O | 0 | 0 | 4286,734 | 10,148 | 10,009 |
| MEA | 0 | 0 |  |  |  |
| CO2 | 0 | 0 | 1585,313 | 1357,797 | 1357,727 |
| N2 | 0 | 0 | 29985,529 | 0,288 | 0,288 |
| O2 | 0 | 0 | 4939,157 | 17,274 | 17,274 |
| HCO3- | 0 | 0 |  | 0,00E+00 | 0,069 |
| MEACOO- | 0 | 0 |  |  |  |
| MEA+ | 0 | 0 |  |  |  |
| CO3-2 | 0 | 0 |  |  |  |
| H3O+ | 0 | 0 |  |  | 0,069 |
| OH- | 0 | 0 |  |  |  |
| Mole Frac |  |  |  |  |  |
| H2O | 0 | 0 | 0,105 | 0,007 | 0,007 |
| MEA | 0 | 0 |  |  |  |
| CO2 | 0 | 0 | 0,039 | 0,98 | 0,98 |
| N2 | 0 | 0 | 0,735 |  |  |
| O2 | 0 | 0 | 0,121 | 0,012 | 0,012 |
| HCO3- | 0 | 0 |  |  |  |
| MEACOO- | 0 | 0 |  |  |  |
| MEA+ | 0 | 0 |  |  |  |
| CO3-2 | 0 | 0 |  |  |  |
| H3O+ | 0 | 0 |  |  |  |
| OH- | 0 | 0 |  |  |  |

| | H2O-MEA2 | LEAN | LEAN-HX | LEAN-P2 | LEAN-REB |
|---|---|---|---|---|---|
| | | | | | |
| **Temperature C** | 55,8 | 38 | 53,3 | 53,3 | 122,2 |
| **Pressure   bar** | 1,013 | 1,3 | 1,8 | 1,9 | 1,9 |
| **Vapor Frac** | 0 | 0 | 0 | 0 | 0 |
| **Mole Flow   kmol/hr** | 1555,861 | 43962,057 | 42104,957 | 42104,957 | 42105,743 |
| **Mass Flow   tonne/hr** | 35,648 | 1045,661 | 1011,586 | 1011,585 | 1011,586 |
| **Volume Flow cum/hr** | 3,66E+01 | 1008,505 | 981,931 | 981,93 | 1036,013 |
| **Enthalpy   MMkcal/hr** | -110,503 | -3085,712 | -2944,828 | -2944,824 | -2882,717 |
| **Mass Flow   tonne/hr** | | | | | |
| **H2O** | 25,669 | 702,233 | 668,941 | 668,941 | 668,818 |
| **MEA** | 2,109 | 190,515 | 191,67 | 191,67 | 192,966 |
| **CO2** | | | | | 0,035 |
| **N2** | | | | | |
| **O2** | 0,018 | | | | |
| **HCO3-** | 0,106 | 0,539 | 0,677 | 0,677 | 1,873 |
| **MEACOO-** | 4,714 | 92,492 | 91,969 | 91,969 | 91,171 |
| **MEA+** | 2,995 | 58,521 | 57,417 | 57,417 | 56,576 |
| **CO3-2** | 0,037 | 1,358 | 0,907 | 0,907 | 0,143 |
| **H3O+** | 0,00E+00 | | | | |
| **OH-** | 0 | 0,004 | 0,005 | 0,005 | 0,004 |
| **Mass Frac** | | | | | |
| **H2O** | 0,72 | 0,672 | 0,661 | 0,661 | 0,661 |
| **MEA** | 0,059 | 0,182 | 0,189 | 0,189 | 0,191 |
| **CO2** | 0 | | | | |
| **N2** | 0 | | | | |
| **O2** | 0,001 | | | | |
| **HCO3-** | 0,003 | 0,001 | 0,001 | 0,001 | 0,002 |
| **MEACOO-** | 0,132 | 0,088 | 0,091 | 0,091 | 0,09 |
| **MEA+** | 0,084 | 0,056 | 0,057 | 0,057 | 0,056 |
| **CO3-2** | 1,00E-03 | 0,001 | 0,001 | 0,001 | |
| **H3O+** | | | | | |
| **OH-** | | | | | |
| **Mole Flow   kmol/hr** | | | | | |
| **H2O** | 1424,859 | 38979,829 | 37131,881 | 37131,882 | 37125,031 |
| **MEA** | 34,523 | 3118,909 | 3137,825 | 3137,826 | 3159,037 |
| **CO2** | 0,002 | 0,001 | 0,004 | 0,004 | 0,79 |
| **N2** | 0,01 | | | | |
| **O2** | 0,563 | | | | |
| **HCO3-** | 1,74E+00 | 8,834 | 11,091 | 11,092 | 30,698 |
| **MEACOO-** | 45,291 | 888,615 | 883,596 | 883,597 | 875,927 |
| **MEA+** | 48,26 | 942,976 | 925,18 | 925,178 | 911,638 |
| **CO3-2** | 0,615 | 22,633 | 15,113 | 15,111 | 2,389 |
| **H3O+** | | | | | |
| **OH-** | 0,002 | 0,26 | 0,267 | 0,267 | 0,234 |
| **Mole Frac** | | | | | |
| **H2O** | 0,916 | 0,887 | 0,882 | 0,882 | 0,882 |
| **MEA** | 0,022 | 0,071 | 0,075 | 0,075 | 0,075 |
| **CO2** | | | | | |
| **N2** | | | | | |
| **O2** | | | | | |
| **HCO3-** | 0,001 | | | | 0,001 |
| **MEACOO-** | 0,029 | 0,02 | 0,021 | 0,021 | 0,021 |
| **MEA+** | 0,031 | 0,021 | 0,022 | 0,022 | 0,022 |
| **CO3-2** | | 0,001 | | | |
| **H3O+** | | | | | |
| **OH-** | | | | | |

|  | LEANCOOL | LEANMIX | MKUP-H2O | MKUP-MEA | RICH-ABS |
|---|---|---|---|---|---|
|  |  |  |  |  |  |
| Temperature C | 38 | 38,3 | 38 | 38 | 51,3 |
| Pressure bar | 1,3 | 1,3 | 1,3 | 1,3 | 1,02 |
| Vapor Frac |  |  |  |  |  |
| Mole Flow kmol/hr | 42104,955 | 43292,774 | 916,3 | 271,519 | 42133,053 |
| Mass Flow tonne/hr | 1011,585 | 1044,678 | 16,507 | 16,585 | 1072,042 |
| Volume Flow cum/hr | 973,706 | 1001,425 | 16,623 | 10,954 | 1129,144 |
| Enthalpy MMkcal/hr | -2958,216 | -3038,03 | -62,34 | -17,474 | -3105,91 |
| Mass Flow tonne/hr |  |  |  |  |  |
| H2O | 668,87 | 685,382 | 16,507 |  | 667,186 |
| MEA | 191,501 | 208,065 |  | 16,585 | 32,289 |
| CO2 |  |  |  |  | 0,017 |
| N2 |  |  |  |  | 0,008 |
| O2 |  |  |  |  | 0,553 |
| HCO3- | 0,508 | 0,487 |  |  | 7,105 |
| MEACOO- | 91,558 | 91,583 |  |  | 222,035 |
| MEA+ | 57,833 | 57,841 |  |  | 141,794 |
| CO3-2 | 1,311 | 1,316 |  |  | 1,054 |
| H3O+ |  |  |  |  |  |
| OH- | 0,004 | 0,005 |  |  |  |
| Mass Frac |  |  |  |  |  |
| H2O | 0,661 | 0,656 | 1 |  | 0,622 |
| MEA | 0,189 | 0,199 |  | 1 | 0,03 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  | 0,001 |
| HCO3- | 0,001 |  |  |  | 0,007 |
| MEACOO- | 0,091 | 0,088 |  |  | 0,207 |
| MEA+ | 0,057 | 0,055 |  |  | 0,132 |
| CO3-2 | 0,001 | 0,001 |  |  | 0,001 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow kmol/hr |  |  |  |  |  |
| H2O | 37127,941 | 38044,461 | 916,3 |  | 37034,477 |
| MEA | 3135,064 | 3406,22 |  | 271,519 | 528,6 |
| CO2 | 0,001 | 0,001 |  |  | 0,39 |
| N2 |  |  |  |  | 0,288 |
| O2 |  |  |  |  | 17,274 |
| HCO3- | 8,32 | 7,983 |  |  | 116,439 |
| MEACOO- | 879,643 | 879,889 |  |  | 2133,206 |
| MEA+ | 931,895 | 932,012 |  |  | 2284,795 |
| CO3-2 | 21,84 | 21,932 |  |  | 17,566 |
| H3O+ |  |  |  |  |  |
| OH- | 0,251 | 0,277 |  |  | 0,018 |
| Mole Frac |  |  |  |  |  |
| H2O | 0,882 | 0,879 | 1 |  | 0,879 |
| MEA | 0,074 | 0,079 |  | 1 | 0,013 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- |  |  |  |  | 0,003 |
| MEACOO- | 0,021 | 0,02 |  |  | 0,051 |
| MEA+ | 0,022 | 0,022 |  |  | 0,054 |
| CO3-2 | 0,001 | 0,001 |  |  |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |

|  | RICH-HX | RICH-P1 | WATER | WATER2 |
|---|---|---|---|---|
|  |  |  |  |  |
| **Temperature C** | 107,3 | 51,3 | 35 | 47 |
| **Pressure   bar** | 1,9 | 2 | 1,1 | 1,02 |
| **Vapor Frac** | 0,015 | 0 | 0 | 0 |
| **Mole Flow   kmol/hr** | 42385,515 | 42133,054 | 989,52 | 24,786 |
| **Mass Flow   tonne/hr** | 1072,042 | 1072,042 | 17,826 | 0,447 |
| **Volume Flow cum/hr** | 11557,31 | 1129,124 | 17,932 | 0,452 |
| **Enthalpy   MMkcal/hr** | -3043,767 | -3105,879 | -67,374 | -1,682 |
| **Mass Flow   tonne/hr** |  |  |  |  |
| **H2O** | 667,376 | 667,186 | 17,826 | 0,446 |
| **MEA** | 63,299 | 32,291 |  |  |
| **CO2** | 11,128 | 0,017 |  |  |
| **N2** | 0,008 | 0,008 |  |  |
| **O2** | 0,553 | 0,553 |  |  |
| **HCO3-** | 7,274 | 7,107 |  |  |
| **MEACOO-** | 196,853 | 222,033 |  |  |
| **MEA+** | 125,302 | 141,794 |  |  |
| **CO3-2** | 0,256 | 1,053 |  |  |
| **H3O+** |  |  |  |  |
| **OH-** | 0,001 |  |  |  |
| **Mass Frac** |  |  |  |  |
| **H2O** | 0,623 | 0,622 | 1 | 0,999 |
| **MEA** | 0,059 | 0,03 |  |  |
| **CO2** | 0,01 |  |  |  |
| **N2** |  |  |  |  |
| **O2** | 0,001 | 0,001 |  | 0,001 |
| **HCO3-** | 0,007 | 0,007 |  |  |
| **MEACOO-** | 0,184 | 0,207 |  |  |
| **MEA+** | 0,117 | 0,132 |  |  |
| **CO3-2** |  | 0,001 |  |  |
| **H3O+** |  |  |  |  |
| **OH-** |  |  |  |  |
| **Mole Flow   kmol/hr** |  |  |  |  |
| **H2O** | 37044,989 | 37034,459 | 989,52 | 24,775 |
| **MEA** | 1036,273 | 528,631 |  |  |
| **CO2** | 252,851 | 0,39 |  |  |
| **N2** | 0,288 | 0,288 |  |  |
| **O2** | 17,274 | 17,274 |  | 0,011 |
| **HCO3-** | 119,208 | 116,468 |  |  |
| **MEACOO-** | 1891,276 | 2133,188 |  |  |
| **MEA+** | 2019,053 | 2284,783 |  |  |
| **CO3-2** | 4,266 | 17,554 |  |  |
| **H3O+** |  |  |  |  |
| **OH-** | 0,037 | 0,018 |  |  |
| **Mole Frac** |  |  |  |  |
| **H2O** | 0,874 | 0,879 | 1 | 1 |
| **MEA** | 0,024 | 0,013 |  |  |
| **CO2** | 0,006 |  |  |  |
| **N2** |  |  |  |  |
| **O2** |  |  |  |  |
| **HCO3-** | 0,003 | 0,003 |  |  |
| **MEACOO-** | 0,045 | 0,051 |  |  |
| **MEA+** | 0,048 | 0,054 |  |  |
| **CO3-2** |  |  |  |  |
| **H3O+** |  |  |  |  |
| **OH-** |  |  |  |  |

Verification of the Aspen Plus model

The parameters shown in the table below were used to verify the simulation of the Nexant propsed capture process. These parameters were chosen since they were considered to be important and have a significant impact on the process simulations.

| | | Aspen Plus | Aspen Hysys | Deviation (percent) |
|---|---|---|---|---|
| Equipment | | Value | Value | Value |
| Absorber | | | | |
| | Number of absorption sections | 8 | 8 | - |
| | Number of washing sections | 3 | 3 | - |
| | Lean mass flow [ton/h] | 1046 | 1112 | 6 % |
| | Alfa [mol CO2/mol MEA] | 0,186 | 0,186 | - |
| Desorber | | | | |
| | Number of desorption sections | 19 | 19 | - |
| | Number of washing sections | 3 | 3 | - |
| | Condenser duty [MW] | 40,6 | 42,5 | 4 % |
| | Reboiler duty [MW] | 78,6 | 78,3 | 0,4 % |
| Blower | | | | |
| | Work [MW] | 3,3 | 3,1 | 6 % |
| CO2 compression | | | | |
| | Number of compression stages | 3 | 3 | - |
| | Number of cooling stages | 3 | 3 | - |
| | Total work [MW] | 5,3 | 3,5 | 51 % |
| | Total cooling duty [MW] | 7,8 | 7,2 | 8 % |

The blower duty, lean mass flow rate, desorber condenser and reboiler, and the work and cooling duty in the $CO_2$ compression were compared between the simulation tools to verify the model constructed in Aspen Plus. The average overall deviation between the Aspen Plus and Aspen Hysys model was 13 %. Except the $CO_2$ compression, which resulted in a high deviation, the average deviation was 4 %. The $CO_2$ compression in the Aspen Plus model was modelled as a MCOMPR unit (Multistage Compressor), whereas in the Aspen Hysys model each compressor was modelled separately (as can be seen in Appendix 3, page 64). The high deviation could be a result of the different simulation approach in the $CO_2$ compression. Comparing the total work given in the table with the total compressor work in the Tel-Tek study made at Borealis (see Appendix 2, page 60) it is 7.8 MW.

The deviations can depend upon several reasons:
- The columns in the Aspen Plus model are rate (kinetic) based compared to the equilibrium based (with Murphree efficiency) columns in the Aspen Hysys model. The rate based model also considers mass and heat transfer.
- The blower isentropic and mechanical efficiency are set by default to 0.72 and 1 respectively (not presented in the Aspen Hysys model).
- The property method for the $CO_2$ compression section used in the Aspen Plus simulation was the PENG – ROBINSON equation of state. Information about property method for the carbon capture process in the Aspen Hysys simulation was not available (more about property methods used in the Aspen Plus simulations are described in Appendix 1, page 56).

The deviations in total work duty in the multistage compressors could not be studied further in detail as time was not enough. For further reading in differences between Aspen Hysys and Aspen Plus the reader is referred to (Øi, 2007).

# Appendix 5 – Flowsheet and stream data for case 1
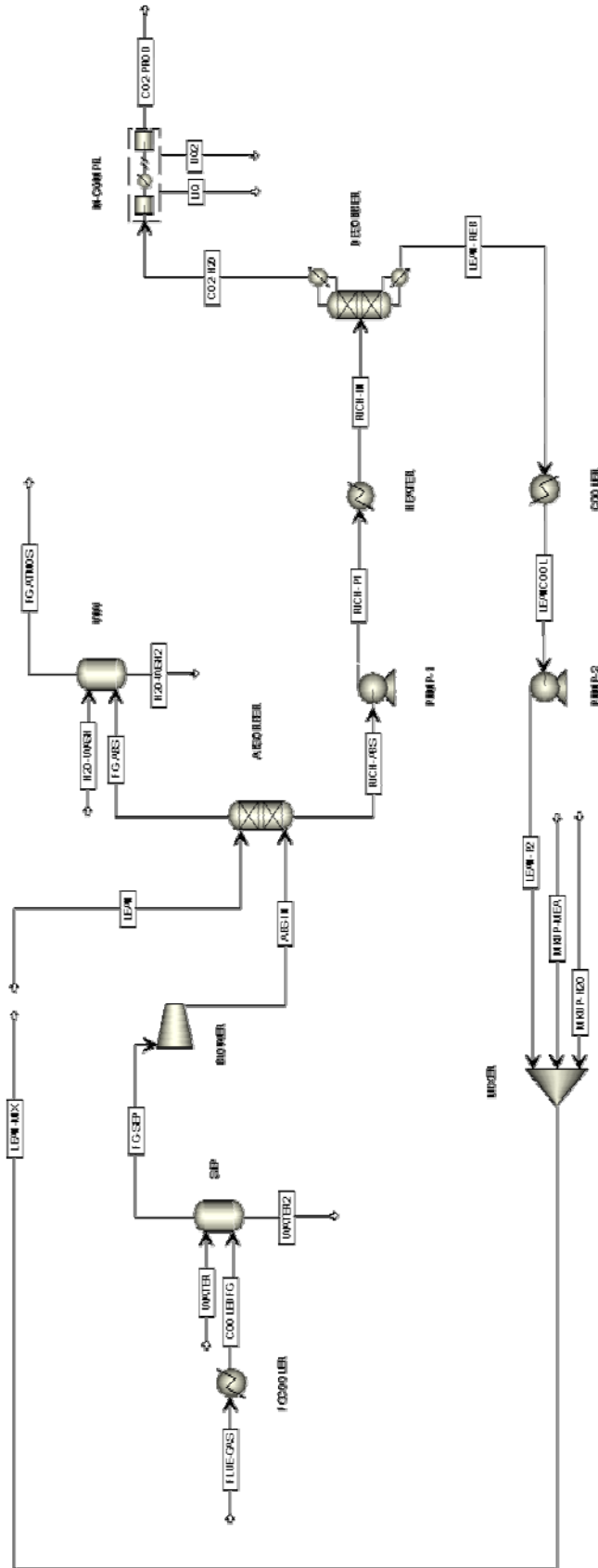
|                         | ABS-IN     | CO2-H2O   | CO2-PROD  | COOLEDFG   | FG-ABS      |
|-------------------------|------------|-----------|-----------|------------|-------------|
|                         |            |           |           |            |             |
| Temperature  C          | 56,4       | 23,6      | 25        | 80         | 59,5        |
| Pressure   bar          | 1,11       | 1,7       | 75        | 1,02       | 1,02        |
| Vapor Frac              | 1          | 1         | 0         | 1          | 1           |
| Mole Flow  kmol/hr      | 25165,93   | 1239,788  | 1223,754  | 28544      | 25570,451   |
| Mass Flow  tonne/hr     | 703,641    | 53,96     | 53,672    | 764,505    | 679,48      |
| Volume Flow cum/hr      | 621278,686 | 17991,601 | 74,039    | 821676,998 | 693346,102  |
| Enthalpy   MMkcal/hr    | -281,215   | -115,466  | -117,517  | -470,523   | -260,171    |
| Mass Flow  tonne/hr     |            |           |           |            |             |
| H2O                     | 47,654     | 0,392     | 0,102     | 108,502    | 76,795      |
| MEA                     |            |           |           |            | 0,167       |
| CO2                     | 62,807     | 53,472    | 53,47     | 62,811     | 9,435       |
| N2                      | 574,924    | 0,008     | 0,008     | 574,924    | 574,916     |
| O2                      | 18,257     | 0,089     | 0,089     | 18,267     | 18,168      |
| HCO3-                   |            |           | 0,002     |            |             |
| MEACOO-                 |            |           |           |            |             |
| MEA+                    |            |           |           |            |             |
| CO3-2                   |            |           |           |            |             |
| H3O+                    |            |           | 0,001     |            |             |
| OH-                     |            |           |           |            |             |
| Mass Frac               |            |           |           |            |             |
| H2O                     | 0,068      | 0,007     | 0,002     | 0,142      | 0,113       |
| MEA                     |            |           |           |            |             |
| CO2                     | 0,089      | 0,991     | 0,996     | 0,082      | 0,014       |
| N2                      | 0,817      |           |           | 0,752      | 0,846       |
| O2                      | 0,026      | 0,002     | 0,002     | 0,024      | 0,027       |
| HCO3-                   |            |           |           |            |             |
| MEACOO-                 |            |           |           |            |             |
| MEA+                    |            |           |           |            |             |
| CO3-2                   |            |           |           |            |             |
| H3O+                    |            |           |           |            |             |
| OH-                     |            |           |           |            |             |
| Mole Flow  kmol/hr      |            |           |           |            |             |
| H2O                     | 2645,172   | 21,733    | 5,661     | 6022,784   | 4262,756    |
| MEA                     |            |           |           |            | 2,729       |
| CO2                     | 1427,111   | 1214,993  | 1214,952  | 1427,2     | 214,382     |
| N2                      | 20523,105  | 0,277     | 0,277     | 20523,136  | 20522,827   |
| O2                      | 570,542    | 2,786     | 2,786     | 570,88     | 567,757     |
| HCO3-                   |            |           | 0,039     |            |             |
| MEACOO-                 |            |           |           |            |             |
| MEA+                    |            |           |           |            |             |
| CO3-2                   |            |           |           |            |             |
| H3O+                    |            |           | 0,039     |            |             |
| OH-                     |            |           |           |            |             |
| Mole Frac               |            |           |           |            |             |
| H2O                     | 0,105      | 0,018     | 0,005     | 0,211      | 0,167       |
| MEA                     |            |           |           |            |             |
| CO2                     | 0,057      | 0,98      | 0,993     | 0,05       | 0,008       |
| N2                      | 0,816      |           |           | 0,719      | 0,803       |
| O2                      | 0,023      | 0,002     | 0,002     | 0,02       | 0,022       |
| HCO3-                   |            |           |           |            |             |
| MEACOO-                 |            |           |           |            |             |
| MEA+                    |            |           |           |            |             |
| CO3-2                   |            |           |           |            |             |
| H3O+                    |            |           |           |            |             |
| OH-                     |            |           |           |            |             |

|  | FG-ATMOS | FG-SEP | FLUE-GAS | H2O-WASH | H2O-WSH2 |
|---|---|---|---|---|---|
| | | | | | |
| **Temperature C** | 54 | 47 | 144 | 35 | 54 |
| **Pressure   bar** | 1,013 | 1,02 | 1,02 | 1,02 | 1,013 |
| **Vapor Frac** | 1 | 1 | 1 | 0 | 0 |
| **Mole Flow   kmol/hr** | 25021,633 | 25165,93 | 28544 | 19185,541 | 19733,244 |
| **Mass Flow   tonne/hr** | 669,447 | 703,641 | 764,505 | 353,77 | 363,803 |
| **Volume Flow cum/hr** | 671927,651 | 656740,154 | 970586,322 | 355,957 | 368,876 |
| **Enthalpy   MMkcal/hr** | -229,585 | -282,926 | -457,055 | -1312,478 | -1343,065 |
| **Mass Flow   tonne/hr** | | | | | |
| **H2O** | 66,977 | 47,654 | 108,502 | 342,917 | 352,688 |
| **MEA** | 0,003 | | | 2,023 | 2,286 |
| **CO2** | 9,384 | 62,807 | 62,811 | 0,001 | 0,002 |
| **N2** | 574,916 | 574,924 | 574,924 | 0,004 | 0,004 |
| **O2** | 18,166 | 18,257 | 18,267 | 0,044 | 0,045 |
| **HCO3-** | | | | 0,939 | 1,176 |
| **MEACOO-** | | | | 3,959 | 3,807 |
| **MEA+** | | | | 3,699 | 3,687 |
| **CO3-2** | | | | 0,185 | 0,107 |
| **H3O+** | | | | | |
| **OH-** | | | | | |
| **Mass Frac** | | | | | |
| **H2O** | 0,1 | 0,068 | 0,142 | 0,969 | 0,969 |
| **MEA** | | | | 0,006 | 0,006 |
| **CO2** | 0,014 | 0,089 | 0,082 | | |
| **N2** | 0,859 | 0,817 | 0,752 | | |
| **O2** | 0,027 | 0,026 | 0,024 | | |
| **HCO3-** | | | | 0,003 | 0,003 |
| **MEACOO-** | | | | 0,011 | 0,01 |
| **MEA+** | | | | 0,01 | 0,01 |
| **CO3-2** | | | | 0,001 | |
| **H3O+** | | | | | |
| **OH-** | | | | | |
| **Mole Flow   kmol/hr** | | | | | |
| **H2O** | 3717,807 | 2645,172 | 6022,784 | 19034,775 | 19577,145 |
| **MEA** | 0,055 | | | 33,111 | 37,426 |
| **CO2** | 213,228 | 1427,111 | 1427,2 | 0,016 | 0,055 |
| **N2** | 20522,824 | 20523,105 | 20523,136 | 0,142 | 0,146 |
| **O2** | 567,72 | 570,542 | 570,88 | 1,38 | 1,417 |
| **HCO3-** | | | | 15,392 | 19,268 |
| **MEACOO-** | | | | 38,034 | 36,575 |
| **MEA+** | | | | 59,599 | 59,417 |
| **CO3-2** | | | | 3,079 | 1,778 |
| **H3O+** | | | | | |
| **OH-** | | | | 0,013 | 0,016 |
| **Mole Frac** | | | | | |
| **H2O** | 0,149 | 0,105 | 0,211 | 0,992 | 0,992 |
| **MEA** | | | | 0,002 | 0,002 |
| **CO2** | 0,009 | 0,057 | 0,05 | | |
| **N2** | 0,82 | 0,816 | 0,719 | | |
| **O2** | 0,023 | 0,023 | 0,02 | | |
| **HCO3-** | | | | 0,001 | 0,001 |
| **MEACOO-** | | | | 0,002 | 0,002 |
| **MEA+** | | | | 0,003 | 0,003 |
| **CO3-2** | | | | | |
| **H3O+** | | | | | |
| **OH-** | | | | | |

|  | LEAN | LEAN-MIX | LEAN-P2 | LEAN-REB | LEANCOOL |
|---|---|---|---|---|---|
| Temperature C | 38 | 38,1 | 38 | 122,2 | 38 |
| Pressure   bar | 1,3 | 1,3 | 1,9 | 1,9 | 1,8 |
| Vapor Frac | 0 | 0 | 0 | 0 | 0 |
| Mole Flow   kmol/hr | 38188,267 | 38200,313 | 36546,221 | 36546,923 | 36546,221 |
| Mass Flow   tonne/hr | 908,329 | 908,329 | 878,53 | 878,53 | 878,53 |
| Volume Flow cum/hr | 876,052 | 876,011 | 846,043 | 900,152 | 846,044 |
| Enthalpy   MMkcal/hr | -2680,447 | -2681,038 | -2568,503 | -2502,96 | -2568,507 |
| Mass Flow   tonne/hr |  |  |  |  |  |
| H2O | 610,004 | 610,271 | 580,49 | 580,442 | 580,49 |
| MEA | 165,493 | 165,602 | 165,581 | 166,865 | 165,581 |
| CO2 |  |  |  | 0,031 |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 0,468 | 0,468 | 0,446 | 1,642 | 0,446 |
| MEACOO- | 80,344 | 80,115 | 80,212 | 79,862 | 80,213 |
| MEA+ | 50,835 | 50,692 | 50,655 | 49,559 | 50,655 |
| CO3-2 | 1,18 | 1,177 | 1,142 | 0,125 | 1,142 |
| H3O+ |  |  |  |  |  |
| OH- | 0,004 | 0,004 | 0,004 | 0,003 | 0,004 |
| Mass Frac |  |  |  |  |  |
| H2O | 0,672 | 0,672 | 0,661 | 0,661 | 0,661 |
| MEA | 0,182 | 0,182 | 0,188 | 0,19 | 0,188 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 0,001 | 0,001 | 0,001 | 0,002 | 0,001 |
| MEACOO- | 0,088 | 0,088 | 0,091 | 0,091 | 0,091 |
| MEA+ | 0,056 | 0,056 | 0,058 | 0,056 | 0,058 |
| CO3-2 | 0,001 | 0,001 | 0,001 |  | 0,001 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow   kmol/hr |  |  |  |  |  |
| H2O | 33860,384 | 33875,216 | 32222,074 | 32219,429 | 32222,076 |
| MEA | 2709,285 | 2711,072 | 2710,724 | 2731,749 | 2710,723 |
| CO2 | 0,001 | 0,001 | 0,001 | 0,703 | 0,001 |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 7,674 | 7,663 | 7,305 | 26,912 | 7,305 |
| MEACOO- | 771,908 | 769,705 | 770,643 | 767,28 | 770,645 |
| MEA+ | 819,129 | 816,818 | 816,226 | 798,564 | 816,226 |
| CO3-2 | 19,661 | 19,612 | 19,031 | 2,085 | 19,03 |
| H3O+ |  |  |  |  |  |
| OH- | 0,226 | 0,226 | 0,216 | 0,201 | 0,216 |
| Mole Frac |  |  |  |  |  |
| H2O | 0,887 | 0,887 | 0,882 | 0,882 | 0,882 |
| MEA | 0,071 | 0,071 | 0,074 | 0,075 | 0,074 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- |  |  |  | 0,001 |  |
| MEACOO- | 0,02 | 0,02 | 0,021 | 0,021 | 0,021 |
| MEA+ | 0,021 | 0,021 | 0,022 | 0,022 | 0,022 |
| CO3-2 | 0,001 | 0,001 | 0,001 |  | 0,001 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |

|  | LIQ | LIQ2 | MKUP-H2O | MKUP-MEA | RICH-ABS |
|---|---|---|---|---|---|
|  |  |  |  |  |  |
| Temperature C | 25 |  | 38 | 38 | 51,7 |
| Pressure   bar | 5,92 |  | 1,3 | 1,3 | 1,02 |
| Vapor Frac | 0 |  | 0 | 0 | 0 |
| Mole Flow   kmol/hr | 15,995 | 0 | 1654,09 | 0,002 | 36571,51 |
| Mass Flow   tonne/hr | 0,288 | 0 | 29,799 | 0 | 932,451 |
| Volume Flow cum/hr | 0,29 | 0 | 30,007 | 0 | 983,966 |
| Enthalpy   MMkcal/hr | -1,099 |  | -112,535 | 0 | -2701,488 |
| Mass Flow   tonne/hr |  |  |  |  |  |
| H2O | 0,288 | 0 | 29,799 | 0 | 578,937 |
| MEA | 0 | 0 |  | 0 | 24,074 |
| CO2 | 0 | 0 |  | 0 | 0,022 |
| N2 | 0 | 0 |  | 0 | 0,008 |
| O2 | 0 | 0 |  | 0 | 0,089 |
| HCO3- | 0 | 0 |  | 0 | 7,292 |
| MEACOO- | 0 | 0 |  | 0 | 195,369 |
| MEA+ | 0 | 0 |  | 0 | 125,762 |
| CO3-2 | 0 | 0 |  | 0 | 0,898 |
| H3O+ | 0 | 0 |  | 0 |  |
| OH- | 0 | 0 |  | 0 |  |
| Mass Frac |  |  |  |  |  |
| H2O | 1 |  | 1 | 0 | 0,621 |
| MEA | 0 |  |  | 1 | 0,026 |
| CO2 | 0 |  |  | 0 |  |
| N2 | 0 |  |  | 0 |  |
| O2 | 0 |  |  | 0 |  |
| HCO3- | 0 |  |  | 0 | 0,008 |
| MEACOO- | 0 |  |  | 0 | 0,21 |
| MEA+ | 0 |  |  | 0 | 0,135 |
| CO3-2 | 0 |  |  | 0 | 0,001 |
| H3O+ | 0 |  |  | 0 |  |
| OH- | 0 |  |  | 0 |  |
| Mole Flow   kmol/hr |  |  |  |  |  |
| H2O | 15,992 | 0 | 1654,09 | 0 | 32135,88 |
| MEA | 0 | 0 |  | 0,002 | 394,122 |
| CO2 | 0 | 0 |  | 0 | 0,494 |
| N2 | 0 | 0 |  | 0 | 0,277 |
| O2 | 0,002 | 0 |  | 0 | 2,786 |
| HCO3- | 0 | 0 |  | 0 | 119,501 |
| MEACOO- | 0 | 0 |  | 0 | 1877,013 |
| MEA+ | 0 | 0 |  | 0 | 2026,458 |
| CO3-2 | 0 | 0 |  | 0 | 14,966 |
| H3O+ | 0 | 0 |  | 0 |  |
| OH- | 0 | 0 |  | 0 | 0,013 |
| Mole Frac |  |  |  |  |  |
| H2O | 1 | 0 | 1 | 0 | 0,879 |
| MEA | 0 | 0 |  | 1 | 0,011 |
| CO2 | 0 | 0 |  | 0 |  |
| N2 | 0 | 0 |  | 0 |  |
| O2 | 0 | 0 |  | 0 |  |
| HCO3- | 0 | 0 |  | 0 | 0,003 |
| MEACOO- | 0 | 0 |  | 0 | 0,051 |
| MEA+ | 0 | 0 |  | 0 | 0,055 |
| CO3-2 | 0 | 0 |  | 0 |  |
| H3O+ | 0 | 0 |  | 0 |  |
| OH- | 0 | 0 |  | 0 |  |

|  | RICH-P1 | RICH-IN | WATER | WATER2 |
|---|---|---|---|---|
|  |  |  |  |  |
| Temperature C | 51,7 | 107 | 35 | 47 |
| Pressure    bar | 2 | 1,9 | 1,1 | 1,02 |
| Vapor Frac | 0 | 0,015 | 0 | 0 |
| Mole Flow   kmol/hr | 36571,511 | 36815,731 | 709 | 4087,069 |
| Mass Flow   tonne/hr | 932,451 | 932,451 | 12,773 | 73,637 |
| Volume Flow cum/hr | 983,95 | 10438,652 | 12,848 | 74,445 |
| Enthalpy   MMkcal/hr | -2701,458 | -2647,415 | -48,274 | -277,378 |
| Mass Flow   tonne/hr |  |  |  |  |
| H2O | 578,936 | 579,364 | 12,773 | 73,621 |
| MEA | 24,077 | 53,147 |  |  |
| CO2 | 0,022 | 10,77 |  | 0,004 |
| N2 | 0,008 | 0,008 |  | 0,001 |
| O2 | 0,089 | 0,089 |  | 0,011 |
| HCO3- | 7,294 | 6,531 |  |  |
| MEACOO- | 195,366 | 172,417 |  |  |
| MEA+ | 125,761 | 109,909 |  |  |
| CO3-2 | 0,897 | 0,224 |  |  |
| H3O+ |  | 0 |  |  |
| OH- |  | 0,001 |  |  |
| Mass Frac |  |  |  |  |
| H2O | 0,621 | 0,621 | 1 | 1 |
| MEA | 0,026 | 0,057 |  |  |
| CO2 |  | 0,012 |  |  |
| N2 |  | 0 |  |  |
| O2 |  | 0 |  |  |
| HCO3- | 0,008 | 0,007 |  |  |
| MEACOO- | 0,21 | 0,185 |  |  |
| MEA+ | 0,135 | 0,118 |  |  |
| CO3-2 | 0,001 | 0 |  |  |
| H3O+ |  | 0 |  |  |
| OH- |  | 0 |  |  |
| Mole Flow   kmol/hr |  |  |  |  |
| H2O | 32135,858 | 32159,577 | 709 | 4086,608 |
| MEA | 394,158 | 870,074 |  |  |
| CO2 | 0,495 | 244,715 |  | 0,087 |
| N2 | 0,277 | 0,277 |  | 0,032 |
| O2 | 2,786 | 2,786 |  | 0,338 |
| HCO3- | 119,534 | 107,027 |  | 0,002 |
| MEACOO- | 1876,989 | 1656,506 |  |  |
| MEA+ | 2026,447 | 1771,013 |  |  |
| CO3-2 | 14,955 | 3,725 |  |  |
| H3O+ |  | 0 |  | 0,002 |
| OH- | 0,013 | 0,031 |  |  |
| Mole Frac |  |  |  |  |
| H2O | 0,879 | 0,874 | 1 | 1 |
| MEA | 0,011 | 0,024 |  |  |
| CO2 |  | 0,007 |  |  |
| N2 |  | 0 |  |  |
| O2 |  | 0 |  |  |
| HCO3- | 0,003 | 0,003 |  |  |
| MEACOO- | 0,051 | 0,045 |  |  |
| MEA+ | 0,055 | 0,048 |  |  |
| CO3-2 |  | 0 |  |  |
| H3O+ |  | 0 |  |  |
| OH- |  | 0 |  |  |

# Appendix 6 – Flue gas condensation

Dew point: 61.6 °C

**Cooler duty dependence of flue gas temperature**

A line chart titled "Cooler duty dependence of flue gas temperature" with the y-axis labeled "Outlet flue gas temperature, °C" ranging from 0 to 140, and the x-axis labeled "Cooler duty, MW" ranging from 0 to 120. The curve, labeled "Flue gas cooling," starts at about 133 °C near 0 MW, drops steeply to about 62 °C at 20 MW, then declines gradually to 0 °C at about 108 MW.

Flue gas inlet temperature to the flue gas cooler is 144 °C. First point in diagram is when the flue gases are cooled to 133 °C.

# Appendix 7 – Flowsheet and stream data for case 3

Simulation convergence problems arise with the makeup flow of MEA. Even though it is << 0,001 ton/h in the stream table result it should be equal to the losses. The only losses of MEA are to the atmosphere i.e. in stream FG-ATMOS. Therefore, the makeup flow of MEA should be equal to the value in FG-ATMOS. This does not affect the simulation results, more than the result of makeup MEA flow, since the flowsheet is not connected in LEAN and LEANMIX. The simulation flowsheet and stream data are presented below.

|  | ABS-IN | CO2-H2O | CO2-PROD | COOLEDFG | FG-ABS |
|---|---|---|---|---|---|
| Temperature C | 56,4 | 23,6 | 25 | 47 | 58 |
| Pressure   bar | 1,11 | 1,7 | 75 | 1,02 | 1,02 |
| Vapor Frac | 1 | 1 | 0 | 0,882 | 1 |
| Mole Flow  kmol/hr | 25166,018 | 1239,803 | 1223,727 | 28543,999 | 25155,06 |
| Mass Flow  tonne/hr | 703,644 | 53,96 | 53,671 | 764,505 | 671,989 |
| Volume Flow cum/hr | 621280,876 | 17993,715 | 74,033 | 656804,001 | 678999,167 |
| Enthalpy   MMkcal/hr | -281,217 | -115,469 | -117,518 | -512,182 | -236,555 |
| Mass Flow  tonne/hr |  |  |  |  |  |
| H2O | 47,654 | 0,392 | 0,102 | 108,502 | 69,314 |
| MEA |  |  |  |  | 0,159 |
| CO2 | 62,808 | 53,472 | 53,47 | 62,811 | 9,429 |
| N2 | 574,924 | 0,008 | 0,008 | 574,924 | 574,916 |
| O2 | 18,259 | 0,087 | 0,087 | 18,267 | 18,171 |
| HCO3- |  |  | 0,002 |  |  |
| MEACOO- |  |  |  |  |  |
| MEA+ |  |  |  |  |  |
| CO3-2 |  |  |  |  |  |
| H3O+ |  |  | 0,001 |  |  |
| OH- |  |  |  |  |  |
| Mass Frac |  |  |  |  |  |
| H2O | 0,068 | 0,007 | 0,002 | 0,142 | 0,103 |
| MEA |  |  |  |  |  |
| CO2 | 0,089 | 0,991 | 0,996 | 0,082 | 0,014 |
| N2 | 0,817 |  |  | 0,752 | 0,856 |
| O2 | 0,026 | 0,002 | 0,002 | 0,024 | 0,027 |
| HCO3- |  |  |  |  |  |
| MEACOO- |  |  |  |  |  |
| MEA+ |  |  |  |  |  |
| CO3-2 |  |  |  |  |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow  kmol/hr |  |  |  |  |  |
| H2O | 2645,181 | 21,774 | 5,661 | 6022,781 | 3847,528 |
| MEA |  |  |  |  | 2,599 |
| CO2 | 1427,126 | 1215,007 | 1214,966 | 1427,198 | 214,244 |
| N2 | 20523,11 | 0,29 | 0,29 | 20523,136 | 20522,82 |
| O2 | 570,601 | 2,732 | 2,732 | 570,88 | 567,869 |
| HCO3- |  |  | 0,039 | 0,002 |  |
| MEACOO- |  |  |  |  |  |
| MEA+ |  |  |  |  |  |
| CO3-2 |  |  |  |  |  |
| H3O+ |  |  | 0,039 | 0,002 |  |
| OH- |  |  |  |  |  |
| Mole Frac |  |  |  |  |  |
| H2O | 0,105 | 0,018 | 0,005 | 0,211 | 0,153 |
| MEA |  |  |  |  |  |
| CO2 | 0,057 | 0,98 | 0,993 | 0,05 | 0,009 |
| N2 | 0,816 |  |  | 0,719 | 0,816 |
| O2 | 0,023 | 0,002 | 0,002 | 0,02 | 0,023 |
| HCO3- |  |  |  |  |  |
| MEACOO- |  |  |  |  |  |
| MEA+ |  |  |  |  |  |
| CO3-2 |  |  |  |  |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |

|  | FG-ATMOS | FG-SEP | FLUE-GAS | H2O-WASH | H2O-WSH2 |
|---|---|---|---|---|---|
| Temperature C | 53,8 | 47 | 144 | 35 | 53,8 |
| Pressure bar | 1,013 | 1,02 | 1,02 | 1,02 | 1,013 |
| Vapor Frac | 1 | 1 | 1 | 0 | 0 |
| Mole Flow kmol/hr | 24970,708 | 25166,018 | 28544 | 7806,818 | 7990,063 |
| Mass Flow tonne/hr | 668,528 | 703,644 | 764,505 | 143,953 | 147,413 |
| Volume Flow cum/hr | 670068,709 | 656742,472 | 970586,322 | 144,843 | 149,458 |
| Enthalpy MMkcal/hr | -226,687 | -282,928 | -457,055 | -534,063 | -543,931 |
| Mass Flow tonne/hr |  |  |  |  |  |
| H2O | 66,06 | 47,654 | 108,502 | 139,537 | 142,771 |
| MEA | 0,003 |  |  | 0,823 | 0,945 |
| CO2 | 9,379 | 62,808 | 62,811 | 0 | 0,001 |
| N2 | 574,916 | 574,924 | 574,924 | 0,002 | 0,002 |
| O2 | 18,171 | 18,259 | 18,267 | 0,018 | 0,018 |
| HCO3- |  |  |  | 0,382 | 0,484 |
| MEACOO- |  |  |  | 1,611 | 1,606 |
| MEA+ |  |  |  | 1,505 | 1,542 |
| CO3-2 |  |  |  | 0,075 | 0,045 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mass Frac |  |  |  |  |  |
| H2O | 0,099 | 0,068 | 0,142 | 0,969 | 0,969 |
| MEA |  |  |  | 0,006 | 0,006 |
| CO2 | 0,014 | 0,089 | 0,082 |  |  |
| N2 | 0,86 | 0,817 | 0,752 |  |  |
| O2 | 0,027 | 0,026 | 0,024 |  |  |
| HCO3- |  |  |  | 0,003 | 0,003 |
| MEACOO- |  |  |  | 0,011 | 0,011 |
| MEA+ |  |  |  | 0,01 | 0,01 |
| CO3-2 |  |  |  | 0,001 |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow kmol/hr |  |  |  |  |  |
| H2O | 3666,86 | 2645,181 | 6022,784 | 7745,47 | 7924,983 |
| MEA | 0,055 |  |  | 13,473 | 15,465 |
| CO2 | 213,122 | 1427,126 | 1427,2 | 0,007 | 0,022 |
| N2 | 20522,819 | 20523,11 | 20523,136 | 0,058 | 0,059 |
| O2 | 567,853 | 570,601 | 570,88 | 0,562 | 0,578 |
| HCO3- |  |  |  | 6,263 | 7,926 |
| MEACOO- |  |  |  | 15,477 | 15,43 |
| MEA+ |  |  |  | 24,252 | 24,85 |
| CO3-2 |  |  |  | 1,253 | 0,743 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  | 0,005 | 0,007 |
| Mole Frac |  |  |  |  |  |
| H2O | 0,147 | 0,105 | 0,211 | 0,992 | 0,992 |
| MEA |  |  |  | 0,002 | 0,002 |
| CO2 | 0,009 | 0,057 | 0,05 |  |  |
| N2 | 0,822 | 0,816 | 0,719 |  |  |
| O2 | 0,023 | 0,023 | 0,02 |  |  |
| HCO3- |  |  |  | 0,001 | 0,001 |
| MEACOO- |  |  |  | 0,002 | 0,002 |
| MEA+ |  |  |  | 0,003 | 0,003 |
| CO3-2 |  |  |  |  |  |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |

|  | LEAN | LEAN-MIX | LEAN-P2 | LEAN-REB | LEANCOOL |
|---|---|---|---|---|---|
|  |  |  |  |  |  |
| Temperature C | 38 | 38,1 | 38 | 122,3 | 38 |
| Pressure   bar | 1,3 | 1,3 | 1,9 | 1,9 | 1,8 |
| Vapor Frac | 0 | 0 | 0 | 0 | 0 |
| Mole Flow   kmol/hr | 41015,644 | 41027,04 | 39788,924 | 39789,936 | 39788,924 |
| Mass Flow   tonne/hr | 995,988 | 995,988 | 973,683 | 973,684 | 973,684 |
| Volume Flow cum/hr | 963,265 | 963,232 | 940,804 | 1001,166 | 940,804 |
| Enthalpy   MMkcal/hr | -2893,527 | -2894,112 | -2809,878 | -2737,526 | -2809,883 |
| Mass Flow   tonne/hr |  |  |  |  |  |
| H2O | 649,041 | 649,294 | 627,004 | 626,932 | 627,004 |
| MEA | 182,213 | 182,313 | 182,295 | 183,84 | 182,295 |
| CO2 |  |  |  | 0,045 |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 0,556 | 0,556 | 0,537 | 1,957 | 0,537 |
| MEACOO- | 99,925 | 99,709 | 99,793 | 99,27 | 99,793 |
| MEA+ | 62,916 | 62,782 | 62,75 | 61,492 | 62,75 |
| CO3-2 | 1,333 | 1,331 | 1,3 | 0,145 | 1,3 |
| H3O+ |  |  |  |  |  |
| OH- | 0,004 | 0,004 | 0,004 | 0,003 | 0,004 |
| Mass Frac |  |  |  |  |  |
| H2O | 0,652 | 0,652 | 0,644 | 0,644 | 0,644 |
| MEA | 0,183 | 0,183 | 0,187 | 0,189 | 0,187 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 0,001 | 0,001 | 0,001 | 0,002 | 0,001 |
| MEACOO- | 0,1 | 0,1 | 0,102 | 0,102 | 0,102 |
| MEA+ | 0,063 | 0,063 | 0,064 | 0,063 | 0,064 |
| CO3-2 | 0,001 | 0,001 | 0,001 |  | 0,001 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |
| Mole Flow   kmol/hr |  |  |  |  |  |
| H2O | 36027,269 | 36041,308 | 34804,01 | 34800,011 | 34804,011 |
| MEA | 2982,998 | 2984,638 | 2984,344 | 3009,648 | 2984,343 |
| CO2 | 0,001 | 0,001 | 0,001 | 1,013 | 0,001 |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- | 9,117 | 9,107 | 8,806 | 32,073 | 8,805 |
| MEACOO- | 960,035 | 957,956 | 958,765 | 953,74 | 958,766 |
| MEA+ | 1013,794 | 1011,635 | 1011,119 | 990,84 | 1011,119 |
| CO3-2 | 22,212 | 22,177 | 21,669 | 2,416 | 21,669 |
| H3O+ |  |  |  |  |  |
| OH- | 0,218 | 0,218 | 0,211 | 0,196 | 0,211 |
| Mole Frac |  |  |  |  |  |
| H2O | 0,878 | 0,878 | 0,875 | 0,875 | 0,875 |
| MEA | 0,073 | 0,073 | 0,075 | 0,076 | 0,075 |
| CO2 |  |  |  |  |  |
| N2 |  |  |  |  |  |
| O2 |  |  |  |  |  |
| HCO3- |  |  |  | 0,001 |  |
| MEACOO- | 0,023 | 0,023 | 0,024 | 0,024 | 0,024 |
| MEA+ | 0,025 | 0,025 | 0,025 | 0,025 | 0,025 |
| CO3-2 | 0,001 | 0,001 | 0,001 |  | 0,001 |
| H3O+ |  |  |  |  |  |
| OH- |  |  |  |  |  |

|  | LIQ | LIQ2 | MKUP-H2O | MKUP-MEA | RICH-ABS |
|---|---|---|---|---|---|
|  |  |  |  |  |  |
| Temperature C | 25 |  | 38 | 38 | 55,9 |
| Pressure   bar | 5,92 |  | 1,3 | 1,3 | 1,02 |
| Vapor Frac | 0 |  | 0 | 0 | 0 |
| Mole Flow  kmol/hr | 16,036 | 0 | 1238,114 | 0,002 | 39814,114 |
| Mass Flow   tonne/hr | 0,289 | 0 | 22,305 | 0 | 1027,604 |
| Volume Flow cum/hr | 0,291 | 0 | 22,461 | 0 | 1083,289 |
| Enthalpy   MMkcal/hr | -1,102 |  | -84,234 | 0 | -2938,183 |
| Mass Flow  tonne/hr |  |  |  |  |  |
| H2O | 0,289 | 0 | 22,305 | 0 | 625,974 |
| MEA |  | 0 |  | 0 | 39,187 |
| CO2 |  | 0 |  | 0 | 0,017 |
| N2 |  | 0 |  | 0 | 0,008 |
| O2 |  | 0 |  | 0 | 0,087 |
| HCO3- |  | 0 |  | 0 | 5,798 |
| MEACOO- |  | 0 |  | 0 | 217,976 |
| MEA+ |  | 0 |  | 0 | 137,679 |
| CO3-2 |  | 0 |  | 0 | 0,877 |
| H3O+ |  | 0 |  | 0 | 0 |
| OH- |  | 0 |  | 0 | 0 |
| Mass Frac |  |  |  |  |  |
| H2O | 1 | 0 | 1 | 0 | 0,609 |
| MEA |  | 0 | 0 | 1 | 0,038 |
| CO2 |  | 0 | 0 | 0 | 0 |
| N2 |  | 0 | 0 | 0 | 0 |
| O2 |  | 0 | 0 | 0 | 0 |
| HCO3- |  | 0 | 0 | 0 | 0,006 |
| MEACOO- |  | 0 | 0 | 0 | 0,212 |
| MEA+ |  | 0 | 0 | 0 | 0,134 |
| CO3-2 |  | 0 | 0 | 0 | 0,001 |
| H3O+ |  | 0 | 0 | 0 | 0 |
| OH- |  | 0 | 0 | 0 | 0 |
| Mole Flow  kmol/hr |  |  |  |  |  |
| H2O | 16,034 | 0 | 1238,114 | 0 | 34746,81 |
| MEA |  | 0 |  | 0,002 | 641,527 |
| CO2 |  | 0 |  | 0 | 0,395 |
| N2 |  | 0 |  | 0 | 0,29 |
| O2 | 0,002 | 0 |  | 0 | 2,732 |
| HCO3- |  | 0 |  | 0 | 95,025 |
| MEACOO- |  | 0 |  | 0 | 2094,214 |
| MEA+ |  | 0 |  | 0 | 2218,487 |
| CO3-2 |  | 0 |  | 0 | 14,614 |
| H3O+ |  | 0 |  | 0 | 0 |
| OH- |  | 0 |  | 0 | 0,021 |
| Mole Frac |  |  |  |  |  |
| H2O | 1 | 0 | 1 | 0 | 0,873 |
| MEA |  | 0 |  | 1 | 0,016 |
| CO2 |  | 0 |  | 0 | 0 |
| N2 |  | 0 |  | 0 | 0 |
| O2 |  | 0 |  | 0 |  |
| HCO3- |  | 0 |  | 0 | 0,002 |
| MEACOO- |  | 0 |  | 0 | 0,053 |
| MEA+ |  | 0 |  | 0 | 0,056 |
| CO3-2 |  | 0 |  | 0 |  |
| H3O+ |  | 0 |  | 0 |  |
| OH- |  | 0 |  | 0 |  |

|  | RICH-IN | RICH-P1 | WATER |
|---|---|---|---|
|  |  |  |  |
| **Temperature C** | 107 | 55,9 | 47 |
| **Pressure   bar** | 1,9 | 2 | 1,02 |
| **Vapor Frac** | 0,01 | 0 | 0 |
| **Mole Flow   kmol/hr** | 39987,383 | 39814,115 | 3377,98 |
| **Mass Flow   tonne/hr** | 1027,604 | 1027,604 | 60,861 |
| **Volume Flow cum/hr** | 7489,718 | 1083,273 | 61,529 |
| **Enthalpy   MMkcal/hr** | -2885,972 | -2938,15 | -229,254 |
| **Mass Flow   tonne/hr** |  |  |  |
| **H2O** | 625,81 | 625,973 | 60,848 |
| **MEA** | 61,554 | 39,189 | 0 |
| **CO2** | 7,643 | 0,017 | 0,003 |
| **N2** | 0,008 | 0,008 | 0,001 |
| **O2** | 0,087 | 0,087 | 0,009 |
| **HCO3-** | 6,997 | 5,8 | |
| **MEACOO-** | 198,998 | 217,974 | |
| **MEA+** | 126,27 | 137,679 | |
| **CO3-2** | 0,241 | 0,876 | |
| **H3O+** | 0 | 0 | |
| **OH-** | 0,001 | 0 | |
| **Mass Frac** |  |  |  |
| **H2O** | 0,609 | 0,609 | 1 |
| **MEA** | 0,06 | 0,038 | |
| **CO2** | 0,007 | 0 | |
| **N2** | 0 | 0 | |
| **O2** | 0 | 0 | |
| **HCO3-** | 0,007 | 0,006 | |
| **MEACOO-** | 0,194 | 0,212 | |
| **MEA+** | 0,123 | 0,134 | |
| **CO3-2** | 0 | 0,001 | |
| **H3O+** | 0 | 0 | |
| **OH-** | 0 | 0 | |
| **Mole Flow   kmol/hr** |  |  |  |
| **H2O** | 34737,736 | 34746,791 | 3377,6 |
| **MEA** | 1007,704 | 641,559 | |
| **CO2** | 173,663 | 0,396 | 0,072 |
| **N2** | 0,29 | 0,29 | 0,026 |
| **O2** | 2,732 | 2,732 | 0,279 |
| **HCO3-** | 114,677 | 95,055 | 0,002 |
| **MEACOO-** | 1911,883 | 2094,194 | |
| **MEA+** | 2034,641 | 2218,476 | |
| **CO3-2** | 4,024 | 14,603 | |
| **H3O+** | 0 | 0 | 0,002 |
| **OH-** | 0,033 | 0,021 | |
| **Mole Frac** |  |  |  |
| **H2O** | 0,869 | 0,873 | 1 |
| **MEA** | 0,025 | 0,016 | |
| **CO2** | 0,004 | 0 | |
| **N2** | 0 | 0 | |
| **O2** | | | |
| **HCO3-** | 0,003 | 0,002 | |
| **MEACOO-** | 0,048 | 0,053 | |
| **MEA+** | 0,051 | 0,056 | |
| **CO3-2** | | | |
| **H3O+** | | | |
| **OH-** | | | |

Values for Figure 20, section 6.4

| No of sections | Alfa | Mole fraction MEA | Mole fraction CO2 | Mole fraction H2O | LEAN flow, ton/h |
|---|---|---|---|---|---|
| | | | | | |
| 2 | 0,01 | 0,118 | 0,00118 | 0,88082 | 771 |
| | 0,05 | 0,118 | 0,0059 | 0,8761 | 857 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | 997 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | 1189 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | 1456 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | 1859 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | 2515 |
| 4 | 0,01 | 0,118 | 0,00118 | 0,88082 | 544 |
| | 0,05 | 0,118 | 0,0059 | 0,8761 | 602 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | 694 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | 818 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | 996 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | 1265 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | 1700 |
| | 0,35 | 0,118 | 0,0413 | 0,8407 | 2397 |
| | 0,4 | 0,118 | 0,0472 | 0,8348 | 3945 |
| 6 | 0,01 | 0,118 | 0,00118 | 0,88082 | 526 |
| | 0,05 | 0,118 | 0,0059 | 0,8761 | 580 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | 664 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | 779 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | 939 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | 1184 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | 1595 |
| 8 | 0,01 | 0,118 | 0,00118 | 0,88082 | 522 |
| | 0,05 | 0,118 | 0,0059 | 0,8761 | 575 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | 658 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | 769 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | 922 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | 1153 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | 1551 |



Absorber optimization
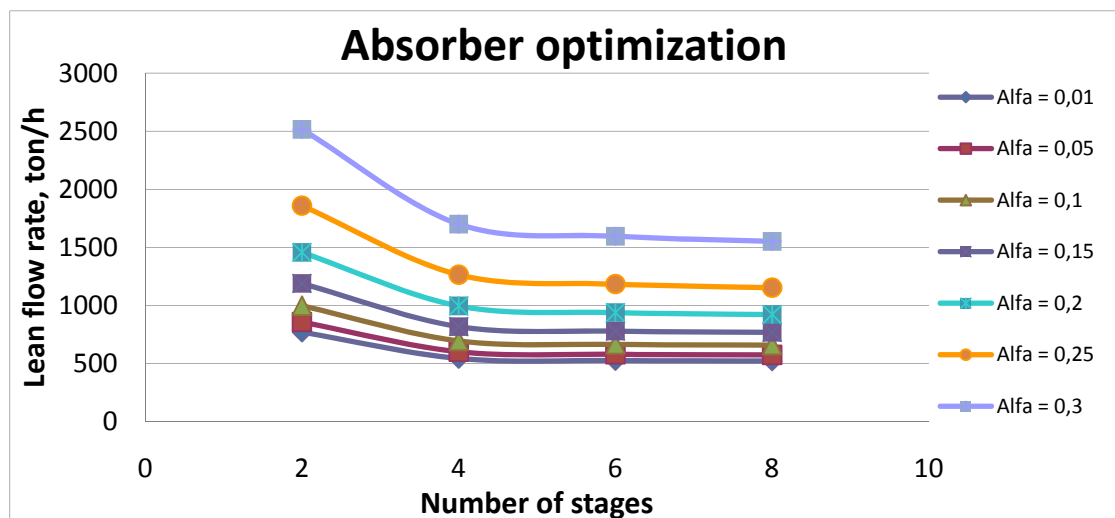
<u>Physical explanation to above figure</u>

The driving force between According to Nusselt's film theory
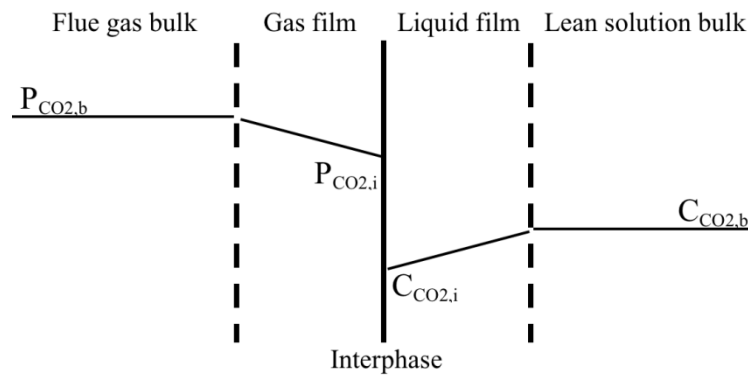


$P_{CO2,b}$ = CO$_2$ partial pressure in the gas bulk

$P_{CO2,i}$ = CO$_2$ partial pressure at the gas-liquid interface

$C_{CO2,b}$ = CO$_2$ concentration in the liquid bulk

$C_{CO2,i}$ = CO$_2$ concentration at the gas-liquid interface

*Figure 22. The theory of how to enhance CO$_2$ absorption. The figure shows pressure and concentrations gradients close to and at the gas-liquid interphase.*

A high CO$_2$ partial pressure in the gas bulk phase results in a high CO$_2$ partial pressure at the gas-liquid interface. The same follows for the concentrations in the liquid phase. Also, the partial pressure and concentration at the interface is lower than in the bulk due to mass transfer resistance in the films. Assuming the flue gases as an ideal gas the partial pressure of CO$_2$ in the bulk can be written as.

$$P_{CO2,b} = \dot{n}_{CO2,b} \cdot R \cdot T_{gas}/\dot{V}_{gas}$$

Where $\dot{n}_{CO2,b}$ is the molar flow rate of CO$_2$ in the gas phase, $R$ is the ideal gas constant, $T_{gas}$ is the gas temperature and $\dot{V}_{gas}$ is the gas volumetric flow rate. Inserting $C_{CO2,b} = \dot{n}_{CO2,b}/\dot{V}_{gas}$ and rewriting above expression the CO$_2$ concentration in the gas bulk is obtained.

$$C_{CO2,b} = P_{CO2,b}/R \cdot T_{gas}$$

The driving force is $C_{CO2,b}$– $C_{CO2,i}$. The expression shows that if the CO$_2$ partial pressure is increased, the CO$_2$ concentration increases and so does the driving force, but the blower work consumption increases as well. Also, decreasing the CO$_2$ concentration in the lean solution i.e. $C_{CO2,b}$ (decreases $C_{CO2,i}$) results in an increased driving force. The expression also shows that a lower temperature increases the driving force i.e. enhances the absorption, as discussed in section 6.3.1.

Values for Figure 21, section 6.4

| No of sections | Alfa | Mole frac MEA in LEAN | Mole frac CO2 in LEAN | Mole frac H2O in LEAN | Condenser duty [MW] | Reflux Ratio (mass) | Reboiler duty [MW] |
|---|---|---|---|---|---|---|---|
| 19 | 0,05 | 0,118 | 0,0059 | 0,8761 | -272,4 | 6,93 | 307,1 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | -121,0 | 3,06 | 156,6 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | -57,7 | 1,45 | 94,6 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | -26,5 | 0,66 | 64,9 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | -20,6 | 0,50 | 60,6 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | -20,6 | 0,50 | 60,8 |
| | 0,35 | 0,118 | 0,0413 | 0,8407 | -20,6 | 0,50 | 53,8 |
| 14 | 0,05 | 0,118 | 0,0059 | 0,8761 | -294,0 | 7,48 | 328,8 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | -129,6 | 3,28 | 165,3 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | -62,4 | 1,57 | 99,3 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | -29,5 | 0,73 | 67,8 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | -20,8 | 0,51 | 60,8 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | -20,6 | 0,50 | 60,8 |
| | 0,35 | 0,118 | 0,0413 | 0,8407 | -20,6 | 0,50 | 53,8 |
| 9 | 0,05 | 0,118 | 0,0059 | 0,8761 | -348,3 | 8,86 | 383,1 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | -169,7 | 4,30 | 205,4 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | -83,5 | 2,11 | 120,4 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | -41,5 | 1,04 | 79,9 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | -24,2 | 0,60 | 64,1 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | -20,9 | 0,51 | 60,9 |
| | 0,35 | 0,118 | 0,0413 | 0,8407 | -20,6 | 0,50 | 53,8 |
| 4 | 0,05 | 0,118 | 0,0059 | 0,8761 | -601,0 | 15,31 | 635,7 |
| | 0,1 | 0,118 | 0,0118 | 0,8702 | -239,3 | 6,08 | 275,0 |
| | 0,15 | 0,118 | 0,0177 | 0,8643 | -118,2 | 2,99 | 155,1 |
| | 0,2 | 0,118 | 0,0236 | 0,8584 | -60,2 | 1,51 | 98,5 |
| | 0,25 | 0,118 | 0,0295 | 0,8525 | -32,7 | 0,81 | 72,6 |
| | 0,3 | 0,118 | 0,0354 | 0,8466 | -22,9 | 0,56 | 63,1 |
| | 0,35 | 0,118 | 0,0413 | 0,8407 | -20,8 | 0,51 | 54,0 |
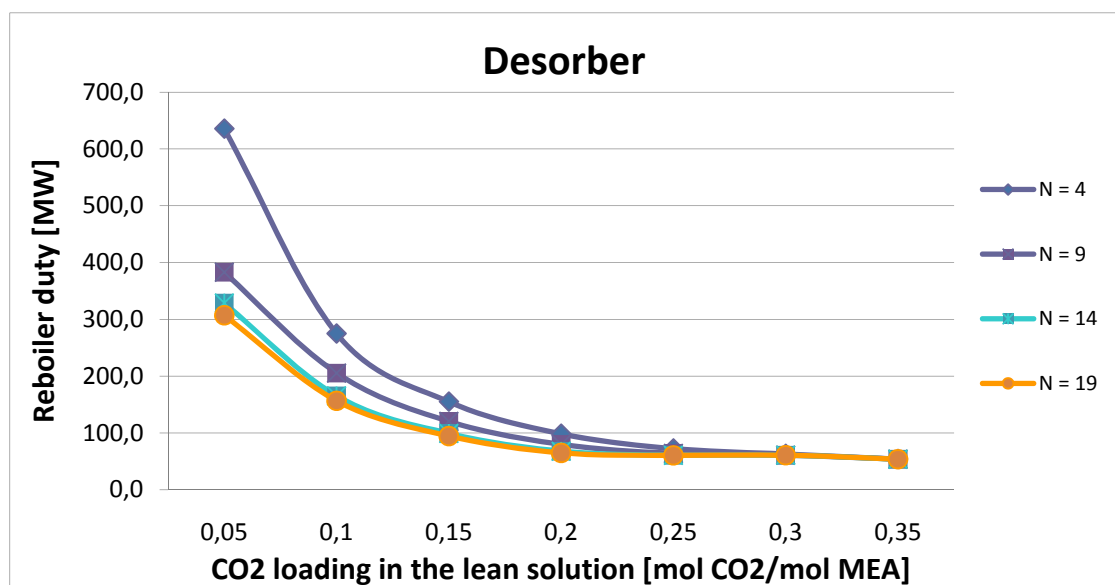
# Appendix 8 – Economic tables for this thesis

Costs for the carbon capture plant are calculated with the six – tenth rule in section 5.4. Index k represents the Tel-Tek study. Index x represents this thesis. Other equipment costs are taken from the Tel-Tek study (Appendix 2, page 60). The currency exchanges used are:

| Year 2010 | | |
|---|---:|---|
| 1 EUR = | 8,1493 | NOK |
| 1 NOK = | 0,1227 | EUR |
| Year 2006 | | |
| 1 EUR = | 8,109 | NOK |
| 1 NOK = | 0,1233 | EUR |
| Year 2006 | | |
| 1 EUR = | 9,1907 | SEK |
| 1 SEK = | 0,1088 | EUR |
| Year 2010 | | |
| 1 EUR = | 9,7135 | SEK |
| 1 SEK = | 0,1029 | EUR |

Chemical engineering plant cost index (CEPCI)

The CEPCI is used to convert the total capital investment cost of the carbon capture plant from 2006 to 2010.

CEPCI, 2006 = 499.6 (mean value of all montly values in 2006)

CEPCI, 2010 = 542.3 (mean value of montly values from January – Mars 2010)

Equipment investment costs:

| Equipment | Unit | Ek (Tel-Tek) | Equipment cost [NOK] | Equipemt cost [NOK] 2010 |
|---|---|---:|---:|---:|
| Flue gas cooler | [m$^2$] | 1048 | 9665000 | 10491052 |
| Blower | [kW] | 3027 | 7284000 | 7906552 |
| Absorber | [m$^3$] | 2847 | 26773000 | 29061245 |
| Rich-Lean heat exchanger | [m$^2$] | 12385 | 7205000 | 7820800 |
| Desorber | [m$^3$] | 736 | 11957000 | 12978945 |
| Reboiler | [m$^2$] | 3530 | 5445000 | 5910375 |
| Condenser | [m$^2$] | 1048 | 7602000 | 8251731 |
| Multistage compressor | [kW] | 7772 | 38828000 | 42146566 |
| | | | | |
| Equipment | Unit | Ex (This study) | Equipment cost [NOK] | Equipment cost [EUR] 2010 |
| Flue gas cooler | [m$^2$] | 1569 | 13915608 | 1707594 |
| Blower | [kW] | 2100 | 6121135 | 751129 |
| Absorber | [m$^3$] | 1071 | 14658705 | 1798780 |
| Rich-Lean heat exchanger | [m$^2$] | 8600 | 6058608 | 743456 |
| Desorber | [m$^3$] | 590 | 11117864 | 1364281 |
| Reboiler | [m$^2$] | 2500 | 4642274 | 569657 |
| Condenser | [m$^2$] | 700 | 6221008 | 763384 |
| Multistage compressor | [kW] | 4700 | 29638604 | 3636973 |
| | | | | |
| **TOTAL EQUIPMENT INVESTMENT COST [EUR]** | | | | **15623236** |

The cost capacity exponent ($n_c$) is 0.7 for the equipments in the table. All equipment costs except the equipment costs shown in the table above, are the same as in the Tel-Tek study (see Appendix 2, page 60). Total equipment investment cost including the costs from the Tel-Tek study is 15.6 M€.

Total capital investment (CAPEX):

| Investment analysis | Absorber cost [EUR] | Desorber cost [EUR] | Mcompr cost [EUR] | Piping cost [EUR] |
|---|---|---|---|---|
| Equipment cost | 4339819 | 5403971 | 5135990 | 3873681 |
| Equipment erection cost | 325424 | 751363 | 379349 | 265998 |
| Piping cost | 1620984 | 3743033 | 1889730 | |
| Electrical cost | 885367 | 2044327 | 1032134 | |
| Instrumentation cost | 1074088 | 2480180 | 1252126 | |
| Ground work cost | 298031 | 688077 | 347389 | 230966 |
| Steel and concrete cost | 919586 | 2123434 | 1071995 | 110954 |
| Isolation cost | 168761 | 389818 | 196821 | |
| Total cost | 9632059 | 17624202 | 11305534 | 4481599 |
| | | | | |
| Planning process cost | 258714 | 597436 | 301668 | 230966 |
| Planning mechanical cost | 124345 | 287030 | 144886 | 110954 |
| Planning piping cost | 483383 | 1116071 | 563475 | |
| Planning electrical cost | 231754 | 535171 | 270152 | |
| Planning instrumentation cost | 332682 | 768103 | 387783 | |
| Planning ground work cost | 48822 | 112791 | 56996 | 43556 |
| Planning steel and concrete cost | 142146 | 328166 | 165660 | |
| Planning isolation cost | 27392 | 63286 | 31960 | |
| Total cost | 1649240 | 3808053 | 1922578 | 385477 |
| | | | | |
| Project purchase | 185784 | 428913 | 216530 | 41425 |
| Project steering | 82436 | 190367 | 96147 | 35964 |
| Construction management | 418575 | 966533 | 487924 | 179951 |
| Project management | 406909 | 939484 | 474341 | 89909 |
| Total cost | 1093704 | 2525297 | 1274942 | 347249 |
| | | | | |
| Plant test run cost | 142802 | 271337 | 167460 | 0 |
| | | | | |
| Total cost except contign. [EUR] | 12517805 | 24228890 | 14670515 | 5214324 |
| | | | | |
| Contignency cost | 2299587 | 4286451 | 2635395 | 1048112 |
| | | | | |
| | | | | |
| **CAPITAL INVESTMENT [EUR]** | **14817392** | **28515341** | **17305910** | **6262437** |
| | | | | |
| **TOTAL CAPITAL INVESTMENT [EUR]** | **66901079** | | | |

The annualized total capital investment cost becomes 5 393 899 €. The equation for this calculation is shown in section 7.1.1.

Utility and chemicals consumption:

Fresh water and cooling water consumptions are actual consumptions from the process. Other consumptions are taken from the Tel-Tek study as these were outside the scope for this thesis (Appendix 2, page 60). The consumptions for fuel and electricity are calculated in this thesis.

| Utility and chemicals | Unit | Total consumption |
|---|---|---|
| Fresh water consumption | [m$^3$/h] | 25 |
| Cooling water consumption | [m$^3$/h] | 4412 |
| MEA consumption | [kg/h] | 30 |
| NA2CO3 | [kg/h] | 92,3 |
| Active coal | [kg/h] | 3,6 |
| Corrosion inhibitor | [kg/h] | 0,25 |
| Destruction | [kg/h] | 83,5 |

The heating efficiency (heat supplied to process/fuel supplied to boiler) is 0.8 and used to calculate the fuel needed to deliver 24.4 MW of heat to the carbon capture process. The power to heat ratio is 0.07. The boiler efficiency is 0.87. (Spetz, 2010)

| | |
|---|---|
| Heat supplied to the process [MW] | 24,4 |
| Supplied fuel energy to steam boilers [MW] | 30,5 |
| Electricity generated in turbo alternator [MW] | 2,1 |
| Electricity supplied from power plant [MW] | 8,9 |

Total operating costs (OPEX):

Operating hours per year for the carbon capture process is 8000 h/year.

| Costs | Price | Total cost [kEUR/yr] |
|---|---|---|
| Electricity from power plant [EUR/MWh] | 56,6 | 4035 |
| Fuel supplied to steam boilers  [EUR/MWh] | 30,9 | 7537 |
| Fresh water [EUR/m3] | 0,01 | 2,5 |
| Cooling water [EUR/m3] | 0,001 | 44 |
| MEA [EUR/kg] | 1,78 | 426 |
| NA2CO3 [EUR/kg] | 0,57 | 423 |
| Active coal [EUR/kg] | 5,43 | 156 |
| Corrosion inhibitor [EUR/kg] | 1,85 | 3,7 |
| Destruction [EUR/kg] | 0,25 | 165 |
| Total variable costs | | 12792 |
| | | |
| Operators [EUR/yr] | 57652 | 346 |
| Officer | - | 84 |
| Maintenance [% of total capital investment] | 4 | 2676 |
| Total fixed costs | | 3106 |
| | | |
| **TOTAL OPERATING COSTS [kEUR/yr]** | | **15898** |

The prices for electricity and fuel are up to date prices from Reine Spetz. Other prices are taken from the Tel-Tek study and converted from NOK (2006) to € (2006).

# Appendix 9 – Carbon capture process stream data used in pinch analysis (case 3)

| Stream | Start temperature [°C] | Target temperature [°C] | Duty [kW] | ΔTmin | F*Cp [kJ/K,s] |
|---|---|---|---|---|---|
| Flue gas cooling/condensation | 144 | 61,9 | 20140 | 10 | 245 |
| Flue gas cooling/condensation | 61,9 | 47 | 43970 | 5 | 2951 |
| Preheating rich solution | 56,2 | 107 | 60680 | 5 | 1194 |
| Cooling lean solution | 122,3 | 38 | 84150 | 5 | 998 |
| CO2-H2O condensation (condenser) | 104,7 | 70 | 21400 | 5 | 617 |
| CO2-H2O condensation (condenser) | 70 | 23,6 | 5140 | 10 | 110 |
| Reboiler | 122,3 | 122,3 | 64910 | 5 | - |

All streams in the carbon capture process that needs to be heated or cooled are presented in the table. Individual temperature differences are used and set to 10 K for gases, 5 K for liquids and condensation/evaporation.

The streams in the cracker plant are shown in the table below. The chosen streams that fulfilled the criteria's presented in section 6.4.2 (page 39) are marked in the table. The hot utility temperatures (steam) are not up to date in the table. The temperatures of the hot utility are those shown in Table 3 in section 3.3.1 (page 8).

| HXTR-No. | Description | T-start [°C] | T-target [°C] | Qref [kW] | Utility |
|---|---|---|---|---|---|
| E-601 B | Raw SCN | 115 | 175 | 639 | Hot oil |
| E-603 A/B | Reboiler to T-601 | 210 | 255,4 | 1333 | Hot oil |
| E-604, E-613 | Overhead stream from T-601 | 117,6 | 21 | 635 | Air |
| E-1 | Stream between R-2201 and R-1 | 75 | 50 | 4582 | CW |
| E-2 | Bottom stream from T-2201 | 90 | 40 | 838 | CW |
| E-2203 | Heating of recirculation from reactors | 65 | 50 | 2697 | CW |
| E-2205 | Overhead stream from T-2201 | 81 | 18 | 944 | CW |
| E-2207 | Reboiler to T-2201 | 94 | 97 | 861 | 1.8bar(g)/180 |
| E-2215 | Overhead stream from T-2202 A | 52 | 45 | 3150 | CW |
| E-2226 | Overhead stream from T-2202 B | 52 | 41 | 3255 | CW |
| E-2214 | Overhead stream from T-2203 and T-2205 | 57 | 22 | 2306 | CW |
| E-2229 | C₄-raff to storage | 63 | 23 | 2468 | CW |
| E-2218 N | Reboiler to T-2202 A | 137 | 138 | 2336 | 8.8bar(g)/250 |
| E-2225 N | Reboiler to T-2202 B | 129 | 131 | 3357 | 8.8bar(g)/250 |
| E-2218 X | Reboiler to T-2204 | 117 | 118 | 329 | 8.8bar(g)/250 |
| E-2222 | Reboiler to T-2205 | 63 | 64 | 899 | 1.8bar(g)/180 |
|  | Flue gas from cracking furnaces | 300 | 150 | 22410 | Air |
| E-735 | LPG heater | -10 | 4 | 688 | 1.8bar(g)/180 |
| E-736 | Propane heater | -30 | 9 | 1031 | 1.8bar(g)/180 |
| E-973 | Propylene heater | -40 | 3 | 667 | 1.8bar(g)/180 |
| E-961/E-967 | Ethane vaporizer (to furnaces) | -85 | 20 | 1046 | 1.8bar(g)/180 |
| E-1609X/ E-1606Y | LPG feed heater/vaporizer | 4 | 73 | 2377 | 8.8bar(g)/250 |
| E-1606Y | LPG feed heater/ vaporizer | 73 | 83 | 3875 | 8.8bar(g)/250 |
| E-1955 | Propane feed heater | 9 | 25 | 1490 | C3/9h |
| E-1608 | Propane feed heater/vaporizer/superheater | 26 | 40 | 299 | 1.8bar(g)/180 |
| E-1608 | Propane feed heater/vaporizer/superheater | 40 | 41 | 2190 | 1.8bar(g)/180 |
| E-1652A-E. E. 1655 / | Quench oil coolers | 187 | 147 | 13000 | BFW2.7bar(g) |
| E-1651 A-H part 1 | Primary fractionator, primary condensers (Top T-1651) | 113 | 105 | 1701 | CW |
| E-1651 A-H part 2 | Primary fractionator, primary condensers (Top T-1651) | 105 | 100 | 1866 | CW |
| E-1651 A-H part 3 | Primary fractionator, primary condensers (Top T-1651) | 100 | 90 | 3876 | CW |
| E-1651 A-H part 4 | Primary fractionator, primary condensers (Top T-1651) | 90 | 80 | 29624 | CW |
| E-1651 A-H part 5 | Primary fractionator, primary condensers (Top T-1651) | 80 | 70 | 18785 | CW |
| E-1651 A-H part 6 | Primary fractionator, primary condensers (Top T-1651) | 70 | 59 | 13353 | CW |
| E-1656 A-H | Primary fractionator, secondary condensers (Top T-1651) | 59 | 35 | 16670 | CW |
| E-1701 AX-DX | First stage after cooler (After C-1701) | 89 | 22 | 12610 | CW |
| E-1702 A-D | Second stage after cooler (After C-1702:1) | 64 | 25 | 8550 | CW |
| E-1703 A/B | Third stage after cooler (After C-1702:2) | 82 | 32 | 8280 | CW |
| E-1707 | Distillate stripper reboiler (T-1701) | 112 | 113 | 570 | 2.7bar(g) |
| E-1731 | Weak caustic heater (reflux T-1702) | 32 | 41 | 766 | 1.8bar(g)/180 |
| E-1704 X/ E-1705 X | Caustic scrubber overhead cooler/ Depropanizer feed chiller (Top T-1702) | 36 | 11 | 4539 | CW |
| E-1733 | Depropanizer feed chiller (After driers) | 10 | -16 | 4280 | C3/-21c |
| E-1709 A/B | Depropanizer reboiler (T-1703A) | 81 | 82 | 4590 | 1.8bar(g)/180 |
| E-1708 A/B | Depropanizer condenser (T-1703A) | -26 | -36 | 4480 | C3/-41c |
| E-1734 | Depropanizer booster compressor Aftercooler (After C-1703) | 56 | 27 | 1210 | CW |
| E-1802 | Acetylene converter feed heater (After C-1801 1st) | 43 | 84 | 4055 | 1.8bar(g)/180 |
| E-1803 | Acetylene converter interstage cooler no 1 (R-1801A) | 95 | 81 | 872 | CW |
| E-1804 | Acetylene converter interstage cooler no 2 (R-1801A) | 87 | 79 | 473 | CW |
| E-1805 | Acetylene converter interstage cooler no 3 (R-1801A) | 84 | 79 | 279 | CW |
| E-1806 A/B | Acetylene converter effluent cooler (R-1801A) | 81 | 27 | 2964 | CW |
| E-1803 B | Acetylene converter interstage cooler no 1 (R-1801B) | 95 | 81 | 796 | CW |
| E-1804 B | Acetylene converter interstage cooler no 2 (R-1801B) | 87 | 79 | 449 | CW |
| E-1805 B | Acetylene converter interstage cooler no 3 (R-1801B) | 84 | 79 | 265 | CW |
| E-1806 C/D | Acetylene converter effluent cooler (R-1801B) | 81 | 27 | 2814 | CW |
| E-1807 AX/BX/ E-18 | Feed booster second stage discharge cooler/ Demethanizer feed/HL propylen | 91 | 16 | 8320 | CW |
| E-1710 | Debutanizer condenser (T-1704) | 50 | 46 | 2930 | CW |
| E-1711 AX/BX | Debutanizer reboiler (T-1704) | 123 | 124 | 2840 | 2.7bar(g) |
| E-1712 A/B | Raw SCN cooler (Bottom T-1704) | 123 | 28 | 1010 | CW |
| E-1848 | Ethane feed heater (To E-1849 C°2) | -81 | -25 | 1235 | C3/-21h |
| E-1851 (4031) | Demethanizer feed chiller no. 4 | -24 | -35 | 2080 | C3/-41c |
| E-1811 | Demethanizer feed chiller no. 1 | 15 | -16 | 1490 | C3/-21c |
| E-1812 | Demethanizer feed chiller no. 2 | -19 | -35 | 2340 | C3/-41c |
| E-1820 AX/BX | Prefractionator reboiler (T-1801X) | 4 | 7 | 5936 | C3/9h |
| E-1820 AX/BX (4221 | Prefractionator reboiler (T-1801X) | 4 | 8 | 464 | C3/9h |
| E-1819 X | Prefractionator condenser (T-1801X) | -34 | -39 | 610 | C2/-62c |
| Z-1800 A/B (4111) | Dephlegmator no. 1 | -43 | -76 | 6070 | C2/-84c |
| E-1882 | Demethanizer reboiler (T-1807) | -8 | -7 | 2160 | C3/9h |
| E-1814 X | Demethanizer feed chiller (Top T-1807) | -53 | -81 | 1250 | C2/-84c |
| E-1881 (4311) | Demethanizer condenser (T-1807) | -90 | -97 | 645 | C2/-100c |
| E-1822 | Deethanizer reboiler (T-1802) | 66 | 67 | 11160 | 1.8bar(g)/180 |
| E-1821 A/B | Deethanizer condenser (T-1802) | -8 | -10 | 10820 | C3/-21c |
| E-1824 A-D | Ethylene/ethane splitter reboiler (T-1803) | -4 | -3 | 16600 | C3/9h |
| E-1823 A-D | Ethylene/ethane splitter condenser (T-1803) | -27 | -28 | 19590 | C3/-41c |
| E-1836 | Ethylene product superheater no. 1 (C″₂ to customers) | -27 | 2 | 710 | C3/9h |
| E-1837 X | Ethylene product superheater no. 2 (C″₂ to customers) | 2 | 20 | 410 | C3/9h |
| E-1834/ E-1835 | Ethylene product chiller no. 1 and 2 (C″₂ to storage) | -27 | -96 | 320 | C2/-62c |
| E-1859 | Hydrofiner feed vaporizer (Before R-1802) | 58 | 59 | 1338 | 2.7bar(g) |
| E-1827/ E-1828 | Hydrofiner effluent cooler (Before T-1804) | 68 | 60 | 330 | CW |
| E-1830 | Propylene rerun tower reboiler(T-1804) | 67 | 68 | 850 | 1.8bar(g)/180 |
| E-1829 A/B | Propylene rerun tower condenser (T-1804) | 43 | 34 | 3510 | CW |
| E-1890 | Secondary deethanizer reboiler (T-1809) | 52 | 53 | 3090 | 1.8bar(g)/180 |
| E-1889 | Secondary deethanizer condenser (T-1809) | 40 | 32 | 2910 | Air |
| E-1845 A/B | Propylene/propane splitter reboiler  (T-1805) | 55 | 56 | 21230 | 1.8bar(g)/180 |
| E-1846 A/B | Propylene/propane splitter overheads condenser (T-1805) | 44 | 32 | 21700 | Air |
| E-1832 X | Propylene product chiller no. 2 (C″₃ product) | 32 | -9 | 790 | C3/-21c |
| E-1833 | Propylene product chiller no. 3 (C″₃ product) | 10 | -34 | 850 | C3/-41c |
|  | Preheating demineralized water | 10 | 44 | 7140 | CircWh |
|  | Preheating demineralized water | 44 | 128 | 17640 | 1.8bar(g)/180 |
| E-1954/51 | Propylene cooling | 90 | 39 | 9000 | CircWc |
| E-1953 | Propylene condensation | 39 | 38 | 30000 | CW |
| E-1953 | Propylene cooling | 39 | 27 | 4000 | CW |
|  | Ethylene cooling | 53 | 24 | 1300 | CW |
|  | Ethylene cooling | 24 | 14 | 500 | C3/9c |
|  | Ethylene cooling | 14 | -15 | 1600 | C3/-21c |
|  | Ethylene cooling | -15 | -24 | 9000 | C3/-41c |
|  | Fuel gas heating for bed regeneration | 5 | 220 | 1000 | 85bar(g)sh |
|  | Fuel gas cooling from bed regeneration | 180 | 30 | 800 | CW |
|  | Product from Cracking furnaces | 850 | 350 | 50800 | BFW85bar(g) |
|  | Product from Cracking furnaces | 850 | 350 | 27300 | BFW8.8bar(g) |
|  | Product from Cracking furnaces | 850 | 350 | 50000 | BFW1.8bar(g) |
|  | Process steam to Cracking furnaces | 169 | 170 | 64600 | 8.8bar(g)/250 |
|  |  |  |  |  | Sum |

# Appendix 10 - Carbon capture process stream data used in pinch analysis (case 4)

| Stream | Start temperature [°C] | Target temperature [°C] | Duty [kW] | ΔTmin | F*Cp [kJ/K,s] |
|---|---|---|---|---|---|
| Flue gas cooling/condensation | 144 | 61,9 | 20140 | 10 | 245 |
| Flue gas cooling/condensation | 61,9 | 47 | 43970 | 5 | 2951 |
| Preheating rich solution | 55,9 | 107 | 60680 | 5 | 1187 |
| Cooling lean solution | 122,3 | 38 | 84150 | 5 | 998 |
| CO2-H2O condensation (condenser) | 260 | 133 | 15500 | 5 | 122 |
| CO2-H2O condensation (condenser) | 133 | 70 | 9740 | 5 | 155 |
| CO2-H2O condensation (condenser) | 70 | 23,6 | 5140 | 10 | 111 |
| Reboiler | 122,3 | 122,4 | 64910 | 5 | - |

All streams in the carbon capture process that needs to be heated or cooled are presented in the table, with respect to the case with MVR. Individual temperature differences are used and set to 10 K for gases, 5 K for liquids and condensation/evaporation.

The streams in the cracker plant are shown in Appendix 9. The chosen streams that fulfilled the criteria's presented in section 6.4.2 (page 38) are marked in the table. The hot utility temperatures (steam) are not up to date in the table. The temperatures of the hot utility are those shown in Table 3 in section 3.3.1 (page 8).

# Appendix 11 – Net $CO_2$ captured

<u>Tel-Tek study</u>

The total amount of flue gases fed into the carbon capture plant in the Tel-Tek study was 82 ton/h. The amount of $CO_2$ from the cracker furnaces was 64 ton/h (2006) and from the proposed steam boiler 18 ton/h. As 85 % of the ingoing $CO_2$ is captured following is obtained:

*Captured $CO_2$ from cracker furnaces = 64 \* 0,85 = 54,4 ton/h*

*Emitted $CO_2$ to the atmosphere from cracker furnaces = 64 – 54,4 = 9,6 ton/h*

*Captured $CO_2$ from steam boiler = 18 \* 0,85 = 15,3 ton/h*

*Emitted $CO_2$ to the atmosphere from steam boiler = 18 – 15,3 = 2,7 ton/h*

The net captured $CO_2$ is the quantity captured from the cracker plant subtracted by the quantity of $CO_2$ vented to the atmosphere from the steam boiler.

*Net captured $CO_2$ = 54,4 – 2,7 = 51,7 ton/h*

*Yearly net captured $CO_2$ = 51,7 \* 8000 [h/yr] = 413 600 ton/yr*

<u>This thesis</u>

The total amount of flue gases fed into the carbon capture plant in this thesis was 62.9 ton/h. Differences arise since up to date data was given in this project. 85 % of the ingoing $CO_2$ is captured.

*Captured $CO_2$ from cracker furnaces = 62,9 \* 0,85 = 53,5 ton/h*

*Emitted $CO_2$ to the atmosphere from cracker furnaces = 62,9 – 53,5 = 9,4 ton/h*

The $CO_2$ generated in the existing steam boilers is not captured. Combustion of fuel gas generates 0.17 ton $CO_2$/MWh. The amount of fuel supplied is 30.5 MW.

*Generated $CO_2$ in the existing steam boilers = 0,17 \* 30,5 = 5,2 ton/h*

*Captured $CO_2$ from steam boiler = 0 ton/h*

*Emitted $CO_2$ to the atmosphere from steam boiler = 5,2 ton/h*

*Net captured $CO_2$ = 53,5 – 5,2 = 48,3 ton/h*

*Yearly net captured $CO_2$ = 48,3 \* 8000 [h/yr] = 386 400 ton/yr*