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Partial CO2 capture simulation and cost estimation

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Summary:

Simulation of CO_2 absorption using amine solutions have been performed for many years at University College of Southeast Norway, both with and without vapor recompression.

In this thesis, partial CO_2 capture was evaluated with a constant heat amount available and a specified exhaust flow. Full-flow (treating all the exhaust) and part-flow (treating only a part) alternatives were simulated with Aspen HYSYS and cost estimated. A standard process and a vapor compression configuration were calculated. The absorption column height varied between 5 and 15 stages (estimated to 1 meter each).

These results show that a standard full-flow CO_2 capture process with a 5-stage absorption column gives the lowest cost at 21 NOK/ton CO_2 removed over a 25-year period. However, a full flow-vapor recompression alternative increased the captured amount of CO_2 considerably while increasing capture cost to 28 NOK/ton CO_2.They also show that a vapor recompression CO_2 capture process with 15 stages in the absorption column is the most energy optimum at 2.88 MJ/kg CO_2 removed, and gives the highest removal rate at 48 %.

A sensitivity analysis of both packing and compressor cost show that with an increased cost, the standard process would still give the lowest NOK/ton CO_2. The same was done by setting a price of 0.2 NOK/kW on the recovered waste heat from the plant, here the vapor recompression case gave the lowest NOK/ton CO_2.

To be more certain about the conclusions in this thesis, further calculations related to costs not included in this report must be performed, e.g. costs like the transport related to a full-flow or partial-flow capture process.

Preface

This master thesis was written during the 2017 spring semester as a part of the master program Process Technology for the Faculty of Technology at University College of Southeast Norway.

The following programs were used in this thesis; Microsoft Office (Word and Excel), Aspen HYSYS V8.6 and Aspen In-Plant Cost Estimator V8.4. The reader should have knowledge to chemical engineering terms and should be familiar with Aspen HYSYS. Knowledge about cost estimation is beneficial, but not necessary.

I would like to thank my supervisor Lars Erik Øi who has been with me on four different projects throughout my 5 years at the University College and has always shown great enthusiasm, supervision, and help in every and any matter. Special thanks also go out Hassan Ali from Tel-Tek for his introduction and guidance in using Aspen In-Plant Cost Estimator.

Porsgrunn, 15/5-17

Erik Sundbø

Nomenclature

Abbreviations

CAPEX	Capital expenditure
CEPCI	Chemical Engineering Plant Cost Index
CO_2	Carbon dioxide
LMTD	Lean Mean Temperature Difference
MEA	Monoethanolamine
Mol%	Mole percentage
NOK	Norwegian Kroner
OPEX	Operational expenditure
PFD	Process flow diagram
SS	Stainless steel
TCM	Technology Center Mongstad
USD	United States Dollar
VR	Vapor recompression
Wt%	Weight percentage

Symbol list

Symbol	Unit	Description
η	[-]	Efficiency
ρ	kg/m3	Density
m	kg/h	Mass flow
Q	kW	Heat transfer rate
THK	m	Thickness
U	W/m^2*K	Heat transfer coefficient
v_g	m/s	Gas velocity
V_{gas}	m^3/s	Volumetric flow rate

Contents

1 Introduction

This chapter gives a short introduction to CO_2 capture and its necessity.

1.1 Climate change

There is a correlation between the increase in CO_2 in the atmosphere and the increase in global average temperature. As of April 2017, the CO_2 level in the atmosphere has increased to 406.17 ppm, the highest it has been for hundreds of thousands of years.
The increased emission of greenhouse gases like CO_2 continue to shrink the ice caps and increase the sea levels. A reduction in emission are the only thing that can slow this progress down [1].

1.2 CO_2 capture

One of the measures that could be taken to slow down global warming is CO_2 capture. Capturing CO_2 before it goes out into the air and store it to reduce the pollution, from sources that releases a lot. The four basic systems for capturing CO2 from plant that use fossil fuels or biomass are [2]:

- Post-combustion capture
- Oxy-fuel combustion capture
- Pre-combustion capture
- Capture from industrial product streams

Post-combustion capture is the system used in this project, a system that capture the flue gas after the combustion. This process usually involves separating the CO_2 from the fuel gas using an absorbent, e.g. monoethanolamine.
A CO_2 capture system is often capable of capturing 90 % of the CO_2 from the flue gas stream, making it a very good option to help reduce the emission of green house gasses.

1.3 CO_2 capture projects in Norway

Norway is a leading nation when it comes to CO_2 capture, and this subchapter looks at some of the projects.

1.3.1 Mongstad

As of 2013 the CO_2 Technology Center Mongstad was considered the largest CO2 capture technology test center. The main goal of TCM of being a platform for testing and improving of CO2 capture [3]. What started as a financial catastrophe ended up as a world leader in testing [4].

1.3.2 Yara Porsgrunn

The Yara ammonia plant on Hærøya in Porsgrunn has an annual emmison of 1.2 million tons of CO2 each full production year. Since 1988 they have captured about 0.8 Mt each year, with 0.2-0.3 Mt that CO2 being sold to other industries, while the remainder of the CO2 has been emitted. They have stated that by using proven, existing technology they would be able to capture 210 kt CO2/year [5]. It was announced in 2017 that Aker Solutions had won the contract to perform a concept study for the Yara ammonia plant[6], and that Gassnova would provide further support in studies [7].

1.3.3 Klemetsrudanlegget

The Klemetsrudanlegget is a waste heat recovery plant in Oslo who uses waste to provide environmentally-friendly district heating and electricity. A feasibility study has been done and found that a full-scale capture plant could be established there. A full-scale plant could remove 90 % of the 350 000 ton of CO2 that is emitted [8]. In 2017, it was announced that Gassnova would provide support for further studies [7].

1.3.4 Norcem Brevik

Norcem cement plant in Brevik has been working on plans to implement a full-scale capture plant. A test facility using technology from Asker Solution was established in 2013, and had a successful test period. It was proven that 400 000 ton per year could be captured, cooled and ready for transport [9]. It was announced in 2017 that Aker Solutions had won the contract to perform a concept study for the Yara ammonia plant [6], and that Gassnova would provide further support in studies[7].

1.4 Partial CO_2 capture

It might not always be realistic to capture >90 % of the flue gas, this is where partial CO_2 capture comes in. Figure 1-1 shows a capture situation like the one chosen for this project, a capture plant with a limited amount of energy to use in the reboiler. The figure shows that it is full-flow, this means that all the flue gas will enter the absorption column so that the MEA can capture as much as possible. A process like this will have significantly lower removal rate than 90 %, but could be a cheaper and more realistic option.

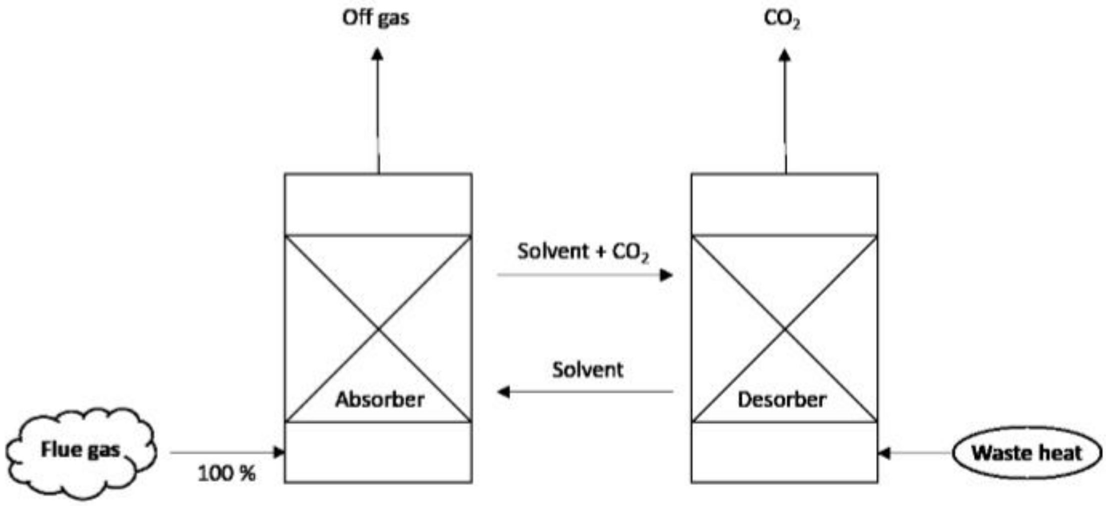

Figure 1-1: Schematic of full-flow capture [10]

Figure 1-2 shows a capture system with part-flow. Here only a percentage, in this case 80 %, of the flue gas enters the absorption column, while the remaining 20 % is released into the air. This could be a good choice of process in plants where the flue gas must be transported a long way or must be further cleaned before entering the absorption column.

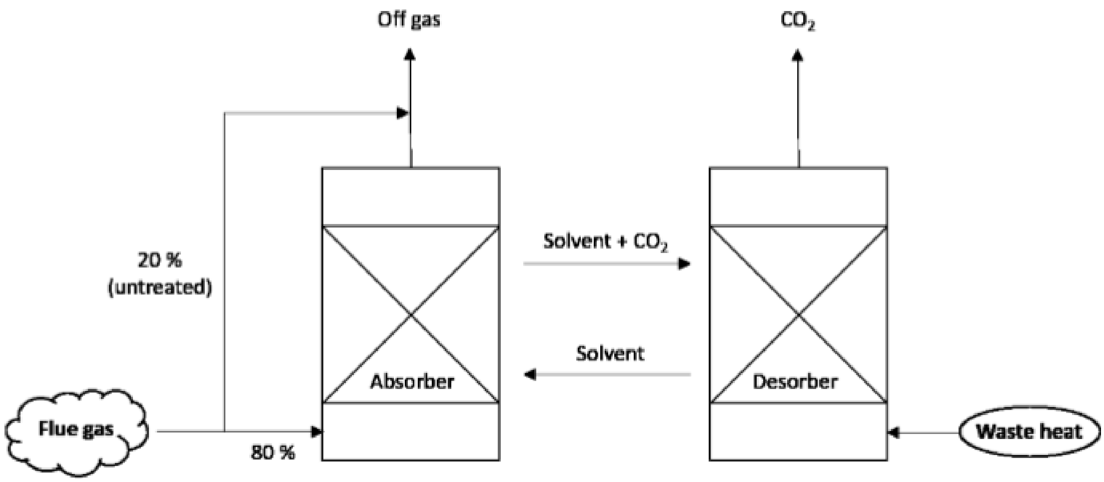

Figure 1-2: Schematic of part-flow capture [10]

1.5 Project description

There is a long history of doing similar projects at the University College of Southeast Norway, and this project follows in that tradition. See the project description in Appendix A for further details for this project.

2 Process description

This chapter describes the process of CO_2 capture and gives a little background with a focus on the case chosen for this project.

2.1 Cement industry

Cement industry is a big contributor when it comes to CO_2 emissions and plants like Brevik has, as mentioned in Chapter 1.3.4, has started to take it seriously and looking to become zero emission plants.

2.2 CO_2 capture process

In this subchapter, both a standard CO_2 capture process and a vapor recompression capture process are described.

2.2.1 Standard case

The flue gas from, e.g. a cement plant, enters through the lower half of the absorption column as shown in Figure 2-1. At the same time lean amine, an amine mixture low in CO_2 content, enters from the top of the column and flows down through the packing inside the column and absorbs the flue gas. The packing helps the absorption into the amine by increasing the surface capture are of the amine. Excess flue gas not captured in the amine exits the top of the absorption column and is released out into the air.
The now rich amine, an amine mixture rich in CO_2 content, travels from the bottom of absorption column and is pumped through a plate heat exchanger which exchanges heat between the rich amine that enters the desorber column and the lean amine that exits the desorber column. After the exchanger, the amine enters the desorber column where it is heated up and CO_2 is released from the amine. The vapor released from the amine exits through the top of the where it goes through a condenser which pumps amine back into the desorber while the now captured CO_2 exit for transport.
Lean amine exits the bottom of the desorber through a reboiler where even more CO_2 is stripped and sent back into the desorber. The remaining lean amine exits the reboiler and exchanges heat with the rich amine in the plate heat exchanger before getting cooled in cooler before re-entering the absorption column.

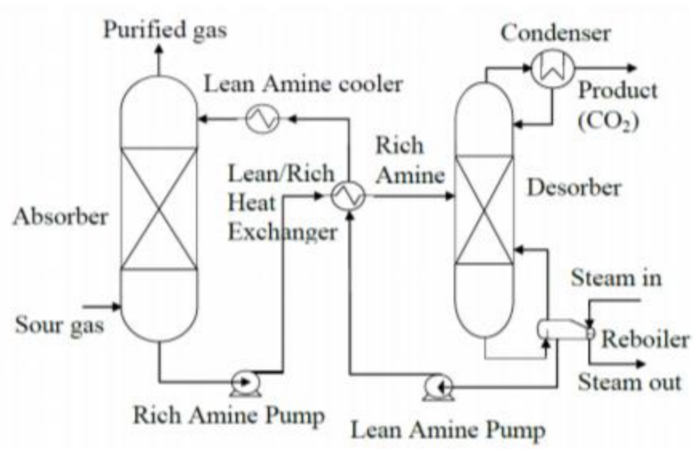

Figure 2-1: Standard CO_2 capture process[11]

2.2.2 Vapor recompression

For vapor recompression, the process is the same as with a standard until the lean amine exits the reboiler. Upon exiting the reboiler the lean amine is flashed through a valve that induces a pressure drop which creates vapor in the mixture. This vapor is the separated from the lean amine in the gas/liquid separator as shown in Figure 2-2, compressed and injected back into the desorber to assist in the regeneration of amine in the bottom of the column.

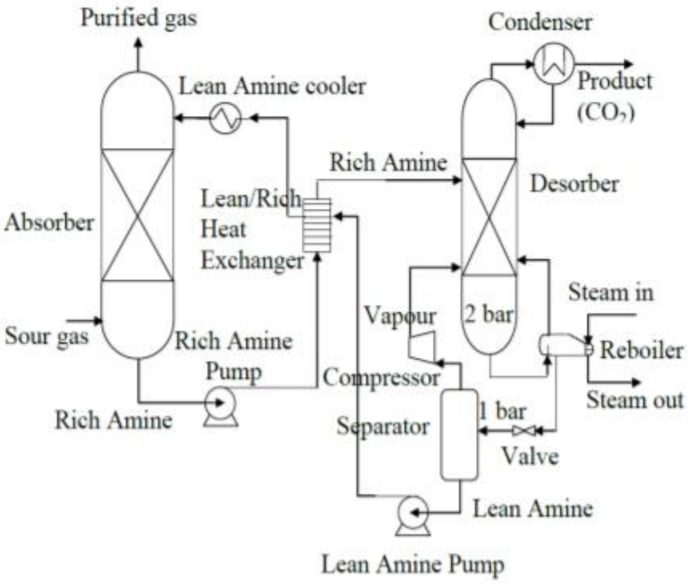

Figure 2-2: CO_2 capture process with vapor recompression[11]

3 Process simulation

This chapter details the specifications set for both base cases alternatives and the simulations in general. It also shows how the simulations are represented in Aspen HYSYS and details the removal rate of different alternatives and which one that is the most energy optimum process.

3.1 Standard base case specifications

The process parameter detailed in Table 3-1 are the specific parameters entered in the standard base case in the Aspen HYSYS simulation model. As detailed in the table below this was a full-flow case, with 10 stages in the absorption column and MEA as the absorbent. The parameters that are **bolded** in the text are parameters that stayed the same throughout every simulation in both the standard cases, vapor compression cases whether they were full-flow or part-flow.

Parts of the specification data are cited from previous master thesis by Kallevik (2010), Svolsbru (2013) and Park (2016) which all had a goal of replicating the real life parameter of a cement plant.

There is also one important more output parameter from the simulations, it was decided early on that the simulation were going to be done based on Brevik Cement Plant. It was also specified that the energy used in the reboiler was going to be waste heat energy from the cement plant itself. This energy is “free” energy that could be extracted from the cement making process and used for other things instead of just being spilled heat.

After a deliberation with Lars Andre Tokheim it was estimated that this waste heat could represent around 25 MW in power [12]. The aim of the simulations was then to see how much CO_2 could be removed from the flue gas of “string one” of the plant, one of two strings (chimneys), using 25 MW as the energy available for the reboiler.

Table 3-1: Standard base case input paramaters

Process input parameters	Unit	Value
Flue gas temperature	[°C]	80
Flue gas pressure	[bar]	1.1
Flue gas rate	[kmol/h]	8974
CO_2 content in flue gas	[mol%]	17.8
H_2O content in flue gas	[mol%]	20.63
Lean amine temperature	[°C]	40
Lean amine pressure	[bar]	1.01
Lean amine rate	[kg/h]	545000
MEA content in Lean amine	[wt%]	29.00
CO_2 content in Lean amine	[wt%]	5.5
Number of stages in absorber column (N_{stages})	[-]	10
Murphree efficiency	[-]	0.15
Pressure increase across Rich pump	[bar]	0.9
Rich pump adiabatic efficiency	[%]	75
Rich amine temperature out of Lean/Rich HX	[°C]	101.2
Number of stages in desorber column	[-]	8
Murphree efficiency in desorber column	[-]	1.0
Reflux ratio in desorber column	[-]	0.3
Reboiler temperature	[°C]	120
Reboiler pressure	[bar]	2

3.2 Standard case overview

Figure 3-1 shows process flow diagram (PFD) of the Aspen HYSYS case that was used for the standard case. This file was a continuation of a file that was created during a semester project [13] that was completed during the 2016 autumn semester. This was done to still be able to use the amine pack that is ideal for CO_2 capture simulation but has since been replaced with another pack that can be problematic to run. The setup of the Aspen HYSYS model is the same as the system described in Chapter 2.2.1, and is a typical setup for these types of simulations.

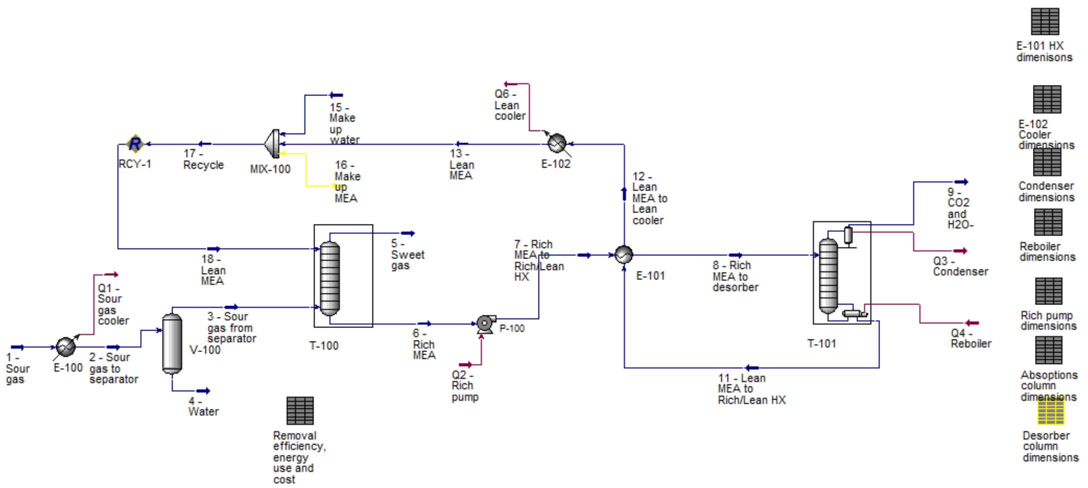

Figure 3-1: Standard CO_2 capture plant in Aspen HYSYS

Removal rate is calculated in a spreadsheet within Aspen HYSYS, doing calculations in these types of spreadsheets allows the user to take parameters directly from the different material flows and equipment. This makes it so that the spreadsheet updates itself as the parameters in the system are being changed making it easy to create a new model with new parameters, e.g. changing the number of stages in the absorption column. Figure 3-2 shows the removal rate for the standard cases, and it show that a 15-stage, full-flow process gives the best removal rate at 41.2 %.

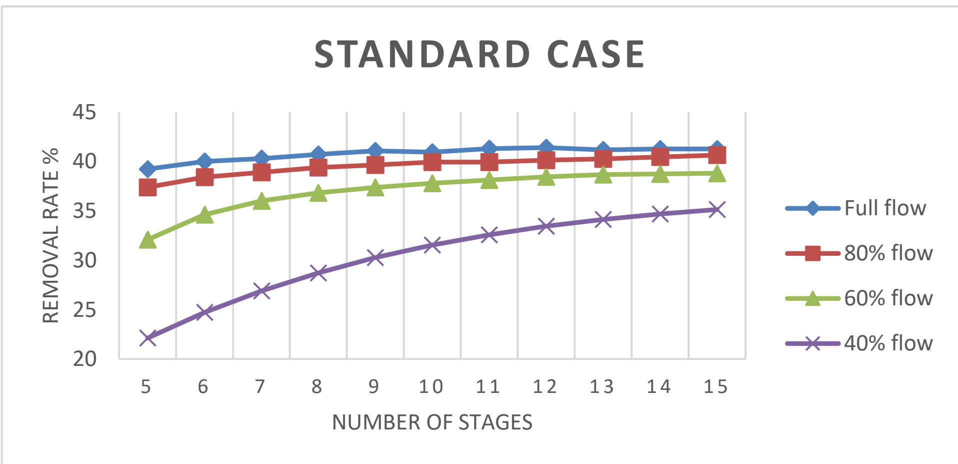

Figure 3-2: Removal rate in standard cases

Figure 3-3 shows the energy use for the standard case with different flows into the absorption column. Full-flow has the lowest energy use and it also shows that a high number of stages in the absorption column is the most energy efficient. The most energy efficient is the full-flow, with 15-stages which have a energy use of 3.1 MJ/kg CO_2.

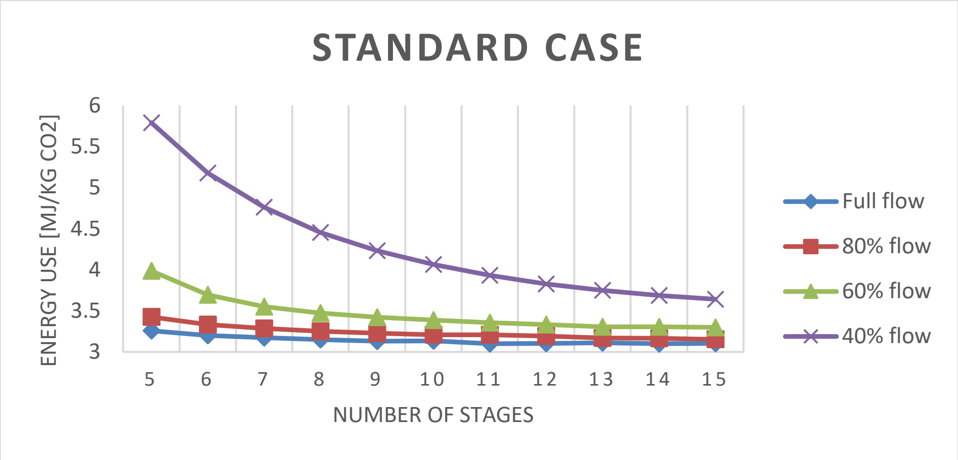

Figure 3-3: Energy use in standard cases

3.3 Vapor recompression base case specifications

As mentioned in Chapter 3.1 the bolded specifications from Table 3-1 also applies to the vapor recompression base case, which also has 10 stages in its absorption column. The parameters in Table 3-2 are all constant throughout the different simulations involving vapor recompression.

Table 3-2: Vapor recompression base case input paramaters

Process input parameters	Unit	Value
Pressure from valve	[bar]	1.2
Compressor Adiabatic Efficiency	[%]	75
Pressure from compressor	[bar]	2
Pressure increase across Return pump	[bar]	0.8
Return pump adiabatic efficiency	[%]	75

3.4 Vapor recompression case overview

The vapors recompression case was built out from the same model that was used in the standard case. The stream from the desorber to the desorber is disconnected and is replaced by the inner loop of the as shown in Figure 3-4. Since there is pressure drop over the valve a return pump is put after the separator to increase it again.

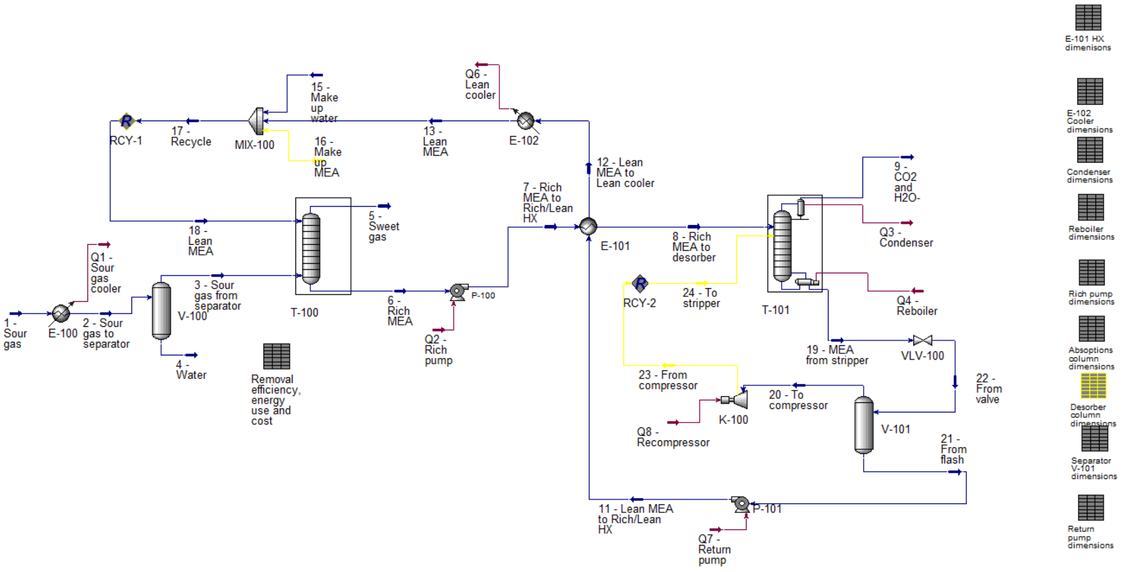

Figure 3-4: CO_2 capture plant with vapor recompression in Aspen HYSYS

Figure 3-5 gives a closer look at the PFD of the vapor recompression process, more clearly showing what was described in Chapter 2.2.2.

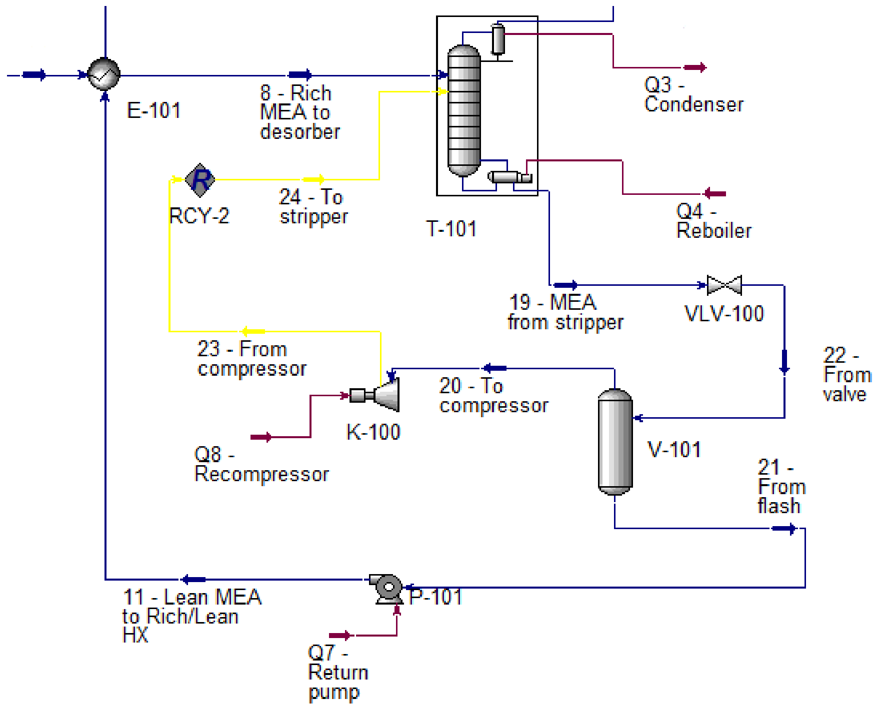

Figure 3-5: Closeup of vapor recompression cycle

The removal rate of the different vapor recompression cases can be found in Figure 3-6. It is apparent that higher number of stages in the absorption column leads to a higher removal rate, and that a full-flow process is the most efficient. With 15 stages in the absorption column, a full-flow capture process can remove 48 % of the CO_2.

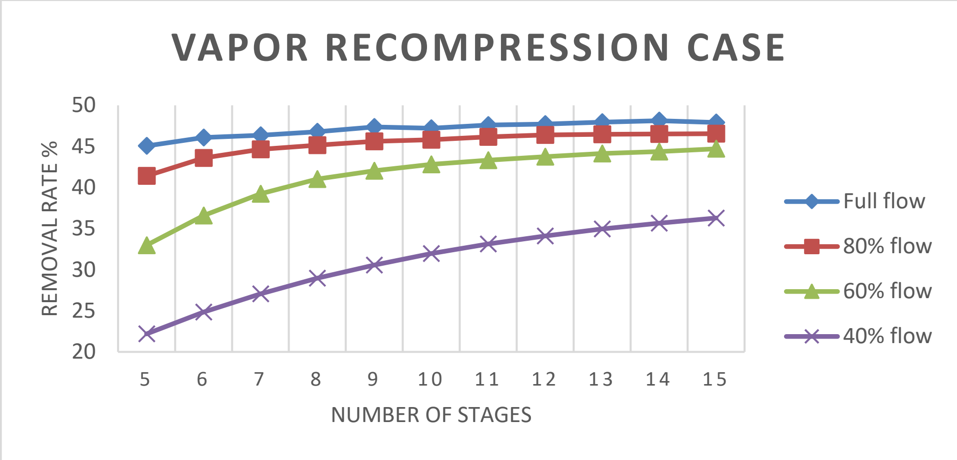

Figure 3-6: Removal rate in vapor recompression cases

For the energy use, a high number of stages in the absorption column is also ideal as shown in Figure 3-7. It also shows again that a full-flow process is the most energy efficient with an energy use of 2.88 MJ/kg CO_2.

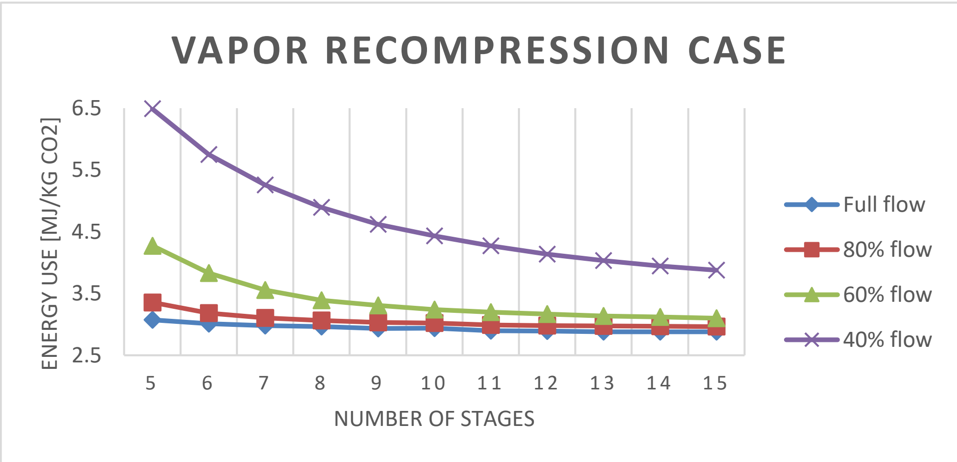

Figure 3-7: Energy use in vapor recompression cases

3.5 Energy optimization

Figure 3-8 show a comparison between the best alternatives from the standard case and the vapor recompression, the full-flow alternatives. There is a clear difference in the removal rate between these two alternatives when going from 5 to 15 stages the removal rate for the standard case goes from 39 % to 41 %, while the vapor recompression case goes from 45 % to 48 %.

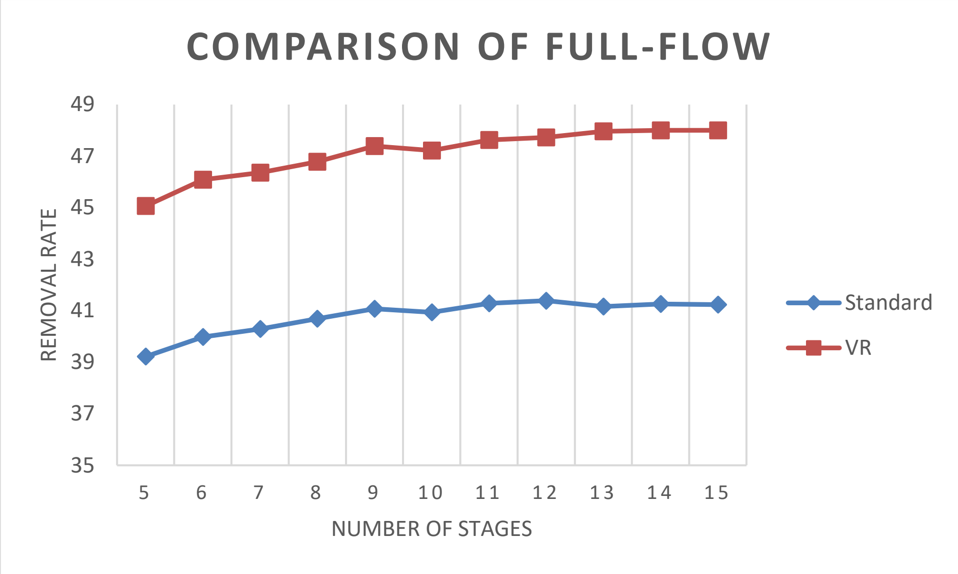

Figure 3-8: Comparison of removal rate for standard and VR

Comparing the two most energy effective alternatives from the standard and vapor recompression cases, both full-flow, it shows in Figure 3-9 that vapor recompression is more energy efficient than the standard case. At 15 stages the vapor recompression case gets down to 2.88 MJ/kg CO_2, while standard case at the same number of stages gives an energy use of 3.1 MJ/kg CO_2. Vapor recompression is the most energy efficient process of the two and the most energy optimum of the vapor recompression cases is the 15-stage process at full-flow.

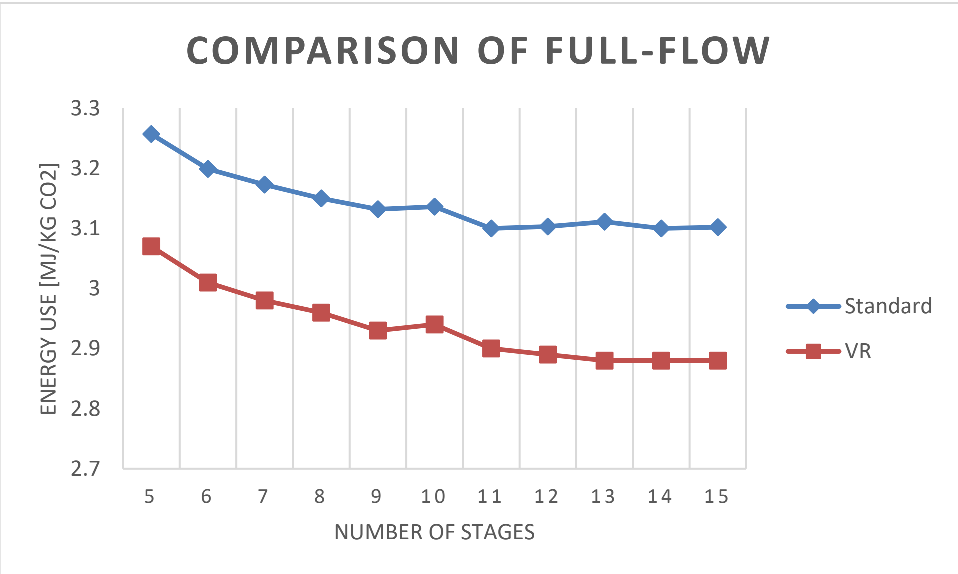

Figure 3-9: Comparison of energy use for standard and VR

4 Equipment dimensioning

This chapter details the work done relating to dimensioning of the different equipment used in the simulation models for both standard and vapor compression cases.

4.1 Absorption column

For the absorption column, there are several different outputs calculated; the weight of the desorber shell, packing volume and the absorber column area. The gas velocity is set to 2 m/s and was used in Equation 4.1 to calculate the absorber column.

$$A = \frac{\dot{V}_{gas}}{v_g} \tag{4.1}$$

The desorber diameter was then found from Equation 4.2 using the calculated area.

$$D = \left[\left(\frac{4}{\pi}\right) * A\right]^{0.5} \tag{4.2}$$

The assumption was then made that the thickness of the shell is 0.01 meter and that the absorption column is the height of the packing, where one stages is 1 meter in height, plus an extra 23 meter. The extra height is to house other internals; liquid distributer, liquid catchers, packing support, gas distributer, liquid and gas inlets. The volume of the shell construct itself was found by Equation 4.3 and then simply multiplied by the weight of SS which is 8000 kg/m^3.

$$V_{shell} = \left[\left(\frac{\pi}{4}\right) * \left(THK_{shell} * 2\right)\right] * H_{absorber} \tag{4.3}$$

Packing volume was calculated from the diameter of the absorption column and the height of the packing using Equation 4.4.

$$V_{packing} = A * N_{stages} \tag{4.4}$$

The most central parameters used in the dimensioning of the absorption column is listed in Table 4-1 below, with unit description and some related notes.

Table 4-1: Absorption column dimensioning parameters

Parameter	Unit	Note
Gas velocity	[m/s]	
Flue gas rate	[m^3/s]	Input from Aspen HYSYS
Absorber column area	[m^2]	
Absorber diameter	[m]	
Total packing height	[m]	
Total height of absorber	[m]	
Volume of absorber shell	[m^3]	
Weight of absorber shell (SS)	[ton]	Outputs from calculations
Packing volume	[m^3]	

4.2 E-101 Heat exchanger

In the heat exchanger dimensioning, the heat transfer area is the main result. The inputs from Aspen HYSYS in these calculations are the cold side duty from the, which is converted from kJ/h to kW to use as the heat transfer rate, Q, in Equation 4.5. A heat transfer coefficient, U, of 1500 W/m^2*K [10] was chosen for the heat exchanger as well.

$$A = \frac{Q}{LMTD * U} \tag{4.5}$$

Table 4-2: E-101 Heat exchanger dimensioning parameters

Parameter	Unit	Note
Cold side duty	[kJ/h]	Input from Aspen HYSYS
Heat transfer rate	[kW]	
LMTD	[°C]	Input from Aspen HYSYS
Total transfer area	[m^2]	Output from calculation

4.3 E-102 Cooler

The transfer area is the main output for the cooler calculations. The lean cooler duty, lean amine temperature in and out are the inputs taken from Aspen HYSYS to do the calculations. The Lean Mean Temperature Difference was calculated using Equation 4.6, with the temperature of the cooling water assumed to the temperatures shown in Table 4-3. Equation 4.5 was again used to calculate the transfer area, this time using a heat transfer coefficient of 800 W/m^2*K [13].

$$LMTD = \frac{(T_{h2} - T_{c2}) - (T_{h1} - T_{c1})}{\ln\left[\frac{(T_{h2} - T_{c2})}{(T_{h1} - T_{c1})}\right]} \tag{4.6}$$

Table 4-3: E-102 Cooler dimensioning parameters

Parameter	Unit	Note
Lean cooler duty	[kW]	Inputs from Aspen HYSYS
Lean amine T in	[°C]	
Lean amine T out	[°C]	
Cooling water in	[°C]	Assumed T_{in}=8°C
Cooling water out	[°C]	Assumed T_{out}=23°C
LMTD	[°C]	
Total transfer area	[m^2]	Output from calculation

4.4 Desorber column

For the desorber column the main outputs are almost the same as the absorption column, mainly the weight of the column and the packing volume as shown in Table 4-4. First the desorber column are was found using Equation 4.7 using a gas velocity of 1 m/s.

$$A = \frac{\left(\frac{\dot{m}_v}{\rho_v}\right) + \left(\frac{\dot{m}_l}{\rho_l}\right)}{v_g} \tag{4.7}$$

The diameter of the desorber was then found using Equation 4.2 and the packing volume was found using Equation 4.4 with the assumption that each stage in the desorber had a height of 1 meter. The weight of the shell was found using Equation 4.3, a total desorber height of 25 meters and the thickness of the desorber wall set to 0.01 meter.

Table 4-4: Desorber column dimensioning parameters

Parameter	Unit	Note
Rich mass flow (vapor)	[kg/h]	Inputs from Aspen HYSYS
Rich mass flow (liquid)	[kg/h]	
Rich density (vapor)	[kg/m^3]	
Rich density (liquid)	[kg/m^3]	
Rich volume flow (vapor)	[m^3/s]	
Rich volume flow (liquid)	[m^3/s]	
Total volume flow of rich amine	[m^3/s]	
Desorber column area	[m^2]	Outputs from calculations
Desorber diameter	[m]	
Number of stages	[-]	
Total height of desorber	[m]	
Volume of desorber shell	[m^3]	
Weight of desorber shell	[ton]	Outputs from calculations
Packing volume	[m^3]	

4.5 Condenser

As with the cooler, the condenser use data from Aspen HYSYS and an assumed temperature of the cooling water (shown in Table 4-5) to calculate the Lean Mean Temperature Difference using Equation 4.6. The main output from the calculations, the total transfer area, is then calculated with Equation 4.5 using a heat transfer coefficient of 2000 W/m^2*K [13].

Table 4-5: Condenser dimensioning parameters

Parameter	Unit	Note
Condenser duty	[kW]	Inputs from Aspen HYSYS
Vapor T in	[°C]	
Vapor T out	[°C]	
Cooling water in	[°C]	Assumed T_{in}=8°C
Cooling water out	[°C]	Assumed T_{out}=23°C
LMTD	[°C]	
Total transfer area	[m^2]	Output from calculation

4.6 Reboiler

Like the other heat exchangers, the main output from the reboiler calculations is the transfer area. Only here superheated stream is used as the opposing fluid to the amine to heat up the amine at temperatures detailed in Table 4-6. Using these temperatures and the temperatures taken from Aspen HYSYS the LMTD is calculated using Equation 4.6. The total transfer area is then calculated using Equation 4.5 and a heat transfer coefficient of 2500 W/m^2*K [13].

Table 4-6: Reboiler dimensioning parameters

Parameter	Unit	Note
Reboiler duty	[kW]	Inputs from Aspen HYSYS
Lean amine T in	[°C]	
Lean amine T out	[°C]	
Superheated steam T in	[°C]	Assumed T_{in}=130°C
Superheated steam T out	[°C]	Assumed T_{out}=120°C
LMTD	[°C]	
Total transfer area	[m^2]	Output from calculations

4.7 Rich pump

For the rich pump, the additional energy needed to compensate for the height of the pump must reach based on the height of the desorber. The inlet height was found using Equation 4.8.

$$H_{inlet} = H_{desorber} * 0.8 \tag{4.8}$$

The additional energy used was then calculated from Equation 4.9 and then added together with the rich pump duty taken from Aspen HYSYS.

$$W_{additional} = \frac{\dot{m} * H_{inlet} * g}{\eta_{pump} * 1000} \tag{4.9}$$

Table 4-7: Rich pump dimensioning parameters

Parameter	Unit	Note
Rich pump duty	[kW]	Input from Aspen HYSYS
Rich amine flow	[kg/s]	
Desorber height	[m]	
Rich amine inlet height	[m]	
Gravitational constant	[kg*m/s2]	
Adiabatic efficiency	[-]	
Additional duty required	[kW]	Output from calculation
Actual rich pump duty	[kW]	

4.8 Separator

The main output from the separator calculations is the diameter of the column. This is found from the gas velocity, here set to 1 m/s, and the volumetric flow rate taken from Aspen HYSYS using Equation 4.10.

$$D = \sqrt{\frac{(4 * \dot{V})}{(\pi * v_g)}} \tag{4.10}$$

Table 4-8: Separator dimensioning parameters

Parameter	Unit	Note
Gas velocity	[m/s]	
Volumetric flow rate	[m3/s]	Output from Aspen HYSYS
L/D ratio	[-]	Set to 4
Vessel height	[m]	
Vessel diameter	[m]	Output from calculation

4.9 Compressor

The output for the compressor from Aspen HYSYS is simply the power usage and is there for the only parameter shown in Table 4-9.

Table 4-9: Compressor dimensioning parameters

Parameter	Unit	Note
Compressor duty	[kW]	Input from Aspen HYSYS

4.10 Return pump

As with the rich pump, the return pump needs additional energy to compensate for the height that the fluid must be carried in order to enter the absorption column. The inlet height was found using Equation 4.11.

$$H_{inlet} = H_{absorber} * 0.8 \tag{4.11}$$

The additional energy used was then calculated from Equation 4.9 and then added together with the return pump duty taken from Aspen HYSYS.

Table 4-10: Return pump dimensioning parameters

Parameter	Unit	Note
Return pump duty	[kW]	Input from Aspen HYSYS
Lean amine flow	[kg/s]	
Absorber height	[m]	
Lean amine inlet height	[m]	
Gravitational constant	[kg*m/s2]	
Adiabatic efficiency	[-]	
Additional duty required	[kW]	Output from calculation
Actual return pump duty	[kW]	

5 Cost estimation

In this chapter, the two methods used for cost estimations and how the installation factors were chosen are detailed.

5.1 Web-based calculator

The first method used for the cost estimation is a web-based calculator [14] delivered by McGraw-Hill, Inc based on the book “Plant Design and Economics for Chemical Engineers” from Peters and Timmerhaus. It is a simple tool that allows the user to select different equipment and input the correct dimension data to get the cost of the equipment and is also simply referred to as “Peters” in the URL of the calculator.
The cost data produced by this calculator is in 2002 USD, and were converted to 2016 USD using the Chemical Engineering Plant Cost Index data found in Appendix C. For these calculations, all components except the pumps could be chosen to be made of SS316, therefor theses are cast iron and were ascribed a different installation factor than the others.
There are also some limitations to the size of the dimensions the calculator is able to handle, so in cases where the dimension data were too large for the calculator, Capacity Factor Method [15] was used to get the cost data using Equation 5.1 with the exponent “e” equal to 0.65.

$$Cost_B = Cost_A * \left(\frac{Capacity_B}{Capacity_A} \right)^e \tag{5.1}$$

The results from these calculations are shown in Figure 5-1 which shows that a capture process with a low number of stages is the most cost effective, more specifically a 5-stage standard process is the cost optimum option here. Both the standard and vapor recompression process had their cost optimum with a 5-stage absorption column, with a cost of 20.5 NOK/ton CO_2 and 31.8 NOK/ton CO_2 respectively.

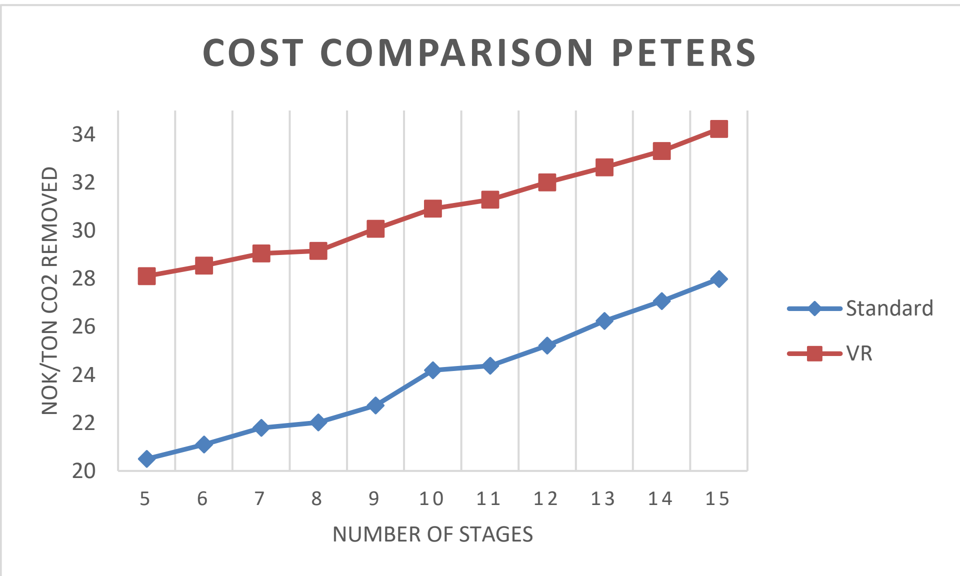

Figure 5-1: Comparison of cost estimations done with Peters

The internals for both the desorber and absorption column was calculated from the same data for both Peters and Aspen In-Plant. With the cost of packing for both desorber and absorption column set to 7600 $\$/m^3$ in 2010 USD. With a cost of 4000 $/m2 for the liquid distributor, 2000 $/m2 for the liquid catcher and 800 $/m2 for packing support grid, all in 2011 USD [16]. This was all converted to 2016 USD using Appendix C.

The separator was assumed to be stainless steel bubble-plate towers with 4 trays. And the valve after the desorber in the vapor recompression case was assumed to be neglectable.

5.2 Aspen In-Plant

The second set of calculations were done using Aspen In-plant Cost Estimator V8.4 which is a cost estimation program made by Aspen Technology, Inc. This is a cost estimation tool that allows the user to create both simple cost estimations using only a few parameters, or detailed cost estimations using a lot of parameters.
For the calculations done here the bare minimum number of parameters needed was inserted to complete a large number of calculations that had to be done within a reasonable timeframe. Aspen In-Plant Cost Estimator V8.4 use data from the first quarter of 2013, therefore, the data from the calculations were converted to 2016 USD using the Chemical Engineering Plant Cost Index data found in Appendix C.

This second calculation method was done to get one low estimate in the web-based calculations, and one high estimation in the form of the Aspen In-Plant calculations. The difference between the two methods was lower than expected, but there was still a difference which in turn gives some numbers to confirm the trend in the results.

Figure 5-2 shows the cost comparison between the estimations done for standard and vapor recompression process in Aspen In-Plant. The trend here is like the one found in Figure 5-1, with the standard process at a low number of stages being the most cost efficient option. At 5 stages the cost of the standard case is 22.1 NOK/ton CO_2 captured, and 31.8 NOK/ton for the vapor recompression case.

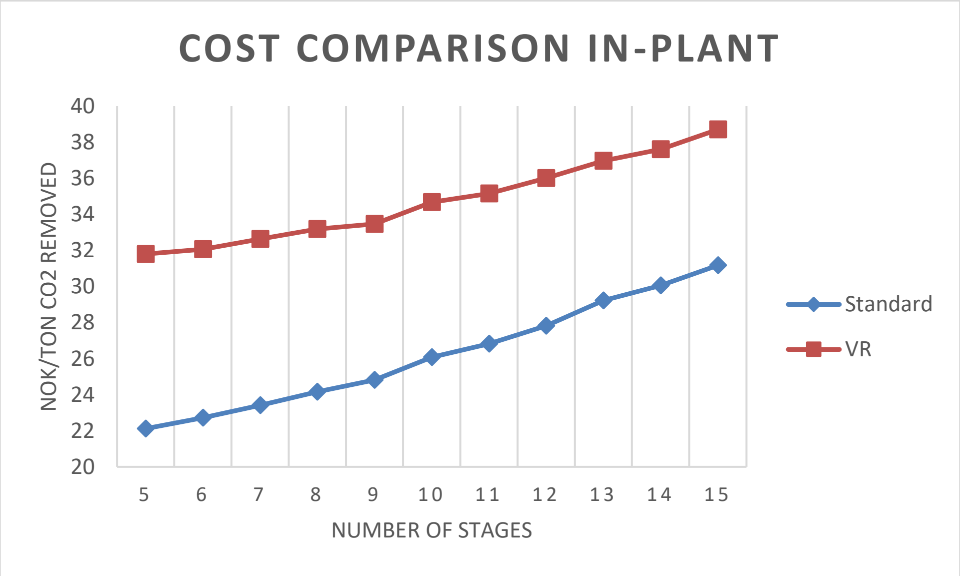

Figure 5-2: Comparison of cost estimations done with Aspen In-Plant

5.3 Installation factor

Installation factors used for the cost estimations were found using Appendix D which contains a table of installation factor created by Nils Henrik Eldrup[15]. Almost every installation factor in the cost estimation was calculated under the assumption that they were made from "Stainless Steel (SS316) Welded" and therefore had a material factor of 1.75.
A new installation factor for this equipment was found based on this using Equation 5.2.

$$f_C = f_{TC} - f_P - f_E + f_M\left(f_P + f_E\right) \tag{5.2}$$

Where:

f_C = Installed factor
f_{TC} = Total installed cost factor
f_P = Piping cost factor for equipment
f_E = Equipment cost factor
f_M = Material cost factor

5.4 OPEX

The following was considered when calculating the operational expenditure of the plant.

5.4.1 Utilities

For the utilities, it was assumed that the waste heat energy used in the reboiler and the cooling water used in the condenser and cooler was all considered free of charge. This would not be the case in a real-world scenario and is something that could be done in the future to get more accurate results. Therefore there are only three different factor considered for the utilities; rich pump, return pump and compressor duty.
They are calculated from an assumed uptime for the capture system of 8000 hours per year and is calculated over a 25-year period using the discount factor calculated from Equation 5.2.

5.4.2 Maintenance

The utilities are not the only operational expenditure, there also must be done periodical maintenance of the equipment. Maintenance is often set to be a certain percentage of the cost of the capital expenditure and that percentage varies over time as the equipment gets older or depending on the size of the project. It may vary from as low as 1.5 % to more than 15 %, but for this project, the maintenance cost was set to 4% yearly [17]. Smaller expenses like raw material, e.g. amine refill, is considered a part of these expenses.

5.4.3 Project economics

For the project economics, it was determined that the project was going to be looked at over 25-year period with an annual rate of return of 7.5 %. The project is assumed to be built in year “0” and has a running time until year “25”. Using Equation 5.2 [18] the discount factor is found to be 11.15 and used to calculate the cost of the OPEX over the 25-year period.

$$D_n = \frac{1}{(1-r)^n} \tag{5.2}$$

where

D_n = discount factor in year ‘n’
r = rate of return
n = year number

5.5 Sensitivity analysis

A sensitivity analysis was also done on three different scenarios to test the sensitivity of the calculations done.

5.5.1 Increased packing cost

The packing cost represents the largest pre-installed cost and could be a big factor in the total investment cost of a CO_2 capture plant. A sensitivity analysis was therefore done where the cost of the packing was doubled, from 7600$/m^3 to 15200$/m^3. Increasing the cost of packing decreases the cost gap between the standard case and the vapor recompression case, while widening the gap between the cost of full-flow and part-flow capture as shown in Table 5-1 below.

Table 5-1: Comparison of alternative with high packing cost

Alternatives	Peters (NOK/ton CO_2)	In-plant (NOK/ton CO_2)
Standard, FF (5 stages)	24.7	26.4
Standard, FF (10 stages)	32.3	34.2
Standard, 80 % flow (8 stages)	26.9	28.1
VR case, FF (5 stages)	31.8	34.9
VR case, FF (10 stages)	38.0	41.7
VR case, 80 % flow (8 stages)	34.4	37.0

5.5.2 Increased compressor cost

There is a lot of uncertainty when it comes to the cost of the compressor. A compressor capable of dealing with amines is required for a vapor recompression system and could be very expensive. In the cost estimations, all components are assumed to be SS316 where possible, this also includes the compressors, but since the compressor must be able to handle amines it might have to be made from some sort of exotic material, and it might have to be a custom build. It would not be unrealistic that the compressor cost would double and Table 5-2 shows that this would only further the gap in cost per ton CO_2 removed between the standard case and the vapor recompression case.

Table 5-2: Comparison of alternative with compressor cost

Alternatives	Peters (NOK/ton CO_2)	In-plant (NOK/ton CO_2)
Standard, FF (5 stages)	20.5	22.1
Standard, FF (10 stages)	24.2	26.1
Standard, 80 % flow (8 stages)	21.5	22.7
VR case, FF (5 stages)	34.0	37.9
VR case, FF (10 stages)	36.4	41.0
VR case, 80 % flow (8 stages)	35.6	39.0

5.5.3 Cost of waste heat

Up until now, the waste heat has been considered free of cost in the cost estimations, in a real-world scenario, this is not likely. There will be cost related to creating the system that recovers the heat from the cement plant, and there will also be additional cost related to maintenance of the recovery system. One of the biggest differences in cost for the two cases over a 25-year period is the operating cost since there only is one pump to consider for the standard case based on the parameters set for this project. Therefore setting a cost on the recovery of the waste heat would as low as 0.2 NOK/kW would, as shown in Table 5-3, be enough to make the vapor recompression case the better option. It does now not only cost less per ton CO_2 removed, but it also removes more CO_2 over the same period making it a far better option in this scenario.

Table 5-3: Comparison of alternative with cost on waste heat

Alternatives	Peters (NOK/ton CO_2)	In-plant (NOK/ton CO_2)
Standard, FF (5 stages)	101.4	103.0
Standard, FF (10 stages)	101.7	103.6
Standard, 80 % flow (8 stages)	102.0	103.3
VR case, FF (5 stages)	98.5	101.6
VR case, FF (10 stages)	98.1	101.9
VR case, 80 % flow (8 stages)	100.0	102.6

5.6 Cost optimization

Cost calculations were also done for two part-flow cases which had a similar removal rate as the lowest cost options for the standard case and vapor recompression case. This was done to get an insight if a part-flow capture process would be cheaper than a full-flow process. Table 5-4 shows a comparison between different the different cost estimations done with both with and without vapor recompression, part-flow, and full-flow.
From the table, it is clear that a standard case is the cost optimum capture method, and it also shows that full-flow is slightly more cost effective than the part-flow if only by a small margin.

Table 5-4: Comparison of cost estimations

Alternatives	Peters (NOK/ton CO_2)	In-plant (NOK/ton CO_2)
Standard, FF (5 stages)	20.5	22.1
Standard, FF (10 stages)	24.2	26.1
Standard, 80 % flow (8 stages)	21.5	22.7
VR case, FF (5 stages)	28.1	31.8
VR case, FF (10 stages)	30.9	34.7
VR case, 80 % flow (8 stages)	29.7	32.3

6 Discussion

In this chapter, there will be discussion around the work that has been done and the results that have been found. Cost estimations found using Peters will be used for the discussion to not create confusion in the text. This can be done since the trends of the two cost are similar.

6.1 Uncertainties in the calculations

There are some minor uncertainties in Aspen HYSYS related to the material streams and equilibrium models. Tuning the model to have a minimum approach of 10 °C in the E-101 heat exchanger and a usage of 25 MW in the reboiler is a very sensitive operation, and a small difference here can have a big impact on both the removal rate and energy use. In the simulation done in this project, the margin of error on these two parameters has been kept to within 1 %, but it is possible that this could have an impact on the results of these calculations.

The biggest uncertainty rests in cost estimations both when it comes to the estimations done on the equipment and the installation factors. Compressor and packing cost are the two biggest uncertainties in the equipment cost estimations since they represent the biggest expenses. A difference in packing cost would have a big impacting in the cost estimations when comparing a full-flow processes against a part-flow process with a similar removal rate since these are close in cost per ton CO_2 removed with the chosen price a fluctuation here could have a big impact.
For the compressor, there is a lot of uncertainty around the cost since it might need to be specially made for the process since it must be able to run amines through it for several years. Therefore it might have to be made from an exotic material, further increasing the cost, which in turn would widen the cost gap between the standard process and a vapor recompression process.
Installation cost could vary from those presented in Appendix D, all the factors that combine to represent the total installation factor are subject to change.

But the calculations done does imply that the conclusions made based on the calculations are not influenced by this to a large extent. The same conclusions would probably be drawn if a different style of calculations had been executed.

6.2 Energy optimum process

An energy optimization has been done and a vapor recompression case with full-flow and 15 stages in the absorber has been found to be the most energy optimum process using 2.88 MJ/kg CO_2 captured. This option also gave the best removal rate of 48 % compared to the 15-stage standard full-flow with a removal rate of 41 % with an energy use of 3.1 MJ/kg CO_2. As mentioned in the chapter above there is some uncertainty related to the calculations of the removal rate and energy use because of the sensitivity of the simulations, but that is likely to have little impact on the conclusion of the energy optimum process.

6.3 Cost optimum process

Calculations were done to find the cost optimum process both with and without vapor recompression. There were also completed four calculations on part-flow alternatives with similar removal efficiencies as the best alternatives with and without vapor recompression to see if a part-flow capture process could be more cost efficient.

A full-flow, 5-stage standard capture process was found to be the cost optimum process at 20.5 NOK/ton CO_2 captured, with the closest alternative being an 80 % part-flow, 8-stage standard process. With a similar removal rate as the cost optimum process, the part-flow process has a cost of 21.5 NOK/ton CO_2 captured making it seemingly almost as good of an alternative. There are things not considered in the cost estimation that could affect the results of this conclusion such as transport and cleaning of the gas before it enters the absorption column.
Since there is a different amount of flue gas entering the absorption column in the two processes there will be a different cost in getting the right amount of gas in there, and there might be a need for different levels of cleaning of the gas before the absorption column as well. There might be a need for removal of NO_x and SO_x from the flue gas before it enters the absorption column since these particles might affect the amines in the system [19]. The more gas that needs to be cleaned, the more expensive equipment is needed.
The vapor recompression capture process is more expensive, with a full-flow, 5-stage process having a cost of 28.1 NOK/ton CO_2 captured. But it does, in turn, capture almost 6 % more CO_2 than the standard process, so it can be said that there is a reasonable incremental increase in cost for the vapor recompression case. It is a viable alternative if there is interest in investing a little bit more to remove more CO_2, and the cost difference between this and the cost optimum alternative will only decrease when a monetary cost is determined for the waste heat.
Uncertainty surrounding the compressor price also stems from the fact that there is very little information around this available. Only one paper[20] related to this was found during the project and that contained a compressor that seemed to be quite under estimated when it came to the cost.

6.4 Further work

To get more certainty in the result more detailed cost estimation should be done on some of the equipment. Further research into compressor and packing cost would be good steps towards improving the accuracy of the calculations, Aspen In-Plant also allows you to go into even more detailed specifications on each equipment which in turn will make the cost estimations more realistic.
Use of different installation factor would also have a profound effect on the cost estimations.
There is also a lot of things not included in the calculations, further work could include these to improve the accuracy of the cost estimation. This could mean including cost of NOx and Sox removal before the flue gas enters the absorption column, and how much it costs to transport the flue gas to the absorption column. This is a cost that could fluctuate based on the amount of gas that is delivered to the absorption column, e.g. if it is a full-flow or part-flow stream of gas. There is also smaller things and equipment that could be taken into account, there are a lot of smaller things that is involved in a CO_2 capture process that does not show up in an Aspen HYSYS model.
The inclusion of the cost of waste heat would also be a logical next step since it is not realistic that this energy would be free, therefore putting a monetary value on the waste heat would help to improve the cost estimation. And it would also, as shown in the sensitivity analysis, reduce the gap in cost between a standard capture process and a vapor recompression capture case.

An abstract that was sent in to SIMS2017 and approved for the conference is included in the report as Appendix B, and represents some work that will be done related to this. Further work can be done related to this and if the paper created for this is approved it will lend further trust in the results from this report.

7 Conclusion

The calculations done shows that a low number of stages in the absorption column is optimal from an economical point of view. It is also optimal with full-flow compared to 80 % part-flow with a similar removal rate as the full-flow capture process.
The cost-optimal process has been shown to be a standard process with full-flow and 5 stages in the absorption column, giving a cost of 20.5 NOK/ton CO_2 removed over a period of 25 years.

Vapor recompression gives a higher removal rate than the standard process with the lower cost option having a removal rate of 45.1 % compared to 39.2 %, it also gives a reasonable incremental increase in cost at 28.1 NOK/ton. This is still a low cost for removal of CO_2, and could, therefore, be a good alternative if someone is willing to spend the extra investment cost and CAPEX to capture more CO_2.

A doubling in cost for either the packing or compressor would still give the same results when it comes to which process that is most cost optimum of standard and vapor recompression but setting a value one the waste heat could affect this. Setting a value of 0.2 NOK/kW would make vapor recompression the cheaper option in NOK/ton CO_2 captured, and a lower cost of the waste heat would still help in reducing the cost gap between the two alternatives.

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Appendices

Appendix A – Project Description

University College
of Southeast Norway

Faculty of Technology, Natural Sciences and Maritime Sciences, Campus Porsgrunn

FMH606 Master's Thesis

Title:

Partial CO_2 capture simulation and cost estimation

USN supervisor: Lars Erik Øi

Co-supervisor(s): Nils Eldrup (USN/Tel-Tek), representative from Tel-Tek

Task description:

The general aim is to develop further models in Aspen HYSYS for calculation, equipment dimensioning, cost estimation and cost optimization of CO_2 capture by atmospheric exhaust gas absorption into an amine solution. A special aim is to evaluate partial CO_2 capture processes.

1. Process description of partial CO_2 capture from one or more industrial sources by absorption into an amine solution.

2. Aspen HYSYS simulations of CO_2 capture alternatives with absorption into an amine solution. Calculation of dependencies of different process alternatives. Energy optimization of different process alternatives.

3. Process equipment dimensioning and equipment cost estimation.

4. Evaluation of the most energy optimum and the most cost optimum process alternatives, preferably using Aspen HYSYS.

Address: Kjølnes ring 56, NO-3918 Porsgrunn, Norway. **Phone:** 35 57 50 00. **Fax:** 35 55 75 47.

Background:

The most promising method for capture of CO_2 from atmospheric exhaust is by the help of amine solutions. Cost estimates for CO_2 removal from natural gas based power plants have been made for Kårstø, Tjeldbergodden, Mongstad and Herøya. Master Projects from 2007 to 2016 at USN (earlier HiT) have included cost estimation in a spreadsheet connected to an Aspen HYSYS simulation. In a Master Thesis (Svolsbru, 2013), Aspen HYSYS simulations of CO_2 capture from a cement plant were performed. In a Master Thesis (Park, 2016), partial capture from a cement plant was evaluated. These projects have used different tools for equipment cost estimation. USN/HiT has collaborated with different companies (Tel-Tek, Statoil, Aker Kværner, Norcem, Yara, Skagerak and Gassnova) which work with plans for CO_2 capture. USN is involved in a large project (Co2stCap) to evaluate partial CO_2 capture from industrial sources, especially from steel, pulp and paper and cement plants.

Student category: PT or EET

Practical arrangements:

The work will mainly be carried out at USN in collaboration with Tel-Tek.

Signatures:

Student (date and signature): 1/2 - 2017 Emil Sundbø

Supervisor (date and signature): 1/2 - 2017 Lars Erik Øi

Appendix B – SIMS Abstract

Simulation and Economic Optimization of Vapour Recompression Configuration for Partial CO_2 capture

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Abstract for SIMS2017 LEØ/Erik/HA/29.3.2017

A standard method for CO_2 capture is by absorption in an amine based solvent like MEA (monoethanol amine) followed by desorption. Such plants are traditionally designed for removal of 85-90 % of the CO_2 from the exhaust gas since this is a reasonable trade-off between acceptable removal efficiency and modest investment cost. The major challenge in its implementation is the high energy demand for CO_2 desorption. However, in many industrial cases, a limited amount of cheap waste heat is available and this makes partial CO_2 capture an interesting option. In case of partial CO_2 capture, it is not obvious whether a high removal efficiency from a part of the exhaust or a low removal efficiency of the total exhaust is the best solution. The aim of this work is to perform simulations of various process configurations including vapour recompression to find the energy optimum and the most cost effective solutions. Especially the focus is to perform a cost-benefit analysis of various configurations to evaluate whether it is cost optimum to treat all the exhaust gas or only a part of it.

Several studies have investigated utilization of waste heat with a standard MEA absorption and desorption process (traditional configuration) but there are few studies which have focused on different process configuration powered by waste heat. A process configuration which has been shown to be very energy efficient is vapour recompression where the regenerated amine from the desorption column is depressurized to a pressure below the desorption pressure, before the liquid is recycled back to the top of the absorber while the gas is compressed and recycled back to the bottom of the desorption column.

Different alternatives were simulated using the process simulation tool Aspen HYSYS version 8.6 using the Kent-Eisenberg vapour/liquid equilibrium model. An exhaust gas from a cement plant was used as an example case. According to available data, the amount of waste heat from the cement plant is up to 25 MW, which is not enough to remove 85-90 % of the total CO_2. This is however an attractive case for partial CO_2 capture. Murphree efficiencies for CO_2 were specified to 0.15 in the absorption column and the stages in the desorption column were specified as equilibrium stages. The number of absorption stages was varied between 5 to 15. The process was simulated with a part of the total exhaust gas (part-flow) from 40 up to 100 % (which is full-flow) of the gas through the absorber column.

The total CO_2 removal efficiency was calculated for all the alternatives. The results showed clearly that the full-flow alternative achieved the highest removal efficiency for both the traditional and for the vapour recompression configuration. Without vapour recompression, the removal efficiency varied between 39 and 41 % where the highest removal efficiency was achieved with 15 absorption stages. With vapour recompression, 45 to 48 % was achieved with 5 and 15 absorption stages, respectively. The solution with the highest removal efficiency with a heat consumption of 25 MW, was regarded as the energy optimum process.

For some of the process alternatives, the process was cost estimated to find the cost optimum alternative. The energy cost for electricity was set to 0.05 Euro/kWh while the 25 MW waste heat was regarded as free. Investment cost was estimated based on equipment cost from literature and detailed installation factors. The interest rate was set to 7.5 % and the calculation period was 25 years. The lowest total cost per ton CO_2 captured was calculated for the standard full-flow process with 5 absorption stages. However, the full-flow process with a vapour recompression configuration and 5 absorption stages had a considerably higher CO_2 removal rate and only a slightly higher total cost per ton CO_2 captured.

Appendix C – Cost Index and Currency Exchange rate

Chemical Engineering Plant Cost Index (CEPCI) [21]

Year	CEPCI
January 2016	556.8
Yearly average 2011	585.7
Yearly average 2010	550.8
January 2002	395.6

Currency exchange rate [22]

Year	Exchange rate
January 2016	8.4 NOK/$

Appendix D – Installation factor

2013-2014

Cost of equipment in Carbon Steel (CS) (kNOK)	Fluid								Solid						
	0-20	20-100	100-500	500-1000	1000-2000	2000-5000	5000-15000	>15000	0-20	20-100	100-500	500-1000	1000-2000	2000-5000	>5000
Equipment	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00
Erection	0,70	0,37	0,20	0,14	0,11	0,09	0,08	0,06	1,55	0,82	0,48	0,34	0,28	0,20	0,17
Piping	2,80	1,51	0,88	0,65	0,51	0,38	0,32	0,23	0,57	0,31	0,17	0,13	0,10	0,08	0,07
Electric	0,81	0,56	0,38	0,32	0,27	0,22	0,20	0,14	1,37	0,86	0,57	0,44	0,37	0,31	0,26
Instrument	2,80	1,51	0,88	0,65	0,51	0,38	0,32	0,23	1,11	0,61	0,36	0,26	0,21	0,14	0,12
Civil work	0,43	0,28	0,20	0,16	0,13	0,11	0,10	0,07	0,99	0,59	0,38	0,29	0,23	0,19	0,16
Steel & concrete	1,41	0,92	0,62	0,50	0,43	0,34	0,31	0,22	1,97	1,22	0,80	0,62	0,52	0,41	0,37
Insulation	0,53	0,27	0,14	0,11	0,09	0,07	0,04	0,03	0,53	0,27	0,14	0,11	0,09	0,07	0,04
Direct Cost	**10,60**	**6,53**	**4,43**	**3,65**	**3,17**	**2,69**	**2,47**	**2,09**	**9,20**	**5,79**	**4,00**	**3,33**	**2,90**	**2,50**	**2,29**
Engineering Process	0,97	0,34	0,19	0,14	0,12	0,10	0,09	0,07	0,97	0,34	0,19	0,14	0,12	0,10	0,09
Engineering Mechanical	0,77	0,19	0,08	0,04	0,03	0,02	0,01	0,01	0,97	0,29	0,13	0,09	0,07	0,04	0,03
Engineering Piping	0,85	0,46	0,27	0,20	0,14	0,11	0,10	0,07	0,17	0,09	0,04	0,03	0,02	0,02	0,02
Engineering Electric	0,82	0,24	0,12	0,09	0,08	0,07	0,04	0,03	0,96	0,32	0,16	0,12	0,10	0,08	0,07
Engineering Instrument	1,46	0,57	0,28	0,20	0,16	0,11	0,10	0,07	0,95	0,28	0,12	0,09	0,07	0,04	0,03
Engineering Civil	0,31	0,09	0,03	0,02	0,02	0,01	0,01	0,01	0,39	0,13	0,07	0,04	0,03	0,02	0,02
Engineering Steel & Concrete	0,46	0,19	0,10	0,08	0,07	0,04	0,04	0,03	0,53	0,22	0,12	0,10	0,09	0,07	0,07
Engineering Insulation	0,21	0,07	0,02	0,01	0,01	0,01	0,01	0,01	0,21	0,07	0,02	0,01	0,01	0,01	0,01
Engineering Cost	**5,83**	**2,11**	**1,09**	**0,78**	**0,63**	**0,48**	**0,41**	**0,30**	**5,14**	**1,73**	**0,86**	**0,63**	**0,50**	**0,39**	**0,34**
Procurement	1,22	0,41	0,16	0,10	0,07	0,03	0,02	0,02	1,22	0,41	0,16	0,10	0,07	0,03	0,02
Project Control	0,29	0,11	0,04	0,03	0,03	0,02	0,02	0,02	0,26	0,09	0,04	0,03	0,02	0,02	0,01
Site Management	0,52	0,33	0,22	0,19	0,16	0,13	0,12	0,09	0,44	0,28	0,20	0,16	0,14	0,12	0,12
Project management	0,70	0,36	0,23	0,19	0,16	0,13	0,12	0,09	0,60	0,31	0,20	0,16	0,13	0,12	0,11
Administration Cost	**2,74**	**1,21**	**0,66**	**0,50**	**0,41**	**0,33**	**0,29**	**0,21**	**2,52**	**1,09**	**0,60**	**0,46**	**0,37**	**0,31**	**0,27**
Commissioning	0,57	0,26	0,13	0,08	0,08	0,04	0,04	0,03	0,49	0,23	0,12	0,09	0,07	0,04	0,03
Total Known Cost	**19,72**	**10,12**	**6,31**	**5,00**	**4,29**	**3,55**	**3,23**	**2,64**	**17,35**	**8,85**	**5,58**	**4,49**	**3,84**	**3,25**	**2,94**
Contingency	3,91	2,02	1,27	1,01	0,87	0,73	0,66	0,48	3,39	1,73	1,11	0,89	0,77	0,66	0,60
Total Cost	**23,63**	**12,13**	**7,57**	**6,02**	**5,16**	**4,28**	**3,89**	**3,11**	**20,73**	**10,59**	**6,69**	**5,38**	**4,60**	**3,90**	**3,54**

Material factors

When using other materials than CS, the factors for equipment and piping must be multiplied with the Material factor.

Material factors:
Stainless Steel (SS316) Welded: 1,75
Stainless Steel (SS316) Machined : 1,30
GRP: 1,00
Exotic: 2,50

Porsgrunn December 2013

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