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Abstract:

The work in this thesis is a continuation of earlier work by students at Telemark University College (TUC) of CO_2 capture simulation and cost estimation in Aspen Hysys.

A Hysys simulation of a CO_2 capture process by absorption in a monoethanol amine (MEA) solution from the flue gas from a 500 MW natural gas power plant has been developed as a verification of earlier simulations at TUC. The major improvements in this work are new calculation methods for make-up water and MEA and simulation of a direct contact cooler (DCC) unit. For cost estimation purposes, calculations of overall heat transfer coefficient and correction factor for heat exchangers have been performed.

On the basis of the base case simulation output, installed cost estimates for equipment have been made. Only equipment related to flue gas cooling and the CO_2 absorption and regeneration process have been included in the simulation and cost estimation scope. Variation in cost changes has been monitored when changing process parameters like minimum approach temperature in the lean/rich heat exchanger, absorber packing height, absorber gas feed temperature. The parametric studies have been performed for CO_2 removal efficiencies of 80, 85 and 90 %. In most of the calculations, one meter of packing was specified with a Murphree efficiency of 0,15. When optimizing feed gas temperature, a temperature dependent efficiency was used.

The base case with an CO_2 removal efficiency of 85 % has been estimated with a specific energy consumption of 3,61 MJ/kg CO_2, and equipment installed cost is estimated to 1400 MNOK. The annual operational utility cost has been found to be 203 MNOK, where 61 % is related to steam consumption in the desorber reboiler. The amine package in Aspen Hysys with Kent Eisenberg was used. The Li-Mather model was checked for comparison with the base case, this resulted in a 1,5 % increase in the annual operational utility cost and 0,8 % increase in the equipment installed cost.

Parametric studies at a CO_2 removal efficiency of 85 % have resulted in optimum minimum approach temperature in the lean/rich heat exchanger between 10-14 K, absorber packing height 15 m, and absorber feed gas temperature approximately 40 ^{0}C. At 90 % efficiency the effect of varied process parameters is greater then at 85 %. Economic parameters like uptime and calculation period also influence on the optimum parameters.

This study shows how significant process parameters are to overall cost of CO_2 capture. Major improvements in cost savings can be made by optimization. Aspen Hysys is a suitable tool for such calculations.

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Preface

This master thesis has been developed during the spring semester of 2010 as part in a master program in Process technology.

Software used in this master thesis is Microsoft Visio 2007, Microsoft Office 2007, the process simulation tool Aspen Hysys 2006 and Adobe Acrobat 9 Pro.

The reader should have knowledge to chemical engineering terms when reading the thesis. Knowledge to Aspen Hysys features like absorber, desorber, adjust, recycle and spreadsheet functions is considered beneficial, but not necessary. Relevant data from the simulations can be found as appendices to the report.

I would like to give a special thanks to my supervisor Lars Erik Øi at Telemark University College for his enthusiasm and good guidance during the work on this thesis.

Porsgrunn, 3. June 2010

Ove Braut Kallevik

1 Introduction

This master thesis is a continuation of previous work performed by students at TUC on simulation and cost estimation of capture of carbon dioxide (CO_2) by post combustion amine absorption.

1.1 Purpose

The purpose is to give insights as to which technical factors that impact the performance and costs with regards to capture of CO_2 post combustion amine absorption. It also wishes to exemplify and illustrate the importance of optimization of process parameters and how this can lead to improved process performance and lower overall costs. Evaluation and improvement to previous studies on simulation and cost estimation performed by students at TUC has been a part of this thesis.

1.2 Background

Currently, combustion of fossil fuels accounts for 60 % of all electricity produced worldwide [1] and 85% of all commercial energy consumption [1]. As the focus on climate change increases, technologies for removing the CO_2 have been the source of many studies. CO_2, accounting for 55 % of the global warming has been given much attention, and technologies are being developed in order to reduce emissions.

For the production of electric energy, three principle categories exist for simultaneous capture of CO_2 from the combustion process for producing power;

- Pre-combustion – conversion of fossil fuel to synthesis gas for further combustion
- Oxy-combustion – combustion of fossil fuel and pure oxygen
- Post-combustion - removal of CO_2 from a conventional exhaust

The two first represent novel technologies and research still remains before they are commercially available. Post combustion is the technology which is considered the most mature for CO_2 capture [3]. It also has the advantage of being retrofitted downstream existing power plants or other industrial sources. However, amine capture has never been implemented on a large scale power plant before. The largest existing capture plant is that of Billington with 0,1 Mt CO_2/yr. Because of the lack experience from previous projects, great uncertainties are assumed related to the scale-up technical performance and cost.

According to Røkke *et al*. [3], the following parameters affect the plant cost for CO_2 post combustion capture by absorption:

- Exhaust gas volume rate
  - This determines dimensioning of process equipment in the gas path – which usually makes up the majority of the equipment cost
- CO_2 content in the flue gas
  - Increasing the CO_2 partial pressure (concentration) lowers energy consumption
- CO_2 removal rate
  - Energy consumption increases with increased removal rate
- Flow rate of amine
  - Allowable CO_2 loading of amine determines amine flow rate and hence size of equipment and utility requirement
- Energy requirement
  - Hot utility – large amount of costly high temperature utility is required in order to reverse chemical reactions between CO_2 and the amine
  - Electricity – in most cases, the flue gas has to be transported through the capture plant, due to the large volume rates the electricity cost is significant

The major challenges and sensitivities regarding an absorption capture process is the nature of absorption of a gaseous component into a liquid. The driving force for the mass-transfer is concentration, or in this case, the partial pressure of the CO_2 to be captured. Flue gas from a natural gas power plant contains as little as 3,5 - 5 mole% CO_2 [3]. This results in small driving forces for absorption, and hence large importance is given the absorbent properties and contact area (absorber size) when removing the CO_2 efficiently.

The advantage of using amines as absorbents are their ability to chemically react with CO_2 to make the absorption go faster, even at low partial pressures of CO_2. The downside is the increased energy required to reverse this reaction. The recovery of the absorbed CO_2 from the solvent is an endothermic reaction and therefore needs the addition of thermal energy. This is the one major operational cost involved in the whole CO_2 capture plant. Another important factor regarding the choice of absorbent is the loading capacity, or cyclic capacity. The loading factor describes how much CO_2 that can be present in the amine solution relative to the amine. The difference between lean and rich loading affects the necessary amount of liquid flow of the amine in the loop, and hence the cost of purchase and operation of the involved equipment. Higher capacity means lower liquid flow rate of amine. The capacity is dependent of the concentration of the amine and how high loading that is achievable based on equilibrium between CO_2 and the absorbent. A type of amine which there has been done extensive research on and frequently used in modeling, is monoethanol amine (MEA). The typical concentration of MEA may be 30 % in an aquatious solution, but the use of corrosion inhibitors may be necessary at these high concentrations [3].

Many studies have been performed on both simulation and cost estimation of CO_2 capture. However, most of these are focused on one design point, not showing the impact of change in process parameters values [4]. This study wishes to investigate the sensitivities of change of process values in order to show the impact on overall cost. Two articles aiming at evaluating technological and economical performance are the work of Abu-Zahra *et al.*[4] and that of Røkke *et al.*[3]. From the first reference, a 600 MW bituminous coal fired power plant with 13,3 mole% CO_2 in the flue gas was simulated in Aspen Plus. The study showed that major cost reduction with regards to optimum MEA concentration, lean loading of the amine and desorber column pressure was possible. For the CO_2 capture plant, it was found that the equipment related to flue gas path contributes to 75 % of total equipment cost. The study also showed that by increasing the MEA concentration of the lean amine from 30 to 40 wt%, the specific energy requirement was reduced from 3,3 to 3,01 MJ/kg CO_2. This change led to a reduction in the cost of the power plants cost of electricity by 5,3 %. The optimum lean amine loading was between 0,25 – 0,33 mole CO_2/mole MEA. The study also pointed out as the process is highly energy demanding, fuel prices may influence the cost of CO_2 removal. It found that doubling the fuel cost would lead to a 23 % increase in the cost per ton CO_2 avoided[1]. The costs per ton of CO_2 removed showed little variation in the range of 80-95 % CO_2 removal efficiency.

The effect of higher CO_2 content in the flue gas in the range 5-20 mole% was studied in Røkke *et al.* [3], the cost per capture CO_2 were found to decrease from 434 to 375 NOK/ton CO_2 in the interval. The same study also performed several studies on cost estimates on CO_2 capture projects from various emissions sources in Norway was performed. They found large variations in the cost estimates, and pointed out some aspects as to why the estimates vary so much:

- Methods and sources for cost estimation
- Variation in result presentation
- Choice of technology
- System boundaries
- Economic calculation assumptions
    - Choice of calculation period
    - Calculation interest rate
    - Currency exchange rates

They also simulated a CO_2 capture process based on several different industrial sources in Norway. The cost per captured CO_2 was found in the range 367 – 865 NOK/ton CO_2. These costs where quite comprehensive and included process equipment, capital costs, taxes and

[1] Cost of CO_2 avoided =(cost of electricity$_{capture}$-cost of electricity$_{reference}$)/(CO_2 emission$_{reference}$-CO_2 emission$_{capture}$)[4]

administrations costs etc. They also pointed out that the major cost driver for installation cost is the absorber, which again is dependent on flue gas flow rate and the CO_2 content. For the capture cost itself, it found that the cost of energy is the most important parameter. The estimates did not however include any possible savings due to integration with nearby process infrastructure.

In the study by Abu-Zahra *et al.* [4], it was concluded that cost of the CO_2 capture is a limiting factor for further full scale build. And in the work of Rao *et al.* [5], a group of selected experts was asked to indicate in which areas R&D resources should be focused in the next years in order to reduce the costs of CO_2 capture. The top priorities were found to be:

- Development of absorbents with lower regeneration heat requirement
- Development of less expensive technologies for CO_2 removal
- Improved heat integration in the capture plant
- Development of power plants with higher efficiency, and hence lower heat rate to the capture plant

The possible reduction of heat consumption in the capture plant was also mentioned by Røkke *et al.*[3]. Here it was claimed that it is plausible to achieve a 30 % reduction in heat requirement in 3-5 years time. This has to be done in conjunction with the proper selection of equipment and materials to withstand corrosion, as this probably involves introducing even more electrolytic amine solutions. As the current cost estimates are in the area of 40-70 €/ton CO_2, the research on this subject should target to get it below 25 €/ton CO_2 [1].

1.3 Objectives

The thesis description and objectives can be found in Appendix 1.

2 Process description and Hysys base case simulation

2.1 Process description

A process flow diagram showing the main components and flows in an absorption plant for removal of CO_2 from a flue gas is shown in Figure 2-1. The figure is made by the author and is based on various sources [7][4].

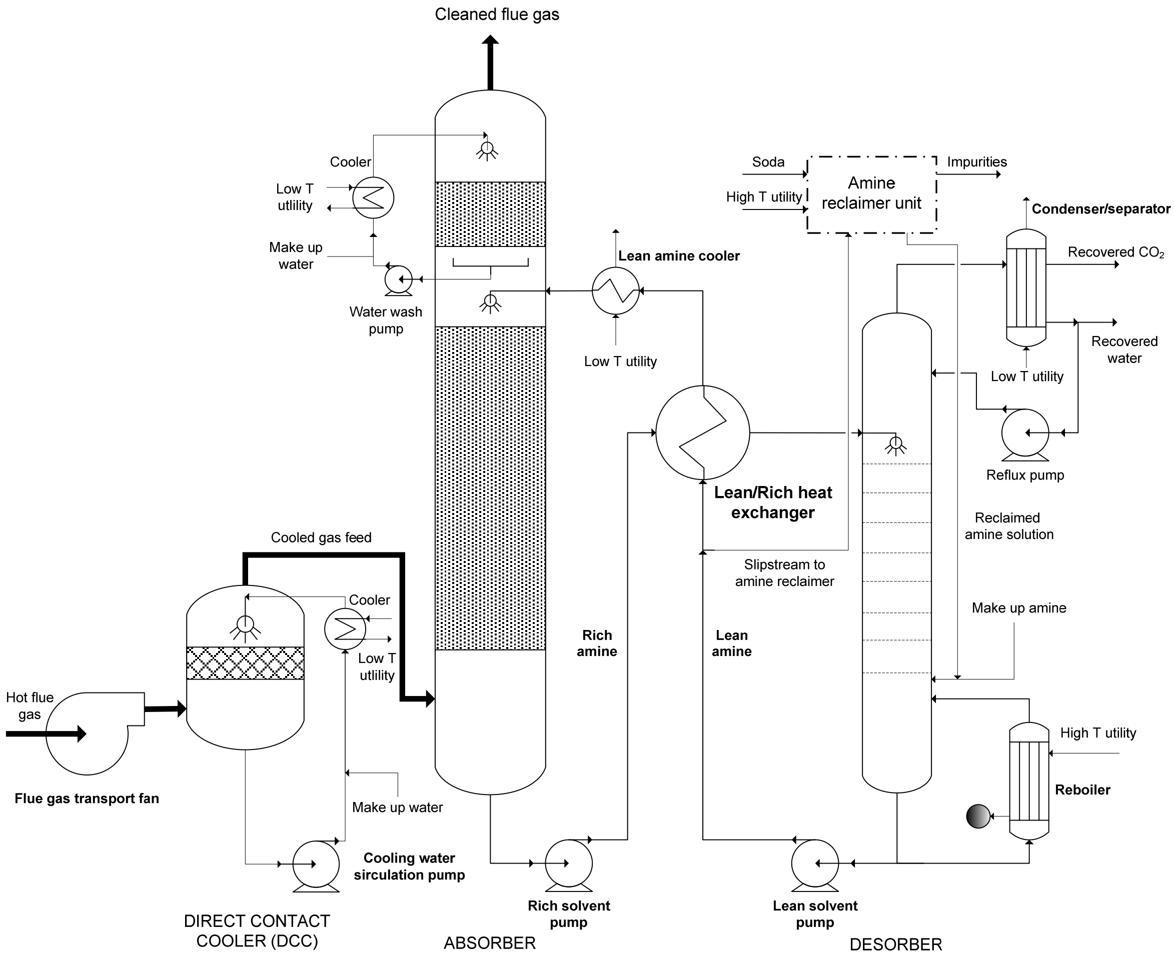

Figure 2-1 General flow diagram of a CO_2 removal process plant

The figure show the gas conditioning part, which involves the transport fan and direct contact cooler which give the flue gas the necessary pressure and temperature prior to the absorption column. The flue gas containing CO_2 is led into contact with the absorber liquid which has the ability to solve the gaseous CO_2 into the liquid and mass-transfer can occur. The storage of the gaseous component in the solvent may be done by either chemical or physical bonding, or a combination. The classification of the solvents depends on their physical function. Amine solutions are classified as chemical solvents.

The reverse process of absorption is called desorption and involves removing the absorbed gas from the absorbent so that the absorber can be re-used for continuous absorption. The following text describes the basic process equipments needed to fulfill this process and their various physical constraints and dependencies.

2.1.1 Flue gas transport fan

The flue gas effluent from a gas power plant is exhausted at about atmospheric pressure and at a temperature in the area of 70-90^{o}C, other industrial sources may have temperatures that differ from these values [3]. In order to provide the driving force necessary to transport it through the direct contact vessel and the downstream absorber column, a fan has to be installed. The pressure drop for a given flow rate of flue gas through a fixed width absorber column is generally dependent on the total height of structured packing in the absorber column. The more packing in the column for a given volume flow, the higher pressure loss has to be overcome and thus increasing energy consumption in the transport fan.

2.1.2 Direct contact cooler

The direct contact cooler (DCC) is a unit comprised from three process equipments; the direct contact vessel, the water circulation pump and circulation water cooler. The flue gas entering the DCC contact vessel will be at a higher temperature than the flue gas source due to the enthalpy increase in the upstream transport fan. With regards to obtaining optimum absorption conditions, the flue gas has to be cooled down. Typical absorber feed gas temperature is typically in the area 40-50^{o}C [3]. In the direct contact vessel, water is distributed over a contact medium which maximizes contact area between the countercurrent flowing water and flue gas. Sensible heat from the flue gas will cause some of the water to evaporate, leading to a temperature reduction in the flue gas. The other mode of energy transport is by transferring sensible and latent heat to the water, which leads to a temperature rise of the water out in the bottom of the contact vessel. This water is cooled in the circulation water cooler by an external cold utility and recycled back into the contact vessel for further cooling.

2.1.3 Absorber column

The flue gas enters the absorption column in the bottom and rises vertically. At the same time the absorber liquid flow counter-current from the top of the column. Inside the column, contact devices are installed to maximize the surface area between the liquid solvent and the flue gas. These devices can, depending on considerations like pressure drop and hydraulic capacity, be trays, random packing or structured packing [7]. The structured packing generally offers the lowest pressure drop. The amine solution also generates heat as it is mixed with

CO_2-rich gas, which will lead to some increase in sensible energy. The temperature increase along the absorber is caused by [7]:

- Heat of solution (condensation, mixing, reaction)
- Heat of solvent when condensing/vaporizing
- Sensible heat transfer between gas and liquid phases
- Temperature losses

This causes the temperature profile to vary along the absorber column height, and because the reaction kinetics between CO_2 and amine vary as a function of the temperature, the absorption equilibrium will vary along the column height.

The amine solution also has to have devices called liquid distributors in order to distribute it over the total surface area of the column and the contact volume of the structured packing. These are important in order to utilize and maximize the contact area between the flue gas and amine solution. Important parameters are necessary pressure drop for maximum liquid distribution and turndown ratio. Inadequate distribution of the absorber liquid over the contact packing volume is referred to as maldistribution, and may drastically reduce packing efficiency. Cases of 2 to 3 times increase of packing height necessary to achieve absorber performance is reported [7].

The flue gas column velocity is limited by a condition known as liquid entrainment. The absorber is designed for the highest velocity possible due to a consequent smaller column diameter requirement and lower column cost. If the velocity gets to high, liquid is entrained in the flue gas flow. This has primarily two negative consequences, the first being the loss of costly amine to the atmosphere which has to be continuously added to the plant in order to make up for these losses. Secondly, amines are reported to have negative impact on organic organisms and are considered a local pollution from the plant. One way of overcoming this problem is by installation of a water wash section downstream the absorber section [3][7]. The water wash section can be integrated at the top of the absorber column, or as a standalone unit. This will recover the entrained MEA in the flue gas by absorbing it into water distributed across the top flow section. Due to the large volume flow rate of flue gas, the absorber tower is the physically largest piece of equipment in the capture plant.

2.1.4 Rich amine pump

The CO_2 loaded amine is collected in the sump of the absorber column. This MEA solution is often referred to as “rich loaded amine”, and is quantified in the terms of moles CO_2 per moles of MEA. The amine has to be transported for further separation of CO_2 from the amine solution so that the amine can be recycled back to the absorber. This transport is done by a pump which is referred to as the “Rich amine pump”. The pump differential pressure may be found from determining:

- Friction loss in piping
- Static height difference between the liquid level in the absorber sump and inlet nozzle in the desorber column
- Pressure loss in lean/rich heat exchanger
- System pressure difference absorber - desorber

The power required is determined from the pressure differences, flow rates and the hydraulic efficiency of the pump [8], the equation is shown in Appendix 2.

2.1.5 Desorber column

The downside to having a high solubility at low partial pressure of CO_2 in the amine solution is a high energy requirement for the reverse process, desorption. In the desorber column, energy is added in the form of a high temperature utility, typically low pressure steam, in order to recover the CO_2 from the circulated amine solution. The hot utility in the stripper are used for three purposes [3]:

- Add sensible heat to the rich amine (this is dependent on the approach temperature in the lean/rich heat exchanger)
- Reverse the absorber reaction in order to remove the CO_2 from the amine. This is an endothermic reaction
- Generation of stripping steam to ensure a driving force for the desorption reaction

Of the three mentioned, the two latter are the two most important.

The primary constituents in the gas phase overhead flow of the desorber are CO_2 and water vapor. The loss of water through this stream can be compensated by both condensation and recovery of the desorber overhead water, or by adding fresh water some other part in the amine loop, or by a combination. The degree of removal of the CO_2 from the desorber feed rich amine solution determines the rest-content of CO_2 in the lean amine solution which is reused for absorption in the absorber. This is referred to as "lean loading", and is also quantified in the terms of moles CO_2 per moles of MEA. The largest consumer of hot utility in the CO_2 capture process is the reboiler connected to the desorber column. In this report, the specific energy consumption per mass CO_2 is connected to hot utility consumption in this reboiler.

2.1.6 Lean amine pump

The lean amine pump transfers the amine solution which is collected at the sump of the desorber column through the lean/rich heat exchanger, the lean amine cooler and finally to the absorber. Like for the rich amine pump, the pumps necessary duty is found from determining:

- Friction loss in piping
- Pressure loss in the lean/rich heat exchangers
- Pressure loss in the lean amine cooler
- Static height difference between liquid level in the desorber sump and inlet nozzle height in the absorber column
- System pressure difference desorber - absorber

As for the rich amine pump, the power required for this pump can be found from the pressure differences, flow rates and the hydraulic efficiency of the pump.

2.1.7 Lean/rich amine heat exchanger

The lean/rich amine heat exchanger (L/R heat exchanger) is a device for recovering energy in the absorber process. As its name implies, its purpose is to transfer sensible heat from the hot lean amine stream to the colder rich amine stream. This will reduce the energy required in the reboiler duty desorber column. Typically, the degree of recovery of energy in this exchanger is a trade-off between operating expenditure (OPEX) in the form of hot utility consumption and capital expenditure (CAPEX) in the size of the lean/rich heat exchanger and reboiler. The degree of energy recovery is quantified by the term minimum temperature approach (ΔT_{min}) in the L/R heat exchanger. The ΔT_{min} is defined as the smallest temperature difference between either the hot inlet stream and the cold outlet stream, or the hot outlet stream and cold inlet stream. In this case, the definition of ΔT_{min} is shown in Equation 2-1.

$$\Delta T_{\min} = T_{Lean\_MEA\_Outlet} - T_{Rich\_MEA\_Inlet} \qquad \textit{Equation 2-1}$$

ΔT_{min} is a critical parameter and is used to illustrate the trade-off between of the degree of energy recovery in any heat exchanger and its size. A heat exchanger with a high energy recovery has a low ΔT_{min}, but will require a larger surface area. Conversely, a higher ΔT_{min} will lead to a lower degree of energy recovery, but would require a smaller surface area [9].

2.1.8 Lean amine cooler

The lean amine cooler may offer additional cooling to the lean amine downstream the L/R heat exchanger in order to achieve the required lean amine temperature upstream the absorber column. In this heat exchanger an external utility like cooling water or some other cold utility may be applied.

2.1.9 MEA reclaimer

Due to the high temperature conditions in the reboiler and reactions in the absorber with contaminants in the flue gas, impurities in the amine solution build up over time. These will reduce the effective amine concentration and consequently solution performance, in addition excessive fouling may occur. This is handled by a reclaimer unit which by using a hot utility boils off the amine and water, while the impurity products mainly remains in the boiler bottoms and is withdrawn as waste product. The waste products consist typically of higher molecular weight organic degradation products, inorganic salts and heat stable salts (HSS). MEA consumption has been experimentally found to be in the range of 1,4 -2,0 kg MEA/ton CO_2 from a post combustion from a coal power plant [10]. The loss of MEA has to be added to the amine loop in order to maintain the required absorbent solution performance.

2.1.10 Water condenser and separator

Overhead products from the desorber column consist primarily of water and recovered CO_2. In order to meet compression specifications for further CO_2 transport and to recover water for the amine circulation solution, the water is removed from the overhead vapor. This may be done by cooling the overhead stream and then separation of the liquid water phase from the gas phase. The recovered water may be recycled back into the amine solution flow loop, to make up losses in the absorber and desorber column. The CO_2 rich gas phase is routed to downstream conditioning and transport before storage. To avoid freezing and corrosion, the water content should be low.

2.2 Simulation of base case in Aspen Hysys

The base case simulation in Aspen Hysys has been developed as a verification and evaluation of an earlier CO_2 removal plant simulated in the master thesis of Blaker [11]. In the following subchapters, specifications and assumptions for the various process equipments necessary to simulate the base case CO_2 capture process is described. In tables Hysys output is denoted accordingly.

2.2.1 Scope of simulation

In Figure 2-2 the base case Hysys simulation process flow sheet (PFD) is shown.

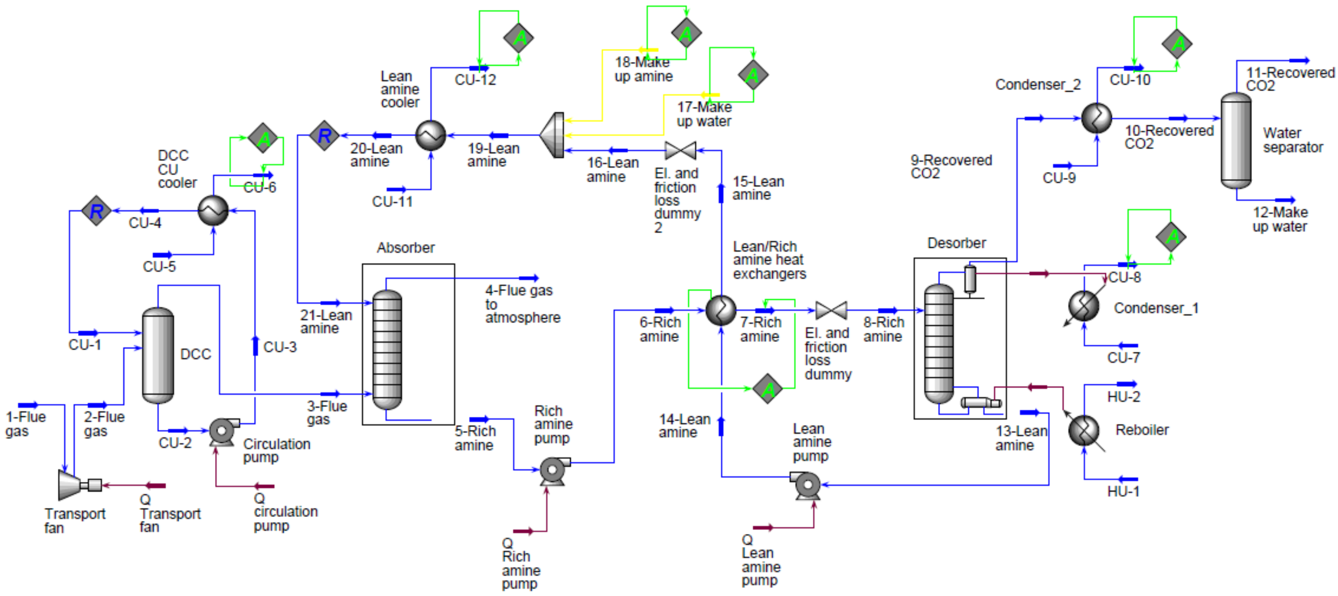

Figure 2-2 PFD of the CO_2 capture process simulated in Aspen Hysys

All streams are notated with a running number followed main function/description. Process equipment has been given a descriptive name. A larger version of the PFD in Figure 2-2 can be found in Appendix 16.

In the simulation, not all process equipment that is necessary in a full-scale plant is included. The simulation scope includes the following:

- Flue gas transport fan
- Flue gas direct contact cooler unit (DCC)
- Absorption column
- Rich amine transport pump
- Desorber column
- Lean amine transport pump
- Lean/rich heat exchanger
- Lean amine cooler
- Water condenser and separator

The equipment not included in this simulation, but which may be critical for the operation and performance of such a plant is:

- Water wash section
- MEA reclaimer
- Equipment for conditioning of make-up water and amine

The water wash section will reduce MEA losses to atmosphere with the purified flue gas, but since the water wash has not been simulated, there will be losses that have to be accounted for by a make-up stream. As there is a net loss of water in the overall process, make-up for water is also necessary.

The MEA reclaimer unit has not been simulated, so the amine solution is assumed not to experience any form of thermal or chemical degradation. The simulation will consequently not consider energy requirement for removing impurities as a function of amine solution flow rate and the necessary make-up rate of MEA due to degradation.

2.2.2 Equilibrium model

When simulating a flow sheet involving absorption of a sour gas into an electrolytic solution, additional software to describe their interaction more precisely is needed. The correct simulation of mass transfer rates between the sour gas and amine solution is important with regards to obtain realistic absorption and desorption performance, and ultimately utility demands and process characteristics.

Within the simulation program, the equilibrium compositions have to be calculated between the amine solution and the flue gas for each stage in the absorber. The amine package contains models originally developed for an amine plant simulator AMSIM, which have been implemented to Aspen Hysys. The model is restricted to 4 acid gases, CO_2, H_2S, COS and CS_2. In this simulation, only CO_2 is simulated and any selectivity for any of the other sour gas over CO_2 to the amine solution is not considered. The CO_2 equilibrium solubility and kinetic parameters when in contact with the amine solution is the main purpose of the amine package. In addition, the reaction between the CO_2 and the amine solution is exothermic, causing heat effects in the absorber. Correlations are in the amine package made so that the heats of solution are set up as a function of composition and the chosen amine solution [14]. However, the initial data which are based on empirical data from several sources have limitations for the applicability of the amine package. When using MEA, the concentration can only be in the range 0-30 wt%, partial pressure in the range 0-20 bar and temperature in the range 25-126 ^{0}C. The data also are not correlated for amine loading above 1 mole CO_2/mole amine. The simulations made in this report are within these ranges.

The amine package has its own efficiency model for simulation of columns which is based on pressure, temperature, phase compositions, flow rates, physical properties, kinetic and mass transfer parameters and geometrical design. The amine stage efficiency has not been used in this thesis, but instead assumed constant Murphree stage efficiencies based on work of Øi [13] have been applied.

The amine package uses the following methods for calculation of vapor- liquid equilibrium (VLE):

- Liquid phase: Kent-Eisenberg or Li-Mather
- Vapor phase: Peng-Robinson
- Enthalpy and entropy: Curve fitting

The Kent-Eisenberg amine fluid package has been chosen for the base case simulation.

2.2.3 Stream specifications

A collection of the significant streams between the CO_2 capture plant and its boundary limits and to the absorber column are shown in Table 2-1.

Table 2-1 Process stream data

Stream	1-Flue gas feed	3-Flue gas	4-Flue gas to atmosphere	11-Recovered CO_2	12-Make up water	21-Lean amine
CO_2 [wt%]	0,059	0,059	0,009	0,984	0,002	0,055
MEA [wt%]	0	0	0,001	0,000	0,004	0,290
H_2O [wt%]	0,043	0,041	0,064	0,016	0,994	0,655
N_2 [wt%]	0,898	0,900	0,926	0,000	0,000	0
P [kPa a]	101	121	106	200	200	101
T [^oC]	100	40,7	48,6	40,0	40	40
Flow [t/h]	3073	3065	2979	155,6	33,6	3600

The flue gas feed and lean amine has been obtained from an earlier simulation in order to verify the simulation of Blaker [11]. The feed gas is specified with a CO_2 content of 3,73 mole%, and should be representative to the flue gas composition and flow rate from a 500 MW natural gas power plant. In real life CO, O_2, NO_x and SO_x will be present in addition to water, N_2 and CO_2 [3], these have however not been included in the simulation. Stream 4, 11 and 12 are results from the convergence of the flow sheet with the required CO_2 removal performance. The lean amine is specified with 29 wt% MEA and the flow rate is adjusted to achieve the base case CO_2 removal grade.

The process requires both cold and hot utilities for external heating and cooling. Also make-up streams compensating for water and MEA losses have to be considered in order for effective convergence of the simulation. These stream compositions are shown in Table 2-2.

Table 2-2 Utility stream specifications

Stream	Cold utility (CU)	Hot utility (HU)	18-Make up amine	17-Make up water
MEA [wt%]	0	0	1	0
H_2O [wt%]	1	1	0	1
P [kPa a]	101	500	301	301
T [^oC]	15	160	15	15
Flow [t/h]	NA	NA	2,38	102

The hot utility has been chosen as low pressure steam at a slightly superheated condition. At a pressure of 500 kPa a, the saturation temperature is 151,3^{o}C. The hot utility source could be available from either nearby infrastructure or from the waste-heat recovery steam generator in the adjacent power plant.

For the cold utility, water at constant initial temperature of 15^{o}C has been chosen. This temperature may be taken from a local fresh-water source or from a sea water source. Possible governmental regulations limiting cold utility outlet temperature is not discussed in this thesis, but there may be limitations to heat flux emissions for full size plant. An adjust function has been applied to all cold utility streams in order to achieve a $\Delta T=T_{CU\ out} - T_{CU\ in}=10^oC$ by adjusting the flow rate of the cold utility. With the selected inlet temperature this implies that all cold utilities are emitted from the plants battery limits at a temperature of 25^{o}C. The assumed cold utility temperatures affect the driving forces and hence the design of the heat exchangers.

The make up streams are simulated as pure component streams. For both water and amine, mass balance spreadsheets and adjust functions have been made to correct for net losses of the respective components for the whole systems boundary limits. The spreadsheet mass balance calculation for the water make-up giving the target value for the adjust function is shown in Table 2-3. Negative notation symbolizes output from process boundaries, while positive notation symbolizes input to process boundaries.

Table 2-3 Water losses spreadsheet and make up calculation

Flue gas feed *(Hysys)* [t/h]	+125
Absorber gas effluent *(Hysys)* [t/h]	-191
Desorber gas effluent *(Hysys)* [t/h]	-36
Sum water losses [t/h]	-227
Water make up [t/h]	+102

The “Water make up” value is the target value for the adjust-function which corrects the water make up stream accordingly.

The spreadsheet mass balance calculation for the amine make-up adjust function is showed in Table 2-4.

Table 2-4 Amine losses spreadsheet and make up calculation

Absorber gas effluent *(Hysys)* [kg/h]	- 2260
Desorber gas effluent *(Hysys)* [kg/h]	- 120
Sum MEA losses [kg/h]	- 2380
MEA make up [kg/h]	+2380

Like the water make up calculation, the “MEA make up” mass flow value is the target value for the adjust function which corrects the MEA make up stream flow rate.

2.2.4 Flue gas transport fan

Specifications for the flue gas transport fan are shown in Table 2-5.

Table 2-5 Flue gas transport data

Inlet temperature [^oC]	100
Inlet pressure [kPa a]	101
Outlet pressure [kPa a]	121
Adiabatic efficiency [%]	80

In order to be able to achieve a low enough temperature upstream the absorber column, the transport fan has been installed upstream the DCC. The outlet pressure has been assumed sufficient in order to transport the flue gas through the DCC and absorption column, and is equivalent to a pressure increase of 20 kPa. The adiabatic efficiency is selected in the higher end of typical values for fans [12].

2.2.5 Direct contact cooler unit

The simulation of the direct contact cooler (DCC) is comprised of a flash separator and a cooling water circulating loop from the bottom of the separator through a cooler before it is recycled back into the flash separator. An overview of the input data for the DCC unit can be seen in Table 2-6.

Table 2-6 Direct contact cooler (DCC) data

Flue gas inlet rate *(Hysys)* [Am^3/s]	833,2
Flue gas inlet temperature *(Hysys)* [^oC]	123,7
Flue gas inlet pressure [kPa a]	121
Gas path pressure loss DCC unit [kPa]	0
Circulation water inlet DCC temperature [^oC]	30
Water flow rate upstream DCC vessel [t/h]	6500
Circulation pump inlet pressure [kPa a]	121
Circulation pump outlet pressure [kPa a]	301
Adiabatic efficiency circulation pump [%]	75
Pressure loss DCC CU cooler tube side [kPa]	179
CU flow rate *(Hysys)* [m^3/h]	6776
Pressure loss shell side (CU) [kPa]	NA

The DCC vessel is simulated as a flash separator i.e. an enthalpy balance is performed for the entering liquid water and flue gas, at its respective temperatures. Hysys calculates changes in sensible heats due to energy transfer due to both latent and sensible heat for the two inlet streams, and then splits the liquid and gas phase. The DCC vessel has been simulated with no pressure loss.

The water circulation loop has a recycle function installed in order to converge the feedback of the cooled circulation water back into the DCC vessel. For the base case, the heated circulation water from the DCC vessel experiences a flow increase of ~8 t/h over the DCC vessel. This is primarily due to water condensation from the flue gas as it is cooled down. The recycle function ignores this increase so that the water entering the DCC vessel is constant at 6500 t/h. A practical application of this phenomenon would be excess water available for make up in the amine flow loop. The CU flow rate required in the DCC CU heat exchanger has been adjusted by Hysys in order to meet the CU temperature difference of 10 K.

2.2.6 Absorber column

An overview of the input data for the absorber column can be seen in Table 2-7.

Table 2-7 Absorber data

Flue gas flow rate *(Hysys)* [Am^3/s]	655,8
Flue gas temperature *(Hysys)* [^oC]	40,7
Inlet flue gas pressure [kPa a]	121
Packing height [m]	16
Packing efficiency [m^{-1}]	0,15
Lean amine inlet temperature [^oC]	40
Lean amine flow rate [t/h]	3600
Lean amine loading *(Hysys)* [mole CO_2/mole MEA]	0,263
Pressure loss [kPa/m packing]	0,94
Outlet flue gas pressure [kPa a]	106

The DCC manages to bring the flue gas temperature upstream the absorber column from 123,7^oC down to 40,7^oC with the base case settings. The effect is also seen in the actual flow rate which is reduced from 833,2 to 655,8 Am^3/s over the DCC unit.

As an adaptation to realistic performance of the equilibrium between CO_2 and the amine solution, Hysys has the possibility for the user to specify the Murphree efficiency for each equilibrium stage in the absorber. The Murphree stage efficiency gives the possible change in stage gas phase composition related to the theoretical composition change [7]. The stage efficiency is considered constant at 15 % per meter of packing for all stages in the absorber and has been adopted from the master thesis of Blaker [11]. The efficiency is considered a good average approximation for the overall conditions in the absorber [13]. In practice the efficiency will vary according to temperature and concentration gradients along the column, among other parameters. The number of absorber stages has been assumed, while the lean amine flow rate has been adjusted in order to meet the base case CO_2 removal efficiency requirement of 85 %.

The pressure loss has been simulated over the absorber as 15 kPa. The outlet pressure have been specified to 106 kPa a, in order to allow for a pressure drop over a downstream water wash section of 5 kPa.

2.2.7 Rich amine pump

Base case specifications and assumptions for the rich amine pump are shown in Table 2-8.

Table 2-8 Rich amine pump data

Flow rate rich amine *(Hysys)* [t/h]	3686
Rich amine inlet temperature *(Hysys)* [^oC]	44
Inlet pressure [kPa a]	121
Outlet pressure [kPa a]	750
Pump differensial pressure [kPa]	629
Adiabatic efficiency [%]	75

The flow rate of rich amine converts to a volume flow of 3498 m^3/h. The pump differential pressure has been set to 629 kPa, with an outlet pressure of 750 kPa a. This is considered a conservative assumption in order to overcome friction and separation losses in piping, pressure loss in downstream lean/rich heat exchangers, static height difference and to overcome the slightly elevated system pressure in the desorber of 200 kPa a. The adiabatic efficiency of the pump has been assumed to be 75 %, a mid-range value for centrifugal pumps [12].

2.2.8 Desorber column

Base case specifications and assumptions for the rich amine pump are shown in Table 2-9.

Table 2-9 Desorber data

Rich amine flow rate *(Hysys)* [t/h]	3686
Rich amine inlet temperature *(Hysys)* [^oC]	106,6
Rich amine inlet pressure [kPa a]	250
Rich amine loading *(Hysys)* [mole CO_2/mole MEA]	0,469
Packing height [m]	12
Packing efficiency [m^{-1}]	0,5
Desorber operation pressure [kPa a]	200
Pressure loss desorber column [kPa/m packing]	0
Reflux ratio [-]	0,4
Reboiler temperature [^oC]	120

The desorber has been simulated with a full reflux condenser, which implies that the overhead condenser only has to provide enough cooling duty in order to condense the reflux stream. The overhead balance then exits the column as a vapor phase. The main reason for this selection is to reduce condenser size and cost. The rich amine is feed into the top stage of the desorber column.

The desorber stage efficiency has been assumed to be 50 % per meter of packing, and assumed constant for the whole column. The stage efficiency has been adopted from the master thesis of Blaker [11]. The pressure in the desorber has been set to 200 kPa, and without pressure loss. The desorber unit is specified with a fixed reflux ratio[2] of 0,4 and a reboiler temperature of 120°C. The reboiler temperature is also adopted from the master thesis of Blaker as previous studies have shown that a high temperature is beneficial to low energy requirement in the reboiler [3].

2.2.9 Lean amine pump

In Table 2-10 specifications for the lean amine pump can be found.

Table 2-10 Lean amine pump data

Flow rate lean amine *(Hysys)* [t/h]	3497
Lean amine inlet temperature [°C]	120
Inlet pressure [kPa a]	200
Outlet pressure [kPa a]	700
Pump differensial pressure [kPa]	500
Adiabatic efficiency [%]	75

The flow rate of rich amine converts to a volume flow of 3613 m^3/h. The pump differential pressure has been set to 500 kPa, with an outlet pressure of 700 kPa. The system resistance and adiabatic efficiency is estimated in the same way as for the rich amine pump in chapter 2.2.7. The lean amine is allocated on the tube side in the downstream lean/rich heat exchanger. The pressure loss is assumed to be lower than compared to the shell side, but as the lean amine pump has to lift the liquid into the higher absorber column, the two amine pumps should have somewhat similar outlet pressures.

[2] Reflux ratio is the molar liquid flow returned to the top stage divided by the sum of the vapor leaving the top stage of the desorber.

2.2.10 Lean/rich amine heat exchanger

Specifications for the process/process heat exchanger for lean- and rich amine are shown in Table 2-11.

Table 2-11 Lean/rich amine heat exchanger data

Flow rate lean amine *(Hysys)* [t/h]	3497
Lean amine inlet temperature *(Hysys)* [^oC]	120,2
Lean amine outlet temperature *(Hysys)* [^oC]	54,3
Lean amine inlet pressure [kPa a]	700
Lean amine outlet pressure [kPa a]	600
Flow rate rich amine *(Hysys)* [t/h]	3686
Rich amine inlet temperature *(Hysys)* [^oC]	44,2
Rich amine outlet temperature [^oC]	109,5
Rich amine inlet pressure [kPa a]	750
Rich amine outlet pressure [kPa a]	450
Minimum approach temperature, ΔT_{min} [^oC]	10
Pressure loss tubeside [kPa]	100
Pressure loss shell side [kPa]	300

The purpose of this heat exchanger is to recover heat from the hot lean amine to the colder rich amine. With ΔT_{min} as the quantitative measure of energy recovery, the heat exchanger was for the base case designed with a temperature on the rich amine outlet giving a ΔT_{min}=10^oC. This was achieved by implementing an adjust function, where the adjusted temperature was the temperature of stream 26. In this case ΔT_{min} is the temperature difference between stream 15 and 6. The pressure drops are assumed values.

2.2.11 Lean amine cooler

The lean amine cooler data can be seen in Table 2-12.

Table 2-12 Lean amine cooler data

Lean amine inlet temperature *(Hysys)* [^oC]	53
Lean amine outlet temperature [^oC]	40,0
Lean amine inlet pressure [kPa a]	150,0
Lean amine outlet pressure [kPa a]	101,0
CU flow rate *(Hysys)* [m^3/h]	3889
Pressure loss tubeside [kPa]	49
Pressure loss shell side [kPa]	NA

The lean amine is cooled further down to the specified 40^{o}C on the cooler outlet, which is base case setting. The inlet temperature of 53^{o}C is somewhat lower than the lean amine exit temperature from the lean/rich heat exchanger. This is due to the introduction of make-up streams of water and MEA. These streams are assumed to hold an ambient temperature of 15^{o}C. The respective flow rates are indicated in Table 2-3 and Table 2-4. The corresponding CU demand is by Hysys calculated to 3889 m^3/h.

2.2.12 Water condenser

As mentioned in the process description, the overhead vapor flow from the desorber column has to be stripped for water content as downstream compression and transport of CO_2 requires very low water content. The water condenser specifications and data, named Condenser_2 in the flow sheet, can be seen in Table 2-13.

Table 2-13 Water condenser data

Overhead flow rate *(Hysys)* [t/h]	189,2
Overhead inlet temperature *(Hysys)* [^oC]	91,0
Overhead inlet pressure [kPa a]	200,0
Outlet temperature lean amine [^oC]	40,0
Pressure loss condenser [kPa]	0
Cold utility inlet temperature [^oC]	15
Cold utility outlet temperature [^oC]	25
CU flow rate *(Hysys)* [m^3/h]	3889

The vapor flow is equal to 22,8 m^3/s, and contains 18,9 wt% water. The condenser is assumed to be without pressure loss. The CU flow rate is by Hysys calculated to 3889 m^3/h, this in order to meet the CU temperature difference criteria of 10 K.

2.2.13 Water separator

The water separator specifications are shown in Table 2-14.

Table 2-14 Water separator data

Inlet rate gas *(Hysys)* [Am^3/s]	13
Flue gas inlet temperature [^oC]	40
Inlet pressure [kPa a]	200
Total pressure loss water separator [kPa]	0

The condensation of water vapor and the lower temperature of the CO_2 stream have reduced the volume flow from 22,3 to 13 Am^3/s. Furthermore, the water separator is simulated as a flash separator which means that the two-phase feed is split perfect. The unit is assumed without any pressure drop. The recovered liquid phase consists primarily of water which in theory is available as make-up water in the capture plant.

3 Dimensioning and selection of equipment

After the base case has been established, temperatures, flow rates and heat and power duties are identified. This lay the foundation for estimating physical size and further on the basis for cost estimation of the various process equipment and external utility requirements. The external utilities in this context are assumed to be steam, cooling water and electricity. Output values from Hysys are denoted accordingly in the following tables, while the remainder are assumed or calculated values. The tables only show some of all the calculations and assumptions which has been made for the dimensioning. For a complete documentation the reader is referred to in Appendix 6 - Appendix 14.

3.1 Flue gas transport fan

Hysys output data for the flue gas transport fan is shown in Table 3-1.

Table 3-1 Flue gas transport fan dimensioning

Outlet temperature *(Hysys)* [^{o}C]	123,7
Duty *(Hysys)* [MW]	21,6

The enthalpy increase of the flue gas due to the pressure increase also results in increased temperature. The adiabatic efficiency of the fan determines how much of this temperature increase will be for a given pressure duty. For the specified adiabatic efficiency and pressure increase, the temperature of the flue gas increase 23,7^{o}C through the fan. Estimation data for fans capable of delivering the specified pressure increase at the high flow rates experienced here was not available. Instead, the installation cost found in the master project by Madsen *et al.*[15] has been used.

3.2 Direct contact cooler unit

System boundary for this process includes:

- DCC contact vessel
- Water circulation pump
- DCC CU cooler

The DCC simulated performance and further dimensioning is shown in Table 3-2. It is pointed out that this table only shows the most significant data. For the complete dimensioning calculation data the reader is referred to Appendix 6.

Table 3-2 DCC unit dimensioning

Flue gas outlet temperature *(Hysys)* [^oC]	40,7
Duty DCC circulation pump *(Hysys)* [kW]	437,5
K-factor DCC vessel [m/s]	0,15
Vertical flow velocity DCC vessel [m/s]	4,7
Vessel diameter [m]	15,1
Vessel total height [m]	15,1
Packing height of contact medium DCC vessel [m]	3
Duty DCC CU cooler *(Hysys)* [MW]	81,1
LMDT uncorrected DCC CU cooler [^oC]	15,4
U*F*A factor DCC CU cooler [kW/K]	5264
Assumed flow velocity tube side DCC CU cooler [m/s]	1
Assumed flow velocity shell side DCC CU cooler [m/s]	1
Calculated DCC CU cooler U [W/m^2K]	1052
Assumed constant DCC CU cooler U*F [W/m^2K]	960
DCC CU cooler area [m^2]	5483
Number of heat exchanger units	1
Pressure loss tubeside [kPa]	9,7
Pressure loss shell side [kPa]	28
Estimated shell diameter [m]	2,2
Estimated shell length [m]	13

From the table it can be seen that the DCC unit reduce the temperature of the flue gas down to 40,7^oC. The DCC vessel is dimensioned as a scrubber containing a 3 meter bed of contact medium in order to increase heat transfer between cooling water and flue gas. The vessel is dimensioned using Souder-Brown equation, with a K-value of 0,15. The value is selected based on considerations on the trade-off of between pressure loss in the vessel, i.e. the flow velocity in the vessel, and the amount of possible liquid carry-over with the flue gas. Liquid carry-over is assumed not to represent any problem in the absorber because free liquid probably will settle in the sump absorber and will not cause any problems downstream anyway. The selected K-value results in a vertical velocity of 4,7 m/s, which gives a vessel diameter of 15,1 meter. The assumed L/D ratio gives a vessel height of 15,1 m. The physical size of the vessel shell plus the packing volume makes up the basis for estimating the cost of the vessel.

The circulation pump has a duty requirement of 437,5 kW, which is the basis for cost estimation of this unit.

For the DCC CU cooler, a calculation scheme has been applied in order to evaluate the overall heat transfer coefficient U and the LMDT correction factor F for the given temperature differences and flow. Physical properties for the process liquids are applied together with assumptions regarding physical sizes and configuration of the heat exchanger. The type of heat exchanger estimated is a shell and tube with one shell pass and two tube passes [9]. For later optimization, parameters will change and consequently the performance of the heat exchanger. Therefore, a constant value of the product of the overall heat transfer coefficient and the LMDT correction factor F has been used for cost estimation purposes, and is referred to as U*F. The required heat exchanger area is then calculated from the heat exchanger equation found in Appendix 2. The calculated area is used as basis for cost estimation of the heat exchanger. The calculated area for the DCC CU cooler is found to be 5483 m^2, correlating to an expected physical size of 2,2 m diameter and a length of 13 m. The calculated pressure loss is lower than estimated pressure loss of the heat exchanger of 179 kPa, which indicates that the simulated pump duty should be dimensioned large enough.

3.3 Absorber column

The performance and dimensioning of the absorber column can be seen in Table 3-3.

Table 3-3 Absorber column dimensioning

Total CO_2 removal efficiency *(Hysys)* [%]	85,2
Rich amine flow rate *(Hysys)* [t/h]	3686
Rich amine temperature *(Hysys)* [^oC]	44
Flue gas outlet temperature *(Hysys)* [^oC]	48,6
Vertical flow velocity [m/s]	3,6
Absorber diameter [m]	15,2
Column total height [m]	50

The overall CO_2 removal rate in the absorber is 85,2 %, which was the set point when initially adjusting the lean amine flow rate under the establishment of the base case. From the streams exiting the absorber column, a temperature increase can be observed for both the flue gas and the rich amine. This is due to exothermic reactions between CO_2 and MEA.

The design parameter vertical gas velocity through the packed column is set to 3,6 m/s, this according to 75 % of the flooding velocity of a structured packing type named Mellapak 250X [3]. The gas velocity at the actual gas flow rate results in an absorber column diameter of 15,2 m. The total height of the column is assumed to be 50 m, this to accommodate the structured packing for CO_2 capture, bulk separation section in the bottom, a water wash section in the top section and liquid distribution equipment for absorber and water wash sections. The absorber vessel size together with the packing volume is the basis for cost estimation of the absorber unit. Costs of the water wash section, liquid distributors and various mechanical supporting is assumed as a percentage addition to the vessel and packing installed cost. The full dimensioning and cost calculation is shown in Appendix 7.

3.4 Amine pumps

The power requirement of the two amine pumps makes up the basis for cost estimation for the units. For complete dimensioning of the pumps, the reader is referred to Appendix 10.

3.4.1 Rich amine pump

Hysys calculations for the rich amine pump are shown in Table 3-4.

Table 3-4 Rich amine pump dimensioning

Rich amine outlet temperature *(Hysys)* [^{o}C]	44,2
Power consumption *(Hysys)* [kW]	815

From the table the outlet temperature of the rich amine is shown, and it can be seen that there is a slight temperature increase due to dissipation of energy applied from the pump. This is considered irrelevant in this context and has no practical consequence. The power consumption of the pump is by Hysys calculated to 815 kW, which is the basis for the cost estimation.

3.4.2 Lean amine pump

Hysys calculations for the lean amine pump are shown in Table 3-5.

Table 3-5 Lean amine pump dimensioning

Lean amine outlet temperature [^{o}C]	120,2
Power consumption [kW]	669

Like the rich amine pump, it is also experienced a slight temperature increase of the fluid over the pump. The power consumption is calculated to 669 kW, which is slightly lower than the rich amine pump. This is caused by a somewhat lower lifting height and lower flow rate of this pump.

3.5 Desorber column

The performance and design parameters for the desorber column are shown in Table 3-6. The complete dimensioning calculations can be seen in Appendix 11.

Table 3-6 Desorber column dimensioning

CO_2 removed from rich amine *(Hysys)* [t/h]	153,2
Lean amine flow rate *(Hysys)* [t/h]	3497
Lean amine temperature *(Hysys)* [^oC]	120
Lean amine loading *(Hysys)* [mole CO_2/mole MEA]	0,265
Column overhead temperature *(Hysys)* [^oC]	91
Vertical flow velocity [m/s]	1,08
Column diameter [m]	5,2
Column total height [m]	30
Reboiler duty *(Hysys)* [MW]	154,2
Condenser duty *(Hysys)* [MW]	26,3
LMDT reboiler [K]	35,8
Reboiler assumed U [W/m^2K]	2500
Reboiler area [m^2]	1723
LMDT condenser [K]	71,1
Condenser assumed U [W/m^2K]	2000
Condenser area [m^2]	185

The mass flow of removed CO_2 from the rich amine in the desorber corresponds to the specified 85,2 % removal rate of CO_2 from the flue gas stream. The lean amine from the desorber has a CO_2 loading of 0,265. The calculation of vertical gas velocity in the column is adapted from the work of Blaker in his master thesis [11]. This results in a column diameter of 5,2 m. The column total height is assumed to be 30 m, including 12 meter of structured packing with a Murphree stage efficiency of 50 % per meter. The vessel dimensions in addition to the cost of the packing volume in the column are the basis for cost estimation.

The reboiler duty is calculated by Hysys to 154,2 MW. The calculated LMDT is found to be 35,8 K. The overall heat transfer coefficient U is found from literature and is assumed to be constant at 2500 W/m^2K [12]. By using the heat exchanger equation the required heat exchanger area is calculated to 1723 m^2.

The condenser duty is calculated by Hysys to 26,3 MW, and the LMDT is found to be 26,3 K. The overall heat transfer coefficient U for the condenser is found in literature and is assumed to be 2000 W/m^2K [12], the heat exchanger equation then gives a required heat exchanger area of 185 m^2. The two respective heat exchanger areas are the basis for their cost estimates.

3.6 Heat exchangers

The performance and dimensioning data of the lean/rich heat exchangers, lean amine cooler and water condenser are collected in this chapter. Like the DCC CU cooler, a calculation scheme has been applied, and all heat exchangers are assumed to be shell and tube heat exchangers with 1 shell pass and 2 tube passes. For all the heat exchangers, tube- and shell side flow velocities are assumed to be 1 m/s, numbers recommended in literature [9][16].

3.6.1 Lean/rich heat exchanger

The calculated duty from Hysys and the major dimensioning parameters are shown in Table 3-7.

Table 3-7 Lean/rich heat exchanger dimensioning

Duty *(Hysys)* [MW]	247
LMDT uncorrected [^oC]	10,3
U*F*A factor [kW/K]	23861
Assumed flow velocity tube side [m/s]	1
Assumed flow velocity shell side [m/s]	1
Calculated U [W/m^2K]	944
Assumed constant U*F [W/ m^2K]	750
Required total heat exchanger area [m^2]	31814
Number of heat exchanger units	6
Specific area [m^2/shell]	5302
Pressure loss tubeside [kPa]	94
Pressure loss shellside [kPa]	265
Estimated shell diameter [m]	2,1
Estimated shell length [m]	12,8

The low ΔT_{min} specification of the lean/rich heat exchanger means that the temperature driving force is low, and excessive heat transfer area are necessary if only on exchanger is used. Calculation wise this is exemplified with a very low correction factor F for this case. Instead, the equations applied in Appendix 8 adjusts the number of heat exchangers in series in order to obtain a correction factor above the recommended value of 0,75. The number of units is calculated to 6.

With the assumed geometric, thermal and hydraulic conditions, and overall heat transfer coefficient U is found to be 944 W/m^2K, with an overall correction factor F of 0,77. As for the DCC CU cooler, these numbers will vary depending operating conditions during later parametric studies. Therefore, an assumed constant value of U*F=750 W/m^2K has been applied for the calculation of required heat exchanger area. With these assumptions, the required heat exchanger area is found to be 31814 m^2. Distributed over 6 units this gives an area of 5302 m^2 per heat exchanger unit. The total heat exchanger area is distributed over several units due to practical limitations in manufacture and on-site maintenance of the heat exchangers [9]. This area is the basis for estimation of lean/rich heat exchanger costs. The physical size of each heat exchanger unit is estimated to 2,1 m diameter with a length of 12,8 m.

The calculated pressure loss for all units on the tube side is 94 kPa, while the shell side is calculated to 265 kPa. This correlates well to the simulation specifications which was set constant in the lean/rich heat exchanger to 100 and 300 kPa respectively.

3.6.2 Lean amine cooler

Performance and calculations for the lean amine cooler is shown in Table 3-8.

Table 3-8 Lean amine cooler dimensioning

Duty *(Hysys)* [MW]	47,8
LMDT uncorrected [^oC]	25,6
U*F*A factor [kW/K]	1810
Assumed flow velocity tube side [m/s]	1
Assumed flow velocity shell side [m/s]	1
Calculated U [W/m^2K]	899
Assumed constant U*F [W/m^2K]	850
Required heat exchanger area [m^2]	2130
Number of heat exchanger units	1
Pressure loss tubeside [kPa]	9
Pressure loss shellside [kPa]	23
Estimated shell diameter [m]	1,6
Estimated shell length [m]	9,4

The calculated duty of the lean amine cooler is 47,8 MW, with and uncorrected LMDT of 25,6^{o}C. For calculation purposes, the product of calculated overall heat transfer coefficient U and correction factor, U*F, has been set constant at 850 W/m^2K. The area requirement of the lean amine heat exchanger is found to be 2130 m^2, located in one unit. For the complete dimensioning, the reader is referred to Appendix 9.

3.6.3 Water condenser

In Table 3-9 the performance and dimensioning assumptions and calculations for the water condenser can be seen. The complete documentation of the dimensioning of the condenser can be seen in Appendix 13.

Table 3-9 Water condenser dimensioning

Condenser duty *(Hysys)* [MW]	25,3
LMDT corrected *(Hysys)* [K]	40,1
Condenser assumed U [W/m^2K]	2000
Condenser area [m^2]	315

As shown in the table, the duty of the condenser is 25,3 MW. The corrected LMDT calculated by Hysys of 40,1 K is used for dimensioning due to the complex temperature profile when the desorber overheads are partially condensed down to 40^oC. The overall heat transfer coefficient U is found from literature and is assumed to be constant at 2000 W/m^2K [12]. Calculation of the heat exchanger equation results in a required heat exchanger area of 315 m^2. The calculated heat exchanger area is the basis for heat exchanger base cost.

3.7 Water separator

The water separator for separation of liquid water from the CO_2 is shown in Table 3-10. Full calculations are shown in Appendix 14.

Table 3-10 Water separator dimensioning

K-factor [m/s]	0,1
Vertical flow velocity [m/s]	1,77
Vessel diameter [m]	3,05
Vessel total height [m]	12,2
Water recovery [t/h]	33,6

The separator has been designed with a K-factor of 0,1 when using Souder-Brown equation for determining the gas vertical velocity. This results in a vessel diameter of 3,05 m, and with the chosen length/diameter ratio given in Appendix 14, a vessel height of 12,2 m. The water recovery rate is by Hysys calculated to 33,6 t/h.

4 Cost estimation methods

Estimation of basis cost and finally installed equipment cost is performed on the basis of the various equipment dimensions and capacities found in chapter 3 Dimensioning and selection of equipment. Further explanations for the methods of cost estimation are covered in this chapter.

4.1 Classification of cost

Expenses are often divided into two categories, capital expenditure (CAPEX) and operational expenditure (OPEX).

CAPEX are funds used for acquisitions of physical assets in order to be able to create a service or product. In this project, CAPEX are the installed costs for process equipment necessary to the CO_2 capture plant. Important costs that are not considered, but will impact considerably on the total plants costs are:

- Acquisition of property
- Ground preparation
- Utility connections to nearby infrastructure
- Administrative buildings, offices, control rooms etc.

OPEX are costs necessary in order to run and maintain equipment, system and to pay worker wages. In this context, operating expenditure is cost related to purchase of utilities like steam, electricity and cooling water. Important OPEX cost that are not accounted for and quantified in this report may be:

- Salary for employees
- Maintenance cost of equipment and buildings
- Taxes and other capital costs
- Raw materials
- Spare parts

As can be seen on the above, there are some costs not accounted for in this report. However, these are assumed to be constant regardless of parametric setting of the process plant, and should be considered as a constant addition to this projects cost estimate.

4.2 Equipment cost calculation

In order to estimate the cost of process equipment, the following data for the equipment has to be known [9]:

- The characteristic size of the equipment
  - Depending on type of equipment, this can be weight, volume, power requirement, physical dimensions, heat transfer area or a mass flow rate
- Materials of construction
  - Depending on service, equipment material has to be adopted to the specific application
- Design pressure and temperature
  - The equipment has to be constructed and sized to be suitable for the application

In this report, the installed equipment cost for the process equipment is found from the following procedure:

- Finding the characteristic size of the capacity measure for the equipment
- Obtaining the base cost from a power law capacity correlation relating it to a known cost for a equipment with a different capacity for a given year
- Adjust the obtained base cost to the correct currency based on the currency exchange rate for the year of the cost correlation
- Adjust for inflation and price change in the time period from the year of cost estimation and to present date by using a cost index. In this report the Chemical Engineering Plant Cost Index (CEPCI) has been used
- Using a installation factor scheme adjusting for material selection in order to find the total installed cost for the equipment [17]

For the DCC and absorption tower, costs for liquid distributors and mechanical supporting have been estimated as a percentage of vessel shell and packing costs. The capture plant operates at low pressures, so no correction factors for high pressure equipment has been considered.

4.2.1 Power law of capacity

The power law capacity correlation is shown in Equation 4-1 [9].

$$C_E = C_B \left(\frac{Q}{Q_B}\right)^M \qquad \textit{Equation 4-1}$$

Where C_E = cost of equipment with capacity Q

C_B = known cost for equipment with capacity Q_B

M = scaling constant for specific type of equipment

The nature of this cost correlation is that the cost of an equipment is over-, under- or proportional, depending on the selection of the scaling constant M.

4.2.2 Currency conversion

All base cost estimates are available in US dollars, and in order to convert this to NOK, the appropriate exchange rates have to be applied. For this project, the historic average exchange rate for the year of the equipment base cost estimate has been used. The equation used is shown in Equation 4-2.

$$C_{NOK} = C_{\$} \cdot R$$ *Equation 4-2*

Where C_{NOK} = Price in NOK for a given year

$C_{\$}$ = Price in \$ for a given year

R = Currency exchange rate for a given year NOK/\$

4.2.3 Cost index

Prices tend to increase over time, both due to inflation but also as a consequence of supply and demand of the different equipment. To account for this calculation methods can be applied. The price index correlation is showed in Equation 4-3 [9].

$$C_1 = C_2 \cdot \frac{Index_1}{Index_2}$$ *Equation 4-3*

Where C_1 = equipment cost in year 1

C_2 = equipment cost in year 2

$Index_1$ = index value in year 1

$Index_2$ = index value in year 2

The CE Plant cost index is a commonly used cost index which can be further subdivided into four subcategories. The CE cost index for equipment has been used during cost estimation in this report. No cost index for 2010 was available, so the equipment cost index for 2009 has been applied. This means that all equipment cost is in 2009 NOK. The base cost is shown for the estimated equipment in Appendix 6 to Appendix 14.

4.2.4 Installation factors

The installed equipment cost, in this context sometimes referred to as CAPEX, are costs involved in purchase and installation of the equipment and includes:

- Direct cost
- Engineering cost
- Administration cost
- Commissioning
- Contingency

The calculation of the total installed cost from the purchase cost of equipment is found by applying Equation 4-4 [17].

$$C_i = C_p \cdot [f_{TC} - f_P - f_E + f_m \cdot (f_P + f_E)] \qquad \textit{Equation 4-4}$$

Where: C_i $= Total\ installed\ cost - CAPEX\ for\ a\ equipment\ [NOK]$

C_p $= Purchase\ cost\ for\ a\ equipment\ [NOK]$

f_{TC} $= Total\ installed\ cost\ factor$

f_P $= Piping\ cost\ factor\ for\ equipment$

f_E $= Equipment\ cost\ factor$

f_m $= Material\ cost\ factor$

An overview of all base costs, installation factors and installed costs can be found in Appendix 15. In the appendix, the selected materials for the various equipments are stated. All choices of materials are assumed, and do not significantly affect the final installed cost estimate.

4.3 Cost of utilities

The cost of steam and electricity are given as cost per energy unit. The steam price is assumed, while electricity cost is derived from the assumption that steam is used to drive a steam turbine producing electricity with an efficiency of 25 %. The cooling water cost is calculated from the cost of transporting the water by using Equation 4-5 with the following assumptions:

- Base case electricity price of 0,4 NOK/kWh
- Pump differential pressure of 3 bar
- Pump efficiency of 100 %

The cooling water cost is found from manipulating the pump power equation found in Appendix 2 and is shown in Equation 4-5:

$$C_{cw} = C_{El} \cdot \frac{\rho g H}{3{,}6 \cdot 10^6} \quad \textit{Equation 4-5}$$

Where: $C_{cw} = Cost\ of\ cooling\ water[NOK/m^3]$

$C_{El} = \ Cost\ of\ electricity\ [NOK/kWh]$

$\rho \ = \ Density\ of\ water\ [kg/m^3]$

$g \ = \ Gravitational\ constant[m/s^2]$

$H \ = \ Lifting\ height\ of\ pump\ [m]$

A summary of the utility costs are found in Table 4-1.

Table 4-1 Utility cost

Cost hot utility /steam [NOK/kWh]	0,1
Cost electricity [NOK/kWh]	0,4
Cost cooling water [NOK/m^3]	0,033

No capital cost investment for equipment for preparation, generation or infrastructure for transportation for the cooling water, steam or electricity has been applied. The total cost of all utilities is referred to as OPEX the report.

4.4 Net present value

When the capital costs involved in installation of the necessary equipment and the operational cost of utilities are known, an evaluation of the total cost considering both over a time period can be made. This is done by the net present value (NPV) method, which is a way of calculating the relative effect of the process solution with respect to the influence of CAPEX versus OPEX on the total cost for a given time period and discount rate. The CAPEX are assumed to occur in year 0, while the operational costs occur each year for a given calculation period. The future operational costs are given in nominal cost, i.e. the value is expressed in the value in year 0. In real life inflation will cause the cost to increase over the time period. In the NPV calculation, the operational costs are discounted from the nominal value to a present value by utilizing an assumed discount rate. The expression for calculation of the NPV of the sum of future operational cost is shown in Equation 4-7 [18].

$$NPV_{OPEX} = a \cdot \frac{(1+r)^n - 1}{(1+r)^n \cdot r} \quad \textit{Equation 4-7}$$

Where: $NPV_{OPEX} \ = Sum\ of\ Net\ present\ value\ of\ operational\ costs\ [NOK]$

a $= Annual\ operational\ cost\ /\ OPEX\ \ [NOK]$

$r \ = Annual\ interest\ rate$

$n \ = Number\ of\ years$

The NPV for the process solution is calculated from Equation 4-8.

$$NPV = CAPEX + NPV_{OPEX} \qquad \textit{Equation 4-8}$$

Where: $NPV \quad = Net\ present\ value\ of\ all\ costs\ [NOK]$

$CAPEX \ = Equipment\ installation\ cost\ [NOK]$

The calculation in Equation 4-8 only consider costs involved with the CO_2 capture, so any NPV mentioned in this report is related to costs. Following this, the process solution with the lowest NPV in this report shows the less expensive process solution. The assumed discount rate and calculation period for NPV calculations is shown in Table 4-2.

Table 4-2 NPV calculation assumptions

Calculation period [years]	20
Discount rate per year [%]	7

Furthermore, the NPV calculation assumes no rest-value of the equipment after the calculation period. In practice there will be decommissioning costs and possible income from sale of salvaged equipment and materials.

5 Results of simulation and cost estimate calculations

5.1 Base case

5.1.1 Process performance

The CO_2 capture efficiency, utility- consumption and costs for the base case process are shown in Table 5-1.

Table 5-1 Base case process results

CO_2 removal [%]	85,2
Energy consumption [MJ/kg CO_2]	3,61
Hot utility consumption [MW]	154,2
Electricity consumption [MW]	23,5
Cold utility consumption [m^3/h]	14938
Uptime [h/yr]	8000
CO_2 removed [Mt/yr]	1,230
Hot utility cost [MNOK/yr]	123,4
Electricity cost [MNOK/yr]	75,3
Cold utility cost [MNOK/yr]	4,0

The energy consumption is derived by dividing the hot utility consumption in the desorber reboiler by the mass flow of CO_2 captured.

5.1.2 Base case equipment cost estimates

The base case equipment installed cost and their relative distributions are shown in Table 5-2.

Table 5-2 Base case equipment installed costs and relative distributions

Equipment	Installed cost [MNOK]	Relative CAPEX [%]
Transport fan	50	3,6
Direct contact cooler	147,7	10,5
Absorber column	858,3	61,3
Lean amine cooler	21,8	1,6
Lean amine circulation pump	12,0	0,9
Lean/rich heat exchangers	220,3	15,7
Desorber column	47,1	3,4
Desorber condenser	5,6	0,4
Desorber reboiler	18,9	1,3
Desorber OH condenser	4,9	0,3
Desorber OH water separator	0,9	0,1
Rich amine circulation pump	13,1	0,9
Total installed cost	1400	100

An overview of each equipment base cost and installation factors are summarized in Appendix 15.

5.2 Sensitivity analysis

A series of case studies in order to investigate the effect of parametric changes on the costs related to the capture of CO_2. The various case studies are discussed in the following subchapters.

5.2.1 Approach temperature in lean/rich heat exchanger

A case study has been performed in order to investigate economical performance when changing the degree of heat recovery in the lean/rich heat exchanger. This is quantified by the term minimum approach temperature, described in chapter 2.1.7. The ΔT_{min} has been adjusted in the range 5 to 25^{0}C. During the study, all parameters regarding the flue gas and absorption column has been held constant for a given overall CO_2 removal efficiency. This is also the case for the lean amine flow rate and composition.

For each study a graph of the NPV and energy consumption for each step during the case is shown. The studies have been notated study 1A, 1B and 1C, reflecting an overall CO_2-removal efficiency of 82, 85 and 90 % respectively.

5.2.2 Absorber packing height

A case study has been performed in order to investigate economical performance when changing the number of stages in the absorber column. One absorber stage is assumed to be 1 meter of packing height. During the case study, the flue gas pressure upstream the absorber column has been adjusted to account for pressure drop as a function of number of stages in the absorber. The pressure drop correlation has been assumed to be proportional with the number of stages in the absorber by the factor 0,94 kPa per meter of packing.

The lean amine composition has been held constant during the studies, but the lean amine flow rate has been adjusted for each study in order to achieve to required CO_2 removal efficiency.

For each study a graph of the NPV and energy consumption for each step during the case is shown. The studies have been notated study 2A, 2B and 2C, reflecting an overall CO_2-removal efficiency of 80, 85 and 90 % respectively.

5.2.3 Absorber feed gas temperature

A case study has been performed in order to investigate economical performance when changing the flue gas inlet temperature to the absorber column. This is achieved by changing the circulation flow rate of cooling water in the DCC, while keeping number of stages in the absorber and hence absorber inlet pressure constant. The lean amine composition has been held constant during the studies, but the lean amine flow rate has been adjusted for each case in order to achieve to required CO_2 removal efficiency. The study has also been accomplished in conjunction with changing the Murphree stage efficiency to adjust for the effects from various temperature profiles in the absorber column at different feed gas temperatures and lean amine flow rates. The calculation scheme for estimation of the Murphree stage efficiency has been developed by Øi [19]. The absorber column temperature profile varies depending on feed gas inlet temperature, lean amine flow rate (and temperature), and the total CO_2 removal efficiency. The input data for the calculations are stage pressure and temperature. From the calculation scheme, only the top-, bottom- and maximum temperature stage has been calculated, and the intermediate stages found from a linearization between the three known points. The temperature profile in the column is typically at its maximum somewhere in the upper part of the top section. The maximum Murphree stage efficiency value has been deducted a value of 0,01.

For each study a graph of the NPV and energy consumption for each step during the case is shown. The studies have been notated study 3A and 3B, reflecting an overall CO_2-removal efficiency of 85 and 90 % respectively.

5.2.4 Uptime and calculation period

A case study has been performed in order to investigate the change in the specific cost per ton of CO_2 removed when changing the cost calculation assumptions for the base case. This is done by changing the calculation period for the plant at various uptime figures.

The study has been performed with a calculation period of 10, 15 and 20 years. Each calculation period has been calculated for and yearly utilization (uptime) of 4000 and 8000 hours.

5.2.5 Amine fluid package

A case study has been performed in order to investigate the change in process performance and change in costs for the base case when changing the amine fluid package (FP). The Li-Mather amine FP has been compared to the Kent-Eisenberg FP used in BC.

5.3 Results from sensitivity analysis

5.3.1 Study 1A

Change in ΔT_{min} – 82 % CO_2 removal

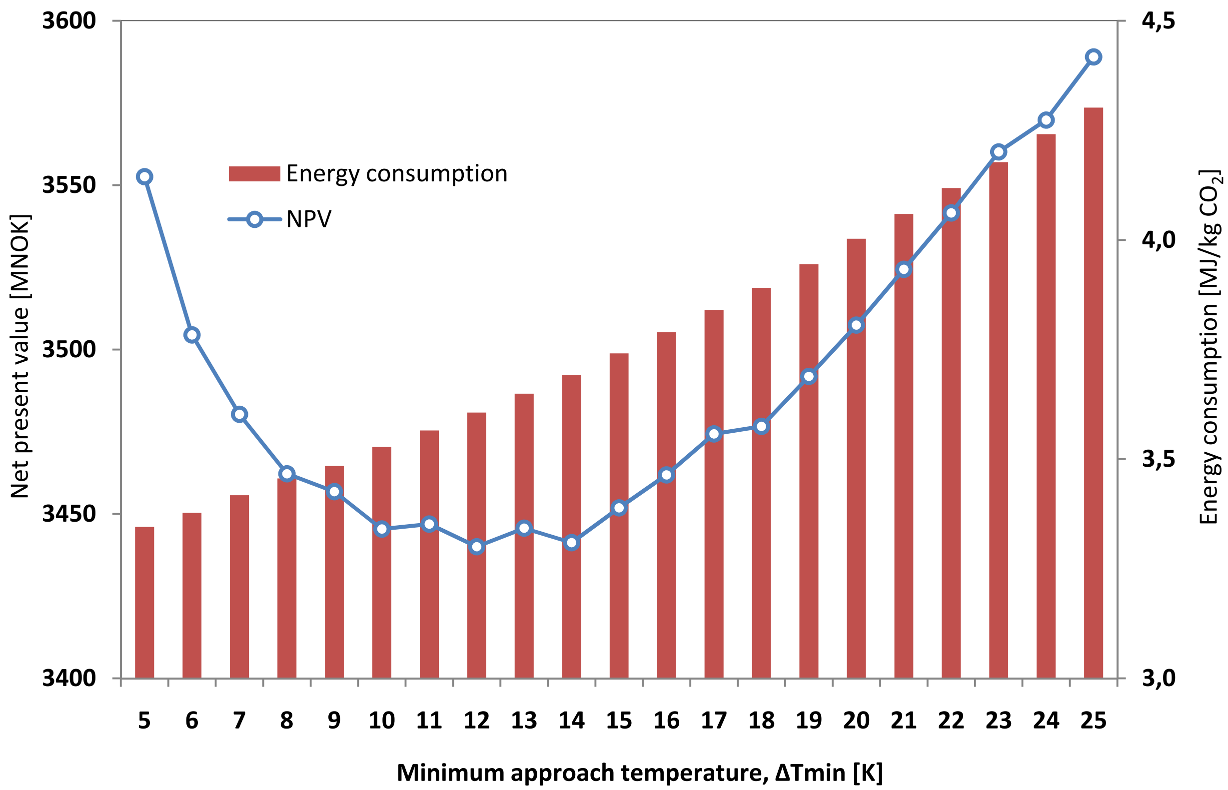

Figure 5-1 Results of NPV and energy consumption as a function of minimum approach temperature in the lean/amine heat exchanger at 82 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.2 Study 1B

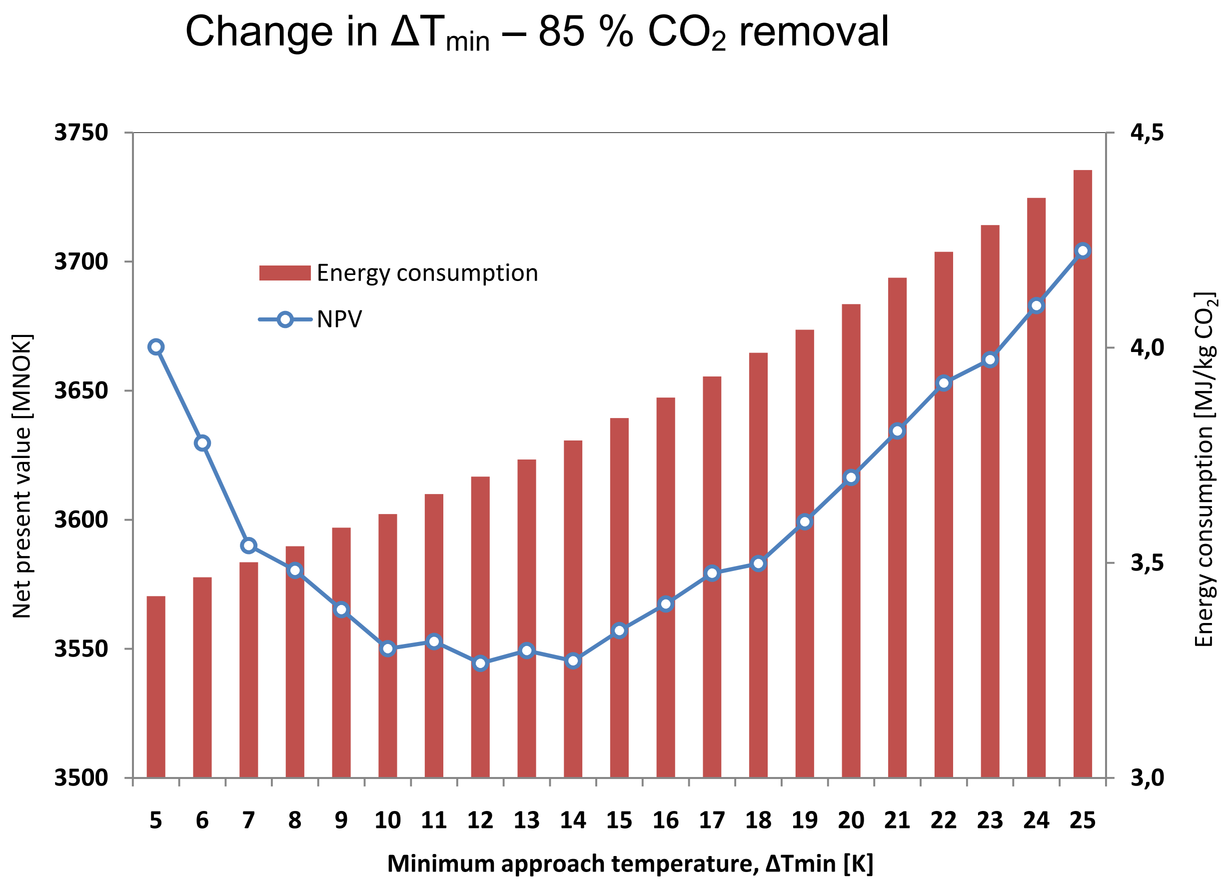

Figure 5-2 Results of NPV and energy consumption as a function of minimum approach temperature in the lean/amine heat exchanger at 85 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.3 Study 1C

Change in ΔT_{min} – 90 % CO_2 removal

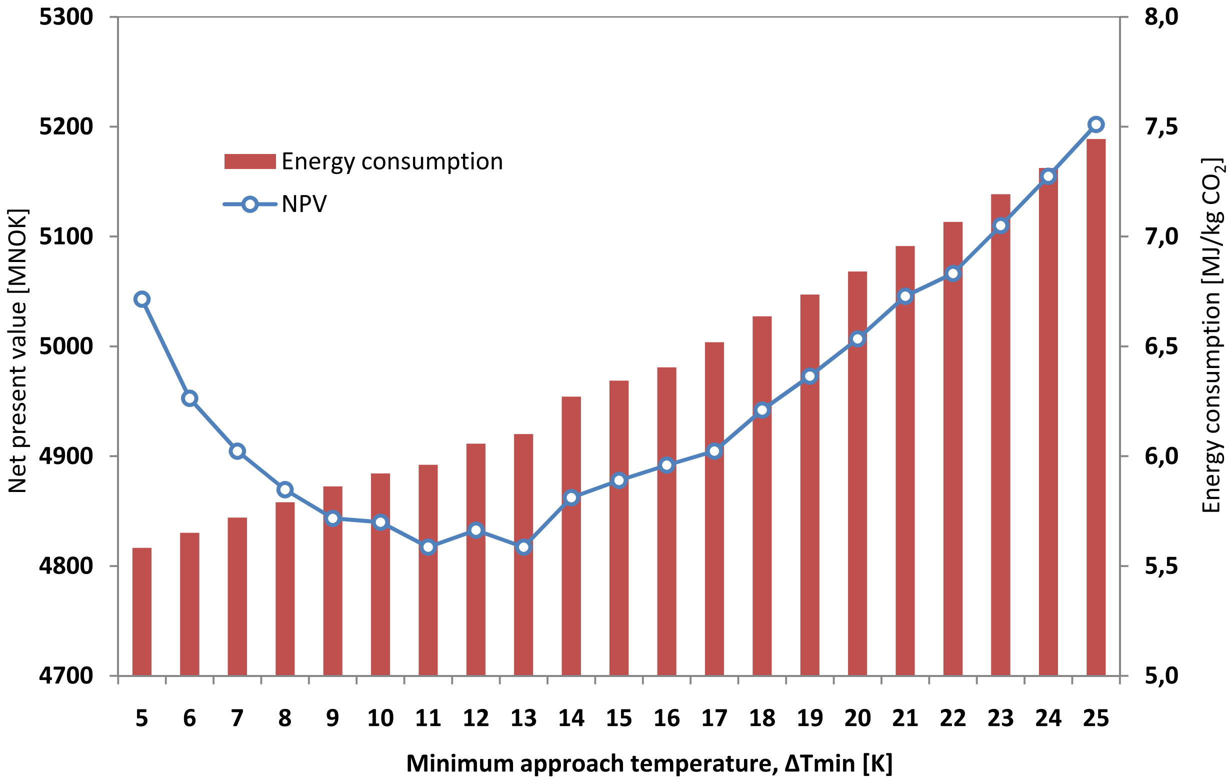

Figure 5-3 Results of NPV and energy consumption as a function of minimum approach temperature in the lean/amine heat exchanger at 90 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.4 Study 2A

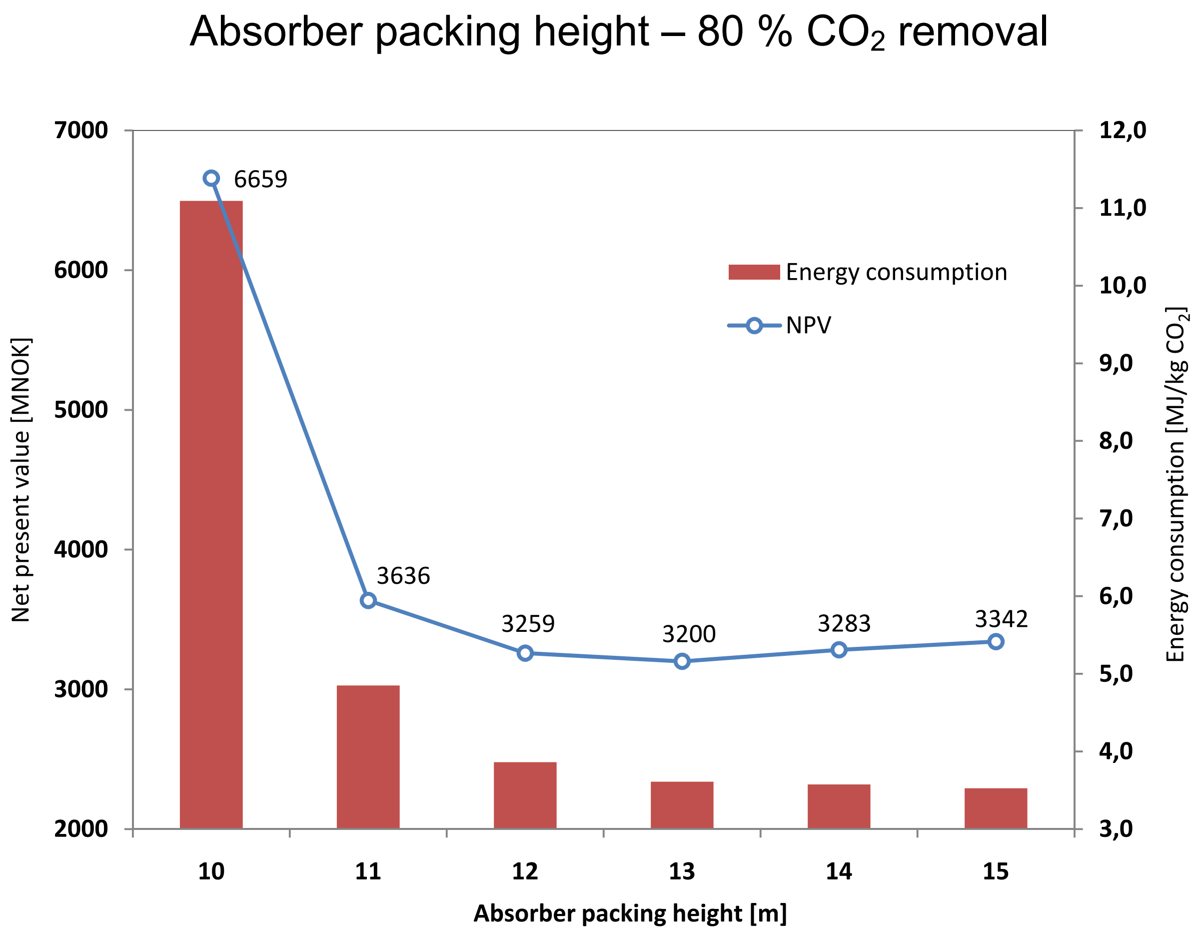

Figure 5-4 Results of NPV and energy consumption as a function of absorber packing height, constant Murphree stage efficiency of 0,15 m^{-1} and 80 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.5 Study 2B

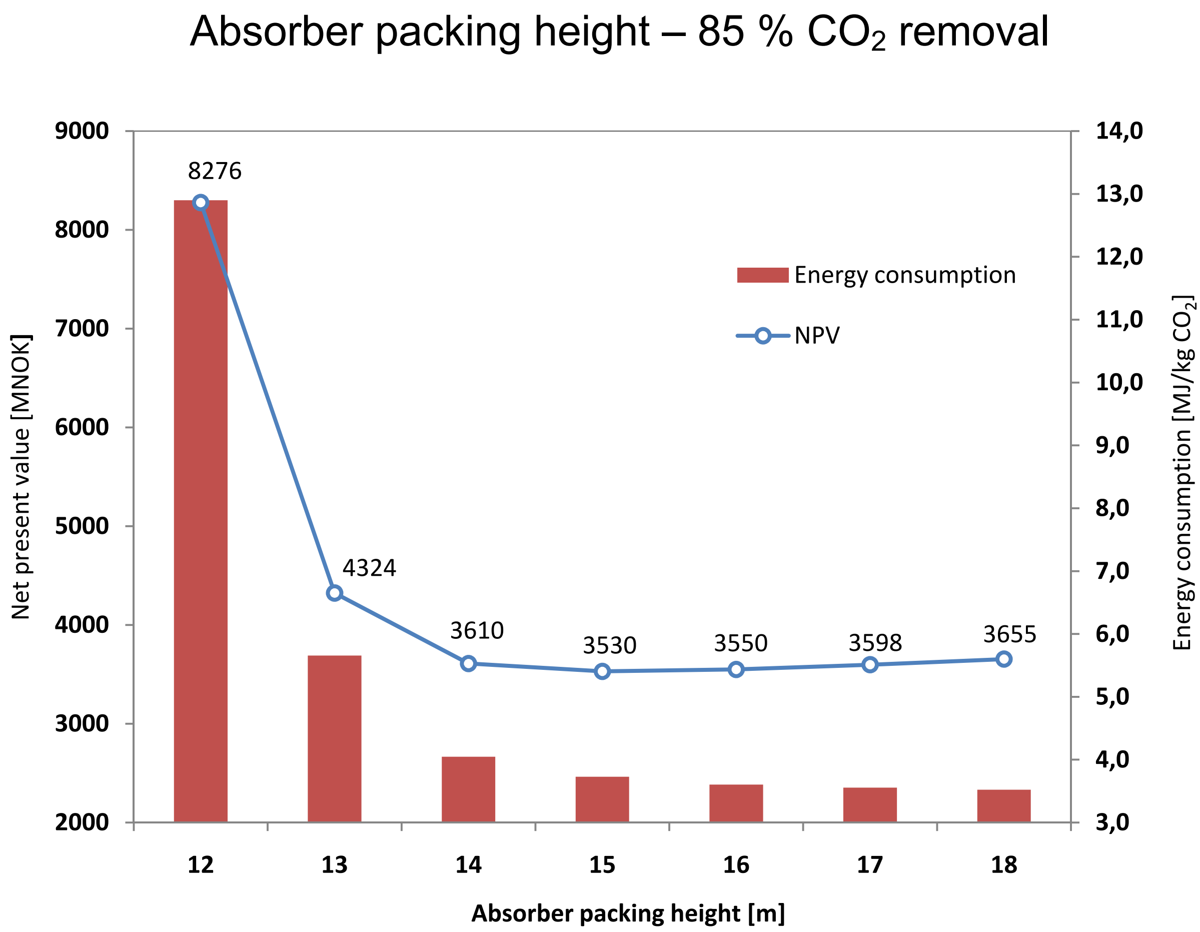

Figure 5-5 Results of NPV and energy consumption as a function of absorber packing height, constant Murphree stage efficiency of 0,15 m^{-1} and 85 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.6 Study 2C

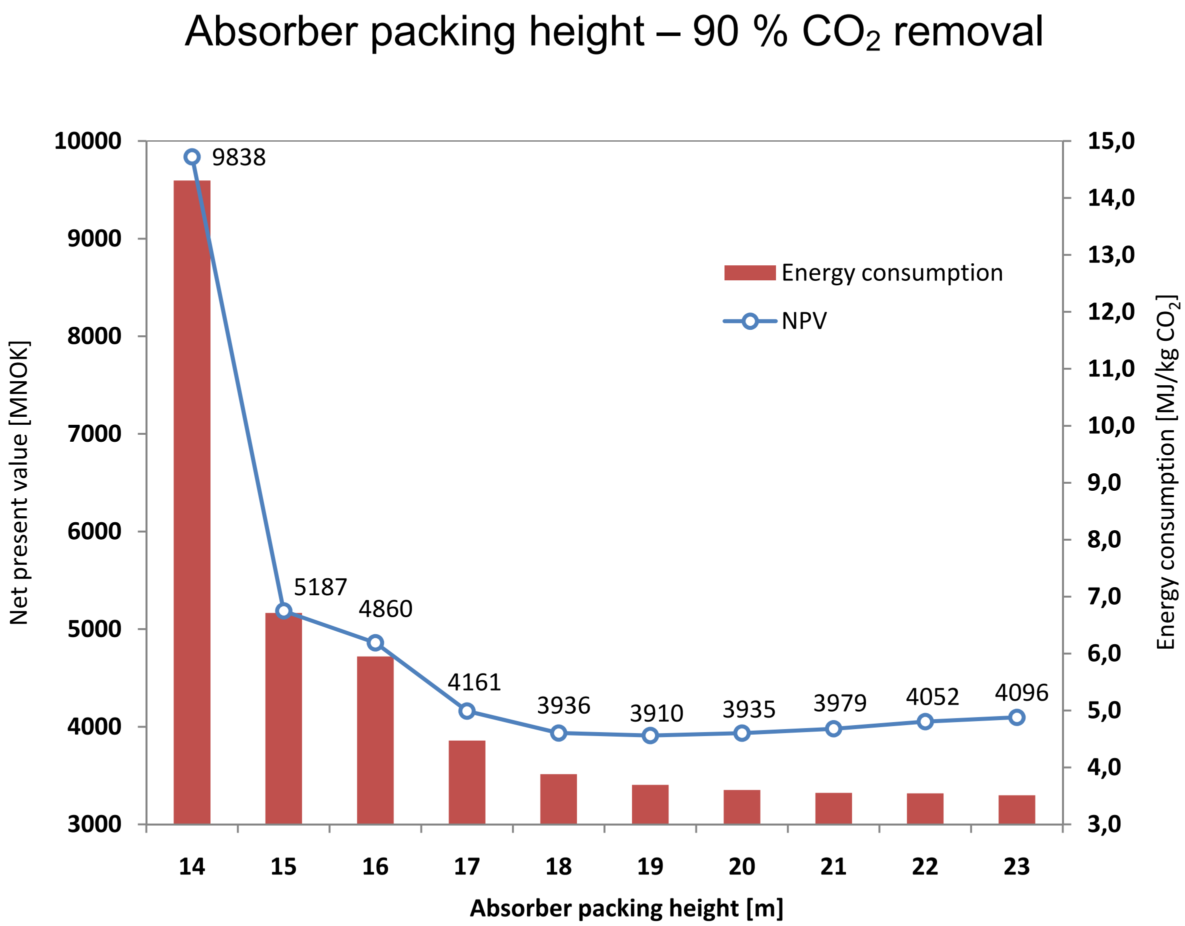

Figure 5-6 Results of NPV and energy consumption as a function of absorber packing height, constant Murphree stage efficiency of 0,15 m^{-1} and 90 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.7 Study 3A

Absorber feed gas temperature 85 % CO_2 removal

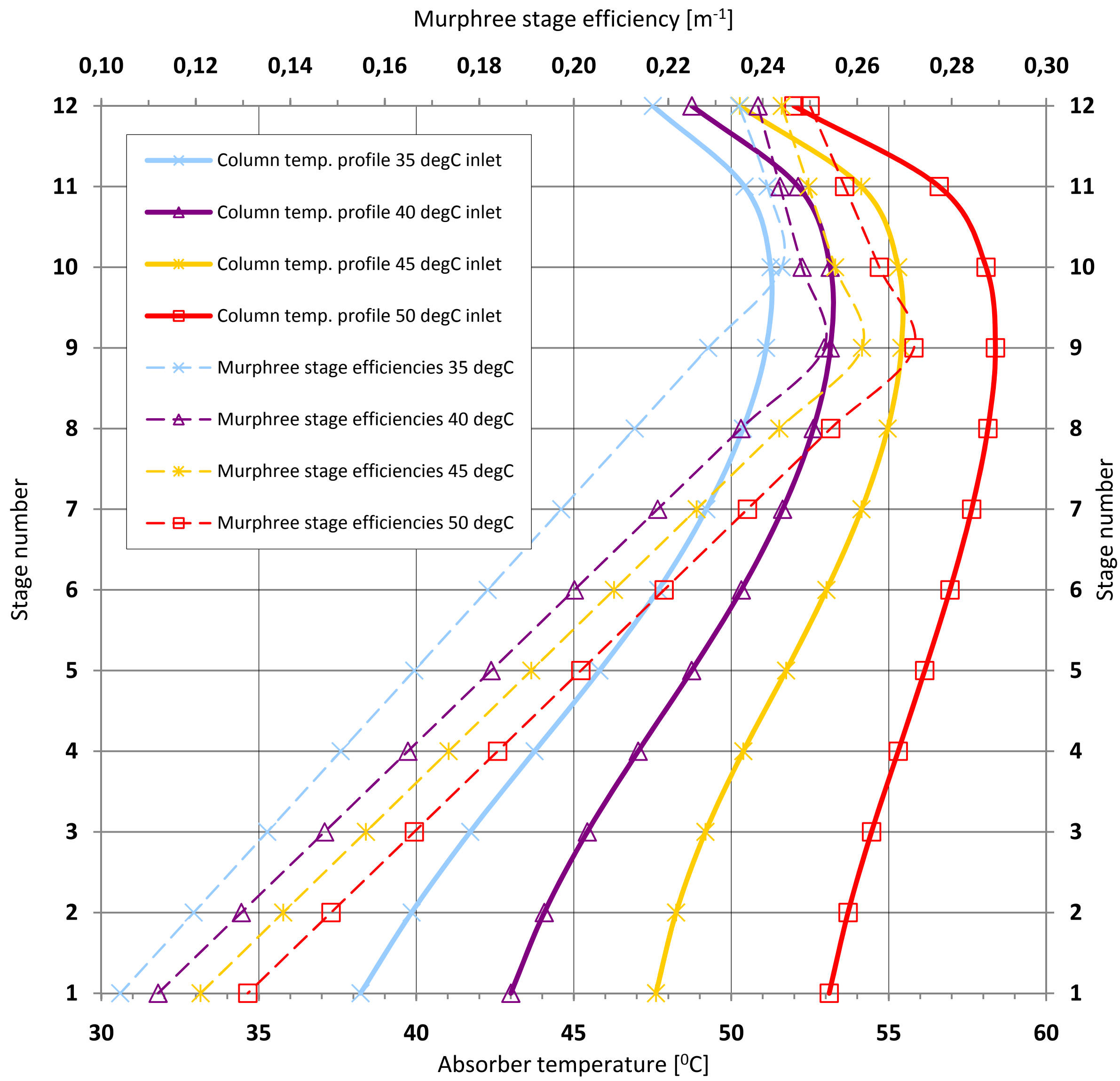

Figure 5-7 Absorber temperature and estimated Murphree stage efficiency profiles at 85 % removal rate

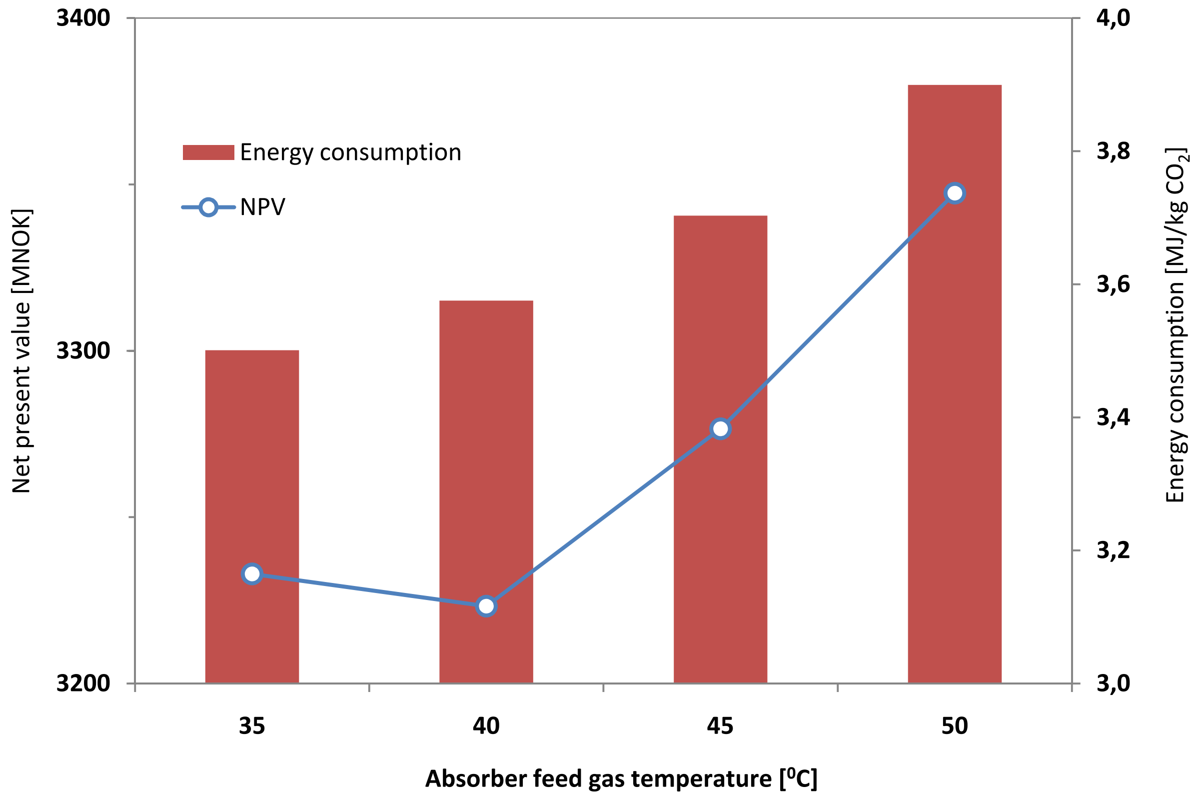

Figure 5-8 Results of NPV and energy consumption as a function of absorber feed gas temperature, 12 meter packing height in absorber, variable Murphree stage efficiency, 85 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.8 Study 3B

Absorber feed gas temperature 90 % CO_2 removal

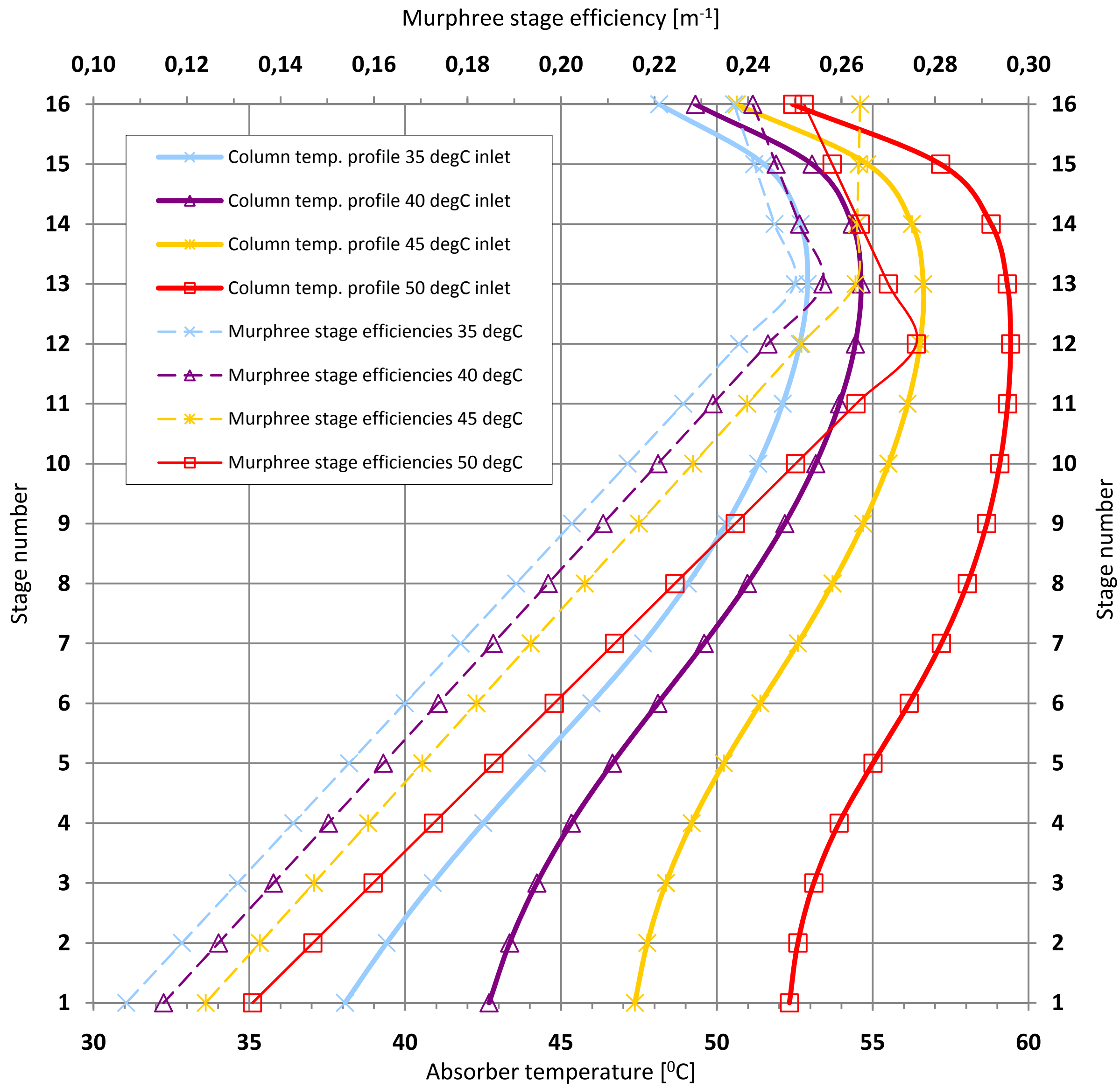

Figure 5-9 Absorber temperature and estimated Murphree stage efficiency profiles at 90 % removal rate

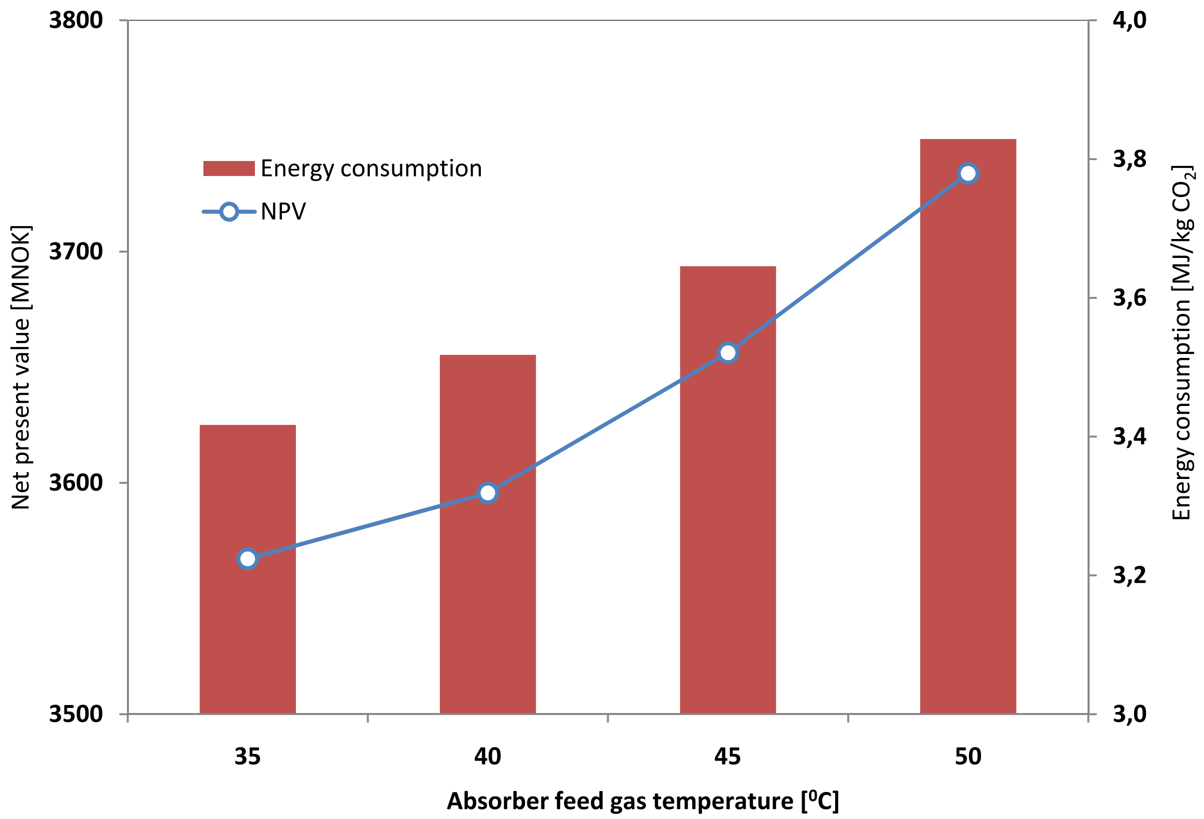

Figure 5-10 Results of NPV and energy consumption as a function of absorber feed gas temperature, 16 meter packing height in absorber, variable Murphree stage efficiency, 90 % removal efficiency, 20 years calculation period and 7 % interest rate

5.3.9 Study 4A Uptime and calculation period

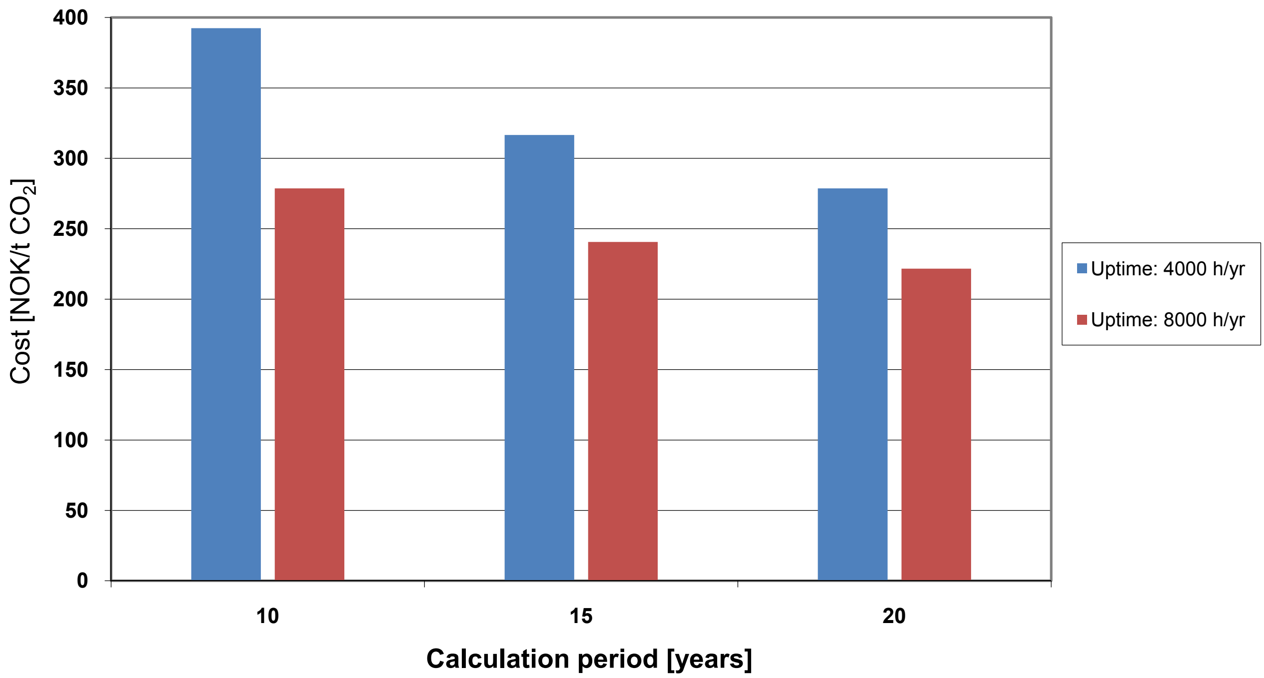

Figure 5-11 Nominal cost per ton CO_2 at base case conditions as a function of uptime and calculation period

5.3.10 Study 5A Amine fluid package

The result of changing the amine fluid package in the base case (BC) condition is shown in Table 5-3.

Table 5-3 Results from changing amine FP on BC

	BC with Kent-Eisenberg FP	BC with Li-Mather FP	Change relative to K-E FP [%]
CO_2 removed [%]	85,2	85,2	0,0
Energy consumption [MJ/kg CO_2]	3,61	3,69	2,2
Lean amine rate [t/h]	3600	3800	5,6
OPEX total [MNOK/yr]	202,7	205,7	1,5
CAPEX [MNOK]	1400	1412	0,8
Total NPV [MNOK]	3548	3591	1,2

6 Discussion

6.1 Base case results

The results from the base case process simulation and cost estimation of utility costs and equipment installed cost can be found in Table 5-1 and Table 5-2.

The base case CO_2 capture process has at a removal efficiency of 85 % and a specific energy consumption of 3,61 MJ/kg CO_2. This is well in line with what is reported in literature. It is lower than what was found in the work of Blaker [11], where for the same flue gas composition and flow rate it was found to be 4,28 MJ/kg CO_2. In this study the ΔT_{min} was somewhat higher at 14 K, however at parametric studies of the minimum approach temperature it did not go below 4,15 MJ/kg CO_2 at an ΔT_{min} of 6 K.

The installed cost of all equipment in the base case is found to be 1400 MNOK with an annual utility cost of 203 MNOK. Of the equipment installed costs, the absorber, the DCC and flue gas transport fan makes up 75 % of total cost, which is in line with previous cost estimates [4]. Other major equipment costs are related to the lean/amine heat exchangers and the desorber column.

Of the utility costs, steam consumption in the desorber reboiler accounts for 61 %. The desorber reboiler duty is calculated to 154,2 MW, while the electricity consumption is 23,5 MW. 92 % of all electricity consumption is consumed by the flue gas transport fan. The total cooling water requirement is found to be 14 938 m^3/h, where 45 % of this is in the DCC CU cooler. The overall cold utility consumption is somewhat higher than the studies found by Svendsen *et al.* [6], but the relative consumption in the DCC unit is roughly the same. The differences are probably attributed higher flow rate of flue gas in this report and assumptions for cold utility feed- and discharge temperatures. The results show that the electricity and cooling water consumption are largest in the flue gas flow path of the capture plant, while steam consumption is related exclusively to the regeneration of the amine solution.

Over a calculation period of 20 years and an uptime of 8000 hours per year, the NPV of all cost is calculated to 3548 MNOK. This is higher than that of Blaker, which found NPV cost to be 2649 MNOK, this however was at a calculation period of 10 years.

6.2 Study 1 ABC ΔT_{min} in lean/rich heat exchanger

The results from study 1A, 1B and 1C can be seen in Figure 5-1, Figure 5-2 and Figure 5-3.

In order to make the simulations in study 1 ABC converge, the recycle function on the lean amine was disabled. Due to limitation in absorber convergence at low flow rate of lean amine, a minimum overall CO_2 removal efficiency of ~82 % was achieved for study 1A for the specified absorber packing height of 16m.

This study is the classical trade-off between heat exchanger area and the requirement of external utility [9]. The main variables affected are the L/R heat exchanger size, the lean amine cooler size and CU consumption, desorber reboiler HU consumption, and to a lesser extent, the desorber- condenser, reboiler and column.

For study 1B, the amine flow rate in the system has been adjusted to 3600 t/h in order to achieve an overall CO_2 removal of ~85 %. The case at ΔT_{min}=10 K corresponds to the base case. The lean amine cooler installed cost increases on average 3,4 % for each increase of one degree Celsius of the ΔT_{min}. The CU consumption in the lean amine cooler simultaneously increases on average 6,5% for each incremental step of one degree increase of the ΔT_{min}. For the L/R heat exchanger, the opposite is the case: on average a reduction of 8,3 % is experienced on the heat exchanger installed cost. Due to reduced inlet temperature of the rich amine to the desorber, the desorber hydraulic load reduces, hence reducing the desorber column size. More steam will have to be added to the desorber reboiler as the energy recovery in the L/R heat exchanger reduces. The steam consumption increases on average 1,3 % for one degree Celsius increase of ΔT_{min}. There is also a minor increase in desorber reboiler installed cost. All costs related to electricity consumption are constant at 75,3 MNOK per year during the variation of ΔT_{min} in the specified range.

The specific energy consumption increases linearly from its lowest value at 3,42 up to 4,41 MJ/kg CO_2 at ΔT_{min} of 5 and 25 ^oC respectively in study 1B. The specific energy consumption is slightly higher in the same ΔT_{min} interval for this study compared to study 1A.

The increase of costs when increasing the overall CO_2 removal efficiency from 82 to 85 % in study 1A and 1B is regarded as minor. The difference is attributed a larger flow rate in the amine loop in study 1B (3600 versus 3340 t/h) in order to increase the CO_2 removal efficiency.

The area of minimum NPV for study 1A and 1B seems to occur in a relatively flat area between ΔT_{min}= 10-14^oC. The approximate value of the NPV in this wide minimum is 3550 MNOK for study B, compared to 3445 MNOK in study 1A. From this it can also be seen that the base case is near optimum with regards to its assumed ΔT_{min} =10 K.

The overall change of NPV during study 1B shows a difference from minimum to maximum value of 160 MNOK in the specified range of ΔT_{min}. The effect on NPV for this design variable seems to be increasing as the CO_2 removal rate increases, as it was 149 MNOK in the same interval in study 1A.

For study 1C the amplitude and absolute value of process performance and installed costs have changed significantly. The lean amine flow has been adjusted to 7850 t/h in order to obtain a CO_2 removal efficiency of 90 %. This is considerably higher than the flow rates in study 1A and 1B. All costs related to electricity are constant during the variation of ΔT_{min} at 81 MNOK per year, which is somewhat higher level than in the 1A and 1B studies. This is primarily due to the higher flow rates of amines in order to meet the required removal rate specification.

At 90 % removal efficiency, the specific energy consumption increases proportional from its lowest value at a 5,58 up to 7,44 MJ/kg CO_2 at ΔT_{min} of 5 and 25 ^oC respectively. The energy requirement for increasing the CO_2 removal rate 5 %-points has a major impact on amine flow rate, and hence hot utility requirement. The area of minimum NPV seems to occur in a relatively wide range of ΔT_{min}= 11-13 ^0C, a slightly narrower range than in the 1A and 1B studies. NPV is approximately 4822 MNOK in the minimum range, which is ~6 % higher than in the optimum region at 85 % removal efficiency.

The overall change of NPV during study 1C shows a difference from lowest to highest value of 385 MNOK, and show that the amplitude within the range of ΔT_{min} seems to increase as the CO_2 removal rate increase. This indicates that optimum ΔT_{min} increases in importance as the CO_2 removal rate, and hence the circulation rate of amine, increases.

For full documentation of the process parameters, performance, OPEX and CAPEX for each step in study 1A, 1B and 1C it is referred to Appendix 17 through Appendix 22.

6.3 Study 2 ABC absorber packing height

The results from study 2A, 2B and 2C can be seen in Figure 5-4, Figure 5-5 and Figure 5-6.

For the various absorber packing heights, the inlet pressure has been varied to simulate the necessary driving pressure. The pressure gradient used is 0,94 kPa/m packing. Another assumption is that the pressure downstream the absorber packing material always should be 106 kPa an in order to allow for further pressure loss in a downstream water wash section. With these assumptions, it means that for example 10 absorber stages requires an absorber inlet pressure of 115,4 kPa a, while 21 stages requires 125,7 kPa a. The implications of these assumptions are that the duty of the upstream transport fan is increased as more stages in the absorber are introduced due to the increased pressure loss. Furthermore, increased compression has the consequence of a higher flue gas temperature downstream the transport fan, which means that the DCC has to increase its cooling performance for a specified feed gas temperature to the absorber column. During the change of absorber number of stages, the absorber Murphree stage efficiency has been kept constant at base case assumptions, 0,15 m^{-1}.

This parametric study is the classic trade-off of any distillation or absorption operation philosophy. For a given feed- and product quality, the absorbent flow rate decreases with increasing number of stages, and the opposite [9]. The major capital changes in this study are the installed costs for the absorber column and the lean/rich heat exchanger.

For study 2B, it was only possible to achieve a removal rate at 85 % with the number of absorber stages in the range of 12 – 18 stages. The case step of 16 stages corresponds to the base case. The NPV at 12 stages is calculated to 8276 MNOK, while it decreases until it reaches its minimum at 15 stages with an NPV of 3530 MNOK. From the minimum value at 15 stages, the NPV steadily increases as the number of stages is increased beyond this point. The effect of reducing the number of stages from 15 to 14 leads to an NPV increase of 80 MNOK. Likewise, the increase of one absorber stage from 15 to 16 stages leads to an NPV increase of 20 MNOK. The case of 16 stages is equal to the BC condition, and these results indicate that the base case process is close to optimum with regards to number of stages in the absorber. The overall energy requirement is also gradually reduced as the number of stages increases. This is seen in connection with the gradual reduction of amine flow rate as the stage numbers are increased. The implications of increased amine flow rates are increased heat exchanger area of both lean amine - and lean/rich heat exchangers and equipment related to the desorber column. The hot utility requirement is reduced as the number of stages are increased, from 12,9 MJ/kg CO_2 at 12 stages, down to 3,52 MJ/kg CO_2 at 18 stages. It is noted that the overall energy consumption and NPV is higher for the 85 % removal efficiency than the 80 %. Beyond 18 stages it is not possible to be able to converge the absorber column due to a minimum lean amine flow rate.

For study 2A it was only possible to achieve 80 % CO_2 removal rate with the number of absorber stages in the range of 10 – 15 stages. Like study 2B at 85 %, the NPV at study 2A with 80 % removal efficiency reaches its maximum at the minimum number of absorber stages. As seen by the NPV function in Figure 5-4, the overall costs increase dramatically as the number of stages is reduced below 11 stages. The overall function of NPV is at its highest at 6659 MNOK when there are 10 stages in the absorber, but reduces dramatically as the number of stages is increased and it reaches its minimum at 13 stages where the NPV is 3200 MNOK. From this point on, the NPV steadily increases as the number of stages is increased beyond this point. The effect of reducing the number of stages from 13 to 12 leads to an NPV increase of 59 MNOK. Likewise, the increase of one absorber stage from 13 to 14 stages leads to an NPV increase of 83 MNOK.

For study 2A, hot utility requirement is reduced as the number of stages are increased, from 11,1 MJ/kg CO_2 at 10 stages, down to 3,53 MJ/kg CO_2 at 15 stages. Beyond this point, it was not possible to achieve convergence in the absorber column due to a minimum lean amine flow rate. At the maximum energy consumption the lean amine flow rate is 19 975 t/h, while it decreases as the number of stages are increased down to 3280 t/h at 15 stages in the absorber. This change is manifested in the lean/rich heat exchangers which has an installed cost of 733 MNOK at the extreme case of 10 stages, while it dramatically reduces as the amine flow is reduced when the number of stages is increased. At 15 stages the lean/rich heat exchanger is at its minimum with an installed cost of 207 MNOK.

The electricity consumption in the study 2A is at its minimum at 11 stages. For the whole range of parameters in this study, the difference from maximum to minimum electricity consumption is 16,4 MNOK per year. The two major cost drivers of electric power consumption is the flue gas transport fan and the amine circulation pumps.

For study 2C it was only possible to achieve a 90 % removal rate with the number of absorber stages in the range of 14 – 23 stages.

Like the study 2A and 2B, the overall function of NPV at 90 % removal efficiency has its maximum at the minimum number of stages. The highest NPV of 9838 MNOK is found at 14 stages in the absorber, but reduces as the number of stages is increased and it reaches its minimum at 19 stages with an NPV of 3910 MNOK. From the minimum number of stages, the NPV steadily increases as the number of stages is increased beyond this point. The effect of reducing the number of stages from 19 to 18 leads to an NPV increase of 26 MNOK. Likewise, the increase of one absorber stage from 19 to 20 stages leads to an NPV increase of 15 MNOK. It appears that the minimum NPV reaches as shallower minimum as the overall CO_2 removal grade is increased when changing the absorber packing height.

Also in study 2C the hot utility requirement is reduced as the number of stages is increased. The energy consumption reduces from its maximum of 14,3 MJ/kg CO_2 at 14 stages down to 3,51 MJ/kg CO_2 at 23 stages. It is noted that the overall energy consumption and NPV is higher for the 90 % removal efficiency than the 85 %. It was not possible to converge the absorber column with more than 23 stages due to minimum lean amine flow rate in the absorber. The extreme changes in hot utility requirements during the study are a consequence of the change in the loop flow of amine flow rate. At 14 stages the lean amine flow rate is 34 973 t/h, while it decreases with increase in number of stages in the absorber. At the maximum number of absorber stages, the lean amine flow rate is at its minimum of 3665 t/h.

A practical effect not showed here but which may affect real life performance is the effect of a maldistributed liquid phase. A maldistributed liquid phase may cause the stage efficiency to drop as much as 2-3 times [7]. This may indicate that the practical optimum for this specific case is an absorber with at a somewhat higher flow rate of amine for a given absorber packing height, this in order to add robustness to the simulated performance.

For full documentation of the process performance, parameters, OPEX and CAPEX for each step in study 2A, 2B and 2C, it is referred to Appendix 23 through Appendix 28.

6.4 Study 3 AB absorber feed gas temperature

The absorber temperature profile and estimated Murphree stage efficiencies for study 3A and 3B can be found in Figure 5-7 and Figure 5-9. The NPV and specific energy consumption for the same studies are shown in Figure 5-8 and Figure 5-10.

In these studies, the temperature of the flue gas entering the absorber column has been varied from 35^{o}C to 50^{o}C in steps of 5^{o}C by varying the circulation rate of CU in the DCC circulation loop. Due to a set temperature of 30^{o}C of the CU entering the DCC, it was not possible to achieve a temperature of 30^{o}C on the absorber feed gas temperature, and hence not tested for economical performance.

The two major cost drivers in these case studies are the amine loop on the one side and the cost related to the DCC cooler and column diameter on the other side. The general effect of increased inlet temperature is reduced size and duty of the DCC unit, but increased size of the absorber column and increased flow rate in the amine loop. Even though the temperature is increasing and thus increasing the stage efficiency, the flow of amine for a given number of stages has to increase as the flue gas temperature increases. As the absorber feed gas temperature increases, the actual gas flow rate increases. With the design assumption using constant vertical velocity in the absorber column, column size and installed cost will increase with increased volume flow.

For study 3A with a CO_2 removal efficiency of 85 %, the maximum temperature in the absorber typically occurs at the 9.-10. stage. In order to make this study converge for the specified range of flue gas inlet temperatures, the number of stages in the absorber column has been set to 12. Over the simulated feed gas temperatures of 35 – 50^{o}C, the NPV varies 124 MNOK from minimum to maximum. The minimum NPV occurs at a feed gas temperature of 40^{o}C, although the case of 35^{o}C inlet temperature is only 10 MNOK (0,3%) higher. The energy requirement increases with the feed gas temperature, and is 3,50 MJ/kg CO_2 at 35^{o}C and increases to 3,90 MJ/kg CO_2 at 50^{o}C inlet temperature.

Compared to the base case, this simulation of differentiated stage efficiencies seems to result in a less costly solution. Technically the solution in this study is somewhat different with regards of having fewer stages and a lower flow rate of amine but it is still able to achieve 85 % removal efficiency of the CO_2.

For study 3B with a CO_2 removal efficiency of 90 %, 16 stages in the absorber have been used during all feed gas inlet temperatures. Typically the maximum temperature occurs at the 12.-13. stage. When comparing study 3B to study 3A, it is apparent that the temperature development in the column for study 3B results in a higher temperature profile through the column stages. The calculated stage efficiencies are consequently somewhat higher in study

3B compared with study 3A. Even though the temperatures are increasing and thus increasing the stage efficiency, the flow of lean amine for a given number of stages has to increase as the flue gas temperature increases in order to maintain the specified overall CO_2 removal efficiency. The lean amine flow rate is at its lowest with an absorber feed gas temperature of 35^{o}C at 3530 t/h, while at an inlet temperature of 50^{o}C, the required lean amine flow rate is 4080 t/h.

Over the simulated feed gas temperatures of 35 – 50^{o}C, the NPV for study 3B varies 167 MNOK from minimum to maximum value, compared to 124 MNOK in study 3A for the same range of feed gas temperature. The minimum NPV occurs at a feed gas temperature of 35^{o}C, although the 40^{o}C inlet temperature is only 29 MNOK (0,8%) higher. This indicates that the absolute variations in NPV are considered not to vary significantly. It is possible that incremental steps of only one degree Celsius for the absorber feed gas temperature would have revealed a different NPV minimum. The energy requirement increases with the feed gas temperature, and is 3,42 MJ/kg CO_2 at 35^{o}C and up to 3,83 MJ/kg CO_2 at 50^{o}C inlet temperature. The specific energy requirement is actually higher in the 3A study with a lower overall CO_2 removal efficiency.

For full documentation of the process performance, OPEX and CAPEX for each step in study 3A and 3B, it is referred to Appendix 29 to Appendix 32.

6.5 Study 4A uptime and calculation period

From Figure 5-11 it can be seen that the cost per ton CO_2 removed is highly dependent on the basis for the NPV calculations. The process performance and physical capacities for the base case process itself is constant for all cases, i.e. the installed equipment cost for all cases are constant. The annual utility costs are the same for a given uptime.

The predominant trend that the specific CO_2 removal cost is reduced the more the plant is utilized, distributing the initial investment cost over a longer time period and consequently more captured CO_2. The NPV increases with accumulated operation hours, with the lowest value at an uptime of 4000 h/yr and a calculation period of 10 years and up to maximum NPV with an uptime of 8000 h/yr and a calculation period of 20 years. An interesting case is for the two cases with the same accumulated number of operational hours and hence total mass of CO_2 captured. This is the case with a calculation period of 10 and 20 years, with an annual uptime of 8000 and 4000 hours, respectively. The latter has the lowest NPV, this is due the discount effect on the halved annual operational costs, while the nominal cost per ton CO_2 is the same due to the calculation method.

The study shows that the choice of calculation period and plant utilization (uptime) is crucial for the determination of the NPV, but probably also when optimizing the process variables.

This is because the process variables affect the size of the equipment cost and the annual operational costs.

6.6 Study 5A amine fluid package

From Table 5-3, the changes in process performance and costs when changing the amine fluid package from the Kent-Eisenberg to the Li-Mather liquid calculation method can be seen. The lean amine flow rate had to be increased from 3600 to 3800 t/h in order to obtain the base case CO_2 removal efficiency. The specific energy consumption is consequently increased by 2,2 % from 3,61 to 3,69 MJ/kg CO_2. This is seen in connection with increased steam consumption and increase of the annual operational cost by 1,5 %.

There is also some slight change in installed equipment cost, mostly related to the increased flow rate of lean amine. The NPV for the Li-Mather case increases by 1,2 % compared to the base case with the Kent-Eisenberg fluid package.

These results indicate that there is some sensitivity with regards to the simulated process performance, and consequently to the estimated cost due to the selection of fluid package. This change with regards to the overall cost however, is not considered significant as the uncertainties in the base cost probably are somewhat higher.

6.7 Comparison to earlier simulations

6.7.1 Optimum ΔT_{min} in l/r heat exchanger

In this study, optimum ΔT_{min} was found to give a minimum NPV in the range of 10-14 K at an overall CO_2 removal efficiency of 85%. Earlier work by Blaker [11] has found this to be 19 K by manual case studies, and 17,7 K by automated optimization. Although the calculation assumptions for NPV are not comparable, it is indicated that the future of full scale CO_2 capture plants are plants with lower external utility demand (lower specific energy consumption). One way of achieving this is to increase the internal heat recovery in the capture plant. A lower ΔT_{min} indicates a higher degree of heat recovery. Higher heat recovery may also make the plant less sensitive to fluctuations in fuel prices. Other the other hand, a too low value would lead to very low driving forces which again would lead to unreasonably large heat exchanger area. It is reasonable that the ΔT_{min} should be in the area 10-20 K, but as indicated in study 4A, this is highly dependent on the planned utilization of the plant reflected in the assumptions for calculation period and annual uptime.

6.7.2 Optimum absorber packing height

At a removal efficiency of 85 %, the optimum number of stages was found to be 15 meter of packing. Earlier work by Blaker [11] has found this to be 16 meters. There is expected to be large uncertainties in the cost estimates for structured packing, however it is reasonable that the optimum number of stages is the area of 15-16 meters. Experimental work regarding liquid distribution and robustness of process solution should be done in with the basis of these simulations.

6.7.3 Optimum absorber feed gas temperature

In this study, optimal absorber feed gas temperature was found to be 40^{o}C with a removal efficiency of 85 %. The study at 90 % showed 35^{o}C. Both studies show little change in absolute cost between 35 and 40^{o}C, which suggest that optimum may be somewhere in between the two values. In the master thesis of Blaker optimization found this to be 35^{o}C [11]. It is plausible that the optimum feed temperature is in the area 35-40^{o}C. The lower feed gas temperature is beneficial with regards to the significant absorber column cost and the overall absorption performance. Optimal temperature is also assumed dependent on available cold utility temperature, determining necessary flow rates of circulation water and cooler size requirement.

6.8 Accuracy and uncertainties

In the report there are several factors which lead to uncertainties, both related to the simulation, the dimensioning and cost estimation assumptions:

- In the cost estimation of process equipment, large uncertainties are expected related to the base equipment cost and validity of the scale up factors in the specified range. Especially for the large cost items in the capture plant like the structured packing for the absorber and desorber columns is this considered large sources of uncertainties. Correct prices for heat exchanger area are also expected to be important in order to reduce the uncertainties.
- The high energy consumption of the CO_2 capture process inevitably leads to sensitivity to energy prices like oil and electricity. The fluctuation in fuel price may be larger than the uncertainty of the energy consumption calculated in Hysys. Lowering the energy consumption also reduces the impact of the fluctuation in price of energy.
- Applicability of the simulated results to a full scale plant with regards to physical phenomena:
    - Uncertainty with regards to how a full scale liquid distribution affects the optimum number of stages and lean amine flow rate. It is difficult to verify if the simulated optimum stage height at a given lean amine flow rate is robust

enough to meet the full-scale performance specifications. As the major costs are in the absorber column, this is assumed to be important with regards to how valid the simulation is. Even thought the process equipment used in the capture plant are widely used in other parts of chemical processing, the large scale of these systems represent risk factors with regards to the scale up performance.

  - The pressure losses in the system are based on assumptions and some calculations. The one major operational cost with regards to pressure loss in the capture plant are related to the flue gas transport fan. The fan outlet pressure is based on assumed pressure drop in the downstream absorber. This is expected to represent large uncertainties with regards to actual pressure drop and consequently the actual electricity consumption. The pressure drop is also expected to vary with lean amine flow rate in the column. The pressure losses in the amine loop are assumed to be minor in this context.
  - As temperatures in the desorption section of the capture plant is expected to be higher than ambient temperature, heat loss is also assumed to occur which may affect utility consumption in practice. This source of error is not expected to be very significant, but can be reduced by increased insulation of equipment. This will however represent additional installation costs.
  - The MEA solution combined with high process temperatures is expected to set limitations to choice of materials. This may affect cost if special materials or specialized equipment design have to be used.
- There is also assumed to be uncertainties in equilibrium calculations in the absorber and desorber column. The study of changing the amine fluid package in Hysys showed only minor difference in both energy consumption and cost. Aspen Hysys calculates equilibrium between vapor and liquid. Other simulation programs like Aspen Plus have models able to combine both equilibrium models with reaction kinetic models which could simulate the practical performance even better giving lower uncertainties.
- As stated in several earlier studies, convergence is difficult to achieve, especially in the absorber and desorber, this made it not possible to fully automate the case studies. In the studies, default convergence criteria have been used. Higher accuracy is expected if these are reduced. Also, because of frequent problem of achieving convergence, an overall optimization for all process parameters in order to find the least costly solution by simultaneously changing several process parameters has not been performed.
- Operational costs are assumed to be well documented in this thesis. Additional operational costs not covered, which may be case specific, are assumed to be costs related to:
  - Removal of impurities in the amine loop

  - Cost for make-up MEA and water. Loss of amine to atmosphere and degradation is assumed to be a function of amine flow rate, and hence should be relatively more dominant in high flow rate amine cases
  - Maintenance - this is often included as a percentage of equipment installed costs, and should be a dominant factor in process solutions with high equipment costs

6.9 Further work

Further work should seek to quantify and minimize the sources of error in order to increase accuracy in the simulations and cost estimations:

- Study the likelihood and consequences of maldistribution when using structured packing in the absorber and desorber columns
- Acquirement of improved basis for cost estimation, especially for the most significant equipment like:
  - Structured packing
  - Heat exchangers
- Improved heat integration in the capture plant
- Literature studies and choice of new absorbents with lower regeneration heat requirement for further simulation
- Choice of new materials for process equipment so that they are capable of handling more corrosive absorbents

7 Conclusion

The work in this thesis is a continuation of earlier work by students at Telemark University College (TUC) of CO_2 capture simulation and cost estimation in Aspen Hysys.

A Hysys simulation of a CO_2 capture process by absorption in a monoethanol amine (MEA) solution from the flue gas from a 500 MW natural gas power plant has been developed as a verification of earlier simulations at TUC. The major improvements in this work are new calculation methods for make-up water and MEA and simulation of a direct contact cooler (DCC) unit. For cost estimation purposes, calculations of overall heat transfer coefficient and correction factor for heat exchangers have been performed.

On the basis of the base case simulation output, installed cost estimates for equipment have been made. Only equipment related to flue gas cooling and the CO_2 absorption and regeneration process have been included in the simulation and cost estimation scope. Variation in cost changes has been monitored when changing process parameters like minimum approach temperature in the lean/rich heat exchanger, absorber packing height, absorber gas feed temperature. The parametric studies have been performed for CO_2 removal efficiencies of 80, 85 and 90 %. In most of the calculations, one meter of packing was specified with a Murphree efficiency of 0,15. When optimizing feed gas temperature, a temperature dependent efficiency was used.

The base case with an CO_2 removal efficiency of 85 % has been estimated with a specific energy consumption of 3,61 MJ/kg CO_2, and equipment installed cost is estimated to 1400 MNOK. The annual operational utility cost has been found to be 203 MNOK, where 61 % is related to steam consumption in the desorber reboiler. The amine package in Aspen Hysys with Kent Eisenberg was used. The Li-Mather model was checked for comparison with the base case, this resulted in a 1,5 % increase in the annual operational utility cost and 0,8 % in the equipment installed cost.

Parametric studies at a CO_2 removal efficiency of 85 % have resulted in optimum minimum approach temperature in the lean/rich heat exchanger between 10-14 K, absorber packing height 15 m, and absorber feed gas temperature approximately 40 ^{0}C. At 90 % efficiency the effect of varied process parameters is greater then at 85 %. Economic parameters like uptime and calculation period also influence on the optimum parameters.

Capture of CO_2 post combustion by absorption in amines has high operational costs. The process equipment used are known equipment and therefore the scale up of CO_2 capture is considered technically possible, however, the high energy consumption makes it sensitive to changes in utility costs. The greatest uncertainties for CO_2 capture are assumed to be in the overall utilization of the plant, cost of the utilities and the fact that there are no plants were the initial lessons can be learnt. As economic parameters are assumed to influence the optimum process parameters, any process optimization on a full scale plant should be made on a good estimation of actual lifetime and uptime basis for that particular plant.

This study shows how significant process parameters are to overall cost of CO_2 capture. Major improvements in cost savings can be made by optimization. Aspen Hysys is a suitable tool for such calculations.

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Appendices

Appendix 33 Case study 4A: Process overview

Appendix 34 Case study 5A: Process overview

Appendix 35 Case study 5A: Cost overview

Appendix 1 Master thesis description

Høgskolen i Telemark
Avdeling for teknologiske fag

FMH606 Master's Thesis

Cost estimation and optimization of CO_2 removal

Ove Braut Kallevik

Responsible professor (at HiT): Lars Erik Øi

Aims:

Improve models for calculation, equipment dimensioning and cost estimation of CO_2 removal from atmospheric exhaust gas.

Extend models to include possibilities for cost optimization (minimization of the sum of capital and operation cost) of different process parameters.

Tasks:

1. Evaluation of earlier projects on process simulation and cost estimation of CO_2 removal plants.

2. Calculations of CO_2 removal with absorption and desorption in an amine solution using Aspen HYSYS. Calculations of dependencies of different removal efficiencies, CO_2 concentrations, equipment dimensions and other assumptions.

3. Extend the models to include possibilities for cost optimization (minimization of the sum of capital and operating cost).

4. Calculate cost optimum values of parameters like exhaust inlet temperature, removal grade, absorber column height, circulating rate and heat exchanger temperature differences.

5. Evaluation of uncertainties in the calculations.

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Appendix 1 Master thesis description

Background:

The most actual method for removal of CO_2 from atmospheric exhaust is by the help of amine solutions. Cost estimates for CO_2 removal from gas based power plants have been made for Kårstø, Tjeldbergodden, Mongstad and Porsgrunn. The details in these cost estimates are normally not open available.

HYSYS has been much used in student projects at Telemark University College for process simulation and as a basis for dimensioning and cost estimation of CO_2 removal. However, these calculations have large uncertainties. Telemark University College has collaborated with different companies (TelTek, Hydro, Aker Kværner, Norcem and Skagerak) which work with plans for CO_2 removal. In autumn 2007, 2008 and 2009, Master student projects have resulted in an improved model for cost estimation using Aspen HYSYS.

Practical arrangements:

The working place will be HiT.

Signatures:

Student (date and signature): 01.05.10

Supervisor (date and signature): 1/3-10

Document name: d10Ove1/1.3.2010

Appendix 2 Relevant formulas

1. Pump power equation

$$P[W] = \frac{\rho[[{}^{kg}/_{m^3}]g[{}^{m}/_{s^2}]H[m]Q[{}^{m^3}/_{s}]}{\eta[-]}$$ Eq. 23.8 [8]

2. LMDT calculation

$$LMDT[K] = \frac{\Delta T_{Hot\,side}[K] - \Delta T_{Cold\,side}[K]}{ln\frac{\Delta T_{Hot\,side}}{\Delta T_{Cold\,side}}}$$ Eq. 15.46 [9]

3. Souder Brown equation

$$v[{}^{m}/_{s}] = K[{}^{m}/_{s}]\sqrt{\frac{\rho_{liquid}[{}^{kg}/_{m^3}] - \rho_{gas}[{}^{kg}/_{m^3}]}{\rho_{gas}[{}^{kg}/_{m^3}]}}$$ [20]

4. Heat exchanger equation

$$Q[W] = U\left[{}^{W}/_{m^2K}\right]A[m^2]F[-]LMDT[K]$$ Eq. 11.9 [21]

Appendix 3 Process equipment base cost

18 *Process Economics*

Table 2.1 Typical equipment capacity delivered capital cost correlations.

Equipment	Material of construction	Capacity measure	Base size Q_B	Base cost C_B ($)	Size range	Cost exponent M
Agitated reactor	CS	Volume (m^3)	1	1.15×10^4	1–50	0.45
Pressure vessel	SS	Mass (t)	6	9.84×10^4	6–100	0.82
Distillation column (Empty shell)	CS	Mass (t)	8	6.56×10^4	8–300	0.89
Sieve trays (10 trays)	CS	Column diameter (m)	0.5	6.56×10^3	0.5–4.0	0.91
Valve trays (10 trays)	CS	Column diameter (m)	0.5	1.80×10^4	0.5–4.0	0.97
Structured packing (5 m height)	SS (low grade)	Column diameter (m)	0.5	1.80×10^4	0.5–4.0	1.70
Scrubber (Including random packing)	SS (low grade)	Volume (m^3)	0.1	4.92×10^3	0.1–20	0.53
Cyclone	CS	Diameter (m)	0.4	1.64×10^3	0.4–3.0	1.20
Vacuum filter	CS	Filter area (m^2)	10	8.36×10^4	10–25	0.49
Dryer	SS (low grade)	Evaporation rate (kg $H_2O \cdot h^{-1}$)	700	2.30×10^5	700–3000	0.65
Shell-and-tube heat exchanger	CS	Heat transfer area (m^2)	80	3.28×10^4	80–4000	0.68
Air-cooled heat exchanger	CS	Plain tube heat transfer area (m^2)	200	1.56×10^5	200–2000	0.89
Centrifugal pump (Small, including motor)	SS (high grade)	Power (kW)	1	1.97×10^3	1–10	0.35
Centrifugal pump (Large, including motor)	CS	Power (kW)	4	9.84×10^3	4–700	0.55
Compressor (Including motor)		Power (kW)	250	9.84×10^4	250–10,000	0.46
Fan (Including motor)	CS	Power (kW)	50	1.23×10^4	50–200	0.76
Vacuum pump (Including motor)	CS	Power (kW)	10	1.10×10^4	10–45	0.44
Electric motor		Power (kW)	10	1.48×10^3	10–150	0.85
Storage tank (Small atmospheric)	SS (low grade)	Volume (m^3)	0.1	3.28×10^3	0.1–20	0.57
Storage tank (Large atmospheric)	CS	Volume (m^3)	5	1.15×10^4	5–200	0.53
Silo	CS	Volume (m^3)	60	1.72×10^4	60–150	0.70
Package steam boiler (Fire-tube boiler)	CS	Steam generation ($kg \cdot h^{-1}$)	50,000	4.64×10^5	50,000–350,000	0.96
Field erected steam boiler (Water-tube boiler)	CS	Steam generation ($kg \cdot h^{-1}$)	20,000	3.28×10^5	10,000–800,000	0.81
Cooling tower (Forced draft)		Water flowrate ($m^3 \cdot h^{-1}$)	10	4.43×10^3	10–40	0.63

CS = carbon steel; SS (low grade) = low-grade stainless steel, for example, type 304; SS (high grade) = high-grade stainless steel, for example, type 316

The table can be found on page 18 in [9].

Appendix 4 Currency factors and cost index

Currency exchange rates for equipment cost estimates [22]

Year	Currency conversion [NOK/$]
1990	6,25
2000	8,81

CEPCI index for equipment cost estimates [23]

Year	CEPCI [-]
1990	357,7
2000	394,1
2009	616

Appendix 5 Installation cost factors

2009/2010 Cost of equipment in Carbon Steel (CS)	Fluid 0-20	20-100	100-500	500-1000	1000-2000	2000-5000	5000-15000	>15000	Solid 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	0.68	0.36	0.19	0.14	0.11	0.09	0.08	0.05	1.50	0.80	0.46	0.33	0.27	0.19	0.16
Piping	2.71	1.46	0.85	0.62	0.50	0.37	0.31	0.23	0.55	0.30	0.16	0.13	0.10	0.08	0.06
Electric	0.79	0.54	0.37	0.31	0.26	0.22	0.19	0.14	1.32	0.83	0.55	0.43	0.36	0.30	0.25
Instrument	2.71	1.46	0.85	0.62	0.50	0.37	0.31	0.23	1.08	0.59	0.34	0.25	0.20	0.14	0.12
Civil work	0.42	0.27	0.19	0.15	0.13	0.11	0.10	0.06	0.96	0.57	0.37	0.28	0.23	0.18	0.15
Steel & concrete	1.37	0.89	0.60	0.48	0.42	0.33	0.30	0.22	1.91	1.18	0.78	0.60	0.51	0.40	0.36
Insulation	0.52	0.26	0.14	0.11	0.09	0.06	0.04	0.03	0.52	0.26	0.14	0.11	0.09	0.06	0.04
Direct Cost	10,26	6,32	4,29	3,53	3,07	2,61	2,39	2,02	8,91	5,61	3,88	3,22	2,81	2,42	2,22
Engineering Process	0.94	0.33	0.18	0.14	0.12	0.10	0.09	0.06	0.94	0.33	0.18	0.14	0.12	0.10	0.09
Engineering Mechanical	0.74	0.18	0.08	0.04	0.03	0.02	0.01	0.01	0.94	0.28	0.13	0.09	0.06	0.04	0.03
Engineering Piping	0.82	0.44	0.26	0.19	0.14	0.11	0.10	0.06	0.16	0.09	0.04	0.03	0.02	0.02	0.02
Engineering Electric	0.80	0.24	0.12	0.09	0.08	0.06	0.04	0.03	0.93	0.31	0.15	0.12	0.10	0.08	0.06
Engineering Instrument	1.41	0.55	0.27	0.19	0.15	0.11	0.10	0.06	0.92	0.27	0.12	0.09	0.06	0.04	0.03
Engineering Civil	0.30	0.09	0.03	0.02	0.02	0.01	0.01	0.01	0.38	0.13	0.06	0.04	0.03	0.02	0.02
Engineering Steel & Concrete	0.44	0.18	0.10	0.08	0.06	0.04	0.04	0.03	0.52	0.22	0.12	0.10	0.09	0.06	0.06
Engineering Insulation	0.20	0.06	0.02	0.01	0.01	0.01	0.01	0.01	0.20	0.06	0.02	0.01	0.01	0.01	0.01
Engineering Cost	5,64	2,05	1,06	0,75	0,61	0,46	0,40	0,29	4,98	1,68	0,83	0,61	0,48	0,38	0,33
Procurement	1.18	0.40	0.15	0.10	0.06	0.03	0.02	0.02	1.18	0.40	0.15	0.10	0.06	0.03	0.02
Project Control	0.28	0.11	0.04	0.03	0.03	0.02	0.02	0.02	0.25	0.09	0.04	0.03	0.02	0.02	0.01
Site Management	0.51	0.32	0.22	0.18	0.15	0.13	0.12	0.09	0.43	0.27	0.19	0.15	0.14	0.12	0.12
Project management	0.68	0.34	0.23	0.18	0.15	0.13	0.12	0.09	0.58	0.30	0.19	0.15	0.13	0.12	0.11
Administration Cost	2,65	1,17	0,64	0,48	0,40	0,32	0,28	0,20	2,44	1,06	0,58	0,44	0,36	0,30	0,26
Commissioning	0.55	0.25	0.13	0.08	0.08	0.04	0.04	0.03	0.47	0.23	0.12	0.09	0.06	0.04	0.03
Total Known Cost	19,09	9,80	6,11	4,85	4,16	3,44	3,12	2,55	16,80	8,57	5,41	4,35	3,72	3,14	2,84
Contingency	3.79	1.96	1.23	0.98	0.84	0.71	0.64	0.46	3.28	1.68	1.08	0.86	0.74	0.64	0.58
Total Cost	22,88	11,75	7,33	5,83	5,00	4,15	3,77	3,02	20,07	10,25	6,48	5,21	4,46	3,78	3,42

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 23/11 2009

Nils Henrik Eldrup

Appendix 6 Dimensioning of direct contact cooler (1 of 3)

DCC VESSEL

Dimensioning		**Note**
Gas flow [Am3/h]	2999405	
Gas flow [Am3/s]	833	
Liquid mass density [kg/m3]	1000	
Gas phase mass density [kg/m3]	1,02	
K factor	0,15	Assumed
L/D ratio	1	Assumed
Vertical velocity [m/s]	4,7	Sounder-Brown eq.
Vessel diameter [m]	15,1	
Vessel length [m]	15,1	
Wall thickness [m]	0,01	Assumed
Density SS [kg/m3]	8000	Approx value
Absorber shell volume [m3]	7,1	
Mass of absorber [tons]	56,9	
Contact medium height [m]	3,0	
Contact medium volume [m3]	533,7	

Base cost estimate DCC shell [$]	65600	[9], table 2.1 CS,$@2000,
Base size [ton]	8	8-300 t
Scaling factor	0,89	[9], table 2.1
Base cost estimate random packing [$/m3]	3046	Pall ring 2" [7], CS $@1990

NOK/$ @1990	6,25	[22]
NOK/$ @ 2000	8,81	[22]
CE index 1990	357,7	[23]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost absorber shell	5,2
Cost packing	15,9

Cold utility circulation pump

Adiabatic efficiency [%]	75	Assumed
Duty [kW]	437	

Base cost estimate one pump unit with motor [$]	9840	[9], table 2.1
Base size [kW]	4	4-700 kW
Scaling factor	0,55	[9], table 2.1

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost circulation pump	1,8

Appendix 6 Dimensioning of direct contact cooler (2 of 3)

DCC COOLER

	Fluid 1 HOT	Fluid 2 COLD	Note
	TUBE	SHELL	
Stream name	**CU-3**	**CU-5**	
Flowrate [m3/s]	1,8	1,9	
Initial temp [degC]	40,7	15,0	
Final temp [degC]	30	25	
Density[kg/m3]	992	1015	
Heat capacity[J/kgK]	4216	4316	
Viscosity[Pas]	6,31E-04	1,15E-03	
Thermal conductivity [W/mK]	0,64	0,60	
Flow velocity [m/s]	1	1	Assumed, tube 1-3, shell 1-2 m/s

Exchanger geometry assumptions			
Tube inner D [m]	1,60E-02		Assumed
Tube outer D [m]	2,00E-02		Assumed
Tube pitch [factor * Tube outer]	1,25		Assumed
Pitch configuration factor [-]	0,866	triangular	Assumed
Baffle cut [fraction height]	0,25		Assumed
Length/Shell diameter - ratio	6	5-10	Assumed
Fraction of thermal effectivity 1T-2S	0,9		Assumed

Dimensioning

R	1,08		Eq. 15.49 [9]
P	0,39		Eq. 15.50 [9]
W	0,92		
Number of shells	0,60	1	Eq 15.62 [9]
Correction factor	0,92	Preferably Ft>0,75	Eq. 15.51 [9]
LMDT, uncorrected [K]	15,4		15.46 [9]
Min. approach, [K]	15		
Prandl number [-]	4,2		
Constant heat transfer, pipe [-]	4907		Eq. 15.17 [9]
Tube inner heat transfer coeff [W/m2K)	4907		15.14 [9]
Fhn	1	Assumed	Correction factor, number of tube rows crossed Pg. 665 [9]
Fhw	1	Assumed	Correction factor, baffle window Page 665 [9]
Fhb	0,8	Assumed	Correction factor, bypass stream factor Page 665 [9]
Fhl	0,8	Assumed	Correction factor, leakage Page 665 [9]
Constant heat transfer, shell [-]	4849		Eq. 15.24 [9]
Shell outer heat transfer coeff [W/m2K)	4849		Eq. 15.21 [9]
Fouling coefficient tubeside [W/m2K]	5000	Aqueous	Table 15.2 [9]
Fouling coefficient shellside [W/m2K]	10000	Aqueous	Table 15.2 [9]
Thermal conductivity tube material [W/mk]	16	Stainless	Table 15.4 [9]
1/Uoverall [m2K/W]	9,50E-04		Eq. 15.13 [9]
Uoverall [W/m2K]	1052		
Pressure loss calculation, tube side			
Kpt1	1,59E-13		Eq. 15.19 [9]
Kpt2	1,47E-06		Eq. 15.20 [9]
Pressure loss tube side per unit [bar]	9,71E-02		Eq. 15.16 [9]
Pressure loss calculation, shell side			
Fpb	0,5	Assumed	Correction factor pressuredrop shell-bundle clearance [9]
Fpl	0,5	Assumed	Corr. factor leakage baffle/tube-to-shell clearance [9]
Kps1	7161,69		Eq. 15.25 [9]
Kps2	1,47		Eq. 15.26 [9]
Kps3	5391,59		Eq. 15.27 [9]
Kps4	2,06		Eq. 15.28 [9]
Pressure loss shell side per unit [bar]	0,28		Eq. 15.22 [9]

Appendix 6 Dimensioning of direct contact cooler (3 of 3)

Heat exchanger summary		
Actual duty [MW]	81,1	
Corrected LMDT [K]	14	
Uoverall [W/m2K]	1052	
Correction factor	0,92	
Uncorrected LMDT	15,40	
F*U	968,3	
Assumed constant F*U [W/m2K]	**960**	Assumed
Total heat exchanger area required [m2]	5483	
Number of shells	1	
Heat exchanger area per shell [m2]	5483	0-4500 m2
Estimated shell diameter	2,2	Eq 15.38 [9]
Estimated Shell length	13	
Summary pressure losses		
Pressure loss all units in series, tube side [bar]	9,71E-02	
Pressure loss all units in series, shell side [bar]	0,28	

Base cost estimate heat exchanger [$]	32800	[9], table 2.1
Base size [m2]	80	80-4000 m2
Scaling factor	0,68	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
DCC CU cooler 1	8,0

Appendix 7 Dimensioning of absorber column (1of 1)

ABSORBER

Dimensioning		**Note**
Flue Gas Volume Flow [Am3/h]	2360705	
Flue Gas Volume Flow [Am3/s]	656	
Allowable gas velocity in absorber [m/s]	3,6	[3]
Absorber diameter [m]	15,2	Not considering liquid
Packing height [m]	16	
Total packing volume [m3]	2914,5	
Absorber total height [m]	50	Assumed
Wall thickness [m]	1,00E-02	Assumed
Density SS [kg/m3]	8000	Approx value
Absorber shell volume [m3]	23,9	
Mass of absorber [tons]	198	

Base estimate		
Base cost estimate absorber shell [$]	65600	[9], table 2.1, CS, 2000
Base size [ton]	8	8-300 t
Scaling factor	0,89	[9], table 2.1
Base cost estimate structured packing [$/m3]	4264	Assumed 140 % 2" Pall Rings, SS, 1990

Currency and cost index		
NOK/$ @1990	6,25	[22]
NOK/$ @ 2000	8,81	[22]
CE index 1990	357,7	[23]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost absorber shell	15,7
Cost packing	133,9

Appendix 8 Dimensioning of l/r heat exchangers (1of 2)

LEAN/RICH HEAT EXCHANGERS

Heat exchanger input

	Fluid 1 HOT	Fluid 2 COLD	**Note**
	TUBE	SHELL	
Stream name	**14-Lean amine**	**6-Rich amine**	
Flowrate [m3/s]	0,98	0,97	
Initial temp [degC]	120	44	
Final temp [degC]	54	109	
Density[kg/m3]	1033	1016	
Heat capacity[J/kgK]	3751	3818	
Viscosity[Pas]	0,0006	0,0007	
Thermal conductivity [W/mK]	0,53	0,53	
Flow velocity [m/s]	1	1	Assumed, tube 1-3, shell 1-2 m/s
Physical properties at film temperature	87	77	

Exchanger geometry assumptions		**Note**	
Tube inner D [m]	1,60E-02		Assumed
Tube outer D [m]	2,00E-02		Assumed
Tube pitch [factor * Tube outer]	1,25		Assumed
Pitch configuration factor [-]	0,866	Triangular config	Assumed
Baffle cut [fraction height]	0,25		Assumed
Length/Shell diameter - ratio	6	Typically 5-10	Assumed
Fraction of thermal effectivity 1T-2S	0,9		Assumed

Dimensioning

R	1,01			Eq. 15.49 [9]
P	0,86			Eq. 15.50 [9]
W	0,99			
Number of shells	5,69		6	Eq 15.62 [9]
Correction factor		Too low Ft		Eq. 15.51 [9]
Recaluclation, IF				
Z	0,94			Eq 15.57 [9]
Pnew	0,51			Eq 15.57 [9]
New correction factor	0,77			Eq. 15.51 [9]
LMDT, uncorrected [K]	10,3			Eq. 15.46 [9]
Min. approach, [K]	10,0			
Prandl number [-]	3,95			
Constant heat transfer, pipe [-]	4557			Eq. 15.17 [9]
Tube inner heat transfer coeff [W/m2K]	4557			Eq. 15.14 [9]
Fhn	1	Assumed		Correction factor, no of tube rows crossed [9]
Fhw	1	Assumed		Correction factor, baffle window [9]
Fhb	0,8	Assumed		Correction factor, bypass stream factor [9]
Fhl	0,8	Assumed		Correction factor, leakage [9]
Constant heat transfer, shell [-]	5100			Eq. 15.24 [9]
Shell outer heat transfer coeff [W/m2K]	5100			Eq. 15.21 [9]

Appendix 8 Dimensioning of l/r heat exchangers (2 of 2)

Fouling coefficient tubeside [W/m2K]	5000	Aqueous salt	Table 15.2 [9]
Fouling coefficient shellside [W/m2K]	5000	Aqueous salt	Table 15.2 [9]
Thermal conductivity tube material [W/mK]	16	Stainless steel	Table 15.4 [9]
1/Uoverall [m2K/W]	1,06E-03		Eq. 15.13 [9]
Uoverall [W/m2K]	944		
Pressure loss calculation, tube side			
Kpt1	3,87E-13		Eq. 15.19 [9]
Kpt2	1,84E-06		Eq. 15.20 [9]
Pressure loss tube side per unit [bar]	0,157		Eq. 15.16 [9]
Pressure loss calculation, shell side			
Fpb	0,5	Assumed	Correction factor pressuredrop shell-bundle clearance, app. C [9] page 663
Fpl	0,5	Assumed	Corr. factor leakage baffle/tube-to-shell clearance, app. C [9] page 663
Kps1	7183		Eq. 15.25 [9]
Kps2	2,57		Eq. 15.26 [9]
Kps3	4848		Eq. 15.27 [9]
Kps4	3,98		Eq. 15.28 [9]
Pressure loss shell side per unit [bar]	0,44		Eq. 15.22 [9]

Heat exchanger summary

Actual duty [MW]	246,5	
Corrected LMDT [K]	8,0	
Uoverall [W/m2K]	944	
Correction factor	0,77	
Uncorrected LMDT	10,3	
F*U	728,8	
Assumed constant F*U [W/m2K]	**750**	
Total heat exchanger area required [m2]	31814	
Number of shells	6	
Heat exchanger area per shell [m2]	5302	0-4500 m2
Estimated shell diameter	2,1	Eq 15.38 [9]
Estimated Shell length	12,8	
Summary pressure losses		
Pressure loss all units in series, tube side [bar]	0,94	
Pressure loss all units in series, shell side [bar]	2,65	

Base estimate

Base cost estimate heat exchanger [$]	32800	[9], table 2.1, CS, $@2000,
Base size [m2]	80	[9], table 2.1
Scaling factor	0,68	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
HA-1 L/R amine	7,8
HA-2 L/R amine	7,8
HA-3 L/R amine	7,8
HA-4 L/R amine	7,8
HA-5 L/R amine	7,8
HA-6 L/R amine	7,8

Appendix 9 Dimensioning of lean amine heat exchanger (1 of 2)

LEAN AMINE COOLER

Heat exchanger input

	Fluid 1 HOT	Fluid 2 COLD	Note
	TUBE	SHELL	
Stream name	**19-Lean amine**	**CU-11**	
Flowrate [m3/s]	1,0	1,0	
Initial temp [degC]	53,0	15,0	
Final temp [degC]	40,0	25,1	
Density[kg/m3]	1052,7	1011,1	
Heat capacity[J/kgK]	3568,9	4314,3	
Viscosity[Pas]	0,0013	0,0010	
Thermal conductivity [W/mK]	0,50	0,60	
Flow velocity [m/s]	1,0	1,0	Assumed, tube 1-3, shell 1-2 m/s
Physical properties at film temperature [degC]	46,5	20,1	

Exchanger geometry assumptions		Note	
Tube inner D [m]	1,60E-02		Assumed
Tube outer D [m]	2,00E-02		Assumed
Tube pitch [factor * Tube outer]	1,25		Assumed
Pitch configuration factor [-]	0,866	Triangular config	Assumed
Baffle cut [fraction height]	0,25		Assumed
Length/Shell diameter - ratio	6	5-10	Assumed
Fraction of thermal effectivity 1T-2S	0,9		Assumed

Dimensioning

R	1,28		Eq. 15.49 [9]
P	0,27		Eq. 15.50 [9]
W	0,76		
Number of shells	0,39	1	Eq. 15.62 [9]
Correction factor	0,97	Preferably Ft>0,75	Eq. 15.51 [9]
LMDT, uncorrected [K]	26,4		Eq. 15.46 [9]
Min. approach, [K]	25,0		
Prandl number [-]	9,4		
Constant heat transfer, pipe [-]	2934,9		Eq. 15.17 [9]
Tube inner heat transfer coeff [W/m2K]	2934,9		Eq. 15.14 [9]
Fhn	1	Assumed	Correction factor, no of tube rows crossed [9]
Fhw	1	Assumed	Correction factor, baffle window [9]
Fhb	0,8	Assumed	Correction factor, bypass stream factor [9]
Fhl	0,8	Assumed	Correction factor, leakage [9]
Constant heat transfer, shell [-]	5078,6		Eq. 15.24 [9]
Shell outer heat transfer coeff [W/m2K]	5078,6		Eq. 15.21 [9]
Fouling coefficient tubeside [W/m2K]	5000	Aqueous salt	Table 15.2 [9]
Fouling coefficient shellside [W/m2K]	10000	Aqueous salt	Table 15.2 [9]

Appendix 9 Dimensioning of lean amine heat exchanger (2 of 2)

Thermal conductivity tube material [W/mk]	16	Stainless steel	Table 15.4 [9]
1/Uoverall [m2K/W]	1,11E-03		Eq. 15.13 [9]
Uoverall [W/m2K]	899		
Pressure loss calculation, tube side			
Kpt1	2,17E-12		Eq. 15.19 [9]
Kpt2	5,64E-06		Eq. 15.20 [9]
Pressure loss tube side per unit [bar]	8,95E-02		Eq. 15.16 [9]
Pressure loss calculation, shell side			
Fpb	0,5	Assumed	Correction factor pressuredrop shell-bundle clearance, app. C [9] page 663
Fpl	0,5	Assumed	Corr. factor leakage baffle/tube-to-shell clearance, app. C [9] page 663
Kps1	6225		Eq. 15.25 [9]
Kps2	3		Eq. 15.26 [9]
Kps3	3836		Eq. 15.27 [9]
Kps4	4		Eq. 15.28 [9]
Pressure loss shell side per unit [bar]	0,23		Eq. 15.22 [9]

Heat exchanger summary		
Actual duty [MW]	47,8	
Corrected LMDT [K]	25,6	
Uoverall [W/m2K]	899	
Correction factor	0,97	
Uncorrected LMDT	26,41	
F*U	870	
Assumed constant F*U [W/m2K]	850	Assumed
Total heat exchanger area required [m2]	2130	
Number of shells	1	
Heat exchanger area per shell [m2]	2130	0-4500 m2
Estimated shell diameter	1,57	Eq 15.38 [9]
Estimated Shell length	9,44	
Summary pressure losses		
Pressure loss all units in series, tube side [bar]	8,95E-02	
Pressure loss all units in series, shell side [bar]	0,23	

Base estimate

Base cost estimate heat exchanger [$]	32800	[9], table 2.1 CS,
Base size [m2]	80	[9], table 2.1
Scaling factor	0,68	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
HA-1 Lean amine cooler	4,2

Appendix 10 Dimensioning of lean and rich amine pumps (1 of 1)

AMINE PUMPS

Rich amine pump

Dimensioning		**Note**
Adiabatic efficiency [%]	75	
Duty [kW]	815	

Lean amine pump

Dimensioning		
Adiabatic efficiency [%]	75	
Duty [kW]	669	

Base estimate		
Base cost estimate one pump unit with motor [$]	9840	[9], table 2.1, CS, $@2000
Base size [kW]	4	4-700 kW
Scaling factor	0,55	[9], table 2.1

Currency and cost index		
NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost rich amine pump	2,5
Cost lean amine pump	2,3

Appendix 11 Dimensioning of desorber column (1 of 1)

DESORBER

Dimensioning		**Note**
Vapour flow top column [Am3/h]	82190	
Vapour flow top column [Am3/s]	22,8	
Liquid flow [m3/s]	0,98	
Mass density for liquid [kg/m3]	967,7	
Mass density for vapour [kg/m3]	2,3	
Vapor massflow [t/h]	189,2	
Liquid massflow [t/h]	3496,7	
Surface tension [dyne/cm]	52,8	
Tray spacing [m]	0,91	[11]
x-axis	0,90	[11]
Cbs [m/s]	4,81E-02	[11]
vf [m/s]	1,20	[11]
vgas [m/s]	1,08	[11]
Calculated desorber diameter [m]	5,20	
Calculated desorber area [m2]	21,19	
Number of stages in absorber [-]	12	
Packing height [m]	12	
Absorber height [m]	30	Assumed
Wall thickness [m]	1,00E-02	Assumed
Density SS [kg/m3]	8000	Approx value
Absorber shell volume [m3]	4,9	
Mass of absorber [tons]	39,2	
Total packing volume [m3]	254,2	

Base estimate

Base cost estimate absorber shell [$]	65600	[9], table 2.1, CS, $@2000
Base size [t]	8	8-300t
Scaling factor	0,89	[9], table 2.1
Base cost estimate structured packing [$/m3]	4264	Assumed 140 % 2" Pall Rings, SS, 1990

Currency and cost index

NOK/$ @1990	6,25	[22]
NOK/$ @ 2000	8,81	[22]
CE index 1990	358	[23]
CE index 2000	394	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost absorber shell	0,42
Cost packing	11,7

Appendix 12 Dimensioning desorber condenser (1 of 1)

DESORBER HEAT EXCHANGERS

REBOILER

Dimensioning

		Note
Reboiler duty [MW]	154,2	Assuming neglible pressure loss
Steam temperature inlet [degC]	160	
Steam temperature outlet [degC]	151,8	
Reboiler boling temp [degC]	120,0	
LMDT [K]	35,8	
U overall [W/m2K]	2500	Assumed constant Table 14-5 [12]
Area[m2]	1723	

CONDENSER

Dimensioning

Condenser duty [MW]	26,3	
Cold utility temperature inlet [degC]	15	
Cold utility temperature outlet [degC]	24,7	
Top stage condensing temp [degC]	91,0	
LMDT [K]	71,1	
U overall [W/m2K]	2000	Assumed constant Table 14-5 [12]
Area[m2]	185	

Base estimate

Base cost estimate heat exchanger [$]	32800	[9], table 2.1, CS, $@2000
Base size [m2]	80	[9], table 2.1
Scaling factor	0,68	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Reboiler	3,6
Condenser	0,8

Appendix 13 Dimensioning of condenser 2 (1 of 1)

Condenser 2 | **Note**

Dimensioning

Assuming neglible pressure loss

Condenser duty [MW]	25,3	
Cold utility temperature inlet [degC]	15	
Cold utility temperature outlet [degC]	24,9	
Top stage condensing temp [degC]	91,0	
LMDT (Hysys) [K]	40,1	
U overall [W/m2K]	2000	Assumed constant Table 14-5 [12]
Area[m2]	315,4	

Base estimate

Base cost estimate heat exchanger [$]	32800	[9], table 2.1, CS, $@2000
Base size [m2]	80	[9], table 2.1
Scaling factor	0,68	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Condenser_2	1,15

Appendix 14 Dimensioning water separator (1 of 1)

Water separator

Dimensioning

		Note
Gas flow [Am3/h]	46613	
Gas flow [Am3/s]	13	
Liquid mass density [kg/m3]	1048	
Vapour mass density [kg/m3]	3,34	
K factor	0,1	Assumed
L/D ratio	4	Assumed
Vertical velocity [m/s]	1,77	Sounder-Brown eq.
Vessel diameter [m]	3,05	
Vessel length [m]	12,21	
Wall thickness [m]	1,00E-02	Assumed
Density SS [kg/m3]	8000	Approx value
Absorber shell volume [m3]	1,17	
Mass of absorber [tons]	9,4	

Base estimate

Base cost estimate absorber shell [$]	65600	[9], table 2.1, CS, $@2000
Base size	8	8-300 t
Scaling factor	0,89	[9], table 2.1

Currency and cost index

NOK/$ @ 2000	8,81	[22]
CE index 2000	394,1	[23]
CE index 2009	616	[23]

Base cost estimate	Base estimate [MNOK]
Cost water separator	0,12

Appendix 15 Equipment base cost and installed costs (1 of 2)

Overview installation factor and costs (1 of 2), all cost in mNOK

													NOTE
FLUE GAS TRANSPORT FAN													
Total installed cost DCC circulation pump												50,0	1
DCC VESSEL	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
Cost absorber shell, mNOK, 2009	5,2	CS	1	2,39	0,4	0,28	0,64	3,77	0,31	1	3,77	19,5	
Cost packing	15,9	SS316	1,75	2,61	0,46	0,32	0,71	4,15	0,37	1	5,18	82,3	
Column bearings, internal support and liquid distributors												3,6	2
Total installed DCC VESSEL												105,4	
DCC CIRCULATION PUMP	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
Cost circulation pump 1	1,8	CS	1	3,07	0,61	0,4	0,84	5	0,5	1	5	9,0	
Total installed cost DCC circulation pump												9,0	
DCC COOLER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
DCC CU cooler 1	8,0	SS304	1,3	2,39	0,4	0,28	0,64	3,77	0,31	1	4,16	33,3	
Total installed cost DCC COOLER												33,3	
DCC UNIT TOTAL COST												147,7	
ABSORBER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
Cost absorber shell, mNOK, 2009	15,7	CS	1	2,39	0,4	0,28	0,64	3,77	0,31	1	4,75	74,7	
Cost packing	133,9	SS	1,75	2,39	0,4	0,28	0,64	3,77	0,31	1	4,75	636,2	
Column bearings, internal support and liquid distributors												20,2	3
Additional cost for water wash section												127,2	4
Total installed cost ABSORBER												858,3	
LEAN AMINE COOLER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
HA-1 Lean amine	4,2	SS	1,75	2,61	0,46	0,32	0,71	4,15	0,37	1	5,18	21,8	
Total installed cost LEAN AMINE COOLER												21,8	
AMINE PUMPS	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost	
Cost rich amine pump	2,5	SS316	1,75	2,61	0,46	0,32	0,71	4,15	0,37	1	5,18	13,1	
Cost lean amine pump	2,3	SS316	1,75	3,07	0,61	0,4	0,28	4,16	0,5	1	5,29	12,0	
Total installed cost AMINE PUMPS												25,0	

Note

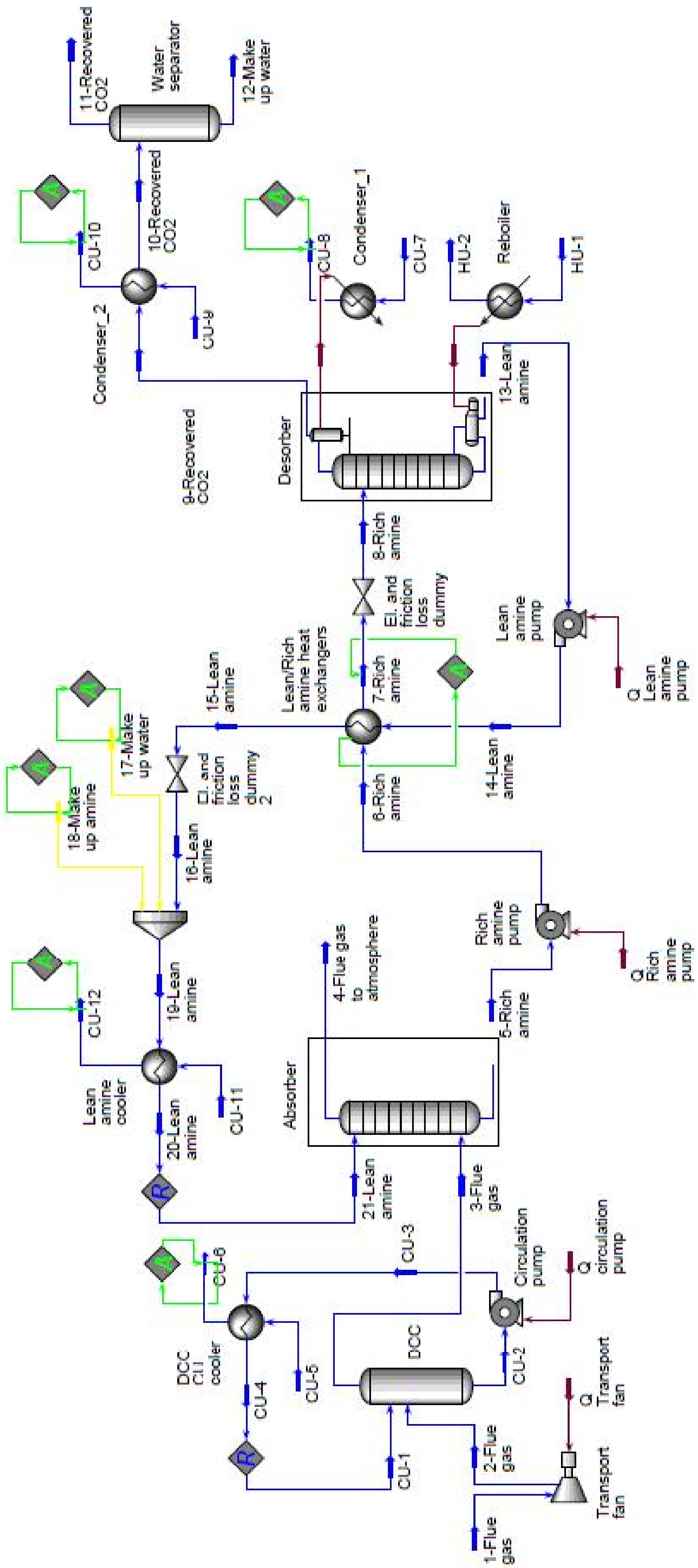
1 Assumed constant
2 10 % of shell cost + 2 % of packing cost
3 10 % of shell cost + 2 % of packing cost
4 20 % of packing cost

Appendix 15 Equipment base cost and installed costs (2 of 2)

Overview installation factor and costs (2 of 2), all cost in mNOK

LEAN/RICH HEAT EXCHANGERS	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
HA-1 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
HA-2 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
HA-3 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
HA-4 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
HA-5 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
HA-6 L/R amine heat exchanger	7,8	SS316	1,75	2,39	0,4	0,28	0,64	3,71	0,31	1	4,69	36,7
Total installed cost LEAN/RICH HEAT EXCHANGERS												220,3
DESORBER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
Cost desorber shell, mNOK, 2009	0,4	CS	1	4,29	1,06	0,64	1,23	7,33	0,85	1	7,33	3,1
Cost packing	11,7	SS	1	2,39	0,4	0,28	0,64	3,77	0,31	1	3,77	44,0
Total installed cost absorber												47,1
DESORBER REBOILER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
Reboiler	3,6	SS316	1,75	2,61	0,46	0,32	0,71	4,15	0,37	1	5,18	18,9
Total installed cost DESORBER REBOILER												18,9
DESORBER CONDENSER	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
Condenser	0,8	SS316	1,75	3,53	0,75	0,48	0,98	5,83	0,62	1	7,05	5,6
Total installed cost DESORBER CONDENSER												5,6
CONDENSER_2	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
Condenser_2	1,2	SS304	1,3	3,53	0,75	0,48	0,98	5,83	0,37	1	6,24	7,2
Total installed cost CONDENSER_2												7,2
WATER SEPARATOR	Base estimate	Material	Material factor	Direct cost factor	Engineering cost factor	Administration cost factor	Contingency factor	Total cost factor	Piping factor	Equipment factor	Installation factor	Installed cost
Cost water separator	0,1	SS304	1,3	4,29	1,06	0,64	1,23	7,33	0,85	1	7,89	0,9
Total installed cost WATER SEPARATOR												0,9

Appendix 16 Hysys base case PFD



Appendix 17 Case study 1A: Process overview

Case step [dTmin in L/R heat exchanger]	BC	5	6	7	8	9	10

Economic evaluation

CAPEX equipment installed cost [MNOK]	1400	1556	1496	1458	1426	1410	1383
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	110	111	112	114	114	116
Operating expenditure - cold utility [MNOK/yr]	4,0	3,6	3,7	3,7	3,7	3,8	3,8
Operating expenditure - electricity [MNOK/yr]	75,3	75,0	75,0	74,9	74,9	75,0	75,0
OPEX total [MNOK/yr]	203	188	190	191	192	193	195
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	3553	3504	3480	3462	3457	3445

Performance

Specific energy consumption [MJ/kg CO2]	3,61	3,35	3,38	3,42	3,46	3,48	3,53
CO2 removed [%]	85,2	81,9	81,9	81,9	81,9	81,9	81,9
CO2 removed [t/y]	1230134	1182546	1182519	1182519	1182519	1182519	1182519
CO2 removed total [Mtons]	24,6	23,7	23,7	23,7	23,7	23,7	23,7
Total specific cost [NOK/t CO2]	222	225	224	223	223	223	223

Process performance

Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,48	0,48	0,48	0,48	0,48	0,48
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	3340	3340	3340	3340	3340	3340
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	112,8	112,1	111,4	110,7	110,0	109,3
dTmin in L/R heat exchanger [degC]	10,0	5,0	6,0	7,0	8,0	9,0	10,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,5	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 17 Case study 1A: Process overview

Case step [dTmin in L/R heat exchanger]	11	12	13	14	15	16	17	18
Economic evaluation								
CAPEX equipment installed cost [MNOK]	1371	1350	1340	1320	1313	1306	1300	1284
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	117	118	120	121	123	124	126	128
Operating expenditure - cold utility [MNOK/yr]	3,9	3,9	3,9	4,0	4,0	4,1	4,1	4,2
Operating expenditure - electricity [MNOK/yr]	75,0	75,0	75,0	75,0	75,0	75,0	75,0	75,0
OPEX total [MNOK/yr]	196	197	199	200	202	204	205	207
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3447	3440	3446	3441	3452	3462	3474	3477
Performance								
Specific energy consumption [MJ/kg CO2]	3,57	3,61	3,65	3,69	3,74	3,79	3,84	3,89
CO2 removed [%]	81,9	81,9	81,9	81,9	81,9	81,9	81,9	81,9
CO2 removed [t/y]	1182519	1182519	1182519	1182519	1182519	1182519	1182519	1182519
CO2 removed total [Mtons]	23,7	23,7	23,7	23,7	23,7	23,7	23,7	23,7
Total specific cost [NOK/t CO2]	224	224	225	225	226	227	229	229
Process performance								
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,48	0,48	0,48	0,48	0,48	0,48	0,48	0,48
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3340	3340	3340	3340	3340	3340	3340	3340
Number of stages in absorber [-]	16	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	108,5	107,8	106,9	106,0	105,1	104,1	103,2	102,2
dTmin in L/R heat exchanger [degC]	11,0	12,0	13,0	14,0	15,0	16,1	17,0	18,0
Number of stages in desorber [-]	12	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500	500

Appendix 17 Case study 1A: Process overview

Case step [dTmin in L/R heat exchanger]	19	20	21	22	23	24	25
Economic evaluation							
CAPEX equipment installed cost [MNOK]	1280	1275	1272	1268	1265	1252	1249
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	130	131	133	135	137	139	141
Operating expenditure - cold utility [MNOK/yr]	4,2	4,3	4,3	4,4	4,5	4,5	4,6
Operating expenditure - electricity [MNOK/yr]	75,0	75,0	75,0	75,0	75,0	75,0	75,0
OPEX total [MNOK/yr]	209	211	213	215	217	219	221
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3492	3507	3524	3542	3560	3570	3589
Performance							
Specific energy consumption [MJ/kg CO2]	3,94	4,00	4,06	4,12	4,18	4,24	4,30
CO2 removed [%]	81,9	81,9	81,9	81,9	81,9	81,9	81,9
CO2 removed [t/y]	1182519	1182519	1182519	1182519	1182519	1182519	1182519
CO2 removed total [Mtons]	23,7	23,7	23,7	23,7	23,7	23,7	23,7
Total specific cost [NOK/t CO2]	231	232	234	235	237	238	240
Process performance							
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,48	0,48	0,48	0,48	0,48	0,48	0,48
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3340	3340	3340	3340	3340	3340	3340
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	101,2	100,2	99,2	98,2	97,2	96,2	95,2
dTmin in L/R heat exchanger [degC]	19,0	20,0	21,0	22,0	23,0	24,0	25,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 18 Case study 1A: Cost overview

Case step [dTmin in L/R heat exchanger]	**Base case**	**5**	**6**	**7**	**8**	**9**	**10**
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	22	15	16	17	18	19	20
Lean amine circulation pump	12	11	11	11	11	11	11
Lean/rich heat exchangers	220	385	324	286	253	236	209
Desorber column	47	45	45	44	44	44	43
Desorber condenser	6	6	6	6	6	5	5
Desorber reboiler	19	19	19	20	20	20	20
Desorber OH condenser	5	5	5	5	5	5	5
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13
Total installed cost [MNOK]	1400	1556	1496	1458	1426	1410	1383
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	154198	137374	138676	140336	141903	143075	144858
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	109,9	110,9	112,3	113,5	114,5	115,9
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	3889	2076	2351	2484	2857	3165	3479
Desorber reflux condenser [m3/h]	2225	2518	2265	2262	2227	2174	2146
Desorber cooler[m3/h]	2098	2079	2355	2240	2129	1999	1891
Total cold utility consumption [m3/h]	14938	13399	13697	13713	13939	14064	14242
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	3,6	3,7	3,7	3,7	3,8	3,8
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	669	617	614	605	606	622	606
Rich amine pump [kW]	815	755	755	755	755	755	755
Total electricity consumption [kWh/yr]	188358556	187461047	187441190	187367140	187369864	187498733	187375922
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	75,0	75,0	74,9	74,9	75,0	75,0
Total utility cost [MNOK/yr]	203	188	190	191	192	193	195

Appendix 18 Case study 1A: Cost overview

Case step [dTmin in L/R heat exchanger]	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Installed cost equipment in MNOK								
Transport fan	50	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858	858
Lean amine cooler	21	21	22	23	24	24	25	26
Lean amine circulation pump	11	11	11	11	11	11	11	11
Lean/rich heat exchangers	196	174	164	144	137	129	122	106
Desorber column	44	43	43	42	42	42	42	41
Desorber condenser	5	5	5	5	5	5	6	6
Desorber reboiler	20	20	20	21	21	21	21	21
Desorber OH condenser	4	4	4	4	4	4	4	3
Desorber OH water separator	1	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13	13
Total installed cost [MNOK]	1371	1350	1340	1320	1313	1306	1300	1284
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Utility consumption								
Hot utility								
Desorber duty [kW]	146400	148077	149845	151591	153620	155604	157692	159757
MEA recalimer duty [MW] **								
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	117,1	118,5	119,9	121,3	122,9	124,5	126,2	127,8
Cold utility								
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	3852	4144	4454	4671	4982	5325	5660	6004
Desorber reflux condenser [m3/h]	2105	2072	2041	2007	1967	1936	1914	1874
Desorber cooler[m3/h]	1777	1685	1571	1466	1369	1256	1160	1055
Total cold utility consumption [m3/h]	14460	14628	14792	14871	15045	15243	15460	15659
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	3,9	3,9	3,9	4,0	4,0	4,1	4,1	4,2
Electricity consumers								
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437	437
Lean amine pump [kW]	617	609	609	608	608	608	609	611
Rich amine pump [kW]	755	755	755	755	755	755	755	755
Total electricity consumption [kWh/yr]	187458273	187398087	187401628	187386608	187388727	187391778	187394045	187415424
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,0	75,0	75,0	75,0	75,0	75,0	75,0	75,0
Total utility cost [MNOK/yr]	196	197	199	200	202	204	205	207

Appendix 18 Case study 1A: Cost overview

Case step [dTmin in L/R heat exchanger]	**19**	**20**	**21**	**22**	**23**	**24**	**25**
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	26	27	28	28	29	29	30
Lean amine circulation pump	11	11	11	11	11	11	11
Lean/rich heat exchangers	101	96	92	88	84	71	68
Desorber column	41	41	41	40	40	40	40
Desorber condenser	6	6	6	6	6	6	6
Desorber reboiler	22	22	22	22	22	23	23
Desorber OH condenser	3	3	3	3	3	2	2
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13
Total installed cost [MNOK]	1280	1275	1272	1268	1265	1252	1249
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	161969	164352	166677	169095	171526	174149	176641
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	129,6	131,5	133,3	135,3	137,2	139,3	141,3
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	6347	6611	6956	7300	7643	7982	8228
Desorber reflux condenser [m3/h]	1843	1820	1798	1776	1745	1731	1718
Desorber cooler[m3/h]	959	859	772	685	598	520	434
Total cold utility consumption [m3/h]	15875	16016	16252	16486	16711	16959	17106
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,2	4,3	4,3	4,4	4,5	4,5	4,6
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	617	609	615	610	616	610	611
Rich amine pump [kW]	755	755	755	755	755	755	755
Total electricity consumption [kWh/yr]	187465610	187401588	187444949	187405945	187457079	187409707	187411789
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,0	75,0	75,0	75,0	75,0	75,0	75,0
Total utility cost [MNOK/yr]	209	211	213	215	217	219	221

Appendix 19 Case study 1B: Process overview

Case step [dTmin in L/R heat exchanger]	**Base case**	**5**	**6**	**7**	**8**	**9**	**10**
Economic evaluation							
CAPEX equipment installed cost [MNOK]	1400	1590	1537	1485	1461	1429	1402
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	117	118	120	121	122	123
Operating expenditure - cold utility [MNOK/yr]	4,0	3,8	3,8	3,8	3,9	3,9	4,0
Operating expenditure - electricity [MNOK/yr]	75,3	75,3	75,3	75,3	75,3	75,3	75,3
OPEX total [MNOK/yr]	203	196	198	199	200	202	203
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	3667	3630	3590	3580	3565	3550
Performance							
Specific energy consumption [MJ/kg CO2]	3,61	3,42	3,47	3,50	3,54	3,58	3,61
CO2 removed [%]	85,2	85,2	85,2	85,2	85,2	85,2	85,2
CO2 removed [t/y]	1230134	1230134	1230134	1230134	1230134	1230134	1230134
CO2 removed total [Mtons]	24,6	24,6	24,6	24,6	24,6	24,6	24,6
Total specific cost [NOK/t CO2]	222	224	223	222	222	222	222
Process performance							
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,47	0,47	0,47	0,47	0,47	0,47
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	3600	3600	3600	3600	3600	3600
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	113,3	112,6	111,9	111,1	110,3	109,5
dTmin in L/R heat exchanger [degC]	10,0	5,0	6,0	7,0	8,0	9,0	10,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 19 Case study 1B: Process overview

Case step [dTmin in L/R heat exchanger]	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Economic evaluation								
CAPEX equipment installed cost [MNOK]	1388	1364	1354	1334	1327	1319	1313	1296
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	125	126	128	129	131	133	134	136
Operating expenditure - cold utility [MNOK/yr]	4,0	4,0	4,1	4,1	4,2	4,2	4,2	4,3
Operating expenditure - electricity [MNOK/yr]	75,3	75,3	75,3	75,3	75,3	75,3	75,3	75,3
OPEX total [MNOK/yr]	204	206	207	209	211	212	214	216
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3553	3544	3549	3545	3557	3567	3579	3583
Performance								
Specific energy consumption [MJ/kg CO2]	3,66	3,70	3,74	3,78	3,84	3,88	3,93	3,99
CO2 removed [%]	85,2	85,2	85,2	85,2	85,2	85,2	85,2	85,2
CO2 removed [t/y]	1230134	1230134	1230134	1230134	1230134	1230134	1230134	1230134
CO2 removed total [Mtons]	24,6	24,6	24,6	24,6	24,6	24,6	24,6	24,6
Total specific cost [NOK/t CO2]	223	223	223	224	225	226	227	228
Process performance								
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,47	0,47	0,47	0,47	0,47	0,47	0,47
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	3600	3600	3600	3600	3600	3600	3600
Number of stages in absorber [-]	16	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	108,6	107,8	106,9	106,0	105,1	104,1	103,2	102,2
dTmin in L/R heat exchanger [degC]	11,0	12,0	13,0	14,0	15,0	16,1	17,0	18,0
Number of stages in desorber [-]	12	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500	500

Appendix 19 Case study 1B: Process overview

Case step [dTmin in L/R heat exchanger]	**19**	**20**	**21**	**22**	**23**	**24**	**25**
Economic evaluation							
CAPEX equipment installed cost [MNOK]	1292	1287	1283	1279	1265	1262	1260
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	138	140	142	144	146	149	151
Operating expenditure - cold utility [MNOK/yr]	4,3	4,4	4,4	4,5	4,6	4,6	4,7
Operating expenditure - electricity [MNOK/yr]	75,4	75,3	75,3	75,3	75,3	75,3	75,3
OPEX total [MNOK/yr]	218	220	222	224	226	228	231
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3599	3616	3634	3653	3662	3683	3704
Performance							
Specific energy consumption [MJ/kg CO2]	4,04	4,10	4,16	4,22	4,28	4,35	4,41
CO2 removed [%]	85,2	85,2	85,2	85,2	85,2	85,2	85,2
CO2 removed [t/y]	1230134	1230134	1230134	1230134	1230134	1230134	1230134
CO2 removed total [Mtons]	24,6	24,6	24,6	24,6	24,6	24,6	24,6
Total specific cost [NOK/t CO2]	230	231	233	234	235	237	239
Process performance							
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,47	0,47	0,47	0,47	0,47	0,47
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	3600	3600	3600	3600	3600	3600
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	101,2	100,2	99,2	98,2	97,2	96,2	95,2
dTmin in L/R heat exchanger [degC]	19,0	20,0	21,0	22,0	23,0	24,0	25,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 20 Case study 1B: Cost overview

Case step [dTmin in L/R heat exchanger]	**Base case**	**5**	**6**	**7**	**8**	**9**	**10**
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	22	17	18	20	20	21	22
Lean amine circulation pump	12	12	12	12	12	12	12
Lean/rich heat exchangers	220	411	358	305	280	249	220
Desorber column	47	49	48	47	47	47	47
Desorber condenser	6	6	6	6	6	6	6
Desorber reboiler	19	20	20	20	21	21	21
Desorber OH condenser	5	6	5	5	5	5	5
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13
Total installed cost [MNOK]	1400	1590	1537	1485	1461	1429	1402
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	154198	146181	148051	149555	151148	152977	154326
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	116,9	118,4	119,6	120,9	122,4	123,5
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	3889	2474	2629	2474	3191	3499	3813
Desorber reflux condenser [m3/h]	2225	2263	2263	2261	2261	2261	2261
Desorber cooler[m3/h]	2098	2703	2612	2703	2351	2230	2098
Total cold utility consumption [m3/h]	14938	14166	14230	14164	14529	14715	14897
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	3,8	3,8	3,8	3,9	3,9	4,0
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	669	671	653	653	654	654	663
Rich amine pump [kW]	815	815	815	815	815	815	815
Total electricity consumption [kWh/yr]	188358556	188370948	188229899	188233277	188236488	188239227	188311753
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	75,3	75,3	75,3	75,3	75,3	75,3
Total utility cost [MNOK/yr]	203	196	198	199	200	202	203

Appendix 20 Case study 1B: Cost overview

Case step [dTmin in L/R heat exchanger]	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Installed cost equipment in MNOK								
Transport fan	50	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858	858
Lean amine cooler	23	24	24	25	26	27	27	28
Lean amine circulation pump	12	12	12	12	12	12	12	12
Lean/rich heat exchangers	206	182	171	151	143	134	128	111
Desorber column	46	46	46	45	45	45	44	44
Desorber condenser	6	6	6	6	6	6	6	6
Desorber reboiler	21	21	21	22	22	22	22	22
Desorber OH condenser	5	5	4	4	4	4	4	4
Desorber OH water separator	1	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13	13
Total installed cost [MNOK]	1388	1364	1354	1334	1327	1319	1313	1296
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Utility consumption								
Hot utility								
Desorber duty [kW]	156325	158050	159743	161635	163860	165891	167998	170346
MEA recalimer duty [MW] **								
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	125,1	126,4	127,8	129,3	131,1	132,7	134,4	136,3
Cold utility								
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	4130	4334	4644	4956	5268	5610	5851	6194
Desorber reflux condenser [m3/h]	2228	2204	2155	2121	2081	2050	2019	1979
Desorber cooler[m3/h]	1967	1861	1746	1632	1509	1404	1299	1186
Total cold utility consumption [m3/h]	15051	15125	15271	15436	15584	15791	15896	16085
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	4,0	4,1	4,1	4,2	4,2	4,2	4,3
Electricity consumers								
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437	437
Lean amine pump [kW]	657	655	660	656	657	658	657	658
Rich amine pump [kW]	815	815	815	815	815	815	815	815
Total electricity consumption [kWh/yr]	188264316	188248557	188284193	188254549	188257680	188267732	188262956	188265738
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	75,3	75,3	75,3	75,3	75,3	75,3	75,3
Total utility cost [MNOK/yr]	204	206	207	209	211	212	214	216

Appendix 20 Case study 1B: Cost overview

Case step [dTmin in L/R heat exchanger]	**19**	**20**	**21**	**22**	**23**	**24**	**25**
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	29	29	30	30	31	32	32
Lean amine circulation pump	12	12	12	12	12	12	12
Lean/rich heat exchangers	105	100	96	91	77	74	71
Desorber column	44	44	43	43	43	43	42
Desorber condenser	6	6	6	6	6	6	6
Desorber reboiler	22	23	23	23	23	24	24
Desorber OH condenser	3	3	3	3	3	3	2
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	13	13	13
Total installed cost [MNOK]	1292	1287	1283	1279	1265	1262	1260
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	172637	175181	177795	180370	183022	185719	188487
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	138,1	140,1	142,2	144,3	146,4	148,6	150,8
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	6537	6802	7147	7490	7834	8075	8419
Desorber reflux condenser [m3/h]	1940	1917	1885	1863	1841	1819	1797
Desorber cooler[m3/h]	1072	972	876	780	693	597	511
Total cold utility consumption [m3/h]	16275	16416	16635	16860	17093	17217	17452
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,3	4,4	4,4	4,5	4,6	4,6	4,7
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	671	668	661	659	659	664	660
Rich amine pump [kW]	815	815	815	815	815	815	815
Total electricity consumption [kWh/yr]	188376589	188348651	188291910	188275631	188277861	188315311	188282052
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,4	75,3	75,3	75,3	75,3	75,3	75,3
Total utility cost [MNOK/yr]	218	220	222	224	226	228	231

Appendix 21 Case study 1C: Process overview

Case step [dTmin in L/R heat exchanger]	Base case	5	6	7	8	9	10
Economic evaluation							
CAPEX equipment installed cost [MNOK]	1400	1973	1856	1781	1718	1664	1637
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	202	204	207	209	212	214
Operating expenditure - cold utility [MNOK/yr]	4,0	7,1	7,1	7,2	7,3	7,4	7,3
Operating expenditure - electricity [MNOK/yr]	75,3	81,0	81,0	81,0	81,0	81,0	81,0
OPEX total [MNOK/yr]	203	290	292	295	297	300	302
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	5043	4953	4905	4869	4843	4840
Performance							
Specific energy consumption [MJ/kg CO2]	3,61	5,58	5,65	5,72	5,79	5,86	5,92
CO2 removed [%]	85,2	90,1	90,1	90,1	90,1	90,1	90,1
CO2 removed [t/y]	1230134	1300706	1300706	1300706	1300706	1300706	1300706
CO2 removed total [Mtons]	24,6	26,0	26,0	26,0	26,0	26,0	26,0
Total specific cost [NOK/t CO2]	222	299	296	295	295	295	295
Process performance							
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,36	0,36	0,36	0,36	0,36	0,36
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	7850	7850	7850	7850	7850	7850
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	114,0	113,2	112,3	111,4	110,5	109,8
dTmin in L/R heat exchanger [degC]	10,0	5,0	6,1	7,1	8,1	9,1	10,1
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,5
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 21 Case study 1C: Process overview

Case step [dTmin in L/R heat exchanger]	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Economic evaluation								
CAPEX equipment installed cost [MNOK]	1598	1576	1542	1522	1509	1498	1467	1458
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	215	219	220	227	229	231	236	240
Operating expenditure - cold utility [MNOK/yr]	7,4	7,5	7,6	7,7	7,8	7,9	7,9	8,0
Operating expenditure - electricity [MNOK/yr]	81,1	81,0	81,2	81,0	81,0	81,1	81,0	81,0
OPEX total [MNOK/yr]	304	307	309	315	318	320	324	329
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	4817	4833	4817	4862	4878	4892	4905	4942
Performance								
Specific energy consumption [MJ/kg CO2]	5,96	6,06	6,10	6,27	6,34	6,40	6,52	6,64
CO2 removed [%]	90,1	90,1	90,1	90,1	90,1	90,1	90,1	90,1
CO2 removed [t/y]	1300706	1300706	1300706	1300706	1300706	1300706	1300706	1300706
CO2 removed total [Mtons]	26,0	26,0	26,0	26,0	26,0	26,0	26,0	26,0
Total specific cost [NOK/t CO2]	295	297	297	301	302	304	306	309
Process performance								
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,36	0,36	0,36	0,36	0,36	0,36	0,36	0,36
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40	40
Lean amine rate [tons/h]	7850	7850	7850	7850	7850	7850	7850	7850
Number of stages in absorber [-]	16	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	108,9	108,0	107,2	106,1	105,2	104,2	103,2	102,2
dTmin in L/R heat exchanger [degC]	11,0	12,0	13,0	14,1	15,0	16,0	17,0	18,0
Number of stages in desorber [-]	12	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,5	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reboiler temperature [degC]	120	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500	500

Appendix 21 Case study 1C: Process overview

Case step [dTmin in L/R heat exchanger]	**19**	**20**	**21**	**22**	**23**	**24**	**25**

Economic evaluation

CAPEX equipment installed cost [MNOK]	1450	1443	1436	1413	1408	1405	1401
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	243	247	251	255	260	264	269
Operating expenditure - cold utility [MNOK/yr]	8,1	8,3	8,3	8,4	8,6	8,7	8,8
Operating expenditure - electricity [MNOK/yr]	81,0	81,0	81,0	81,1	81,0	81,1	81,1
OPEX total [MNOK/yr]	333	336	341	345	349	354	359
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	4973	5007	5046	5066	5110	5155	5202

Performance

Specific energy consumption [MJ/kg CO2]	6,74	6,84	6,96	7,07	7,19	7,31	7,44
CO2 removed [%]	90,1	90,1	90,1	90,1	90,1	90,1	90,1
CO2 removed [t/y]	1300706	1300706	1300706	1300706	1300706	1300706	1300706
CO2 removed total [Mtons]	26,0	26,0	26,0	26,0	26,0	26,0	26,0
Total specific cost [NOK/t CO2]	311	314	317	319	323	326	330

Process performance

Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,36	0,36	0,36	0,36	0,36	0,36	0,36
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	7850	7850	7850	7850	7850	7850	7850
Number of stages in absorber [-]	16	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	101,2	100,2	99,2	98,2	97,2	96,2	95,2
dTmin in L/R heat exchanger [degC]	19,0	20,0	21,0	22,0	23,0	24,0	25,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 22 Case study 1C: Cost overview

Case step [dTmin in L/R heat exchanger]	Base case	5	6	7	8	9	10
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	22	37	39	40	42	43	44
Lean amine circulation pump	12	18	18	18	18	18	18
Lean/rich heat exchangers	220	701	584	508	444	389	364
Desorber column	47	92	91	91	90	90	88
Desorber condenser	6	8	7	7	7	7	8
Desorber reboiler	19	29	29	30	30	30	30
Desorber OH condenser	5	10	10	9	9	9	8
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	20	20	20	20	20	20
Total installed cost [MNOK]	1400	1973	1856	1781	1718	1664	1637

Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	154198	252136	255237	258344	261515	264761	267440
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	201,7	204,2	206,7	209,2	211,8	214,0
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	3889	8661	9004	9651	10337	10985	11766
Desorber reflux condenser [m3/h]	2225	3913	3913	3913	3913	3820	3917
Desorber cooler[m3/h]	2098	7311	7019	6705	6390	6076	5132
Total cold utility consumption [m3/h]	14938	26610	26662	26994	27366	27607	27542
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	7,1	7,1	7,2	7,3	7,4	7,3
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	669	1451	1447	1448	1449	1450	1453
Rich amine pump [kW]	815	1793	1793	1793	1793	1793	1793
Total electricity consumption [kWh/yr]	188358556	202439468	202407262	202415344	202423336	202431228	202455367
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	81,0	81,0	81,0	81,0	81,0	81,0
Total utility cost [MNOK/yr]	203	290	292	295	297	300	302

Appendix 22 Case study 1C: Cost overview

Case step [dTmin in L/R heat exchanger]	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Installed cost equipment in MNOK								
Transport fan	50	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858	858
Lean amine cooler	45	47	48	49	50	51	52	53
Lean amine circulation pump	19	18	19	18	18	19	18	18
Lean/rich heat exchangers	321	301	263	246	232	219	189	179
Desorber column	90	88	90	86	86	87	85	85
Desorber condenser	8	8	7	7	7	7	7	7
Desorber reboiler	30	31	31	32	32	32	32	33
Desorber OH condenser	8	8	7	7	7	6	6	6
Desorber OH water separator	1	1	1	1	1	1	1	1
Rich amine circulation pump	20	20	20	20	20	20	20	20
Total installed cost [MNOK]	1598	1576	1542	1522	1509	1498	1467	1458

Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Utility consumption								
Hot utility								
Desorber duty [kW]	269202	273539	275513	283232	286515	289243	294400	299753
MEA recalimer duty [MW] **								
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	215,4	218,8	220,4	226,6	229,2	231,4	235,5	239,8
Cold utility								
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	12303	12970	13582	14311	15055	15798	16442	17185
Desorber reflux condenser [m3/h]	3916	3916	3884	3756	3707	3535	3451	3296
Desorber cooler[m3/h]	4814	4568	4241	3990	3752	3412	3165	2919
Total cold utility consumption [m3/h]	27759	28180	28433	28783	29240	29471	29784	30126
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	7,4	7,5	7,6	7,7	7,8	7,9	7,9	8,0
Electricity consumers								
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437	437
Lean amine pump [kW]	1498	1469	1516	1456	1457	1489	1459	1460
Rich amine pump [kW]	1793	1793	1793	1793	1793	1793	1793	1793
Total electricity consumption [kWh/yr]	202818822	202583654	202962843	202481397	202489696	202743529	202505477	202510984
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	81,1	81,0	81,2	81,0	81,0	81,1	81,0	81,0
Total utility cost [MNOK/yr]	304	307	309	315	318	320	324	329

Appendix 22 Case study 1C: Cost overview

Case step [dTmin in L/R heat exchanger]	19	20	21	22	23	24	25
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148
Absorber column	858	858	858	858	858	858	858
Lean amine cooler	54	55	56	57	58	58	59
Lean amine circulation pump	18	18	18	18	18	19	18
Lean/rich heat exchangers	170	162	154	129	124	118	113
Desorber column	85	85	85	86	85	87	86
Desorber condenser	7	7	7	7	7	7	7
Desorber reboiler	33	33	34	34	35	35	35
Desorber OH condenser	6	5	5	5	5	4	4
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	20	20	20	20	20	20	20
Total installed cost [MNOK]	1450	1443	1436	1413	1408	1405	1401
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	304216	308934	314176	319165	324836	330230	336174
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	243,4	247,1	251,3	255,3	259,9	264,2	268,9
Cold utility							
DCC CU cooler [m3/h]	6726	6726	6726	6726	6726	6726	6726
Lean amine cooler [m3/h]	17928	18671	19314	20056	20747	21491	22137
Desorber reflux condenser [m3/h]	3230	3128	3036	2925	2850	2757	2693
Desorber cooler[m3/h]	2653	2413	2184	1956	1753	1556	1373
Total cold utility consumption [m3/h]	30537	30938	31260	31663	32077	32530	32929
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	8,1	8,3	8,3	8,4	8,6	8,7	8,8
Electricity consumers							
Transport fan [kW]	21623	21623	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	437	437	437	437	437	437
Lean amine pump [kW]	1461	1461	1462	1474	1463	1484	1477
Rich amine pump [kW]	1793	1793	1793	1793	1793	1793	1793
Total electricity consumption [kWh/yr]	202518161	202524543	202530103	202627520	202541061	202703321	202650805
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	81,0	81,0	81,0	81,1	81,0	81,1	81,1
Total utility cost [MNOK/yr]	333	336	341	345	349	354	359

Appendix 23 Case study 2A: Process overview

Case step [Stages in absorber]	Base case	10	11	12	13	14	15
Economic evaluation							
CAPEX equipment installed cost [MNOK]	1400	1951	1274	1232	1252	1297	1338
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	355	156	124	114	115	114
Operating expenditure - cold utility [MNOK/yr]	4,0	11,4	5,1	4,0	3,7	3,7	3,7
Operating expenditure - electricity [MNOK/yr]	75,3	78	62	63	66	69	72
OPEX total [MNOK/yr]	203	444	223	191	184	188	189
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	6659	3636	3259	3200	3283	3342
Performance							
Specific energy consumption [MJ/kg CO2]	3,61	11,09	4,85	3,86	3,61	3,57	3,53
CO2 removed [%]	85,2	79,7	80,1	80,1	79,0	80,1	80,3
CO2 removed [t/y]	1230134	1150725	1156700	1157460	1141325	1157344	1159833
CO2 removed total [Mtons]	24,6	23	23	23	23	23	23
Total specific cost [NOK/t CO2]	222	471	248	219	216	218	221
Process performance							
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,30	0,40	0,45	0,47	0,47	0,48
Absorber gas feed temperature [degC]	40,7	39,8	39,9	40,1	40,2	40,3	40,5
Absorber gas feed pressure [kPa]	121	115,38	116,31	117,25	118,19	119,13	120,06
Absorber gas feed flowrate [Am3/h]	2360705	2469370	2450509	2431748	2413287	2395117	2377422
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	19975	5200	3730	3330	3330	3280
Number of stages in absorber [-]	16	10	11	12	13	14	15
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	110,1	109,6	109,5	109,4	109,4	109,3
dTmin in L/R heat exchanger [degC]	10,0	10,0	10,1	10,1	10,1	10,0	10,0
Number of stages in desorber [-]	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500

Appendix 24 Case study 2A: Cost overview

Case step [Stages in absorber]	**Base case**	**10**	**11**	**12**	**13**	**14**	**15**
Installed cost equipment in MNOK							
Transport fan	50	50	50	50	50	50	50
Direct contact cooler	148	149	149	148	148	148	148
Absorber column	858	590	636	682	727	771	815
Lean amine cooler	22	72	31	23	21	20	20
Lean amine circulation pump	12	31	15	12	11	11	11
Lean/rich heat exchangers	220	733	278	223	208	209	207
Desorber column	47	231	62	47	43	43	43
Desorber condenser	6	8	6	5	5	5	5
Desorber reboiler	19	43	24	21	20	20	20
Desorber OH condenser	5	11	7	5	5	5	5
Desorber OH water separator	1	1	1	1	1	1	1
Rich amine circulation pump	13	34	16	13	13	13	12
Total installed cost [MNOK]	1400	1951	1274	1232	1252	1297	1338
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000
Utility consumption							
Hot utility							
Desorber duty [kW]	154198	443222	194794	155202	143047	143651	142002
MEA recalimer duty [MW] **							
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	354,6	155,8	124,2	114,4	114,9	113,6
Cold utility							
DCC CU cooler [m3/h]	6726	6300	6375	6450	6526	6557	6557
Lean amine cooler [m3/h]	3889	22309	6657	4300	3431	3431	3427
Desorber reflux condenser [m3/h]	2225	4839	2517	2086	2095	2095	2104
Desorber cooler[m3/h]	2098	9090	3643	2274	1903	1891	1857
Total cold utility consumption [m3/h]	14938	42537	19192	15110	13955	13973	13945
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	11,4	5,1	4,0	3,7	3,7	3,7
Electricity consumers							
Transport fan [kW]	21623	15825	16798	17776	18749	19716	20667
DCC CU circulation pump [kW]	437	459	456	453	450	447	444
Lean amine pump [kW]	669	3713	955	680	605	605	608
Rich amine pump [kW]	815	4531	1183	846	754	753	741
Total electricity consumption [kWh/yr]	188358556	196219879	155130726	158041987	164463744	172171184	179686821
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	78,5	62,1	63,2	65,8	68,9	71,9
Total utility cost [MNOK/yr]	203	444	223	191	184	188	189

Appendix 25 Case study 2B: Process overview

Case step [Stages in absorber]	**Base case**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Economic evaluation								
CAPEX equipment installed cost [MNOK]	1400	2454	1459	1355	1370	1404	1439	1478
Income CO2 [NOK/t CO2]	0	0	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	441	193	138	127	123	122	120
Operating expenditure - cold utility [MNOK/yr]	4,0	13,5	6,5	4,4	4,0	4,0	3,9	3,9
Operating expenditure - electricity [MNOK/yr]	75,3	95	71	70	72	75	78	81
OPEX total [MNOK/yr]	203	550	270	213	204	203	204	205
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	8276	4324	3610	3530	3550	3598	3655
Performance								
Specific energy consumption [MJ/kg CO2]	3,61	12,90	5,66	4,05	3,73	3,60	3,56	3,52
CO2 removed [%]	85,2	85,2	85,2	85,2	85,2	85,2	85,2	85,2
CO2 removed [t/y]	1230134	1230171	1230715	1230580	1230723	1230832	1231225	1230422
CO2 removed total [Mtons]	24,6	25	25	25	25	25	25	25
Total specific cost [NOK/t CO2]	222	547	279	228	221	222	224	227
Process performance								
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,29	0,37	0,44	0,46	0,47	0,47	0,48
Absorber gas feed temperature [degC]	40,7	40,1	40,2	40,3	40,5	40,6	40,7	40,8
Absorber gas feed pressure [kPa]	121	117,25	118,19	119,13	120,06	121	121,94	122,88
Absorber gas feed flowrate [Am3/h]	2360705	2431748	2413287	2395117	2377422	2359566	2341976	2324645
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	28000	6929	4250	3770	3600	3520	3475
Number of stages in absorber [-]	16	12	13	14	15	16	17	18
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	110,2	109,6	109,6	109,6	109,6	109,4	109,3
dTmin in L/R heat exchanger [degC]	10,0	10,0	10,1	10,0	10,0	9,9	10,0	10,0
Number of stages in desorber [-]	12	12	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500	500	500

Appendix 26 Case study 2B: Cost overview

Case step [Stages in absorber]	**Base case**	**12**	**13**	**14**	**15**	**16**	**17**	**18**
Installed cost equipment in MNOK								
Transport fan	50	50	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	148	148	148
Absorber column	858	682	727	771	815	858	900	942
Lean amine cooler	22	88	40	26	23	22	21	20
Lean amine circulation pump	12	37	17	13	12	12	12	12
Lean/rich heat exchangers	220	927	335	244	227	222	218	216
Desorber column	47	410	79	53	48	47	46	46
Desorber condenser	6	9	7	6	6	6	6	6
Desorber reboiler	19	50	28	23	21	21	21	20
Desorber OH condenser	5	12	8	6	5	5	5	5
Desorber OH water separator	1	1	1	1	1	1	1	1
Rich amine circulation pump	13	40	19	14	13	13	13	13
Total installed cost [MNOK]	1400	2454	1459	1355	1370	1404	1439	1478
Uptime [h/yr]	8000	8000	8000	8000	8000	8000	8000	8000
Utility consumption								
Hot utility								
Desorber duty [kW]	154198	551017	241674	172948	159305	153995	152038	150459
MEA recalimer duty [MW] **								
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	440,8	193,3	138,4	127,4	123,2	121,6	120,4
Cold utility								
DCC CU cooler [m3/h]	6726	6450	6526	6557	6557	6635	6719	6795
Lean amine cooler [m3/h]	3889	29259	9707	5078	4126	4069	3778	3625
Desorber reflux condenser [m3/h]	2225	4820	3058	2275	2128	2128	2137	2146
Desorber cooler[m3/h]	2098	9979	5005	2712	2240	2105	2025	1971
Total cold utility consumption	14938	50508	24295	16622	15051	14937	14659	14536
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	13,5	6,5	4,4	4,0	4,0	3,9	3,9
Electricity consumers								
Transport fan [kW]	21623	17776	18749	19716	20667	21623	22574	23520
DCC CU circulation pump [kW]	437	453	450	447	444	443	441	439
Lean amine pump [kW]	669	5213	1278	776	687	673	639	643
Rich amine pump [kW]	815	6345	1580	965	854	815	796	785
Total electricity consumption	188358556	238297532	176458613	175232836	181219601	188433227	195605395	203103042
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	95,3	70,6	70,1	72,5	75,4	78,2	81,2
Total utility cost [MNOK/yr]	203	550	270	213	204	203	204	205

Appendix 27 Case study 2C: Process overview

Case step [Stages in absorber]	**Base case**	**14**	**15**	**16**	**17**

Economic evaluation

CAPEX equipment installed cost [MNOK]	1400	3083	1677	1641	1539
Income CO2 [NOK/t CO2]	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	512	243	215	161
Operating expenditure - cold utility [MNOK/yr]	4,0	15,4	8,7	7,7	5,5
Operating expenditure - electricity [MNOK/yr]	75,3	111	80	81	80
OPEX total [MNOK/yr]	203	638	331	304	247
Uptime [h/yr]	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	9838	5187	4860	4161

Performance

Specific energy consumption [MJ/kg CO2]	3,61	14,31	6,71	5,95	4,47
CO2 removed [%]	85,2	89,1	90,0	90,1	90,0
CO2 removed [t/y]	1230134	1287327	1300438	1301535	1300382
CO2 removed total [Mtons]	24,6	26	26	26	26
Total specific cost [NOK/t CO2]	222	615	319	297	249

Process performance

Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,29	0,35	0,36	0,41
Absorber gas feed temperature [degC]	40,7	40,3	40,5	40,6	40,7
Absorber gas feed pressure [kPa]	121	119,13	120,06	121	121,94
Absorber gas feed flowrate [Am3/h]	2360705	2395117	2377179	2359321	2341728
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40
Lean amine rate [tons/h]	3600	34973	9500	7900	5200
Number of stages in absorber [-]	16	14	15	16	17
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	110,2	109,7	109,7	109,6
Minimum approach temp in L/R heat exchanger	10,0	10,0	10,0	10,0	10,0
Number of stages in desorber [-]	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500

Appendix 27 Case study 2C: Process overview

Case step [Stages in absorber]	**18**	**19**	**20**	**21**	**22**	**23**
Economic evaluation						
CAPEX equipment installed cost [MNOK]	1528	1550	1580	1616	1653	1688
Income CO2 [NOK/t CO2]	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	140	134	130	128	129	127
Operating expenditure - cold utility [MNOK/yr]	4,6	4,4	4,2	4,2	4,2	4,1
Operating expenditure - electricity [MNOK/yr]	82	85	88	90	93	96
OPEX total [MNOK/yr]	227	223	222	223	226	227
Uptime [h/yr]	8000	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3936	3910	3935	3979	4052	4096
Performance						
Specific energy consumption [MJ/kg CO2]	3,88	3,69	3,60	3,55	3,55	3,51
CO2 removed [%]	90,1	90,2	90,1	90,1	90,6	90,1
CO2 removed [t/y]	1302007	1302369	1302244	1301612	1308655	1301964
CO2 removed total [Mtons]	26	26	26	26	26	26
Total specific cost [NOK/t CO2]	233	231	231	233	236	239
Process performance						
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,45	0,46	0,47	0,47	0,47	0,48
Absorber gas feed temperature [degC]	40,8	40,9	41,0	41,1	41,2	41,3
Absorber gas feed pressure [kPa]	122,88	123,81	124,75	125,69	126,63	127,56
Absorber gas feed flowrate [Am3/h]	2324396	2307496	2290662	2274071	2257717	2241764
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40
Lean amine rate [tons/h]	4250	3940	3800	3720	3720	3655
Number of stages in absorber [-]	18	19	20	21	22	23
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,6	109,6	109,5	109,5	109,3	109,2
Minimum approach temp in L/R heat exchanger	10,0	9,9	10,0	9,9	10,0	10,1
Number of stages in desorber [-]	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500

Appendix 28 Case study 2C: Cost overview

Case step [Stages in absorber]	**Base case**	**14**	**15**	**16**	**17**
Installed cost equipment in MNOK					
Transport fan	50	50	50	50	50
Direct contact cooler	148	148	148	148	148
Absorber column	858	771	815	858	900
Lean amine cooler	22	101	49	44	31
Lean amine circulation pump	12	42	20	18	15
Lean/rich heat exchangers	220	1079	416	366	278
Desorber column	47	768	105	90	63
Desorber condenser	6	9	7	7	6
Desorber reboiler	19	55	33	30	25
Desorber OH condenser	5	12	10	9	7
Desorber OH water separator	1	1	1	1	1
Rich amine circulation pump	13	46	22	20	16
Total installed cost [MNOK]	1400	3083	1677	1641	1539
Uptime [h/yr]	8000	8000	8000	8000	8000
Utility consumption					
Hot utility					
Desorber duty [kW]	154198	639462	303149	268849	201870
MEA recalimer duty [MW] **					
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	511,6	242,5	215,1	161,5
Cold utility					
DCC CU cooler [m3/h]	6726	6557	6564	6649	6724
Lean amine cooler [m3/h]	3889	35661	14892	12804	7236
Desorber reflux condenser [m3/h]	2225	5053	4110	3738	2895
Desorber cooler[m3/h]	2098	10403	6863	5848	3579
Total cold utility consumption [m3/h]	14938	57674	32429	29038	20433
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	15,4	8,7	7,7	5,5
Electricity consumers					
Transport fan [kW]	21623	19716	20667	21623	22574
DCC CU circulation pump [kW]	437	447	446	444	442
Lean amine pump [kW]	669	6518	1758	1460	953
Rich amine pump [kW]	815	7920	2169	1805	1182
Total electricity consumption [kWh/yr]	188358556	276809620	200319947	202654964	201213733
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	110,7	80,1	81,1	80,5
Total utility cost [MNOK/yr]	203	638	331	304	247

Appendix 28 Case study 2C: Cost overview

Case step [Stages in absorber]	**18**	**19**	**20**	**21**	**22**	**23**
Installed cost equipment in MNOK						
Transport fan	50	50	50	50	50	50
Direct contact cooler	148	148	148	148	148	147
Absorber column	942	983	1023	1063	1103	1142
Lean amine cooler	25	23	22	22	22	21
Lean amine circulation pump	13	12	12	12	12	12
Lean/rich heat exchangers	245	235	228	227	225	222
Desorber column	55	51	49	48	48	48
Desorber condenser	6	6	6	6	6	6
Desorber reboiler	23	22	22	21	21	21
Desorber OH condenser	6	5	5	5	5	5
Desorber OH water separator	1	1	1	1	1	1
Rich amine circulation pump	14	14	13	13	13	13
Total installed cost [MNOK]	1528	1550	1580	1616	1653	1688
Uptime [h/yr]	8000	8000	8000	8000	8000	8000
Utility consumption						
Hot utility						
Desorber duty [kW]	175444	167028	162956	160601	161084	158699
MEA recalimer duty [MW] **						
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	140,4	133,6	130,4	128,5	128,9	127,0
Cold utility						
DCC CU cooler [m3/h]	6808	6884	6968	7043	7119	7203
Lean amine cooler [m3/h]	5312	4688	4299	4053	4003	3801
Desorber reflux condenser [m3/h]	2555	2462	2397	2383	2380	2350
Desorber cooler[m3/h]	2666	2365	2205	2142	2119	2053
Total cold utility consumption [m3/h]	17341	16398	15869	15622	15620	15407
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,6	4,4	4,2	4,2	4,2	4,1
Electricity consumers						
Transport fan [kW]	23520	24450	25386	26316	27241	28152
DCC CU circulation pump [kW]	440	439	437	435	434	432
Lean amine pump [kW]	798	721	702	676	675	674
Rich amine pump [kW]	964	892	860	841	841	826
Total electricity consumption [kWh/yr]	205778547	212017921	219077205	226143428	233528292	240668609
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	82,3	84,8	87,6	90,5	93,4	96,3
Total utility cost [MNOK/yr]	227	223	222	223	226	227

Appendix 29 Case study 3A: Process overview

Case step [Absorber feed gas temperature]	Base case	35	40	45	50
Economic evaluation					
CAPEX equipment installed cost [MNOK]	1400	1232	1222	1235	1240
Income CO2 [NOK/t CO2]	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	119	122	126	133
Operating expenditure - cold utility [MNOK/yr]	4,0	4,0	3,9	3,6	3,5
Operating expenditure - electricity [MNOK/yr]	75,3	66	63	63	62
OPEX total [MNOK/yr]	203	189	189	193	199
Uptime [h/yr]	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	3233	3223	3277	3347
Performance					
Specific energy consumption [MJ/kg CO2]	3,61	3,50	3,58	3,70	3,90
CO2 removed [%]	85,2	85,1	85,1	85,1	85,1
CO2 removed [t/y]	1230134	1227414	1228704	1229131	1229207
CO2 removed total [Mtons]	24,6	24,5	24,6	24,6	24,6
Total specific cost [NOK/t CO2]	222	204	203	207	212
Process performance					
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,48	0,47	0,46	0,45
Absorber gas feed temperature [degC]	40,7	34,8	39,7	44,6	49,8
Absorber gas feed pressure [kPa]	121	117,25	117,25	117,25	117,25
Absorber gas feed flowrate [Am3/h]	2360705	2350337	2426047	2511638	2622909
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40
Lean amine rate [tons/h]	3600	3460	3540	3710	3980
Number of stages in absorber [-]	16	12	12	12	12
Murphree efficiency in absorber [m^-1]	0,15	variable	variable	variable	variable
Rich amine pump pressure [kPa a]	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	109,3	109,4	109,5	109,7
dTmin in L/R heat exchanger [degC]	10,0	10,0	10,0	10,0	9,9
Number of stages in desorber [-]	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500

Appendix 30 Case study 3A: Cost overview

Case step [Absorber feed gas temperature]	**Base case**	**35**	**40**	**45**	**50**
Installed cost equipment in MNOK					
Transport fan	50	50	50	50	50
Direct contact cooler	148	169	150	134	113
Absorber column	858	662	680	701	729
Lean amine cooler	22	16	21	26	30
Lean amine circulation pump	12	12	12	12	13
Lean/rich heat exchangers	220	233	219	217	206
Desorber column	47	45	46	48	51
Desorber condenser	6	6	6	6	6
Desorber reboiler	19	20	21	21	22
Desorber OH condenser	5	5	5	5	5
Desorber OH water separator	1	1	1	1	1
Rich amine circulation pump	13	13	13	13	14
Total installed cost [MNOK]	1400	1232	1222	1235	1240
Uptime [h/yr]	8000	8000	8000	8000	8000
Utility consumption					
Hot utility					
Desorber duty [kW]	154198	149207	152538	158038	166438
MEA recalimer duty [MW] **					
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	119,4	122,0	126,4	133,2
Cold utility					
DCC CU cooler [m3/h]	6726	8771	6614	4061	508
Lean amine cooler [m3/h]	3889	2190	3675	5142	7629
Desorber reflux condenser [m3/h]	2225	2149	2149	2269	2430
Desorber cooler[m3/h]	2098	1944	2014	2191	2493
Total cold utility consumption [m3/h]	14938	15053	14451	13663	13059
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	4,0	3,9	3,6	3,5
Electricity consumers					
Transport fan [kW]	21623	17776	17776	17848	17776
DCC CU circulation pump [kW]	437	1297	481	197	18
Lean amine pump [kW]	669	628	643	690	747
Rich amine pump [kW]	815	778	800	845	914
Total electricity consumption [kWh/yr]	188358556	163831047	157604636	156644460	155645399
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	65,5	63,0	62,7	62,3
Total utility cost [MNOK/yr]	203	189	189	193	199

Appendix 31 Case study 3B: Process overview

Case step [Absorber feed gas temperature]	Base case	35	40	45	50
Economic evaluation					
CAPEX equipment installed cost [MNOK]	1400	1391	1407	1426	1437
Income CO2 [NOK/t CO2]	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	123	127	132	138
Operating expenditure - cold utility [MNOK/yr]	4,0	4,1	4,0	3,8	3,7
Operating expenditure - electricity [MNOK/yr]	75,3	78	76	75	75
OPEX total [MNOK/yr]	203	205	207	211	217
Uptime [h/yr]	8000	8000	8000	8000	8000
Number of years	20	20	20	20	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	3548	3567	3596	3656	3734
Performance					
Specific energy consumption [MJ/kg CO2]	3,61	3,42	3,52	3,65	3,83
CO2 removed [%]	85,2	90,1	90,1	90,1	90,1
CO2 removed [t/y]	1230134	1300282	1300610	1301555	1300990
CO2 removed total [Mtons]	24,6	26,0	26,0	26,0	26,0
Total specific cost [NOK/t CO2]	222	211	213	217	222
Process performance					
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,48	0,48	0,47	0,46
Absorber gas feed temperature [degC]	40,7	35,0	40,2	45,0	50,0
Absorber gas feed pressure [kPa]	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2276971	2353540	2435784	2536764
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40
Lean amine rate [tons/h]	3600	3530	3670	3840	4080
Number of stages in absorber [-]	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	variable	variable	variable	variable
Rich amine pump pressure [kPa a]	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	108,7	109,4	109,5	109,6
dTmin in L/R heat exchanger [degC]	10,0	10,0	9,9	10,0	10,0
Number of stages in desorber [-]	12	12	12	12	12
Murphree efficiency in desorbe [m^-1]	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500

Appendix 32 Case study 3B: Cost overview

Case step [Absorber feed gas temperature]	Base case	35	40	45	50
Installed cost equipment in MNOK					
Transport fan	50	50	50	50	50
Direct contact cooler	148	169	149	134	116
Absorber column	858	831	856	883	916
Lean amine cooler	22	15	21	26	30
Lean amine circulation pump	12	12	12	12	13
Lean/rich heat exchangers	220	223	225	223	210
Desorber column	47	47	48	50	52
Desorber condenser	6	6	6	6	6
Desorber reboiler	19	21	21	22	23
Desorber OH condenser	5	5	5	5	5
Desorber OH water separator	1	1	1	1	1
Rich amine circulation pump	13	13	13	14	14
Total installed cost [MNOK]	1400	1391	1407	1426	1437
Uptime [h/yr]	8000	8000	8000	8000	8000
Utility consumption					
Hot utility					
Desorber duty [kW]	154198	154265	158871	164758	172965
MEA recalimer duty [MW] **					
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	123,4	127,1	131,8	138,4
Cold utility					
DCC CU cooler [m3/h]	6726	9091	6941	4404	1111
Lean amine cooler [m3/h]	3889	2101	3741	5482	7717
Desorber reflux condenser [m3/h]	2225	2182	2290	2277	2528
Desorber cooler[m3/h]	2098	1908	2061	2241	2530
Total cold utility consumption [m3/h]	14938	15282	15033	14404	13886
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	4,1	4,0	3,8	3,7
Electricity consumers					
Transport fan [kW]	21623	21623	21623	21623	21623
DCC CU circulation pump [kW]	437	1271	471	204	39
Lean amine pump [kW]	669	651	671	710	751
Rich amine pump [kW]	815	792	829	874	936
Total electricity consumption [kWh/yr]	188358556	194694764	188754379	187283808	186789861
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4
Cost electricity [MNOK/yr]	75,3	77,9	75,5	74,9	74,7
Total utility cost [MNOK/yr]	203	205	207	211	217

Appendix 33 Case study 4A: Process overview

Case step [Calculation period]	**10**	**15**	**20**	**10**	**15**	**20**
Economic evaluation						
CAPEX equipment installed cost [MNOK]	1400	1400	1400	1400	1400	1400
Income CO2 [NOK/t CO2]	0	0	0	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4	0,4	0,4	0,4
Operating expenditure - hot utility	62	62	62	123	123	123
Operating expenditure - cold utility	2,0	2,0	2,0	4,0	4,0	4,0
Operating expenditure - electricity	37,7	37,7	37,7	75,3	75,3	75,3
OPEX total [MNOK/yr]	101	101	101	203	203	203
Uptime [h/yr]	4000	4000	4000	8000	8000	8000
Number of years	10	15	20	10	15	20
Calculation discount rate	0,07	0,07	0,07	0,07	0,07	0,07
Total NPV [MNOK]	2112	2323	2474	2824	3246	3548
Performance						
Specific energy consumption [MJ/kg CO2]	3,61	3,61	3,61	3,61	3,61	3,61
CO2 removed [%]	85,2	85,2	85,2	85,2	85,2	85,2
CO2 removed [t/y]	615067	615067	615067	1230134	1230134	1230134
CO2 removed total [Mtons]	6,2	9,2	12,3	12,3	18,5	24,6
Total specific cost [nominal tot NOK/t CO2]	392	317	279	279	241	222
Process performance						
Lean amine MEA wt%	0,29	0,29	0,29	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,47	0,47	0,47	0,47	0,47
Absorber gas feed temperature [degC]	40,7	40,7	40,7	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360705	2360705	2360705	2360705	2360705
CO2 in inlet gas [mole%]	0,037	0,037	0,037	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067	0,067	0,067	0,067
Lean amine temperature	40	40	40	40	40	40
Lean amine rate [tons/h]	3600	3600	3600	3600	3600	3600
Number of stages in absorber [-]	16	16	16	16	16	16
Murphree efficiency in absorber [m^-1]	0,15	0,15	0,15	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750	750	750	750
Heated rich amine temperature [degC]	109,5	109,5	109,5	109,5	109,5	109,5
dTmin in L/R heat exchanger [degC]	10,0	10,0	10,0	10,0	10,0	10,0
Number of stages in desorber [-]	12	12	12	12	12	12
Murphree efficiency in desorber [m^-1]	0,5	0,5	0,5	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120	120	120	120
Pressure top desorber [kPa a]	200	200	200	200	200	200
Pressure in reboiler [kPa a]	200	200	200	200	200	200
Lean amine pump pressure [kPa a]	700	700	700	700	700	700
Cold utility inlet temperature [degC]	15	15	15	15	15	15
Cold utility outlet temperature [degC]	25	25	25	25	25	25
Hot utility inlet temperature [degC]	160	160	160	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500	500	500	500

Appendix 34 Case study 5A: Process overview

Case step [Amine fluid package]	Base case	Li-Mather directly applied to BC	Li Mather 85,2
Economic evaluation			
CAPEX equipment installed cost [MNOK]	1400	1398	1412
Income CO2 [NOK/t CO2]	0	0	0
Cost hot utility [NOK/kWh]	0,1	0,1	0,1
Cost cold utility [NOK/m3]	0,033	0,033	0,033
Cost electricity [NOK/kWh]	0,4	0,4	0,4
Operating expenditure - hot utility [MNOK/yr]	123	121	126
Operating expenditure - cold utility [MNOK/yr]	4,0	3,9	4,1
Operating expenditure - electricity [MNOK/yr]	75,3	75,3	75,6
OPEX total [MNOK/yr]	203	201	206
Uptime [h/yr]	8000	8000	8000
Number of years	20	20	20
Calculation discount rate	0,07	0,07	0,07
Total NPV [MNOK]	3548	3523	3591
Performance			
Specific energy consumption [MJ/kg CO2]	3,61	3,60	3,69
CO2 removed [%]	85,2	84,0	85,2
CO2 removed [t/y]	1230134	1213118	1230434
CO2 removed total [Mtons]	24,6	24,3	24,6
Total specific cost [NOK/t CO2]	222	223	225
Process performance			
Lean amine MEA wt%	0,29	0,29	0,29
Lean loading [mole CO2/mole MEA]	0,26	0,26	0,26
Rich loading [mole CO2/mole MEA]	0,47	0,46	0,45
Absorber gas feed temperature [degC]	40,7	40,7	40,7
Absorber gas feed pressure [kPa]	121	121	121
Absorber gas feed flowrate [Am3/h]	2360705	2360706	2360706
CO2 in inlet gas [mole%]	0,037	0,037	0,037
Water in inlet gas [mole%]	0,067	0,067	0,067
Lean amine temperature	40	40	40
Lean amine rate [tons/h]	3600	3599	3800
Number of stages in absorber [-]	16	16	16
Murphree efficiency in absorber [%/m]	0,15	0,15	0,15
Rich amine pump pressure [kPa a]	750	750	750
Heated rich amine temperature [degC]	109,5	109,5	109,6
Minimum approach temp in L/R heat exchanger [degC]	10,0	10,1	10,0
Number of stages in desorber [-]	12	12	12
Murphree efficiency in desorber [%/m]	0,5	0,5	0,5
Reflux ratio in desorber [-]	0,4	0,4	0,4
Reboiler temperature [degC]	120	120	120
Pressure top desorber [kPa a]	200	200	200
Pressure in reboiler [kPa a]	200	200	200
Lean amine pump pressure [kPa a]	700	700	700
Cold utility inlet temperature [degC]	15	15	15
Cold utility outlet temperature [degC]	25	25	25
Hot utility inlet temperature [degC]	160	160	160
Hot utility inlet pressure [kPa a]	500	500	500

Appendix 35 Case study 5A: Cost overview

Installed cost equipment in MNOK	BC w/K-E	Li-Mather on BC	Li-Mather 85,2
Transport fan	50	50	50
Direct contact cooler	148	148	148
Absorber column	858	858	858
Lean amine cooler	22	22	23
Lean amine circulation pump	12	12	12
Lean/rich heat exchangers	220	220	228
Desorber column	47	46	48
Desorber condenser	6	6	6
Desorber reboiler	19	19	19
Desorber OH condenser	5	5	5
Desorber OH water separator	1	1	1
Rich amine circulation pump	13	13	13
Total installed cost [MNOK]	1400	1398	1412

Uptime [h/yr]	8000	8000	8000

Utility consumption

Hot utility

Desorber duty [kW]	154198	151675	157617
MEA recalimer duty [MW] **			
Cost hot utility [NOK/kWh]	0,1	0,1	0,1
Cost hot utility [MNOK/yr]	123,4	121,3	126,1

Cold utility

DCC CU cooler [m3/h]	6726	6726	6726
Lean amine cooler [m3/h]	3889	3889	4172
Desorber reflux condenser [m3/h]	2225	2220	2220
Desorber cooler[m3/h]	2098	1944	2077
Total cold utility consumption [m3/h]	14938	14779	15195
Cost cold utility [NOK/m3]	0,0334	0,0334	0,0334
Cost cold utility [MNOK/yr]	4,0	3,9	4,1

Electricity consumers

Transport fan [kW]	21623	21624	21624
DCC CU circulation pump [kW]	437	437	437
Lean amine pump [kW]	669	657	694
Rich amine pump [kW]	815	815	861
Total electricity consumption [kWh/yr]	188358556	188261436	188931952
Cost electricity [NOK/kWh]	0,4	0,4	0
Cost electricity [MNOK/yr]	75,3	75,3	76

Total utility cost [MNOK/yr]	203	201	206