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FMH606 Master's Thesis 2021

Energy and Environmental Technology

Process simulation and automated cost optimization of CO2 capture using Aspen HYSYS

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www.usn.no

Course: FMH606 Master's Thesis, 2021

Title: Process simulation and automated cost optimization of CO_2 capture using Aspen HYSYS

Number of pages: 81

Keywords: CO_2 capturing, Aspen HYSYS and Cost estimation

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Summary:

This thesis is based on earlier work conducted by the USN, HSN, and TUC on cost estimation and optimization of CO_2 capturing from flue gas using monoetahnol amine absorption (MEA).

A simulation model has been implemented in the Aspen HYSYS V10 to simulate the CO_2 removal process by using the calculations in spreadsheets. Spreadsheets have been used to compute capital expenditure (CAPEX), operational expenditure (OPEX), equipment dimensioning, and removal efficiency. Prices for the base cases were calculated in Aspen In-Plant Cost Estimator V10, and the power-law equation was applied to account for new equipment dimensions. The tools case study, Aspen simulation workbook, and Visual Basic for Application (VBA) in Excel have been used as solutions to automate the simulation. The chance of making a mistake when selecting the appropriate installation factor and subfactors for each equipment has been eliminated through the VBA code, which does it automatically.

The best trade-off between heat exchanger area and energy consumption has been obtained for the minimum approach temperature (ΔT_{min}) in a lean-rich heat exchanger at 9 degrees Celsius (°C). In addition, the optimal number of absorber stages in the process has been determined to be 15 stages, and the gas through the absorber has an optimal superficial velocity of 2 to 2.2 m/s.

With this model, iterative cost estimation of CO_2 absorption and desorption processes can be implemented automatically and instantly. Human errors in selecting installation factors and subfactors for different equipment are also eliminated.

Preface

This project was written as the thesis for the online master's degree program Energy and Environmental Technology(EET) at the University of South-Eastern Norway (USN).

I would like to take the opportunity to express my gratitude to the people who helped me in completing my thesis.

First of all, I would like to thank Lars Erik Øi for their supervision. Based on their constructive feedback and support, I was challenged to broaden my view on the subject, and their ideas gave me new insights for my thesis. They also helped me with formulating the thesis, which does the thesis as it is now. Thereafter, I would like to thank Solomon Aforkoghene Aromada, my co-supervisor, for answering my questions and assisting me to complete this thesis successfully.

Finally, I like to thank my family and friends for their encouragement and support during this period. This thesis indeed would not have been possible without your support.

Porsgrunn, 30.09.2021

Pouya Rahmani

Contents

Nomenclature

CAPEX[€], Capital expenditure

CS, Carbon steel

DCC, Direct contact cooler

ΔT_{min}[°C], Minimum approach temperature

EDF, Enhanced Detailed Factor

k€, ×1000 Euro

MEA, Monoethanolamine

OPEX, Operational expenditure

PFD, Process Flow Diagram

SS, Stainless steel

VBA, Visual Basic for Application

1 Introduction

This study is based on the previous studies completed in the USN, HSN, and TUC in the context of the estimation of carbon dioxide (CO2) capture by monoethanolamine (MEA) absorption in the post-combustion.

1.1 Background

The main goal of the Paris Agreement is to establish a worldwide framework for avoiding catastrophic climate change by keeping global warming far below 2 degrees Celsius (°C) and pursuing efforts to keep it below 1.5°C. Reducing Greenhouse Gas (GHG) emissions has been identified as one of the most effective ways to achieve this goal [1].

CO2 emissions, as one of the main GHG emissions, should be managed and decreased to the greatest extent. The carbon capture and storage (CCS) process, which consists of the capturing and storing of CO2, is a proper method to encourage industrial plants to reduce their CO2 emissions. CO_2 capturing, which is generated by fossil and biomass power plants and industrial facilities, is one of the activities that can be helpful in the reduction of the greenhouse gas effects on the world. Nowadays, the world's capacity for capturing and storing CO_2 is around 40 million tons [2].

Amine-based CO_2 absorption has been identified as the most appropriate solution for combustion-based power plants, especially in the power plants with low CO_2 concentrations in the flue gas. This solution has been tested, and acceptable results have been achieved. There is a technical similarity with the end-of-pipe control system as well as, extensive studies in this area can be found for optimizing the CO_2 capturing process [3].

The flue gas flow rate, CO_2 concentration in the flue gas, CO_2 removal efficiency, amine flow rate, steam, and power price are the factors that can affect the plant's CO_2 capturing total price [4]. The optimal value for different equipment of the process should be measured to minimize the plant's CO_2 capturing price with the consideration of the required CO_2 removal efficiency.

They have investigated the effects of different parameters to optimize and reduce the total price of the CO2 capturing process. They have calculated the cost of power for an additional amount of the lean loading under the different CO2 capture efficiency scenarios. They have found an interval for lean loading to optimize the cost of power in the power plant [5].

CO_2 removal efficiency has been defined as a function of the amine flow rate, the height of the absorber packing, the temperature of the absorption process, and the temperature of the hot utility. For various forms of structured packing, the optimum gas velocity and pressure drop in the absorption column have been explored. Murphree efficiency (E_M) in the absorption calculation is considered as a constant. While E_M in the HYSYS is computed automatically and it is varying between 0.08 and 0.13 [6].

For all packing options, the trade-off between capital expenditure (CAPEX) and operational expenditure (OPEX) is investigated. The CAPEX includes a shell, packing, column internals, liquid distributor, packing support, and flue gas fan prices. For different types of packing, specific ranges for velocity and pressure drop have been identified between 2 and 2.5 m/s and 10 to 15 mbar, resppectively. The pressure drop in stages has been determined by subtracting the pressures at the top and bottom of each stage. The total pressure drop in the absorber is calculated by the summation of these pressure drops with the atmospheric pressure [7].

It is required to mention that the cost estimation method employed in the CO_2 capturing plant should be exact and accurate. Adding extra safety factors is the reason for increasing the plant's capital cost, and these calculations are not applicable in the real world. The Enhanced Detailed Factor (EDF) is a method for determining the installation factor in the plant's capital cost that has shown acceptable results. The cost of carbon capture using the EDF method with ΔT_{min} =15°C was computed at 66 €/tCO2, whereas other methods showed 69 to 79 €/tCO2. These findings show that this method can be used to calculate capital costs for a variety of plan types and conditions [8].

Automation of the cost estimation and optimization of the equipment are the topics that have received more attention in recent years. Automation of the process has been investigated to determine the trade-off between the area of the lean-rich heat exchanger and energy consumption. Another study has been done for the automation of the process in order to update the number of stages automatically in the cost estimation. They have added an adjust operation in the Aspen HYSYS model for achieving the target ΔT_{min} in the lean-rich heat exchanger and removal efficiency based on lean amine flow. These operations have significant impacts on the decreasing of the modeling time and achieve more accurate results. Prices for removal efficiency of 85 percent and 90 percent were compared to each other. The simulation's robustness should be considered as one of the suggestions to obtain better results [9].

In the simulation procedure, some assumptions and methods have been used that can be developed or replaced by better assumptions and methods in order to get more accurate results. Automation of the process is the other assumption that should also be investigated.

1.2 Objective

This master's project has many primary aims, which are highlighted below:

- Optimizing the amine-based CO2 capture in the Aspen HYSYS,
- Dimensioning and cost estimating by spreadsheet,
- Automating the optimization process,
- And evaluating the limitations in the estimation and optimization process.

Appendix A is a detailed explanation of the project's overall goal.

2 Process summary and simulation of the base case in the Aspen HYSYS

In this chapter CO_2 capturing process will be explained briefly. All the main components and their roles in the process are defined. In the second part, the specifications and assumptions of the base case are shown in Table 2-1.

2.1 Process summary

The main components of the process, the CO_2 removal procedure, and the general flow diagram of the CO2 removal power plant are shown in Figure 2-1.

A production plant's flue gas is transferred to the carbon-capturing facilities. The flue gas is driven through the pre-cooler, separator, and absorber by a fan. It should provide the flue gas with the required pressure and flow. In the pre-cooler, the temperature of the flue gas should be reduced to around 40°C. When lean-amine comes into touch with flue gas in the absorber, the CO_2 in the flue gas is absorbed. The solution passes through the lean-rich heat exchanger. The temperature of the solution is increased after the lean-rich heat exchanger by absorbing heat from the lean-amine flow. CO_2 is removed from the mixture via a stripper or desorber, and absorbent flows toward the lean-rich heat exchanger. Before entering the absorber, the temperature of the mixture is decreased to around 40°C in the lean-amine cooler. Due to the bonding, amine solutions are categorized as chemical absorbents. Physical bondings are another category for solvents families [4]. The flow diagram, which is shown in Figure 2-1, is based on the previous study [10].

2.1.1 Absorber column

In the absorber, liquid and gas have flown countercurrent. The main aim is to absorb the gas mixture to the liquid or solvent by providing contacting surface. The mass transfer happens on these surfaces(stages). The following are the major steps in the design of the absorber:[11]

- Choosing the Solvent
- Finding the most cost-effective gas velocity(absorber diameter)
- Calculating the height of the absorber, which includes the number of stages in the absorber
- Calculating the best solvent circulation rate
- Calculating temperature of streams
- Finding the operating pressure in the absorber
- Designing of the mechanical components

Designing the absorber is one of the most important parts of CO2 capturing, and here only some parts of the absorber have been explained [11].

The absorber's solvent can be chemical or physical. The amount of solubility for the desired solute is one of the key reasons for choosing the solvent's type, which is influenced by the temperature and pressure. MEA is the chemical solvent that is used in this project.

Tray and packed towers are two types of contactors used in the absorber. Structured packing is one of the most popular packed towers in commercial practice. Low-pressure drop is a key factor in making it more attractive. Structured packing is utilized in this project [11].

Another important aspect to consider while designing an absorber is the gas velocity. As the gas velocity changed, the diameter of the packing altered as well. When the diameter of the packing is decreased, the pressure drop and energy consumption increase. The flue gas's high velocity in the absorber has two negative effects:

- Loss of MEA
- Local pollution due to MEA losses

Installing a water wash downstream of the absorber can assist to reduce these negative consequences[11].

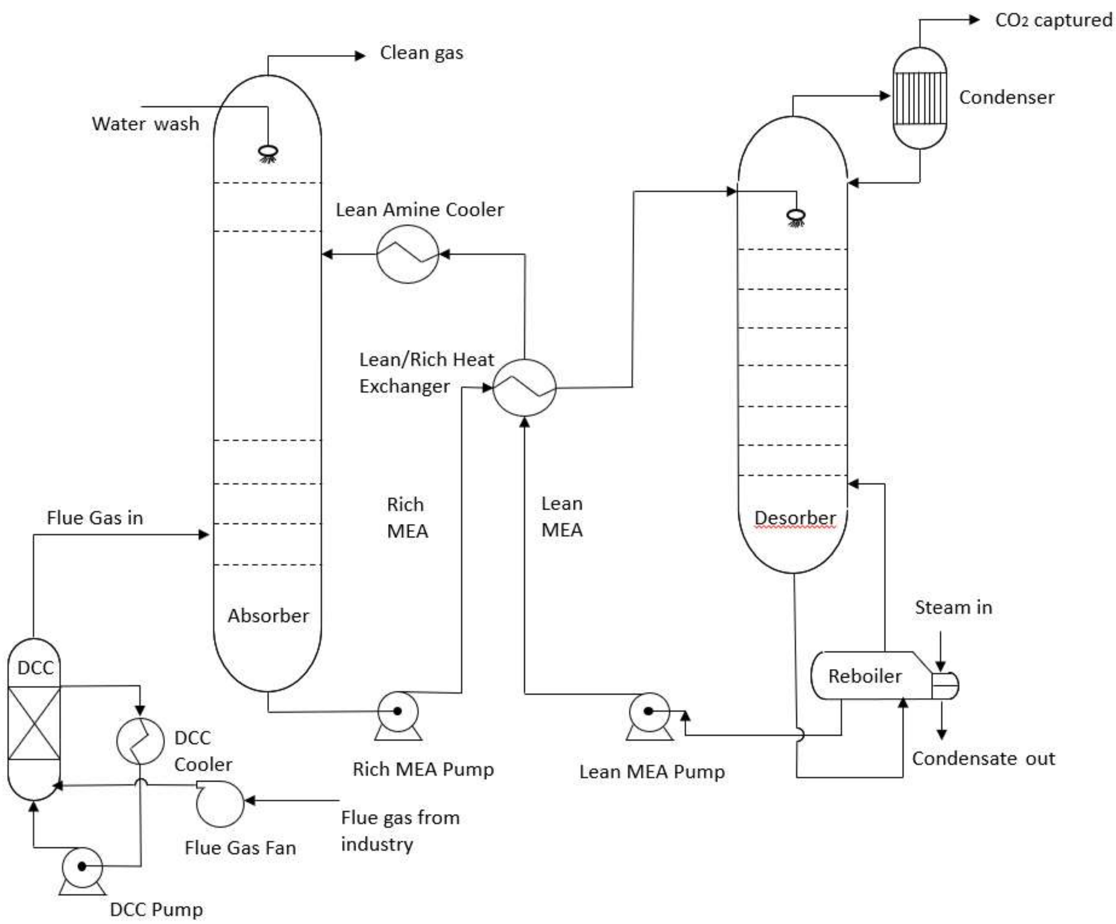

Figure 2-1: Flow diagram of the standard amine-based CO_2 capture plant[10].

2.1.2 Desorber column

Structured packing, reboiler, and condenser are the key components of the desorber. In the desorber, CO_2 is removed from the circulated amine solution. For regeneration of the amine from the circulating solution should be added heat. The liquid solution flows from the bottom of the desorber to the reboiler, where the heat from the steam is absorbed as a hot resource (kallevik].[4] A thermosiphon vertical fixed tube sheet(V-FXD) reboiler has been used for cost estimation in this project. This sort of reboiler is quite common, and it's usually used in one of the following situations: [12]

- The constant head above the reboiler
- Low operating pressure
- The reboiler feed contains a high concentration of volatiles(over 5%)

Effluent flow from the desorber includes water and CO_2. Water can be condensed and return to the process. Lean amine solution returned to the process from the bottom of the reboiler [4].

2.1.3 Water separator

There may be a small amount of water in the flue gas before it goes through the absorber due to the temperature reduction in the pre-cooler. In the separator, water can be separated from the gas.

2.1.4 Lean-rich heat exchanger

In the lean-rich heat exchanger, the heat added to the lean solution in the reboiler can be transferred to the rich amine solution. This heat exchanger is playing the role of pre-cooling for the lean solution. One of the most expensive pieces of equipment in this CO_2 capture process is the lean-rich heat exchanger. It is required to find the optimum $\Delta T_{min,}$ which can be calculated based on the trade-off between heat exchanger area and energy consumption. An increasing trend in ΔT_{min} leads to a decreasing trend in the surface area and heat recovery, and vice versa [4].

2.1.5 Lean amine cooler

The lean cooler is able to reduce the temperature of the lean solution to around 40°C. External cooling water resources are applicable to decrease the temperature. The total area of this heat exchanger is usually lower than the lean-rich heat exchanger, and all the assumptions are the same for both [4].

2.1.6 Pumps

Different pumps are needed in this process. The main pumps are rich-MEA pump, lean-MEA pump, pump in the condenser of the desorber, and cooling water pumps. All pumps, in general, should have enough head to overcome all process losses. The required head of the pump should be determined by considering the losses in the pipes, in the absorber, desorber, heat exchangers, pressure differential between the absorber and desorber, and type of the solution or liquid [4].

2.1.7 Fan

The flue gas pressure from the process is around 1atmosphere (atm), and the temperature is about 110°C. The fan should compensate for all the pressure drop in the pre-cooler, separator, and absorber. Pressure drop in the absorber can be affected by the number of packing stages and cross-sectional area of the packing. The minimum driving force of the fan is determined after the calculation of pressure drop for this equipment. The fan is chosen based on the pressure drop and the volume flow of the flue gas. Due to the large volume flow, two or more fans in parallel may be employed [4].

2.1.8 Pre-cooler before absorber

This pre-cooler is to reduce the temperature of flue gas to around 40°C.

2.2 Aspen HYSYS simulation of the base case

The process diagram of CO_2 capture, which is included in the calculations, is shown in Figure 2-1.

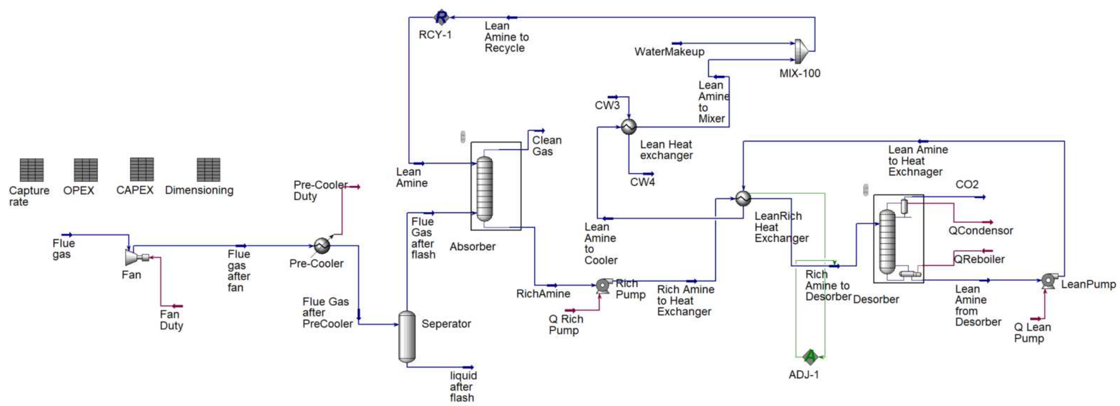

Figure 2-2: PFD (Process Flow Diagram) of the CO_2 capturing process simulated in the Aspen HYSYS.

In the PFD, all the streams have been labeled with the name of a fluid that flows to the next equipment. PFD Main equipment has been marked with name, and adjust operation has been added to the simulation in order to achieve the required ΔT_{min} in the flowsheet. Table 2-1 presents an overview of the required data for Aspen HYSYS' base case simulation.

Table 2-1: Specification and assumption for base model simulation.

Parameters	Value	Source
Co_2 capture efficiency(%)	85	[10]
Flue gas		
Temperature	110°C	Assumed
Pressure	101kPa	[4]
CO_2 mole-fraction	0.033	Assumed
H_2O mole-fraction	0.069	Assumed
N_2 mole-fraction	0.898	Assumed
Molar flow rate	10,910 kmol/h	Assumed

Inlet temperature to Absorber	40°C	[10]
Inlet pressure to absorber	120kPa	Assumed
Lean-Rich heat exchanger		
ΔT_{min}	10°C	[10]
Lean MEA		
Temperature	40°C	[10]
Pressure	110kPa	Assumed
Molar flow rate	132,100kmol/h	Calculated
Mass fraction of MEA	0.225	Assumed
Mass fraction of CO_2	0.035	Assumed
Absorber		
No. of stages	10	[9]
Murphree Efficiencies(%)	25	[13]
Packing type	M76YB	Aspen In-Plant
Desorber		
No. of stages	6	[9]
Pressure	200kPa	[9]
Murphree Efficiencies(%)	50	[10]
Reflux ratio	0.3	[10]
Temperature into desorber	103.2°C	Assumed
Reboiler temperature	120°C	[10]
Pumps		
Adiabatic efficiencis(%)	75	Assumed
Fan		
Adiabatic efficiencis(%)	75	Assumed

Structured packings M76YB are made of stainless steel 304 and have a 45° vertical orientation angle and a geometric area of 250 m2/m3. Table 2-2 summarizes an overview of the assumptions and foundation for designing the equipment dimensions. The dimensioning spreadsheet in the Aspen HYSYS has been added to Appendix D. the Table is based on a study from Aromada [10].

Table 2-2: Equipment dimensioning factors and assumptions

Equipment	Assumption
Absorber	Superficial velocity of 2.5m/s, TT=40m, 1m packing height per stage
Desorber	Superficial velocity of 1m/s, TT=15m, 1m packing height per stage
Packing	Structured packing: SS304 Mellapak 250YB
Lean/Rich heat exchanger Reboiler Condenser Coolers Intercooler pressure drop	U = 0.5kW/m^{2}K U = 0.8kW/m^{2}K U = 1kW/m^{2}K U = 0.8kW/m^{2}K 0.5bar
Pumps	Centrifugal
Flue gas fan	Centrifugal
Separators	Corrosion allowance of 0.001m; joint efficiency of 0.8; stress 2.15×10^8 Pa; TT=3Do

3 Cost estimation method

The main aim of the cost estimation is to calculate the total cost of the project and the uncertainties. Calculations are based on the dimensioning from the Aspen HYSYS V10. Prices for all the parts in the base case process have been computed from the Aspen In-Plant Cost Estimator V10. In this chapter, the calculation technique for the different dimensioning ports will be discussed.

3.1 Clasification of expenses

CAPEX and OPEX are the main expenses in the CO_2 capturing projects. CAPEX includes the cost of purchasing and installation of equipment, piping, instruments and control, electrical equipment, buildings, land, engineering and supervision, construction costs, contingency, and start-up expenses. Land, delivering the essential utility to the site, administrative buildings, and control rooms expenses are excluded from this project. OPEX consists of the operation cost, utility cost, and maintenance cost. Spare parts, building maintenance, raw material, and employee's salary are not considered in the calculations [4].

In this project, CAPEX has solely taken into account the cost of equipment. The equipment expenses at the Aspen-In Plant are calculated using costs from the first quarter of 2016. It includes an estimate for labor costs that are not included in the cost estimates.

3.2 Design data for a price estimation in Aspen-In Plant

In Aspen-In Plant software, some design data must be provided in order to achieve more accurate results. Designing parameters are explained in the four main categories. These categories are towers or column-trayed/packed, heat exchangers, pumps, and blowers.

1. Designing items in the Towers or column-trayed/packed are:

- Application
- Dimensions
- Shell material
- Vessel diameter
- Vessel tangent to tangent height
- Packing type
- Number of packed sections
- Total packing height

2. Designing items in the heat exchangers are:

- Heat transfer area
- Number of shells
- Tube material
- Shelle material

3. Designing items in the pumps are:

- Casting material
- Liquid flow rate

4. Designing items in the fans are:

- Material
- Actual gas flow rate

All these items must be specified for calculation in the Aspen In-Plant, however other input data can be used from the default data.

3.3 Equipment cost calculation

The technique which has been used for estimating the capital cost is EDF Estimation. A table related to this method has been added in Appendix C [4].

The procedure of the cost estimation in this project is based on the following steps:

- Finding the new dimensions of the equipment, which can be volume, area, heat transfer area, or duty,
- Calculating the cost of the new equipment by using the power law, which is explained in the 3.3.1and Table 3-1,
- Finding the material factor which is added in Appendix C,
- Computing the price of carbon steel through the division of the equipment price by the material factor,
- Extracting the installation factor for 2020, equipment cost, and piping from the table, which includes the equipment cost adjustment. The installation factor can be calculated from 3.3.3,
- Adjusting the prices for 2020 by using 3.3.1
- Calculating the final price for each equipment by multiplication the installation cost and the number of equipment.

Table 3-1: sizing factor for different equipmentp[10]

<table>
<tr><th>Equipment</th><th>Sizing Factors</th></tr>
<tr><td>Absorber</td><td rowspan="3">Tangent-to-tangent height(TT), packing height, inner and outer diameters</td></tr>
<tr><td>Desorber</td></tr>
<tr><td>Packing</td></tr>
<tr><td>Lean/Rich heat exchanger
Reboiler
Condenser
Coolers
Intercooler pressure drop</td><td>Heat transfer area(m^2)</td></tr>
<tr><td>Pumps</td><td>Flow rate(L/s) and power(kW)</td></tr>
</table>

Flue gas fan	Flow rate(m^3/h) and power(kW)
Separators	Outer diameters; tangent-to-tangent height(TT)

3.3.1 Power law

The cost of the new facility is derived from a similar facility with a different capacity. This relation is shown in eq.(3.1)

$$C_E = C_B(\frac{Q}{Q_B})^M \tag{3.1}$$

Where C_E = equipment cost with capacity Q

C_B = known base cost for equipment with capacity Q_B

M = constant depending on equipment type

Exponent power(e) can vary from 0.6 until 1.7, according to the facility type. [4]

3.3.2 Index adjustment

The cost of the equipment used in the process may vary from year to year due to inflation and other variables. To update the base costs published in open literature or other resources, cost indexes should be used. The price indexes are added in Appendix B. The price index equation is shown in Eq.(3.2) [4].

$$C_1 = C_2.\frac{Index_1}{Index_2} \tag{3.2}$$

Where C_1 = equipment cost in year 1

C_2 = equipment cost in year 2

$Index_1$: cost index in year 1

$Index_2$: cost index in year 2

Index adjustment data from the SSB website, which has the main role in providing these data, is used in this study and is included in Appendix B. The base costs have been extracted from Aspen In-Plant, based on a database from 2016.

3.3.3 Installation factor

Using the table of installation factor 2020 [] in AppendixC, the total price for the plant can be calculated from the CAPEX of the equipment. This table includes the direct cost, engineering, administration, commissioning, and contingency. The total installation cost is computed by applying[4].

$$C_i = C_P.[f_{TC} - f_P - f_E + f_m.(f_p + f_E)] \tag{3.3}$$

Where: C_i = Total installed cost factor for carbon steel[€]

C_P = Purchase cost for a equipment for carbon steel[€]

f_{TC} = Total installation cost factor

f_P = Piping cost factor for equipment

f_E = Equipment cost factor

f_m = Material cost factor

3.4 Assumptions for CAPEX

All the assumptions for CAPEX estimation are shown in Table 3-2. This table is drawn by the author but the idea is from Aromada [10].

Table 3-2: CAPEX assumptions

Parameter	Value	Source
Cost year	2020, January	[10]
Cost currency	Euro(€)	[10]
Method of CAPEX estimation	EDF method	[10]
Plant location	Rotterdam	[10]
Project life	25	[10]
Duration of construction	0	[4]
Discount rate	8.5%	Assumed
Material conversion factor(SS to CS)	1.75 Welded; 1.3 Machined	[4]
Annual maintenance	4% of CAPEX	[4]
Cost data year	2016, January	Aspen In-Plant

3.5 Cost of utilities or OPEX

In this project, OPEX for one year includes electricity cost, steam cost, and cooling water cost for 8000 hours per year. The one-year maintenance price is considered to be 4% of the total CAPEX price, which is added to the one-year OPEX price.

Table 3-3: Utility and maintenance cost

Electricity cost	0.06 €/kWh
Steam cost	0.015€/kWh
Cooling water	0.02€/m^3
Maintenance cost	0.04 of CAPEX

3.6 Net present value

Net present value (NPV) is a method of calculating a project's overall cost by considering capital and operating costs over a given time period. The capital cost in this calculation covers all installation expenses for the main equipment in the CO_2 capture process. Operational costs involve utility expenditure. CAPEX is assumed to start from year zero in this calculation, and the OPEX is computed from year zero. [4]

$$NPV_{OPEX} = \sum_{N=0}^{N=end} \left\{ (a) \times \frac{1}{(1+i)^N} \right\} \tag{3.4}$$

$$NPV_{OPEX} = a \times \sum_{N=0}^{N=end} \left\{ \frac{1}{(1+i)^N} \right\} \tag{3.5}$$

Where: NPV_{OPEX} = Total OPEX price for calculation period [€]

i = annual interest rate

a = annual operation cost [€]

N = number of years

The annual OPEX price is assumed constant in the project calculation period.

$$NPV = CAPEX + NPV_{OPEX} \tag{3.6}$$

The calculation period for this project is 20 years, and the discount rate is 8.5%.

4 Sensitivity Analysis

This chapter will investigate the effect of changing the different items in order to determine the best trade-off for the CO_2 capture plant. These items are the absorber packing height, ΔT_{min}, and superficial velocity in the absorber.

4.1 Approach temperature in the lean-rich heat exchanger

The goal of this subchapter is to determine the minimum approach temperature for the lean-rich heat exchanger with the lowest NPV. The heat exchanger is one of the most expensive parts of the CO_2 capture process. Optimizing the dimension of this equipment leads to a reduction in the price. By changing the lean-amine flow in the simulation, the capture rate remains constant. The NPV values are calculated in the Aspen HYSYS spreadsheet for each ΔT_{min}, and the results are shown in Appendix I. All other parameters such as flue gas temperature after pre-cooler, number of stages in the absorber, and essential items have remained constant. NPV can be computed for different ΔT_{min} values by defining a case study or manually in the Aspen HYSYS. The first case in the Aspen HYSYS case study achieved about 85% CO2 removal efficiency just by changing lean-amine flow. The energy consumption is determined by dividing the reboiler's energy consumption by the mass flow of CO_2 captured. The explained situation has been shown in Figure 4-1.

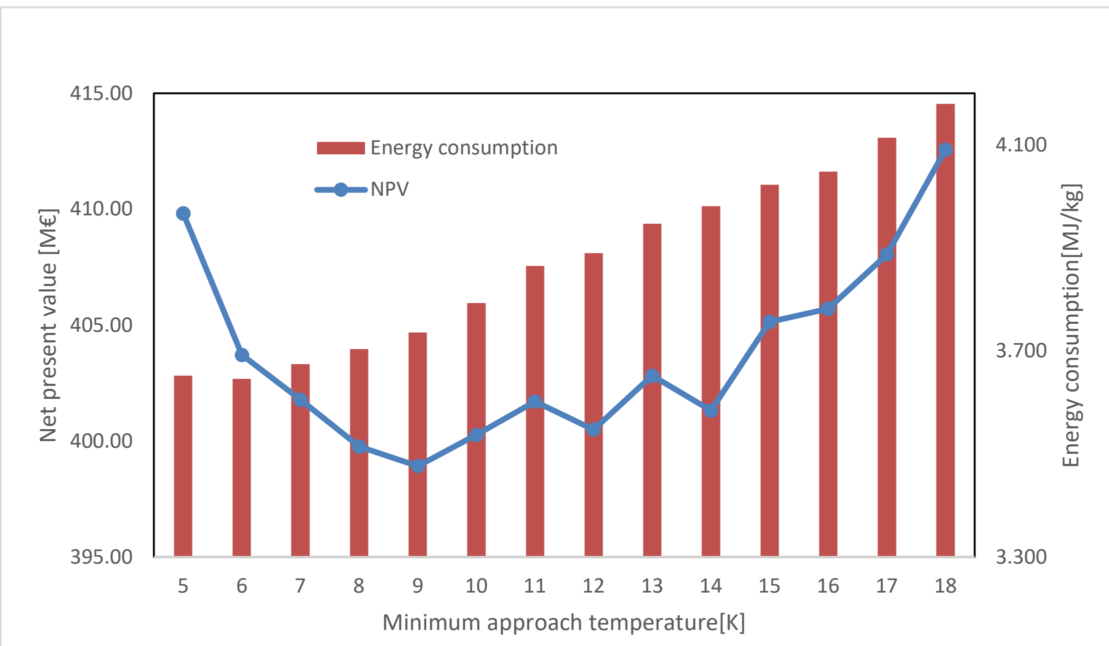

Figure 4-1: NPV and energy consumption as a function of ΔT_{min} with 85% capture rate, E_M=0.25, 20 years calculation period, and 8.5% interest rate

4.2 Absorber packing height

The optimal number of stages in the absorber can be determined in this analysis. The height of each stage is considered to be 1 meter (m) with a pressure drop of 1 kilopascal (kPa) per meter of packing. In order to reach 85%, CO_2 removal efficiency should be adjusted according to the lean-amine flow rate. Pressure in the stream from the fan to the absorber should be adjusted based on the new number of stages in the absorber. Murphree efficiency has been set to 0.15 in all stages. Aspen HYSYS automatically assigns a value of 1 to new stages, which should be updated to 0.15 in this simulation. For simulations in 4.2 and 4.3, several assumptions in the base case have been changed as follows:

- Absorber's number of stages has been changed from 10 to 16
- E_M has been set to 0.15 instead of 0.25.
- Number of stages in the desorber has been increased from 6 to 10 stages.

New capital costs have been calculated by using Aspen In-plant for the base case.

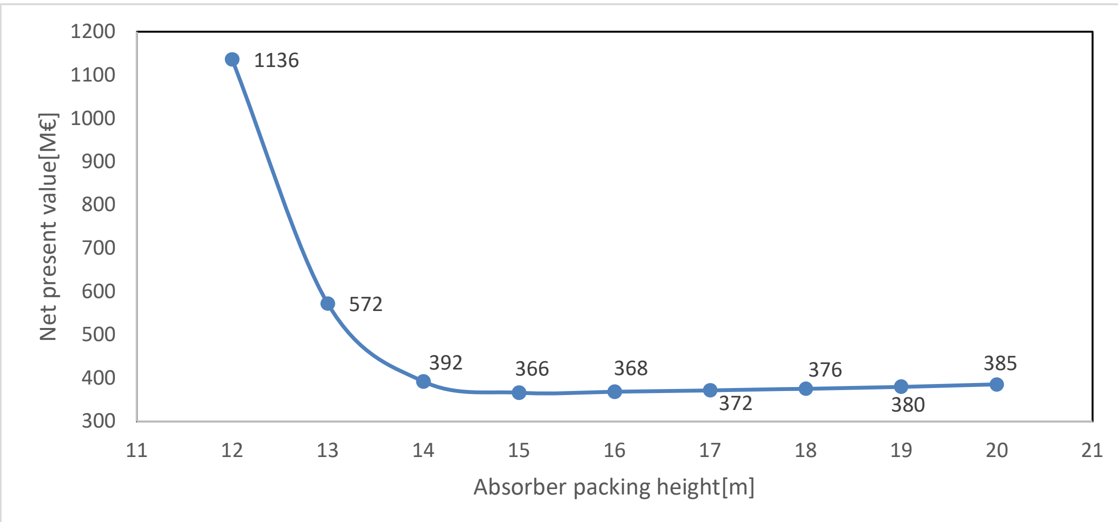

Figure 4-2: NPV as a function of absorber packing height with removal efficiency 85%, E_M=0.15, 20 years calculation period, and 8.5% interest rate

4.3 Superficial velocity in the absorber

This section has been investigated the variation of NPV by changing the gas velocity in the absorber. Only the OPEX and CAPEX of the fan and absorber are considered in the NPV. The simulation's packing type is M76YB (structured packing). Pressure drop per meter of packing for different velocities have been estimated from Appendix J. In this part, a column for the absorber's internal price has been added to the CAPEX. The costs for unit liquid distributor, packaging support, and liquid catcher were 4000, 800, and 2000$/m2 correspondingly, according to data from Dejanovic [7].

In Figure 4-3, Murphree efficiency is set at 0.15 in the simulation, and packing is assumed to remain constant for different velocities. The flue gas total pressure to the absorber has been adjusted in the Aspen HYSYS simulation based on the new pressure drop per meter of the packing. Appendix I-3 summarizes the simulation results for different velocities in the absorber.

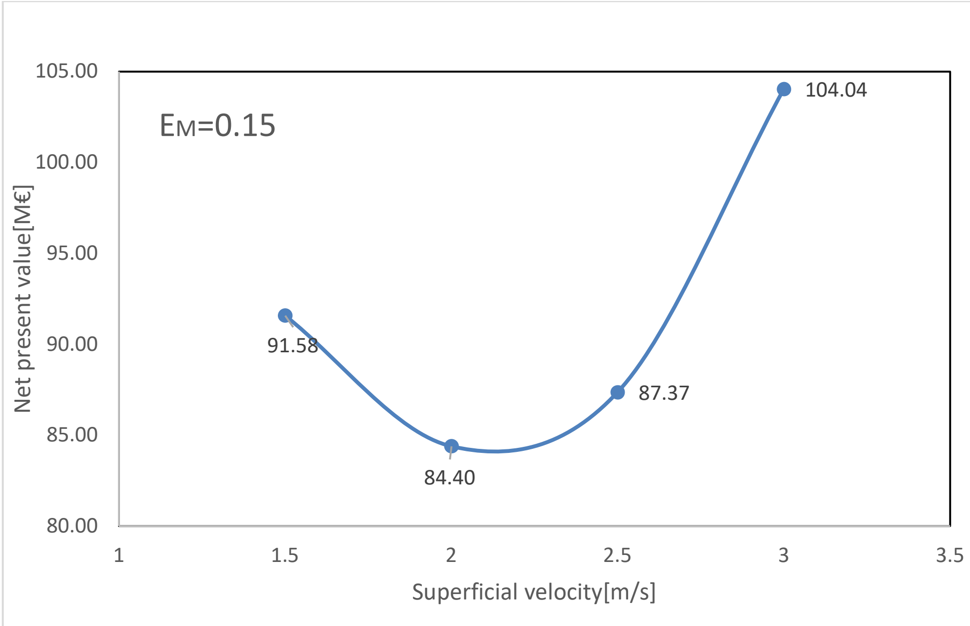

Figure 4-3: NPV as a function of superficial velocity with 85% capture rate, 20 years calculation period, and 8.5% interest rate, and constant volume in the packing

In the subsequent analysis, E_M (Base Case) for velocity 2.5 m/s is assumed to be constant at 0.15. The E_M (Base Case) value for each velocity has been computed using the percentage of the differences between the corresponding E_M value and the E_M value of 2.5m/s. For velocities of 1.5, 2, 2.5, and 3 m/s, EM is computed using the pseudo-first-order method, with area correction factors taken from Debrito-Billet provides a summary of the E_M and E_M of the base case (Base Case) [13]. Background, formulation, and the excel sheet related to these calculations have been depicted in Appendix K. Results have been presented in Appendix I-4.

Table 4-1: calculation of E_M

Superficial gas velocity(m/s)	mole flow rate, mol/(m2s)	a_{EFF}	E_M	E_M (Base Case)
3	110	0.75	0.212	0.135
2.5	92	0.7	0.234	0.15
2	74	0.65	0.265	0.17
1.5	55	0.6	0.317	0.20

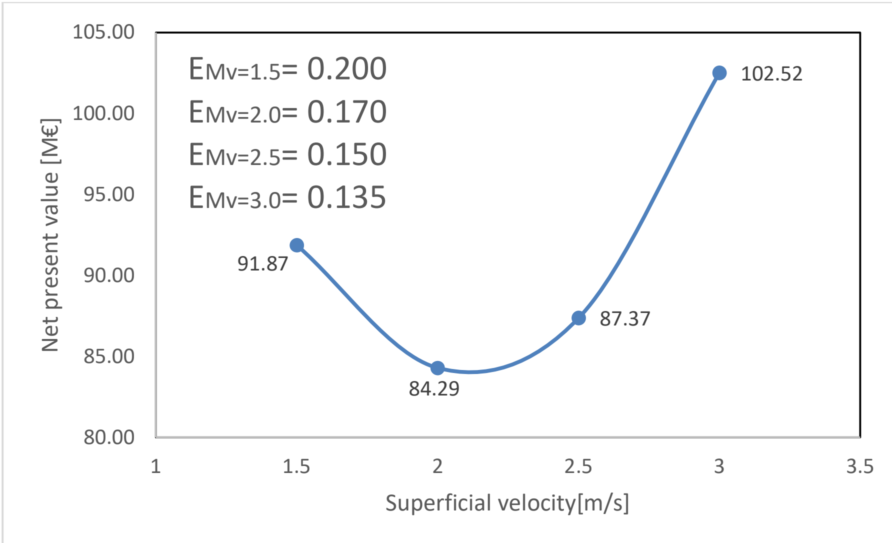

Figure 4-4: NPV as a function of superficial velocity with 85% capture rate, 20 years calculation period, and 8.5% interest rate and constant volume in the packing

5 Automation

Automation of the simulation has been investigated in this chapter, and results have been compared with manual simulation. Some of the input data should be changed in the simulations manually, which is time-consuming. Connecting Excel and Aspen HYSYS to transfer the data is the first step toward automating the process. In order to make the connection, there are different ways, including the Aspen simulation workbook and programming in visual basic. In addition, defining a case study in the Aspen HYSYS can be useful for automating the simulations.

5.1 Aspen simulation workbook

The Aspen simulation workbook is an Excel feature that can be activated through Excel's settings. The Aspen HYSYS simulation model should be linked to Excel, and it has to be done under the simulation tab in the Aspen simulation workbook. Variables in Aspen HYSYS simulation can be copied from a spreadsheet or other parts of the process and put into the organizer under the Aspen simulation workbook. There are two options to run the simulation 1) create the profile table 2) create the scenario table. Creating the profile table is more manual and should be updated with the new input data each time to run the simulation. In the second option, the scenario table, all of the input data are collected once, and the simulation runs one at a time. Results will be displayed when all of the simulations have been completed. It should be noted that there is the possibility to save the Aspen HYSYS model for each simulation. In this procedure, all the processes are automatic, but the input data should be added manually[14].

In the approach temperature simulation, ΔT_{min} is considered as input in the lean-rich heat exchanger. As well, capture rate and NPV are considered as outputs. In order to fix the capture efficiency at about 85%, a controller should be added to the simulation model. With this option, simulation is more automated. Table 5-1 summarizes the results of the automated simulation by changing ΔT_{min}. Aspen HYSYS model for $\Delta T_{min} = 5°C$ has been used as the base for other simulations. Required ΔT_{min} have been written as input in Table 5-1, and a scenario function has been run for the input data. The results will be presented in Excel. The controller for the capture rate percentage has been set to 85±0.05.

The pressure of the flue gas to the absorber, the number of stages, the pressure in the last stage, and E_M are all input variables for the simulation of changes in the height of the absorber packing. The number of stages changes the absorber's input pressure. The pressure drop in each stage is assumed to be 1kPa. NPV can be considered as an output.

Table 5-1: simulation results

		Input	Output		
Scenario	Active	ADJ-1.Target Value.Target Value	OPEX.Cell Matrix.L.18.18	Capture rate.Cell Matrix.B.3.3	Status
		C			
Case 1	*	5	417.5818	84.97652	Ready
Case 2	*	6	427.2614	84.97652	Ready
Case 3	*	7	419.8426	84.99438	Ready
Case 4	*	8	411.0337	84.99438	Ready
Case 5	*	9	409.403	84.99438	Ready
Case 6	*	10	407.2608	84.99438	Ready
Case 7	*	11	405.1125	84.99438	Ready
Case 8	*	12	413.5633	84.96599	Ready
Case 9	*	13	414.0344	84.96599	Ready
Case 10	*	14	412.6077	84.96599	Ready
Case 11	*	15	416.0912	84.98471	Ready
Case 12	*	16	416.656	84.98471	Ready
Case 13	*	17	418.3534	85.03405	Ready
Case 14	*	18	420.0309	84.99674	Ready
Case 15	*	19	420.8641	85.01249	Ready
Case 16	*		420.8641	85.01249	Ready

5.2 Case study

The case study is another solution for the automation of the process. In this method, a provided Aspen HYSYS model is used as the base, and a new case should be added to the model's case study folder. Independent and dependent variables are imported from this model to the new case study. The start point, endpoint, and step size should be added in the setup tab. The simulation has been done for the specified ranges, and results can be exported to Excel. Figure 5-1 compares the results for manual and case study (automatic) simulation. Setting the goal of 85% on capture efficiency leads to more accurate results, and consequently, the process will be more automated. The results are closer to the manual results by adding this controller, especially when the temperature is from 5°C to 12°C. These diagrams have the same minimum value for NPV in different ΔT_{min}. In both of the diagrams, the minimum value for NPV is between 8 and 9.

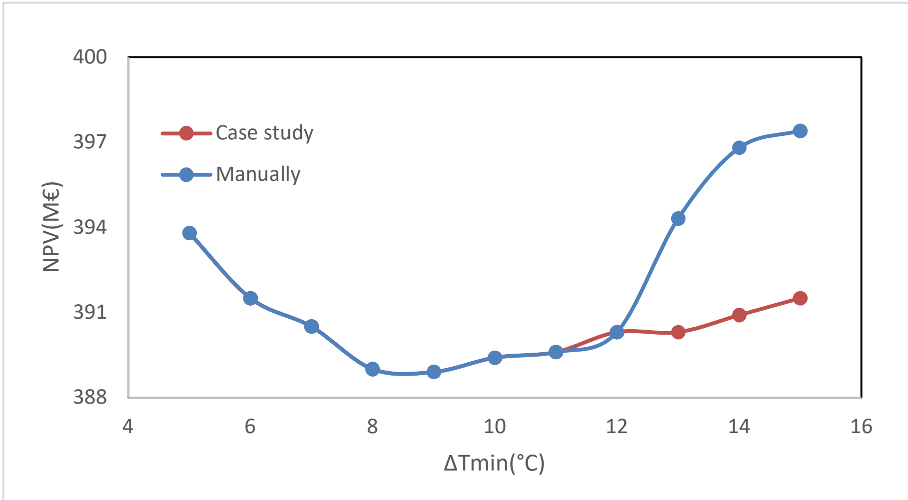

Figure 5-1: NPV as a function of ΔT_{min} with removal efficiency 85%, E_M=0.25, 20 years calculation period and 8.5% interest rate for case study(automatic) and manually, by using Aspen HYSYS model for ΔT_{min}=5°C

5.3 Visual basic for application

VBA programming language in Excel is another method for automating the process and cost estimation in Aspen HYSYS. Aspen HYSYS library can be activated in Excel from the developer tab, visual basic, tools, and preference [15]. Aspen HYSYS root should be inserted into an Excel sheet and updated for different models. In the entire process, Aspen HYSYS model is closed. One of the most time-consuming in cost estimations is determining the correct installation factor from the EDF table (Appendix C). The CAPEX spreadsheet must be updated with new equipment costs by using the new variables from the EDF table. A VBA code is written for coupling Aspen HYSYS spreadsheet and Excel. The code reads equipment prices from the Aspen HYSYS spreadsheet and imports them into an excel spreadsheet. In parallel, the total installation factors, equipment factors, and piping factors have been read from Excel to the Aspen HYSYS spreadsheet. A copy of the EDF table should be made in Excel. Appendix K contains the VBA code.

5.4 limitations

Several limitations have been identified for the automated cost estimation of CO_2 capturing by using a spreadsheet in Aspen HYSYS, which are as follows:

- Different results will be achieved for the other base models in the case study (Figure 5-2 and Figure 5-3).
- Other base models produce different results in the Aspen simulation workbook.
- It is not possible to use case study in the sensitivity analysis for different absorber heights and superficial velocity (E_M, the pressure of flue gas into the absorber, and pressure in the last stage of the absorber should be updated for each case).
- Correction to the installation factors in the CAPEX spreadsheet and Aspen simulation workbook for the case study.
- Before running the model, the input data for the Aspen simulation workbook should be imported manually in Excel.

- The coupling of the VBA code for updating the EDF installation factors with the case study and Aspen simulation workbook should be integrated.
- VBA code will be complicated for running the Aspen HYSYS directly from Excel and updating installation factors from the EDF table.
- The connection of Aspen HYSYS and Excel to achieve different running times might be regarded as the final limitation that has been found.

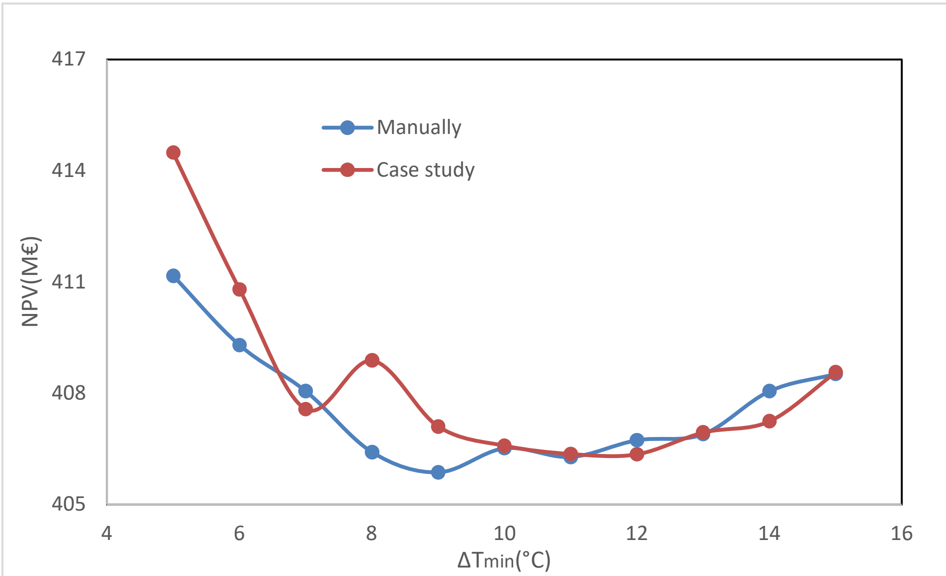

Figure 5-2:NPV as a function of ΔT_{min} with removal efficiency 85%, E_M=0.25, 20 years calculation period and 8.5% interest rate for case study(automatic) and manually, by using Aspen HYSYS model for ΔT_{min}=16°C

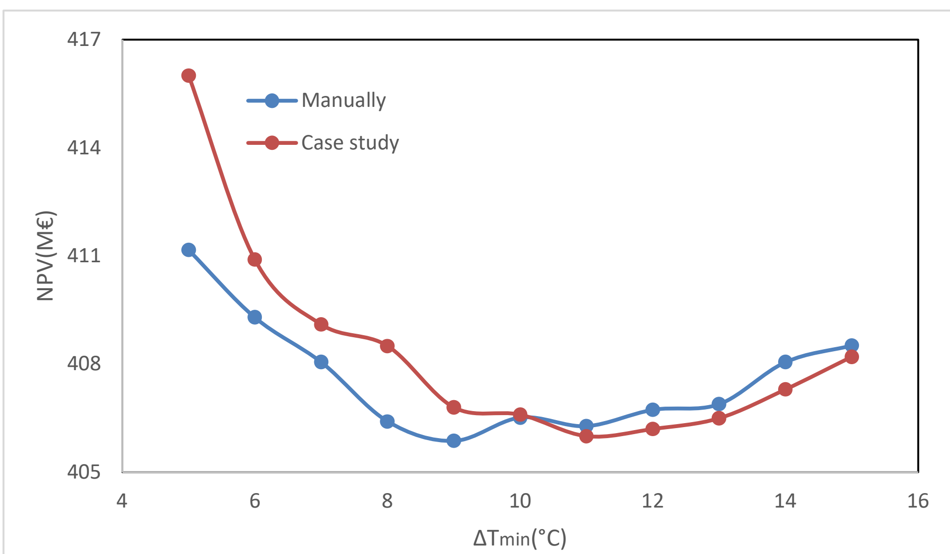

Figure 5-3:NPV as a function of ΔT_{min} with removal efficiency 85%, E_M=0.25, 20 years calculation period and 8.5% interest rate for case study(automatic) and manually, by using Aspen HYSYS model for ΔT_{min}=10°C

6 Discussion

In this chapter, different results will be compared and discussed with previous studies. In the last subchapter, the achievements from the present work will be explained, and avenues for future works will be recommended.

6.1 Comparison of the results with previous works

In this study, two different base cases have been used for simulation. Differences between these case studies have been explained in section 4.2. In The total equipment price for 2020 has been compared with the Aromada[8]. The total price of the equipment in the CO2 capture plant without direct contact cooler (DCC) has been reported 122 million euros (M€). CAPEX in this study is 120.2 M€ for 10 stages and 117 M€ for 16 stages. Absorber, fan, and lean-rich heat exchanger are the most expensive parts of the process, and they have been considered around 80% to 85% of the total equipment cost of the process.

Table 6-1, simulation results are compared with the current literature, and the results show differences between this simulation with the other simulations. Reboiler-specific heat is used for validation of the simulation. CO_2 mole fraction, absorber height, and E_M are the factors that can explain the differences. Despite these differences in the simulation, energy regeneration for simulations with ΔT_{min}=10°C were in the range 3.2 and 5 MJ/kgCO_2 as reported in several articles [16].

The total equipment price for 2020 has been compared with the Aromada[8]. The total price of the equipment in the CO_2 capture plant without direct contact cooler (DCC) has been reported 122 million euros (M€). CAPEX in this study is 120.2 M€ for 10 stages and 117 M€ for 16 stages. Absorber, fan, and lean-rich heat exchanger are the most expensive parts of the process, and they have been considered around 80% to 85% of the total equipment cost of the process.

Table 6-1: Comparison simulation results[8]

	CO_2 (mol%)	Capture rate(%)	No. absorber stages	ΔT_{min}	Rich loading	Reboiler specific heat (MJ/kgCO_2)	E_M
[8]EDF	3.75	85.06	15	10	0.5	3.71	0.11-0.21
Present work	3.3	85.08	16	10	0.5	3.64	0.15
Amrollahi[17]	3.8	90	13	8.5	0.47	3.74	n.a
N.Sipocz[18]	4.2	90	26.9(Height)	10	0.47	3.93	n.a
Øi2007[6]	3.75	85	10	10	n.a	3.65	0.25
Present work	3.3	85.12	10	10	0.5	3.79	0.25

This is a trade-off between the heat exchanger area and the process's external utility requirements (steam, power, and cooling water).[Robin S. chemical] Figure 4-1shows the minimum value of NPV, which is at ΔT_{min}=9°C. CAPEX for the process has been steadily reduced, and at the same time, whereas OPEX has been gradually increased (Appendix I-1) for ΔT_{min} from 5°C to 18°C. In this simulation, the main part that has effects on the CAPEX is the lean-rich heat exchanger. The heat transfer between lean amine and rich amine has been decreased due to the smaller lean-rich heat exchanger area, and on the other side, the required heat in the reboiler will be increased. The main reason for the rising trend in OPEX is the increase ΔT_{min}. When ΔT_{min} is 9°C, NPV has the lowest value, as seen in Figure 4-1. The range has been determined from 10°C to 14°C for ΔT_{min}[4]. As well, based on the results of another study, the value of ΔT_{min} has a nearly same range of 10°C to 15°C [9]. It should be noted that there are also some differences between the results of these two studies and the present work. The number of stages, E_M, and the flue gas content are the main differences in the present work case in comparison with the Kallevik [4]. As well, the removal efficiency and CO_2 content in the flue gas are the main differences between the present work and Øi [9].

The next analysis in Figure 4-1 is changing the energy in the reboiler per kg CO_2, which has been captured in the process. The reduction of the lean-rich heat exchanger area is the reason for the increasing trend in energy consumption in the reboiler, as indicated in the diagram.

CAPEX and OPEX can be affected by the packing height in the absorber. Each stage of the absorber has a given pressure drop per meter, and the pressure of the flue gas to the absorber should be updated for each simulation. Total equipment price has been changed due to the changing cost from the absorber to the lean-rich heat exchanger. The primary cause for increasing OPEX in the calculations is increasing the pressure drop in the absorber column. Calculation results and diagram are presented in appendix I-2 and Figure 4-2, respectively Figure 4-2. The optimal number of stages has been reported 15 for both this study and Kallevik [4].

Figure 4-3 depicts the NPV value in the absorber for various gas velocities. In this analysis, it has been assumed that the packing volume is equal for different velocities. Based on this assumption, the cross-sectional area plays an important role in the calculations. The NPV of the superficial velocity with the lowest NPV is between 2 and 2.2. The cross-sectional area of the packing decreases as the flue gas velocity increases. By reducing the packing diameter, the total price of the absorber is reduced, while at the same time, the fan price is increased. Lower velocity leads to a decrease in the pressure drop in the packing. Therefore, the fan should be able to overcome these pressure drops. Energy consumption (power for the fan) has been raised for a higher velocity of flue gas in the absorber, and the main reason was the high-pressure drop in the packing. A trade-off between energy consumption and the price of equipment for different gas velocities is around 2.2 m/s. E_M is considered to be constant for all the velocities in the simulation. Pressure drop per meter packing has been extracted from Appendix J for packing type 250Y and different velocities.

The only difference between this following analysis and the previous one is the impacts of the calculated E_M on the results. The most significant change is an increase in the price of absorber packing; however, the optimum superficial velocity for flue gas in the absorber is about 2.2 m/s. Results are presented in Figure 4-4 and Appendix I-4.

According to the previous studies, gas velocity in the absorber is between 2 and 2.5 m/s, which is consistent with the results of this study [7].

6.2 Comparison of the automation methods

Automation of the cost estimation for CO_2 capturing is helpful to reduce the required time for simulation or check new changes in the process. Possibilities for automated cost estimation and optimization have been investigated by Øi [9]. They have added adjust operation to their Aspen HYSYS model to achieve the required ΔT_{min} and capture efficiency. The potentials for automated cost estimation and optimization were studied by Øi [9]. To attain the required ΔT_{min} and capture efficiency, they have incorporated adjust operation to their Aspen HYSYS model. These operations are beneficial to reduce the required time for performing the defined requirement. The model should be iterated until all of the constraints are fulfilled, and the model results can be calculated using the Aspen HYSYS spreadsheet. All simulations in this study used the adjust operation for ΔT_{min}, and the Aspen simulation workbook used the adjust operation for removal efficiency. Case study and Aspen simulation workbook are two ways for automating the simulation in the Aspen HYSYS. Aspen simulation workbook offers more flexibility in terms of automating simulation. The case study in Aspen HYSYS can only be set up for some simulations, such as different ΔT_{min}, but changing the absorber height is not straightforward. The pressure to the absorber should be adjusted for the new absorber height, which is challenging in the case study, while it is quite simple in the aspen simulation workbook. VBA is another way to connect Aspen Hsysy and Excel. It is possible to import from Excel to Aspen HYSYS spreadsheet by total installation factors in the EDF method. A VBA code has been written and added to Appendix K. The Aspen HYSYS library can be quickly imported into Excel and utilized in simulations upon request.

6.3 Future work

In the current study, the calculation of the dimension and price of the different equipment has been done by using the spreadsheet in the Aspen HYSYS. The required Aspen HYSYS spreadsheets are defined and formulized for dimensioning and calculating of CAPEX, OPEX (NPV), and removal efficiency. For each change in equipment size, all the items have been updated automatically. The base case prices have been calculated from Aspen In-Plant Cost Estimator V10, and these values were used to determine the other prices using the power law equation. Gas inlet temperature to the absorber is assumed to be constant at 40°C. The optimal ΔT_{min} for lean/rich heat exchangers and the optimal number of stages in the absorber have been determined. Furthermore, the effects of various superficial velocities with a constant packing volume have been described. In one of the cases, E_M is considered as a constant, whereas in the other case, E_M is calculated by the Excel spreadsheet, which is presented in Appendix L. The data which are applied for the calculation of the pressure drop per meter packing is extracted from the diagram in Appendix J.

The following approaches have been used to automate the process:

- Case study
- Aspen simulation workbook
- VBAfor application in the Excel

The first approach may be configured in Aspen HYSYS, with the results exported to Excel. The second is an Excel-Aspen HYSYS interface, which allows Aspen HYSYS to be run through Excel and the results to be shown in Excel. In order to execute the Aspen HYSYS simulation in the Aspen simulation worksheet, different input data can be added to Excel. VBA is an Excel programming language that can be used to make the cost estimation fully automated. A VBA code was created to link the Aspen HYSYS spreadsheet and the Excel file of the installation factors. This code has enabled the link between the CAPEX spreadsheet and the Excel table of installation factors

The input data (number of stages, pressure drop per stage, different ΔT_{min}) is read and imported into the Aspen HYSYS model from Excel using VBA code. Therefore, the automation of the whole process can be an interesting future study for researchers. Implementing all of the simulations in the Aspen simulation workbook with different stages and improving the robustness of the Aspen HYSYS simulations are the other suggestions that can be valuable for future studies.

7 Conclusion

MEA, followed by desorption, is one of the conventional techniques for removing CO2 from industry flue gas [9]. Absorption, desorption, and circulation system are the main considerations in this study. Aspen HYSYS V10 is used for CO_2 capturing simulation. By defining four spreadsheets for dimensioning, CAPEX, OPEX, capture rate and all of the required items for simulation and optimization are available. Aspen In-Plant Cost Estimator V10 with the database from 2016 was used to estimate the cost of the base case equipment.The other required prices have been calculated by utilizing of power law equation.

Due to the analysis for determining the trade-off between the heat exchanger area and energy consumption, minimum approach temperature has been investigated in the analyses. The optimal ΔT_{min} has been obtained equal to 9°C. Absorber height optimization with the optimal number of stages equal to 15 is another item that has been investigated in this study. Superficial velocity for constant packing volume has also been examined. Pressure drop and E_M were collected from two different sources, and the optimal velocity was found to be between 2 and 2.2 m/s in both cases.

Case study, Aspen simulation workbook, and VBA have been examined for the automation of cost estimation and optimization. The case study can be defined in the Aspen HYSYS simulation model, and it can be automatically calculated by adding adjust operation to the ΔT_{min} and removal efficiency. In the Aspen simulation workbook, all the input data, which consists of pressure drop in the different number of stages, E_M, and other inputs data, are defined in Excel. After that, all the data have been read from Excel, and finally, the simulation is performed using these updated data. VBA is one of the best solutions to provide a connection between Aspen HYSYS and Excel. In the present study, a VBA code has been written for importing installation factors from the table to the Aspen HYSYS spreadsheet.

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Appendices

Appendix A – project description

University of South-Eastern Norway

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

FMH606 Master's Thesis

Title: Process simulation and automated cost optimization of CO_2 capture using Aspen HYSYS

USN supervisor: Lars Erik Øi and Solomon Aromada (co-supervisor and PhD student USN)

External partner: Nils Eldrup (SINTEF Tel-Tek)

Task background:
Master projects from 2007 at the University of South-Eastern Norway and Telemark University College have included cost estimation in a spreadsheet connected to an Aspen HYSYS simulation. USN (HSN and TUC) has collaborated with different companies (SINTEF Tel-Tek, Statoil/Equinor, Aker Solutions, Norcem, Yara, Skagerak and Gassnova) working on CO_2 capture.

Task description:
The general aim is to develop further models in Aspen HYSYS especially for cost optimization of CO_2 capture by amine absorption. A special aim is to utilize the spreadsheet facility in Aspen HYSYS to optimize the process.

1. Literature search on cost estimation and optimization of amine based CO_2 capture. Of special interest is optimization based on dimensioning, cost estimation and optimization from process simulation.

2. Aspen HYSYS simulation, dimensioning and cost estimation of different alternatives utilizing the spreadsheet facility in Aspen HYSYS.

3. Process optimization of process parameters and possibly automated optimization. Typical parameters are gas inlet temperature, temperature approach in the main heat exchanger and packing height in the absorption column. Possible challenges are optimization of the gas velocity and the pressure drop through the absorber.

4. Evaluation of limitations for cost optimization in cost estimation and cost optimization of amine based CO_2 absorption.

Student category: EET or PT

The task is suitable for online students (not present at the campus): Yes (but it must be possible to run the Aspen HYSYS program)

Practical arrangements:
The work will be carried out mainly at USN or from home.

Supervision:
As a general rule, the student is entitled to 15-20 hours of supervision. This includes necessary time for the supervisor to prepare for supervision meetings (reading material to be discussed, etc).

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

Signatures:

Supervisor (date and signature): 6/9-21

Student (write clearly in all capitalized letters): POUYA RAHMANI

Student (date and signature): 20.08.2021

Appendix B – Price index and cumulative discount factors[19]

Year	Price index
2020	111.3
2019	109.3
2018	106
2017	104.3
2016	101.5
2011	92.7

	Discount rate (% per year)											
No. of years	2.5	5.0	7.5	10.0	12.5	15.0	17.5	20.0	22.5	25.0	27.5	30.0
0	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	1.98	1.95	1.93	1.91	1.89	1.87	1.85	1.83	1.82	1.80	1.78	1.77
2	2.93	2.86	2.80	2.74	2.68	2.63	2.58	2.53	2.48	2.44	2.40	2.36
3	3.86	3.72	3.60	3.49	3.38	3.28	3.19	3.11	3.03	2.95	2.88	2.82
4	4.76	4.55	4.35	4.17	4.01	3.85	3.72	3.59	3.47	3.36	3.26	3.17
5	5.65	5.33	5.05	4.79	4.56	4.35	4.16	3.99	3.83	3.69	3.56	3.44
6	6.51	6.08	5.69	5.36	5.05	4.78	4.54	4.33	4.13	3.95	3.79	3.64
7	7.35	6.79	6.30	5.87	5.49	5.16	4.87	4.60	4.37	4.16	3.97	3.80
8	8.17	7.46	6.86	6.33	5.88	5.49	5.14	4.84	4.57	4.33	4.12	3.92
9	8.97	8.11	7.38	6.76	6.23	5.77	5.38	5.03	4.73	4.46	4.23	4.02
10	9.75	8.72	7.86	7.14	6.54	6.02	5.58	5.19	4.86	4.57	4.32	4.09
11	10.51	9.31	8.32	7.50	6.81	6.23	5.74	5.33	4.97	4.66	4.39	4.15
12	11.26	9.86	8.74	7.81	7.05	6.42	5.89	5.44	5.06	4.73	4.44	4.19
13	11.98	10.39	9.13	8.10	7.27	6.58	6.01	5.53	5.13	4.78	4.48	4.22
14	12.69	10.90	9.49	8.37	7.46	6.72	6.12	5.61	5.19	4.82	4.52	4.25
15	13.38	11.38	9.83	8.61	7.63	6.85	6.21	5.68	5.23	4.86	4.54	4.27
16	14.06	11.84	10.14	8.82	7.78	6.95	6.28	5.73	5.27	4.89	4.56	4.28
17	14.71	12.27	10.43	9.02	7.92	7.05	6.35	5.77	5.30	4.91	4.58	4.29
18	15.35	12.69	10.71	9.20	8.04	7.13	6.40	5.81	5.33	4.93	4.59	4.30
19	15.98	13.09	10.96	9.36	8.15	7.20	6.45	5.84	5.35	4.94	4.60	4.31
20	16.59	13.46	11.19	9.51	8.24	7.26	6.49	5.87	5.37	4.95	4.61	4.32
21	17.18	13.82	11.41	9.65	8.33	7.31	6.52	5.89	5.38	4.96	4.61	4.32
22	17.77	14.16	11.62	9.77	8.40	7.36	6.55	5.91	5.39	4.97	4.62	4.32
23	18.33	14.49	11.81	9.88	8.47	7.40	6.57	5.92	5.40	4.98	4.62	4.33
24	18.88	14.80	11.98	9.98	8.53	7.43	6.60	5.94	5.41	4.98	4.63	4.33

Appendix C – Installation cost factor

Equipment cost (CS) in kEUR from:	0	10	20	40	80	160	320	640	1280	2560	5120
to:	10	20	40	80	160	320	640	1280	2560	5120	10240
Equipment costs	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00	1,00
Erection cost	0,49	0,33	0,26	0,20	0,16	0,12	0,09	0,07	0,06	0,04	0,03
Piping incl. Erection	2,24	1,54	1,22	0,96	0,76	0,60	0,48	0,38	0,30	0,23	0,19
Electro (equip & erection)	0,76	0,59	0,51	0,44	0,38	0,32	0,28	0,24	0,20	0,18	0,15
Instrument (equip. & erection)	1,50	1,03	0,81	0,64	0,51	0,40	0,32	0,25	0,20	0,16	0,12
Ground work	0,27	0,21	0,18	0,15	0,13	0,11	0,09	0,08	0,07	0,06	0,05
Steel & concrete	0,85	0,66	0,55	0,47	0,40	0,34	0,29	0,24	0,20	0,17	0,15
Insulation	0,28	0,18	0,14	0,11	0,08	0,06	0,05	0,04	0,03	0,02	0,02
Direct costs	7,38	5,54	4,67	3,97	3,41	2,96	2,59	2,30	2,06	1,86	1,71
	-	-	-	-	-	-	-	-	-	-	-
Engineering process	0,44	0,27	0,22	0,18	0,15	0,12	0,10	0,09	0,07	0,06	0,05
Engineering mechanical	0,32	0,16	0,11	0,08	0,06	0,05	0,03	0,03	0,02	0,02	0,01
Engineering piping	0,67	0,46	0,37	0,29	0,23	0,18	0,14	0,11	0,09	0,07	0,06
Engineering el.	0,33	0,20	0,15	0,12	0,10	0,08	0,07	0,06	0,05	0,04	0,04
Engineering instr.	0,59	0,36	0,27	0,20	0,16	0,12	0,10	0,08	0,06	0,05	0,04
Engineering ground	0,10	0,05	0,04	0,03	0,02	0,02	0,01	0,01	0,01	0,01	0,01
Engineering steel & concrete	0,19	0,12	0,09	0,08	0,06	0,05	0,04	0,04	0,03	0,03	0,02
Engineering insulation	0,07	0,04	0,03	0,02	0,01	0,01	0,01	0,01	0,00	0,00	0,00
Engineering	2,70	1,66	1,27	0,99	0,79	0,64	0,51	0,42	0,34	0,28	0,23
	-	-	-	-	-	-	-	-	-	-	-
Procurement	1,15	0,38	0,48	0,48	0,24	0,12	0,06	0,03	0,01	0,01	0,00
Project control	0,14	0,08	0,06	0,05	0,04	0,03	0,03	0,02	0,02	0,01	0,01
Site management	0,37	0,28	0,23	0,20	0,17	0,15	0,13	0,11	0,10	0,09	0,09
Project management	0,45	0,30	0,26	0,22	0,18	0,15	0,13	0,11	0,10	0,09	0,08
Administration	2,10	1,04	1,03	0,94	0,63	0,45	0,34	0,27	0,23	0,20	0,18
	-	-	-	-	-	-	-	-	-	-	-
Commissioning	0,31	0,19	0,14	0,11	0,08	0,06	0,05	0,04	0,03	0,02	0,02
	-	-	-	-	-	-	-	-	-	-	-
Identified costs	12,48	8,43	7,11	6,02	4,91	4,10	3,49	3,02	2,66	2,37	2,13
	-	-	-	-	-	-	-	-	-	-	-
Contingency	2,50	1,69	1,42	1,20	0,98	0,82	0,70	0,60	0,53	0,47	0,43
	-	-	-	-	-	-	-	-	-	-	-
Installation factor 2020	14,98	10,12	8,54	7,22	5,89	4,92	4,19	3,63	3,19	2,84	2,56

Fluid handling equipment Installation factors

Adjustment for materials:

SS316 Welded: Equipment and piping factors multiplies with 1,75

SS316 rotating: Equipment and piping factors multiplies with 1,30

Exotic Welded: Equipment and piping factors multiplies with 2,50

Exotic Rotating: Equipment and piping factors multiplies with 1,75

Porsgrunn September 2020
Nils Henrik Eldrup

Appendix D–1: Dimensioning of absorber and desorber

ABSORBER			
Flue gas volumetric flow[m3/h]	2904558	ITEM DESCRIPTION	Absorber
Flue gas volumetric flow[m3/s]	807	APPLICATION	ABSORB
Superficial gas velocity[m/s]	2.5	SHELL MAT.	SS316
Area[m2]	323	VESSEL DIAM.(m)	20.43
Inner-diameter of Absorber[m]	20.27	TANGENT-TANGENT	30
Wall thickness [m]	0.02	PACKING HEIGHT	10
Outer-diameter of Absorber[m]	20.31	PACKING TYPE	M76YB
Packing height per stage [m/stage]	1	No. PACKED SECTIONS	2
Number of stages	10	Area[m2]	327
Total packing height [m]	10	Volume_Old[m3]	3274
Total packing volume [m3]	3227	Volume_New[m3]	3227
Tangent to tangent height [m]	40	NEW COST	
EQUIPMENT AND SETTING		EQUIPMENT AND SETTING	
Material(kEUR)	10505	Material(kEUR)	10355
Packing Cost(kEUR)	9915	Packing Cost(kEUR)	9774
Total Cost	20420		

DESORBER			
Vapour volumetric flow[m3/h]	169120	ITEM DESCRIPTION	Desorber
Vapour volumetric flow[m3/s]	47	APPLICATION	ABSORB
Superficial gas velocity[m/s]	1	SHELL MAT.	SS316
Area[m2]	47	VESSEL DIAM.(m)	7.55
Inner-diameter of Absorber[m]	7.73	TANGENT-TANGENT	15
Wall thickness [m]	0.03	PACKING HEIGHT	6
Outer-diameter of Absorber[m]	7.79	PACKING TYPE	M76YB
Packing height per stage [m/stage]	1	No. PACKED SECTIONS	2
Number of stages	6	Area[m2]	45
Total packing height [m]	6	Volume_Old[m3]	269
Total packing volume [m3]	282	Volume_New[m3]	282
Tangent to tangent height [m]	15	NEW COST	
EQUIPMENT AND SETTING		EQUIPMENT AND SETTING	
Material(kEUR)	1402	Material(kEUR)	1445
Packing Cost(kEUR)	812	Packing Cost(kEUR)	850
Total Cost	2214		

Appendix D–2: Dimensioning of reboiler and condenser

REBOILER			
Reboiler Duty, Q[kW]	142093	ITEM DESCRIPTION	Reboiler(Thermosiphon)
T(out,cold), Lean Amine from reboiler	120	Heat transfer Area(m2)	768
T(in, cold), Lean Amine in to the reboiler	117	No. shells	1
T(in,Hot) Saturated steam In	160	Tube material	316LW
T(Out,Hot) Saturated steam out	152	Shell material	SS316
LMTD	38	Actual area per unit_Old[m2]	770
Overall Heat Transfer coefficient, U [kW/m2.k]	0.8	Actual area per unit_New[m2]	946
Total Heat Transfer Area[m2]	4730		
Max. area per of unit[m2]	1000		
Calculated Number of unit[m2]	5		
Actual number of unit	5		
Actual Area per unit[m2]	946		
EQUIPMENT AND SETTING		NEW COST	
Material(kEUR) PER UNIT	260	Material(kEUR) PER UNIT	298
Material(kEUR) TOTAL COST	1042	Material(kEUR) TOTAL COST	1488

DESORBER CONDENSER			
Heat Transfer rate, Q [kW]	18343		
T(in, Hot) Vapour in to Condenser	102		
T(out, Hot) Vapour out of Condenser	94	ITEM DESCRIPTION	Desorber-Condenser
T(in, cold) cooling water	15	Heat transfer Area(m2)	232
T(out, cold) cooling water	25	No. shells	1
LMTD	78	Tube material	316LW
Overall Heat Transfer coefficient, U [kW/m2.k]	1	Shell material	SS316
Total Heat Transfer Area[m2]	235	Total area_Old[m2]	233
Max. area per of unit[m2]	1000	Total area_New[m2]	235
Calculated Number of unit[m2]	0.24		
Actual number of unit	1		
Cp	4.19		
Mass Flow of Cooling Water required [kg/s]	438		
Density of water[kg/s]	1000		
Volumetric Flow[m3/s]	0.4		
Volumetric flow[m3/h[	1577		
EQUIPMENT AND SETTING		NEW COST	
Material(kEUR)	110	Material(kEUR) TOTAL COST	111

Appendix D–3: Dimensioning of heat exchangers

Lean/Rich HEAT EXCHANGER			
Cold dutyp[kJ/h]	650490187	ITEM DESCRIPTION	
Cold duty[kW]	180692	Heat transfer Area(m2)	983
LMTD(Hysys)	13	No. shells	1
Overll U [kW/m2.K]	0.5	Tube material	316LW
Area of heat exchanger[m2]	27564	Shell material	SS316
max shell&tube area	1000	Actual area per unit_Old[m2]	973.4
Total number of exchanger	28	Actual area per unit_New[m2]	984
Real number of heat exchanger[m2]	28		
Actual area per unit[m2]	984		
EQUIPMENT AND SETTING		NEW COST	
Material(kEUR) PER UNIT	324.8	Material(kEUR) PER UNIT	327
Material(kEUR) TOTAL COST	6171	Material(kEUR) TOTAL COST	9161

LEAN HEAT EXCHANGER			
Heat Transfer rate,Q [kJ/h]	110390510	Heat transfer Area(m2)	899
Heat Transfer rate,Q [kW]	30664	No. shells	1
T(in, Hot) Lean Amine to cooler	52	Tube material	316LW
T(Out, Hot) Lean Amine to Mixer	40	Shell material	SS316
T(in, Cold) Cooling Water 3	15	Actual area per unit_Old[m2]	825
T(Out, Cold) Cooling Water 4	25	Actual area per unit_New[m2]	730
LMTD	26		
LMTD(Hysys)	25		
Overall Heat Transfer Coefficient, U[kW/m2.K]	0.8		
Total Heat Transfer Area [m2]	1465		
Max. Area per unit[m2]	1000		
Calculated number of units	1		
Actual number of unit	2		
Area per unit	732		
Cp	4		
Mass flow of cooling water[kg/s]	732		
Density of water[kg/m3]	1000		
Volumetric Flow [m3/s]	732		
EQUIPMENT AND SETTING		NEW COST	
Material(kEUR) PER UNIT	302	Material(kEUR) PER UNIT	279
Material(kEUR) TOTAL COST	604	Material(kEUR) TOTAL COST	559

Appendix D–4: Dimensioning of pumps, fan and seperator

RICH PUMP			
Flow[m3/h]	2530	Casing material	SS316
Flow[l/s]	703	Duty_Old[kW]	76
Duty[kW]	75	Duty_New[kW]	75
Equipment(kEUR)	269	New Cost(kEUR)	264

LEAN PUMP			
Flow[m3/h]	2630	Casing material	SS316
Flow[l/s]	731	Duty_Old[kW]	99
Duty[kW]	97	Duty_New[kW]	97
Equipment(kEUR)	280	New Cost(kEUR)	276

FLUE GAS FAN			
Actual Flue Gas Flow [m3/h]	3086115	Casing material	CS
Max. Flow Rate [m3/h]	1529000	Duty Per unit_Old[kW]	9084
Calculated No. of Units	2	Duty Per unit_New[kW]	22732
Required Units	3		
Duty [kJ/h]	81834285		
Duty [kW]	22732		
Actual Flue Gas per unit[m3/h]	1543057	New cost	
Equipment per unit(kEUR)	1226	Equipment per unit(kEUR)	2045
Total Equipment price(kEUR)	2452	Total Equipment price(kEUR)	6136

SEPERATOR			
Flue Gas Flow Rate	3441618	Vessel Cross-Sectional Area (Vo/Va)	215.5
Flue Gas Flow Rate (Vo) [m3/s]	956	Packing height [m]	4
Liquid Phase Mass Density [kg/m3]	999	Volume_Old [m3]	862
Gas Phase Mass density [kg/m3]	1	Volume_New [m3]	916
K Factor, Sounder-Brown Velocity	0.15		
Allowable Vapour Velocity, Va	4		
Vessel Cross-Sectional Area (Vo/Va)	229		
Vessel Inner-Diameter (Di) [m]	17		
Stress [Pa]	220000000		
CA (Corrosion) [m]	1.00E-03		
E (Joint Eff.) [-]	0.85		
Design Pressure [bar]	1		
Design Pressure [Pa]	100000		
Wall Thickness [m]	5.56E-03		
Vessel Outer Diameter (Do =Di + 2*Wall Thickne	17		
Packing height [m]	4		
Total Vessel Height [m]	15		
Material Cost/unit(kEUR)	1490	New Cost(kEUR)	1583

Appendix E – 1: CAPEX calculation for base case with 10 stages and E_M=0.25

EQUIPMENT	2016 MATERIAL TYPE	EQ COST in SS (kEUR)	MATERIAL FACTOR	CARBON STEEL EQ COST in CS (kEUR)	CARBON STEEL INSTALLATION FACTOR in CS	EQUIPMENT FACTOR	PIPING FACTOR	STAINLESS STEEL INSTALLATION FACTOR IN SS	TOTAL INSTALLED COST kEUR (CAPEX 2016)	INSTALLED COST kEuros CAPEX 2020/UNIT	UNIT (i.e.) NUMBER OF EQUIPMENT	INSTALLED COST kEuros CAPEX 2020	Total [k€]	EQUIPMENT
Absorber Shell	SS	10505	1.75	6003	2.56	1	0.19	3.45	20725	21644	1	21644	42072	Absorber Shell
Absorber Packing	SS	9915	1.75	5666	2.56	1	0.19	3.45	19561	20428	1	20428		Absorber Packing
Desorber Shell	SS	1402	1.75	801	3.63	1	0.38	4.67	3738	3904	1	3904	6471	Desorber Shell
Desorber Packing	SS	811.8	1.75	464	4.19	1	0.48	5.3	2459	2568	1	2568		Desorber Packing
Desorber Reboiler	SS	298	1.75	170	4.92	1	0.6	6.12	1041	1087	5	5434	5434	Desorber Reboiler
Desorber Condenser	SS	110	1.75	63	7.22	1	0.96	8.69	546	570	1	570	570	Desorber Condenser
LEAN/RICH HEX	SS	327	1.75	187	4.92	1	0.6	6.12	1144	1195	28	33459	33459	LEAN/RICH HEX
LEAN MEA COOLER	SS	279	1.75	160	5.89	1	0.76	7.21	1151	1202	2	2405	2405	LEAN MEA COOLER
RICH PUMP	SS	268.5	1.3	207	4.92	1	0.6	5.4	1115	1165	1	1165	1165	RICH PUMP
LEAN PUMP	SS	280.4	1.3	216	4.92	1	0.6	5.4	1165	1216	1	1216	1216	LEAN PUMP
FLUE GAS FAN	CS	2045	1	2045	3.63	1	0.38	3.63	7424	7754	3	23261	23261	FLUE GAS FAN
SEPERATOR	SS	1490	1.75	851	3.63	1	0.38	4.67	3972	4148	1	4148	4148	SEPERATOR

Appendix E – 2: CAPEX calculation for base case with 16 stages and E_M=0.15

EQUIPMENT	2016 MATERIAL TYPE	EQ COST in SS (kEUR)	MATERIAL FACTOR	CARBON STEEL EQ COST in CS (kEUR)	CARBON STEEL INSTALLATION FACTOR in CS	EQUIPMENT FACTOR	PIPING FACTOR	STAINLESS STEEL INSTALLATION FACTOR IN SS	TOTAL INSTALLED COST kEUR (CAPEX 2016)	INSTALLED COST kEuros CAPEX 2020/UNIT	UNIT (i.e.) NUMBER OF EQUIPMENT	INSTALLED COST kEuros CAPEX 2020	Total [k€]	EQUIPMENT
Absorber Shell	SS	10570	1.75	6040	2.56	1	0.19	3.45	20853	22805	1	22805	63316	Absorber Shell
Absorber Packing	SS	16032	1.75	9161	2.56	1	0.19	3.45	31629	34590	1	34590		Absorber Packing
Desorber Shell	SS	999	1.75	571	4.19	1	0.48	5.3	3026	3309	1	3309	5697	Desorber Shell
Desorber Packing	SS	721	1.75	412	4.19	1	0.48	5.3	2184	2388	1	2388		Desorber Packing
Desorber Reboiler	SS	316	1.75	181	4.92	1	0.6	6.12	1106	1209	4	4836	4836	Desorber Reboiler
Desorber Condenser	SS	102	1.75	58	7.22	1	0.96	8.69	506	553	1	553	553	Desorber Condenser
LEAN/RICH HEX	SS	327	1.75	187	4.92	1	0.6	6.12	1142	1249	25	31227	31227	LEAN/RICH HEX
LEAN MEA COOLER	SS	245	1.75	140	5.89	1	0.76	7.21	1010	1104	2	2209	2209	LEAN MEA COOLER
RICH PUMP	SS	262	1.3	201	4.92	1	0.6	5.4	1086	1188	1	1188	1188	RICH PUMP
LEAN PUMP	SS	269	1.3	207	4.92	1	0.6	5.4	1119	1223	1	1223	1223	LEAN PUMP
FLUE GAS FAN	CS	544	1	544	4.19	1	0.48	4.19	2281	2494	2	4988	4988	FLUE GAS FAN
SEPERATOR	SS	485	1.75	277	4.92	1	0.6	6.12	1697	1856	1	1856	1856	SEPERATOR

Appendix F – 1: OPEX calculation for base case with 10 stages and EM=0.25

	CAPEX	ELECTRICITY	STEAM	COOLING WATER	MTCE COST	TOTAL OPEX/YEAR	TOTAL
UNIT		[€/kWh]	[€/kWh]	[€/(m3)]	%		
COST/UNIT		6.00E-02	1.50E-02	2.00E-02	4.00E-02		
OPERATIONAL HOURS/YEAR		8000	8000	8000			
	[k€]	[€]	[€]	[€]	[€]	[€]	[k€]
ABSOBER	42072	0	0	0	1683	1683	1.7
DESORBER	6471	0	0	0	259	259	0.3
Desorber Reboiler	5434	0	17051214	0	217	17051431	17051
Desorber Condenser	570	0	0	0	23	23	0.0
LEAN/RICH HEX	33459	0	0	0	1338	1338	1
LEAN MEA COOLER	2405	0	0	466207	96	466303	466
RICH PUMP	1165	35987	0	0	47	36033	36
LEAN PUMP	1216	46752	0	0	49	46801	47
FLUE GAS FAN	23261	10911238	0	0	930	10912168	10912
SEPERATOR	4148	0	0	0	166	166	0.2
TOTAL	120201	10993977	17051214	466207	4808		28516

Appendix F – 2: OPEX calculation for base case with 16 stages and EM=0.15

	CAPEX	ELECTRICITY	STEAM	COOLING WATER	MTCE COST	TOTAL OPEX/YEAR	TOTAL
UNIT		[€/kWh]	[€/kWh]	[€/(m3)]	%		
COST/UNIT		6.00E-02	1.50E-02	2.00E-02	4.00E-02		
OPWERATIONAL HOURS/YEAR		8.00E+03	8.00E+03	8.00E+03			
	[k€]	[€]	[€]	[€]	[€]	[€]	[k€]
ABSOBER	63316	0	0	0	2533	2533	2.5
DESORBER	5697	0	0	0	228	228	0.2
Desorber Reboiler	4836	0	16391841	0	193	16392035	16392.0
Desorber Condenser	553	0	0	236493	22	236515	236.5
LEAN/RICH HEX	31227	0	0	0	1249	1249	1.2
LEAN MEA COOLER	2209	0	0	424120	88	424209	424
RICH PUMP	1188	40950	0	0	48	40998	41
LEAN PUMP	1223	44430	0	0	49	44479	44.5
FLUE GAS FAN	4988	1935898	0	0	200	1936098	1936
SEPERATOR	1856	0	0	0	74	74	0.1
TOTAL	117093	2021278	16391841	660613	4684		19078

Appendix G–1: Aspen In-Plant Cost Estimator results for absorber

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Absorber

Item Code: DTW PACKED
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	PACKED	
Number of identical items	1	
EQUIPMENT DESIGN DATA		
Application	ABSORB	
Liquid volume	13111.49	M3
Design gauge pressure	100.002	KPAG
Design temperature	340.000	DEG C
Operating temperature	305.000	DEG C
COLUMN DATA		
Shell material	SS316	
Diameter option	OD	

Vessel diameter	20.430	M
Vessel tangent to tangent height	40.000	M
Head type	HEMI	
MECHANICAL DESIGN DATA		
Wind or seismic design	W+S	
Weld efficiency	85.000	PERCENT
Thickness Average	17.824	MM
Corrosion allowance	0.000000	MM
THICKNESSES REQUIRED		
Thickness for internal pressure	29.361	MM
Wind or seismic design thickness	31.007	MM
PACKING DATA		
Number of distributor plates	0	
Number of packed sections	2	
Section height	5.0000	M
Cross sectional area	327.813	M2
SECTION 1		
Packing type	M76YB	
Total packing height	10.000	M
Packing volume	3278.151	M3
Packing volume per unit height	327.815	M3/M
VESSEL SKIRT DATA		
Skirt material	CS	
Skirt height	9.5000	M
Skirt thickness	55.000	MM
NOZZLE AND MANHOLE DATA		
Nozzle ASA rating	150	CLASS
Nozzle material	SS316	
Nozzle A Quantity	4	
Nozzle A Diameter	1200.000	MM DIAM
Nozzle A Location	S	
Nozzle B Quantity	10	
Nozzle B Diameter	50.000	MM DIAM
Nozzle B Location	S	
Number of manholes	8	
Manhole diameter	900.000	MM
PROCESS DESIGN DATA		
WEIGHT DATA		

Shell	366500	KG
Heads	236500	KG
Nozzles	41	KG
Manholes and Large nozzles	7200	KG
Skirt	265100	KG
Base ring and lugs	36800	KG
Ladder clips	200	KG
Platform clips	540	KG
Fittings and miscellaneous	70	KG
Total weight less packing	913000	KG
VENDOR COST DATA		
Packing cost	9915112	EURO
Material cost	5245216	EURO
Field fabrication cost	1228875	EURO
Fabrication labor	36189	HOURS
Shop labor cost	657467	EURO
Shop overhead cost	676113	EURO
Office overhead cost	1327305	EURO
Profit	1370315	EURO
Total cost	20420400	EURO
Cost per unit weight	22.366	EUR/KG
Cost per unit height or length	2042040.	EUR/M
Cost per unit volume	6229.244	EUR/M3
Cost per unit area	62292.80	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	20420400.	334508.	5603
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**20420400**	**334508**	**5603**

Appendix G–2: Aspen In-Plant cost estimation results for desorber

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Desorber

Item Code: DTW PACKED
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	PACKED	
Number of identical items	1	
EQUIPMENT DESIGN DATA		
Application	ABSORB	
Liquid volume	1118.207	M3
Design gauge pressure	100.002	KPAG
Design temperature	340.000	DEG C
Operating temperature	305.000	DEG C
COLUMN DATA		
Shell material	SS316	
Diameter option	OD	

Vessel diameter	7.5470	M
Vessel tangent to tangent height	25.000	M
Head type	HEMI	
MECHANICAL DESIGN DATA		
Wind or seismic design	W+S	
Fluid volume	20.000	PERCENT
Weld efficiency	85.000	PERCENT
Thickness Average	7.3636	MM
Corrosion allowance	0.000000	MM
THICKNESSES REQUIRED		
Thickness for internal pressure	8.8353	MM
Wind or seismic design thickness	12.364	MM
PACKING DATA		
Number of distributor plates	0	
Cross sectional area	44.734	M2
SECTION 1		
Packing type	M76YB	
Total packing height	6.0000	M
Packing volume	268.406	M3
Packing volume per unit height	44.734	M3/M
VESSEL SKIRT DATA		
Skirt material	CS	
Skirt height	9.5000	M
Skirt thickness	19.000	MM
NOZZLE AND MANHOLE DATA		
Nozzle ASA rating	150	CLASS
Nozzle material	SS316	
Nozzle A Quantity	2	
Nozzle A Diameter	900.000	MM DIAM
Nozzle A Location	S	
Nozzle B Quantity	2	
Nozzle B Diameter	750.000	MM DIAM
Nozzle B Location	S	
Nozzle C Quantity	8	
Nozzle C Diameter	50.000	MM DIAM
Nozzle C Location	S	
Number of manholes	6	
Manhole diameter	900.000	MM

PROCESS DESIGN DATA		
WEIGHT DATA		
Shell	35000	KG
Heads	12200	KG
Nozzles	34	KG
Manholes and Large nozzles	4800	KG
Skirt	33800	KG
Base ring and lugs	4500	KG
Ladder clips	140	KG
Platform clips	400	KG
Fittings and miscellaneous	70	KG
Total weight less packing	90900	KG
VENDOR COST DATA		
Packing cost	811822	EURO
Material cost	636050	EURO
Field fabrication cost	245121	EURO
Fabrication labor	6466	HOURS
Shop labor cost	79066	EURO
Shop overhead cost	81541	EURO
Office overhead cost	177102	EURO
Profit	182899	EURO
Total cost	2213600	EURO
Cost per unit weight	24.352	EUR/KG
Cost per unit height or length	368933.34	EUR/M
Cost per unit volume	8247.209	EUR/M3
Cost per unit area	49483.54	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	2213600.	48967.	820
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0

Appendix G–3: Aspen In-Plant cost estimation results for reboiler

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Desorber-Reboiler

Item Code: DRB THERMOSIPH
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	THERMOSIPH	
Number of identical items	1	
GENERAL DESIGN DATA		
Thermosiphon type	V-FXD	
TEMA type	BEM	
Heat exchanger design option	STAND	
Heat exchanger design+cost tool	ECON	
Heat transfer area	768.000	M2
Number of shells	1	
Number of tube passes	1	
Number of shell passes	1	

Vendor grade	HIGH	
SHELL DATA		
Shell material	SS316	
Shell diameter	1125.000	MM
Shell length	9.0000	M
Shell design gauge pressure	1000.001	KPAG
Shell design temperature	340.000	DEG C
Shell operating temperature	340.000	DEG C
Shell corrosion allowance	0.0	MM
Shell wall thickness	9.0000	MM
ASA rating Shell side	300	CLASS
Number of baffles	16	
Shell fabrication type	PLATE	
Expansion joint	NO	
TUBE DATA		
Tube material	316LW	
Number of tubes per shell	1070	
Tube outside diameter	25.000	MM
Tube length extended	9.0000	M
Tube design gauge pressure	1000.001	KPAG
Tube design temperature	340.000	DEG C
Tube operating temperature	340.000	DEG C
Tube corrosion allowance	0.0	MM
Tube wall thickness	1.2000	MM
Tube gauge	18	BWG
Tube pitch symbol	TRIANGULAR	
Tube pitch	32.000	MM
Tube seal type	SEALW	
TUBE SHEET DATA		
Tube sheet material	316L	
Tube sheet thickness	70.000	MM
Tube sheet corrosion allowance	0.0	MM
Channel material	316L	
PROCESS DESIGN DATA		
Duty	9.6909	MEGAW
Heat of vaporization	350.000	KJ/KG
Vaporization	20.000	PERCENT
Specific gravity tower bottoms	0.500000	

Molecular weight Bottoms	100.000	
HEAD DATA		
Head material Tube side	316L	
ASA rating Tube side	300	CLASS
Head thickness Tube side	9.0000	MM
WEIGHT DATA		
Shell	2400	KG
Tubes	7300	KG
Heads	530	KG
Internals and baffles	1000	KG
Nozzles	1200	KG
Flanges	1700	KG
Base ring and lugs	21	KG
Tube sheet	680	KG
Saddles	160	KG
Fittings and miscellaneous	100	KG
Total weight	15100	KG
VENDOR COST DATA		
Material cost	155149	EURO
Shop labor cost	29452	EURO
Shop overhead cost	33456	EURO
Office overhead cost	20479	EURO
Profit	21864	EURO
Total cost	260400	EURO
Cost per unit weight	17.245	EUR/KG
Cost per unit area	339.062	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	260400.	2805.	47
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0

Appendix G–4: Aspen In-Plant cost estimation results for condenser

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Desorber-Condenser

Item Code: DHE FIXED T S
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	FIXED T S	
Number of identical items	1	
GENERAL DESIGN DATA		
TEMA type	BEM	
Heat exchanger design option	STAND	
Heat exchanger design+cost tool	ECON	
Heat transfer area	232.000	M2
Number of shells	1	
Number of tube passes	1	
Number of shell passes	1	
Vendor grade	HIGH	

SHELL DATA		
Shell material	SS316	
Shell diameter	775.000	MM
Shell length	6.0000	M
Shell design gauge pressure	1000.001	KPAG
Shell design temperature	340.000	DEG C
Shell operating temperature	340.000	DEG C
Shell corrosion allowance	0.0	MM
Shell wall thickness	9.0007	MM
ASA rating Shell side	300	CLASS
Number of baffles	16	
Shell fabrication type	PIPE	
Expansion joint	NO	
TUBE DATA		
Tube material	316LW	
Number of tubes per shell	485	
Tube outside diameter	25.000	MM
Tube length extended	6.0000	M
Tube design gauge pressure	1000.001	KPAG
Tube design temperature	340.000	DEG C
Tube operating temperature	340.000	DEG C
Tube corrosion allowance	0.0	MM
Tube wall thickness	1.2000	MM
Tube gauge	18	BWG
Tube pitch symbol	TRIANGULAR	
Tube pitch	32.000	MM
Tube seal type	SEALW	
TUBE SHEET DATA		
Tube sheet material	316L	
Tube sheet thickness	48.000	MM
Tube sheet corrosion allowance	0.0	MM
Channel material	316L	
HEAD DATA		
Head material Tube side	316L	
ASA rating Tube side	300	CLASS
Head thickness Tube side	7.0000	MM
WEIGHT DATA		
Shell	1100	KG

Tubes	2200	KG
Heads	250	KG
Internals and baffles	480	KG
Nozzles	360	KG
Flanges	850	KG
Base ring and lugs	10	KG
Tube sheet	250	KG
Saddles	90	KG
Fittings and miscellaneous	50	KG
Total weight	5600	KG
VENDOR COST DATA		
Material cost	61280	EURO
Shop labor cost	12473	EURO
Shop overhead cost	13821	EURO
Office overhead cost	10749	EURO
Profit	11677	EURO
Total cost	110000	EURO
Cost per unit weight	19.643	EUR/KG
Cost per unit area	474.138	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	110000.	2382.	40
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**110000**	**2382**	**40**

Appendix G–5: Aspen In-Plant cost estimation results for lean-rich heat exchanger

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Lean/Rich Heat exchanger

Item Code: DHE FIXED T S
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	FIXED T S	
Number of identical items	1	
GENERAL DESIGN DATA		
TEMA type	BEM	
Heat exchanger design option	STAND	
Heat exchanger design+cost tool	ECON	
Heat transfer area	983.000	M2
Number of shells	1	
Number of tube passes	1	
Number of shell passes	1	
Vendor grade	HIGH	

SHELL DATA		
Shell material	SS316	
Shell diameter	1275.000	MM
Shell length	9.0000	M
Shell design gauge pressure	1000.001	KPAG
Shell design temperature	340.000	DEG C
Shell operating temperature	340.000	DEG C
Shell corrosion allowance	0.0	MM
Shell wall thickness	10.000	MM
ASA rating Shell side	300	CLASS
Number of baffles	18	
Shell fabrication type	PLATE	
Expansion joint	NO	
TUBE DATA		
Tube material	316LW	
Number of tubes per shell	1369	
Tube outside diameter	25.000	MM
Tube length extended	9.0000	M
Tube design gauge pressure	1000.001	KPAG
Tube design temperature	340.000	DEG C
Tube operating temperature	340.000	DEG C
Tube corrosion allowance	0.0	MM
Tube wall thickness	1.2000	MM
Tube gauge	18	BWG
Tube pitch symbol	TRIANGULAR	
Tube pitch	32.000	MM
Tube seal type	SEALW	
TUBE SHEET DATA		
Tube sheet material	316L	
Tube sheet thickness	80.000	MM
Tube sheet corrosion allowance	0.0	MM
Channel material	316L	
HEAD DATA		
Head material Tube side	316L	
ASA rating Tube side	300	CLASS
Head thickness Tube side	11.000	MM
WEIGHT DATA		
Shell	3000	KG

Tubes	9300	KG
Heads	750	KG
Internals and baffles	1500	KG
Nozzles	1200	KG
Flanges	2100	KG
Base ring and lugs	31	KG
Tube sheet	970	KG
Saddles	200	KG
Fittings and miscellaneous	100	KG
Total weight	19200	KG
VENDOR COST DATA		
Material cost	195353	EURO
Shop labor cost	36570	EURO
Shop overhead cost	42091	EURO
Office overhead cost	24661	EURO
Profit	26126	EURO
Total cost	324800	EURO
Cost per unit weight	16.917	EUR/KG
Cost per unit area	330.417	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	324800.	3705.	62
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**324800**	**3705**	**62**

Appendix G–6: Aspen In-Plant cost estimation results for lean heat exchanger

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Lean Heat Exchanger

Item Code: DHE FIXED T S
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	FIXED T S	
Number of identical items	1	
GENERAL DESIGN DATA		
TEMA type	BEM	
Heat exchanger design option	STAND	
Heat exchanger design+cost tool	ECON	
Heat transfer area	899.000	M2
Number of shells	1	
Number of tube passes	1	
Number of shell passes	1	
Vendor grade	HIGH	

SHELL DATA		
Shell material	SS316	
Shell diameter	1225.000	MM
Shell length	9.0000	M
Shell design gauge pressure	1000.001	KPAG
Shell design temperature	340.000	DEG C
Shell operating temperature	340.000	DEG C
Shell corrosion allowance	0.0	MM
Shell wall thickness	10.000	MM
ASA rating Shell side	300	CLASS
Number of baffles	18	
Shell fabrication type	PLATE	
Expansion joint	NO	
TUBE DATA		
Tube material	316LW	
Number of tubes per shell	1252	
Tube outside diameter	25.000	MM
Tube length extended	9.0000	M
Tube design gauge pressure	1000.001	KPAG
Tube design temperature	340.000	DEG C
Tube operating temperature	340.000	DEG C
Tube corrosion allowance	0.0	MM
Tube wall thickness	1.2000	MM
Tube gauge	18	BWG
Tube pitch symbol	TRIANGULAR	
Tube pitch	32.000	MM
Tube seal type	SEALW	
TUBE SHEET DATA		
Tube sheet material	316L	
Tube sheet thickness	75.000	MM
Tube sheet corrosion allowance	0.0	MM
Channel material	316L	
HEAD DATA		
Head material Tube side	316L	
ASA rating Tube side	300	CLASS
Head thickness Tube side	10.000	MM
WEIGHT DATA		
Shell	2800	KG

Tubes	8500	KG
Heads	670	KG
Internals and baffles	1400	KG
Nozzles	1200	KG
Flanges	2000	KG
Base ring and lugs	30	KG
Tube sheet	860	KG
Saddles	190	KG
Fittings and miscellaneous	100	KG
Total weight	17800	KG
VENDOR COST DATA		
Material cost	181521	EURO
Shop labor cost	33960	EURO
Shop overhead cost	38801	EURO
Office overhead cost	23054	EURO
Profit	24464	EURO
Total cost	301800	EURO
Cost per unit weight	16.955	EUR/KG
Cost per unit area	335.706	EUR/M2

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	301800.	3705.	62
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**301800**	**3705**	**62**

Appendix G–7: Aspen In-Plant cost estimation results for lean pump

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Lean Pump

Item Code: DCP CENTRIF
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	CENTRIF	
Number of identical items	1	
EQUIPMENT DESIGN DATA		
Casing material	SS316	
Design temperature	50.000	DEG C
Design gauge pressure	1000.000	KPAG
Fluid head	70.000	M
ASA rating	150	CLASS
Brake horsepower	618.897	KW
Driver power	630.000	KW
Speed	1500.000	RPM

Driver type	MOTOR	
Motor type	TEWAC	
Pump efficiency	82.000	PERCENT
Seal type	SNGL	
PROCESS DESIGN DATA		
Liquid flow rate	740.000	L/S
Fluid specific gravity	1.0000	
Fluid viscosity	1.0000	MPA-S
Power per liquid flow rate	0.851351	KW/L/S
Liquid flow rate times head	51800	L/S -M
WEIGHT DATA		
Pump	2400	KG
Motor	1800	KG
Base plate	490	KG
Fittings and miscellaneous	420	KG
Total weight	5100	KG
VENDOR COST DATA		
Motor cost	81122	EURO
Material cost	21506	EURO
Shop labor cost	51432	EURO
Shop overhead cost	52461	EURO
Office overhead cost	35109	EURO
Profit	38770	EURO
Total cost	280400	EURO
Cost per unit weight	54.980	EUR/KG
Cost per unit liquid flow rate	378.919	EUR/L/S
Cost per unit power	445.079	EUR/KW

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	280400.	16298.	272
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0

Appendix G–8: Aspen In-Plant cost estimation results for rich pump

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Rich Pump

Item Code: DCP CENTRIF
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	CENTRIF	
Number of identical items	1	
EQUIPMENT DESIGN DATA		
Casing material	SS316	
Design temperature	50.000	DEG C
Design gauge pressure	1000.000	KPAG
Fluid head	70.000	M
ASA rating	150	CLASS
Brake horsepower	596.316	KW
Driver power	600.001	KW
Speed	1500.000	RPM

Driver type	MOTOR	
Motor type	TEWAC	
Pump efficiency	82.000	PERCENT
Seal type	SNGL	
PROCESS DESIGN DATA		
Liquid flow rate	713.000	L/S
Fluid specific gravity	1.0000	
Fluid viscosity	1.0000	MPA-S
Power per liquid flow rate	0.841516	KW/L/S
Liquid flow rate times head	49910	L/S -M
WEIGHT DATA		
Pump	2300	KG
Motor	1800	KG
Base plate	460	KG
Fittings and miscellaneous	400	KG
Total weight	5000	KG
VENDOR COST DATA		
Motor cost	78501	EURO
Material cost	20511	EURO
Shop labor cost	48907	EURO
Shop overhead cost	49884	EURO
Office overhead cost	33627	EURO
Profit	37070	EURO
Total cost	268500	EURO
Cost per unit weight	53.700	EUR/KG
Cost per unit liquid flow rate	376.578	EUR/L/S
Cost per unit power	447.500	EUR/KW

Summary Costs

Item	**Material(EUR)**	**Manpower(EUR)**	**Manhours**
Equipment&Setting	268500.	16006.	267
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0

Appendix G–9: Aspen In-Plant cost estimation results for fan

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Flue Gas Fan

Item Code: EFN CENTRIF
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	CENTRIF	
Material	CS	
Actual gas flow rate	1406000.	M3/H
Application	HVY	
Speed	1500.000	RPM
Driver power	1120.000	KW
Source of quote	SG	
Driver type	MOTOR	
Total weight	24900	KG

Item	Material(EUR)	Manpower(EUR)	Manhours
Equipment&Setting	1225900.	13135.	218
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**1225900**	**13135**	**218**

Appendix G–10: Aspen In-Plant cost estimation results for seperator

Project : MASTER THESIS PR
Scenario : CO2 CAPTURING

Seperator

Item Code: DVT CYLINDER
Sizing Data
Design Data
Summary Costs

Sizing Data

Description	Value	Units

Design Data

Parameter	Value	Units
Item type	CYLINDER	
Number of identical items	1	
EQUIPMENT DESIGN DATA		
Liquid volume	754.052	M3
Design gauge pressure	100.000	KPAG
Design temperature	340.000	DEG C
Operating temperature	340.000	DEG C
Fluid specific gravity	1.0000	
SHELL DATA		
Shell material	A 516	
Diameter option	OD	

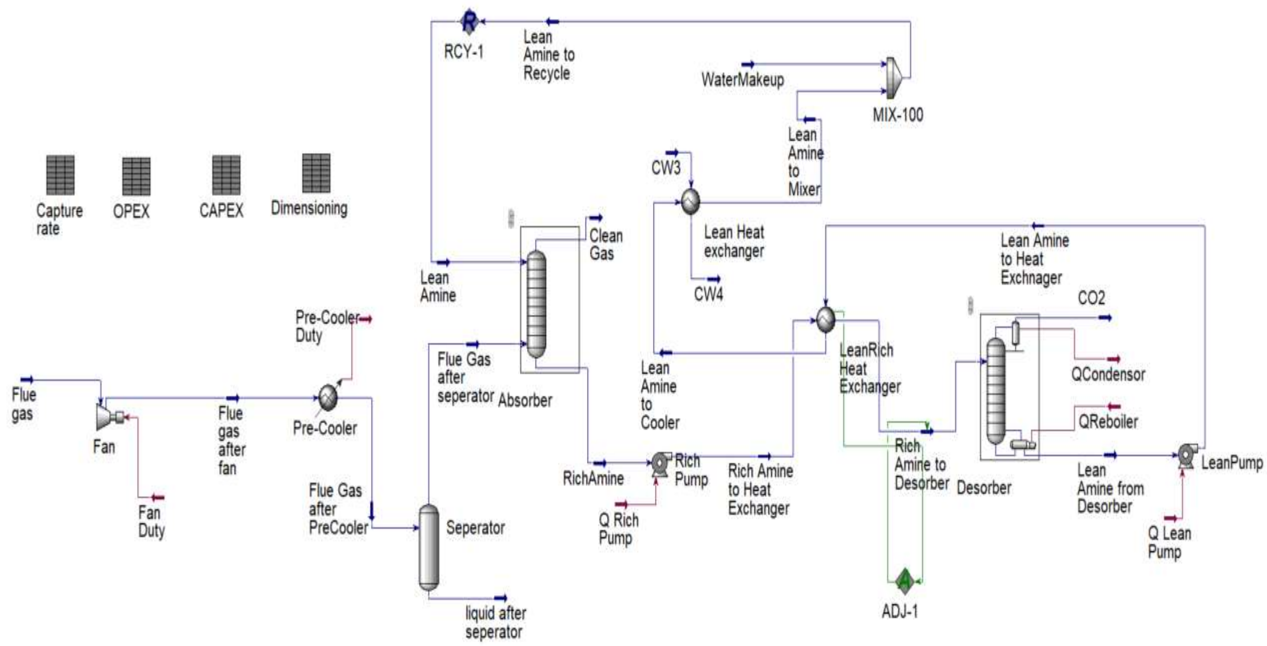
Vessel diameter	8.0000	M
Vessel tangent to tangent height	15.000	M
Head type	HEMI	
MECHANICAL DESIGN DATA		
Wind or seismic design	W+S	
Weld efficiency	85.000	PERCENT
Base material thickness	8.0003	MM
Corrosion allowance	3.0002	MM
Head thickness Top	8.0002	MM
Head thickness Bottom	8.0002	MM
THICKNESSES REQUIRED		
Thickness for internal pressure	4.6162	MM
Wind or seismic design thickness	5.0000	MM
VESSEL SKIRT DATA		
Skirt material	CS	
Skirt height	9.5000	M
Skirt thickness	18.000	MM
NOZZLE AND MANHOLE DATA		
Nozzle ASA rating	150	CLASS
Nozzle material	A 516	
Nozzle A Quantity	1	
Nozzle A Diameter	750.000	MM DIAM
Nozzle A Location	S	
Nozzle B Quantity	1	
Nozzle B Diameter	900.000	MM DIAM
Nozzle B Location	S	
Nozzle C Quantity	1	
Nozzle C Diameter	600.000	MM DIAM
Nozzle C Location	S	
Nozzle D Quantity	1	
Nozzle D Diameter	200.000	MM DIAM
Nozzle D Location	S	
Nozzle E Quantity	7	
Nozzle E Diameter	50.000	MM DIAM
Nozzle E Location	S	
Number of manholes	1	
Manhole diameter	450	MM
WEIGHT DATA		

Shell	23700	KG
Heads	12600	KG
Nozzles	60	KG
Manholes and Large nozzles	990	KG
Skirt	33900	KG
Base ring and lugs	4100	KG
Ladder clips	100	KG
Platform clips	330	KG
Fittings and miscellaneous	70	KG
Total weight	75900	KG
VENDOR COST DATA		
Material cost	139371	EURO
Field fabrication cost	101537	EURO
Fabrication labor	2364	HOURS
Shop labor cost	52542	EURO
Shop overhead cost	56016	EURO
Office overhead cost	59409	EURO
Profit	57327	EURO
Total cost	466200	EURO
Cost per unit weight	6.1423	EUR/KG
Cost per unit liquid volume	618.260	EUR/M3

Summary Costs

Item	Material(EUR)	Manpower(EUR)	Manhours
Equipment&Setting	466200.	33882.	568
Piping	0.	0.	0
Civil	0.	0.	0
Structural Steel	0.	0.	0
Instrumentation	0.	0.	0
Electrical	0.	0.	0
Insulation	0.	0.	0
Paint	0.	0.	0
Subtotal	**466200**	**33882**	**568**

Appendix H – Aspen HYSYS PFD for base case



Appendix I – Results for all the simulation

Table I-1: ΔT_{min}

ΔTmin	CAPEX(M€)	OPEX/yr(M€)	OPEX(8.5%)(M€)	NPV	MJ/kg CO2
5	138.20	27.66	271.62	409.82	2.71
6	131.80	27.69	271.92	403.72	2.75
7	128.00	27.88	273.78	401.78	2.80
8	124.80	28	274.96	399.76	2.84
9	122.10	28.19	276.83	398.93	2.89
10	120.20	28.52	280.07	400.27	2.95
11	118.20	28.87	283.50	401.70	3.05
12	116.60	28.91	283.90	400.50	3.11
13	114.80	29.33	288.02	402.82	3.18
14	113.30	29.33	288.02	401.32	3.17
15	112.80	29.77	292.34	405.14	3.26
16	112.00	29.91	293.72	405.72	3.30
17	110.90	30.26	297.15	408.05	3.40
18	111.20	30.69	301.38	412.58	3.49

Table I-2: Absorber packing height for different number of stages

No. Stages	Capex(M€)	Opex/yr(M€)	Opex 20year, 8.5%	NPV (M€)	MJ/kgCO2
12	449.4	69.95	686.91	1136	11.85
13	195	38.39	376.99	572	6.353
14	121.4	27.6	271.03	392	4.09
15	109.6	26.19	257.19	367	3.72
16	108.9	26.4	259.25	368	3.64
17	108.8	26.78	262.98	372	3.62
18	107.9	27.27	267.79	376	3.6
19	107.8	27.7	272.01	380	3.58
20	107.6	28.28	277.71	385	3.59

Table I-3: Calculation of NPV for different velocities

Velocity	Height of packing per stage(m)	Absorber (M€)	Fan(M€)	OPEX/yr(M€)	stages	pressure drop	Height	NPV
3	1.19	57.67	11.02	3.60	16	0.38 kPa/m	19.04	104.04
2.5	1	63.32	5.0	1.94	16	0.2 kPa/m	16	87.37
2	0.79	69.38	3.5	1.17	16	0.12 kPa/m	12.64	84.40
1.5	0.6	81.56	2.36	0.78	16	0.08 kPa/m	9.6	91.58

Table I-4: Calculation of NPV for different velocities

Velocity	Murphree efficiencies	Height of packing per stage(m)	Absorber cost(M€)	Fan(M€)	OPEX/yr(M€)	stages	pressure drop	Height	NPV
3	0.135	1.19	58.42	8.156	3.66	16	0.38 kPa/m	19.04	102.52
2.5	0.15	1	63.32	5.0	1.94	16	0.2 kPa/m	16	87.37
2	0.17	0.79	69.37	3.5	1.16	16	0.12 kPa/m	12.64	84.29
1.5	0.2	0.6	81.72	2.40	0.79	16	0.08 kPa/m	9.6	91.87

Appendix J – Pressure drop for different packings[7]

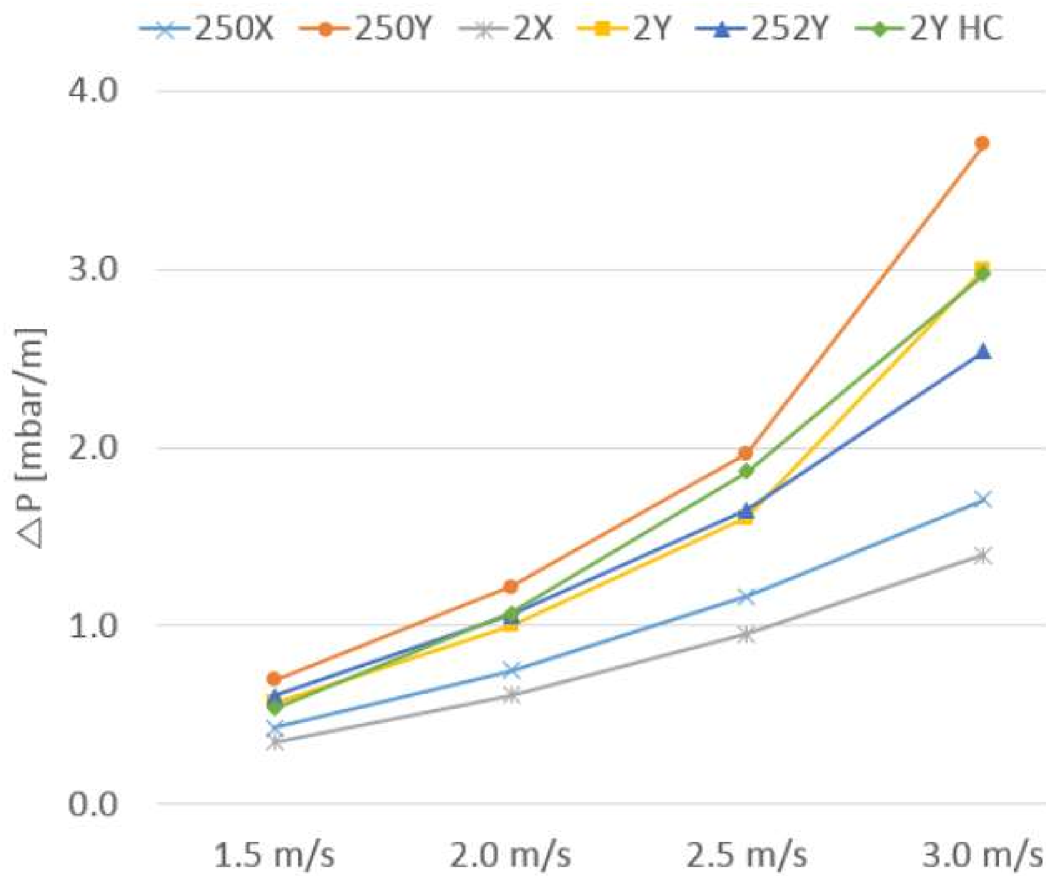


Figure 8. Pressure drop as a function of gas velocity for the different packings

Appendix K – Macro code for importing and exporting data between Aspen HYSYS and Excel[20]

```
Sub PRC_OPT()
  ' HYSYS main objects
    Dim hyApp As HYSYS.Application
    Dim hyCase As HYSYS.SimulationCase
    Set hyApp = CreateObject("HYSYS.Application")
    Set hyCase = hyApp.ActiveDocument

    ' Check if the hyCase is open or it is neccesary to find it in the path
    ' specified in the Sheets("SetUp").Range("B4")
    If hyCase Is Nothing Then
        Dim hyPath As String
        hyPath = Sheets("SetUp").Range("B4").Value2
        If hyPath = "FALSE" Or hyPath = "" Then
            MsgBox ("The Cell B4 is empty.")
        Else
            Set hyCase = GetObject(hyPath)
        End If
    End If

    'Read and write properties
    Set hySS = hyCase.Flowsheet.Operations.Item("CAPEX")
    Dim x As Variant
    Dim y As Variant
    Dim j As Integer
    Dim k As Integer
    ' This loop is to extract prices and name
    j = 18
    For i = 16 To 27
       x = 6
       y = i - 1
       Set hyCell = hySS.Cell(x, y)
       Set hyCell_part = hySS.Cell(x - 2, y)
       Cells(j, 3) = hyCell.CellValue
       Cells(j, 2) = hyCell_part.CellText
       j = j + 1
    Next i
    ' This loop is to import all the claculated factors from from excel to Hysys
    Set hySS = hyCase.Flowsheet.Operations.Item("CAPEX")
    x = 8
    i = 8
    For l = 1 To 3
        x = x + 1
        i = i + 1
        j = 18
        For k = 15 To 26
           y = k
           Set hyCell = hySS.Cell(x, y)
           hyCell.CellValue = Cells(j, i)
           j = j + 1
        Next k
    Next l
End Sub
```

Appendix L – Calculation of Murphree efficiency

Background:

In the simulation of the CO2 capturing process by using MEA(monoethanolamine), stages are assumed ideal in the absorber column. Murphree efficiencies(E_M) should be applied to the absorber calculation to get more accurate results. E_M is defined for a specific height of packing in the absorber. Packings in the absorber are one of the most expansive parts and should be tried to design more accurately to reduce the costs. Packings are designed with respect to maximum efficiency and minimum pressure drop. The first item, efficiency, caused an increase in the number of stages, and it has a direct effect on the CAPEX. Pressure drop has impacted the electricity that is used in the fan and the size of the fan. These explanations have shown that it has an impact on CAPEX and OPEX simultaneously. It has been investigated the differences between rigorous and less complicated methods for simulation of CO_2 absorption. Due to the challenges in convergence, complexity, and computing impact of different assumptions on the accuracy, it has been preferred to use methods with less complexity for simulations. If absorption fulfilling the conditions in the pseudo first-order regime, E_M can be used for calculating. The level of uncertainty in these methods is almost the same.[21]

The equation for Murphree efficiencies:

Formula of E_M for 1m packing height for the structured packing has been derived by Øi.[13] The main formulas for E_M in the current work are:

Overall tray efficiency and Murphree tray efficiency[Murphree, 1925] are connected in the general Coulson and Richardson[1991] equation. By rearranging the equation, E_M can be calculated from eq.(L.1) E_M computed from the eq when the m.V/L is specified

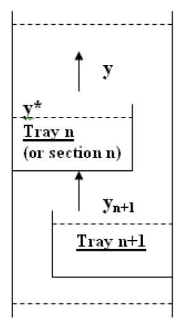


Figure L.1: mole fractions in the Murphree efficiency

$$E_M = \frac{\exp\left[\frac{H_{ELEM}}{HTU_G}.(m.V/L-1)\right]-1}{(m.V/L-1)} \tag{L.1}$$

By combining Van Krevelsen and Hoftijer [1948] with the absorption rate formula for the pseudo 1st order, which depends on the reaction, the overall mass transfer can be calculated and used in the HTU_G formula.

$$K_G a = \frac{1}{\frac{1}{k_G a} + \frac{He}{a.\sqrt{k_2.D_{CO2}.C_{Am}}}} \tag{L.2}$$

$$HTU_G = \frac{G}{K_G a.P} \tag{L.3}$$

$K_G a$: overall mass transfer [mol/(m^3.s.bar)]

G: molar gas flow per cross-section [mol/(m^2.s)]

V: molar vapor flow rates

L: molar liquid flow rates

m: slope of the equilibrium curve

H_{ELEM}: height of packing element

HTU: height of transfer unit

NTU: number of a transfer unit

P: pressure

C_{Am}: amine concentration

D_{CO2}: diffusivity coefficient[m^2/s]

H_e: Henry's constant

k_G: gas side mass transfer coefficient[m/s]

k_2: 2nd order reaction constant[m^3/(kmol.s)]

Murphree efficiencies can be calculated by importing $K_G a$ and HTU_G to the eq.(L.1). Some correction factors have shown up for Henry's constant, fraction effective area, and diffusivity coefficients in the calculations. For computing the percentage difference between E_M in different superficial velocities, for the constant volume of the packing, the effective area coefficient,a_{EFF}, should be updated in the calculating excel sheet. [13]

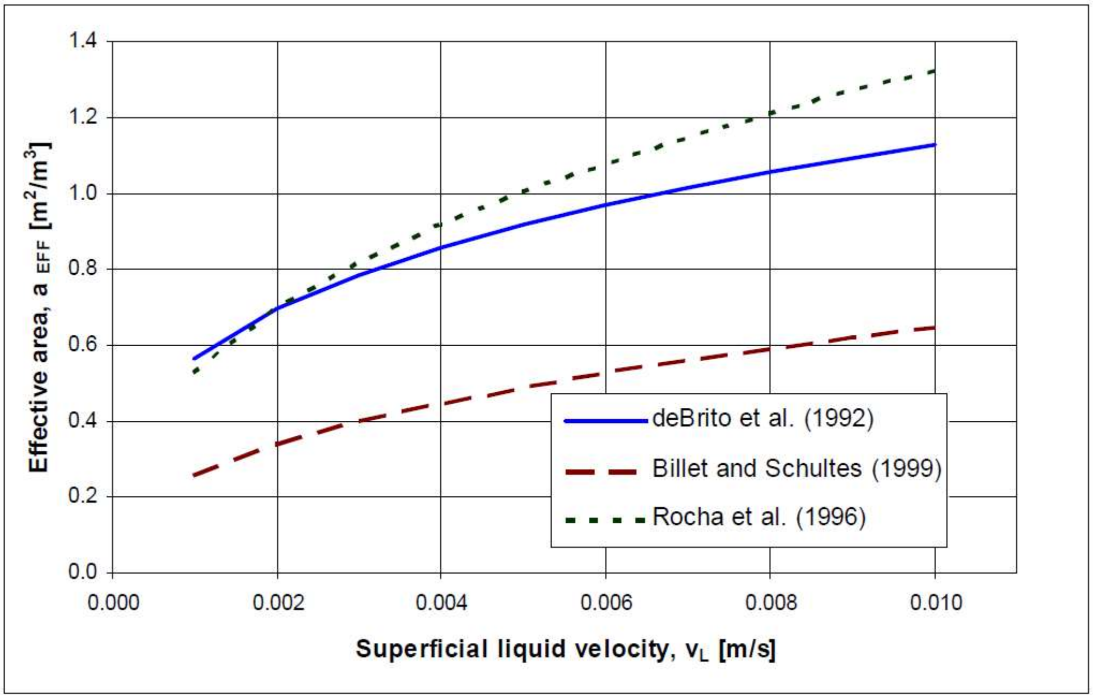


Figure L.2: Relation between effective area and superficial liquid velocity in the CO_2 absorption column[13]

Table L-1: Murphree efficiency calculation excel spreadsheet[13]

CALCULATION OF Enh from DeCoursey, SecorBeutler, % deviation from Pseudo 1st order				
CALCULATION OF E(Murphree)		DeCorSbTh8KE40t.xls		29/08/2021
Calculation based on typical conditions at top conditions (from Øi: SizeTop2).				
Equilibrium conditions are taken from Kent Eisenberg.				
Input parameters:				
Temp, C	Temperature	40		
Ptot, Pa	Pressure(total)	1.01E+05		1,0 atm (out)
PCO2out, Pa	Pressure(CO2out)	530		In: 3,5%
Load, molCO2/molMEA	Loading	0.25		
Rho, kg/m3	Density	1065		25 C
MfracMEA	MEA, Mass fraction	0.3		30 wt%
MwMEA, kg/mol	MwMEA	0.061		
G, mol/(m2s)	MoleFlow(tot)/Crossection	55		3 m/s
a, m2/m3	a(nominal)	250		250Y
kG, mol/(m2sPa)	Gas side kG	2.00E-05		Corrls 3 m/s
mV/L	Factor in Emurph calc.	0.01		SmallDummy
Corr(He)	Correction factor He(CO2)	1.3		Browning
Corr(my)	Correction factor my(CO2)	3		Weiland
Corr(a)	Correction factor area	0.6		deBrito/Billet
Calculations:				
T, K	Temp+273,15		313.15	
He(CO2inWat),Pa/(mol/m3)	2,82e6*exp(-2044/T)		4125.836506	Versteeg
D(CO2inWat), m2/s	2,35e-6*exp(-2119/T)		2.70593E-09	Versteeg
He(CO2), Pa/(mol/m3)	He(CO2inWat)*Corr(He)		5363.587458	
D(CO2), m2/s	D(CO2inWat)/Corr(my)^0.8		1.12E-09	
FracMEA	C(MEA)/C(AminTot)		0.504	From KentE
C(MEA), mol/m3	0,30*Rho/MwMEA*FracMEA		2639.803279	
C(CO2), mol/m3	PCO2out/He(CO2)		0.09881446	
k2, m3/(mol,s)	4,4e8*exp(-5400/T)		14.26990936	Versteeg
Ha (estim)	SQRT(k2*Corr(He)*D(CO2)*C(MEA))/kL		**74.1784552**	kL=0.0001
DeCoursey based on irreversible reaction (all diffusivities assumed equal):				
Enhi (irreversible rx, eq.diff)	1+C(MEA)/(2*C(CO2))		13358.37336	Check >>Ha
M	(Ha*Ha)		5502.443216	
Em1	Enhi-1		13357.37336	
EdeCorsey(ir)	Eq 20 from DeCoursey(1974)		**73.98228744**	**0.003**

DeCoursey based on reversible reaction (all diffusivities assumed equal):			
C(CO2)0		0.0049	KentE
q	C(MEA)/(C(CO2)i-C(CO2)0)	28108.59233	
gam1	C(Carb)/C(MEA)	0.488	KentE
gam2	C(HMEA+)/C(MEA)	0.496	KentE
K, dimless	Eq Constant, k-1/k1	124000	KentE
Em1rev	Eq 40 from DeCoursey (1982)	19577.93159	
EdeCrev	Eq 20 from DeCoursey(1974)	**74.04669606**	**0.002**
Secor and Beutler based on unequal diffusivities, reversible reaction:			
Psi	Eq 46 from Secor and Beutler, DN/DB and DM/DB=1)	1848.110438	
EnhSBi(uneqdiff but ratio=1	Eq 41 from Secor and Beutle(1967,1111)	19679.65685	
EnhSB(uneqdiff but ratio=1)	Eq 20 from DeCoursey(1974)	**74.04741129**	**0.002**
EnhSBi(uneqdiff, ratio 0,6)	Eq 41 from Secor and Beutler (sqrt(D/D))	15244.02205	
EnhSB(uneqdiff)	Eq 20 from DeCoursey(1974)	**74.00736898**	**0.002**
KgaLmol/m3s	(a*Corr(a)/HeCO2)*sqrt(D(CO2)*k2*Corr(He)*C(MEA))	2.59E-04	
KgaTot,mol/m3s	1/(1/KgaLiq+1/(kG*a*Corr(a))	2.43E-04	**0.065**
HTUg, m	G/(Kga*Ptot)	4.20	
Emurphree (1m)	(exp((He/HTUg)*(mV/L-1))-1)/(mV/L-1))	**0.212**	
Guess Enh and check DeCorsey/Thring equations 19 and 27			
Enh		17916.97784	(guess)
Beta (eq 19)	DB/DA=0,6	-0.645692679	
Theta (eq 27)(1,2,1,1)	DC/DA=0,5, DD/DA=0,5	1.000731735	~1,00
EdeCrev1211	Eq 20 from DeCoursey(1974)	**74.03387027**	**0.002**
Enh		8835.975786	(guess)
Beta (eq 19)	DB/DA=0,6	0.188466716	
Theta (eq 26)(1,1,1,1)	DC/DA=0,5, DD/DA=0,5	0.999956222	~1,00
EdeCrev1111	Eq 20 from DeCoursey(1974)	**73.8786799**	**0.004**
HTUdeC, m	HTUg*(1+error in EdCrev1111)	4.22	
EmurphreedeCrev (1m)	(exp((He/HTUdeC)*(mV/L-1))-1)/(mV/L-1))	**0.211**	

Hogendoorn (1997) revised Secor and Beutler equations solved by iteration:

C(Carb)		C(MEA)*gam1	1288.224		
C(HMEA+)		C(MEA)*gam2	1309.342426		
Guess C(Ci)			2375.757995		
C(Bi)	(1,2,1,1)	C(MEA)+2*(C(Carb)-C(Ci))	464.7352889		
C(Di)		C(HMEA+)+1*(C(Ci)-C(Carb))	2396.876421		
K(interface)		C(Ci)*C(Di)/(C(Bi)*C(CO2i))	124000.0009		
Error = K-K(i)			-0.000899677		
Enh(i)		1+sqrt(0,6)*(C(Ci)-C(Carb)/(C(CO2)0-C(C(	8970.866912		
ESBHogrev1211		Eq 20 from DeCoursey(1974)	**73.88327917**		**0.004**
Guess C(Ci)			3125.424713		
C(Bi)	(1,1,1,1)	C(MEA)+1*(C(Carb)-C(Ci))	802.602566		
C(Di)		C(HMEA+)+1*(C(Ci)-C(Carb))	3146.543139		
K(interface)		C(Ci)*C(Di)/(C(Bi)*C(CO2i))	124000		
Error = K-K(i)			2.1563E-07		
Enh(i)		1+sqrt(0,6)*(C(Ci)-C(Carb)/(C(CO2)0-C(C(	15154.03978	ca 682	
ESBHogrev1111		Eq 20 from DeCoursey(1974)	**74.00631438**		**0.002**