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Simulation and cost estimation of CO_2 capture processes using different solvents/blends

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Summary:

There has been a growing trend toward removing CO_2 emissions from the industry with different methods. One of the most mature methods for carbon capture is to absorb CO_2 in an amine-based (MEA) post-combustion technology. Shortcomings of MEA make other solvents and their blends more interesting in CO_2 removal plants.

The work in this master thesis is absorption-desorption CO_2 capture process simulated in Aspen HYSYS for different solvents/blends than MEA. Moreover, cost estimation methods for simulated cases have been performed to provide a complete cost estimation package. The data for cost estimation is provided with Aspen In-Plant Cost Estimator program.

A base case simulation model consisting of a simplified carbon capture unit including a 10-stage absorber, 6-stage desorption column, 85% CO_2 removal efficiency and minimum approach temperature for the lean/rich heat exchanger of 10 °C has undergone different solvents/blends of MEA, MDEA and PZ. The results indicate that adding 5 – 10 wt.% of piperazine to base case (30 wt.%) could offer a blend of solvents with lower regeneration energy than base case. Also, this matter was accurate for adding 5 – 20 wt.% MDEA to base case. Optimization of suggested range of blends has been performed in term of regeneration energy. Optimized concentrations could be as 30% MEA + 5% PZ (wt.%) and 30% MEA + 15% MDEA (wt.%) where lead into 4.9% and 7.5% lower regeneration energy than base case with 3.77 [MJ/kg CO_2]. These blends, also, have been simulated for vapor recompression configuration. Lean, rich and cyclic loadings for suggested blends in both standard and VR configurations have been discussed.

Aspen In-Plant Cost Estimator, applying Enhanced Detail Factor (EDF) method, was used for the cost estimation of processes. based on conducted cost estimations, plant with suggested blends presents cost savings rather than standard base case. Hopefully, the results in this thesis contribute to perform cost optimization more efficiently.

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Preface

This master thesis has been written during the spring of 2021 as a part of the master's program of "Process Technology" at the University of South-Eastern Norway.

The main focus of current project is on simulations of absorption-desorption CO_2 capture process in Aspen HYSYS with different solvents/blends than MEA and cost estimation with different methods which originates from defining a base case from real data from Statoil CO_2 Capture Study at Mongstad (2005).

Firstly, I want to show my appreciation to main supervisor, Lars Erik Øi, Professor at USN for great inspiration and help as well as valuable experiences from him in the process of conducting master thesis. Not only he attended regular meetings, but also, he shared his knowledge and helpful advice while writing this report. His review, feedback and motivation have been of great significance.

The information and reliable communication gathered from Solomon Aromada, PhD student at University of South-Eastern Norway, was invaluable. Especially the guidance on the cost estimation and the use of Aspen In-Plant Cost Estimator. Therefore, I would like to show my appreciation to him as well.

I have, also, chance to appreciate Neda Razi due to her valuable comments during the online meetings.

Finally, I would like to appreciate my dear wife, Sara, for her patience and support during this master thesis. I hope that we will have more time together the years to come.

It is highly recommended understand the CO_2 capture process, and to have knowledge about the Aspen HYSYS simulation tool before reading this report.

Porsgrunn, 2021

Sina Orangi

Nomenclature

CCUS	Carbon Capture Utilization and Storage
CH_4	Methane
CO_2	Carbon Dioxide
SS	Stainless Steel
CS	Carbon Steel
DCC	Direct Contact Cooling
EUR	Euro
GHG	Greenhouse Gases
HEX	Heat Exchanger
IEA	International Energy Agency
MEA	Monoethanolamine
MDEA	Methyl diethanolamine
PZ	Piperazine
N_2O	Nitrous Oxide
OPEX	Operational expenditures
CAPEX	Capital expenditure
USN	University of South-Eastern Norway
CEPCI	Chemical Engineering Plant Cost Index
VR	Vapor Recompression
ME	Murphree Efficiency
wt	Weight
LRHEX	Lean rich heat exchanger
Mol	Mole

Table of contents

1. Introduction

This part of the work covers a brief background for the project where the importance of topic will be discussed. In addition, health and environmental issues resulted from CO_2 will be pointed.

1.1. Background for the interest in CO_2 removal

Due to everyday industrialization, there is an increasing trend for global greenhouse gas emissions in the world which brings severe problems including environmental and health issues. Among greenhouse gases, Carbon Dioxide has the largest share, with more than 76% (Center for Climate and Energy Solutions , 2019).

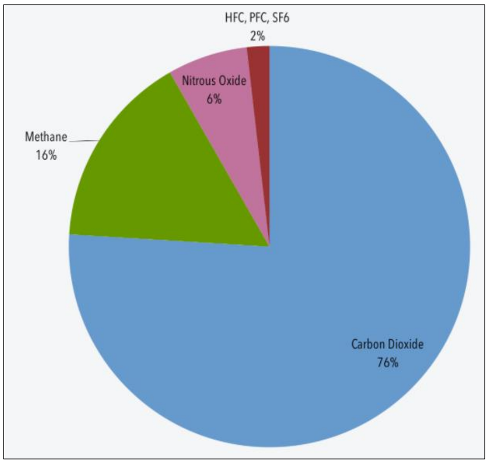

Figure 1.1: Contribution of greenhouse gases in emissions *(Center for Climate and Energy Solutions , 2019)*

Monthly Carbon Dioxide measured at Mauna Loa Observatory, Hawaii is displayed below. This figure emphasizes the increasing trend for CO_2 emissions from different sectors.

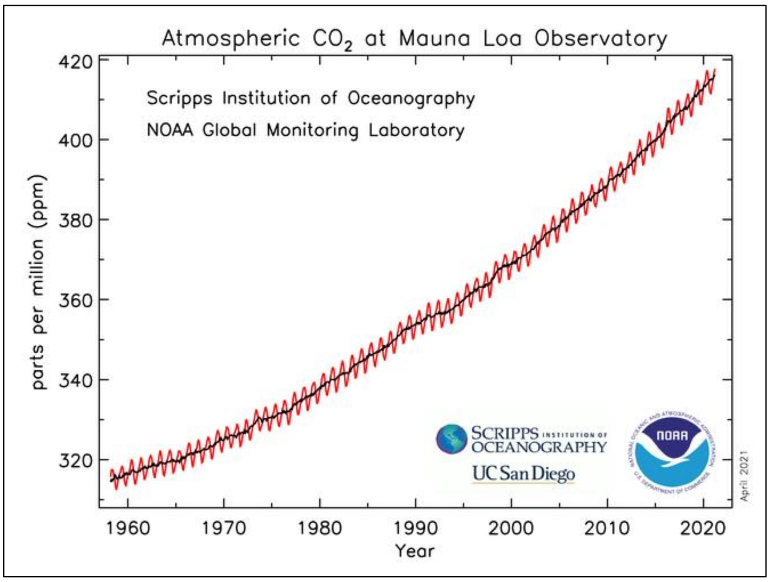

Figure 1.2: Monthly recorded CO_2 emissions at Mauna Loa Observatory, Hawaii *(Global Monitoring Laboratory, 2021)*

Contributions to CO_2 emissions vary sector by sector. A general view of these contributions is depicted in figure below.

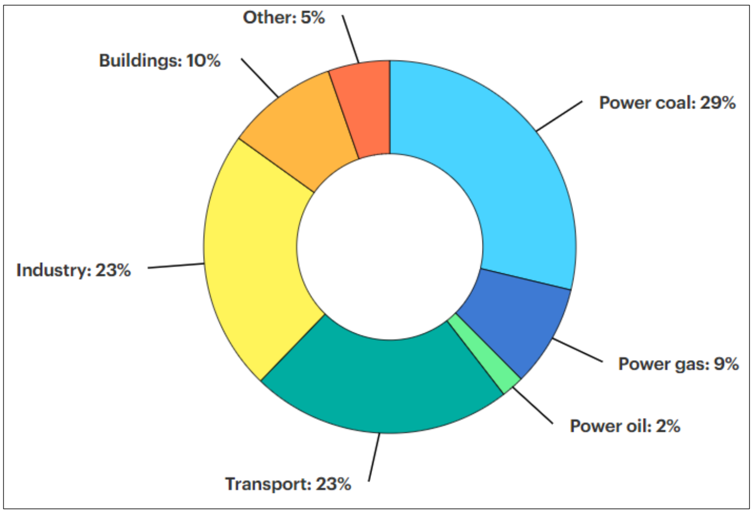

Figure 1.3: Global energy-related CO_2 emissions by sector *(Global energy-related CO2 emissions by sector, 2021)*

Amine-based gas cleaning in one of the most common and oldest procedures where an amine solvent is applied to scrub CO_2 from exhaust gases of plants. Different kinds of solvents might be used to satisfy this aim. Each solvent has advantages and disadvantages. As the most common, MEA solvent has been discussed in various works. New ones or their blends with more positive points will be studied in this work to bring more advantages e.g. lower energy in reboiler to plant, consequently lower cost.

1.2. Scope of thesis

The main aim for conducting this project is to suggest concentrations of other solvents, MDEA and piperazine, or their blends which provide CO_2 removal processes with lower regeneration energy comparing to base case where 30 wt% MEA was used. The importance for reducing thermal energy in reboiler is to its high share in operating and total costs of CO_2 removal plants. (Mudhasakul, Ku, & Douglas, 2013) estimated that approximately 70% of the operating cost arises from regeneration energy.

An optimized absorption-desorption process known as standard base case including 10-stage in absorber, 6-stage of stripper and 10℃ minimum approach temperature difference in lean/rich heat exchanger had been simulated. 30 wt.% monoethanolamine (MEA) was applied as solvent to reach 85% removal efficiency. The present work tends to investigate other solvents, MDEA and PZ, and their blends. Explained process is simulated in Aspen HYSYS version 10 and 12 to suggest those solvents/blends for lower regeneration energy.

Furthermore, cost estimation for simulated cases will be performed based on dimensioning and Enhanced Detail Factor (EDF) method to give insights for total installing costs for such removal plants as well as improvements in costs due to applying other solvents/blends than base case. The applied program which provides data for equipment is Aspen In-Plant Cost Estimator.

In addition, other important parameters including lean CO_2 loading, rich CO_2 loading and cyclic loading will be investigated within this work. Mentioned parameters highly affect the removal process.

As the final work, a study to suggest optimized concentrations for each blend in term of regeneration energy will be performed.

1.3. Outline of thesis

The first chapter of present work includes a brief introduction of current situation of CO_2 emission as well as health and environment issues resulted from Carbon Dioxide.

Chapter 2 begins with a summary for applicable methods where CO_2 is captured totally or partially. The chapter proceeds with explanations for two applicable configurations of amin-based method including standard and vapor recompression. Configurations will be completed with explanation for required equipment for each one. Classifications of different amines and their advantages, disadvantages and properties for each solvent, also, are included in chapter 2 as well as literature review for relevant works which are in line with current work.

In chapter 3, specifications for each mentioned configuration are tabulated.

Chapter 4 covers dimensioning and material selection for each piece of equipment in the removal plants. In chapter 5, mainly, cost estimation methods will be discussed. CAPEX and OPEX, in addition, are included.

The work will end up with chapter 6 where results and discussions are mentioned.

2. Description of CO_2 removal processes

This chapter describes general classification of CO_2 removal technologies, following with detail explanations for absorption-desorption removal process. The chapter continues with including those works in which other solvents/blends than MEA have been discussed.

2.1. Carbon capture technologies

This part covers main classifications of Carbon Dioxide removal technologies with a brief explanation for each one. Carbon capture technologies can be defined as processes or unit operations that separate CO_2 from gas mixtures to produce a CO_2-rich stream to be subsequently stored or utilised (Oreggioni, 2016).

Possible Carbon Dioxide removal methods are classified into three different procedures (Fagerheim, 2019) (Haukås, Helvig, Hæstad, & Lande, 2019) including

- Pre-combustion

where fossil fuel is converted to the synthesis gas for further combustion. In fact, a pre-combustion system involves converting solid, liquid or gaseous fuel into syngas without combustion, so that CO_2 can be removed from the mixture before the H_2 is used for combustion (Fagerheim, 2019)

- Oxy-combustion

where pure Oxygen, instead of air, completes combustion. This oxygen-rich, nitrogen-free atmosphere results in final flue-gases consisting mainly of CO_2 and H_2O (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020).

- Post-combustion

where removal of CO_2 from a conventional exhaust is conducted by chemical and physical processes. Chemical processes include solvent based configurations, calcium looping and enzymes (Oreggioni, 2016). Also, physical removal methods include adsorption or membrane process (Oreggioni, 2016). The absorbed CO_2 is compressed for transportation, storage or utilization (Haukås, Helvig, Hæstad, & Lande, 2019). Absorption-desorption process from post-combustion category is currently the most mature process for CO_2 capture (Fagerheim, 2019) (N.Borhani & Wang, 2019).

The figure below presents all explained processes.

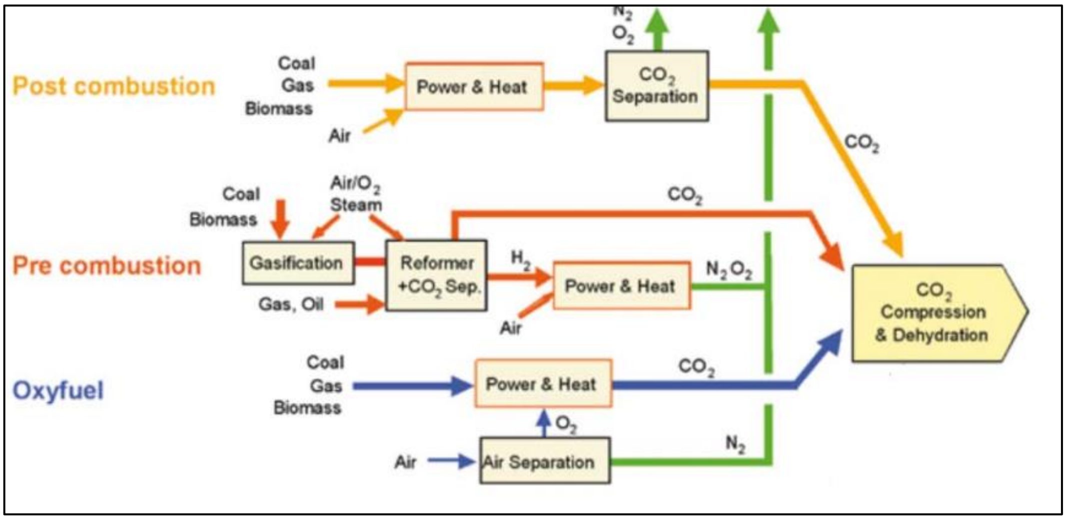

Figure 2.1: Classification and generic schematics for carbon capture technologies *(Oreggioni, 2016)*

2.2. Description of amine-based CO_2 capture process

As mentioned earlier, there are several CO_2 removal technologies developed based on different physical/chemical process. Among of them, absorption-based process involving amine solution is the most applicable method (Øi L. E., Removal of CO2 from exhaust gas) (Øi L. E., Aspen HYSYS Simulation of CO2 Removal by Amine Absorption from a Gas Based Power Plant, 2007), depicted in figure 2.2. The equipment involved in the process is absorption column, desorption, heat exchangers and auxiliary equipment (Øi L. E., Removal of CO2 from exhaust gas) (Øi L. E., Aspen HYSYS Simulation of CO2 Removal by Amine Absorption from a Gas Based Power Plant, 2007).

The main process involved is absorption into a mixture of an amine and water where the simplest and most popular amine to satisfy Carbon Dioxide removal is MEA (Øi L. E., Removal of CO2 from exhaust gas) (Øi L. E., Aspen HYSYS Simulation of CO2 Removal by Amine Absorption from a Gas Based Power Plant, 2007).

Apart from figure 2.2, the main processes within the Carbon Dioxide removal are based on absorption and desorption.

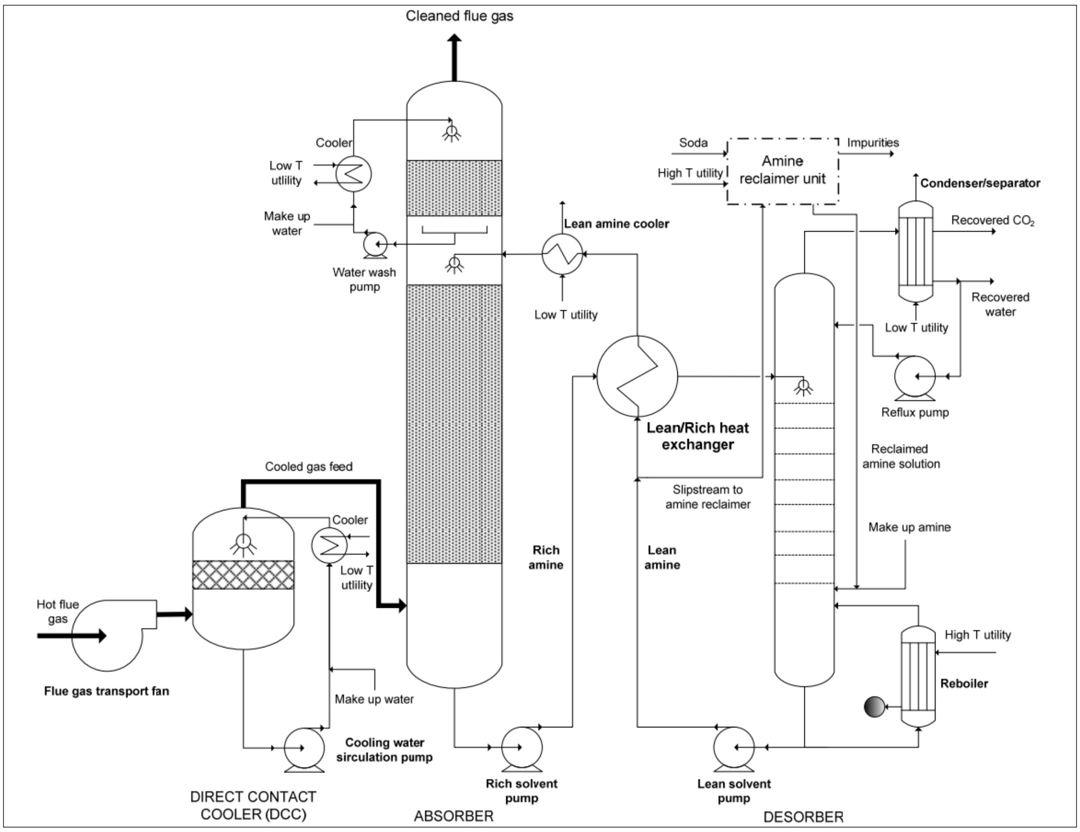

Figure 2.2: General flow diagram of CO_2 removal process plant *(Øi L. E., Removal of CO2 from exhaust gas) (Kallevik, 2010)*

2.3. Description of equipment in amine-based CO_2 capture plant

The process of scrubbing Carbon Dioxide into amine solution requires equipment mainly absorption and desorption columns. Other pieces of equipment are needed to complete a cyclic process including heat exchangers, pumps, fans and etc. Brief explanation for each piece of equipment can be found below.

2.3.1. Direct contact cooler (DCC)

DCC cools down the flue gas coming from power plant before the gas enters absorbing column. The reason for that is to ease the absorbing process. DCC unit includes three parts including direct contact vessel, water circulation pump and circulation water cooler (Kallevik, 2010). Enthalpy and consequently temperature of flue gas rise after passing through fans. Hence, flue gas is carried through cooling water to be reached a lower temperature.

2.3.2. Absorber column

CO_2 gas absorption and other chemical reaction happen in absorber. Flue gas enters absorption column from bottom while a mixture of solvent and water comes from top. The column is equipped with contact devices in order to maximize surface area between liquid solvent and flue gas (Kallevik, 2010). As the mixing of amine solution and CO_2-rich gas is exothermic, temperature alongside the absorber column slightly rises (Kallevik, 2010). In addition, the pressure in absorber column decreases from bottom to top.

2.3.3. Rich and lean amine pump

"Rich amine solution"[1] carrying high amount of absorbed Carbon Dioxide comes out of the bottom of absorber column. The solution should be sent to stripper column in order to separate CO_2 from amine solution. The required pressure for this process is supplied by rich pump.

Besides, regenerated solvent from stripper should be sent back to absorption column. This liquid contains lower amount of CO_2. That is why, this is called "Lean Amine". Lean amine pump performs this process.

2.3.4. Lean/rich heat exchanger

The rich amine solution from absorber requires to be heated before entering desorption column. The lean amine from stripper, also, requires to be cooled before entering the absorber. That is why, both stream exchanges heat in cross flow heat exchangers (Fagerheim, 2019). This reduces the duty of the reboiler in the desorption column as well as duty of the lean amine cooler which is responsible to reduce the temperature of lean amine to absorber (Haukås, Helvig, Hæstad, & Lande, 2019).

Figure 2.3 explains more regarding inlet and outlet streams into/out of lean rich heat exchanger.

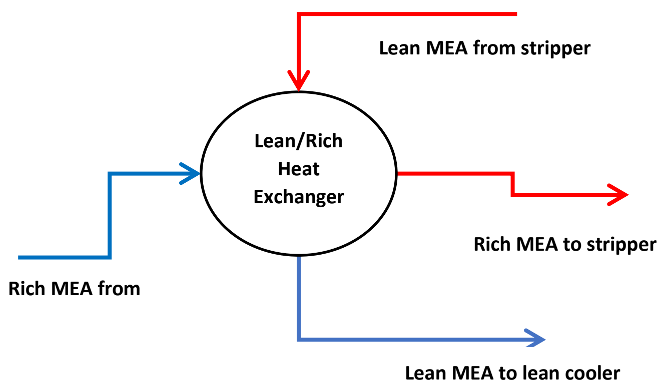

Figure 2.3: Schematic configuration of lean rich heat exchanger *(Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020)*

2.3.5. Stripper

Stripper or desorption column separates CO_2 from the amine solution with applying steam. Separated CO_2 gas leaves stripper from top, meanwhile regenerated solvent, lean amine, leaves column from the bottom.

In desorption column, pressure alongside of column can be assumed to be constant while the temperature decreases from the bottom to the top.

2.3.6. Reboiler

The amount of required heat to regenerate amine solution is the biggest part of the operational cost in absorption-based CO_2 removal plants. This amount of heat for the process is supplied by reboiler. In

[1] Also, it is called "rich amine loaded"

fact, reboiler is a sort of heat exchanger where the steam enters as hot stream to provide the required heat for the flow in the bottom of stripper.

2.3.7. Lean amine cooler

Lean amine cooler cools lean amine from lean/rich heat exchanger. The reason for reduction in the temperature is that optimized temperature of lean solvent to absorber is approximately 40℃ (Park & Øi, 2017).

2.4. Description of solvents and blends

Selection of solvent is imperative in CO_2 capture subject. Because, total cost and efficiency of the removal process are directly affected as results of CO_2 absorption capacity, size of equipment and regeneration energy (N.Borhani & Wang, 2019). That is why, this part of work is dedicated to review different solvents or their blends which can be applied to remove CO_2. Also, their properties, advantages and disadvantages for each one will be included. This part ends up with reviewing some works in which different solvents have been discussed.

2.4.1. Classification of solvents

(N.Borhani & Wang, 2019) classified solvents into three different groups of

- Chemical solvents
- Physical solvents
- Mixture solvents

The first group is known as Chemical solvents due to chemical reaction of solvent(s) with Carbon Dioxide. Amines, salt solutions and ammonia are some common examples of this type.

From reaction view, chemical solvents increase absorption rate of CO_2 at interface between gas and liquid phases.

From advantages view of chemical solvents, relative insensitivity to acid gases partial pressure, capture level of acid gases up to ppm and high absorption and desorption mass transfer coefficient (N.Borhani & Wang, 2019).

Some relevant disadvantages are high energy requirement for solvent regeneration, poor selectivity between acid gases, high price of materials, high heat of absorption, high corrosion, existence of side reactions, environmental damages (N.Borhani & Wang, 2019).

Table below includes some common chemical solvents and their properties.

Table 2.1: Physical Characteristics chemical solvents *(Arachchige & Melaaen, 2012) (N.Borhani & Wang, 2019)*

Family	Name	Formula	MW (g/mol)	Density (g/cm^3)	Melting Point (℃)	Boiling Point (℃)
Amine	MEA/Primary	C_2H_7NO	61.08	1.012	283.4	443
	MDEA/Tertiary	$C_{12}H_{17}NO_2$	119.16	1.038	-21	274.1
	DGA/Primary	$C_4H_{11}NO_2$	105.14	1.056	-12.5	221

	DEA/Primary	$C_4H_{11}NO_2$	105.14	1.097	28	271.1
	DIPA/Secondary	$C_6H_{15}N$	133.19	0.772	-61	84
	TEA/Tertiary	$C_6H_{15}NO_3$	149.19	1.124	21.60	335.4
	PZ/Cyclic diamine	$C_4H_{10}N_2$	86.136	1.1	106	146
Ammonia	-	NH_3	17.031	0.769	-77.73	-33.34
Salt solutions	Potassium carbonate	K_2CO_3	138.210	2.428	981	-
	Potassium bicarbonate	$KHCO_3$	100.12	2.170	292	-
	Sodium carbonate	Na_2CO_3	105.988	2.540	851	-

Pros and cons of those solvents used in this work, also, will be discussed following.

Monoethanolamine (MEA) is grouped as primary group of amine. This amine is proper to remove low amount of CO_2 from flue gases. The solution capacity of that is high as well as high reactivity with CO_2. In addition, process of production is easy. On the other hand, its shortcomings are high corrosiveness, poor thermal stability, low capacity for CO_2 absorption, high heat of reaction with CO_2 and high energy consumption for regeneration. Also, this sort of amine is not suitable for high pressure gas streams (N.Borhani & Wang, 2019) (Arachchige & Melaaen, 2012).

MDEA is another amine from tertiary group with different reaction mechanism with primary and secondary ones. On other words, no carbamate is formed. That is why, pure MDEA does not react effectively with CO_2 due to lack of N-H bonds (Arachchige & Melaaen, 2012) (Øi L. E., Removal of CO2 from exhaust gas). MDEA is highly resistant to degradation with lower corrosiveness than MEA. Heat of reaction with CO_2 and H_2S is low. Other improvement than MEA is to have higher CO_2 loading. Also, as MDEA does not react with COS and CS_2, solvent has lower lost. On the other hand, reaction rate of CO_2 with this amine is slow (N.Borhani & Wang, 2019) (Arachchige & Melaaen, 2012) (Hosseini-Ardali, Hazrati-Kalbibaki, & Fattahi, 2020).

Further, piperazine (PZ) is a cyclic secondary amine. Its advantages comparing with MEA are faster kinetics and higher capacity (N.Borhani & Wang, 2019). Also, it is more resistant to oxidative and thermal degradation (Nwaoha, et al., 2017). Due to high reactivity, PZ is usually added to other solvents as promoter (Ghalib, Ali, Ashri, Mazari, & Saeed, 2017).

As it can be seen, each solvent has favorable characteristics. Thus, combining them could use the positive features of each solvent.

2.4.2. Explanations for governing parameters

To better understanding conducted work, some definitions, firstly, are needed to be explained.

- Loading capacity

This parameter can be calculated by

$$\alpha = \frac{n_{CO_2}}{n_{amine}} \tag{2.1}$$

Where n_{CO_2} corresponds to the number of moles foe CO_2 component and n_{amine} represents number of moles for amine (Gomas & Santos, 2015). This parameter is defined as rich and lean.

- Absorption capacity

This term is defined as the moles of the absorbed CO_2 in 1 liter CO_2 loaded aqueous solution at equilibrium status (Zhang R. , et al., 2017). This parameter indicates the potential CO_2 carrying capacity of an amine and can be calculated by

$$Absoprtion\ Capacity\ (AC) = \alpha_{rich} \times C \tag{2.2}$$

Where α_{rich} is the CO_2 equilibrium loading of an amine solution, and C is the molar concentration of the amine solution. Both AC and C have similar unit of [mol/L] and α_{rich} is unitless.

- Cyclic capacity

This parameter refers to the amount of desorbed CO_2, and can be extracted from CO_2 loading in the liquid phase as

$$Cyclic\ Capacity\ (CC) = (\alpha_{rich} - \alpha_{lean}) \times C \tag{2.3}$$

In equation above, α_{rich} shows CO_2 loading of the initial amine solution and α_{lean} is the CO_2 loading of amine solution after regeneration (Nwaoha, et al., 2017).

- Capacity loading

This term is defined as difference of rich and lean CO_2 loadings. So,

$$Cyclic\ laoding = (\alpha_{rich} - \alpha_{lean}) \tag{2.4}$$

Both defined parameters, absorption capacity and cyclic capacity, should be made bigger to reach an ideal solvent.

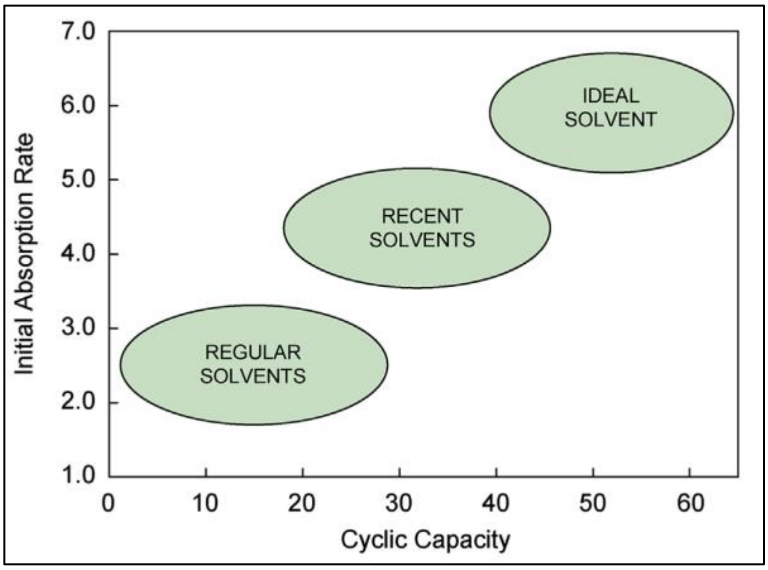

Figure 2.4: effect of absorption rate and cyclic capacity parameters to reach an ideal solvent *(Nwaoha, et al., 2017)*

- Regeneration energy

Regeneration energy is defined as a ratio of energy supplied from the reboiler and the mass rate of CO_2 released from the stripper (Li, Wang, & Chen, 2013). So,

$$Q_{reg} = \frac{H_{reboiler} - H_{loss}}{\dot{m}_{CO_2}} \tag{2.5}$$

Where Q_{reg} is the regeneration energy, $H_{reboiler}$ is the heat duty of the reboiler, H_{loss} is the loss of energy from reboiler and $\dot{m}_{CO_2}$ is mass flow rate of absorbed Carbon Dioxide which comes out from stripper. Commonly, H_{loss} is low comparing with heat of reboiler, so it can be neglected.

Further, $\dot{m}_{CO_2}$ can be calculated based on

$$\dot{m}_{CO_2} = \dot{m}_{solvent} C_{amine} (\alpha_{rich} - \alpha_{lean}) MW_{CO_2} \tag{2.6}$$

In expression above, $\dot{m}_{solvent}$ is the mass flow rate of rich solvent, C_{amine} is the molar concentration of amine and MW_{CO_2} indicates molecular weight of Carbon Dioxide.

The regeneration energy is defined is summation of

$$Q_{reg} = Q_{des,CO_2} + Q_{sen} + Q_{vap,H_2O} \tag{2.7}$$

$$Q_{reg} = \Delta H_{abs,CO_2} + \frac{\rho_{solvent} \dot{V} C_P (T_{reb} - T_{feed})}{\dot{m}_{CO_2}} + \frac{\dot{m}_{H_2O} \Delta H_{H_2O}^{vap}}{\dot{m}_{CO_2}} \tag{2.8}$$

Where $\Delta H_{abs,CO_2}$ shows the heat of reaction, C_P is the heat capacity of rich solvent, T_{reb} and T_{feed} are reboiler and feed solvent temperature to stripper respectively. $\dot{m}_{H_2O}$ is the mass flow rate of water vaporized from stripper and $\Delta H_{H_2O}^{vap}$ is the heat of vaporization (Li, Wang, & Chen, 2013).

Specific heat capacity of amine solvents can be assumed to be constant (Nwaoha, et al., 2017). Thus, the sensible heat of any amine solution can be believed to be influenced by their cyclic loading, amine concentration and density.

Regarding heat of vaporization, higher concentration amine benefits from having a smaller water concentration in solution, resulting into less latent heat of water vaporization (Nwaoha, et al., 2017). At the same time, heat of vaporization highly depends on regeneration temperature.

(Li, Wang, & Chen, 2013) investigated experimentally effect of each term in regeneration energy. They found out that heat of reaction and sensible heat are the main contributors to regeneration energy. Also, (Zhang R. , et al., 2017) indicates that Q_{des,CO_2} is main contributor while Q_{vap,H_2O} consists of 10% of total regeneration energy and 15-20% goes for Q_{sen}.

2.4.3. Literature Review on different solvents and blends

As discussed before, regeneration energy in Dioxide Carbon capture is extremely high, so CO_2 capturing process is regarded as an energy-intensive process. That is why many attempts are being made to reduce the regeneration energy. The importance of reducing regeneration energy requirement is that this parameter accounts for a large share of operational cost where according to (Nwaoha, et al., 2017) it is as high as 70% to 80% of OPEX. (Zheng, Ahmar, Simond, Ballerat-Busserolles, & Zhang, 2020) claimed that this value is 50% - 60% of total operating expense, OPEX, in CO_2 treatment process.

There are various parameters which affect regeneration energy requirement in CO_2 treatment process. This study intends to focus on implementation of other solvents or their blends than MEA via simulation of plant in Aspen HYSYS program to investigate the effects on regeneration energy penalty, consequently CAPEX and OPEX.

Many experimental and simulation studies have been conducted to analyse different solvents and blends to improve CO_2 capturing process. This part attempts to cover some of them.

(Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007) applied ASPEN Plus to present an optimal solution where amine lean solvent loading and MEA solution (wt.%) were 0.3 and 40 respectively, resulting into 23% reduction in thermal energy requirement than a base case with amine lean solvent loading of 0.242 [mol CO_2/mol MEA] and 30 wt.% MEA. Furthermore, (Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007) investigated the effect of MEA content on thermal energy requirement. According to the provided graph, more amount of solvent causes reduction in reboiler duty of process. It is worthy to mention that higher amount of solvent most likely leads to corrosion problems as well as more necessity for good washing section.

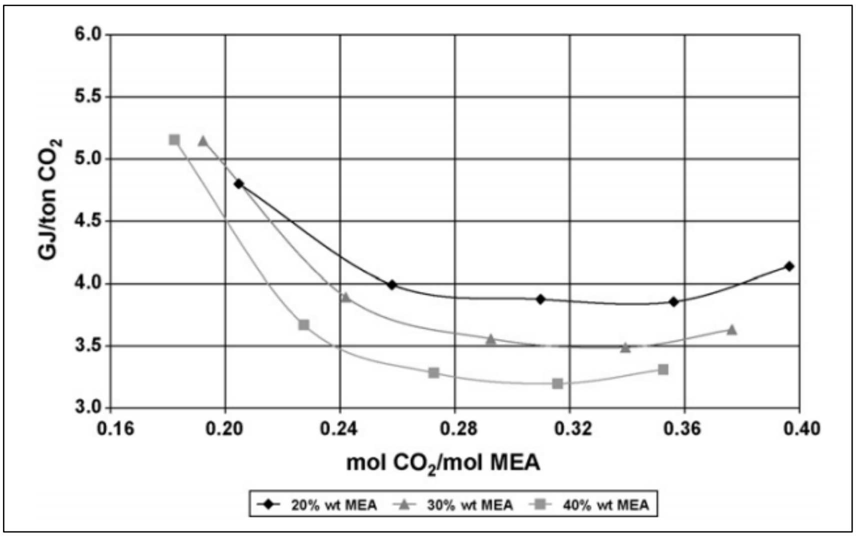

Figure 2.2: Thermal energy requirement at various CO_2/amine lean solvent loading for different MEA (wt%) *(Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007)*

Additionally, (Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007) investigated the relation of lean solvent temperature with regeneration energy requirement. The results are presented below through a diagram. According to provided figure, although lower temperature of lean solvent to absorber causes lower duty for reboiler, this imposes higher duty on the cooling water, consequently, rise in expense. Thus, a trade-off between both reboiler and cooling water duty is required.

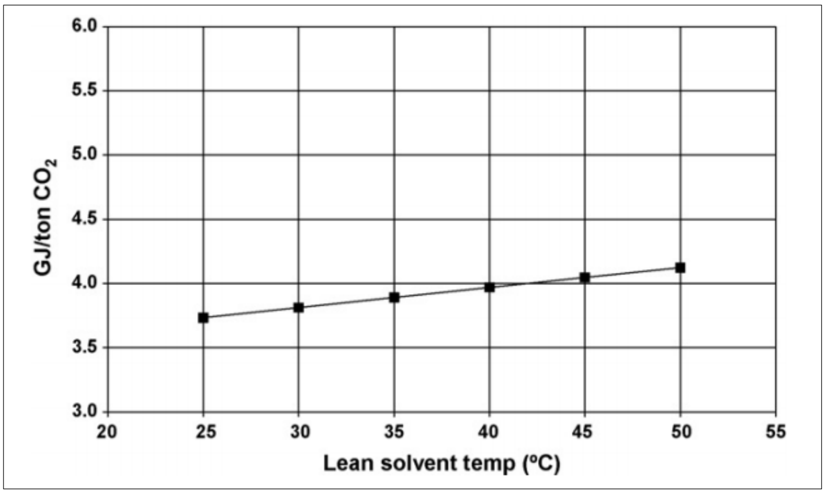

Figure 2.3: Thermal energy requirement for different lean solvent temperatures *(Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007)*

(Zhang R. , et al., 2017) applied experimental setup to compare trio-amine blend of MEA-MDEA-PZ with MEA, DEA, AMP and PZ. The total concentration of the blends was 6M combined in 3 different ways of, 3M MEA-2.5M MDEA-0.5M PZ (blend1), 3M MEA-2M MDEA-1M PZ (blend2) and 3M MEA-1.5M MDEA-1.5M PZ (blend3). Their results summarize as below.

- Blend3 had better performance in case of CO_2 equilibrium solubility, CO_2 absorption rate and absorption capacity comparing to other blends and 5M MEA.

- Applying Gibbs-Helmholtz equation to calculate absorption heat showed lower value for the blends rather than each individual solvent. The reason for mentioning heat of absorption is that lower value for absorption heat causes lower regeneration energy in the process.
- Analysing CO_2 desorption performance showed lower relative energy consumption for the blends compared to 5M MEA, 15.22-49.92% reduction compared to 5M MEA. Such study is presented below.

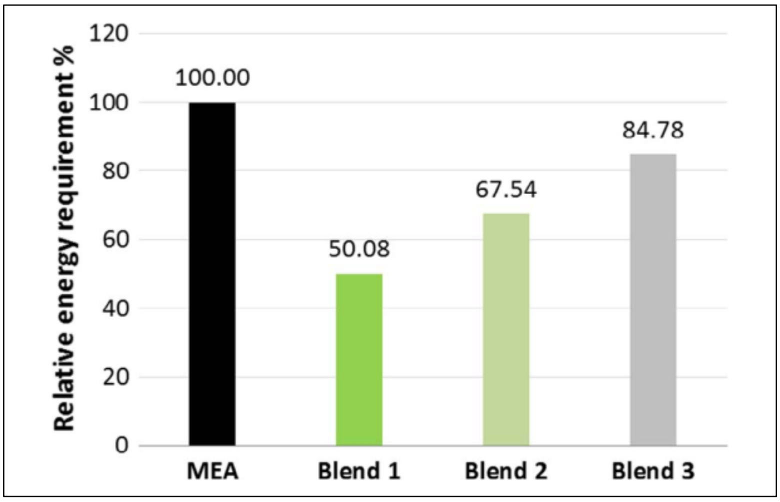

Figure 2.4: Relative energy consumption for concentrated MEA-MDEA-PZ blends compared to 5M MEA (as a benchmark with 100%) *(Zhang R. , et al., 2017)*

(Lee, et al., 2020) executed new blending recipes of a polyamine-based solvents in a 2MW coal-fired pilot-scale carbon capture process to investigate regeneration energy requirement. The recipes consist of IBD, MEA, AMP

and BAE amines which are combined in three different ways.

- IMP consists of 30 mass% IBD + 20 mass% MEA + 6 mass% PZ
- IAP consists of 40 mass% IBD + 6 mass% AMP + 6 mass% PZ
- IBP consists of 40 mass% IBD + 6 mass% BAE + 6 mass% PZ

PZ, also, was added as the reaction rate enhancer. The baseline process was 30 mass% MEA solvent and defined to be compared with presented polyamine-based solvents.

The results show a reduction of 0.7 GJ/ton CO_2 for IAP rather than MEA solvent. The study excluded the investigation for optimal ratio of solvents in blends.

(Arachchige & Melaaen, 2012) simulated a blended solvent of MDEA/MEA with 4:1 mixing ratio in weight basis via Aspen Plus to find out an optimal solution of 85% removal efficiency. In similar processes, the regeneration energy requirement of ca. 3.8 MJ/kg CO_2 was investigated for the MEA solvent versus 2.9 MJ/kg CO_2 for mentioned blended solvent.

(Ghalib, Ali, Ashri, Mazari, & Saeed, 2017) developed a thermodynamic model to predict the vapor liquid equilibrium of CO_2 in aqueous mixtures of MDEA/PZ. It was found out that addition of PZ as an activator to MDEA rise up the solubility of CO_2. However, volatility of amine system increased in low partial pressure of Dioxide Carbon.

(Idem, et al., 2006) evaluated the benefits of a 4:1 molar ratio blended solvent of MEA/MDEA in terms of heat requirement for solvent regeneration via pilot-scale capture plant. The results were compared with MEA solvent and found out a huge reduction in energy requirement.

(Mudhasakul, Ku, & Douglas, 2013) simulated the acid gas removal unit of an actual natural gas sweetening process via Aspen Plus. It was found out that piperazine has a significant impact on the process performance. For instance, every 1 wt% increase in PZ enhance the CO_2 recovery by ca. 10%. Also, the best trade-off between CO_2 recovery and energy consumption occurred at 5 wt% concentration of PZ and 45 wt% aqueous MDEA solvent.

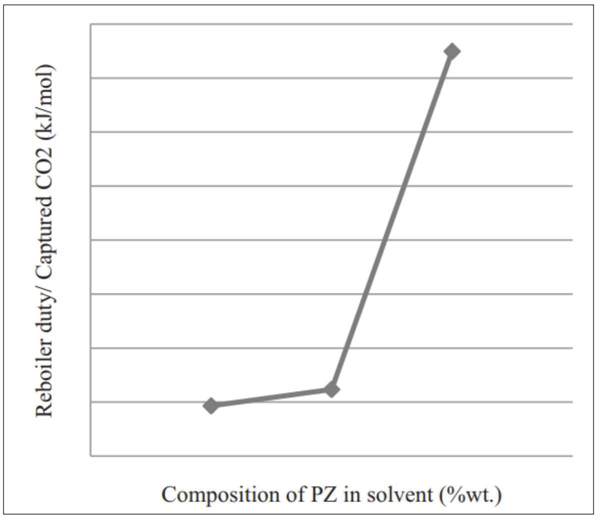

Figure 2.5: Effect of PZ concentration in solvent on the reboiler duty per captured CO_2 of the stripper *(Mudhasakul, Ku, & Douglas, 2013)*. The lowest point refers to 5 wt% PZ in blend, the middle point shows 7.5 wt% PZ and 10 wt% PZ goes with third point.

(Mangalapally & Hasse, 2011) experimented two amine solvents from EU-project CESAR in a pilot plant for 90% removal efficiency. They compared their work with MEA solvent for similar conditions. It was found out 20% reduction in regeneration energy. The applied solvent compositions consist of

- MEA : 0.3 g/g Monoethanolamine + 0.7 g/g water
- CESAR1 : 0.28 g/g AMP + 0.17 g/g PZ + 0.55 g/g Water
- CESAR2 : 0.32 g/g EDA + 0.68 g/g Water

Their results for the MEA solvent showed 3.8 [GJ/ton CO_2] regeneration energy, while this value for the CESAR1 and CESAR2 were 3.0 [GJ/ton CO_2] and 3.45 [GJ/ton CO_2] respectively.

(Zheng, Ahmar, Simond, Ballerat-Busserolles, & Zhang, 2020) experimentally investigated CO_2 behaviour in an aqueous solution for 3 different cases of 30 wt.% MDEA, 50 wt.% MDEA and a blended solvent of 40 wt% MDEA + 10 wt% PZ. It was found out that heat of absorption was enhanced by adding PZ. Also, PZ has no effect on CO_2 capture capacity.

(Khan, et al., 2020) presented a simulation for large-scale 650 MW coal power plant based on MDEA/PZ solvent via Aspen Plus V.10. It was found out an appropriate concentration of 35 wt% MDEA and 15 wt% PZ results in an optimal solution form minimization energy view, simulating 4 different cases including 45/5, 40/10, 35/15 and 30/20 wt%. The effect of applying optimal blended solvent on the regeneration energy led to 24.6% reduction. The 3 remaining cases, also, presented lower energy requirement comparing with MEA base case.

(Abd & Naji, 2020) simulated a real process of acid gas CO_2 capture in Aspen HYSYS V.8.8 to determine the effects of adding different concentrations of activators up to 10% with maintaining the constancy of the entire amine strength of 45%. PZ and Sulfolane were selected as an activator to be added to MDEA. The results showed 5% activator and 40% MDEA has better performance regarding energy consumption. Also, it was investigated that addition of 5% PZ improves the absorption of CO_2 by 92.1%. another result of their work presented more effectivity of Sulfolane than PZ from minimization energy view. Their work for MDEA/PZ consisted of four different cases for solvent concentration like

- Case1: Piperazine 0.02 mol% + MDEA 0.43 mol% + Water 0.55 mol%
- Case2: Piperazine 0.05 mol% + MDEA 0.4 mol% + Water 0.55 mol%
- Case3: Piperazine 0.07 mol% + MDEA 0.38 mol% + Water 0.55 mol%
- Case4: Piperazine 0.1 mol% + MDEA 0.35 mol% + Water 0.55 mol%

Table 2.2: Investigation of improvements in absorption-desorption CO_2 removal

Reference	Study method	Amin/Blend content	Conditions	Effectiveness
(Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007)	Aspen Plus simulation	40 wt% MEA	210 kPa stripper	3.0 GJ/ton CO_2 for 90% removal (23% reduction compared to base case with 30% MEA)
(Zhang R. , et al., 2017)	Experimental setup	3M MEA, 2.5M MDEA, 0.5M PZ 3M MEA, 2M MDEA, 1M PZ 3M MEA, 1.5M MDEA, 1.5M PZ	-	16% to 50% reduction of regeneration energy compared to base case with 5M MEA
(Lee, et al., 2020)	Experimental of 2 MW pilot-scale coal-fired plant	30 wt% IBD, 20 wt% MEA, 6 wt% PZ 40 wt% IBD, 6 wt% AMP, 6 wt% PZ 40 wt% IBD, 6 wt% BAE, 6 wt% PZ	150 Nm3/h flue gas	0.7 GJ/ton CO_2 reduction for IAP case compared to 30% MEA base case
(Arachchige & Melaaen, 2012)	Aspen Plus simulation of 500 MW coal-fired plant	10% MEA, 40% MDEA	674 kg/s flue gas	2.93 GJ/ton CO_2 compared to 3.80 GJ/ton CO_2 for base case with 25% MEA
(Idem, et al., 2006)	Pilot-scale tests for coal-fired plant	MEA/MDEA 4:1	5 kmol/m^3 aqueous blend	Huge reduction compared to MEA case
(Mudhasakul, Ku, & Douglas, 2013)	Aspen Plus simulation	45% MDEA, 5% PZ	-	The case is optimal solution from regeneration energy view with total 50 wt.% content.
(Mangalapally & Hasse, 2011)	Pilot plant tests	28 wt% AMP, 17 wt% PZ, 55 wt% Water 32 wt% EDA, 68 wt% Water	30 -110 kg/h flue gas	3 GJ/ton CO_2 and 3.45 GJ/ton CO_2 respectively compared to 4.1 GJ/ton CO_2 for 30% MDA as base case
(Zheng, Ahmar, Simond, Ballerat-Busserolles, & Zhang, 2020)	Experimental study	50 wt% MDEA 40 wt% MDEA, 10 wt% PZ	Pressure from 0.5 to 4 MPa	Addition of PZ increases heat of absorption
(Khan, et al., 2020)	Aspen Plus large-scale coal power plant	45 wt% MDEA, 5 wt% PZ 40 wt% MDEA, 10 wt% PZ 35 wt% MDEA, 15 wt% PZ 30 wt% MDEA, 20 wt% PZ	775 ton/hr flue gas 2.3 bar pressure at stripper	MDEA 35wt% + PZ 15wt% presents the best regeneration energy, 3.235 GJ/ton CO_2, compared to other cases
(Abd & Naji, 2020)	Aspen HYSYS simulation	43 wt% MDEA, 2 wt% PZ 40 wt% MDEA, 5 wt% PZ 38 wt% MDEA, 7 wt% PZ 35 wt% MDEA, 10 wt% PZ	1133 kmol/hr flue gas	40 wt% MDEA, 5 wt% PZ case has lower regeneration energy among others. Also, Sulfolane acts better than PZ.
(Dubois & Thomas, Comparison of	Aspen HYSYS simulation	30 wt% MEA 40 wt% PZ 10 wt% MDEA, 30 wt% PZ	3997 m^3/h flue gas	Regeneration energy of 3.36, 3.14 and 2.75 GJ/ton CO_2 respectively

varipous configurations of the absorbtion-regereration process using different solvents for the post-combustion CO2 capture applied to cement plant flue gas, 2018)			Stripper pressure for second and third cases at 600 kPa	

2.5. Description of other configurations than standard process

Except standard process depicted in figure 2.2, other configurations e.g. vapor recompression and spit-stream are common processes. Consumption of heat in desorption column is high, so these processes are suggested to reduce this heat consumption (Øi, et al., 2014).

The difference of mentioned configuration with standard one mainly refers absorption and desorption columns. The standard process has simple absorber and desorber.

Vapor recompression configuration is depicted in figure 2.8 below where a regenerated amine solution from the bottom of desorber passes a valve to have a reduction in pressure and enters flash tank. In the flash tank, the liquid lean amine from the bottom of separator leaves the tank to be recirculated back to absorber with a lean pump. Gaseous phase of flow leaves the top of flash tank. This vapor is compressed with a compressor to reach the same pressure with stripper and is sent to desorber. It should be, also mentioned that used valve in this process is a linear one with 50% opening to decrease the pressure of outlet flow form stripper from 200 kPa to 100 kPa as the entry pressure for separator.

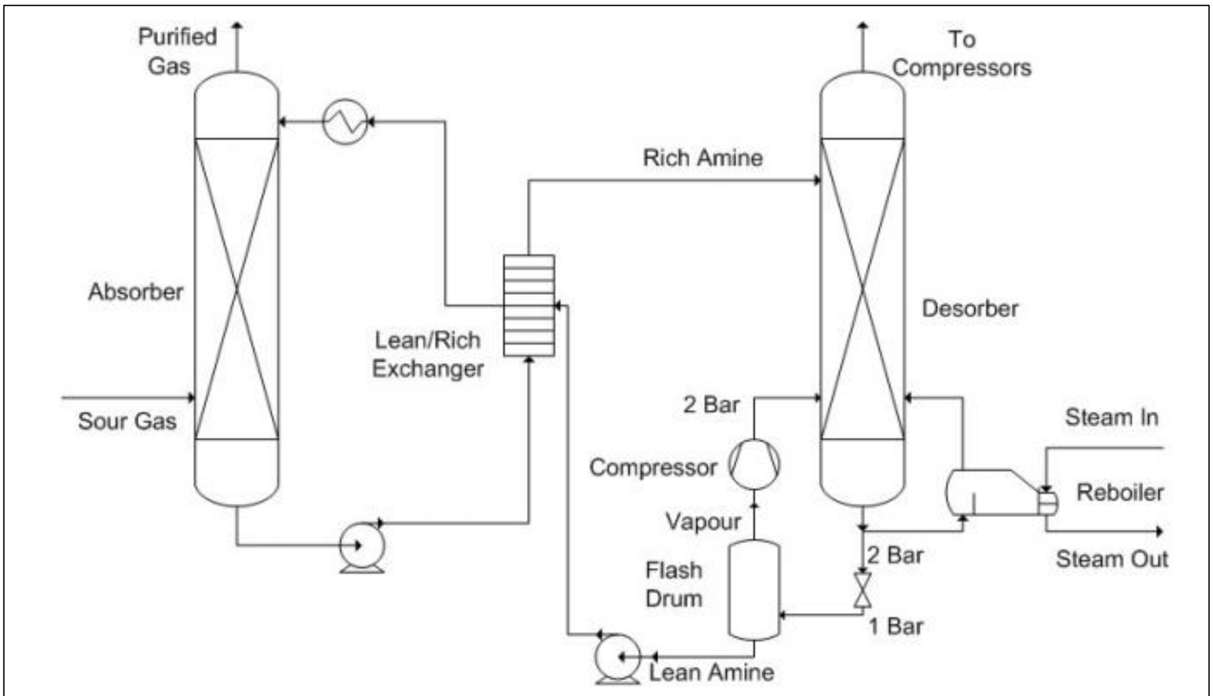

Figure 2.8: Principle for a CO_2 capture process with vapor recompression *(Øi, et al., 2014)*

The temperature in the process should not exceed 120℃ when MEA is the solvent because amine will degenerate (Øi, et al., 2014).

Other solvent, MDEA and PZ, can have different temperature in stripper than MEA. (Dubois & Thomas, Comparison of varipous configurations of the absorbtion-regereration process using different solvents for the post-combustion CO2 capture applied to cement plant flue gas, 2018) simulated the standard process of removal for PZ with 150℃ in the stripper. (Khan, et al., 2020) for the mixture of MDEA and PZ (MDEA 45 wt% + PZ 5 wt%) simulated the process with 125℃ in the stripper.

Temperature in the stripper is significant parameter because high temperature in the stripper leads to degradation of amine where irreversible chemical reaction affects solvent. This phenomenon imposes severe problems for the process including higher corrosion rate, increased amine make-up, significant increase in viscosity which leads to higher duty for pumps, mass transfer limitation (Nwaoha, et al., 2017).

Figure 2.9 presented below includes some difficulties resulted from degradation.

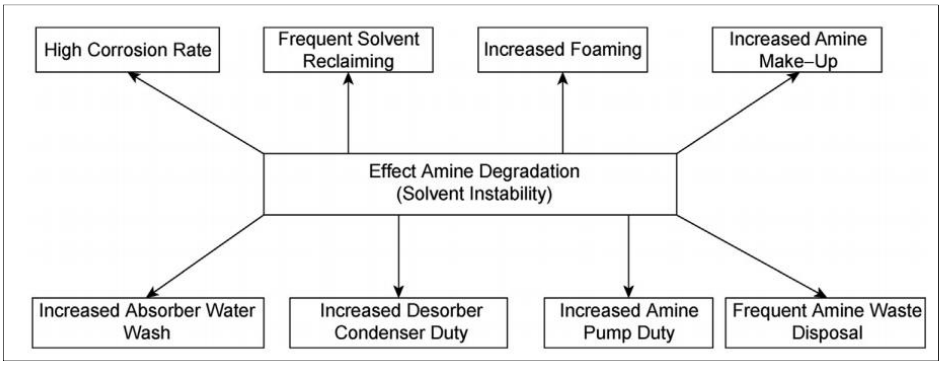

Figure 2.9: Effect of amine degradation towards plant operating costs *(Nwaoha, et al., 2017)*

(Nwaoha, et al., 2017), also, mentioned that though the blends of amines can provide CO_2 removal processes with lower regeneration energy, degradation for mixture of solvents is more reported than individual solvents. In addition, they suggest that degradation can be minimized with flue gas conditioning, development of amine solvents, lower temperature for regeneration.

Split-stream configuration is another process with lower regeneration energy than standard process. A partly regenerated amine solution (semi-lean amine solution) is extracted from the middle of stripper and is sent to the middle of the absorber with the aid of a pump. This process schematically is depicted below in figure 2.10.

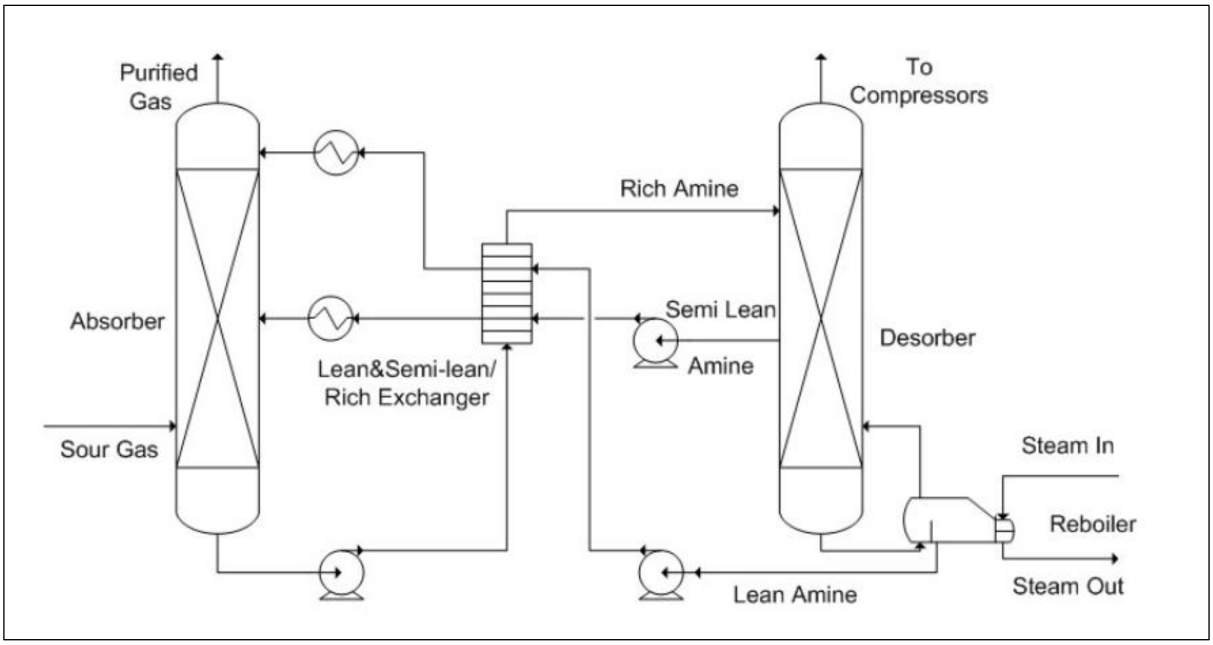

Figure 2.10: Flow diagram of split-stream removal process *(Øi, et al., 2014)*

3. Simulations in Aspen HYSYS

This part, firstly, covers procedures of simulations. Standard removal process and vapor recompression ones are the processes in this work as they have been explained in previous chapter. the work begins with defining a base case where 30 wt% MEA as solvent enters the absorber. Relevant specifications for this case will be included following. The work proceeds with applying other solvents and their blends as solvent with different concentrations. The reason for this is to have CO_2 removal processes with improvements rather than base case. Also, it will be tried to find an optimal blend which improves the removal process in governing parameters especially regeneration energy.

This chapter, additionally, covers specifications for vapor recompression process. Also, this process will be simulated with blends of solvents and improvements will be explained.

3.1. Specification of base case simulation

The simulation begins with selecting properties where components participating in the process should be chosen. Afterwards, package should be defined for the program. The current work bases on chemical solvents – acid gas package where the amines and their blends are being supported. One of the limitations for the chemical solvents-acid gas package in this work is that this package does not include tri-amine blends of MEA, MDEA and PZ.

The process proceeds with defining pieces of equipment as well as inlet and outlet streams for each one. To be able of comparing effects of applying other solvents and their blends, firstly, it is needed to define a feasible base case. The base case for this work is defined from (Øi L. E., Aspen HYSYS Simulation of CO2 Removal by Amine Absorption from a Gas Based Power Plant, 2007) work where an optimized process with 30% MEA solvent removes CO_2 from flus gas. 85% removal efficiency and minimum approach temperature of 10°C in lean/rich heat exchanger, also, have been assumed for the base case.

Carbon Dioxide removal processes have been simulated in Aspen HYSYS version 10 and 12 programs.

Table 3.1 provides specifications corresponding to base case.

Table 3.1: Specification to the base process of CO2 removal for simulation *(Øi L. E., Aspen HYSYS Simulation of CO2 Removal by Amine Absorption from a Gas Based Power Plant, 2007)*

Parameter	Value	Unit
Inlet flue gas temperature to process	40	℃
Inlet flue gas pressure to process	101.0	kPa
Inlet flue gas flow rate	1.091e5	$kgmol/h$
CO_2 content in inlet gas	3.30	$mol\%$
Water content in inlet gas	6.90	$mol\%$
Lean amine temperature before and after pump	120	℃
Amine pressure before rich pump	200	kPa
Amine pressure after rich pump	300	kPa
Lean amine pressure to absorber	101	kPa
Lean amine rate to absorber	1.175e5	$kgmol/h$
CO_2 content in lean amine	2.98	$mole\%$
Number of stages in absorber	10	-
Rich amine pressure before pump	110	kPa
Rich amine pressure after pump	200	kPa

Number of stages of stripper	6 + Reboiler + Condenser	-
Reboiler temperature	120	℃
Efficiency of stages in absorber	0.25	-
Efficiency of stages in stripper	1	-

Based on Table 3.1, efficiency of stages in absorber for base case was assumed to be 0.25, but the base case was simulated, also, with efficiency of stages as 0.17. This value for base case was a suggestion form Aspen HYSYS program while it can not be realistic.

In addition to base case with 30 wt% MEA as solvent, CO_2 removal process was simulated for MEA concentration of 35 wt%, 40 wt% and 45 wt%.

The Aspen HYSYS flowsheet for base case simulation is shown below.

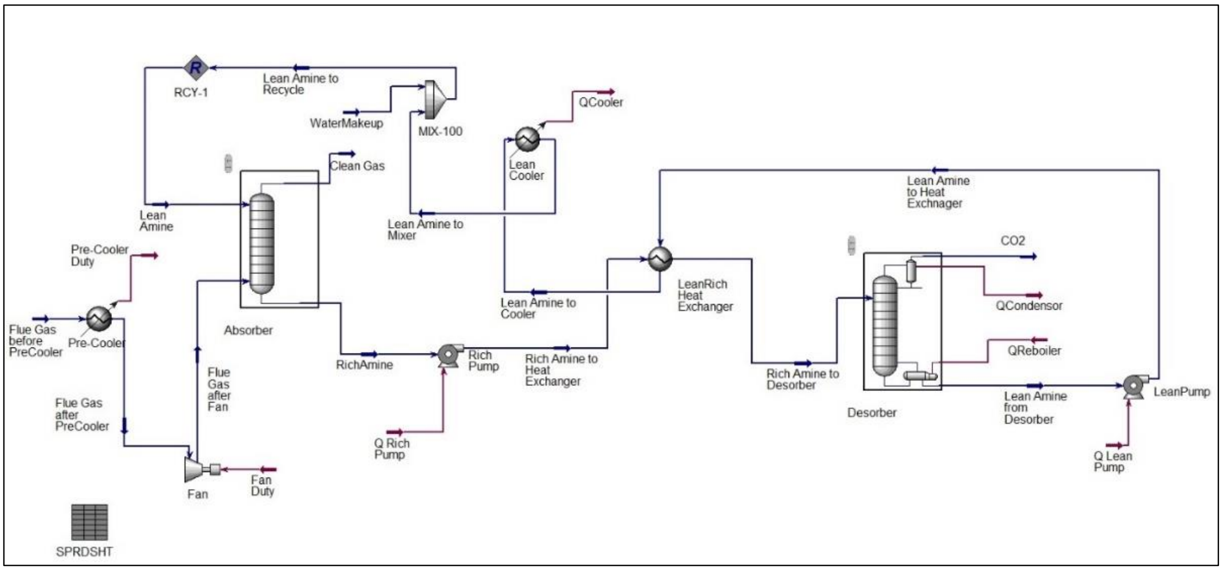

Figure 3.1: Aspen HYSYS flowsheet of standard CO_2 removal process

3.2. Specification to other solvents and blends

Other solvents, MDEA and piperazine (PZ), and their blends including MEA+MDEA and MEA+PZ have been applied to simulate the standard removal process shown in figure 2.2. In many cases, the same specifications of base case have been used for processes. The simulated cases are tabulated below. All cases have removal efficiency of 85% and 10℃ as minimum approach temperature difference.

All simulated standard removal processes are grouped in seven classes like

1. Standard removal process with 30 wt% amine blends of MEA and PZ
2. Standard removal process with 40 wt% amine blends of MEA and PZ
3. Standard removal process with 30 wt% amine blends of MEA and MDEA
4. Standard removal process with 40 wt% amine blends of MEA and MDEA
5. Standard removal process with 50 wt% amine blends of MEA and MDEA
6. Standard removal process with 40 wt% amine blends of MDEA and PZ
7. Standard removal process with 50 wt% amine blends of MDEA and PZ

Table 3.2: Simulated processes with other solvents/blends for standard process

30 wt% MEA+PZ	**40 wt% MEA+PZ**	**30 wt% MEA+MDEA**	**40 wt% MEA+MDEA**	**50 wt% MEA+MDEA**	**40 wt% MDEA+PZ**	**50 wt% MDEA+PZ**
30% MEA	40% MEA	30% MEA	40% MEA	50% MEA	40% MDEA	50% MDEA
27.5% MEA+2.5% PZ	35% MEA+5% PZ	25% MEA+5% MDEA	35% MEA+5% MDEA	45% MEA+5%MDEA	35% MDEA+5% PZ	45% MDEA+5%PZ
25% MEA+5% PZ	30% MEA+10% PZ	22.5% MEA+7.5% MDEA	30% MEA+10% MDEA	40%MEA+10%MDEA	30% MDEA+10% PZ	40%MDEA+10% PZ
22.5% MEA+7.5% PZ	25% MEA+15% PZ	20% MEA+10% MDEA	25% MEA+15% MDEA	35%MEA+15%MDEA	25% MDEA+15% PZ	35%MDEA+15% PZ
20% MEA+10% PZ	20% MEA+20% PZ	15% MEA+15% MDEA	20% MEA+20% MDEA	30%MEA+20%MDEA	20% MDEA+20% PZ	30%MDEA+20% PZ
15% MEA+15% PZ	15% MEA+25% PZ	10% MEA+20% MDEA	15% MEA+25% MDEA	25%MEA+25%MDEA	15% MDEA+25% PZ	25%MDEA+25% PZ
10% MEA+20% PZ	10% MEA+30% PZ	5% MEA+25% MDEA	10% MEA+30% MDEA	20%MEA+30%MDEA	10% MDEA+30% PZ	20%MDEA+30% PZ
5% MEA+25% PZ	5% MEA+35% PZ	30% MDEA	5% MEA+35% MDEA	15%MEA+35%MDEA	5% MDEA+35% PZ	15%MDEA+35% PZ
30% PZ	40% PZ		40% MDEA	10%MEA+40%MDEA	40% PZ	10%MDEA+40% PZ
				5%MEA+45%MDEA		5%MDEA+45%PZ
				50%MDEA		50%PZ

All simulated standard processes were listed above. The efficiency of stage in absorber for all cases was assumed to be 0.25. In addition to table 3.2, two other cases including blend of 30 wt% MEA+5 wt% PZ with the same specifications with tabulated cases and ME of 0.3 and 0.35 were simulated.

3.3. Specification to vapor recompression process

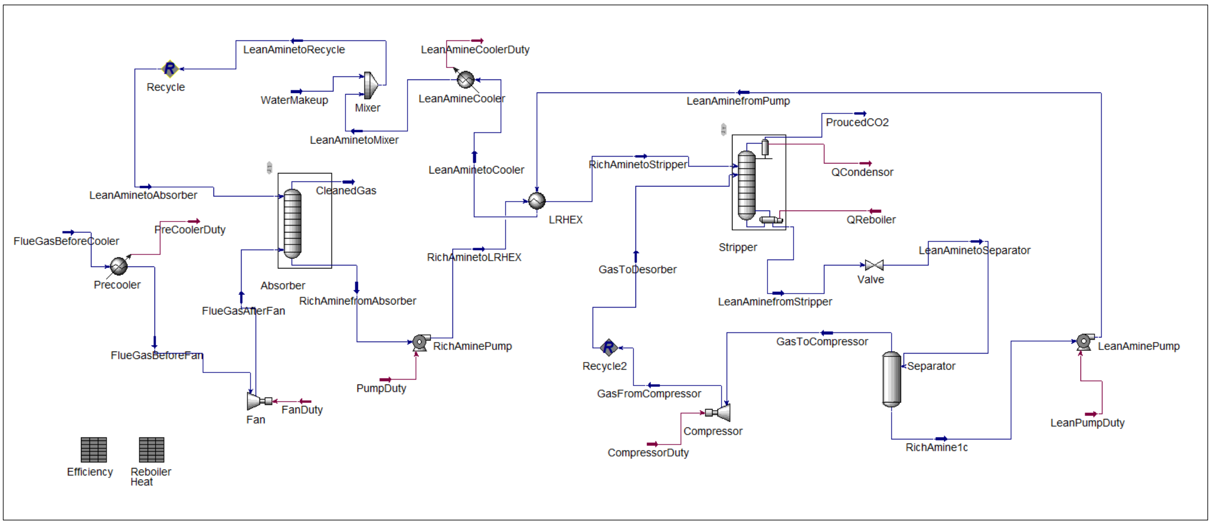

Figure 6: Aspen HYSYS flowsheet of vapor recompression CO_2 removal process

Vapor recompression process was simulated with 30 wt% MEA solvent. The removal efficiency for simulated case was assumed to be 85% as well as 10℃ lean minimum approach temperature. Also, the efficiency of stages in absorber was assumed to be 0.25.

Vapor recompression more and less has the same specifications of standards process. The main differences in this work refer to adding a valve, a separator and a compressor to standards process. Regarding differences in specifications of process, the valve reduces the pressure of outlet stream of stripper from 200 kPa to 100 kPa. This reduction in pressure is conducted by a linear valve with 50% opening. The separator is performing in 100 kPa. The liquified part of inlet stream into separator leaves from the bottom of separator item in 100 kPa pressure. From top of separator, vapor leaves in 100 kPa pressure. Pressure of liquid and gas will rise up to proper values with the aid of lean pump and compressor respectively. In fact, liquid requires to be reached to 300 kPa and vapor should be reached to 200 kPa.

In addition to above, vapor recompression process was simulated for amine blends of (30 wt% MEA+15 wt% MDEA) and (30 wt% MEA+5 wt% PZ). Bott mentioned blends have the same specifications to simulation above in terms of LMTD, removal efficiency and the efficiency of stages in absorber.

Also, vapor recompression for blend of 30 wt% MEA+15 wt% MDEA was simulated when the LMTD is 5℃ in lean rich heat exchanger.

3.4. Simulation results

All mentioned simulations in this chapter have been carried out in Aspen HYSYS version 10. The simulated cases have been investigated in terms of regeneration energy, lean amine loading, rich amine loading, cyclic loading. Extracted results will be explained in chapter 6 dedicated to results and discussion.

The main aim for large number of simulations in this work is to find those blends of amines providing lower regeneration energy in order to reduce total costs for removal plants as well as other improvements.

4. Dimensioning and equipment

This part, firstly, includes dimensioning where size and material for various pieces of equipment in the CO_2 removal process are explained. Also, the reasons for selecting each component of plant will be explained. The chapter, also, includes the relevant formula for each piece of equipment.

Extracted data from dimensioning provides required data for the economy of project. In other words, the data are used in Aspen In-Plant Cost Estimator to calculate the total cost for the simulated plant.

Each component in the plant will be briefly explained following.

4.1. Absorption column

Absorption as the most important process in CO_2 removal plant is defined as the transfer of species of gaseous phase to a liquid solvent. This process is, also, known as gas absorption, gas scrubbing or gas washing. This process is used to separate gas mixtures, recover chemicals or remove impurities (Seader & Henley, 2006).

Absorption bases on counter-current movement of flue gas and liquid solvent. In other words, flue gas including CO_2 enters the column from the bottom while solvent comes into the column from top. Inside the column, contact devices are installed to provide maximum surface area between the liquid solvent and the flue gas (Kallevik, 2010).

Different sorts of absorption column can be found. In this work packing type is used.

Required specifications for absorber in this project is presented in table 4.1 below.

Table 4.1: Specifications to absorption column

Parameter	Symbol	Unit	Specification	Note
Column Type	-	-	Packed	Specified
Gas velocity	v_{gas}	$[{}^{m}/_{s}]$	2.5	Specified
Actual Volume flow rate	$\dot{V}$	$[{}^{m^3}/_{s}]$	Extracted from simulation	Specified
Number of stages	N_{stages}	-	10	Specified
Height of Packing	$h_{packing}$	$[{}^{m}/_{stage}]$	1	Specified
Volume of packing	$V_{packing}$	$[m^3]$	Based on formula 4.2	Calculated
Column height	H	$[m]$	40	Estimated
Column Diameter	D	$[m]$	19.4	Calculated
Shell Material	-	-	SS316	Specified
Packing Type	-	-	Structured	Specified

Calculating of absorption column can be conducted through formula 4.1 below. In addition to the packing height, bottom liquid reservoir and water wash column should be accommodated in the column. That is why, (Kallevik, 2010) and (Øi L. E., Removal of CO2 from exhaust gas) suggested to add extra height to column to provide sufficient room for inlet, top and bottom outlets and auxiliary equipment.

Total number of equilibrium stages in the column defines the height for packing section. It should be noted that based on (Øi L. E., Removal of CO2 from exhaust gas) one stage is assumed equal to one meter of height. Thus, number of stages should be multiplied to one meter to have the total packing height.

(Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) (Øi L. E., Removal of CO2 from exhaust gas) suggested that inlet gas velocity to absorption column is 2.5 $[^{m}/_{s}]$. This assumption leads to calculating diameter of column by formula 4.1.

$$D = \sqrt{4 \cdot \frac{\dot{V}}{\pi \cdot v_{gas}}} \tag{4.1}$$

Where, $D\ (m)$ indicates calculated diameter of absorption column, $\dot{V}\ (^{m^3}/_{s})$ is actual gas volumetric flow rate to column and $v_{gas}\ (m/s)$ is the assumed inlet gas velocity to absorber.

In this work, the actual gas volumetric flow rate is highly big, leading into c.a. 20 meter for column diameter. So, the column is divided to smaller ones where have the same packing volume with the big one.

Total volume of packing is calculated through formula 4.2.

$$V_{packing} = \frac{\pi \cdot D^2}{4} \cdot N_{stages} \tag{4.2}$$

$V_{packing}\ (m^3)$ indicates the volume of packing in absorber, $D\ (m)$ is the calculated diameter and N_{stages} is the number of stages in column.

(Arachchige & Melaaen, 2012) explained that two different sorts of packing materials exist for gas absorption including random packing and structured packing which presented in figure 4.1 below. In large scale CO_2 absorption column, structured packing is better than random ones due to higher efficiency, higher capacity and lower pressure drop along column (Øi L. E., Removal of CO2 from exhaust gas) (Brickett, 2015). Based on figure 4.1, it is obvious that structured packing has potential to provide the removal process with larger contact area between gas and liquid, leading into bigger overall mass transfer coefficient.

Figure 4.1: Random (left) and Structured (right) packing *(Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020)*

4.2. Desorption column

Desorption column requires heat to recover CO_2 from solution. The recovered Carbon Dioxide leaves the column from top and regenerated amine comes back to cyclic removal process from bottom. The column is equipped with condenser at top and reboiler at bottom. The structure and function of this column has high similarity with absorption column. Dimensioning is, also, for this piece of equipment is calculated with formula 4.1 and 4.2. (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of

CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) and (Øi L. E., Removal of CO2 from exhaust gas) suggested 1.0 ($^{m}/_{s}$) is reasonable for inlet velocity of flow to column.

The column can be formed on specifications in table 4.2.

Table 4.2: Specifications to desorption column

Parameter	**Symbol**	**Unit**	**Specification**	**Note**
Column Type	-	-	Packed	Specified
Flow velocity to column	v_{gas}	$[^{m}/_{s}]$	1	Specified
Actual volume flow rate	$\dot{V}$	$[A\,^{m^3}/_{s}]$	Table K.2	HYSYS
Number of stages	N_{stages}	-	6 + reboiler + condenser	Specified
Height of packing	$h_{packing}$	$[^{m}/_{stage}]$	1	Specified
Volume of packing	$V_{packing}$	$[m^3]$	Based on formula 4.2	Calculated
Column height	H	$[m]$	20	Assumed
Column Diameter	D	$[m]$	Based on formula 4.1	Calculated
Shell Material	-	-	SS316	Specified
Packing Type	-	-	Structured	Specified

4.3. Lean / Rich Heat Exchanger

The purpose of using lean rich heat exchanger in removal process is to recover heat. In other words, the rich amine from the bottom of absorber is warmed up by a hot solvent, lean amine, leaving the sump of stripper. Increase of rich amine temperature reduces reboiler duty (Kallevik, 2010).

Shell and tube heat exchanger type as shown in figure 4.2 is used in this work to satisfy the required aims. This type of heat exchanger is the most robust (Aromada, Eldrup, Normann, & Øi, 2020).

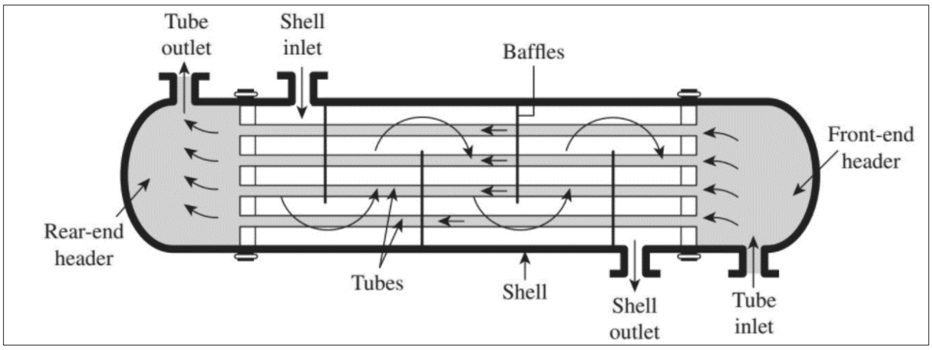

Figure 4.2: A diagram of shell and tube heat exchanger *(Cavallo , 2011) (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020)*

Lean rich heat exchanger is sized with formula 4.3. Other specifications for lean rich heat exchanger are listed in table 4.3 below.

$$A = \frac{\dot{Q}}{U \cdot \Delta T_{lm}} \tag{4.3}$$

where, $A\ (m^2)$ indicates the total area for the heat exchanger, $\dot{Q}\ (kW)$ is heat exchanger duty and $U\ (^{W}/_{m^2K})$ is overall heat transfer coefficient. Equation 4.4 calculates ΔT_{lm}, logarithmic mean temperature difference, as following.

$$\Delta T_{lm} = \frac{\Delta T_{out} - \Delta T_{in}}{\ln \frac{\Delta T_{out}}{\Delta T_{in}}} \tag{4.4}$$

where, ΔT_{lm} (°C) indicates logarithmic minimum approach temperature, ΔT_{out}(°C) is defined as difference in cold and hot inlet streams to heat exchanger, $(T_{hot,in} - T_{cold,in})$, and ΔT_{in} (°C) is the difference of hot and cold outlet streams , $(T_{hot,out} - T_{cold,out})$.

It should be noted that overall heat transfer coefficient, U, for lean rich heat exchanger is assumed to be 500 ($W/_{m^2K}$) (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

Relevant specifications for lean rich heat exchanger are listed in table 4.3 below.

Table 4.3: Specifications to lean rich heat exchanger

Parameter	Symbol	Unit	Specification	Note
Heat Exchanger Type	-	-	U-Tube Flow	Specified
TEMA type	-	-	BEU	Specified
Heat transfer rate / Duty	$\dot{Q}$	$[kW]$	Based on simulations	HYSYS
Overall heat transfer coefficient	U	$[W/_{m^2K}]$	500 (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)	Specified
Temperature of hot flow into the heat exchanger	$T_{hot,in}$	[°C]	Based on simulations	HYSYS
Temperature of hot flow out of heat exchanger	$T_{hot,out}$	[°C]	Based on simulations	HYSYS
Temperature of cold flow into heat exchanger	$T_{cold,in}$	[°C]	Based on simulations	HYSYS
Temperature of cold flow out of heat exchanger	$T_{cold,out}$	[°C]	Based on simulations	HYSYS
Logarithmic mean temperature difference	ΔT_{lm}	[°C]	Based on formula 4.4	HYSYS/Caculated
Total area required	A	$[m^2]$	Based on formula 4.3	Calculated
Shell material	-	-	SS316	Specified
Tube Material	-	-	316LW	Specified

In this work, the calculated total area for lean rich heat exchanger is highly big. That is why, it is divided to smaller ones. The assumption for this division is 1000 m^2 per shell as most common available in the market. So, the number for smaller heat exchangers are calculated.

Scaling factor of 0.65 to be applied in cost estimation of the total smaller heat exchanger for all cases.

4.4. Reboiler

Reboiler is a kind of heat exchanger placed in the bottom of desorption column to boil the liquid and generate vapor. Most reboilers are shell and tube heat exchangers and typically steam is used to provide required heat. Dimensioning for the reboiler is completely identical to lean rich heat exchanger. So, recent presented formulas, 4.3 and 4.4, calculate total area for reboiler.

Other specifications for the reboiler in the project is tabulated in 4.4 below.

Overall heat transfer coefficient, U, for Reboiler is assumed to be 800 ($^{W}/_{m^2K}$) (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

Table 4.4: Specification to reboiler

Parameter	Symbol	Unit	Specification	Note
Heat Exchanger type	-	-	U-Tube Flow	Specified
TEMA type	-	-	BEU	Specified
Heat transfer rate / Duty	$\dot{Q}$	$[kW]$	Based on simulations	HYSYS
Overall heat transfer coefficient	U	$[^{W}/_{m^2K}]$	800 (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)	Specified
Temperature of steam to heat exchanger	$T_{hot,in}$	[℃]	131	Assumed
Temperature of steam hot from heat exchanger	$T_{hot,out}$	[℃]	130	Assumed
Temperature of cold flow into heat exchanger	$T_{cold,in}$	[℃]	Based on simulations	HYSYS
Temperature of cold flow out of heat exchanger	$T_{cold,out}$	[℃]	Based on simulations	HYSYS
Logarithmic mean temperature difference	ΔT_{lm}	[℃]	Formula 4.4	Calculated
Total area required	A	$[m^2]$	Formula 4.3	Calculated
Shell material	-	-	SS316	Specified
Tube material	-	-	316LW	Specified

4.5. Condenser

Condenser is, also, a sort of heat exchanger in which gaseous phase of substance cools down and switch to liquid phase. This process of condensing needs to loss heat to another substance. That is why, cooling water is used to remove the heat from gaseous stream.

Regarding dimensioning of condenser, relevant formula for reboiler and lean rich heat exchanger are applicable. Thus, the required heat transfer area for condenser is calculated by formulas 4.3 and 4.4. Overall heat transfer coefficient, U, for condenser is assumed to be 1000 ($^{W}/_{m^2K}$) (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

Table 4.5 lists all parameters and specification for condenser.

Table 4.5: Specification to condenser

Parameter	Symbol	Unit	Specification	Note
Heat Exchanger type	-	-	U-Tube Flow	Specified
TEMA type	-	-	BEU	Specified
Heat transfer rate / Duty	$\dot{Q}$	$[kW]$	Based on simulations	HYSYS
Overall heat transfer coefficient	U	$[^{W}/_{m^2K}]$	1000 (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)	Specified

Temperature of inlet hot flow	$T_{hot,in}$	[℃]	-	HYSYS
Temperature of outlet hot flow	$T_{hot,out}$	[℃]	-	HYSYS
Temperature of inlet cooling water	$T_{cold,in}$	[℃]	8	Assumed
Temperature of outlet cooling water	$T_{cold,out}$	[℃]	23	Assumed
Logarithmic mean temperature difference	ΔT_{lm}	[℃]	-	Calculated
Total area required	A	$[m^2]$	-	Calculated
Shell material	-	-	SS316	Specified
Tube material	-	-	316LW	Specified

4.6. Lean amine cooler

Lean amine cooler is a type of heat exchanger which reduces the temperature of lean amine stream to an optimized value with the aid of using cooling water. The reason for adjusting lean amine temperature to optimized value is that this temperature is widely affecting efficiency of removal process. In many studies, inlet temperature of lean amine to absorber is assumed to be 40℃ as an optimized parameter. (Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007) investigated effect of lean amine temperature to absorber and suggested that there should be a trade-off between the required cooling water in lean amine cooler and regeneration energy in stripper. They simulated 90% an amine-based removal process with 30 wt% MEA as solvent and found that higher temperature of lean amine to absorber leads to higher regeneration energy for the process as well as lower cooling duty for lean amine cooler. On the other hand, lower temperature of lean amine to absorber results into lower regeneration energy but higher cooling duty. So, an optimized temperature will be needed.

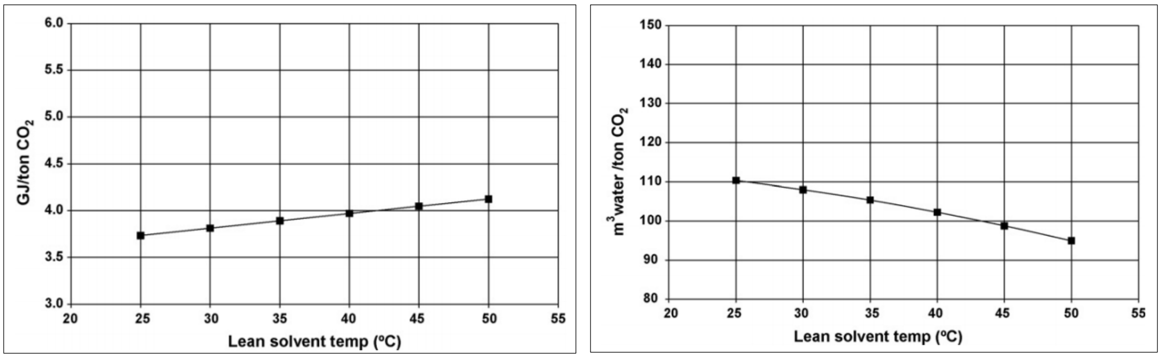

Figure 4.3: Assessment of lean amine temperature and regeneration energy (left) and cooling water (right) for 90% MEA-based removal process *(Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007)*

Total required area of lean amine cooler is calculated by formulas 4.3 and 4.4. Overall heat transfer coefficient for lean amine cooler is assumed to be 800 ($^{W}/_{m^2K}$) (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

Table 4.6 lists required parameters and specification of lean amine cooler for dimensioning.

Table 4.6: Specifications to lean amine cooler

Parameter	**Symbol**	**Unit**	**Specification**	**Note**
Item type	-	-	U-Tube Flow	Specified
TEMA type	-	-	BEU	Specified
Heat transfer duty	$\dot{Q}$	$[kW]$	Based on simulations	HYSYS
Overall heat transfer coefficient	U	$[^{W}/_{m^2K}]$	800 (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation	Specified

			of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)	
Temperature of inlet hot flow	$T_{hot,in}$	[℃]	Based on simulations	HYSYS
Temperature of outlet hot flow	$T_{hot,out}$	[℃]	40℃	Specified
Temperature of inlet cooling water	$T_{cold,in}$	[℃]	8	Specified
Temperature of outlet cooling water	$T_{cold,out}$	[℃]	23	Specified
Logarithmic mean temperature difference	ΔT_{lm}	[℃]	Formula 4.4	Calculated
Total area required	A	$[m^2]$	Formula 4.3	Calculated
Shell material	-	-	SS316	Specified
Tube material	-	-	316LW	Specified

4.7. Pumps

Pumps are devices that supply required energy to transfer fluids from one position to another one. Generally, pumps are classified endlessly based on size, type or applications.

Amine-based CO_2 capture process generally requires two pumps including rich pump and lean pump. Rich pump transfers rich amine flow from the sump of absorber to the top of stripper. Besides, lean pump supplies required energy to send lean amine flow from the bottom of desorption column to the top of absorption column. Totally, these pumps should supply energy to overcome different losses including:

- Friction losses in the pipes.
- Pressure loss in the lean rich heat exchanger placed between two columns.
- Pressure difference between absorption and desorption columns.
- Pressure loss in lean amine cooler.

Both rich and lean pumps in this work are assumed to be centrifugal ones with adiabatic efficiency of 75%.

Dimensioning of a pump can be conducted by required power, fluid head and volumetric flow rate of fluid which are extracted from simulations. Fluid head for both rich and lean pumps are assumed to be 50 (m) and 70 (m) respectively.

Table 4.7 below lists required parameters and specifications to rich and lean pumps.

Table 4.7: Specification to rich and lean pumps

Parameter	**Symbol**	**Unit**	**Specification**	**Note**
Item type	-	-	Centrifugal	Specified
Driver type	-	-	Motor	Specified
Driver power	P	$[kW]$	Based on simulations	HYSYS
Volumetric flow rate	$\dot{V}$	$[{}^{l}/_{s}]$	Based on simulations	HYSYS
Pump material	-	-	SS316	Specified
Pump adiabatic efficiency	-	%	75	Specified
Fluid head for lean amine	$\Delta H_{lean\ pump}$	$[m]$	70	Specified
Fluid head for rich amine	$\Delta H_{rich\ pump}$	$[m]$	50	Specified

4.8. Fan & Compressor

Fans and compressors are classified as turbomachinery equipment which increase driving force by adding energy to a fluid.

CO_2 removal process includes a fan to supply required driving force of flue gas to overcome pressure drop in absorption column. In this work, a centrifugal fan with adiabatic efficiency of 75% is assumed.

In addition of the fan, in vapor recompression process a compressor is, also, required to increase the pressure of flow by stripper pressure. In this work, stripper pressure is equal to 200 (kPa). This aim is satisfied with a centrifugal compressor with 75% adiabatic efficiency.

Relevant parameters and specifications for dimensioning of fan is listed below.

Table 4.8: Specification to fan

Parameter	Symbol	Unit	Specification	Note
Item type	-	-	Centrifugal	Specified
Driver type	-	-	Motor	Specified
Driver power	P	$[kW]$	Based on simulations	HYSYS
Actual volumetric flow rate	$\dot{V}$	$[{}^{m3}/_{hr}]$	Based on simulations	HYSYS
Material	-	-	CS	Specified
Adiabatic efficiency	-	%	75	Specified
Speed	-	$[rpm]$	1800	Specified

Table 4.9 lists the required specifications for compressor in vapor recompression process.

Table 4.9: Specification to compressor

Parameter	Symbol	Unit	Specification	Note
Item type	-	-	Centrifugal	Specified
Driver type	-	-	Motor	Specified
Driver power	P	$[kW]$	Based on simulations	HYSYS
Actual volumetric flow rate	$\dot{V}$	$[{}^{m3}/_{hr}]$	Based on simulations	HYSYS
Material	-	-	CS	Specified
Design pressure inlet	-	$[kPa]$	Based on simulations	HYSYS
Design pressure outlet	-	$[kPa]$	Based on simulations	HYSYS

In this work, high flow rate of flue gas results into a big fan in size. So, it is assumed that two fans are working to satisfy the required driving force for the process. Scaling factor of this division is assumed to be 1 (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020).

4.9. Separator

As it was mentioned earlier, vapor recompression process requires a separator item to separate the two-phase flow leaving desorption column (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). This aim is satisfied with using a vertical separator. The reason for selecting vertical one is that liquid is dominant phase of flow (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). Sounder – Browns (K_s) approach is applied for dimensioning of the separator as equation (Moshfeghian, 2015)

$$V_{Gmax} = K_s \sqrt{\left(\frac{\rho_L - \rho_G}{\rho_G}\right)} \tag{4.5}$$

where, V_{Gmax} $(\mathrm{m/s})$ indicates maximum allowable gas velocity, K_S $(\mathrm{m/s})$ defines as sizing parameter, ρ_L (kg/m^3) shows density of liquid phase and ρ_G (kg/m^3) is density for gaseous phase of flow.

Recommended value of K_s can be assumed to be 0.081 (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). Also, density for liquid and gas phases of stream are extracted from simulated processes. Thus, based on formula 4.5, V_{Gmax} is calculated, resulting into diameter of the separator from formula 4.6.

$$D_{min} = \sqrt{\frac{(4/\pi)\, q_a}{(F_g V_{Gmax})}} \tag{4.6}$$

where, D_{min} (m) is the minimum required diameter for separator, q_a $(\mathrm{m^3/s})$ is gas flow rate at the actual flowing condition and F_G is fraction of cross section area available for gas flow where for vertical separator this term is assumed to be 1 (Moshfeghian, 2015) (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). The height of separator can be calculated from defined ratio of (Moshfeghian, 2015)

$$L/_{D_{min}} = 2.5 \tag{4.7}$$

where L (m) is defined as height of the application.

All required specifications for separator are listed below.

Table 4.10: Specifications to separator item

Parameter	Symbol	Unit	Specification	Note
Separator orientation	-	-	Vertical	Specified
Diameter	D_{min}	$[m]$	Based on 4.6	Calculated
Length	L	$[m]$	Based on 4.7	Calculated
Material	-	-	SS316	Specified

4.10. Non-listed equipment

As this work aims to investigate a preliminary study of CO_2 removal process with other amines than MEA and their blends, only major equipment are listed. In other words, some items like direct contact cooler (DCC), mixer, control valve are not included in this work.

This work proceeds with cost estimation. Thus, for having more realistic ideas from the economy of processes, 20% of total cost for each process is added to total calculated cost to cover non-listed equipment.

5. Economy of project

This part of work, firstly, includes relevant explanations for cost estimation methods, CAPEX and OPEX contents. The chapter proceeds with conducted cost estimation for simulated processes. The economy of project is based on extracted data from Aspen In-Plant Cost Estimator version 10. This program provides cost for different pieces of equipment for year 2016.

5.1. OPEX and CAPEX

Basically, expenditures are divided in two different groups including Capital Expenditure (CAPEX) and Operational Expenditure (OPEX). In fact, CAPEX can be defined as one-time payments to obtain services, acquisitions of physical assets and goods. Companies consider CAPEX as a sort of important investment (Dikov, 2020) because they have a long-term impact on the business like production capacity. Another reason for their importance is their high initial costs.

(Kallevik, 2010) classified CAPEX into

- Acquisition of property
- Ground preparing
- Utility connections to near infrastructure
- Administrative buildings, rooms, offices

In this work, CAPEX excludes funds for ground and buildings.

On the other hand, OPEX, which stands for operating expenses or expenditure, refers to ongoing cost for running a product, business or system (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). Relevant funds for maintaining the systems, also, are classified in this type. As general examples of OPEX, workers' salaries and wages, taxes, raw material, spare items and rent cost can be mentioned (Kallevik, 2010).

5.2. CAPEX for current work

CAPEX calculations for this work starts with cost estimation of each item in the plant which is sized and specified in dimension part. There are different ways to estimate or obtain equipment cost in the plant. Quoted offer form vendor, budgeted prices, in-house data from other projects, commercial databases, books and Internet are different sources used in capital cost estimations (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

It is highly preferable and reliable to have access to latest data for each item in the plant from manufactures but in many cases is not possible. (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) suggested that in-house data from other project are good choices to analyse the economy of project.

As mentioned above, commercial databases are, also, available and can be used for cost estimation of equipment. CAPEX in this project is based on extracting cost of equipment from Aspen In-Plant Cost Estimator version 10. Provided costs of this version refers to year 2016.

Specifications to each item of the removal plant are important to obtain CAPEX because they provide required data for Aspen In-Plant Cost Estimator. The size of each item, the number of required items and material in the construction process which were defined in the previous chapter affect cost of each item. Choice of material is completely dependent on the operation conditions such as pressure, temperature, corrosion and sort of fluid (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

5.3. Enhanced detailed factor (EDF) method

Extracted data from Aspen In-Plant Cost Estimator indicate equipment costs but CAPEX is not only limited to this one. Other costs have to be included like direct cost, engineering costs, administration cost and those costs covering commissioning and contingency. Each mentioned term, also, includes smaller categories. For instance, only direct cost is divided to more detailed elements including erection, piping, electric, instrument, civil work, steel & concrete and insulation. These items are listed below.

Table 5.1: Main elements constituting installing factor *(Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)*

Direct costs	**Engineering costs**	**Administration costs**	**Other costs**
Equipment	Process	Procurement	Commissioning
Erection	Mechanical	Project control	Contingency
Piping	Piping	Site management	
Electric	Electric	Project management	
Instrument	Instrument		
Civil work	Civil		
Steel & concrete	Steel & concrete		
Insulation	Insulation		

Thus, extracted equipment costs from Aspen In-Plant Cost estimator requires to be adjusted by some coefficient to include all main items in table 5.1. Afterwards, the results indicate total installed cost for each piece of equipment.

5.4. Material factor

As it mentioned earlier, each item in the removal plant is specified to a sort of material. In this work, stainless steel and carbon steel are used. So, material factor should be implemented for each item. The reason for this adjustment is that the EDF table is formed on CS material.

Material factors for different categories are listed below.

Table 5.2: Material factor for different kinds of construction *(Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019)*

Sort of material	**Material factor**
Stainless steel (SS316) welded	1.75
Stainless steel (SS316) machined	1.30
Glass-reinforced plastic	1.0
Exotic materials	2.50

By a closer look to table 5.2 above, stainless steel is divided into two different categories including welded and machined. Those items in the CO_2 capture removal plant listed as welded stainless steel are (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020)

- Rich amine pump
- Lean amine pump

whereas, machined ones include:

- Absorber
- Stripper
- Lean Rich Heat Exchanger
- Lean Amine Cooler
- Condenser
- Reboiler

Specifying material and material factor to any piece of equipment are listed in the table below. This table is based on reference (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019).

Table 5.3: Specifications of material and material factor to equipment

List of equipment	Material	Material factor
Absorber	SS316	1.75
Stripper	SS316	1.75
Lean Rich Heat Exchanger	SS316	1.75
Lean Amine Cooler	SS316	1.75
Condenser	SS316	1.75
Reboiler	SS316	1.75
Fan	CS	1.0
Rich amine pump	SS316	1.3
Lean amine pump	SS316	1.3
Compressor	CS	1
Separator	SS316	1.75

Equipment cost for all items should be divided to material factor to adjust the material to EDF table. This table is attached to appendix. So,

$$equipment\ cost_{CS} = \frac{equipment\ cost_{cost}}{f_{mat}} \tag{5.1}$$

where, f_{mat} can be extracted from table 5.3. The resulted value, presently, can be used to find appropriate total installed cost coefficient from EDF table for that particular item. This coefficient is applicable for carbon steel material. So, the coefficient ought to be converted to a suitable case for SS material. This work is done through formula 5.2.

$$f_{t,SS} = f_{t,CS} + [f_{m,SS} - f_{m,CS}] + [f_{p,SS} - f_{p,CS}] \tag{5.2}$$

where, $f_{t,SS}$ is total cost factor of stainless steel, $f_{t,CS}$ indicates total cost factor of carbon steel, $f_{m,SS}$ shows material cost factor of stainless steel, $f_{m,CS}$ is material cost factor of carbon steel, $f_{p,SS}$ shows piping cost factor of stainless steel and $f_{p,CS}$ indicates piping cost factor of carbon steel.

By multiplying calculated factor to equipment cost from Aspen In-Plant Cost Estimator the total installed cost for each piece of equipment is obtained. Sum of CAPEX for each item results into total CAPEX.

5.5. Chemical Engineering Plant Cost Index

Provided equipment costs in part 5.4 refer to year 2016, while the current project is conducting in 2021. Thus, a cost adjustment should be done from one period to another due to changes in value of money. This aim is satisfied with chemical engineering plant cost index, CEPCI, which employed to update the capital cost of a chemical plant from a past time slot to later one.

Since Aspen In-Plant Cost Estimator version 10 provides data for 2016 and current project is conducting in 2021, CEPCI for these two years should be applied to update CAPEX.

Table 5.4: Chemical Engineering Plant Cost Index *(Chemical Engineering, n.d.)*

Year	Chemical Engineering Plant Cost index (CEPCI)
2016	542
2021	655.7

This adjustment is done with

$$Cost_{\ 2021} = Cost_{2016} \times \frac{CEPCI_{2021}}{CEPCI_{2016}} \tag{5.3}$$

5.6. Power law

Power law is a functional relationship where calculates a quantity based on a proportional relationship with other quantity. This relationship can be formulated as 5.4 where $Cost_1$ is the calculated cost for equipment with capacity Q_1, $Cost_2$ is the intended cost for equipment with capacity Q_2, Q_1 is the capacity of item 1, Q_2 is the capacity for item 2, And e is called scaling constant or cost exponent (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020).

As more explanation, if cost estimation of an absorber column with capacity Q_1 is $Cost_1$, this cost for the similar absorber colum with different capacity, Q_2, will be related in the way like:

$$Cost_2 = Cost_1 \left(\frac{Q_2}{Q_1}\right)^e \tag{5.4}$$

Scaling constant normally has a valve of $0.4 < e < 0.9$, but assumed averagely 0.65.

5.7. OPEX for current work

As it mentioned in chapter 5.1, CAPEX might include different items. In this work, OPEX is limited to steam, electricity, cooling water utilities as well as raw material. Raw material in this work is the required solvents or their blends.

Steam utility in this project should be supplied to satisfy the aim for reboiler. Additionally, cooling water utility fulfils the requirements in condenser and lean amine cooler items. Lastly, electricity is used for lean and rich pumps, fan and compressor.

Before conducting OPEX calculations for simulated processes, some relevant parameters are needed to be explained.

5.7.1. Plant lifetime

Time is an important parameter for each project. So, it should be considered for analysing the economy of project. Based on some works, lifetime for CO_2 capture plants can be assumed 20 years. In some other works, other suggestions have been given. For instance, (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) and (Aromada, Eldrup, Normann, & Øi, 2020) assumed 25 years as plant lifetime, whereas (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020), (Razi, Svendsen, & Bolland, 2013) and (Kallevik, 2010) economically analysed Carbon Dioxide removal plants for 20 years.

In this work, 20 years is assumed for lifetime of project.

5.7.2. Discount rate

Value of money during time is not constant. In addition, Carbon Dioxide removal plants are being designed for a considerable time slot. So, it is reasonable to include changes in value of money during the project lifetime.

The range for interest rate is 7%-14%. In this work this parameter is assumed to be 7.5 % (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) (Aromada, Eldrup, Normann, & Øi, 2020) which is assumed to be equal for all time interval of the project. Other assumption can be found in other work. For instance, discount rate per year in (Kallevik, 2010) work is assumed to be 7%.

5.7.3. Maintenance cost

Maintenance costs refer to any cost incurred by an individual or business to keep the assets in good working conditions. Example for this sort of cost is to spend money for repairing a machinery.

This cost, generally, is assumed as a constant percentage of the total installed cost (total CAPEX). This factor varies from 2% to 5%. (Hasan, Baliban, Elia, & Floudas, 2012) assumed this factor as 5% of total CAPEX in their work, while 4% is suggested based on (Aromada, Eldrup, Normann, & Øi, 2020) and (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019). This work assumes 4% to calculate OPEX cost for removal processes (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020).

5.7.4. Utilities cost

As it mentioned before, some pieces if equipment in the CO_2 removal plants require utilities. Used utilities in this work are included in 3 groups, electricity for pumps and compressor, cooling water for condenser and lean amine cooler and steam as the most important one in reboiler. Also, it was mentioned utilities are classified in OPEX calculation because they are used continuously.

Prices for mentioned utilities vary across studies. (Kallevik, 2010) assumed the cost of electricity equals 0.4 (NOK/kWh), 25% this value for the steam, 0.1 (NOK/kWh) and 0.033 (NOK/m^3) for cooling water. Prices for utilities in this work are listed in table 5.5.

Table 5.5: Assumed prices for the utilities project based on Euro *(Aromada, Eldrup, Normann, & Øi, 2020)*

Utility	Value	Unit
Electricity	0.132	[€/kWh]
Steam	0.032	[€/kWh]
Cooling water	0.022	[€/m^3]

5.7.5. Annual hours of operation

Annual hours of operation mean that the total number of hours, the whole plant is used for all commercial purposes.

This time varies in different studies based on capture unit. (Normann, Skagestad, Bierman, Wolf, & Mathisen, 2018) tabulated this parameter for steel industry as 8322 [$hours/year$], for pulp as 7840 [$hours/year$], for cement 7320 [$hours/year$] and for silicon 8760 [$hours/year$]. Also, based on steam generation this parameter is assumed to be 8000 [$hours/year$]. (Khan, et al., 2020) assumed this term 7450 hours per year.

In the project, it is assumed to be 8000 hours in each year (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorbtion Plants for CO2 Mitigation - Method and Assumptions, 2019) (Fagerheim, 2019) (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020).

5.7.6. Cost of solvent

As it was mentioned earlier in this work, three different solvent have been used including MEA, MDEA and PZ. MEA and DEA are grouped as the least expensive amines. The price for some amines are listed below.

Table 5.6: Price of amines *(Gomas & Santos, 2015)*

Amine	**Value**	**Unit**
Diethylamine	66.40	[$€/litre$]
MEA	30.50	[$€/litre$]
PZ	68.70	[$€/litre$]
EDA	31.70	[$€/litre$]
MDEA	51.60	[$€/litre$]
DEA	25.70	[$€/litre$]

In this work, the price for MEA is extracted from (Aromada, Eldrup, Normann, & Øi, 2020) work, 2069 [$€/m^3$].

Cost for MDEA and PZ is calculated proportionally from table 5.6 which lead to 4660 [$€/m^3$] for PZ and 3494 [$€/m^3$] for MDEA.

5.7.7. Location

Rotterdam is set as default place for cost analysis in Aspen In-Plant Cost Estimator program. Due to differences in working conditions from place to place like wages for workers, price for land, insurances, taxes and similar parameters, construction of CO_2 capture plants in other places than Rotterdam requires implementing location factor.

In this work, the place is kept as default. So, the location factor based on similar work of (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020) is assumed to be 1.

5.8. Cost estimation for simulated cases

Regarding CAPEX, each simulated case is, firstly, specified in term of dimension, material and number based on explanations in chapter 4. Also, Aspen In-Plant Cost Estimation version 10 was used to obtain equipment cost for Carbon Dioxide removal plant.

Each process, also, was adjusted in term of installing cost factor based on EDF method and specified material. The total CAPEX resulted from the summation of all pieces of equipment in the plant.

The calculation of OPEX as it mentioned requires considering different parameters. A brief explanation for important ones was given in part 5.7. As summery, explained parameters are listed in table 5.7 below.

Table 5.7: Important parameters for OPEX calculations

Parameter	Value	Unit
Plant lifetime	20	[year]
Discount rate	7.5%	-
Maintenance cost	4% of total CAPEX	-
Electricity cost	0.132	[€/kWh]
Steam cost	0.032	[€/kWh]
Cooling water cost	0.022	[€/m^3]
MEA cost	2069	[€/m^3]
MDEA cost	3494	[€/m^3]
PZ cost	4660	[€/m^3]
Annual operational time	8000	[hours/year]
Location factor	1	(same as Rotterdam)
CEPCI in 2016	542	-
CEPCI in 2020	655.7	-
Scaling constant	0.65	-

As OPEX is day to day costs, it should be calculated for a whole year (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020). Each variable operating cost is estimated using equation 5.5 (Aromada, Eldrup, Normann, & Øi, 2020)

$$Variable\ operating\ cost\ (€/year) = consumption\ (\text{unit}/hour) \times unit\ price\ (€/unit) \times opeational\ hours\ (hour/year) \tag{5.5}$$

The calculated CAPEX is for lifetime of each process. So, it should be calculated for each year. This can be done by implementing annualized factor (Aromada, Eldrup, Normann, & Øi, 2020)

$$Annualized\ factor = \sum_1^n \frac{1}{(1+i)^n} \tag{5.6}$$

where, i indicates the interest rate and n is plant lifetime.

Interest rate for this work is assumed to be constant for all calculation years, 7.5%. Also, plant lifetime is assumed to be 20 years.

It should be mentioned that in some work a minor part of plant lifetime is dedicated to construction time. For instance, (Aromada, Eldrup, Normann, & Øi, 2020) assumed 25 years as total plant lifetime where 2 years is dedicated for construction. This assumption in this work is not included.

Implementation of annualized factor to total CAPEX leads into annualized CAPEX.

$$annualized\ CAPEX = \frac{Total\ CAPEX}{annualized\ factor} \tag{5.7}$$

5.9. Analysis of economy for current work

The economy of each simulated process is estimated with total annual cost where

$$total\ annual\ cost\ (€/year)\ =\ annualized\ CAPEX\ (€/year) + annual\ OPEX\ (€/year) \tag{5.8}$$

For current work, base case is assumed as a reference to be able to compare the total annual cost for other process where MDEA and PZ or their blends are used as solvents.

Another parameter can be considered to compare different processes from the economy view is CO_2 capturing cost which is defined as

$$Annual\ cost\ of\ captured\ CO_2\ =\ \frac{total\ annual\ cost\ (€/year)}{mass\ flow\ rate\ of\ captured\ CO_2\ (kg/year)} \tag{5.9}$$

Both defined parameters in 5.8 and 5.9 are used to recognize economical improvements due applying solvents or their blends rather than base case where 30 wt% MEA is used as solvent.

As it mentioned earlier, the amine-based CO_2 removal process is energy-intensive process which uses huge amount of energy in reboiler as regeneration energy. So, this parameter can be obtained from a simulated process (Orangi , Farsi Madan, Fajferek, Sæter, & Bahri, 2020)

$$regeneration\ energy = \frac{applied\ steam\ in\ process\ [\frac{MJ}{h}]}{Mass\ flow\ rate\ of\ caprtured\ CO_2\ [\frac{kg}{h}]} \qquad [\frac{MJ}{kg}] \tag{5.10}$$

This parameter is, also, calculated for all simulated processes to find out the effect of other solvents than base case and their blends in this term.

5.10. Tips for cost estimation in Aspen In-Plant Cost Estimator

Analysing the economy of the current work has been faced some difficulties. Here, those problems will be listed as well as some practical adjustments for them.

- Due to high value for the flue gas, the calculated diameter of absorber was large in size, around 20 meters. Having such huge absorber in size to satisfy the aim of process imposes a considerable cost to plant. So, it is reasonable to have smaller items whit the same packing volume. In this work, the big absorber is divided to 16 smaller ones.
- Lean rich heat exchanger as one of the most important equipment in plant requires large contact area to satisfy the aims for plant. This big area can not be provided with one heat exchanger, so, smaller ones should be considered. Relevant assumption is 1000 m^2 per shell. So, large area for lean rich heat exchanger is divided into smaller ones where all of them are identical.
- As mentioned earlier, flow rate of the flue gas in this work is large. So, assumed fan for the process requires to be powerful enough, consequently large in size to increase driving force to desired value. The sized fan for this work leads into "out of range size" error for cost estimation. This error can be solved by assuming two smaller and identical fans working parallelly.
- In many studies, generally one case is analysed in term of economy and other cases are calculated based on power law. This process might be highly simpler but imposes higher uncertainties to results. In this work, all simulated cases are estimated completely for all applied equipment in the plant. Such procedure takes more time but it is more precise.
- Economy of present work is limited to main pieces of equipment in the plant. Some other items can be considered in CAPEX to be closer to reality. Also, about the OPEX, human resources like workers and engineers should be added to calculations.

6. Results and discussion

This part of work presents results as well as discussion about them. The chapter proceeds with uncertainties for the work. the chapter ends up with some search opportunity for future in line with the topic.

6.1. Discussion about regeneration energy

Numerous works can be found where the effect of other solvents than MEA and their blends have been investigated for the removal process, while a limited number of studies are available which include the economy of a total removal plant with other solvents or their blends. (Khan, et al., 2020) investigated the effect of MDEA+PZ blend has the potential of total cost saving of 0.67 M$ per year.

Finding an optimal concentration of blends for a solvent is not easy because many governing parameters should be taken into account. Since the main aim of this work is to follow cost estimation method, only some important parameters have been assumed including lean and rich amine loadings, regeneration energy and cyclic loading.

Since amine-based CO_2 removal process is energy-intensive, regeneration energy, as it mentioned earlier, is one of the most influential parameters in OPEX and consequently the economy of project. Applying other solvents than conventional one, MEA, can improve both conventional and vapor recompression CO_2 removal processes in term of regeneration energy and consequently the economy of removal plants.

Various concentrations for different blends including (MEA+MDEA), (MDEA+PZ) and (MEA+PZ) in this work have been tested with standard process. These simulations are tabulated in detail at table 3.2. To being able to recognize the improvement in regeneration energy, a base case is defined. The base case and its specifications have been explained before. Base case in this work requires 3.75 [MJ/ kg absorbed CO_2].

Based on conducted simulations, adding piperazine to MEA to have a blend of MEA and piperazine can present some improvements for the Carbon Dioxide removal plants. This resulted from simulation of plant with different concentration of PZ as additive to 30 wt% MEA to make a blend of MEA and PZ. In fact, different concentration of piperazine is added to 30 wt% MEA to calculate required energy in reboiler in a converged process. Investigations indicate that adding 5-10 wt% piperazine can provide a process with improvement in regeneration energy variable. So, the blend of 30 wt% MEA + 5-10 wt% PZ enhances the standard process in term of regeneration energy. Other investigation is to reach that percentage of piperazine which offers lowest required energy for reboiler. Among simulated cases, adding 5 wt% piperazine can be suggested as the best one. Thus, the process where 30 wt% MEA + 5 wt% PZ is used as solvent has the lowest regeneration energy among other ones. Regeneration energy for this process is 4.5% lower than the base case.

In addition, 5 wt% piperazine is used as additive to 30 wt%, 45 wt% and 40 wt% MEA to assess the effect of that for other concentration of MEA. The results show that this amount, 5 wt% PZ, in all blends cases presents lower regeneration energy comparing with having individual MEA.

It is noted that efficiency of stages in absorber for all simulated processes above was assumed to be 0.25.

Results for all simulations for the blend of MEA+PZ are listed below.

Table 6.1: Investigation of regeneration energy for adding 5 wt% to different concentration of MEA

Concentration [wt%]	**Regeneration energy [MJ/ kg CO_2]**	**improvement compared to base case [%]**	**improvement compared to individual MEA [%]**
30% MEA (base case)	3.75	-	-
25% MEA + 5% PZ	3.67	2.2%	2.2%
40% MEA	3.60	4.2%	-
35% MEA + 5% PZ	3.56	5.3%	1.2%
45% MEA	3.58	4.8%	-
40% MEA + 5% PZ	3.55	5.6%	0.9%

In addition to this, the plant was simulated with the blend of 30 wt% MEA+5 wt% PZ when the efficiency of stages in absorber is 0.30 and 0.35. For both simulations, required energy in reboiler was 3.47 and 3.43 [MJ/kg CO_2] which are 8.1% and 9.3% lower than base case.

Furthermore, the standard process has been tested for 10 wt% PZ as additive to MEA. The results show that adding 10 wt% piperazine to MEA can not present a CO_2 removal process with lower regeneration energy comparing with the same case where individual MEA acts as solvent.

The results of these simulations are listed below.

Table 6.2: Investigation of regeneration energy for adding 10 wt% piperazine to MEA

Concentration [wt%]	**Regeneration energy [MJ/ kg CO_2]**
30% MEA (base case)	3.75
20% MEA +10% PZ	3.78
40% MEA	3.60
30% MEA + 10% PZ	3.70
45% MEA	3.58
35% MEA + 10% PZ	3.82

From table 6.2, it is obvious that 10 wt% PZ can not improve the standard process in term of regeneration energy. It might be questioned that changes for this addition are negligible. It can be explained from cost view where the price of PZ is approximately 2.3 times more than MEA. So, although the regeneration energy for blends in table 6.2 are close but from the cost view, processes with the blend of MEA+10 wt% PZ imposes large costs to the economy of process.

Some literature investigated other improvements of adding PZ to have blend of MEA+PZ but these are out of scope of this work. For instance, (Rochelle, et al., 2011) mentioned advantages of PZ rather than MEA as more resistant to oxidative degradation, less volatility than MEA and no corrosive to stainless steel.

All simulated cases with the blend of MEA+PZ are shown in figure 6.1.

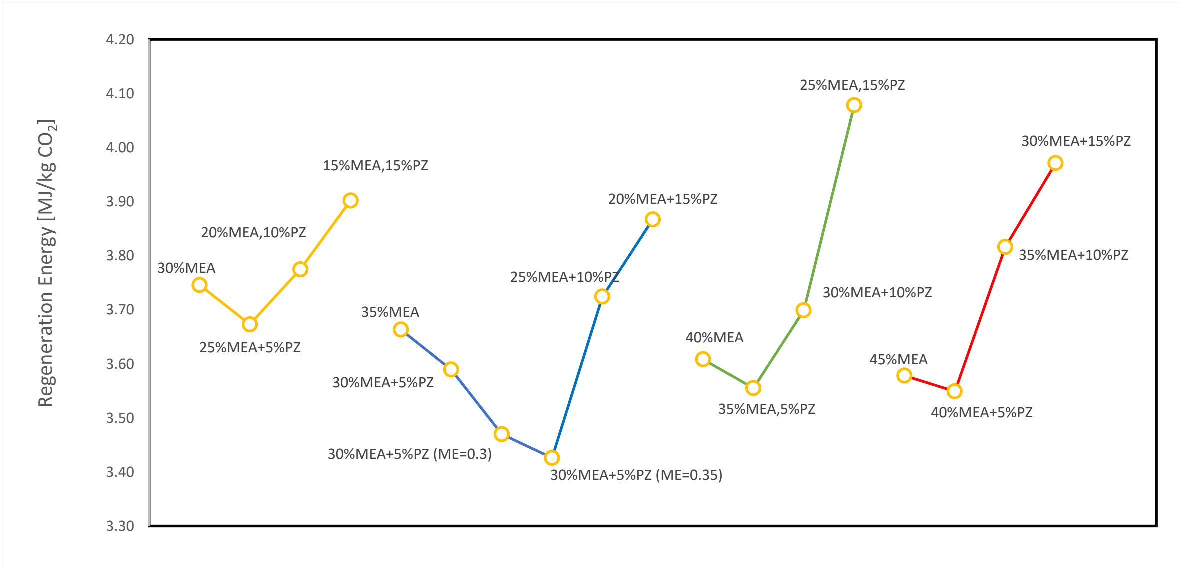

Figure 6.1: Investigation of adding piperazine to different concentrations of MEA for regeneration energy

Regarding the blend of MEA+MDEA, adding 5-20 wt% MDEA to base case (30 wt% MEA) enhances CO_2 removal process in term of regeneration energy. The results for these simulations are listed below.

Table 6.3: Investigation of regeneration energy for adding MDEA to 30 wt% MEA

Concentration [wt%]	**Regeneration energy [MJ/ kg CO_2]**	**improvement compared to base case [%]**
30% MEA (base case)	3.75	-
30% MEA + 5% MDEA	3.61	3.9%
30% MEA + 10% MDEA	3.52	6.5%
30% MEA + 15% MDEA	3.49	7.4%

As it is obvious from table 6.3, the blend of 30 wt% MEA+ 15 wt% MDEA has the lowest value for regeneration energy which is 7.4% lower than base case. (Li, Wang, & Chen, 2013) compared the MEA with blend of MEA+MDEA experimentally from different view. They suggest that this blend can be replaced with individual MEA due to fall in regeneration energy.

Some simulated processes with the blend of MEA+MDEA are depicted in figure 6.2 below.

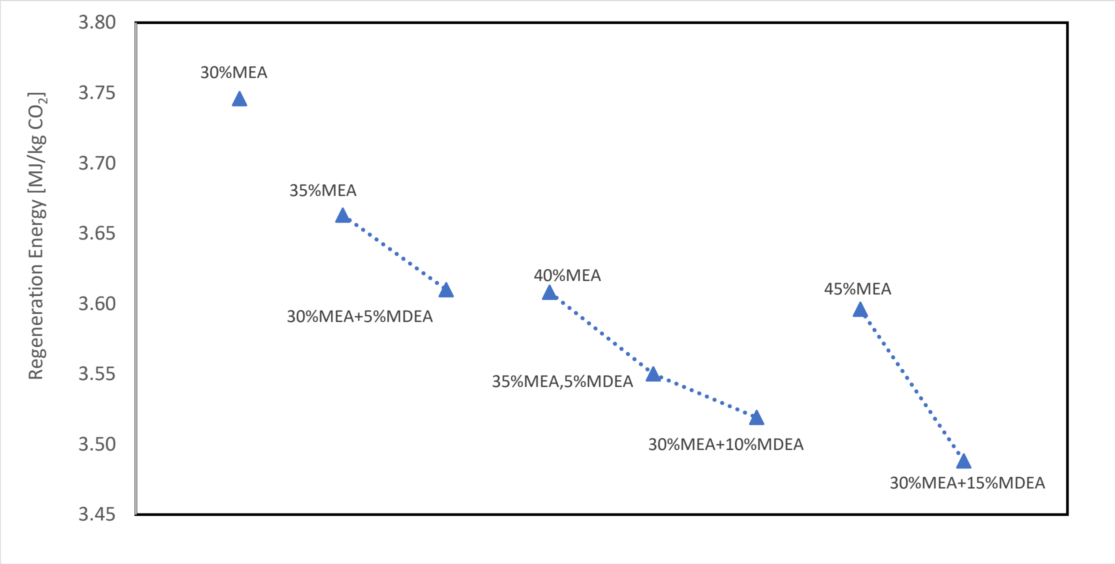

Figure 6.2: Investigation of adding MDEA to different concentrations of MEA for regeneration energy

Regarding the blend of MDEA and PZ, different portions of MDEA and PZ have been tested to assess a blend of MDEA+PZ in term of regeneration energy. According to the simulations, three blends of (20 wt% MDEA+30 wt% PZ), (15 wt% MDEA+35 wt% PZ) and (20 wt% MDEA+40 wt% PZ) requires lower energy for regenerating amine in the stripper comparing with the base case. Among three mentioned cases, the blend of 10 wt% MDEA+40 wt% PZ presents the lowest value, 3.59 [MJ/kg CO_2], which is 4.4.% improvement.

(Dubois & Thomas, Comparison of varipous configurations of the absorbtion-regereration process using different solvents for the post-combustion CO2 capture applied to cement plant flue gas, 2018) assessed the blend of 10 wt% MDEA+ 30 wt% PZ. They suggested that a blend with 5-10 wt% MDEA and 30-35 wt% PZ has potential to reduce regeneration energy than base case. Required energy for mentioned blends of MDEA+PZ are listed below.

Table 6.4: Investigation of MDEA+PZ blend for regeneration energy of standard process compared to individual solvent

Concentration [wt%]	**Regeneration energy [MJ/ kg CO_2]**	**improvement compared to base case [%]**
30% MEA (base case)	3.75	-
20% MDEA + 30% PZ	3.69	1.6%
15% MDEA + 35% PZ	3.65	2.7%
10% MDEA + 40% PZ	3.59	4.6%

Figure below depicts mentioned improvements for regeneration energy.

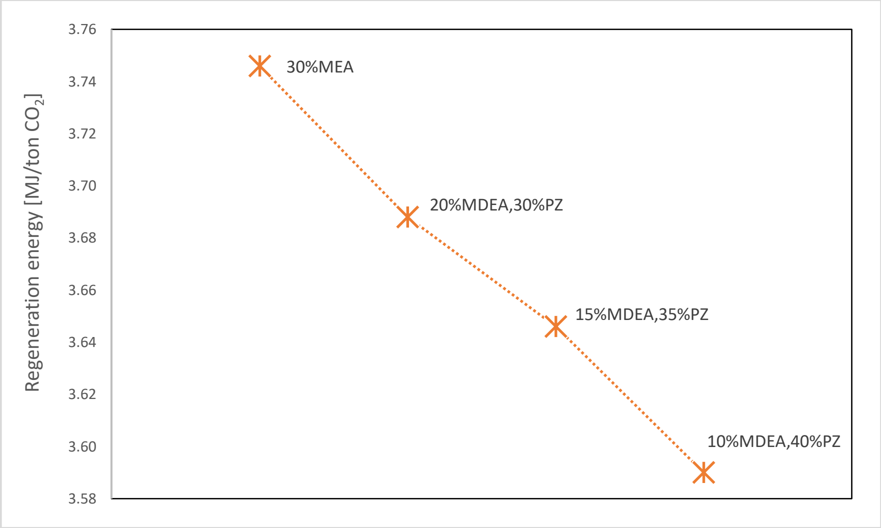

Figure 6.3: Investigation of MDEA+PZ blend in term of regeneration energy for standard process comparing with MEA

The work proceeds with simulation of vapor recompression (VR) process in which 3.14 [MJ/kg CO_2] requires to satisfy 85% removal efficiency with 30 wt% MEA as solvent to absorber. Thus, there is 19.4% fall in regeneration energy compared to base case.

Vapor recompression process, also, has been tested with two blends of (30 wt% MEA+15 wt% MDEA) and (30 wt% MEA+5 wt% PZ). These blends improved the process in term of energy 5.7% and 1.7% respectively compared to VR process with 30 wt% MEA. VR process with mentioned blends, also, enhanced regeneration energy 26.8% and 21.4%.

The table below includes regeneration energy of vapor recompression process for two blends of MEA+PZ and MDEA+MEA.

Table 6.5: Assessment of MDEA+MEA and MEA+PZ blends for regeneration energy in VR process

Concentration [wt%]	**Regeneration energy [MJ/ kg CO_2]**	**improvement compared to standard base case [%]**
30% MEA (base case)	3.75	-
30% MEA (VR)	3.14	19.4%
30% MEA + 5% PZ	3.09	21.4%
30% MEA + 15% MDEA	2.97	26.8%

Figure below compares data from table 6.5.

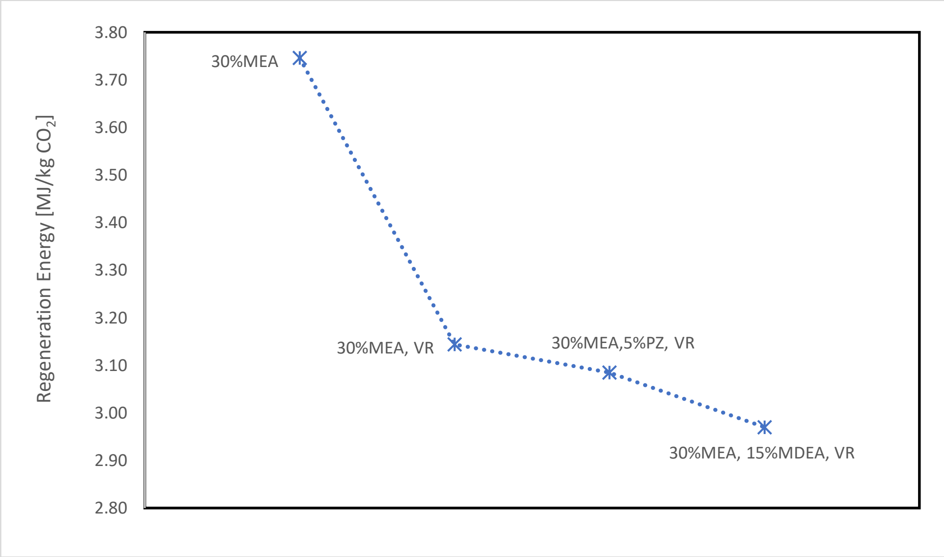

Figure 6.4: Investigation of MDEA+PZ and MEA+MDEA blends in term of regeneration energy for VR process comparing with 30 wt% MEA

6.2. Discussion about lean, rich and cyclic loading

Lean, rich and cyclic loading for MDEA and PZ and blends of MEA, MDEA and PZ are different with individual MEA. Also, different concentration of each amine in the blends can vary rich, lean and cyclic loadings. The differences should be investigated from reaction mechanism view.

Lean loading is one the governing parameters to reduce the regeneration energy (Abu-Zahra, Schneiders, Niederer, Feron, & Versteeg, 2007).

Cyclic capacity, difference of lean and rich loadings, is highly influential on regeneration energy. That is why, this variable is investigated in this work.

Lean, rich and cyclic capacities for standard base case are listed below. As previous section, base case is defined to have a reference for comparing other cases with that.

Table 6.6: Lean, rich and cyclic loading of standard base case process

Parameter	**Loading [mol CO_2/mol amine]**
lean loading	0.267
rich loading	0.495
cyclic loading	0.228

Regarding the blend of MEA+PZ, lean, rich and cyclic loading have been measured from simulations. Table below includes these data for MEA+PZ blend.

Table 6.7: Lean, rich and cyclic loading for MEA+PZ blend for standard process

Solvent/blend (wt%)	**Lean loading [mol CO_2/mol amine]**	**Rich loading [mol CO_2/mol amine]**	**Cyclic loading [mol CO_2/mol amine]**
30% MEA (Base case)	0.267	0.495	0.228
25% MEA+5%PZ	0.252	0.501	0.249
35%MEA	0.297	0.494	0.197

30%MEA+5%PZ	0.269	0.494	0.225
30%MEA+5%PZ (ME = 0.3)	0.278	0.509	0.231
30%MEA+5%PZ (ME = 0.35)	0.276	0.514	0.238
40% MEA	0.323	0.493	0.170
35% MEA+5%PZ	0.298	0.488	0.19
45%MEA	0.350	0.491	0.141
40%MEA+5%PZ	0.323	0.483	0.160

Based on table 6.7, it is obvious that adding 5 wt% piperazine to MEA rises cyclic loading. This change mainly results from reduction in lean loading. These data are depicted in figure 6.5.

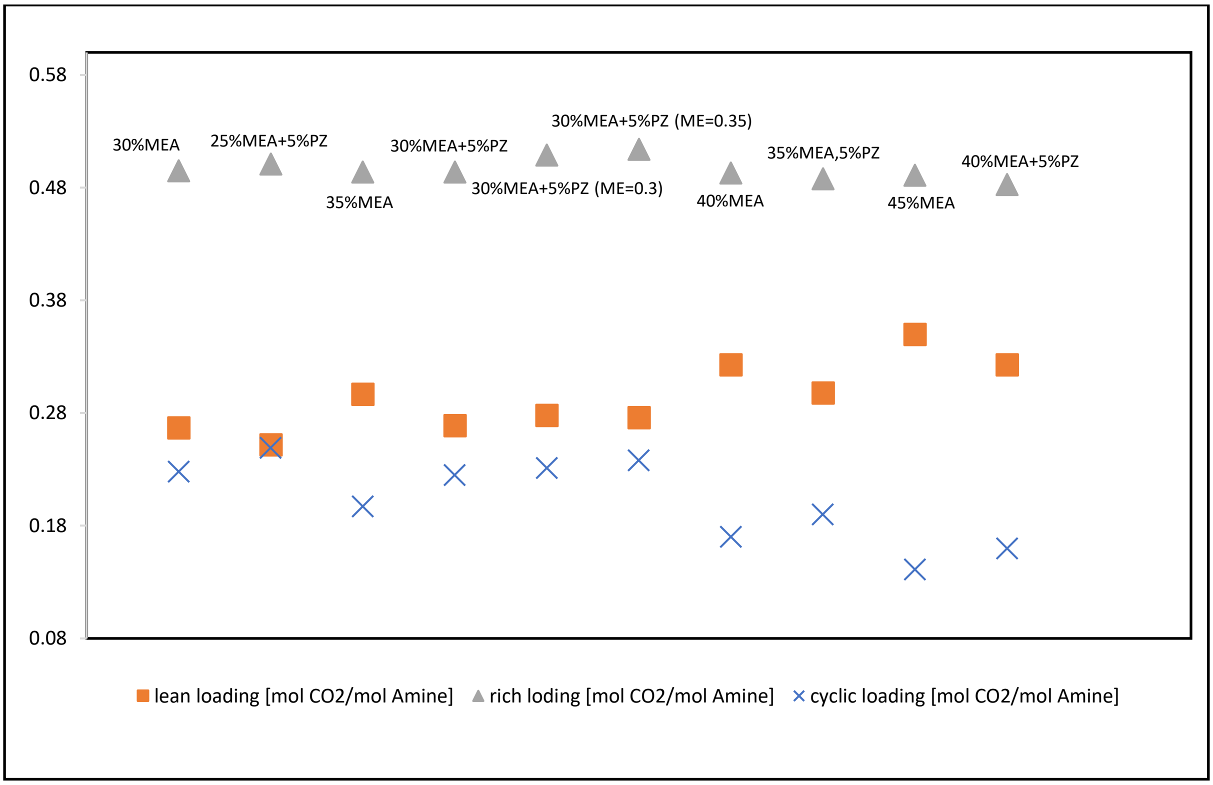

Figure 6.5: Investigation of lean, rich and cyclic loading for different concentration of MEA+PZ blend (standard process) compared to base case

The blend of MEA+MDEA, also, has been investigated for lean, rich and cyclic loading for the standard process.

Presented table below includes these data.

Table 6.8: Investigation of lean, rich and cyclic loading for MDEA+MEA blend in standard process compared to base case

Solvent/blend (wt%)	**Lean loading [mol CO_2/mol amine]**	**Rich loading [mol CO_2/mol amine]**	**Cyclic loading [mol CO_2/mol amine]**
30% MEA (Base case)	0.267	0.495	0.228
35% MEA	0.297	0.494	0.197
30% MEA+5% MDEA	0.262	0.479	0.217

40% MEA	0.323	0.493	0.170
35% MEA+5% MDEA	0.294	0.481	0.187
30% MEA+10% MDEA	0.257	0.459	0.202
45% MEA	0.350	0.491	0.141
30% MEA+15% MDEA	0.252	0.436	0.184

All mentioned data for the blend of MDEA+MEA are depicted below.

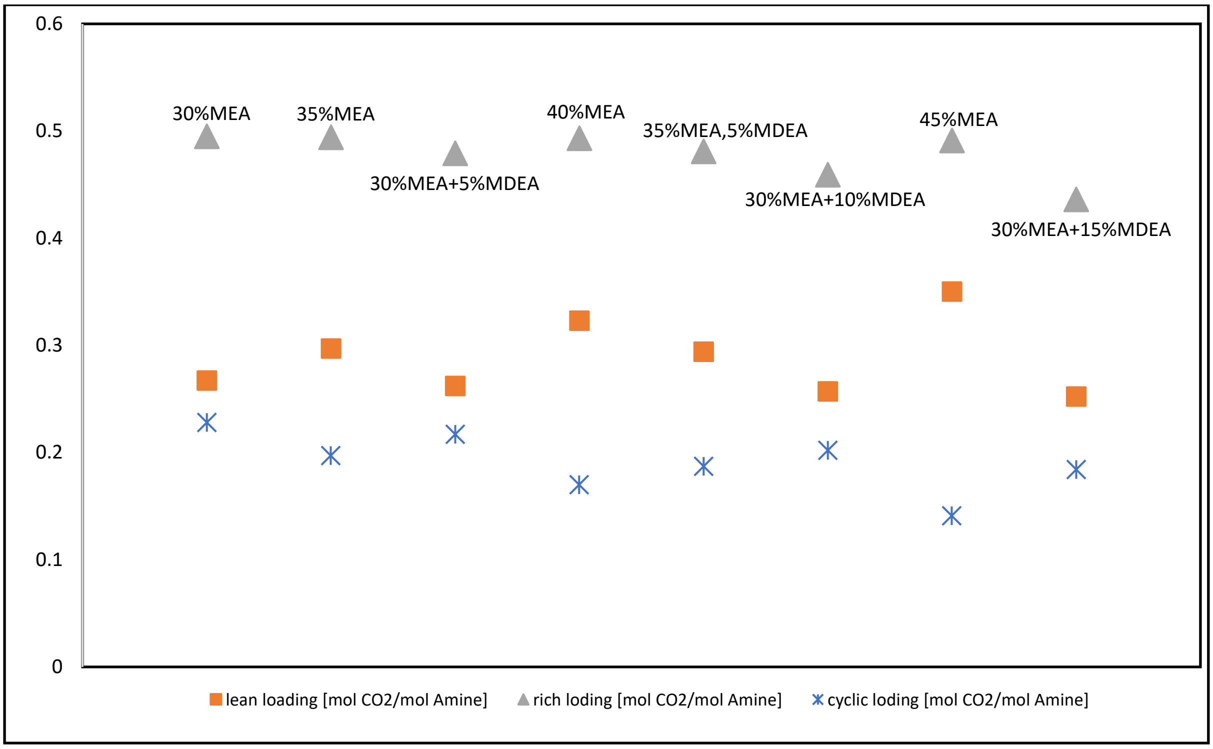

Figure 6.6: Investigation of lean, rich and cyclic loading for different concentration of MEA+MDEA blend (standard process) compared to base case

It is clear the presented blends of MDEA+MEA in figure 6.6 have higher cyclic loading than the same concentration of individual MEA.

In addition to MDEA+MEA and MEA+PZ blends, blend of MDEA+PZ has been investigated in standard process. Total concentration of mentioned blend has been assumed to be 50 wt% due to lower regeneration energy based on explained in previous sections.

The data of lean, rich and cyclic loading for this blend are listed below.

Table 6.9: Investigation of lean, rich and cyclic loading for 50 wt% MDEA+PZ blend (standard process)

Solvent/blend (wt%)	Lean loading [mol CO_2/mol amine]	Rich loading [mol CO_2/mol amine]	Cyclic loading [mol CO_2/mol amine]
30% MEA (Base case)	0.267	0.495	0.228
20% MDEA+30% PZ	0.119	0.579	0.460
15% MDEA+35% PZ	0.117	0.625	0.506
10% MDEA + 40% PZ	0.281	0.652	0.371

Table 6.9 is depicted as below to be comparable more.

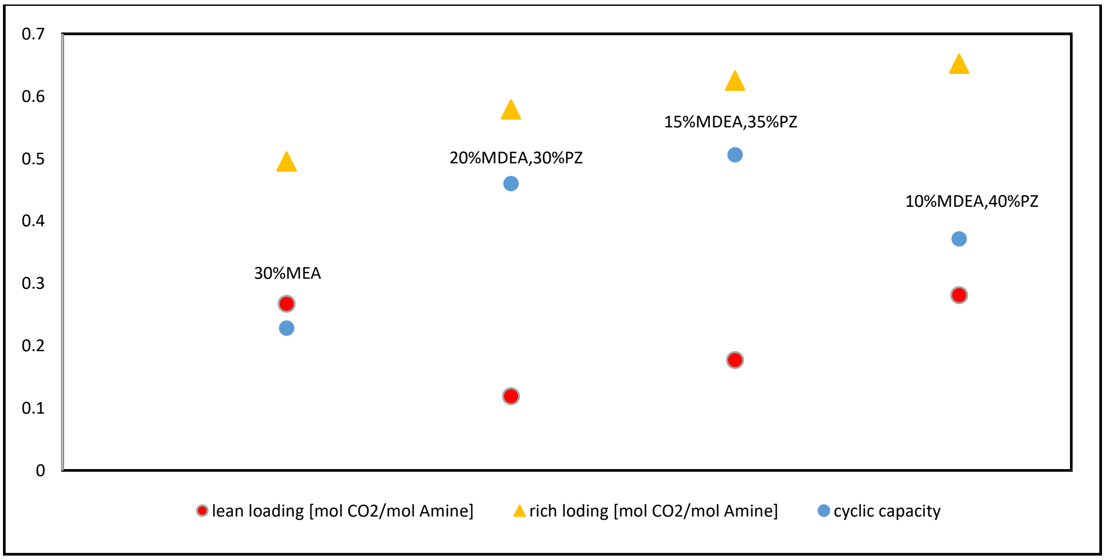

Figure 6.7: Investigation of lean, rich and cyclic loading for 50 wt% concentration of MDEA+PZ blend (standard process) compared to base case

6.3. Discussion about economy of the current work

The procedure of cost estimation for each removal process was completely explained in chapter 5. In this work, 11 different processes have been economically analysed including

1. Standard base case with 30 wt% MEA as solvent with efficiency of stage in absorber as 0.25.
2. Standard process with a blend of 25 wt% MEA+5 wt% PZ as solvent with efficiency of stage of absorber as 0.25.
3. Standard process with a blend of 30 wt% MEA+5 wt% PZ as solvent with efficiency of stage of absorber as 0.25.
4. Standard process with a blend of 30 wt% MEA+5 wt% PZ as solvent with efficiency of stage of absorber as 0.35.
5. Standard process with a blend of 30 wt% MEA+10 wt% PZ as solvent with efficiency of stage of absorber as 0.25.
6. Standard process with a blend of 30 wt% MEA+5 wt% MDEA as solvent with efficiency of stage of absorber as 0.25.
7. Standard process with a blend of 30 wt% MEA+10 wt% MDEA as solvent with efficiency of stage of absorber as 0.25.
8. Standard process with a blend of 30 wt% MEA+15 wt% MDEA as solvent with efficiency of stage of absorber as 0.25.
9. Vapor recompression base case process with a blend of 30 wt% MEA as solvent with the efficiency of stage in absorber as 0.25.
10. Vapor recompression base case process with a blend of 30 wt% MEA+5 wt% PZ as solvent with the efficiency of stage in absorber as 0.25.
11. Vapor recompression base case process with a blend of 30 wt% MEA+15 wt% MDEA as solvent with the efficiency of stage in absorber as 0.25.

It should be mentioned that for all listed processes the removal efficiency is 85% and pinch temperature in the lean rich heat exchanger is kept at 10°C.

6.3.1. The economy of standard configuration

The standard base case has been economically investigated. Presented figure below displays detailed CAPEX of this analysis.

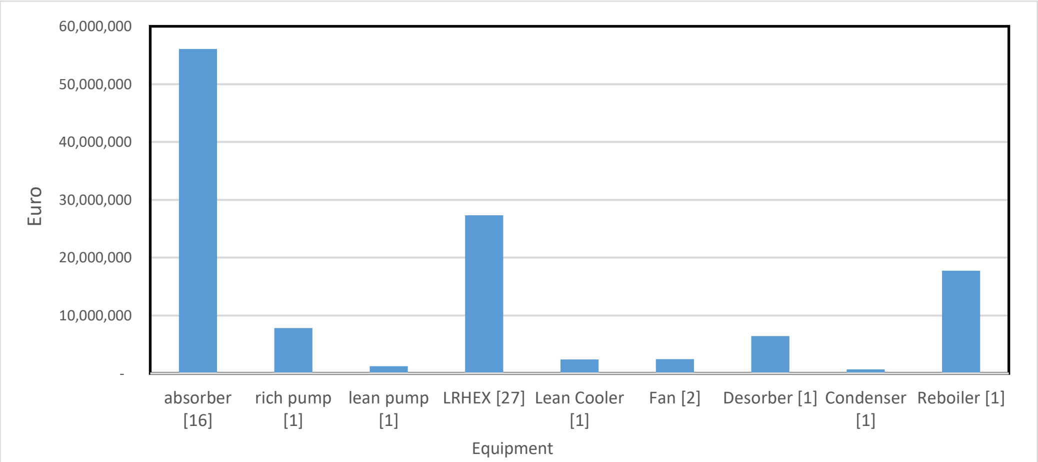

Figure 6.8: Distribution of CAPEX for applied equipment in the standard base case (year 2016), numbers for each item is written in []

The total cost for the standard base case is depicted below.

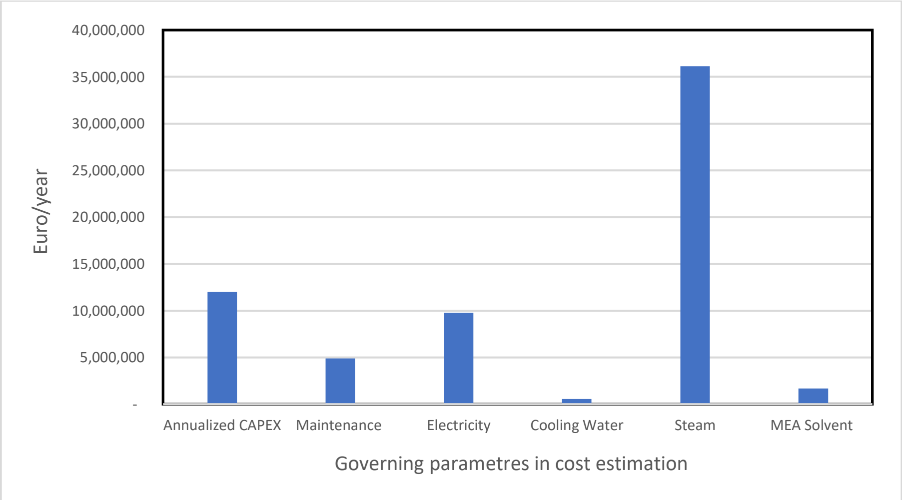

Figure 6.9: Distribution of annualized cost estimation for governing parameters in standard base case (costs are for year 2016)

As it is clear form figure 6.9, steam impose the largest part of costs to process, approximately 56% of total annualized cost.

6.3.2. The economy for blend of MEA and PZ in standard configuration

As it was mentioned earlier, adding 5 wt% PZ to MEA resulted into fall in regeneration energy of the process. Here, cost estimation of four different cases simulated with different concentrations of MEA and PZ are displayed. To be able of understanding the effect of piperazine on the economy of plant, standard base case is, also, mentioned.

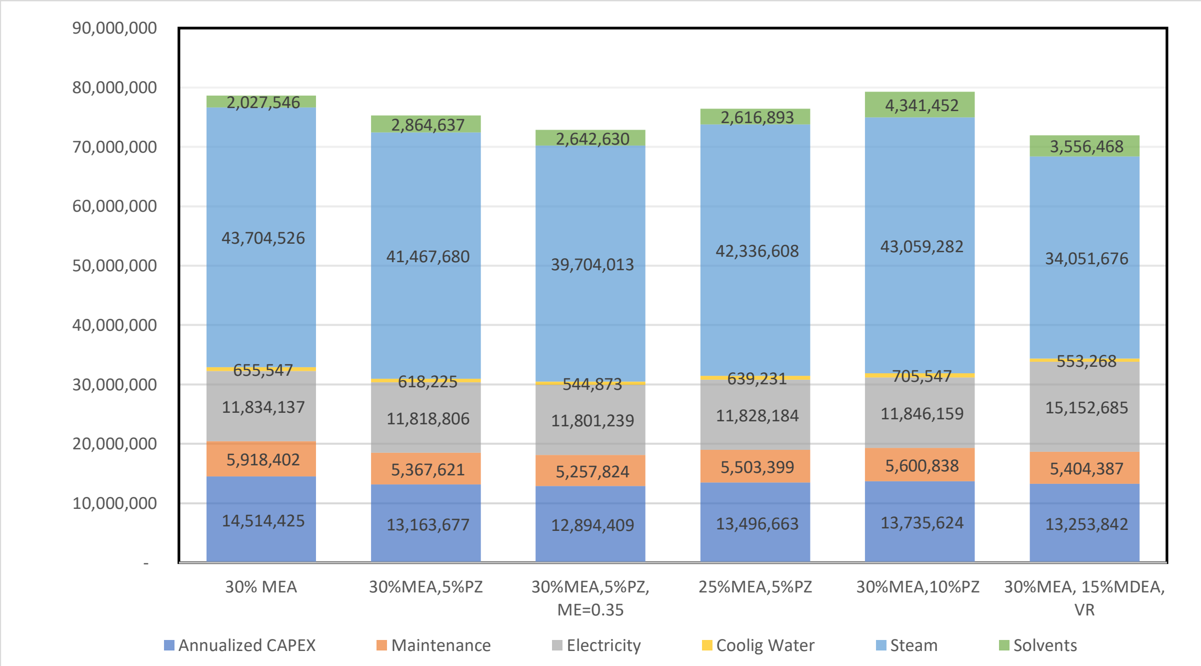

Figure 6.10: Distribution of annualized costs for governing parameters of economy in standard process, blend of MEA+PZ (all costs are based on Euro)

As it can be seen, except the process with 30 wt% MEA+ 10 wt% PZ, other cases are estimated to have lower cost than standard base case.

Total annualized cost for the standard base case was estimated to be around 78.7 million euro per year. Based on the analysing the economy of the plant with MEA+PZ blend, saving in the total annualized cost can be calculated. These savings are depicted below.

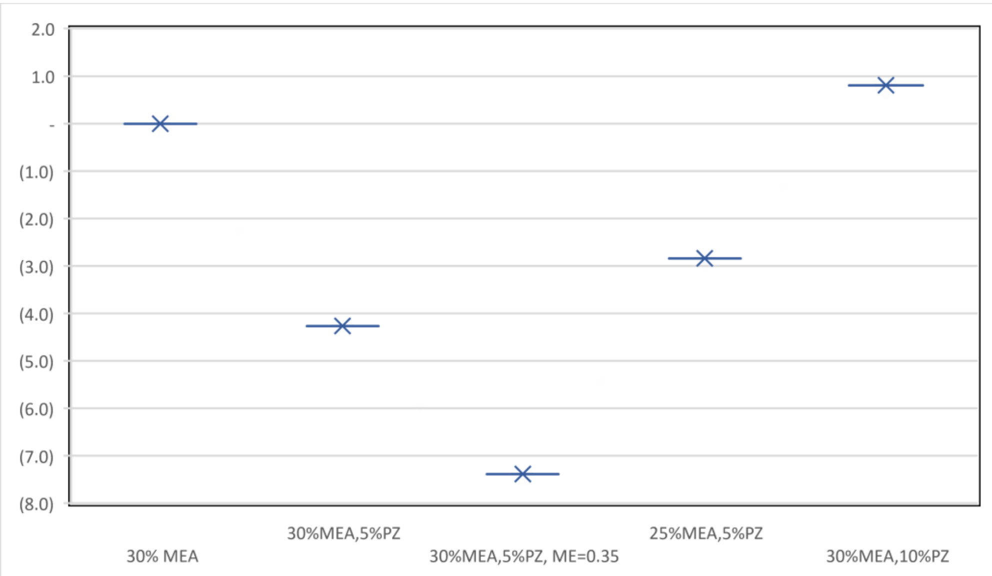

Figure 6.11: Percentages of cost savings in MEA+PZ simulations compared to standard base case (minus values indicate value of savings for simulations)

As it is clear, the blend of 30 wt% MEA+5 wt% PZ brings the highest value of saving, more than 7% of total annualized cost, for the standard process compared to base case.

6.3.3. The economy for blend of MEA and MDEA in standard configuration

Analysing the economy of current work proceeds with the cost estimation for different concentrations of MEA+MDEA blend. To be comparable with the standard base case, this case is, also, will be depicted in figure below.

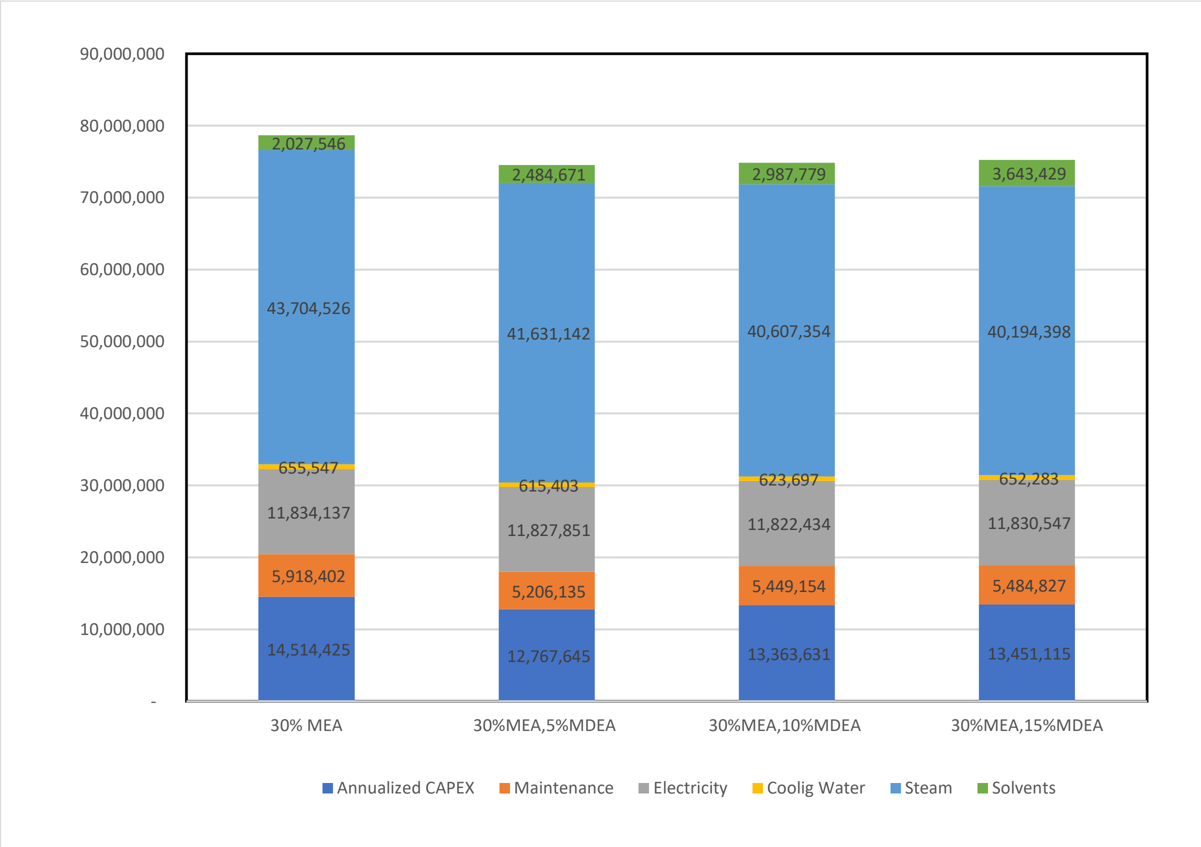

Figure 6.12: Distribution of annualized costs for governing parameters of economy in standard process, blend of MEA+MDEA (all costs are based on Euro)

A close look to conducted cost estimations in figure 6.12 indicates that all suggested processes with a blend of MEA+MDEA requires lower investments than standard base case. The percentages for cost saving compared to the standard base case are depicted below.

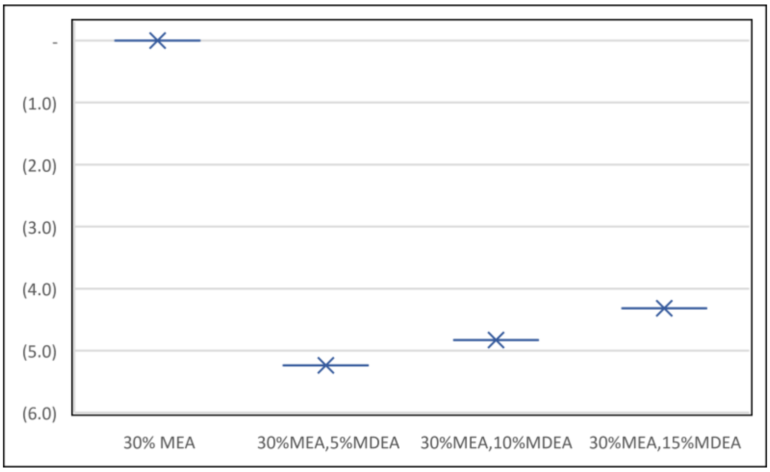

Figure 6.13: Percentages of cost savings in MEA+MDEA simulations compared to standard base case (minus values indicate value of savings for simulations)

Figure 6.13 displays that the blend of 30 wt% MEA + 5 wt% MDEA presents the highest percentage of cost saving, more than 5%. From regeneration energy view, for this combination of amines, the blend of 30 wt% MEA+15 wt% MDEA had the highest energy saving.

6.3.4. The economy of vapor recompression configuration

Vapor recompression process is one of the practical methods to save energy in CO_2 removal process. This process, in addition to equipment of conventional process, requires a compressor and separator (flash tank). The cost for mentioned items in this work are estimated to be approximately 1 million euro for separator and 9 million euro for compressor. Although, these costs have been imposed to the project due to switch to VR configuration, there have been other cost savings especially form steam.

In addition to increase in equipment cost, vapor recompression increases electricity in OPEX calculation. Because a large value of electricity is needed to satisfy the required compressor in the plant.

In this work, besides the standard process, vapor recompression configuration has been simulated for different solvents and their blends. Firstly, vapor recompression process was simulated with 30 wt% MEA as solvent. Removal efficiency and pinch temperature in the lean rich heat exchanger are identical to the standard base case. Efficiency of stage in the absorber is 0.25. The reason to simulate VR process with 30 wt% MEA is to be able to investigate the effect of other solvents/blends for this configuration.

Cost estimation for three simulated processes of VR method has been conducted. According to cost estimation, total annualized cost for vapor recompression process is around 74.5 million euro per year which is 5.5% lower than the standard base case. VR process, also, for the suggested blends saves costs rather than standard base case as well as VR base case.

Conducted cost estimation for vapor recompression is displayed below.

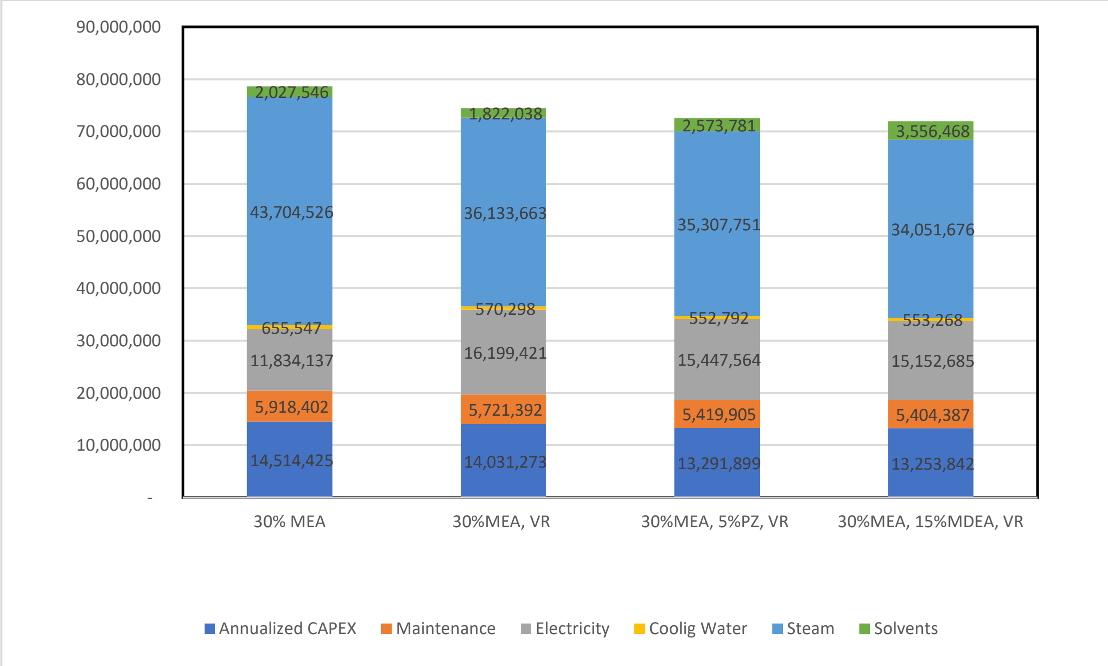

Figure 6.14: Distribution of annualized costs for governing parameters of economy in vapor recompression process, blends of MEA+MDEA and MEA+PZ (all costs are based on Euro)

As mentioned earlier, vapor recompression process in all simulated cases resulted into cost savings compared to standard base case. These savings are depicted below.

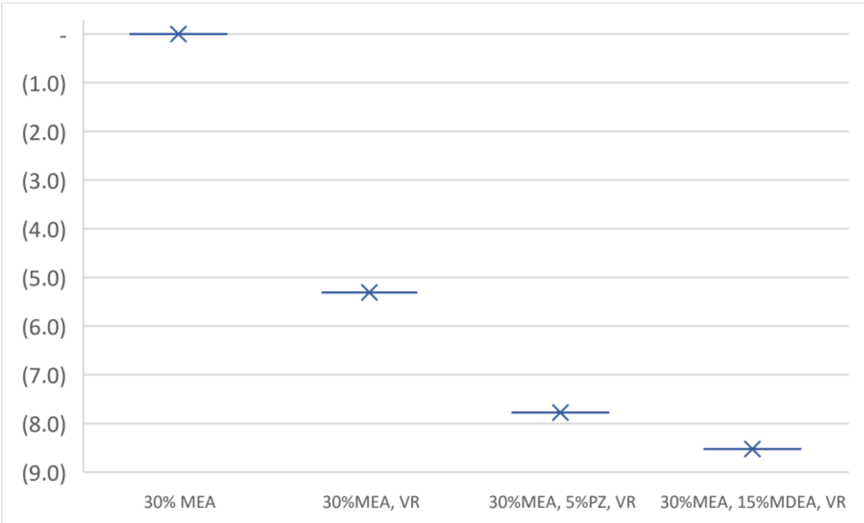

Figure 6.15: Percentages of cost savings in vapor recompression process for different blends (minus values indicate value of savings for simulations)

6.4. Uncertainties

In this part of work, identified uncertainties in different parts of the current project will be explained. Considering these uncertainties definitely brings improvement for future works.

6.4.1. Uncertainties about process simulation

The work is based on dada from simulations conducted in Aspen HYSYS. Simulations can be conducted in other programs like Aspen Plus and ProMax to compare validation of data.

Also, Aspen HYSYS includes different packages. For this work, chemical acid gas package has been selected to cover different solvents and their blends. Other packages can be investigated.

In the literature, the tri-blend of MEA+MDEA+PZ has been attractive from some views including lower regeneration energy, higher absorption rate, higher absorption capacity compared to individual MEA. Chemical acid gas package in Aspen HYSYS is not able to adjust the process for this solvent.

Some components in flue gas have been neglected in this work e.g. H_2S, while in reality these components should be included in the simulations.

Simulations in this work are based on equilibrium model, while rated-base model is another one which gives more accurate data.

6.4.2. Uncertainties about equipment dimensioning

Different literature has been investigated dimensioning and specifications to removal processes. Although, in this work, it was tried to apply those ones which have more credits, there are still some uncertainties.

In case of absorber and desorber, an extra height to accommodate washing section and enough space for flooding have been assumed. These values are selected from installed columns in plants. The total height for the absorber for different cases was assumed to be 40 m and regarding the stripper was 20 m. These values for each simulation should be optimized to present more accurate data. Also, rate of absorption for each solvent/blend is completely different with other ones. For instance, since individual MDEA has low reaction rate, higher absorption column might be applied. The reason is that solvent has sufficient time and space to react in absorber.

The efficiency of stages in absorber for CO_2 was assumed to be 0.25 for all simulations except for two cases with 30 wt% MEA+5 wt% PZ where efficiency of stages was assumed to be 0.3 and 0.35. In the literature, efficiency of stages has not been much discussed, while it affects the simulation extremely and consequently dimensioning.

Material specification to absorber and desorber was assumed to be stainless steel. This assumption for MEA solvent can be found in different works. In this work, the same material specification for the absorber and desorber has been assumed but other possibilities should be investigated. For instance, some literature investigated that PZ solvent does not have problems like corrosion like MEA. Other materials can save CAPEX costs.

One mandatory subject in the plant is to have spare items, especially for rotating items. As the current work more concentrates on the cost estimation procedure, spare items have not been included. In addition, lifetime for the project was assumed to be 20 years. This lifetime was extended to all equipment. In reality, the working lifetime for each item will be different with other type. So, to have more precise cost estimation, it is suggested that lifetime for each piece of equipment is studied separately.

Regarding the applied heat exchanger in this work, all required heat exchangers have been assumed to be shell and tube ones, while other types might be applicable. for instance, gasketed-plate heat exchanger has potential for cost saving than shell and tube one.

6.4.3. Uncertainties about CAPEX estimation

Besides uncertainty in design, uncertainties of commercial aspect impact on CAPEX estimation. CAPEX estimation in this project has been conducted based on some assumptions resulted from literature and Aspen In-Plant Cost Estimator.

Equipment prices have been extracted from Aspen In-Plant Cost Estimator program version 10 but as mentioned earlier, the best source for cost equipment is vendors, or similar items in other projects.

Equipment costs have been adjusted to total installed cost with the aid of EDF table. These coefficients might be changed from time to time based on market and place. So, it is more reasonable to apply other methods as well and compare the results.

Location factor is highly important factor in CAPEX calculation which can be implemented with a factor to total CAPEX. This work is based on location factor of 1, but it should be adjusted precisely.

CAPEX calculation only includes main items in the plant. That is why, a constant percentage of non-listed items has been assumed in this work. To be more precise, CAPEX should be included more items in the plant. Because the non-listed items has been assumed to be constant for all cases.

Absorber, fan and lean rich heat exchanger due to being highly large in size have been divided into smaller items in size. For each deviation, a particular scaling factor was assumed. These scaling factors are extracted from literature and might impose uncertainties to project.

6.4.4. Uncertainties about OPEX estimation

OPEX is mainly made of different utilities. The price for the utilities is varying in literature as well as time, place and availability. To be more precise it is recommenced that consider different prices for utilities as well as changes in costs in one year.

The workers and engineers for the plant in this work are not included. So, these assumptions change the calculated OPEX.

Heat exchanger will be faced fouling increasing linearly against time. This phenomenon reduces heat transfer performance. So, much more utilities will be required to compensate this reduction. It should be, also, mentioned that some new solvents/blends have shown better performance in fouling rather than MEA.

Depending on the place of plant, some utilities can be supplied cheaper or free of charge. For instance, cooling water in those areas in the vicinity of seas or lakes might be provided cheaper than assumed for the project.

6.4.5. Uncertainties about solvents/blends

Various studies have studied possible improvements from different aspects in CO_2 removal process experimentally or based on simulations. Actually, finding an optimal solvent or blend is not easy because numerous parameters should be taken into account. That is why, many suggestions from literature can be found while I some cases they are not in line with each other. Suggested blends of amines in this work are extracted from conducting lots of simulations and testing different parameters. It is suggested that those solvents or blends in this work would been tested experimentally and compered with results of simulations.

It seems there are still uncertainties about portion of each solvent in blends, specification of efficiency of stage in absorber, temperature and pressure in columns especially stripper.

6.5. Future work

Definitely, CO_2 removal will still be an interesting topic for more exploration in the future because the increasing trend of CO_2 emission. Among current methods for CO_2 capturing, amine-based has been the most applicable method especially with MEA solvent. Current works are attempting to introduce new solvents or blends to cover shortcomings for each solvent.

In addition to the removal process, the economy of the project should be considered. Gathering more data form offline and online sources or vendors, applying other cost estimation methods increase the accuracy for the project.

Other packages for the simulation are available in Aspen HYSYS program. Those packages can be investigated. In addition, this work is based on equilibrium method while rate-base method can be assumed for simulations.

Tri-solvent of MEA+MDEA+PZ is one of the most common among blends which brings improvements for the process from different sides. This blend is suggested to be studied.

Governing parameters, like pressure and temperature, in the plant should be optimized for other solvents or blends. For instance, for individual MEA temperature in the desorber was assumed to be 120°C due to avoiding degradation, while according to literature and experiments, PZ can handle higher temperature in the reboiler without degradation.

It is, also, reasonable to study corrosion matter for MDEA and PZ in the plant.

Efficiency of stages might be interesting for study because it seems this parameter for each solvent or blends should be optimized.

6.6. Conclusion

There has been a growing trend toward removing CO_2 emissions from the industry with different methods. One of the most mature methods for carbon capture is to absorb CO_2 in an amine-based (MEA) post-combustion technology. Shortcomings of MEA make other solvents and their blends more interesting in CO_2 removal plants.

The work in this master thesis is absorption-desorption CO_2 capture process simulated in Aspen HYSYS for different solvents/blends than MEA. Moreover, cost estimation methods for simulated cases have been performed to provide a complete cost estimation package. The data for cost estimation is provided with Aspen In-Plant Cost Estimator program.

A base case simulation model consisting of a simplified carbon capture unit including a 10-stage absorber, 6-stage desorption column, 85% CO_2 removal efficiency and minimum approach temperature for the lean/rich heat exchanger of 10 °C has undergone different solvents/blends of MEA, MDEA and PZ. The results indicate that adding 5 – 10 wt.% of piperazine to base case (30 wt.%) could offer a blend of solvents with lower regeneration energy than base case. Also, this matter was accurate for adding 5 – 20 wt.% MDEA to base case. Optimization of suggested range of blends has been performed in term of regeneration energy. Optimized concentrations could be as 30% MEA + 5% PZ (wt.%) and 30% MEA + 15% MDEA (wt.%) where lead into 4.9% and 7.5% lower regeneration energy than base case with 3.77 [MJ/kg CO_2]. These blends, also, have been simulated for vapor recompression configuration. Lean, rich and cyclic loadings for suggested blends in both standard and VR configurations have been discussed.

Aspen In-Plant Cost Estimator, applying Enhanced Detail Factor (EDF) method, was used for the cost estimation of processes. based on conducted cost estimations, plant with suggested blends presents cost savings rather than standard base case. Hopefully, the results in this thesis contribute to perform cost optimization more efficiently.

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Appendices

Appendix A – Submitted “Extended Abstract” for SIMS 2021 conference

Appendix B – Enhanced Detail Factor method table

Appendix C – Cost estimation data for some simulations

Appendix A:

Simulation and economic analysis of MEA+PZ and MDEA+MEA blends in post-combustion CO_2 capture plant

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Extended Abstract for SIMS2021

Key words: Simulation, Carbon capture, optimization, Aspen HYSYS, cost estimation, CAPEX, OPEX

Abstract

Amine based carbon capture is regarded as the most mature process to decrease or remove CO_2 emission from coal- and gas fired power plants. The process is based upon applying an amine, especially monoethanolamine (MEA) as the most actual amine (Øi L. E., CO2 removal by absorption, chanllenges in modeling), to dissolve CO_2 from flue gas in an absorption column shown in figure 1. The outlet solution from the bottom of absorber, rich amine, is sent to a stripper column to be regenerated and sent back to the absorber. The process can be controlled by numerous parameters. That is why various simulations and experimental studies have been conducted to improve performance of the process.

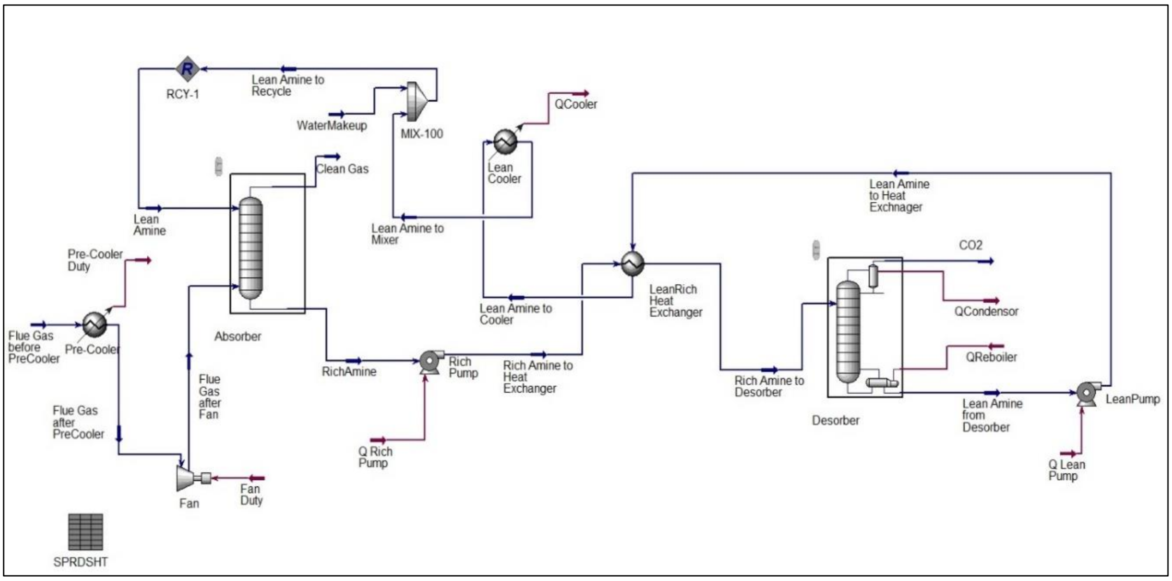

Figure 7: Simulated conventional process of removal CO_2 in Aspen HYSYS version 12

Generally, process improvements can be classified into three different categories, including (Dubois & Thomas, Comparison of various configurations of the absorption-regeneration process using different solvents for the post-combustion CO2 capture applied to cement plant flue gases, 2018):

- Different configuration of removal process e.g. vapor recompression
- Optimization of operational conditions e.g. pressure and temperature of absorber and stripper column
- Switch from monoethanolamine (MEA) to other solvents or their blends

Several projects have been conducted at Telemark University College and University of South-Eastern Norway to reach an optimal removal simulation known as base case where 30% MEA solvent absorbs CO_2 from flue gas (Øi, Sundbø, & Ali, Simulation and Economic Optimization of Vapor Recompression Configuration for Partial CO2 capture, 2017). MEA is one of the most important absorber liquids and the least expensive (Hasan, Abbas, & Ghavami Nasr, 2020).

The conventional simulated process, figure 1, has been performed in a 10-stage absorber, a 6-stage desorber and 10℃ as minimum different approach temperature in the lean rich heat exchanger. The removal efficiency is 85%.

The explained process could be performed with other sorts of solvents or their blends. Primary and secondary amines, like MEA, have fast reaction kinetics with CO_2 but with high energy consumption to regenerate amine in the stripper. Tertiary amines, like MDEA, require less regeneration energy but they absorb CO_2 slowly (Zhang R. , et al., 2017) (Borhani & Wang, 2019). In addition, corrosion and solvent degradation are drawbacks of MEA while for MDEA maximum loading capacity, lower corrosion and oxidative degradation than MEA are positive (Borhani & Wang, 2019). Piperazine (PZ) is added to increase the reaction rate. Thus, mixing amines could provide blends with less shortcomings. Other important parameters as heat of absorption, cyclic loading, CO_2 lean and rich loadings are not the same for different solvents and blends. For instance, (Zhang R. , et al., 2017) experimented heat of absorption for pure amines of MEA and MDEA where MDEA solvent had lower heat of absorption and consequently lower regeneration energy.

The most influential parameter for the total cost of removal plants is regeneration energy. Based on (Lee, Eslick, Miller, & Kitchin, 2013), this parameter accounts for up to 70% of energy demand. This study intends to simulate the effect of adding piperazine and MDEA to MEA in term of regeneration energy, cyclic capacity and CO_2 loading. Carbon Dioxide removal plant process have been simulated with 3 different concentrations of (MEA+PZ) where 5 wt%, 10% wt% and 15 wt% piperazine is added to 30 wt% MEA (base case).

The work proceeded with 5 different cases of MEA+MDEA blends where 5 wt%, 10 wt%, 15 wt%, 20 wt% and 25 wt% MDEA have been added and simulated to 30 wt% MEA. The results show that a blend of 30 wt% MEA + 5 wt% PZ is optimum in term of regeneration energy compared to other concentrations of MEA+PZ. Furthermore, 30 wt% MEA + 15 wt% PZ provides the lowest amount of regeneration energy among simulated cases for MEA+MDEA blends. The results are presented in figure 2 and figure 3 below.

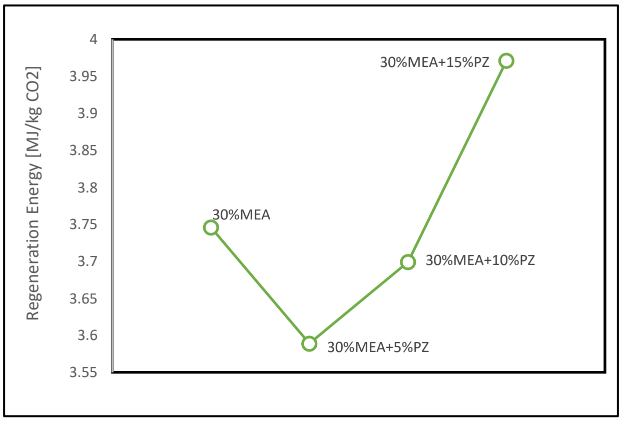

Figure 8: assessment of adding different concentration of piperazine to MEA in term of regeneration energy

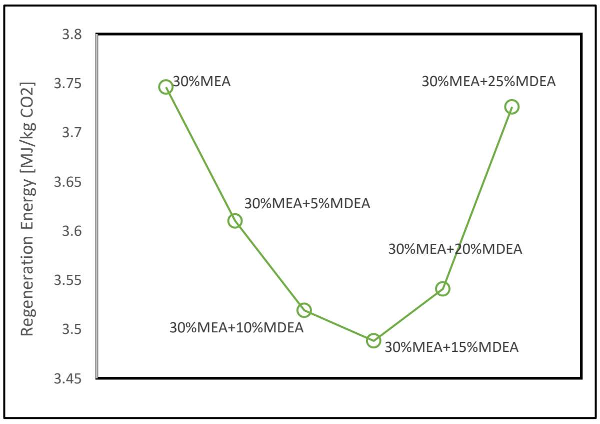

Figure 9: assessment of adding different concentration of MDEA to MEA in term of regeneration energy

Amine blends of (30 wt% MEA+ 5% wt% PZ) and (30 wt% MEA + 15 wt% PZ) led to a decline by 4.9% and 7.5% in regeneration energy compared to base case (30 wt.% MEA) with 3.771 MJ/ kg absorbed CO_2.

An economical study for whole simulated processes has been performed. These studies originate from mass and energy balance equations, resulting in dimensioning all equipment pieces in the plant. Aspen In-Plant Cost Estimator has been used for cost analysis. Calculated CAPEX updating material and other relevant expenses, e.g. engineering costs, direct costs and the Enhanced Detail Factor (EDF) method was applied. Besides, OPEX was calculated with the aid of extracted data from (Ali, Eldrup, Normann, Skagestad, & Øi, Cost Estimation of CO2 Absorption Plants for CO2 Mitigation - Method and Assumptions, 2019). Summation of CAPEX and OPEX forms total installed costs. The applied Aspen In-Plant Cost Estimator provides data for 2018, whereas the project should be updated to 2021 so that CEPCI (Chemical Engineering Plant Cost Index) was implemented.

Furthermore, both suggested blends have potential to improve the economy in a removal plant. Total amount, including OPEX and annualized CAPEX, for base case is 72.1 million Euro per year. According to economic analysis for simulated cases, both blends, (30 wt% MEA + 5 wt% PZ) and (30 wt% MEA + 15 wt% MDEA), lead to approximately 1.5% and 3.8% savings in total costs for a Carbon Dioxide removal plant which is mainly coming from reduction in required steam.

Appendix B:

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 C:

Cost estimation calculations for simulations. The unit for costs is [euro]

		30% MEA	30%MEA 5%PZ	30%MEA 5%PZ ME=0.35	25%MEA 5%PZ	30%MEA 10%PZ	30%MEA 5%MDEA	30%MEA 10%MDEA	30%MEA 15%MDEA	30%MEA VR	30%MEA 5%PZ VR	30%MEA 15%MDEA VR
Total installed cost (2016)	Absorber	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600	56,068,600
	Rich pump	7,864,617	1,069,239	971,043	1,105,180	1,274,615	1,095,553	1,073,731	1,122,508	1,003,133	919,699	914,565
	Lean pump	1,267,920	1,198,260	1,313,765	1,310,580	1,668,680	1,303,020	1,205,820	1,256,040	1,155,060	1,260,495	1,259,212
	LRHEX	27,357,514	24,344,880	21,404,486	26,402,968	27,560,689	25,980,063	24,371,970	24,904,740	21,775,845	18,387,338	18,705,645
	Lean Cooler	2,406,384	2,126,088	1,869,048	2,371,500	2,649,348	2,124,864	2,303,568	2,496,960	2,067,336	1,974,924	1,974,924
	Fan	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800	2,450,800
	Desorber	6,451,229	6,058,436	6,058,436	6,220,311	5,757,543	6,058,436	6,233,373	6,233,373	6,220,311	5,649,800	5,649,800
	Condenser	692,593	663,047	659,571	692,593	686,510	686,510	660,440	656,964	660,440	658,702	657,833
	Reboiler	17,738,306	16,937,270	17,852,023	17,099,810	17,619,035	11,811,819	18,233,121	18,148,571	16,365,535	15,904,886	15,364,269
	Separator									1,277,182	1,063,463	1,025,596
	Compressor									9,182,700	7,658,300	7,605,100
Annual OPEX (2016)	Maintenance	4,891,919	5,367,621	5,257,824	5,503,399	5,600,838	5,206,135	5,449,154	5,484,827	5,721,392	5,419,905	5,404,387
	Electricity	9,781,633	11,818,806	11,801,239	11,828,184	11,846,159	11,827,851	11,822,434	11,830,547	16,199,421	15,447,564	15,152,685
	Cooling Water	541,849	618,225	544,873	639,231	705,547	615,403	623,697	652,283	570,298	552,792	553,268
	Steam	36,124,444	41,467,680	39,704,013	42,336,608	43,059,282	41,631,142	40,607,354	40,194,398	36,133,663	35,307,751	34,051,676
	Solvent(s)	1,675,890	2,864,637	2,642,630	2,616,893	4,341,452	2,484,671	2,987,779	3,643,429	1,822,038	2,573,781	3,556,468
Annualized CAPEX (2021)		14,514,425	13,163,677	12,894,409	13,496,663	13,735,624	12,767,645	13,363,631	13,451,115	14,031,273	13,291,899	13,253,842
Total annualized cost (2021)		78,654,582	75,300,646	72,844,987	76,420,978	79,288,902	74,532,847	74,854,051	75,256,600	74,478,084	72,539,338	71,949,763