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Process simulation and cost estimation of CO_2 capture configurations in Aspen HYSYS

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Summary:

The use of amine-based carbon capture technology to capture CO_2 is a renowned technique. To evaluate and potentially utilize CO_2 capture techniques, economic assessments are very necessary and the performance of equipment in different configurations can potentially have a significant impact on capital and operating expenditures.

In this project, it is aimed to investigate the impact of varying flashing pressures on the cost of steam consumption and electricity in a vapor recompression system. By analyzing these parameters, the study aims to optimize the operating expenditure (OPEX) and the capital expenditure (CAPEX) associated to the system, contributing to the advancement of cost-effective process designs.

In this study, a base case was created in Aspen HYSYS V12 comprising a desorber packing height of 6 m and an absorber packing height of 10 m. The minimum temperature approach (ΔT_{min}) for the lean/rich amine heat exchanger was maintained at 10 °C. The vapour recompression case was simulated for comparisons, in which the inlet gas flow was 85000 [kmol/h] with a lean MEA rate of 120000 [kmol/h]. The estimation of the equipment units' dimensions was done based on the simulation results. The EDF method was utilised in conjunction with Aspen In-Plant Cost Estimator V12 to calculate the total, operational, and capital costs for the two scenarios. In order to assess the economic viability of vapour recompression and examine its effects, the sensitivity analysis for this project sets the flash separator pressure at 80–120 kPa. The aim is to determine how this pressure range affects CAPEX, OPEX, and the cost of capturing CO_2. Then, it is compared two cost estimation methods (the EDF and Nazir-Amini) together.

The EDF technique was used to determine the TPCs (CAPEX) for the base case and vapour recompression scenario, resulting in values of 64.34 M€ and 118.68 M€ in 2021. The projected yearly operational expenditure (OPEX) for the base case is around 32.14 M€/yr, but for the vapour recompression scenario at 100 kPa, it amounts to 28.9 M€/yr.

As part of the sensitivity analysis for the vapour recompression, the pressure in the flash separator was varied in order to determine how this variation affected the total cost. The optimal performance of the flash separator was determined to be 120 kPa. In this pressure, OPEX and CAPEX are determined at 28.3 M€/yr and 98.8 M€ respectively.

The base case, the vapour recompression (100 kPa), and the vapour recompression (120 kPa) scenario were projected to have CO_2 capture costs of 37.34 €/ton CO_2, 33.34 €/ton CO_2, and 31.42 €/ton CO_2, respectively. According to the investigation, the cost of carbon capture can be decreased by raising the flash pressure in the vapour recompression. Therefore, the results illustrate that the most economical scenario is the vapour recompression case at optimal pressure of 120 kPa.

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Preface

This master's thesis was completed in the autumn of 2023 at the University of South-Eastern Norway as part of the master's programme.

Sincere gratitude to my supervisors, Lars Erik Øi and Solomon Aromada, for providing me with this opportunity to expand my knowledge in this area.

Lastly, I would like to express my profound gratitude to my parents for their unwavering support and selfless efforts in guaranteeing that I am afforded the most favorable prospects in life.

Porsgrunn, December 2023

Alireza Razzaghianarmarzi

Contents

Nomenclature

Abbreviation	Description
MEA	Monoethanolamine
CAPEX	Capital expenditure
OPEX	Operational expenditures
CO_2	Carbon dioxide
N_2	Nitrogen
O_2	Oxygen
H_2O	Water
CCS	Carbon capture and storage
EDF	Enhanced Detail Factor
LMTD	Logarithmic mean temperature difference
CS	Caron steel
yr	Year
A	Cross-sectional area
D	Internal diameter
ΔT_{min}	The minimum temperature difference of lean/rich heat exchanger
SS	Stainless steel
$C_{Eq,x}$	Cost of the equipment in material x
f_M	Material factor
f_y	The factor for the y cost in equipment installation
$F_{T,x}$	Total installation factor in material x
$f_{Eq.}$	Equipment factor
$f_{pp.x}$	Piping factor in material x
$C_{EIC,x}$	The total installation cost for the equipment in material x
N	Number of identical units in the type of equipment
n	Plant lifetime in years
BEC	Bare Erected Cost
EPCC	Engineering Procurement Construction Costs
i	The discount rate in %
TPC	Total plant cost
CI_a	Cost inflation index for the year a
$Cost_a$	Cost of the equipment in year a

€	Euro
t	tons
Mt	Metric ton
M€	Million Euro

1 Introduction

Carbon dioxide (CO_2) is generated in significant quantities in industry due to the swift progress of modern civilization. Examples of such processes are the combustion of coal, coke and natural gas, the fermentation of carbohydrate materials and sugars, and the production of cement and gypsum, among others. Each year, in fact, over 30 billion tonnes of carbon dioxide are introduced into the atmosphere. However, the emission of carbon dioxide (CO_2), a significant greenhouse gas, has generated considerable apprehension regarding the correlation between human-caused CO_2 emissions and global warming. Potential contributors to urban pollution, acid rain, and health issues include CO_2 emissions [1]. Carbon dioxide (CO_2) capture from industrial flue gases has evolved into an urgent environmental concern. Increased concentrations of greenhouse gases in the atmosphere pose a substantial risk of climate change [2]. Consequently, it is becoming more and more important to create technologies that may lower CO_2 emissions from burning fossil fuels.

Typically, technologies utilised in the remediation of anthropogenic CO_2 fall into three distinct categories: precombustion CO_2 capture, post-combustion CO_2 capture, and oxy-fuel combustion. Post-combustion technology involves the combustion of fossil fuels in the same manner as traditional energy generation, followed by the capture of CO_2 from the effluent gas. This intuitive method is the subject of extensive research because it is retrofittable to existing power facilities. During precombustion CO_2 collection, the petroleum product gets gasified and reacts in a water gas transition reactor to create H_2 and CO_2. While the CO_2 is captured, the H_2 is converted into energy. Oxy-fuel combustion involves the utilisation of pure or nearly pure CO_2 as fuel, resulting in the production of predominantly CO_2 and HO_2 [2].

1.1 Background of interest for CO_2 capture

The IEA study [3] emphasizes the substantial importance of carbon capture, application, and store in facilitating the shift towards a low-carbon energy system. This technology is crucial for capturing emissions from industrial processes, serving as a means of removing carbon dioxide, and reducing emissions resulting from the utilisation of fossil fuels. The amount of carbon capture, use, and storage (CCUS) is projected to range from 4-6 Gt CO_2 by 2050 in the Accelerated and Net Zero scenarios, However, the New Momentum scenario predicts that the expected amount of CO_2 will be 1 Gt .Due to the extensive time required for the development of storage sites and their associated transport infrastructure, the majority of this capacity is typically finished during the latter part of the Outlook period. Around 15% of the carbon capture, usage, and storage (CCUS) facilities operational in 2050 are specifically allocated to capturing and storing non-energy process emissions from cement manufacturing. This is due to the low availability of other methods for reducing carbon emissions in this industry.

Significant apprehensions have been generated regarding the correlation between anthropogenic carbon dioxide emissions and the rise in worldwide temperatures, colloquially known as "global warming." This may result in more severe weather patterns, the melting of snow cover and ice formations, and an increase in sea level [4]. Acid rain, pollution, and alterations in the food supply are additional adverse environmental consequences associated with CO_2 emissions [1].

Furthermore, even at modest concentrations, CO_2 emissions pose direct dangers to human health. Kidney and bone complications, in addition to a decline in cognitive performance, are among the health issues that can arise from CO_2 exposure at concentrations as low as 1000 parts per million (ppm). Consequently, it is critical to prevent carbon dioxide emissions [5].

1.2 Literature review

The aim of this chapter is to examine the most relevant literature on the design, simulation, sizing, and cost optimisation of the CO_2 capture process. Multiple research initiatives have been carried out with the aim of reducing the cost of CCS.

Lars Erik Øi, Andrea Haukås, worked on the utilisation of MEA-based CO_2 collection, with a specific emphasis on three primary process configurations: standard process, vapour recompression case, and simple split-stream case. A series of parametric investigations were conducted to examine the impact of several factors on the effectiveness of carbon dioxide capture. The absorber height, the split ratio, the flash pressure, and the minimum approach temperature were among these variables. The investigations were conducted with a specific focus on achieving an 85% capture efficiency threshold.

The main results of the study reveal that the ideal parameters for cost efficiency are a packing height of 15 metres and a minimum approach temperature of 13°C for the conventional procedure. Additionally, a flash pressure at 150 kPa is recommended for the vapour recompression [6].

Øi et al. modelled a simple combined cycle gas power plant using Aspen HYSYS to assess the effectiveness of a monoethanol amine (MEA)-based carbon dioxide (CO_2) removal process. In this study, Peng Robinson and Amines Property Package models were employed to calculate thermodynamic properties. The natural gas-based power plant had a total thermal efficiency of 58% without CO_2 removal, and this dropped to around 50% once CO_2 removal was implemented. Heat consumption was assessed as a an outcome of many factors and, in the CO_2 removal with 85%, he determined to be 3.7 MJ per kilogramme CO_2 removed in this investigation, which is about the 4.0 MJ/kg CO_2 value mentioned in the literature [7].

Hasan Ali, in his doctoral dissertation, set out to create a framework for techno-economic analysis, with the goal of pinpointing key elements and illuminating how different technology and economic assumptions influence the expenditure of a capture plant. An amine-based post-combustion CO_2 extraction scenario (85% capturing efficiency) from a cement plant's exhaust gases was used to demonstrate the applicability of the proposed techno-economic analysis technique. The price per ton of CO_2 captured was 63 €. The most important factors influencing the baseline outcomes are the costs of steam, energy, and capital. Cost drivers were identified applying the Enhanced Detailed Factor (EDF) methodology because the Lang factor method was not intended to do so. In this study, natural gas-based steam production is anticipated to be more economical than coal- or biomass-based steam generation; nevertheless, the predicted steam cost is quite sensitive to market factors like fuel prices, which vary greatly globally [8].

Five distinct designs for aqueous absorption/stripping were examined by Karimi et al. (2011) in terms of initial investment and operating costs. For the lowest overall capture cost and CO_2 averted cost, he recommended the vapour recompression design, then the semi-lean amine is cooled in a split-stream setup [9].

In a study conducted by Shirdel (2022), an assessment was made about the variation in the inlet temperature of flue gas to the absorber. The temperature was examined within a range of 30 to 50℃, with increments of 5℃, in a 15-stage absorber. The use of a reduced the rate of amine flow, which enhances the efficiency of conventional stages, resulted in a decrease of about 2% in the expected collected cost compared to the study conducted in the Base Case scenario. The cost experienced a reduction of almost 4% with the use of comparable research in a hypothetical scenario including a 13-stage absorber. The study determined that the most favourable input temperature for the Basic Case was 34 °C, leading to a 39.6 €/ton CO_2 estimated. This finding was obtained by reducing the step size to 1℃ [10].

Aromada (2021). The installation elements of the approach used to estimate chemical plant costs rely on equipment costs. The Enhanced Detailed Factor (EDF) approach was used to assess the impact of equipment installation parameters on the capital cost of amine-based CO_2 collection plants. Plant construction characteristic variables account for diverse plant construction features. The EDF technique estimates new capital cost and this cost for modified plants, small and big, and adjusts for plant circumstances. He Studied the installation variables of several factorial cost estimating techniques on the capital cost and capture cost of an amine-based CO_2 capture system [11].

Aromada (2023). This research examined a cement plant setup that extracts CO_2 using both rich and lean vapour compression. In comparison to conventional procedures and straightforward rich vapour compression and lean vapour compression designs, its energy consumption, real CO2 emission reduction, and potential cost savings were evaluated. Both natural gases combined cycle power plants and renewable sources like water were looked at as ways to get energy. The three vapour compression designs did better than the normal CO_2 absorption configuration in terms of how much energy they needed, how much CO_2 they saved, and how much CO_2 they saved money on. The best results came from a configuration that used both rich and lean vapour compression. This design reduced costs by 24–30% for reboiler heat, 16–18% for equivalent heat, and 13–16% for CO_2 avoided. However, when comparing energy consumption and CO_2 emissions reduction, the combined process was only marginally superior to the straightforward lean vapor compression configuration. The hybrid method was likewise the best choice, according to the economic sensitivity analysis, however it was marginally superior to the lean vapour compression setup [12].

The technical and financial feasibility of four distinct methods to gather CO_2 after combustion from natural gas-fired power plants was examined by Gatti et al. It was determined that Molten Carbonate Fuel Cells (MCFCs) were the most appealing technology because of their 49 $/t CO_2 averted cost and 0.31 MJ LHV/kg CO_2 avoided SPECCA. The other technologies that were taken into consideration were CO_2 permeable membranes, pressurized CO_2 absorption, and supersonic flow-driven CO_2 anti-sublimation and inertial separation. The research demonstrated that CO_2 collection was possible with the integrated MCFC-NGCC systems at a significantly lower energy penalty and lower costs [13].

Masoumeh Dehghanizadeh (2023). This study evaluates an amine-based carbon capture technology using Aspen HYSYS. A base case with a 15-meter absorber and a 6-meter desorber, a minimal temperature approach of 10°C and a removal effectiveness of 85% were determined. Two additional scenarios were created to investigate the effects of increasing flue gas flow rates on the plant. Dimensioning and cost estimation were performed using Aspen HYSYS spread sheets. For the basic case, the case with doubled feed gas, and the case with two absorbers, the EDF technique yielded TPCs (CAPEX). In the event of increased feed, the predicted yearly OPEX is 83.1 MEuro, compared to 42.5 MEuro for the basic case. 52.4 €/ton, 51.8 €/ton, and 50.5 €/ton were the expected costs for carbon capture, respectively. The study demonstrated that increasing the flue gas flow rate can reduce the cost of carbon capture; the most cost-effective solution was found to be double the feed gas flow [14].

Aromada (2017). Aspen HYSYS Version 8.0 was utilised to simulate the standard process, vapour recompression, and vapour recompression combined with split-stream configurations for 85% CO2 capture from exhaust gas. The process details are predicated on the capture of CO_2 from a Mongstad. The energy optimum option is determined to be the vapour recompression alternative containing twenty absorber stages, nine desorber stages, 120 kPa flash pressure, and a minimum approach temperature (T_{min}) of five degrees Celsius. However, the vapour recompression technique with 15 absorber stages, 10 desorber stages, 130 kPa flash pressure, and 13 °C T_{min} is selected by the cost optimization study [15].

These process systems are compared to a MEA-based baseline for energy consumption reductions. Fernandez et al. (2012), he assessed lean vapor compression (LVC) and optimized to maximize the net present value (NPV) of process scheme savings instead of energy demand in the form of equivalent work. Two cases were examined. In case one, the capture plant was completely LVC-adapted. LVC is adapted to a simple capture plant in the second case. The net present value (M€) of the process scheme over the plant life was evaluated as a function of LVC operating conditions for both scenarios. The LVC process plan always has a positive and financially appealing NPV. The first LVC application case is most appealing. This technique reveals that reducing equivalent effort does not necessarily equal maximizing net present value, even while design factors and financial assumptions affect savings [16].

1.3 Objectives

The goal of this master's thesis is to discover the most cost-effective way to use the amine-based solution method for capturing CO_2 in the vapour recompression scenario by focusing on the best performance of the flash separator to reach the economic potential of the vapour recompression process. The investigation of capital and installation costs is subsequently conducted utilizing the Enhanced Detail Factor (EDF) method. The Nazir-Amini method is employed to estimate costs in a comparative analysis. The expected outcomes of this investigation are intended to offer significant perspectives that can improve the financial viability of CO_2 capture techniques.

It looks at the standard base case, which has a 10-stage absorber column, a 6-stage stripper, and a minimum approach temperature difference of 10 ℃ in the lean/rich heat exchanger. It uses a 29 wt% (MEA) solvent, inlet gas temperature 40 ℃ and a CO_2 inlet gas 3.73 mole%. Flue gas data comes from Lars Erik Øi's work [2].

2 Processes description of CO_2 removal

In the next part, it discusses several alternative carbon capture and storage (CCS) technologies, the usual procedure for amine-based processes, and an overview of the equipment used in the process.

2.1 Carbon capture technologies

Detailed explanations of the various carbon dioxide removal systems are provided in this section, which includes the primary classes of these technologies. Carbon capture methods may be characterized as processes or unit operations that remove carbon dioxide (CO_2) from gas mixtures to generate a stream that is high in and can then be stored or used [17]. The technologies for CO_2 capture are classified as illustrated in Figure 2.1 [18].

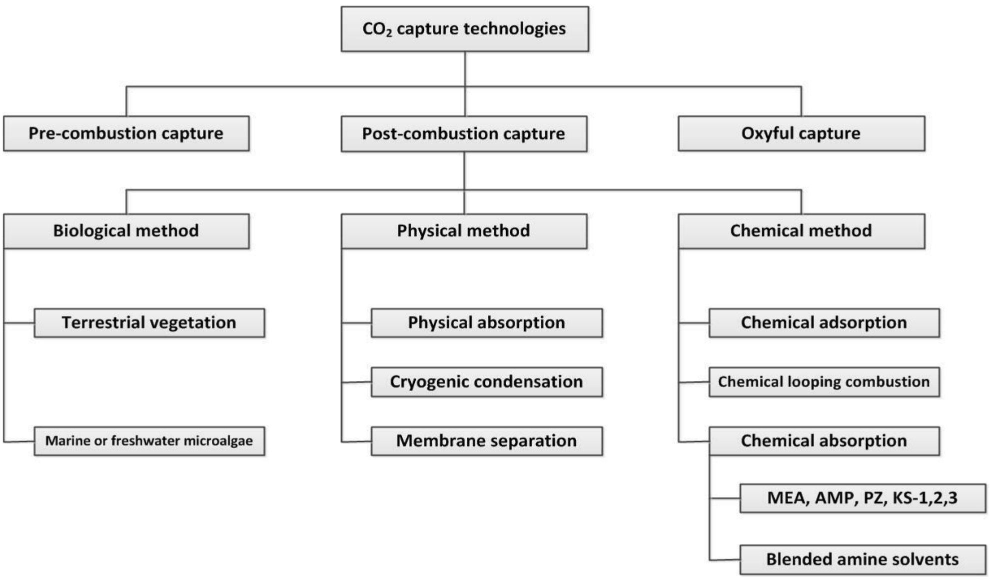

Figure 2.1: Technologies for removing carbon [18]

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2.1.1 Pre-combustion CO_2 capture methods

There is a process that takes place in which fossil fuel is transformed into synthesis gas for further combustion. Pre-combustion systems are methods of turning fuel—solid, liquid, or gaseous—into syngas without requiring combustion. The elimination of carbon dioxide from the mixture is made possible as a result of this, all before the hydrogen is used for burning [19].

2.1.2 Oxyfuel combustion CO_2 capture methods

The desired gas component (oxygen) is extracted from air entering the Air separation unit (ASU) with a high purity of 95–97%. When anthracite coal and oxygen are introduced into the firebox, combustion takes place. The combustion reaction generates flue gas, a byproduct comprising water and carbon dioxide, in addition to impurities including SO_x, HCL, HF, and fly ash. As a result of the comparatively low sulphur content of this coal variety, however, sulphur components are generated in negligible quantities and are therefore deemed insignificant. Following this, the combustion vapour undergoes several treatment procedures to eliminate unwanted contaminants. By recirculating 80% of the flue gas into the furnace, the temperature of the reactor is regulated and the concentration of CO_2 in the flue gas is increased, thereby enhancing CO_2 capture. The vapour generated during the combustion process has the potential to be utilised in power generation. The CO_2 and H_2O mixture are passed through a series of purification devices, which facilitate the removal of water via dehydration. Ultimately, only the captured CO_2 remains. Compression, transportation, and storage of the CO_2 constitute the final phase [20].

2.1.3 Post-combustion CO_2 capture methods

Post-combustion CO_2 capture refers to the process of capturing carbon dioxide (CO_2) via the atmosphere or flue gases produced by the burning of fossil fuels. Basically, carbon dioxide (CO_2) is extracted from the flue gas in power plants or other significant emission sources. Monoethanolamine (MEA) is the preferred solvent due to its superior reactivity with CO_2 compared to other secondary or tertiary amines. This enables absorption to occur in a more compact column [18]. The process flow diagram is shown as follows figure 2.2.

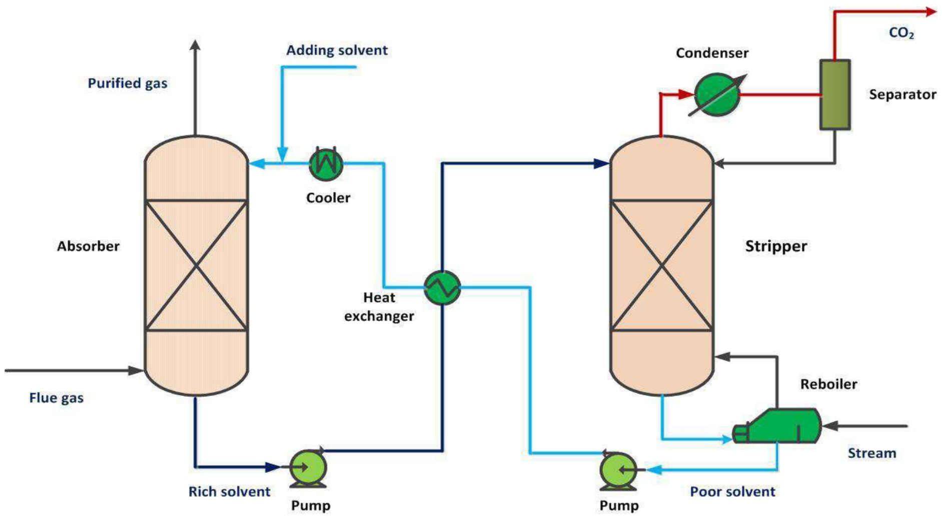

Figure 2.2 : The design of the CO_2 removal process by MEA [18]

MEA is the solvent most frequently employed in CO_2 capture systems. The primary components through which the absorption and desorption processes occur, respectively, are the absorber and stripper.

The packed column is often used as the absorber and stripper in CO_2collection systems due to its stability and high efficiency. In the chemical absorption process, the flue gas is introduced into a packed bed absorber from the bottom and enters to contact with a CO_2 lean solvent in a counter-current manner for absorption. Afterwards, the CO_2 rich solvent is sent to a stripper for thermal regeneration. Following the process of solvent regeneration, the resulting solvent with reduced CO_2 content is returned to the absorber for the purpose of absorbing CO_2 once again [18].

MEA is consistently selected as the absorbent in chemical absorption processes involving CO_2 due to its accelerated reaction rate, which enables absorption to occur within a reduced column length. However, the conventional MEA process encounters several drawbacks when it comes to separating $CO_2$from flue gas: (1) a low capacity for capturing carbon dioxide (g CO_2/g absorbent); (2) amine degradation caused by SO_2, NO_X, and oxygen present in the flue gas, resulting in a significant rate of absorbent makeup; (3) elevated devices corrosion rate; and (4) substantial energy consumption for solvent regeneration [18].

2.2 Description of Amine solution technology

Among different solvents to be used in the post-combustion CO_2 capture method, using an amine-based solution is currently the most advanced and cost-effective way because of the reversible reactions with CO_2. This process involves passing flue gas through an amine solution based on fluids, which absorbs and traps carbon dioxide (CO_2). After being moved to a stripper, the CO_2-enriched amine solution is heated with steam to cause the CO_2 to be freed from the solution. Figure 2.3 shows the schematic of an amine unit [4]. Mono-methanol amine, often known as MEA, is what is the most typical amine for this method.

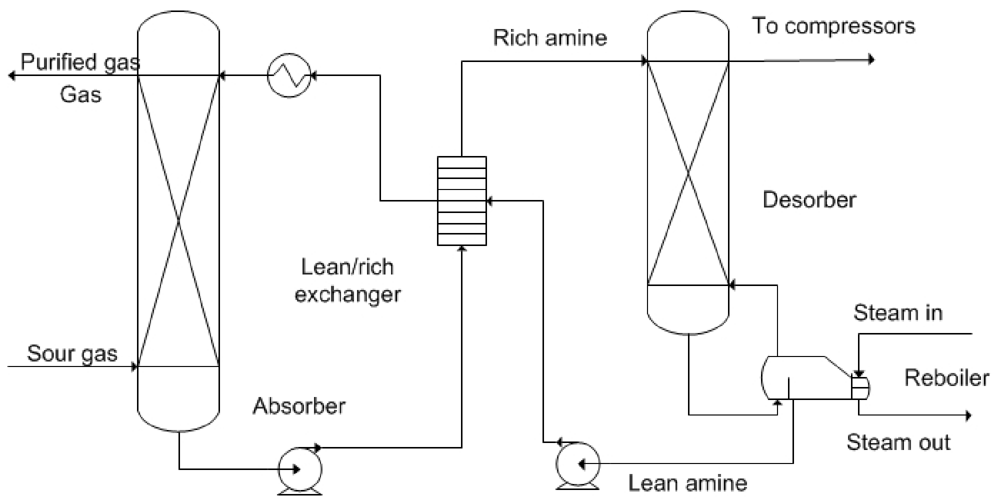

Figure 2.3 :Schematic of amine unit [21]

2.3 Description of equipment in amine-based CO_2 capture plant

Within the framework of this gas scrubbing procedure, the absorber and the desorber column are two of the most important pieces of apparatus. The cyclic process also includes other pieces of machinery, such as pumps, heat exchangers, reboilers, cooler and condensers, among numerous other components.

2.3.1 Absorber

The absorption column is where the primary reaction between CO_2 and amino acid solution occurs. The Amin solution is applied from the top of the column, whereas the discharge gas from the power plant enters from the bottom. Packing material is utilised to enhance the interface area between the liquid solvent and the discharge gas within the column. Due to the exothermic property of the chemical reaction, the column will experience a corresponding decrease in temperature and pressure as the reaction progresses.

The absorber design is a critical component in the process of CO_2 capture. The solvent of the absorbent may be physical or chemical. When selecting a solvent type, one of the primary considerations is the solubility of the desired solute. This solubility is subject to variation depending on temperature and pressure. The chemical solvent utilised in this endeavour is MEA. The absorber employs two distinct varieties of contactors, namely trays and packed towers. Structured packaging is among the most widely used packed structures in industry. A low-pressure decline is a critical element in enhancing its appeal. The implementation of structured packaging is employed in this undertaking. Gas velocity is another crucial factor to consider when designing an absorber. Changes in gas velocity resulted in corresponding modifications to the packing's diameter. A reduction in packaging diameter results in a concomitant decrease in pressure and an increase in energy consumption [22].

2.3.2 Desorber/ stripper column

By utilising steam, a stripper or desorption column is utilised to separate CO_2 from the amine solution. As the separated CO_2 gas exits the stripper from the top, the column is evacuated from the bottom by the regenerated solvent, lean amine.

It is reasonable to presume that the pressure along the side of the desorption column remains constant, whereas the temperature decreases in an ascending direction [23].

2.3.3 Rich and Lean Amine Pumps

These two pumps are used in the process to compensate for the pressure loss in the absorber and desorber. Rich MEA pump increases the pressure of liquid from the bottom of the absorber to the desorber, and when CO_2 and MEA are separated in the desorption column, then the free MEA is sent back to the absorption column. In principle, each pump should possess adequate head to offset all process losses. The pressure disparity between the absorber and desorber, heat exchanger losses, the nature of the liquid or solution, as well as losses in the pipelines, absorber, and desorber, should all be considered when determining the required head of the pump [24].

2.3.4 Reboiler

One of the most significant contributors to the overall operational expenses of absorption-based CO_2 removal facilities is the quantity of heat that is necessary to renew amine solution. During the procedure, the reboiler is responsible for supplying this quantity of heat. To be more specific, a reboiler is a type of heat exchanger that allows steam to enter as a hot stream to supply the necessary heat for the flow that occurs in the bottom of the stripper chamber [23].

2.3.5 Condenser

The carbon dioxide that is expelled from the top of the desorber is cooled down in the condenser, which is used to condense the amine or water that is carried over from the previous step in the stripping process. After that, the condensate is sent back to the desorber in the form of reflux. For the purpose of absorbing the heat from the hot fluid, water is utilised as a cooling medium [25].

2.3.6 Flash Separator

In the second simulation, the separator is adjusted to remove any remaining CO_2 in the lean MEA stream and direct it to the compressor to offset the pressure decrease. The stream goes into a two-phase filter, with vapour on top and liquid on the bottom. As the vapour product goes into the reboiler, it is squeezed, which raises the temperature and pressure.

The primary heat exchanger receives the bottom output from the two-phase separator, which is pumped through it and mixed with the rich stream to share heat. Before returning to the absorber, the lean flow is cooled further [21].

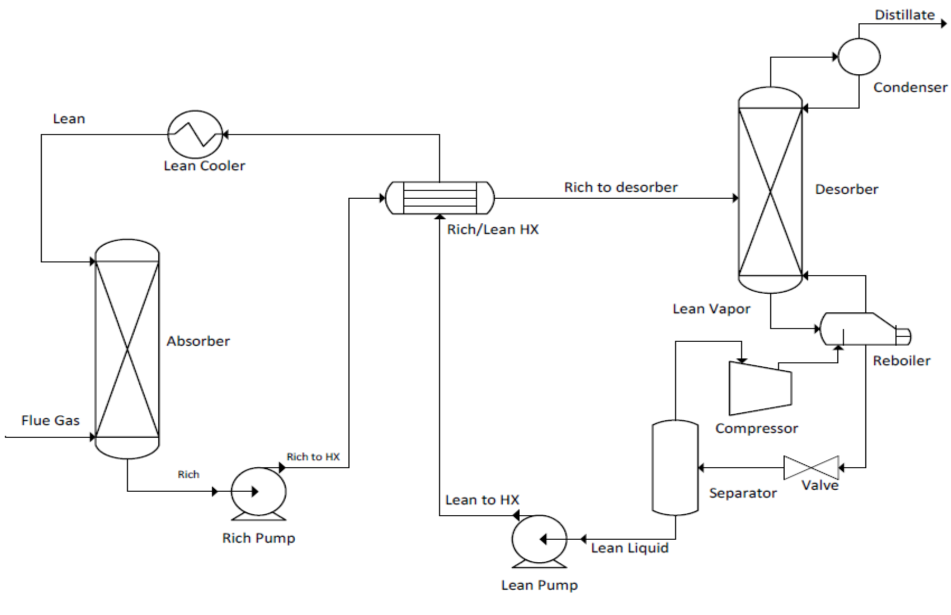

Figure 2.4 : Vapour recompression flow chart [21].

2.3.7 Heat-Exchanger

The rich amine solution is heated up before it enters the stripper column by the process of crossflow heat exchange with the lean amine solution that is coming from the stripper column figure 2.5. This helps to minimize the duty of the reboiler that is in the desorption column. Additionally, to achieve the temperature that is necessary for the lean amine solution, the duty of the amine cooler is decreased. Lean-rich heat exchangers are among the costliest pieces of equipment in this CO_2 collection process. They are also among the most efficient. However, it is necessary to determine the optimal ΔT_{min}, which may be determined by calculating the trade-off between the area of the heat exchanger and the amount of energy that is used. At the same time as there is a downward trend in the surface area and heat recovery, there is also a downward trend in the ΔT_{min} with an increasing tendency [25].

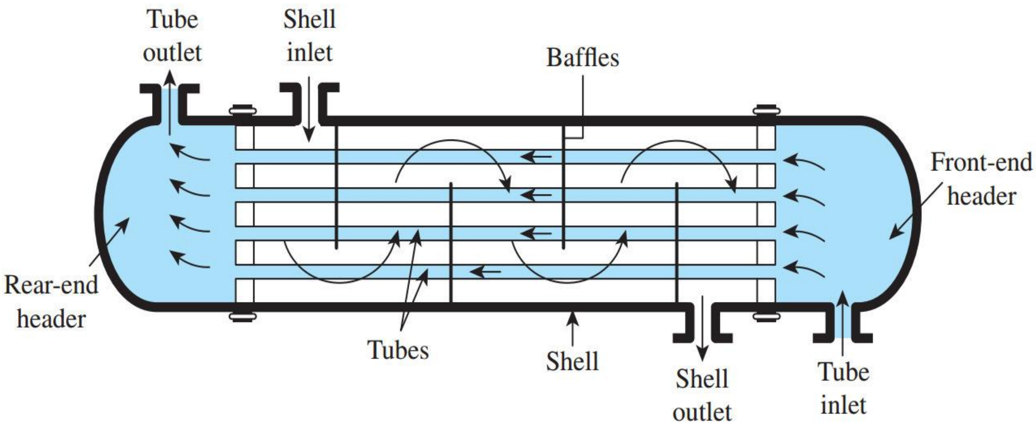

Figure 2.5: representation of a shell and tube heat exchanger schematic [26]

2.3.8 Compressor

In the second design (vapor recompression case), a compressor is employed to elevate the pressure of the CO_2 gas and then recompress it, thus enhancing the desorption efficiency.

2.3.9 Lean amine cooler

Lean amine is cooled by a lean amine cooler, which is in the lean/rich heat exchanger. The reason for the decrease in temperature is that the optimal temperature difference between the lean solvent and the absorber is roughly 40 °C [28].

3 Simulation in Aspen HYSYS

This chapter focuses on the process simulation of the amine-based CO_2 capture plant in Aspen HYSYS V12. This simulation consists of two cases. The first one is known as Base Case, and the second one is the Vapor Recompression Case. This chapter also covers the specifications of these two cases.

3.1 Base Case

In the context of this simulation, a plausible base case has been defined based on the data that has been supplied. The fluid package that was utilized in this investigation was acid gas, which is a chemical solvent. An illustration of the specification of this base case may be seen in Table 3.1 [7]. In addition, the Aspen HYSYS flowsheet of the simulated process is displayed in figure 3.1, and this figure also displays the Aspen HYSYS flowsheet of the simulated process.

In the most basic scenario, the minimum approach temperature for the rich/lean heat exchanger is equivalent to 10 °C. Within the scope of this investigation, the temperature of the rich MEA stream that is directed towards the desorber is altered in order to assess alternative minimum approach temperatures for the rich/lean heat exchanger. Detailed explanations of the findings are provided in the following chapters.

Table 3.1: Specifications for the basic scenario of modelling the CO2 capture process [7].

Sour Gas temperature	40 °C
Sour Gas pressure	110 kPa
Sour Gas flow rate	85000 kgmole/h
The CO_2 content in Sour Gas	3.73 mole%
Water content in Sour Gas	6.71 mole%
Lean MEA temperature	40 °C
Lean MEA pressure	110 kPa
Lean MEA flow rate	120000 kgmole/h
MEA content in Lean MEA	29 mole%
CO_2 in Lean MEA	5.5 mole%
Number of stages in the absorber	10
Murphree's efficiency in the absorber	0.25
Rich MEA pump discharge pressure	200 kPa
Rich MEA to desorber temperature	104.9 °C
Number of stages in the desorber	6
Murphree's efficiency in the desorber	1.0
Reflux ratio in the desorber	0.3
Reboiler temperature	120 °C
Lean MEA pump discharge pressure	200 kPa
ΔT_{min} in Rich/Lean Heat Exchanger	10 °C

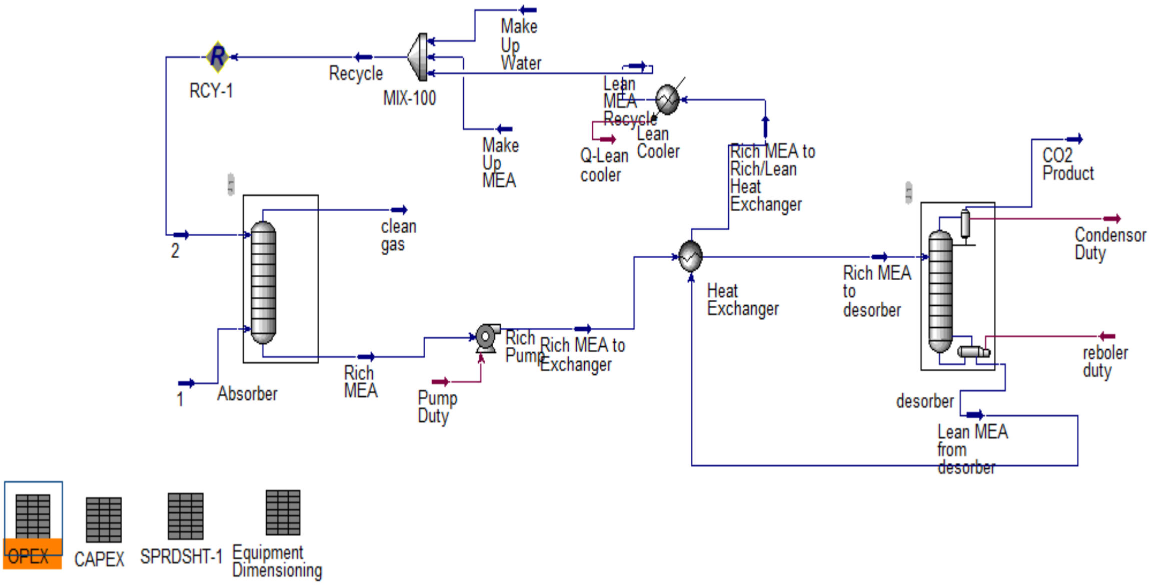

Figure 3.1: The basic case's Aspen Hysys flowsheet.

3.2 Vapor Recompression Case

When the vapour recompression process is carried out, the lean MEA stream that originates from the desorber is depressurized from 200 kPa to 100 kPa by means of a valve. Following this, the stream is introduced into a separator at a pressure of 100 kPa. A liquid Lean MEA is pumped to the lean pump at the bottom of the container. At the very top, the vapour is introduced into a compressor, where it undergoes a pressure rise to 200 kPa before being fed back to the desorber. For the vapour recompression scenario, the Aspen HYSYS flow sheet is depicted in Figure 3.2 respectively[21].

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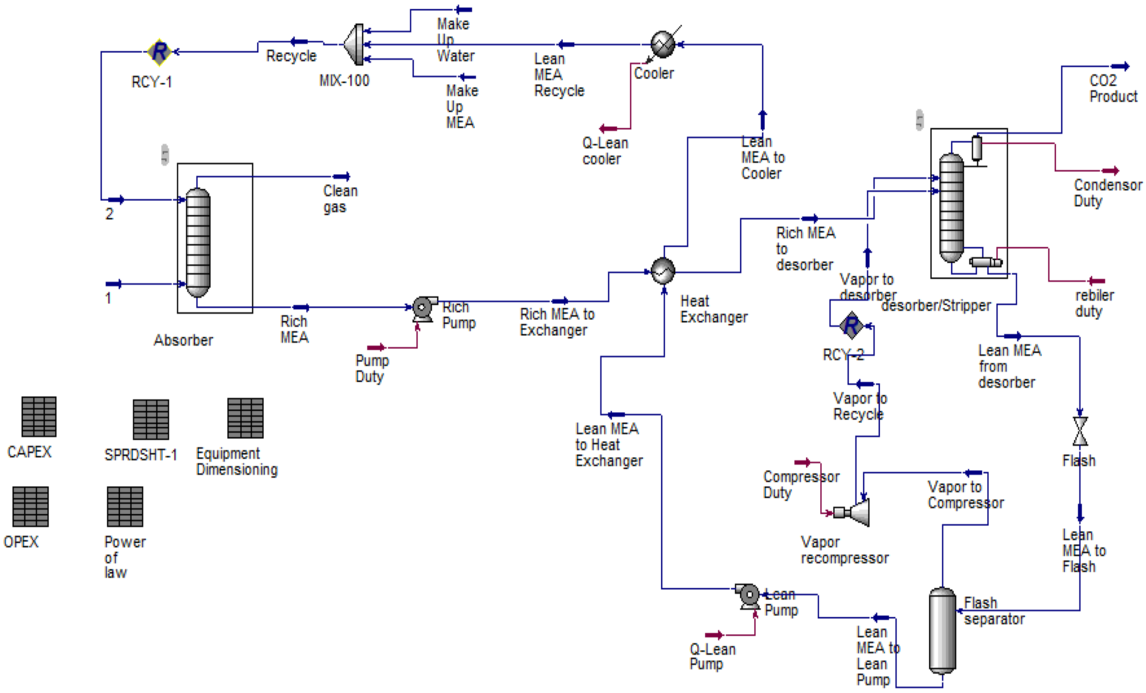

Figure 3.2: Aspen Hysys flow sheet of the vapor recompression scenario.

4 Dimensioning

The sizing of process equipment in both the base case and the vapor recompression scenario is the topic of discussion in this section.

4.1 Absorber

The velocity of the gas contained inside the column must be stated in order to establish the diameter of the absorption column. The usual range of gas velocity is between 2 and 2.5 m/s per second [6]. It is feasible that the cross-sectional area of the column be calculated by using Equation (4.1). The letter A denotes the cross-sectional area, the letter V gas denotes the volumetric flow rate, and the letter $\dot{V}_{gas}$ denotes the gas velocity. We may calculate the diameter of the column using Equation (4.2), where D indicates the diameter of the interior of the column. Table 4.1 displays the gas flow and absorber dimensioning values.

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

$$D = \sqrt{\frac{4 \times A}{\pi}} \tag{4.2}$$

Table 4.1: Calculation of the absorber diameter for base-case and vapor recompression

v (gas)	2.5	m/s	Assumption
Volumetric flow rate	2010030	m^3/h	HYSYS
	558.34	m3/s	HYSYS
Cross section area	223.34	m^2	Equation (4.1)
Diameter	16.86	m	Equation (4.2)

4.2 Desorber (Stripper)

On the same level as the absorption column, we can determine the cross-section area (A). The symbol V represents the velocity of the gas, which is assumed to be 1 meter per second [20]. A look at Table 4.2 reveals the dimensioning characteristics that are associated with the desorber.

Table 4.2: Desorption column diameter calculation for base-case and recompression method.

	Base-case	Recompression case	
v (gas) (m/s)	1	1	Assumption
Amine density (kg/m^3)	1087	1072	HYSYS
Amine mass flow (kg/h)	246100	1659748	HYSYS
Number of stages	6	6	HYSYS
Diameter (m)	9.33	6.196	Equation (4.2)

4.3 Lean/Rich MEA Heat Exchanger

To determine the appropriate size of the heat exchanger, it's necessary to compute the total surface area for heat transfer using Equation (4.4). This was stated that the total heat transfer values for the rich/lean heat exchanger would be 500 W/m^{2}k [11]. When ΔT_{min} is set to 10 °C, LMTD is found. The total heat transfer area is found based on the heat load. The area of each unit is 915.2 m^2 for base case and 980 m^2 for vapor recompression.

Regarding the lean/rich heat exchanger, LMTD is possible to obtain directly from Aspen HYSYS. However, for other equipment, we rely on Equation (4.5) to calculate the LMTD. In this equation, The symbol ΔT_{in} denotes the difference in temperature between the hot and cool intake streams ($T_{hot,in} - T_{cold,in}$), the temperature difference between the hot and cold exit streams is represented by ΔT_{out} ($T_{hot,out} - T_{cold,out}$).

$$A = \frac{\dot{Q}}{U \times \Delta T_{lm}} \tag{4.4}$$

$$\Delta T_{lm} = \frac{\Delta T_{out} - \Delta T_{in}}{\ln(\Delta T_{out}/\Delta T_{in})} \tag{4.5}$$

To determine the quantity of heat exchanger units, it is assumed that the maximum heat transfer area for each unit is 1000 m^2. Therefore, it is necessary to establish a specific criterion for the heat transfer area of each unit, ensuring that the area of each unit does not exceed 1000 m^2. Detailed information on the dimensions and characteristics of the Lean/Rich MEA Heat Exchanger is shown in Table 4.3.

Table 4.3: Dimensioning of the lean/rich heat exchanger.

Parameter	Base-case	Recompression case	
Heat duty (kJ/h)	1.318E+08	1.273E+08	HYSYS
U (W/m^{2}k)	500	500	Assumption
LMTD (°C)	36	9.277	HYSYS
Total Area (m^2)	7321	27440	Equation (4.4)
Number of units	8	28	
Area per unit (m^2)	915.2	980	

4.4 Reboiler

Since the reboiler can be dimensioned using the same manner as the lean/rich MEA heat exchanger, it can be done to apply the equations given in section 4.3 to it. The Reboilers characteristics and dimensions are listed in Table 4.4. It is assumed that the total heat transfer coefficient is 1200 W/ m^2K [6]. The reboilers dimensions and characteristics are listed in Table 4.4.

Table 4.4: Dimensioning of the reboiler.

Parameter	Base-case	Recompression case	
Reboiler duty (kJ/h)	7.194E+08	5.094E+08	HYSYS
U (W/m^2K)	1200	1200	Assumption
LMTD (°C)	27.7	35.10	Equation (4.5)
Total area (m^2)	6000	12090	Equation (4.4)
Number of units	7	13	
Area per unit (m^2)	858.8	1008	

4.5 Condenser

The duty of the condenser is determined using Aspen HYSYS. The LMDT is computed by utilizing the temperatures of the condenser's cold and heated surfaces. The literature provides the overall heat transmission coefficient U, which is assumed to remain constant at 1000 W/(m^2. K) [28]. The heat exchanger equation is employed to determine that one condenser unit is necessary and that a total heat exchanger area of 173.6 and 248.4 m^2 is required for base case and vapor recompression case respectively.

Table 4.5 presents the dimensions and parameters of the condenser.

Parameter	Base-case	Recompression case	
Condenser duty (kJ/h)	6.034E+07	6.466E+07	HYSYS
U (W/m^2K)	1000	1000	Assumption
Vapor inlet temp (°C)	117.2	97.81	HYSYS
Condensed vapor outlet temp (°C)	116	86.81	HYSYS
LMTD (°C)	96.56	72.31	Equation (4.5)
Total area (m^2)	173.6	248.4	Equation (4.4)

4.7 Lean MEA Cooler

For the dimensioning of the lean MEA cooler, again, we can use the equations in section 4.3. Dimensioning and specs for the Lean MEA Cooler are shown in Table 4.6. The inlet coolers were specifically constructed with a total heat transfer coefficient of 800 W/ (m^2. K) [28]. The logarithmic mean temperature difference (LMTD) is determined by taking the minimum temperatures of the cold and hot sides of the input cooler into account. The overall heat transfer area is computed based on the acquired heat duty. Considering the overall heat transfer area and the maximum size of each inlet cooler unit in the first scenario, it is determined that five inlet cooler units are necessary. The calculated actual area for each unit is 896.3. Additionally, two inlet cooler units with an actual area of 888.4 are also computed.

Table 4.6: Specifications and dimensions of the lean MEA cooler.

Parameter	Base-case	Recompression case	
Cooler duty (kJ/h)	5.025E+08	3.803E+07	HYSYS
U (W/m^2K)	800	800	Assumption
CW inlet temp (°C)	20	20	Assumption
CW outlet temp (°C)	20	20	Assumption
MEA inlet temp (°C)	87.16	54.88	HYSYS
MEA outlet temp (°C)	40	40	Assumption
LMTD (°C)	38.93	26.76	Equation (4.5)
Total area (m^2)	4482	830.97	Equation (4.4)
Number of units	5	2	
Area per unit (m^2)	896.3	888.4	

4.6 Pumps

The dimensions of the lean pump and the rich pump are discussed in this section. An adiabatic efficiency of 75% is the goal of the pump design in Aspen HYSYS. While the Aspen In-Plant cost calculator employs volumetric flow to ascertain equipment prices, the duty is utilized as a sizing criteria for pumps purchased from Aspen HYSYS [29]. A comparison of the two pumps' specifications can be seen in the below Table 4.7.

Table 4.7: Specifications of the Lean MEA Pump and Rich MEA Pump.

Parameters	Lean pump: Vapor recompression case	Rich pump: base-case	Rich pump: Vapor recompression case
Power (kW)	99.32	127.9	87.28
Act. vol flow (m^3/h)	2813	3487	2618
Actual. No. of Units	1	1	1

4.7 Compressor

The centrifugal compressor is the type of compressor that is utilized in the vapor recompression scenario. To demonstrate the specifications of the compressor that was utilized in the vapor recompression scenario, Table 4.8 provides an illustration of its adiabatic efficiency, which is estimated to be 75% [29].

Table 4.8: The specifications of the compressor utilized in the vapor recompression scenario are listed

Power (kW)	4137	HYSYS
Inlet act. vol flow (m^3/h)	1.482E+05	HYSYS
Adiabatic efficiency	75%	Assumption
Outlet pressure (kPa)	200	Assumption

4.8 Flash Separator

A vertical separator is used as part of the structure of the gas recompression process. So that you can figure out what size the divide is. In this equation 4.6, V_{Gmax} (m/s) represents the maximum allowable gas velocity, the value of Ks that is recommended is 0.081. Density values for the gas and liquid components of the stream are also retrieved from these modelled operations. So, V_{Gmax} is determined using formula 4.6, and the diameter of the separator is derived using formula 4.7. and $\rho_L(kg/m^3)$ and ρ_G $\rho_L(kg/m^3)$ represent the densities of the gas phase and liquid phase, respectively [23].

$$V_{Gmax} = K_s \sqrt{\frac{\rho_L - \rho_G}{\rho_G}} \tag{4.6}$$

Equation (4.7) can be used to find the minimum required diameter. Where D_{min} is the minimum required diameter in metres, q_a is the real volumetric gas flow rate in metres per second, and F_g is the fraction of usable cross-sectional area for gas flow. F_g is thought to be 1 for a vertical divider [23].

$$D_{min} = \sqrt{\frac{\left(\frac{4}{\pi}\right) q_a}{V_{Gmax} F_g}} \tag{4.7}$$

$$L/D_{min} = 2.5 \tag{4.8}$$

Equation (4.8) can be used to compute the separator height, where L (m) is the separator height. Table 4.9 displays the following are the essential characteristics for the separator:

Table 4.9: The vertical separator specifications

Liquid density (kg/m^3)	1010	HYSYS
Gas density (kg/m^3)	0.612	HYSYS
K_S (m/s)	0.081	Assumption
maximum allowable gas velocity (m/s)	3.2	Equation (4.6)
Act. vol gas flow rate (m^3/h)	92.24	HYSYS

F_g	1	Assumption
Minimum required diameter (m)	5.9	Equation (4.7)
Height (m)	14.9	Equation (4.8)
Flash separator	Vertical	Specified

5 Cost estimation Procedure

The purpose of this section is to provide an estimate of the overall cost that the facility will incur for the CO_2 capture method that has been developed. The calculations for the cost estimate are based on the sizing that was obtained from the findings of the Aspen HYSYS. In the basic case model, the costs for each component are computed by using the Aspen In-Plant Cost Estimator version 12 software. Two distinct approaches, namely the EDF method and the Nazir-Amini methodology, have been employed in the project to compute the Total Plant Cost (TPC). We are going to next to compare these two methodologies.

For the purpose of estimating the total cost of the plant based on the model that is simulated in Aspen HYSYS, the below procedure is utilized [12]:

1. Using the equipment sizing data for the Base Case and the vapor recompression case and applying these data to calculate the cost of each piece of equipment by using Aspen In-Plant cost estimator.
2. Utilizing the Enhanced Detailed Factor (EDF) to determine the overall installation cost
3. Calculating present value by applying cost index correction (year conversion).
4. Annualized capital expenditure (CAPEX) is calculated by considering the discount rate and the expected lifespan of the plant, calculation of annual operational expenditure (OPEX)

5.1 Capital expenditure for current project (CAPEX)

The cost estimate of each item in the plant that is sized and described in the dimension component is the first step in the CAPEX calculations for this operation. There are several methods for calculating or obtaining the cost of the plant's equipment. A variety of sources are utilized in capital cost predictions, including budgeted pricing, estimated offer forms from vendors, internal data from previous projects, commercial databases, publications, and the Internet [28].

As previously stated, commercial databases are also accessible and can be utilised to estimate equipment costs. The CAPEX for this task is determined by importing equipment costs from version 12 of Aspen In-Plant Cost Estimator. The costs listed for this version are accurate as of 2019.

5.1.1 Enhanced detailed factor (EDF) method

For the purpose of evaluating the Bare Erected Cost (BEC) and Total Plant Cost (TPC), an enhanced detailed factor (EDF) approach has been developed. The EDF technique has various benefits, including excellent accuracy in early-stage cost estimations, specific process equipment optimization, and the capacity to do techno-economic evaluations of new technologies or expansion projects. To work properly, the EDF technique requires fundamental data such as a streamlined equipment list and equipment cost. Equipment costs can be estimated using past data from a comparable plant or procedure, or the Aspen In-plant Cost Estimator. To estimate equipment expenses, the factorial method is not employed by this programme. Instead, it obtains equipment costs directly from equipment makers of things [30]. When determining the cost of the equipment, it is essential to make sure that it is adjusted to reflect the appropriate dimensions, year, and material of construction [14].

5.1.2 Material Factor (f_M)

It is important to use material factors (f_M) to change the cost of the equipment into carbon steel if it is not made of carbon steel. This is since the Detailed Factor sheet (Appendix B) was made using carbon steel material costs. It's essential to know that the only things that will be changed are the pipelines and equipment. For each type of material, Table 5.1 shows the material factors [14].

Table 5.1: Technical parameters of the material and the material factor suitable to the equipment [11].

List of equipment	**material**	**Material factor (f_M)**
Absorber	SS304 welded	1.75
Disrober	SS304 welded	1.75
Lean Rich Heat Exchanger	SS304 welded	1.75
Cooler	SS304 welded	1.75
Reboiler	SS304 welded	1.75
Condenser	SS304 welded	1.75
Rich amine pump	SS304 machined	1.3
Lean amine pump	SS304 machined	1.3
Compressor	SS304 machined	1.3
Separator	SS304 welded	1.75

The cost of the equipment in Carbon Steel can be calculated using Equation (5.1) [14].

$$C_{Eq,CS} = \frac{C_{Eq,SS.}}{f_M} \tag{5.1}$$

$C_{Eq,CS}$: Equipment cost in carbon steel

$C_{Eq,SS.}$: Stainless steel equipment cost

f_M : Material factor

5.1.3 Total installation cost factor (F_T)

The overall installation cost factor comprises the sub-factors for direct expenses, engineering costs, administrative costs, and commissioning and contingency costs. Every individual piece of equipment is susceptible to incurring these expenses. To get the total installation factor for carbon steel ($F_{T,CS}$), we employ Equation (5.2). However, all the factors can be obtained from the EDF table in Appendix B, taking into account the equipment cost for carbon steel ($C_{Eq,CS}$) [11].

$$F_{T,CS} = f_{direct} + f_{administation} + f_{commissioning} + f_{contingency} \tag{5.2}$$

f_{direct} : Factor for direct installation cost

$f_{engineering}$: Factor for engineering cost in installation

$f_{commissioning}$: Factor for commissioning cost in installation

$f_{contingency}$: Factor for contingency cost in installation

For the purpose of this research, it is necessary to convert $F_{T,CS}$ to $F_{T,SS}$ by utilising Equation (5.3) and the subfactors EDF table situated in Appendix B. This adjustment needs to be done for each piece of equipment that is constructed from SS, by adding piping factor in carbon steel and equipment factor.

$$F_{T,SS} = \left[F_{T,CS} + \left\{(f_M - 1) \cdot \left(f_{Eq.} + f_{pp,CS}\right)\right\}\right] \tag{5.3}$$

$F_{T,SS}$: Total installation factor in stainless steel

$f_{Eq.}$: Equipment factor

$f_{pp,CS}$: Piping factor in carbon steel

Equation (5.4) can be used to calculate the total equipment installed cost [11].

$$C_{EIC,SS} = C_{Eq,SS} \cdot F_{T,SS} \tag{5.4}$$

$C_{EIC,SS}$: Total installation cost for the equipment in stainless steel

$C_{Eq,SS.}$: Cost of the equipment in stainless steel

To derive the total installed cost for each piece of equipment, multiply the calculated factor by the equipment cost from the Aspen In-Plant Cost Estimator. The sum of each item's CAPEX contributes to the total CAPEX [28].

5.1.4 Total plant cost (TPC)

The total of all equipment installation costs is known as the Total Plant Cost (TPC), Equation (5.6) can be used to compute it.

$$TPC = \sum (C_{EIC,SS}) \tag{5.6}$$

5.1.5 Currency and inflation index adjustment

All expense calculations in this analysis are denominated in Euros (€). Utilizing the Aspen In-Plant cost calculator, the equipment's cost in euros is determined. The EDF approach's factor table additionally includes the currency conversion of equipment costs into euros [11].

The equipment expenses are assessed using data received in 2019 in Version 12 of the Aspen In-Plant cost estimator. This signifies that in order to obtain a current and precise cost estimate, the expense must be indexed to inflation. For 2020, the data utilised in the detail factor table to compute the installed cost factors was collected. Thus, it is necessary to initially revise the apparatus cost to incorporate cost data as of 2020. The EDF method will subsequently be utilised to estimate the overall cost of the installation. Ultimately, the cumulative implemented cost from 2020 to the subsequent year must be adjusted for inflation [29].

Table 5.2: Cost inflation indices for the years 2019 and 2020 [11].

Year	**Cost inflation index**
2019	110.1
2020	112.2
2021	116.1

It implies that an adjustment for inflation must be made to the expense in order to obtain a current and accurate cost estimate.

$$Cost_a = Cost_b \cdot \left(\frac{Cost\ index\ _a}{Cost\ index\ _b}\right) \tag{5.7}$$

The variable $Cost_a$ stands for the cost of equipment in year 'a', while $Cost_b$ represents the cost of equipment in year 'b'. On the other hand, CI_a denotes the cost inflation index in year 'a', and CI_b represents the cost inflation index in year 'b'.

5.1.6 CAPEX Assumptions

Table 5.3 summarizes all the assumptions considered for CAPEX estimation.

Table 5.3: Assumptions in the CAPEX method.

Parameter	**Value**	**Source**
Method of CAPEX estimation	EDF method, 2020	[11]
Cost data year	2019	Aspen In-Plant
Cost currency	Euro (€)	[11]
Project lifetime	20 years	[6], [23]
Discount rate	7.5%	[6], [23]
Discount factor 20 years	10.19	

5.1.7 Power law

Modifying parameters to conduct a sensitivity analysis on the simulation will result in a modification of the equipment's dimensions. The expense of the equipment will be evaluated to the initial price determined using the EDF approach for the vapour recompression instance in this study. This will be done using a spreadsheet called Power of Law. According to the Power Law, the link between equipment dimensions or performance and costs is not necessarily linear. Instead, costs are dictated by the product of capacity and an exponential component. The representation of this may be achieved using Equation (5.8) [23]:

$$\left(\frac{Cost\ of\ B}{Cost\ of\ A}\right) = \left(\frac{Capacity\ of\ B}{Capacity\ of\ A}\right)^{e} \tag{5.8}$$

The exponential size factor, indicated as e, typically varies between 0.35 and 1.70, depending on the equipment type [33]. In the current research, the absorber and desorber columns are considered to have an exponential factor of 1.1, while the rest of the equipment such as pumps, flash separator, compressor, cooler, heat exchanger and reboiler has a factor of 0.65.

5.2 Operating expenditure (OPEX)

A significant proportion of total expenses are devoted to operations and maintenance. It is customary to allocate OPEX expenses as either fixed or variable. An example of fixed expenditures is operational labour and maintenance costs. A proportion of equipment installation cost (EIC) ranging from 2% to 6% is frequently used to estimate maintenance expenses; in this study, 4% was utilized. The annual number of hours worked and the quantity of personnel on staff comprise the operational labour cost. Variable costs consist of consumables such as raw materials, electricity, steam, and solvents. It is possible to determine the annual cost of the utilities that have been given by using equation (5.8) [34], [10], [28].

Table 5.4: Assumptions for OPEX.

Parameter	**Value**	**Unit**	**Source**
Electricity cost	0.136	[€/kWh]	[35]
Steam cost	0.015	[€/kWh]	
Annual operational time	8000	[hours/year]	[30]
Plant lifetime	20	[year]	
Maintenance	4% of CAPEX		[30]
Location	Rotterdam		

Equation (5.8) is utilized in the computation of the annual OPEX.

$$Annual\ utility\ cost \left[\frac{€}{yr}\right] = Consumption \left[\frac{unit}{hr}\right] \cdot unit\ price \left[\frac{€}{hr}\right] \cdot operating\ hours \left[\frac{hr}{yr}\right] \quad (5.8)$$

5.3 Annual total costs and CO_2 capture expenditure

There are several approaches to calculate the overall expense of a CO_2 collection plant. Two methodologies are employed to assess the project's economic viability in this instance. The entire yearly cost can be computed using equation (5.9) [24]. And also The CO_2 capture cost is an additional metric utilized in economic research, and it can be defined as in equation (5.10) [11].

$$Total\ annual\ cost \left[\frac{€}{yr}\right] = Annualized\ CAPEX \left[\frac{€}{yr}\right] + Annualized\ OPEX \left[\frac{€}{yr}\right] \quad (5.9)$$

$$CO_2\ capture\ cost \left[\frac{€}{t}\right] = \frac{Total\ annual\ cost \left[\frac{€}{yr}\right]}{CO_2\ removal\ rate\ \left[\frac{t}{yr}\right]} \quad (5.10)$$

The estimated CAPEX in section 5.1 takes the project's tenure into consideration. Therefore, annualization of CAPEX is required when estimating the total annual cost. To determine annualized CAPEX, equations (5.11) and (5.12) are utilized [24].

$$Annualized\ CAPEX \left[\frac{€}{yr}\right] = \frac{CAPEX}{Annualized\ fator} \quad (5.11)$$

$$Annualized\ factor = \sum_{i=1}^{n} \frac{1}{(1+i)^n} \quad (5.12)$$

Time is a crucial factor for every project. Therefore, it should be taken into consideration to assess the project's economic aspects. According to several studies, the estimated lifespan of

CO_2 capture plants is 20 years. The value of money fluctuates over time due to inflation. Furthermore, there is ongoing development of CO_2 removal facilities that are being created for a significant duration. Therefore, it is logical to incorporate fluctuations in financial value during the duration of the project. The interest rate level varies from 7% to 14%. The parameter (i) in this project is considered to have a value of 7.5% in this study [36].

5.4 Total plant cost estimation using Nazir-Amini technique

To compare cost estimation with a prior approach, the Nazir-Amini method is regarded as the preferred method. The methodology for estimating the total plant cost (TPC) using this approach is detailed in Table 5.4. The cost allocation for process contingencies is influenced by the development level of the technology used for the capturing procedure which is represented by the value of "μ" in the table. "μ" is allocated the value 10 for the MEA process, which is regarded as commercially viable. In this project, to calculate the project contingency which is between 15 and 30 % assumes the value of 20 % [14], [37].

This method utilizes BEC (Base Erected Cost) as the foundation for calculating total plant cost (TPC). To derive a TPC that could be juxtaposed with the one acquired via the EDF method; the current investigation computed BEC utilizing the Aspen In-plant cost estimator while making the necessary adjustments for the year index [14].

Table 5.5: Methodology to estimate TPC by Nazir-Amini method [37]

Component	Definition
Bare Erected Cost (BEC)	Sum of installed cost of equipment
Engineering Procurement Construction Costs (EPCC)	10% OF BEC
Process Contingency	μ% of BEC
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)
Total Contingencies	Process Contingency + Project Contingency
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies

6 Results and Discussion

In this chapter, results for both the base case and the vapor recompression scenario are presented, and an evaluation of the optimum vapor recompression performance is also carried out.

6.1 Base case evaluation

The entire cost (CAPEX and OPEX) is estimated using the technique, equations, and data provided in chapter.

6.1.1 Total cost

Appendix C shows the CO_2 capture plant's total CAPEX and OPEX cost for the base case.

The absorber costs the most equipment at 29.31 million euros (45.5 % of the total CAPEX). Additionally, the reboiler and heat exchanger are costly components, costing 12.44 M€ (19.3 % of total CAPEX) and 9.64 M€ (15 % of total CAPEX), respectively. In figure 6.1, the CAPEX cost for each piece of equipment relative to the total CAPEX is displayed.

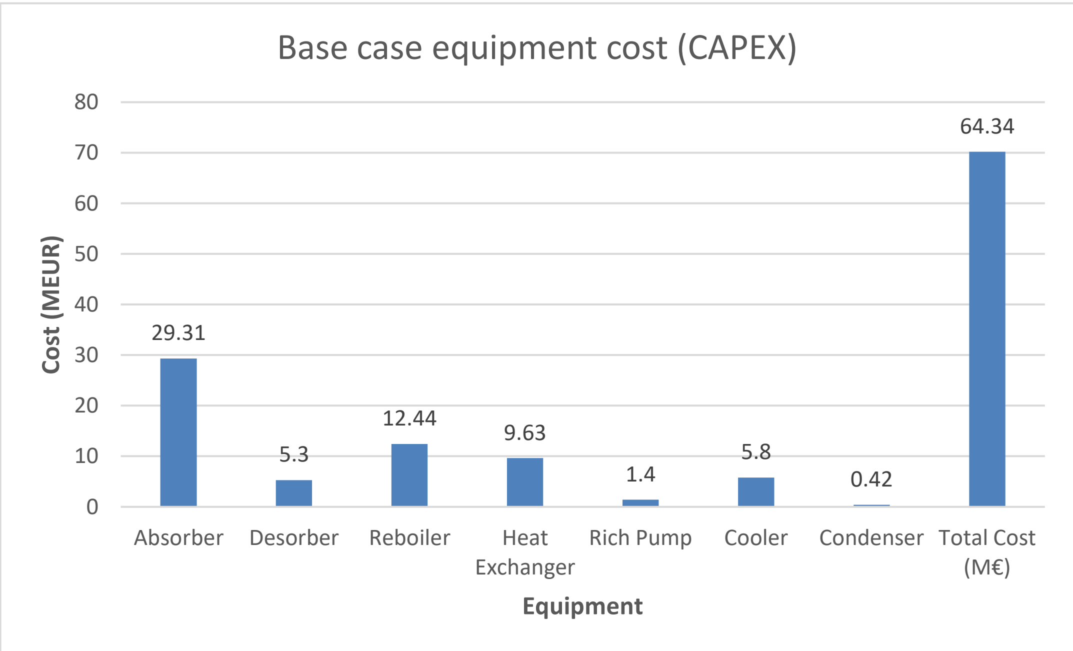

Figure 6.1: Total CAPEX and the operation cost for the vapor by EDF method

As it is shown in tables in Appendix C and figure 6.2, the reboiler has almost all the OPEX cost with 24.98 M€ (75.7 % of the total OPEX). The reason is the high duty used in the reboiler, which is 7.194E+08 (kJ/h) in the base case. A small amount of OPEX is also for the electricity price of rich pumps 0.139 M€, and the OPEX cost of MEA solution is calculated at about 3.8 M€.

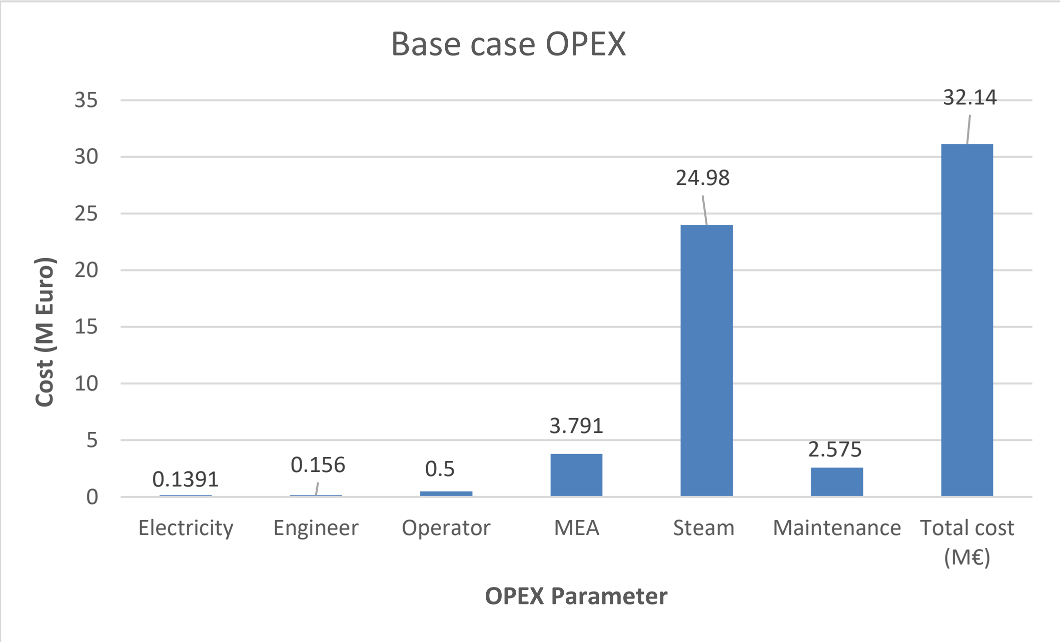

Figure 6.2: Total OPEX and the operation cost for the base case

6.2 Vapor recompression evaluation

The overall expenditure (CAPEX and OPEX) for the vapour recompression scenario is computed using the data from Aspen HYSYS and Aspen In-Plant, in addition to the procedure and equations described in Chapter 5.

6.2.1 Total cost

The heat exchanger and absorber are the most expensive pieces of equipment, costing approximately 36.5 and 29.31 million euros, respectively. Furthermore, the compressor, which plays a crucial role in the vapor recompression scenario, is an expensive component, with a price tag of 24.17 M€ (equivalent to 20% of the total CAPEX). Figure 6.3 presents the CAPEX cost of individual equipment components, including the flash separator, desorber, rich and lean pumps, and cooler, in relation to the total CAPEX.

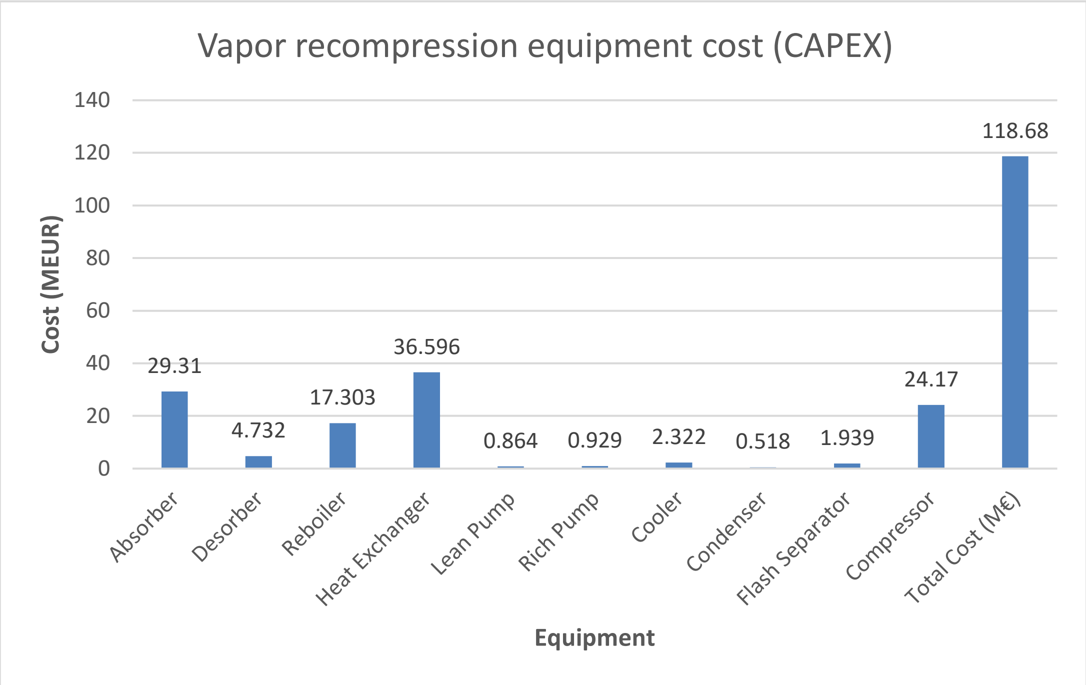

Figure 6.3: Total CAPEX and the operation cost for the vapor recompression by EDF method

The steam in reboiler comprises over half of the OPEX costs, 16.98 million euros, as shown in table C1 of Appendix C and figure 6.4. This represents a reduction in comparison to the base case. As a result of the compressor being added in the vapor recompression scenario, electricity consumption rises and reaches 4.7 million euros. The MEA price, however, was reduced by 1.89 million euros.

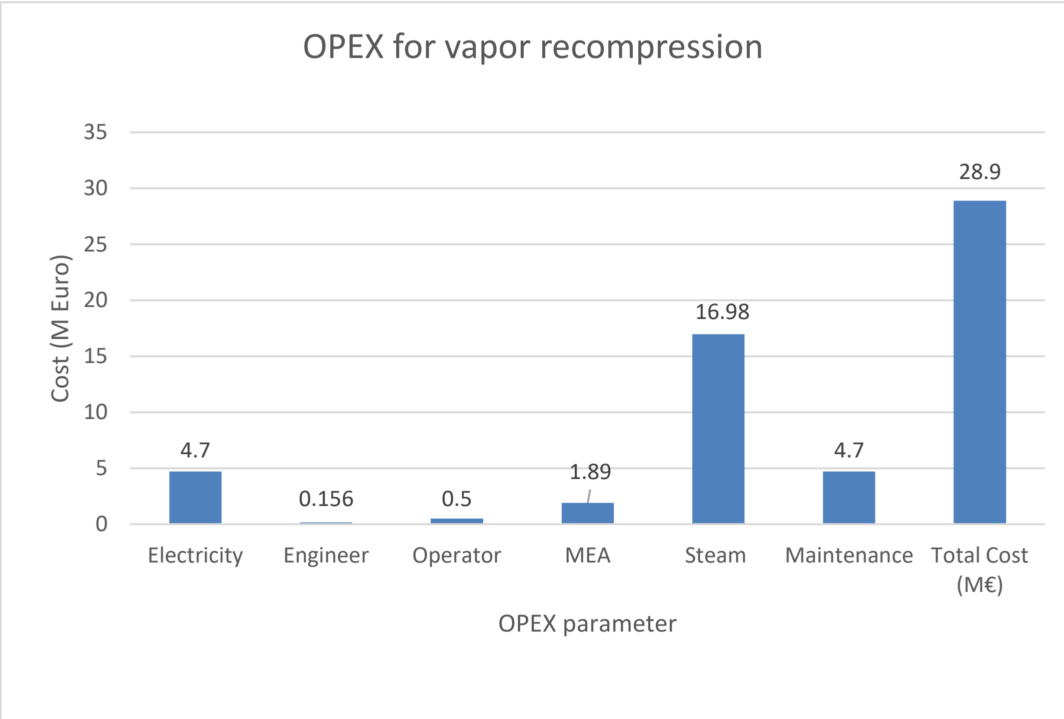

Figure 6.4: Total OPEX and the operation cost for the vapor recompression.

6.3 Comparison between the EDF and Nazir-Amini methods

Initially, the calculation of total plant cost is done by using Nazir-Amini, then a comparison between Nazir-Amini and the EDF methods. The Bare Erected Cost for the vapor recompression study is illustrated in Figure 6.5. By utilizing the EDF method and Aspen In-plant Cost Estimator V12, the BEC was determined. The total cost of equipment as determined by the EDF method is 84.6 M€.

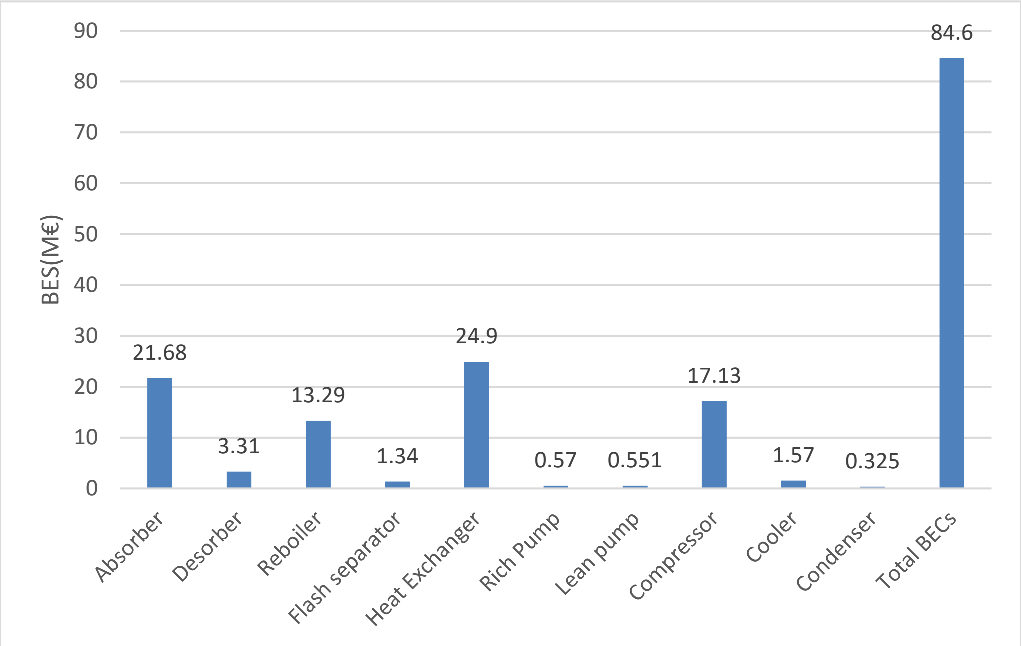

Figure 6.5: BEC for the vapor recompression case using EDF

The TPC was determined using the Nazir-Amini technique. To calculate TPC, the Aspen In-plant cost estimator provides the BEC. For the purpose of cost estimations, a 20% average project contingency was assumed. The data presented in Table 6.1 displays the estimated total plant cost derived from the EFD method.

Table 6.1: TPC for vapor recompression Scenario applying Nazir-Amini Method

Component	Definition	Cost (M Euro)
Bare Erected Cost (BEC)	Sum of installed cost of equipment	84.6
Engineering Procurement Construction Costs (EPCC)	10% OF BEC	8.4
Process Contingency	µ% of BEC	8.4
Project Contingency	15-30 % of (BEC + EPCC + Process Contingency)	20.3
Total Contingencies	Process Contingency + Project Contingency	28.7
Total Plant Costs (TPC)	BEC + EPCC + Total Contingencies	121.8

The TPC estimated using the EDF and Nazir-Amini methods is compared in the vertical chart 6.6 below.

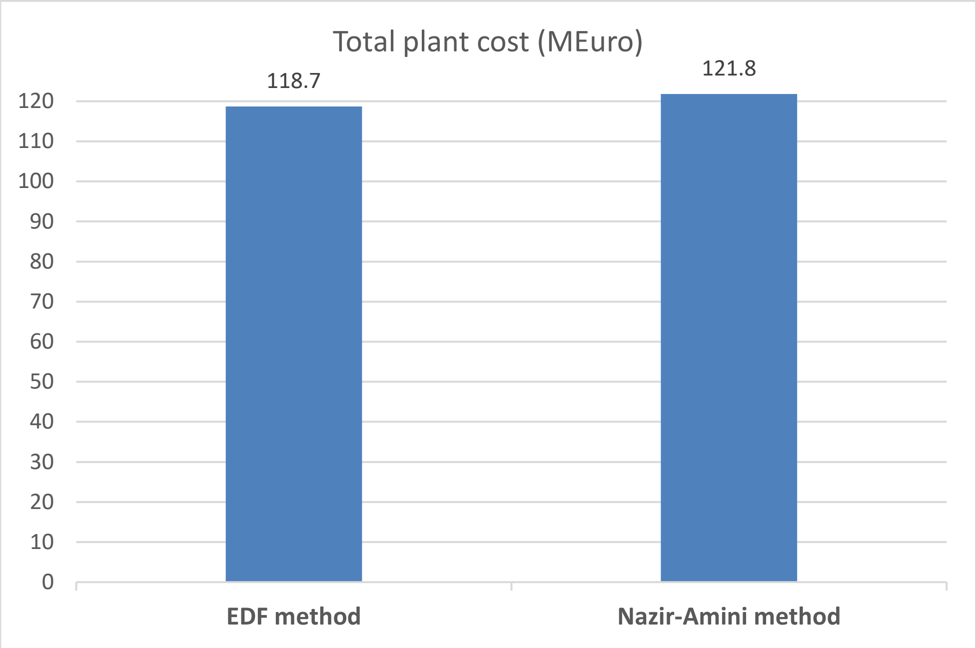

Figure 6.6: Analysis of TPC Calculation Methods

6.3.1 Advantages and disadvantages of the EDF and Nazir-Amini methods

The EDF technique modifies installation factors based on the cost of each piece of equipment, preventing expensive equipment from being overestimated and less expensive equipment from being underestimated. Because of this sensitivity to equipment expenses, total cost estimates can be more accurate. The EDF regards each individual piece of equipment as an independent project, offering a comprehensive analysis of the specific contribution of each item to the overall capital cost. Such a detailed level of analysis aids in identifying the main factors that contribute to costs, allowing for optimization. The EDF technique provides a completely free method for doing technical and economic assessments, compared to using costly software like Aspen Process Economic Analyzer [14].

The Nazir-Amini technique only estimates the Total Project Cost (TPC) and cannot be used for the assessment of each piece of equipment’s installation prices. In addition, The Nazir-Amini technique requires the computation of the Bare Erected Cost before calculating TPC; in this study, the EDF is employed to compute BEC. Moreover, Inaccurate assessment of the project's contingency range can result in less precise estimates using the Nazir-Amini method in comparison to the EDF. It facilitates cost computations by considering the minimum and maximum limits of the project contingency range, thereby offering valuable insights into possible fluctuations in costs.

6.4 Comparison between Base case and vapor recompression

The overall capital expenditure (CAPEX) and operational expenditure (OPEX) expenses for the basic scenario, using the values for the year 2021, amount to 64.34 million Euros and 32.14 million Euros per year, respectively. The absorber is the most expensive piece of equipment in this scenario, accounting for 29.31 M€ (45.5 % of the total CAPEX). The reboiler and the heat exchanger are similarly pricey pieces of equipment, with 12.44 M€ (19.3 % of the total CAPEX) and 9.36 M€ (15 % of the total CAPEX), respectively. It should be mentioned that in this research floating head shell and tube exchanger is selected.

However, the total capital expenditure (CAPEX) and operational expenditure (OPEX) for the vapor recompression scenario in 2021 amount to 118.68 M€ and 28.9 M€/yr, respectively.

For a reasonable cost optimization of the CO_2 capture plan, it is important to compare the base case with the vapor recompression scenario in the total yearly cost for 20 years. The entire annualized base case cost is computed as 38.45 million euros, whereas the corresponding figure for the vapor recompression scenario amounts to 40.6 million euros.

Consequently, the comparison between these two scenarios demonstrates that the total purchased cost (CAPEX) is more cost-effective in the basic case than in the vapor recompression scenario. However, the operating expenditure (OPEX) is lower in the vapor recompression scenario compared to the base case. Also, the entire annualized base case cost has a lower cost than the vapor recompression based on our simulation and cost assessment. Indeed, our research indicates that the basic scenario is a more cost-effective option.

6.5 Optimizing flash separator in the vapor recompression case

In this section, the impact of the pressure on the flash separator is investigated concerning OPEX and CAPEX, to achieve the economic optimum performance of flash. Moreover, another purpose is to assess the expense of CO2 capture during the vapour recompression and the base case. Flah pressure in this simulation assumes 100 kPa pressure in this work. But for sensitivity analysis in this project, the range of pressure variation between 80 and 120 kPa has been considered.

6.5.1 Sensitivity study of flash pressure on CAPEX

Figure 6.7 shows the calculated CAPEX for the vapor recompression case. At the pressure of 120 kPa, CAPEX reaches the minimum value that is 98.86 M€. Pressure variation significantly impacts both the heat exchanger and compressor. The main reason for the capital expenditure is associated with modifying the area of the lean/rich heat exchanger and altering the capacity of the compressor. At 120 kPa, this area decreases from 27,440 m^2 at 100 kPa to 17510 m^2, and the compressor capacity declines from 4137 kW to 2217 kW.

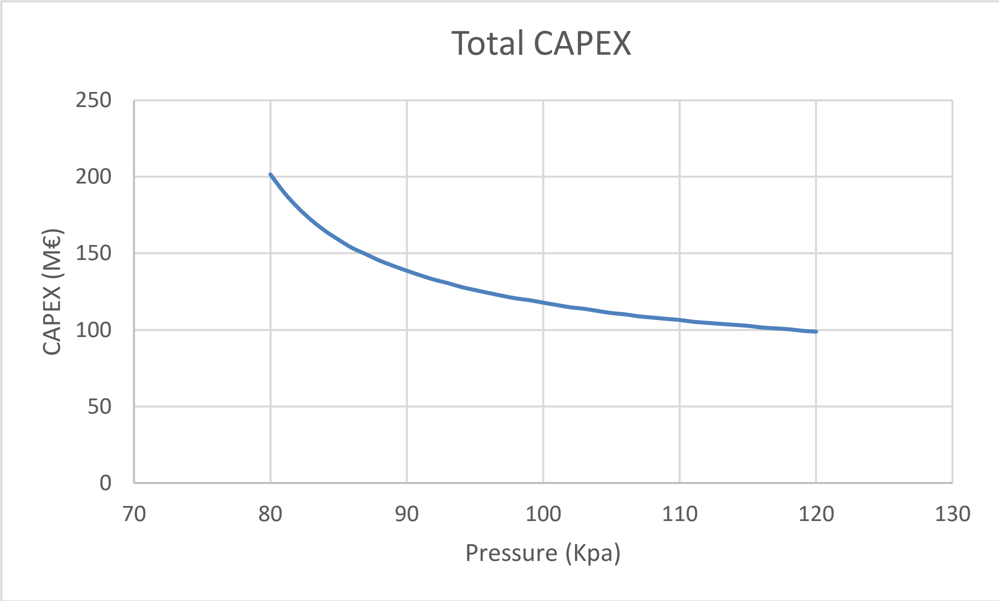

Figure 6.7: CAPEX by changing pressure in the flash

Figure 6.8 shows the heat exchanger purchased price in different pressures. The purchased cost of heat exchanger at 100 kPa is 36.58 M€, this amount decreases 23.41 M€ at 120 kPa. The main reason for this decline is the heat transfer area.

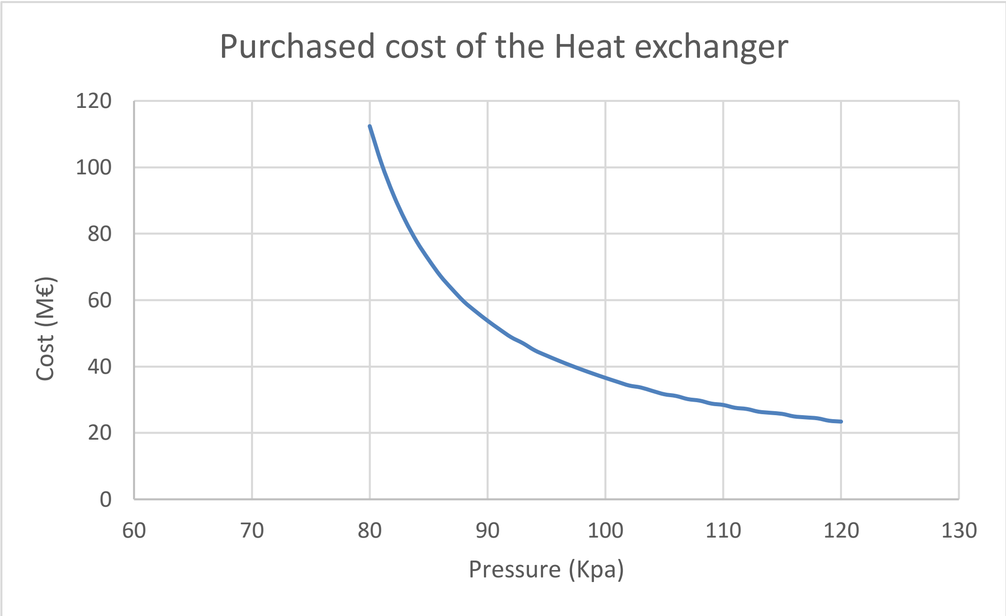

Figure 6.8: Heat exchanger price in different pressure

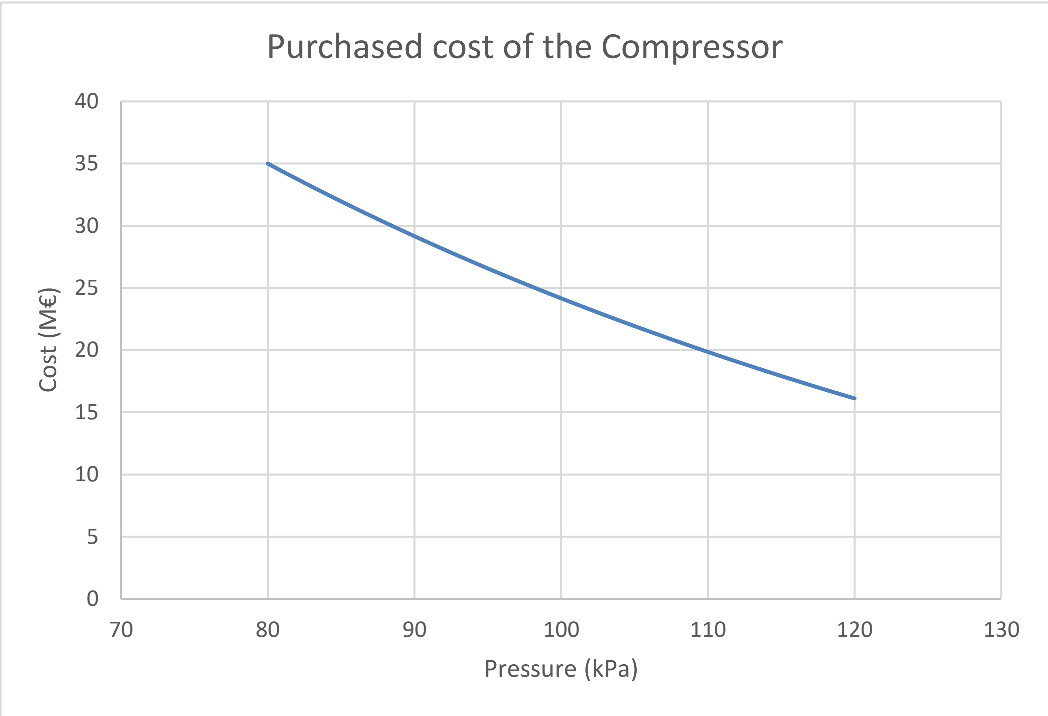

Figure 6.9: Installation compressor price in different flash pressure

Figure 6.9 illustrates the cost of a compressor at various pressures. The compressor directly influences the CAPEX cost. At 100 kPa, the cost of the compressor was 24.16 M€; at 120 kPa, this amount decreases by 16.11 M€.

6.5.2 Sensitivity study of flash pressure on OPEX

As can be seen in figure 6.10, the flash pressure directly affects the compressor and reboiler. At 120 kPa, the compressor operates more efficiently and results in lower power consumption. While this pressure causes to increase steam consumption in reboiler.

At 120 kPa, the OPEX cost reaches 28.39 million euro per year, it presents a reduction of about 0.56 M€. This amount is insignificant compared to the changing pressure in the CAPEX.

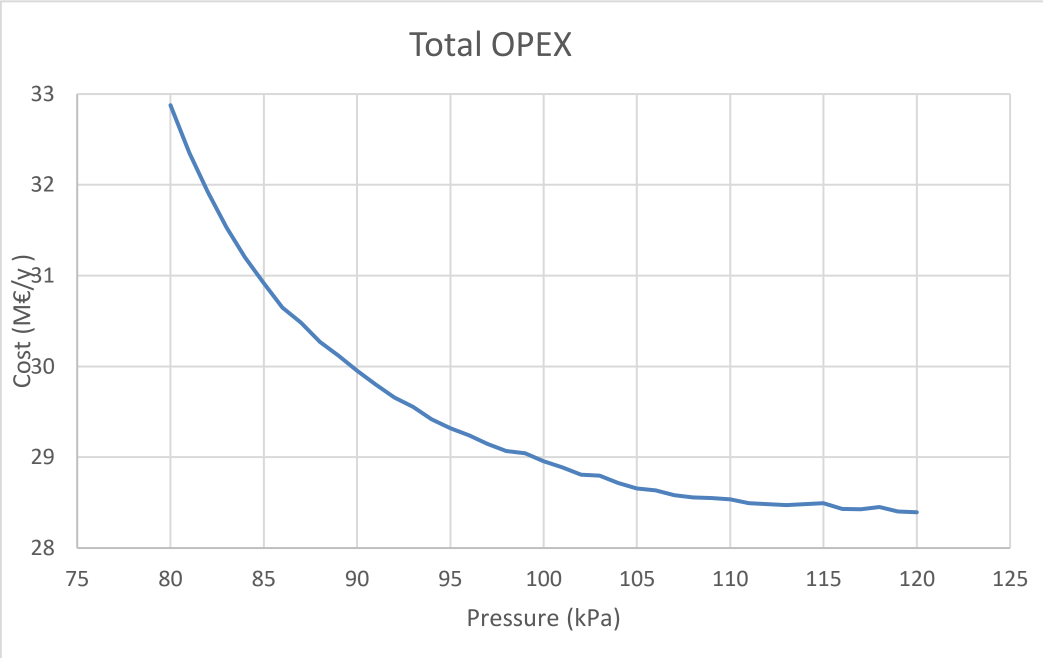

Figure 6.10: Total OPEX by changing pressure in the flash separator

The energy consumption in the compressor and reboiler are two important factors in determining the amount of OPEX. As can be seen in the figure below (6.11), reboiler steam consumption was 16.95 M€/y at 100 kPa, although after increasing pressure at 120 kPa, it has 2 M€/y rise in consumption of steam. In contrast, energy price in the compressor decreases from 4.5 M€/y at 100 kPa to 2.4 M€/y at 120 kPa.

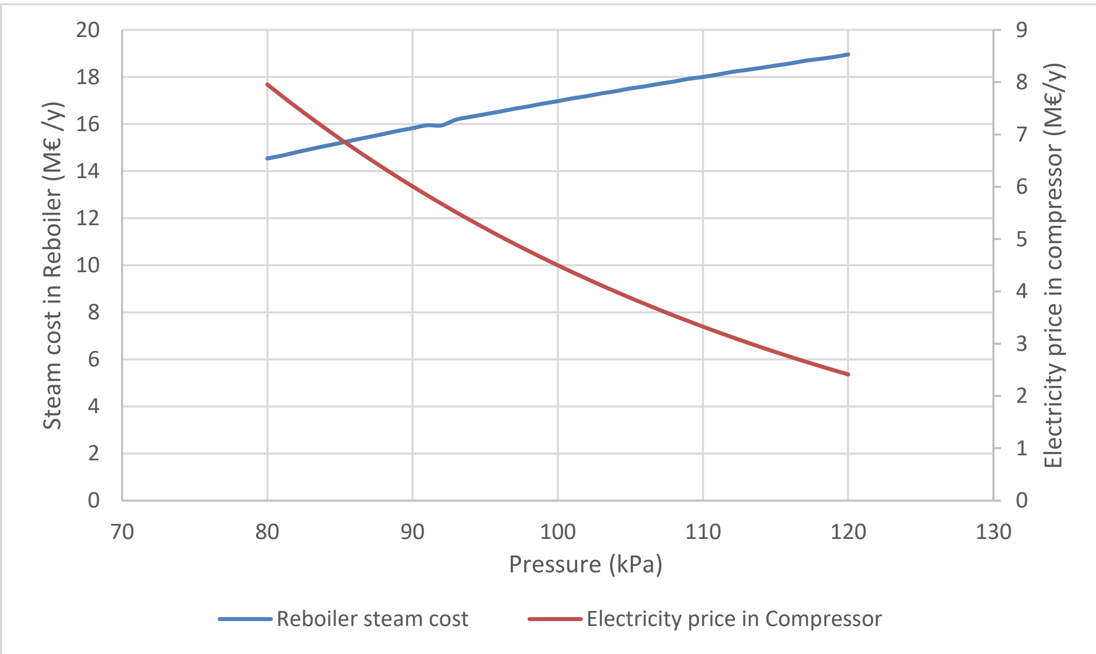

Figure 6.11: Reboiler steam cost and electricity in compressor in different pressure

The objective of this study is to assess the economic potential of vapor recompression. Therefore, our analysis and optimization of the process will focus on the vapor recompression scenario in the following sections.

6.6 Comparison between CO_2 capture cost in Base case and vapor recompression

Figure 6.12 displays the various costs of carbon capture in three different situations. According to the figure, the carbon capture cost via vapor recompression at 120 kPa is lower than the cost of carbon capture in the base case and vapor recompression at 100 kPa. The CO_2 capture cost (€/ton CO_2) is influenced by two factors: the overall annual cost (€/year) and the CO_2 removal rate (ton CO_2/year), as stated in Equation (5.10).

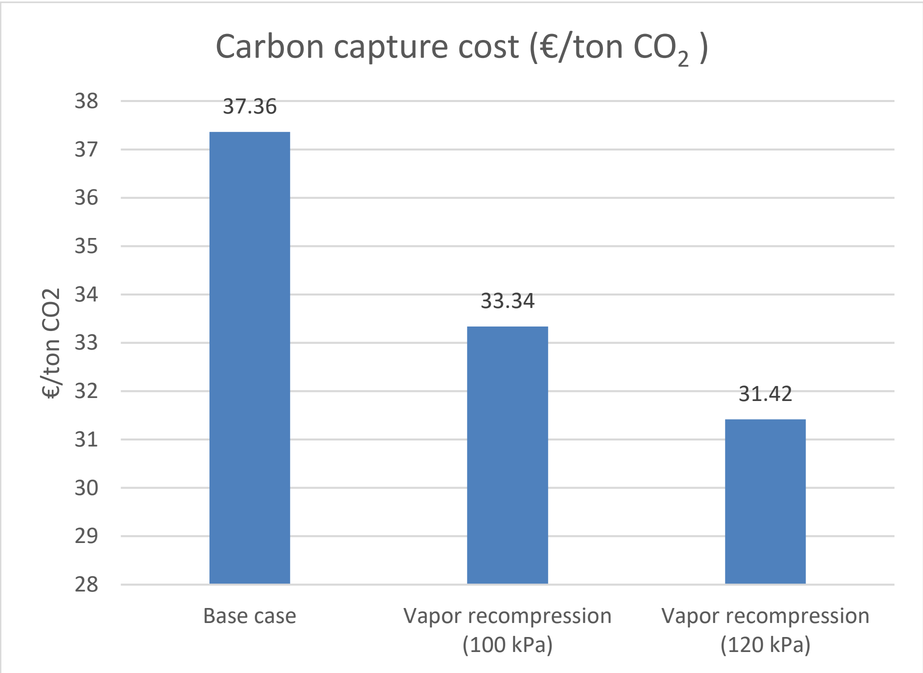

Figure 6.12: CO_2 Capture Cost in Different Scenarios

Also, in the below figure 6.13, the rate of CO_2 capture in different pressures can be seen. It shows at 120 kPa, the rate of CO_2 Capture is the highest in this specific pressure rate and is 151542 kg/h.

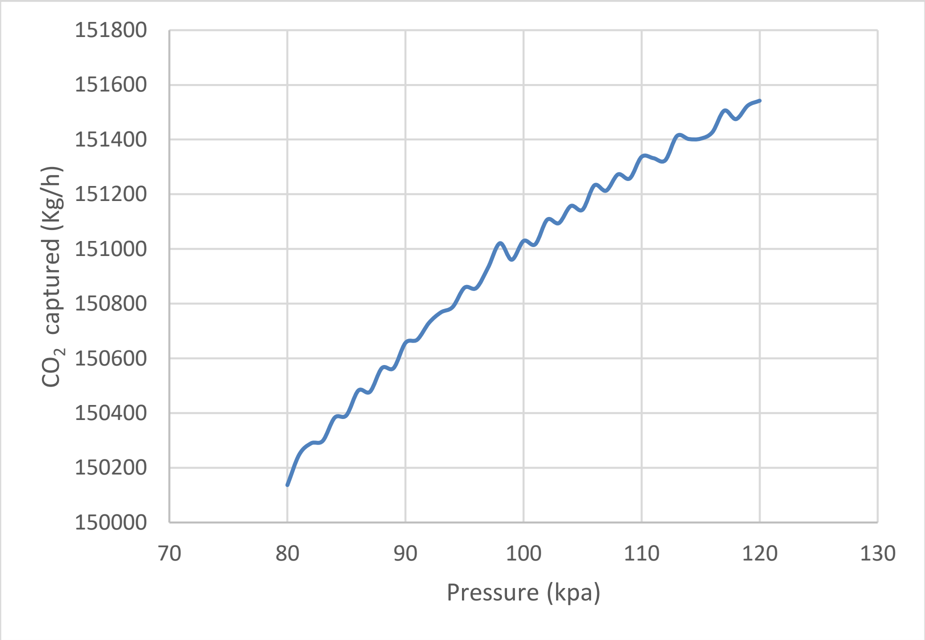

Figure 6.13: CO_2 Captured rate in different pressures of the vapor recompression

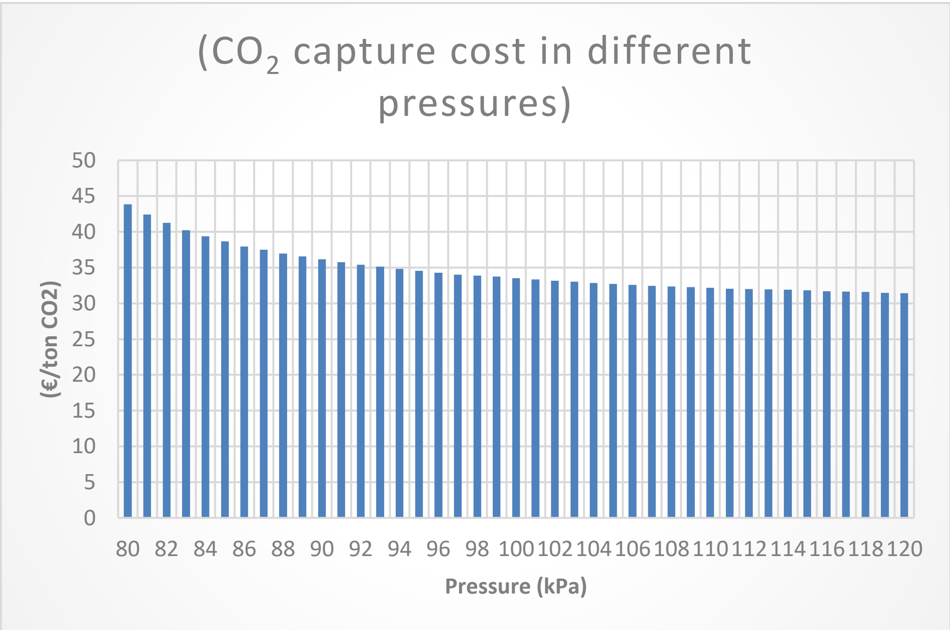

Figure 6.14: CO_2 capture costs variation under different pressure conditions

Figure 6.14 shows the CO_2 capture cost in the pressure range from 80 to 120 kPa. It is illustrated that the CO_2 Capture cost decreased with an increase in the flash pressure on the flash due to the CO_2 capture rate and total annual cost in the different pressure.

6.7 Total plant cost at the optimum flash pressure

In figure 6.15, it shows the compression of total installed costs in the optimum flash pressure (120 kPa) and initial flash pressure (100 kPa). The overall installed cost drops from 118.68 to 98.8 M€ when the flash separator is optimized; this optimization results in a reduction of about 20 M€ in total CAPEX.

The compressor and heat exchanger were the two significant pieces of equipment that were essential to this decrease. As the following figure illustrates, a significant contribution to this evaluation was made by the heat exchanger (13 M€) and the compressor (about 8 M€ reduction).

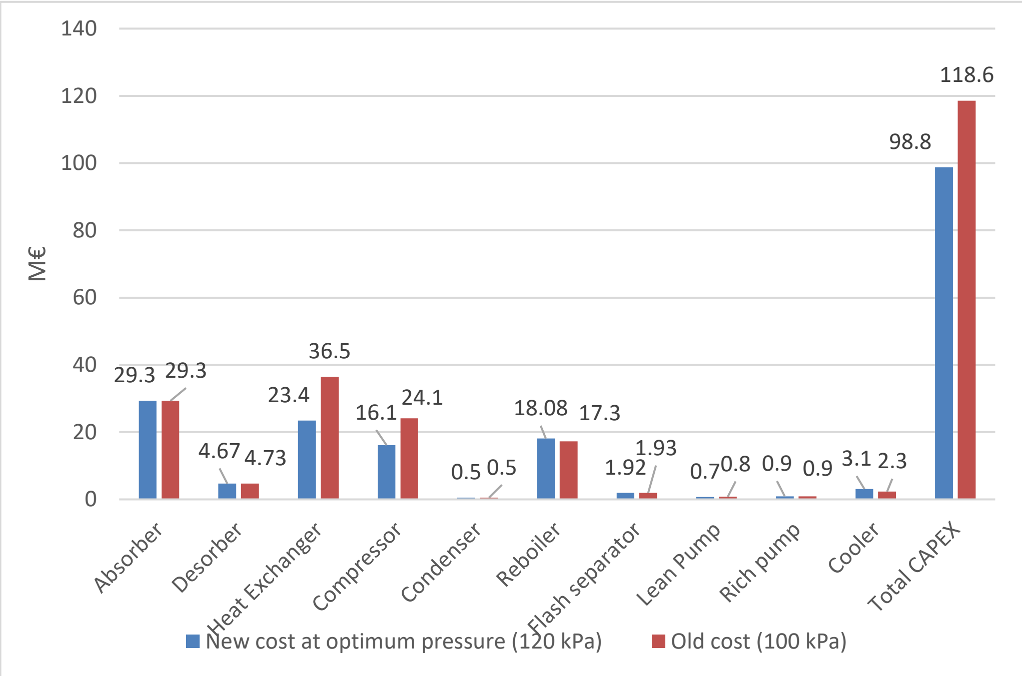

Figure 6.15: TPC comparison for optimum flash separator and the initial installation equipment cost

Consequently, the annualized CAPEX of the vapor recompression case in the optimal flash pressure is 9.7 M€/yr and 11.64 M€/yr for the initial flash pressure in the same case based on equation (5.11).

6.8 Comparison between the Base case and vapor recompression at optimum pressure

As stated in section 6.4, the total annualized costs for the base case scenario come to 38.45 M€/yr. For the vapour recompression scenario at the optimum pressure (120 kPa), the total capital expenditure (CAPEX) and operating expenditure (OPEX) are 98.8 M€ and 28.39 M€/yr, respectively. Thus, over 20 years, the total annualized cost in this scenario is 38 M€/yr. This optimization saves 2.6 M€/yr than vapor recompression (100kPa). This investigation shows that the vapor recompression case in the optimum pressure (120 kPa) is the most cost-effective way for this research.

6.9 Comparison of current work with previous studies

In this session, the effect of flashing pressure in CAPEC and OPEX is compared with previous studies.

6.9.1 Effect of flashing pressure on CAPEX

A key objective is to determine the total CAPEX of the plant. The optimization of the vapor recompression scenario was conducted by altering the pressure of the flash separator. The lean/rich heat exchanger and compressor were shown to have a significant impact on the capital expenditure (CAPEX) and total cost. An analysis of the heat exchanger cost, compressor cost (CAPEX), and total cost in relation to previous investigations would give helpful perspectives into this study, as these two pieces of equipment are identified as the 1st and 3rd major capital cost drivers.

Some prior research and a comparison of the current study are presented in Table 6.2.

Table 6.2: Compression of current work and previous studies

Study	**CO_2 capture rate (%)**	**Heat ex cost (M€)**	**Heat ex. Contribution in TPC**	**Compressor Cost (M€)**	**Compressor Contribution in TPC**
Present work (Vapor rec) (100 kPa)	92.6	36.59	1st	24.17	3rd
Present work (120kPa)	92.6	23.41	2st	16.11	3rd
Aromada et al. [11]	85	38.95	2nd	14.61	3rd
Aromada et al.[38]	90	51	1st	44	2nd
Aromada et al. [39]	85	41.5	1st	-	-
H. Ali et al. [30]	85	25	1st	-	-

6.9.2 Effect of flashing pressure on OPEX/Energy

Another important factor in the cost optimization in a process is the operating cost (OPEX). When the pressure of the flash separator is changed between 80 and 120 kPa, the requirement of compressor into the electricity is varied highly from 7.95 M€/y at 80 kPa to 2.41 M€/y. while the steam in the reboiler has a different trend and it increases from 14.54 M€/y at 80 kPa to 18.94 M€/y. This investigation shows the total OPEX at 120 kpa is the lowest operating cost with 28.39 M€/y.

According to Øi [6] , The optimization involved a column height of 15 meters, a minimum temperature approach of 10 K, and an estimate of the cost-optimal flash pressure at 120 kPa. In this pressure, equivalent energy consumption [MJ/kg CO_2] had the lowest amount of steam consumption about 3.38 (MJ/kg CO_2).

Fernandez et al. [16] states that based on the variables that are being studied with CO_2 capture 90 percentage and solvent 30 wt % MEA, the optimum operating pressure for scenarios is determined to be 1.2 bar. Furthermore, it was shown that optimizing operational conditions alone through energy analysis might provide different outcomes.

6.10 Uncertainties

Unknown factors in the calculations and data utilized in this study have a notable effect on both the calculated equipment price and operating cost. This section primarily addresses uncertainties arising from various components utilized in the calculation process, as well as those associated with the use of Aspen HYSYS and Aspen In-Plant cost estimator.

- Specifications and assumptions can have a substantial effect on the estimated cost. For example, the cost of the absorber and desorber is directly influenced by the packaging type and the number of stages selected. In a similar vein, the heat transfer area in my research was influenced by attributes such as the overall heat transfer coefficient and ΔT_{min} of the heat exchangers, which were both 10 ^oC. Placing the absorber (0.25) and desorber (1) at a constant Murphree efficiency could introduce inaccuracies into the amine circulation flow calculations, consequently affecting the projected cost of the facility [14].
- The power law method was utilized to forecast substantial uncertainties in the cost assessment associated with the scale-up factor. In this the task, the equipment scaling exponent assumed to be 0.65 per piece of equipment introduced an element of ambiguity into the calculation.
- Concerning the equilibrium calculations in the assimilate and desorb columns, uncertainties are also presumed to exist. The calculation of the equilibrium between vapor and liquid is performed in Aspen HYSYS. Combining equilibrium models and reaction kinetic models, the Aspen Plus simulation programme provides models that more precisely simulate the practical performance, thereby increasing the degree of precision [24].
- The equipment costs were modified to reflect the total installed costs utilizing the EDF table. It is conceivable that these coefficients may undergo periodic fluctuations contingent upon the market and geographical location. Hence, it would be more suitable to compare the outcomes of alternative methodologies.
- The Nazir-Amini methodology, project contingency is assumed to be 20 %. This consumption can have a direct effect on TPC which range is from 15 to 30 %.
- One of the parameters which can have an impact on the total CAPEX is the type of heat exchanger. In this project, a floating head shell and tube exchanger was selected. Other types of heat exchangers such as U-tube shell and tube exchanger or fixed tube sheet shell and tube exchanger can have a different price.

6.11 Some suggestions for future work

However, due to a lack of time and knowledge, a great number of various operations and simulations have been postponed till the future. Future work can focus on doing a more in-depth study of the models and approaches. For the purpose of enhancing the resilience and accuracy of simulation and cost estimates, the following recommendations are recommended for further research in this area:

- Examine the compressed vapor return stage in the desorber to see which step provides the lowest energy output.
- Murphree efficiency in the absorber stage can have vital impact on the absorber performance and be considered in the future work.
- Inlet pressure in flue gas to check CO_2 removal efficiency and performance of compressor and reboiler.
- Assessing ΔT_{min} heat exchanger which is an important factor to consider in CO_2 capture plant cost estimation.
- Check the temperature of the reboiler to evaluate the performance of the flash separator.
- Change the type of heat exchanger.
- Change MEA flowrate.

7 Conclusion

The process was simulated and implemented using Aspen HYSYS V12. The equipment was dimensioned based on the simulation's outcomes. The Aspen In-Plant Cost Estimator V12 was utilized to estimate equipment acquisition costs based on 2019 cost data. After that, the CAPEX was calculated using the EDF technique, and the costs were adjusted to the cost index factor for 2021. The vapor compression configuration was then added to the base scenario, and CAPEX was computed. OPEX values for each scenario were computed. The total CAPEX and OPEX in the base case were 64.34 M€ and 32.14 M€/year respectively and these amounts for the vapor recompression case at 100 kPa were 118.68 and 28.9 M€/year. Therefore, the purchased cost in the base case and the operating expenses in the vapor recompression (100kPa) are cost effective.

The aim of this work was on evaluating the only influential parameter in the flash separator, pressure, to determine the ideal value for economic performance. The flash separator's sensitivity analysis revealed that at 120 kPa, CAPEX achieves its lowest point, which is 98.86 M€. The purchased cost of heat exchanger and compressor are the key reasons for this total CAPEX cost reduction. Hence, the purchased cost of the heat exchanger was determined to be 36.58 M€ at a pressure of 100 kPa and decreases to 23.4 M€, at 120 kPa and the compressor from 24.16 to 16.11 M€. In addition, the effect of pressure on OPEX revealed that this quantity reaches its minimum at 120 kPa, and that the electricity in the compressor and the steam in the reboiler were two crucial parameters in reducing OPEX. After optimizing flash separator at 120 kPa, CAPEX and OPEX reached 98.8 M€ and 28.3 M€/year respectively.

The cost of CO_2 capture is an additional economic analysis parameter utilised in most prior research. The base scenario resulted in 37.36 (€/ton CO_2), while the vapour recompression case at 100 kPa was found 33.34 (€/ton CO2) and vapour recompression in the optimum pressure 120 kPa was 31.42 (€/ton CO_2). Consequently, the findings show that the vapour recompression at an optimum pressure of 120 kPa is the most cost-effective option.

The EDF and Nazir-Amini methods are configured to evaluate in the vapor recompression case in order to compare different cost estimates. BEC is required to use the Nazir-Amini method. As a result, it is employed the EDF method to compute the BEC in the Nazir-Amini which amount was 84.6 M€. Considering 20 % project contingency and 10% process contingency, the total plant cost in the Nazir-Amini methodology was 121.8 M€, while in the EDF method it was 118.7 M€. The comparison between the EDF and the Nazir-Amini demonstrated that the EDF method provides a more precise overall evaluation through including equipment cost adjustments into its cost estimation process. In contrast, the Nazir-Amini approach focuses solely on Total Project Cost (TPC) without detailing individual equipment expenses, relying on Bare Erected Cost (BEC) computation. While EDF enables comprehensive analysis and cost optimization, Nazir-Amini provides insights into cost fluctuations but lacks equipment-specific breakdowns. Overall, EDF's granularity enhances accuracy, while Nazir-Amini offers broader cost perspectives.

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Appendices

Appendix A: Project description

Appendix B: Detailed factor table

Appendix C: CAPEX and OPEX for the base case and vapor recompression applying the EDF and the Nazir-Amini

Appendix D: OPEX calculation-Base case and vapor recompression

Appendix E: CO_2 capture cost, CO_2 removal efficiency

Appendix A - Project description

University of
South-Eastern Norway

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

FMH606 Master's Thesis

Title: Simulation and Cost estimation of CO_2 capture including Vapour Recompression

USN supervisor: Lars Erik Øi, Cosupervisor: Solomon Aromada, Equinor

External partner: Possibly SINTEF (by Nils Eldrup) or TCM Mongstad (by R.K. Putta)

Task background:

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

Task description:

The general aim is to develop models in Aspen HYSYS and Aspen Plus

1. Literature search on methods for cost estimation of amine based CO_2 capture including Vapour Recompression based on process simulation.

2. Process simulation of a traditional CO_2 capture process based on absorption into a monoethanol amine solvent. Cost estimation based on Aspen In-plant followed by a detailed factor method and possibly based on Aspen Economy Evaluator.

3. Comparisons of different cost estimates on the standard amine based process and the vapour recompression based process, and evaluate the economical potential of the vapour recompression process.

4. Evaluation of advantages and drawbacks with the different cost estimation methods.

Student category: PT or EET

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

Practical arrangements:

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

Supervision:

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

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

Appendix B: Detailed factor table (2020)

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

The purchased cost of all equipment for the base case with applying EDF method (Total CAPEC)

Equipment	Number of units	Equipment cost for one unit in 2019	CAPEX for one unit in 2021 (EDF method)	CAPEX for all identified units in 2021
Absorber	1	12,930,000 €	29,310,000 €	29,310,000 €
Desorber	1	1,672,00 €	5,340,000 €	5,340,000 €
Reboiler	7	482000 €	1,777,000 €	12,439,000 €
Heat Exchanger	8	326,600 €	1,204,000 €	9,632,000 €
Rich Pump	1	284,500 €	1,412,000 €	1,412,000 €
Cooler	5	315,800 €	1,165,000 €	5,825,000 €
Condenser	1	68,600 €	421,800 €	421,800 €
Total cost (M€)		14.40		64.34

Total OPEX for the base case

Parameter	**OPEX for one year**
Electricity	139100 €
Engineer	156000 €
Operator	500000€
MEA	3791000€
Steam	24980000 €
Maintenance	2575000 €
Total Cost (€)	31,141,000 €
Total cost (M€/yr)	32.14

The purchased cost of all equipment for the vapor recompression case in EDF method (Total CAPEX)

Equipment	Number of units	Equipment cost for one unit in 2019	CAPEX for one unit in 2021 (EDF method)	CAPEX for all identified units in 2021
Absorber	1	12,930,000 €	29,310,000 €	29,310,000 €
Desorber	1	1,482,000 €	4,732,000 €	4,732,000 €
Reboiler	13	416,700 €	1,331,000 €	17,303,000 €
Heat Exchanger	28	354,300 €	1,307,000 €	36,596,000 €
Lean Pump	1	204,000 €	864,080 €	864,080 €
Rich Pump	1	184,700 €	929,200 €	929,200 €
Cooler	2	314,900 €	1,161,000 €	2,322,000 €
Condenser	1	99,000 €	518,400 €	518,400 €
Flash Separator	1	607,300 €	1,939,040 €	1,939,040 €
Compressor	1	10,770,000 €	24,170,00 €	24,170,000 €
Total cost (M€)		27.4		118.68

Total OPEX for the vapor recompression

Parameter	**OPEX for one year**
Electricity	4,703,000 €
Engineer	156000 €
Operator	500000 €
MEA	1,989,000 €
Steam	16,980,000 €
Maintenance	4,747,000 €
Total Cost (€)	28,980,000 €
Total cost (M€/yr)	28.98

BEC for the vapor recompression case by Applying EDF for Nazir-Amini method

Equipment	Number of units	Equipment cost for one unit in 2019	BEC for one unit in 2021 (Nazir-Amini)	BEC for all identified units in 2021
Absorber	1	21,680 ,500 €	21,680 ,500 €	21,680 ,500 €
Desorber	1	3,303,000 €	3,303,000 €	3,313,000 €
Reboiler	13	1,022,800 €	13,296,400 €	13,296,400 €
Flash separator	1	1,354,000 €	1,345,000 €	1,345,000 €
Heat Exchanger	28	888,170 €	24,888,000 €	24,889,000 €
Rich Pump	1	570,000 €	570,000 €	570,000 €
Lean pump	1	551,000	551,000 €	551,000 €
Compressor	1	17,137,000	17,137,000 €	17,137,000 €
Cooler	2	789,900 €	1,579,200 €	1,579,800 €
Condenser	1	325,200 €	325,200 €	325,200 €
Total cost (M€)		25.94		84.6

Procedure of CAPEX calculations for the Base case applying EDF

Parameter	**Description**	**Absorber**	**Desorber**
Packing height (m)	Simulation	10	6
Column height (m)	Simulation	25	15
Diameter (m)	Sizing	16.86	9.33
Sell material		SS-304	SS-304
Equipment cost per unit SS 2019 (kEuro)	Aspen In-plant cost estimator	12930	1704
Equipment cost per unit SS 2020 (kEuro)	convert to 2020	13176	1736.3
Equipment cost per unit CS 2020 (kEuro)	convert to CS	7529	993
Direct cost factor CS 2020	Detail factor Table 2020	2.84	3.63
Total installation cost factor	Detail factor Table 2020	3.762	4.66
Cost 2020 (KEuro)		28900	4543
Equipment cost per unit CS 2021 (kEuro)	convert to 2021	29310	5300

Procedure of CAPEX calculations for the Base case applying EDF

Parameter	Description	Reboiler	Heat Ex	Cooler	Condensor
Total heat transfer area (m2)	Sizing	6011	7321	4482	173.6
Max. Area per unit (m2)	Assumption	1000	1000	1000	1000
Actual number of units		7	8	5	1
LMDT	Sizing	27.7	36	38.93	96.56
Material		SS- 304	SS- 304	SS- 304	SS- 304
Equipment cost per unit SS 2019 (kEuro)	Aspen In-plant cost estimator	491.2	326.6	315.8	68.6
Equipment cost per unit CS 2020 (kEuro)	convert to 2020	491.2	332.8	321.8	69.91
Equipment cost per unit CS 2020 (kEuro)	convert to CS	280.7	190.2	183.9	39.95
Direct cost factor CS 2020	Detail factor Table 2020	4.92	4.92	4.92	8.54
Totall installation cost factor	Detail factor Table 2020	6.12	6.12	6.12	10.2
Cost 2020 (KEuro)		1718	1164	1125	407.7
Equipment cost per unit CS 2021 (kEuro)	convert to 2021	1777	1204	1165	421.8

Procedure of CAPEX calculation for the vapor recompression case

Parameter	Description	Absorber	Desorber
Packing height (m)	Simulation	10	6
Column height (m)	Simulation	25	15
Diameter (m)	Sizing	16.86	6.2
Sell material		SS-304	SS-304
Equipment cost per unit SS 2019 (kEuro)	Aspen In-plant cost estimator	12930	1482
Equipment cost per unit SS 2020 (kEuro)	convert to 2020	13176	1510
Equipment cost per unit CS 2020 (kEuro)	convert to CS	7529	862.9
Direct cost factor CS 2020	Detail factor Table 2020	2.84	4.19
Totall installation cost factor	Detail factor Table 2020	3.762	5.3
Cost 2020 (KEuro)		28900	4573
Equipment cost per unit CS 2021 (kEuro)	convert to 2021	29310	4732

Procedure of CAPEX calculation for the vapor recompression case applying EDF

Parameter	Description	Reboiler	Heat Ex	Cooler	Condensor
Total heat transfer area (m2)	Sizing	12091.7	27423.5	1778	248.1
Max. Area per unit (m2)	Assumption	1000	1000	1000	1000
Actual number of units		13	28	2	1
LMDT	Sizing	35.09	9.27	26.76	72.31
Material		SS- 304	SS- 304	SS- 304	SS- 304
Equipment cost per unit SS 2019 (kEuro)	Aspen In-plant cost estimator	416.7	354.3	314.9	99
Equipment cost per unit CS 2020 (kEuro)	convert to 2020	424.6	361.1	302.9	100.9
Equipment cost per unit CS 2020 (kEuro)	convert to CS	242.7	206.3	183.4	57.65

Direct cost factor CS 2020	Detail factor Table 2020	4.19	4.92	4.92	7.22
Totall installation cost factor	Detail factor Table 2020	5.3	6.12	6.12	8.69
Cost 2020 (KEuro)		1286	1263	1122	501
Equipment cost per unit CS 2021 (kEuro)	convert to 2021	1331	1307	1161	518.4

Procedure of CAPEX calculation for the vapor recompression case applying EDF

Parameter	Description	Flash separator	Rich pump	Lean Pump	Comprrssor
Important factors		D=5.9 m , H=14.9 m	p=87.27 kw	P=99.30kw	P= 4136 kw
Material		SS- 304	SS- 304	SS-304	SS-304
Equipment cost per unit SS 2019 (kEuro)	Apen In-plant cost estimator	607.3	184.7	204.3	10771.3
Equipment cost per unit CS 2019 (kEuro)	convert to 2020	618.9	188.2	208.2	11145.7
Equipment cost per unit CS 2019 (kEuro)	convert to CS	353.6	144.8	160.2	8286
Direct cost factor CS 2020	Detail factor Table 2020	4.19	5.89	4.92	2.56
Totall installation cost factor	Detail factor Table 2020	5.3	6.418	5.4	2.917

Equipment cost per unit CS 2021 (kEuro)	convert to 2021	1939	929.2	864.8	24169

The CAPEX of compressor and heat exchanger in different pressure range.

Pressure (kPa)	Compressor price (M€)	Heat exchanger price (M€)
80	35	112.41
81	34.37	101.14
82	33.75	91.84
83	33.14	84.09
84	32.55	77.59
85	31.96	72.20
86	31.38	67.29
87	30.81	63.26
88	30.25	59.54
89	29.70	56.56
90	29.16	53.80
91	28.62	51.24
92	28.10	48.83
93	27.58	47.02
94	27.07	44.88
95	26.56	43.30
96	26.07	41.81
97	25.58	40.40
98	25.10	39.06
99	24.63	37.79
100	24.16	36.58
101	23.70	35.42
102	23.25	34.30
103	22.80	33.68
104	22.36	32.64
105	21.93	31.64
106	21.50	31.13
107	21.08	30.19
108	20.66	29.73
109	20.26	28.84
110	19.85	28.42
111	19.45	27.58
112	19.06	27.21
113	18.67	26.39
114	18.29	26.06

115	17.91	25.74
116	17.54	24.97
117	17.18	24.68
118	16.82	24.40
119	16.46	23.67
120	16.11	23.41

The steam price and electricity price in different pressure range in

Pressure (kPa)	Reboiler steam cost (M€)	Electricity price in Compressor (M€)
80	14.53	7.95
81	14.65	7.73
82	14.80	7.52
83	14.93	7.31
84	15.06	7.11
85	15.19	6.91
86	15.32	6.72
87	15.44	6.54
88	15.57	6.35
89	15.70	6.18
90	15.82	6.00
91	15.94	5.83
92	15.94	5.67
93	16.18	5.51
94	16.30	5.35
95	16.41	5.20
96	16.52	5.05
97	16.64	4.91
98	16.75	4.77
99	16.86	4.63
100	16.97	4.49
101	17.08	4.36
102	17.18	4.24
103	17.29	4.11
104	17.39	3.99
105	17.51	3.87
106	17.59	3.76
107	17.70	3.64
108	17.80	3.53
109	17.91	3.43
110	17.99	3.32
111	18.10	3.2

112	18.21	3.12
113	18.29	3.02
114	18.38	2.93
115	18.48	2.84
116	18.57	2.75
117	18.68	2.66
118	18.76	2.57
119	18.84	2.49
120	18.95	2.41

Appendix D: OPEX calculation-Base case

Steam	
Steam consumption	208,166.31 KW
Steam price (Eur/Kwh)	0.015
Total price (KEur/yr)	24980.6

Electricity	
Electricity price (Eur/kwh)	0.136
Rich pump electricity consumption	127.86
Total price (KEur/yr)	0.139118162

MEA	
MEA in solution	2125713.456
Make up MEA	61.08372
MEA Price (Eur/Ton)	1450
Total MEA price (KEur/yr)	3790.8

Maintenance	
CAPEX	64384.3
Maintenance price (4% of CAPEX)	2575 (Keur/y)
Engineer	
Number of Engineers	1
engineering salary (KEur/yr)	156.7

Operator	
Number of operators	6
Salary (KEur/yr)	80.41
Total operator salary (KEuro/yr)	482.46

Total annual OPEX (KEur/yr)	32140.1
Total annual OPEX (M€/yr)	

OPEX calculation-Vapor recompression

Steam	
Steam consumption	141448.88 KW
Steam price (Eur/Kwh)	0.015
Total price (KEur/yr)	16973.86

Electricity	
Electricity price (Eur/kwh)	0.136
Rich pump electricity consumption (KW)	87.27
Lean pump electricity consumption (KW)	99.3
Compressor electricity consumption (KW)	4135.90
Total price (Keur/yr)	4702.8

MEA	
MEA in solution	820232.2
Make up MEA	61.08
MEA Price (Eur/Ton)	1450
Total MEA price (Keur/yr)	1897.91

Maintenance	
CAPEX	118672.51
Maintenance price (4% of CAPEX)	4746.90 (KEur/y)
Engineer	
Number of Engineers	1
engineering salary (KEur/yr)	156.7

Operator	
Number of operators	6
Salary (Keur/yr)	80.41
Total operator salary (KEuro/yr)	482.46

Total annual OPEX (KEur/yr)	28960.73

Appendix E: CO_2 capture cost, CO_2 removal efficiency

CO_2 Captured cost:

CO_2 captured (ton/yr)	1208286.08
Total annualized CAPEX (MEuro/yr)	11.720134
Total annualized OPEX (MEuro/yr)	28.96
CO_2 captured cost (Euro/ton)	33.34

CO_2 removal efficiency and reboiler duty:

CO_2 removal efficiency (%)	92.62
Reboiler duty (kJ/h)	509,215,990.03
Reboiler duty per kg CO_2 captured (MJ/ ton CO_2)	3.371