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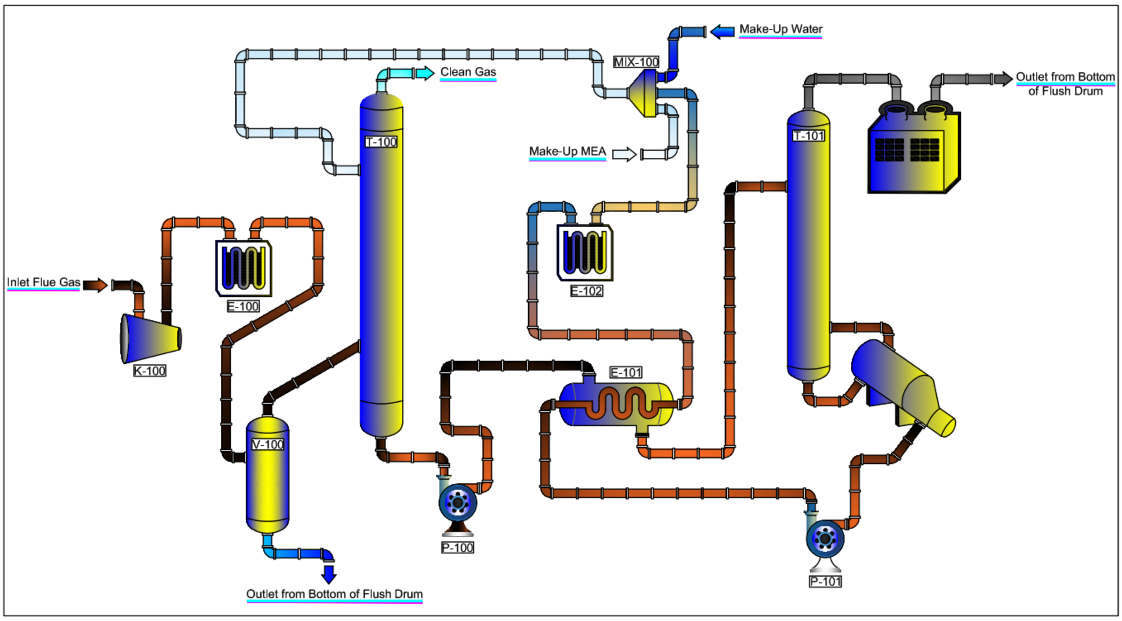
Process Technology

Process design of CO_2 capture from the gas turbine at an oil platform



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Summary:

Human activities and the ever-increasing growing need for energy, have significantly increased the concentration of carbon dioxide. Fossil fuels have established a huge impact to provoke these emissions especially in an offshore application. This tragedy and its consequences, such as global warming and climate change, temperature increase, and also sea level increase, have prompted international organizations all over the world, to set an international framework to control and reduce the adverse effects of greenhouse gas emissions.

In order to remove the CO_2 emissions, several methods have been proposed, but among all, capturing CO_2 by amine-based absorption column seems to be one of the best alternatives. In this project, a standard amine-based CO_2 removal process plant for an offshore application as a request of Moreld Apply has been investigated in Aspen HYSYS software. Due to the restriction in space and weight on the offshore platforms, a wide optimization has been conducted and the impact of parameter changes has been evaluated.

In this study, to meet Moreld Apply AS needs for one of their projects on an offshore platform in the North Sea in Norway, a simplified standard CO_2 capture process plant has been simulated and designed in Aspen HYSYS. This unit has been missioned to remove the CO_2 content of flue gas with 90 % efficiency from the Waste Heat Recovery Unit (WHRU) package which is itself working with two parallel gas turbines, steam turbine, reboiler, condenser and a heat exchanger. The WHRU package was in scope of two bachelor students at USN working in parallel with this study, and this work covers the CO_2 removal process plant simulation and design, dimensioning, cost estimation, as well as optimization.

Having this purpose, a base case model has been developed in which a minimum approach temperature (ΔT_{min}) of 10 °C has been assumed in the main heat exchanger and was specified to have 90 % removal efficiency, packing height in an absorber column was calculated to 16 meter and 3.6 MJ energy in the reboiler per mass of CO_2 captured. Since the CO_2 capture plant is an expensive process especially in an offshore application, a sensitivity analysis has been conducted to investigate the process dependency on some parameter changes, one is changing number of packing height in the absorber column from 10 to 16 meter, and also change of minimum approach temperature (ΔT_{min}) in range of 10, 15, and 20 °C. During these parameter changes, circulation rate, energy demand per kilogram of CO_2 removal, the ratio of CO_2 content to the amine content from the bottom outlet in the absorber column, steam and electricity consumption have been evaluated. As a suggestion for a more optimum solution of this study, minimum approach temperature of 15 °C and an absorber column with 13-meter packing height with 87 % efficiency have been suggested which requires 5.5 MJ energy in the reboiler unit to remove one kilogram of CO_2 during the process.

Contents

Nomenclature

Abbreviation	Description
CAPEX	Capital expenditure
CCS	Carbon capture and storage
CO_2	Carbon dioxide
C.S.	Carbon steel
€	Euro
EDF	Enhanced detailed factor
$EIC_{C.S.}$	Total carbon steel equipment installation cost
f_{mat}	Material factor of stainless steel
$F_{T,C.S.}$	Total carbon steel installation factor
fx	A general factor for the component x
H_2O	Water
H_2S	Hydrogen Sulfide
MEA	Monoethanolamine
n	Operation life time
NGCC	Natural gas combined cycle
NOK	Norwegian Kroner
N_{stage}	Number of packing stages
OPEX	Operational expenditure
PFD	Piping flow diagram
r	Interest rate
S.S.	Stainless steel

Symbol	Description	Unit
A	Cross sectional area	[m^2]
C_P	Specific heat capacity	$\left[\frac{kJ}{kg \cdot K}\right]$
D	Diameter	[m]
H	Height	[m]
$h_{packing}$	Packing height	[m]
h_{shell}	Shell height	[m]
h_{stage}	Packing height of each stage	[m]
$\dot{m}$	Mass flow	$\left[\frac{kg}{h}\right]$
$\dot{Q}$	Duty	$[kW]$
ΔT_{lm}	Logarithmic mean temperature difference	$[K]$
U	Overall heat transfer coefficient	$\left[\frac{kW}{m^2 \cdot K}\right]$
V	Volume	[m^3]
$V_{packing}$	Packing volume	[m^3]
V_{shell}	Shell volume	[m^3]
$\dot{V}$	Volumetric flow rate	$\left[\frac{m^3}{s}\right]$
v_{gas}	Gas velocity	$\left[\frac{m}{s}\right]$
W	Weight	[kg]

1 Introduction

Today, the risk of greenhouse gas pollution is not hidden from anyone and among all carbon dioxide (CO_2) is at the top of list. According to the Intergovernmental Panel on Climate Change (IPCC), fossil fuel has established a significant impact of 57 % in provoking of greenhouse gas (GHG) pollution which results in climate change and global warming, atmospheric temperature increase, and even worse increasing sea level. Despite this threat, the continuous growing demand for energy exacerbates this risk more and more [1], [2].

The climate change and its consequences caused by increasing greenhouses gas specially CO_2 emission from fossil fuels, is not limited to the Norway, it has also become a major concern all over the world. So, by the purpose of reducing and controlling CO_2 emissions, government increased the taxes on CO_2 emission and Norway is one the strictest one. In Norway, it is proven that offshore gas turbines play the most significant role in this disaster [3], [4].

One of the best solutions to mitigate the CO_2 emitted by oil and gas offshore platforms in a large scale, is CO_2 capture and storage (CCS). It is predicted by International Energy Agency (IEA) that through continuous implementing CO_2 capture technology, the CO_2 emitted by the energy sector will experience a 50 % deduction by 2050 [3].

Although CO_2 capture process has known as a golden alternative, running a CO_2 capture process plant is considerably expensive especially in an offshore platform, and there is usually severe restriction in terms of space and weight in an offshore installation [3]. So, in this project in order to reduce cost and specially space (major issue in an offshore application), an optimization on the CO_2 capture process plant has been performed.

1.1 Background

The present project is done as a request of Moreld Apply AS for one of their offshore platform projects in Norway. Due to the demolition of an old Waste Heat Recovery Unit (WHRU) in one of the offshore platforms located in the North Sea, a shortage in the platform's heating medium is forecasted. So then, to compensate for the shortage, a new energy source is demanded for the heating medium. In addition, a CO_2 capture plant is requested to not only reduce the emission (due to the regulations), but also, capturing CO_2 for further investigation. Figure 1.1 is shown an overview of the foreseen new plant.

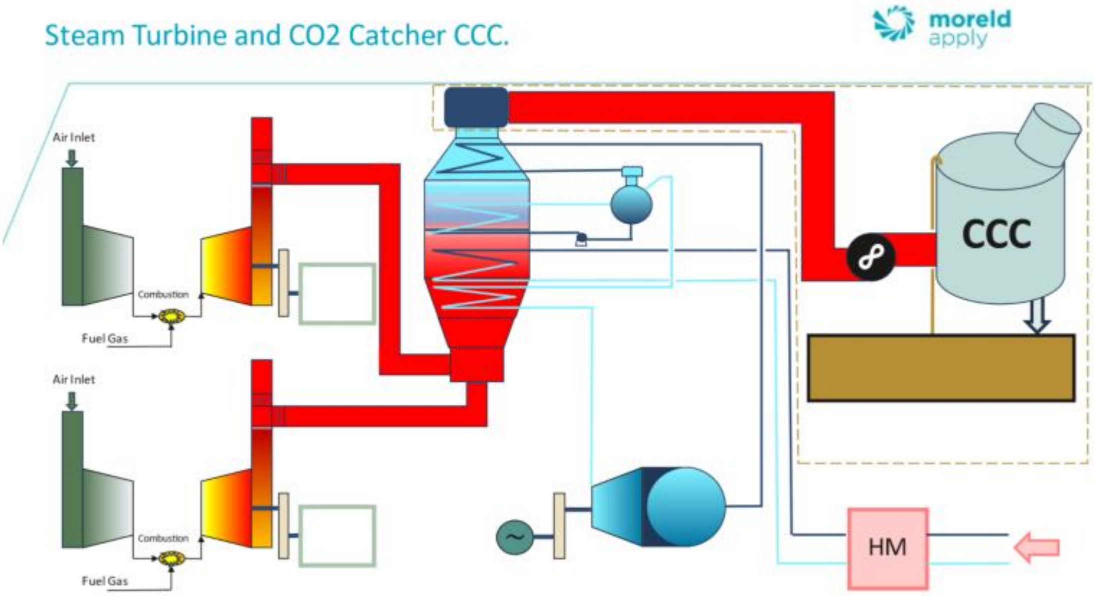

Figure 1.1: Overview on new plant.

So, in this regard, a team including two bachelor students were responded to work on the new WHRU package consisting of gas turbine, steam turbine, reboiler, condenser and a heat exchanger. Regarding the CO_2 capture part, this report will cover a detailed design simulation, dimensioning, and also cost estimation of participated equipment.

2 Literature report and review

This chapter will discuss the most relevant available literature in the area of CO_2 capturing process design and simulation, dimensioning, and cost estimation. This chapter also includes literature review more specifically on the power production through combined cycle processes with emphasis on the offshore application.

2.1 General literature study

Anand B. Rao et al. in 2002 [5], have performed a study on a post combustion process including an absorption amine-based CO_2 capture system, in order to remove CO_2 from the power plant flue gas working with fossil fuels. He integrated the simulation model result with an existing power plant project to evaluate the feasibility study and carbon capture cost estimation.

To remove CO_2 from exhaust gas of a power plant working with fossil fuel in a combined cycle, an amine-based absorption and desorption process was designed by Lars Erik Øi in 2007 [6]. The combined cycle in this study was consisted of both gas and steam turbines followed by a compressor to compress the air needed for combustion in a combustion chamber, a condenser for recirculating the cooling water, and also steam generators. The simulation has been done in Aspen HYSYS tool and the model has been developed for different cases by varying rate of MEA circulation, height of absorption unit, and also absorption and reboiler temperature.

Ove Braut Kallevik in 2010 [7] in his master's thesis, presented an amine-based absorption process of CO_2 capture from natural gas power plant developed in Aspen HYSYS, to introduce a new technique in the direct contact cooler simulation, and also make-up MEA and make up water. He also expanded his study through performing dimensioning and cost estimating to have a better understanding of the subject. So, in this regard, for all heat exchangers participated in the process, a wide calculation on overall heat transfer as well as the correction factor were carried out. He concluded that in order to have an 85 % efficiency in capturing CO_2, 15 stages in absorber column and a minimum approach temperature on 10 to 14 K are needed in the lean-rich heat exchanger in optimum case. He in his study tracked the effects of variation in ΔT_{min}, packing height in the absorber column, and the temperature of inlet gas into the absorption column, for the different removal efficiency of 80 %, 85 %, and also 90 %.

Lars Erik Øi in his thesis for the degree of Doctor Philosophiae in 2012 [8], perform an extensive study and simulation in both Aspen HYSYS and Aspen Plus, on CO_2 capture process of exhaust gas from a power plant working with natural gas. In his work, he calculated the CO_2 removal efficiency based on functionality of the absorber column temperature, circulation rate as well as other factors. By the purpose of optimization, a dimensioning of participated equipment as well as the cost estimation have been conducted.

Supposing CO_2 emission reduction, Xiaobo Luo in 2016 through his PhD thesis [9], has modeled, simulated and optimized a post combustion process working with the MEA amine as the solvent, in order to optimize the design as well as the operation economically. In his model, a combined cycle working with natural gas (NGCC) has been considered. He used Aspen HYSYS and Aspen Plus software for modelling and simulation in his work and at the end, he compared the result from simulation with an experimental data to validate his work.

A techno-economic methodology analysis has been proposed by Hassan Ali in his PhD thesis [10], in 2019, in which the influence of all assumptions whether technical or economic assumptions on the cost estimation were evaluated to have an overview of major costly component and their sensitivity. Having this purpose, he proposed an amine-based absorber process design which was simulated in Aspen HYSYS tool. In his study for the base case, he claimed that CAPEX, steam and electricity cost played the role of cost drivers. For the cost estimation analysis, they used both method of cost estimation in Aspen In-Plant cost estimator and also EDF (Enhanced detail factor) method and calculated CAPEX and OPEX in both methods. In OPEX, they assumed the OPEX was consist of electricity consumption, heat energy usage and the maintenance. Three parameters were defined as variables which were a part of cost evaluation to assess their effects on costs through variation. These three variables are including the removal efficiency of the CO_2 from the flue gas, the number of the packing stages in the absorber column, and the minimum approach temperature (ΔT_{min}) in the main heat exchanger. The most optimized ΔT_{min} in the heat exchanger was calculated to be between 10 and 15 °C, and 12 stages in absorber unit resulted in the most economically optimized solution.

In 2021, Lars Erik Øi et al. [11], by the purpose of calculating the most economically optimized process parameters and also the possibility evaluation of an automated cost estimation as well as optimization, performed a simulation of the CO_2 capture plant in Aspen HYSYS and have done a wide study on the cost estimation and optimization of the designed process. Traditional monoethanol amine (MEA) was chosen to dispose and remove the CO_2 from the exhaust gas that came from a cement plant.

Solomon et al. in 2021 [12], have performed a wide investigation on CO_2 capturing plant (working based on an amine-based solution) cost estimation by implementing Enhanced Detailed Factor method (EDF method) to evaluate how the installation factor of equipment as well as the characteristic factor of plant construction impact on CAPEX cost. After evaluating seven methodologies in his study, he concluded that since the Enhanced Detailed Factor method (EDF) evaluates each equipment as an independent project, the total installation factor for each equipment directly relates to the equipment cost, so that higher equipment's cost caused lower installation factor. So, they considered the EDF method as a suitable method of cost estimating for both establishing new CO_2 capture plant and also modification of existing plants.

In 2021 Sina Orangi [13], in his master thesis developed a traditional amine-based CO_2 capture process to remove CO_2 emission and evaluate the process characters for different solvents. In his study he also performs a cost estimation by the help of Aspen In-Plant tool and also Enhanced detail Factor (EDF) method, to have a better judgment on their effects and find the most optimized solution.

In 2022 Shirvan Shirdel et al. [14], have evaluated a standard CO_2 capture plant missioned to removing carbon dioxide from the flue gas coming from waste burning facility of Fortum at Klemetsrud in Norway. Having cost-optimizing as the main purpose in this study, the simulation has conducted in Aspen HYSYS software and the cost estimation has been implemented with two method of Enhanced Detailed Factor (EDF) and Power Law method. A base case model has been defined in which the absorber packing has 20-meter height with a 90 % efficiency, and the minimum approach temperature set to be 10 °C. The cost of CO_2 capturing for the base case has been calculated as 37.5 $\frac{€}{ton}$. To find out the most optimum solution, a sensitivity analysis has been investigated and as a result, the minimum CO_2 captured cost is calculated based on EDF and Power Law method 37.3 $\frac{€}{ton}$ and 37.1 $\frac{€}{ton}$ respectively with regards of the case has a packing absorber with 19-meter height. With 87 % efficiency working with the temperature difference of 15 °C in the heat exchanger, the cost of CO_2 captured 37 $\frac{€}{ton}$ and 36.7 $\frac{€}{ton}$ have been calculated respectively for the EDF and Power Law methods.

2.2 Specific literature review on power production followed by CO_2 capture, with emphasis on offshore application

In this study, there was a big challenge to find out open literature on process simulation of CO_2 capture from a gas turbine offshore application. There are very few literature references that contain both the CO_2 capture process from offshore gas turbines in a combined cycle by process simulation. However, table 2.1 shows the list of most relevant literature available in this area.

Table 2.1: Literature overview.

Row	Literature Title	Year	Location	Reference
1	Efficient CO_2 capture through a combined steam and CO_2 gas turbine cycle	1992	Belgium	[15]
2	Assessment of power generation concepts on oil platforms in conjunction with CO_2 removal	1994	Norway	[16]
3	Separation of carbon dioxide from offshore gas turbine exhaust	1995	Norway	[17]
4	A quantitative comparison of gas turbine cycles with CO_2 capture	2007	Norway	[18]
5	Redesign, optimization, and economic evaluation of a natural gas combined cycle with the best integrated technology CO_2 capture	2009	Germany	[19]
6	Pre-combustion CO_2 capture analysis of integrated reforming combined cycle	2010	Norway	[20]
7	Off-design simulations of offshore combined cycle	2012	Norway	[21]
8	Optimization of combined cycles for offshore oil and gas installations	2013	Norway	[22]
9	Design and off-design simulations of combined cycles for offshore oil and gas installation	2013	Norway	[23]

10	CCS on offshore oil and gas installation design of post-combustion capture system and steam cycle	2017	Norway	[4]
11	Transient performance of combined cycle power plant with absorption based post-combustion CO_2 capture: dynamic simulations and pilot plant testing	2018	Norway	[24]
12	Simulation of a combined cycle gas turbine power plant in Aspen HYSYS	2019	Singapore	[25]
13	Energy optimization of offshore gas field	2021	Italy	[2]

Here is a brief description on each literature mentioned in table 2.1.

In order to dispose and capture CO_2 for a long-term life time, J. De Ruyck et al. [15] in 1992 proposed a combined cycle with higher efficiency and more economical, in CO_2 capturing process. In their proposal cycle, they considered a combined steam-CO_2 gas turbine and to evaluate and assess the proposed cycle behavior, a viable project was defined. After performing their concept, they concluded that the combined steam-CO_2 gas turbine cycle was viable with no major issues in practical manner and on the other hand, it somehow resulted in no CO_2 emission for a certain application.

Yngvil et al. [16], through their study in 1994, assessed six alternatives in power generation with the purpose of reducing CO_2 emission from exhaust gas released from gas turbine in an offshore natural gas combustion. In their study, they targeted to find out the best possible concept in terms of the highest efficiency, the highest CO_2 concentration, the lowest weight and space, and also the lowest exhaust gas flow. At the top, they focused on weight and space more, since these two factors have a crucial role in an offshore platform project cost. Not having conflict with oil production process when CO_2 capture plant is under maintenance, is the second factor they considered more seriously and the last constrain they took into account more in their study, was the possibility of CO_2 compression and storage. Having these scopes, they considered a simplified CO_2 capture process and defined the six concepts of exhaust gas recycling, a combined cycle with supplementary fired, an exhaust gas recycling with STIG (Steam Injected Gas Turbine), the combination of combined cycle with Oxygen spiking, a traditional steam turbine, and gas engine in the power generation process. After performing and comparing the six defined cycles, they concluded that the first option, exhaust gas recycling in the combined cycle, was the best alternative in the CO_2 capture process in terms of their concerns and it resulted in the highest efficiency in CO_2 removal, and second lightest option.

In 1995 Olav et al. [17], have performed a feasibility study on an optimization of CO_2 removal process for an offshore platform by implementing a combined cycle (including Gas turbine, Heat recovery unit and power generation). To investigate the optimum CO_2 removal process, they have focused on optimization of absorption unit, desorption unit and optimization of the layout. In terms of absorption unit optimization, they have considered three possible processes which are including CO_2 absorption by the help of an absorption column, a Rotating contacting unit and a gas absorption membrane. Regarding the layout optimization, since offshore platforms are very expensive, it is crucial to take extra attention to the space and layout in order to control the cost. So, in this regard, they have made a detailed 3D modelling of the platform with interaction with other units especially absorption and desorption unit, since these two equipment have directly impact on the size and weight and consequently the cost. Through this study, they concluded five results in which:

1. MEA-amine absorption process is concluded as the best process
2. By the purpose of reduction in exhaust gas flow rate and have higher CO_2 concentration, a 40 % recycled line for exhaust gas to the gas turbine is predicted.
3. Although the traditional absorption column is more conventional, gas turbine membrane is concluded the most optimum alternative.
4. The suggested gas absorption membrane has to be tested in order to be sure about the process.
5. The presented study has shown a fascinating internal rate of return (IRR) for the future use if the proposed technology could be run.

In 2007, Hanne M. Kvamsdal et al. [18], considered nine strategies of gas turbine cycles followed by CO_2 capture process in which energy efficiency as well as CO_2 emission through heat and mass balance simulation were investigated. In their study, the focused on the concepts in combination of power plant working with natural fired gas with a reference case of 400 MW, and CO_2 capture process then after. The nine defined concepts were including both pre and post combustion and also oxy fuel process. Through their study they understood that novel technology's concepts resulted in the best performance in terms of both CO_2 emission and the energy efficiency while in both pre and post combustion CO_2 capture processes, they experienced lower efficiency in capturing rate of 90 %.

Cristina Botero et al. in 2009 [19], have conducted a study to find out the best integrated concept of CO_2 emission capturing in a post combustion process of a combined cycle worked with natural gas in a power plant with a capacity of 400 MW. To have a 90 % CO_2 removal efficiency in an amine-based solution (namely monoethanolamine solution) of 30-wt %, the power plant has been redesigned and an optimization and cost estimating on participation of recirculation of exhaust gas has been performed. He concluded that the best approach was a combined system of gas turbine, steam turbine and HRSG unit in which the obtained power was 361 MW with the 50 % LHV efficiency.

In 2010, Lars Olof Nord [20], by the purpose of reducing the increase in atmospheric temperature, has proposed process in which have captured the CO_2 emission from the gas in a power plant working with natural gas for a pre- combustion process including heat recovery steam generator (HRSG), in his PhD thesis. So in this regard, a thermodynamic analysis has performed and the rate of CO_2 capturing and the efficiency of plant were the characters that he focused on his process simulations. In his study, he implemented process simulations by the help of several tools including Aspen Plus, GT PRO, and GTMASTER which are verified by AspenTech, and Thermoflow respectively. He put all efforts to investigate in case of operating condition change, how much the plant is flexible.

Øystein Flatebø in his master thesis in 2012 [21], has evaluated several offshore combined cycle configurations to not only meet the power need for the offshore installation and accordingly reduce the cost, but also control and decrease the emission from the offshore platform operation. In his study, the possible combined cycles have been discussed both technically and thermodynamically, and then an existed offshore combined cycle has studied. A simulation study of two defined case studies also has performed in GTPRO software and the plant performance has validated in GTMASTER software. One of the case studies focused on the designing of the offshore installation, and another assigned to obtain high efficiency. In order to detect a proper balance between weight and efficiency, a sensitivity analysis also has carried out. Overall, the simulated model of the offshore installation case study was recorded an improved result of 50.3 MW and 50.3 % as power output and plant efficiency respectively with the same weight of existing plant. For another case study, higher power of 2.4 MW has generated, however, 209 ton has been added to the weight in this case. He also found out that the ambient temperature has no effect on the gross plant efficiency and the highest efficiency belonged to the design setting.

In continuous work done by Øystein Flatebø in 2012 [21], Alexander Svae Sletten in 2013 [22], has done his master thesis on investigation and optimization of implementing the combined cycle method in practical use in an offshore platform as a source of power. During his thesis, he was assigned to select an objective function in MATLAB, Microsoft Excel, and GT PRO software to optimize the design parameters. He developed an objective function as well as a review on potential MATLAB software optimizing method and then after, an improvement on the optimized solution has been performed. He concluded that optimizing of a combined cycle in offshore application could be a good alternative to improve power generation as well as saving weight.

In 2013, Lars Olof Nord et al. [23], with the purpose of improving net plant efficiency and decreasing the CO_2 emission cost has performed a wide study on offshore combined cycle with oil and gas application in comparison with simple gas turbine cycle. A 26-33 % improvement in plant efficiency and a 20-25 % reduction have recorded for the Combined cycle rather than the simple GT cycle. Despite of these benefits, the combined cycle caused higher ratio of weight to power by 60-70 %, in comparison with the simple GT cycle. In this study also, a layout plant of a combined cycle for an offshore application has been conducted in which both generator and mechanical drive gas turbines.

Lars Olof Nord et al. [4] in 2017, has raised the significant role of gas turbines as a power production resource in an offshore oil and gas installation in increasing the level of CO_2 emission. Absorption based CO_2 capture system and storage, has known as an effective option to control and reduce the CO_2 concentration. So, in order to supply energy demand in the desorber reboiler, a heat resource is required. A simple gas turbine cycle, which is existed in most of the offshore platforms, is not an option to meet power needs in CCS system since no extra steam in this system will remain to donate. To compensate the energy demand entirely or partially, he suggested an additional system of a compact steam bottoming cycle that can supply power need of equipment including in the CO_2 capture system. In his work, as well as designing of a post-combustion CO_2 capture plant, three individual case studies with more focus on system cycle design have been investigated. Through his study, a weight evaluation for main equipment has been performed to investigate a simple relation between outlet gas turbine mass flow and size of the oil and gas installation.

Rubén Mocholí Montañés in 2018 in his PhD thesis [24], has performed a wide study on Chemical absorption process performance analysis to attenuate climate change and greenhouse gas through capturing CO_2 emitted by the combined cycle power plants working with natural gas. To validate and evaluate his work, he also performed a large-scale pilot plant through dynamic process simulation. From the result, he understood the control structure with the best performance is one that desorber bottom output temperature is governed by the flow rate of solvent and reboiler duty is the one that controlled the rate of CO_2 capturing rate. It is also concluded that a scenario with a rapid load change, can be resulted in rejection in gas flow rate.

Zuming Liu et al. in 2019 [25] have performed a design simulation for a combined gas turbine cycle of a power plant, in Aspen HYSYS. Since their work was the first study in this area and there was no similar literature to compare, they validated their work in Aspen HYSYS with an equivalent work in GateCycle model. They found out that on average, the relative deviation of less than 2 % between the obtained power and the thermal efficiency of the gas and steam turbine, and the combined cycle. They also concluded that Aspen HYSYS is a better choice and more accurate than GateCycle for several of reasons. For example, Aspen tool uses fluid package of real gas Peng-Robinson and also possibility of easier integration through its versatile energy system, as well as possibility of making dynamic model to predict the plant's real time behavior.

Adel Ramadan in 2021 [2], in his master's thesis, has performed a wide technical study on the various methods of energy supply and their impact on the environment. In his work he focused on cumulative CO_2 emitted through energy supply cycles during the operation life-time of an offshore platform, and the methods he mainly evaluated were consist of a gas turbine (GT), also combined cycle as the energy supply source, and also supplying electrical power demand through onshore power. Having this purpose, he conducted a process design simulation in Aspen HYSYS software for different possibilities including a combination of gas turbine followed by a WHRU unit, also a combined cycle consisting a gas turbine unit works with a steam bottoming cycle, and finally Importing power from the onshore source to electrify the platform. He concluded that the third option is the most desired option since through this 85.8 % of CO_2 emission will reduce comparing with other methods.

2.3 Lack of information in open literature

There are a lot of open studies investigating CO_2 capture process plants mostly from the gas power plants with an onshore application but not in an offshore installation. There are very few studies on offshore CO_2 capture analysis and almost none about the suggested solutions with parameter values selected in the dimensioning, and also in the optimization. However, in this work, similar procedure of designing, simulating, and cost estimation for the CO_2 capture process have been performed and regarding the optimization part in addition of the CO_2 capture package itself, more consideration regarding the offshore application have been taken into account.

Absorption packing height, Minimum approach temperature (ΔT_{min}) in the main heat exchanger, and the removal efficiency were selected as variable parameters in Lars Erik Øi study [11]. In this study, he concluded the packing height of 12 meter, the ΔT_{min} of 13 °C, while in the study of Ove Braut Kallevik [7], the optimum case has 15 stages in the absorber and the heat exchanger works with 10 to 14 °C temperature difference. Hassan Ali [10], with the same variable as Lars Øi [11], ended up with an optimum case which has ΔT_{min} in the heat exchanger between 10 and 15 °C, and 12 stages in absorber unit. In Shirvan et al. study [14] also the packing absorber with 19-meter height and a temperature difference of 15 °C in the heat exchanger has presented as the most optimum case study.

Lars Øi in both of his studies [6], [11] and also Ove Braut Kallevik [7] all agreed with the removal efficiency 85 % as the most optimum case. However, Shirvan et al. [14] found out the 87 % as the most optimized efficiency.

3 Process description

This chapter covers a general process description of the CO_2 capture process from the WHRU unit and also gives an overall description on upstream process in which inlet flue gas to the absorption process go through before CO_2 will capture in the next step.

3.1 Upstream process before CO_2 capture

Figure 3.1 shows the process that flue gas traveled before CO_2 capturing process. Since in an offshore platforms power energy demanding by different facilities and equipment is huge, in this project, this combined process cycle is assigned to meet the platform's need for power energy. This combination is a set of heat engines collaborate sequentially together with the same heat source and for a typical offshore project is included a set of gas turbine, a WHRU unit, and a generator followed by a condenser and a pump. Having this purpose, the flue gas enters into the two parallel gas turbines which are assigned to produce power. Then after, release heat will recover in a waste heat recovery unit (WHRU) which work in conjunction with the gas turbines to supply additional heat and generate more energy as a supplementary power for the entire system on the platform. One of the exhaust gasses from WHRU unit will send into the reboiler and steam turbine to generate electricity and then, the produced water will send into the condenser for getting ready to send back into the WHRU to complete the cycle[2], [8].

Another exhaust gas from WHRU will send into the heat exchanger to cool it down and make it ready for further actions in CO_2 capture plant to remove the CO_2 from the gas.

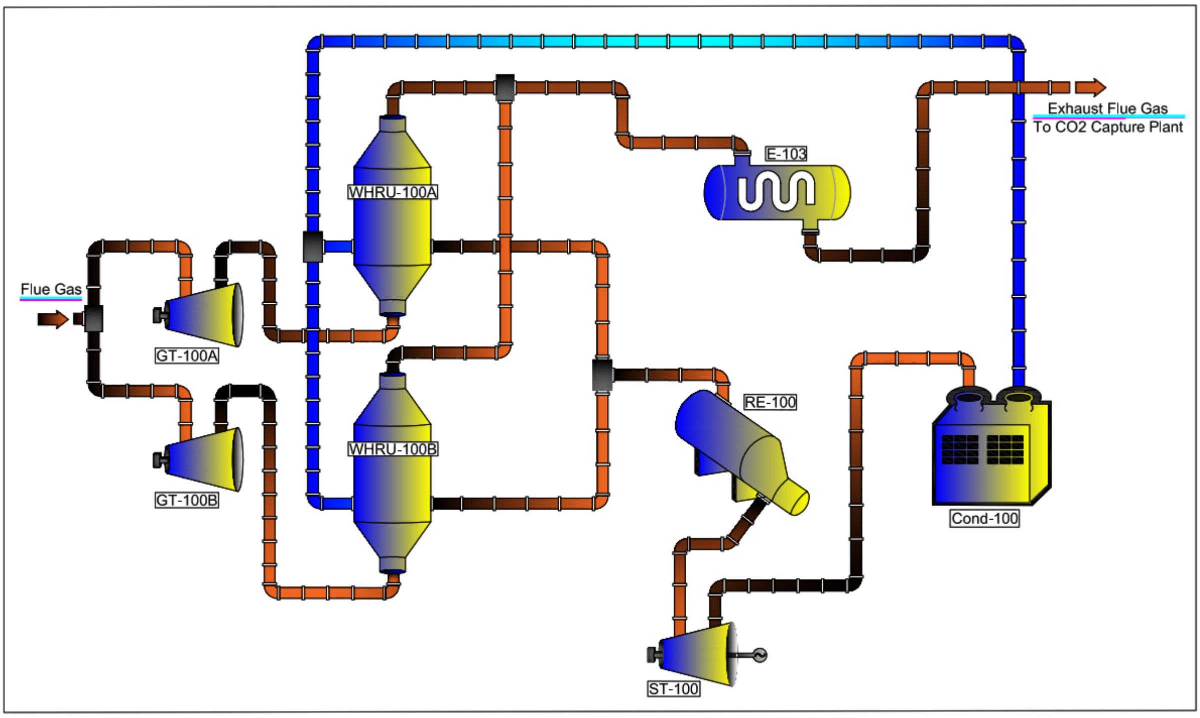

Figure 3.1: Upstream process before CO_2 capture.

3.2 CO_2 capture process

Figure 3.2 shows the CO_2 capture process for the defined project. The exhaust gas from WHRU unit after cooling down through a heat exchanger, by the purpose of capturing CO_2 will carry out to the absorber unit. But before entering the absorption column, the pressure, and temperature of the flue gas needed to be adjusted. To have a saturated vapor flue gas, a flush drum after the fan and the cooler is considered to achieve this importance.

Then after, to proceed with the CO_2 removal process, the exhaust gas will lead to an absorber where the CO_2 will remove from the gas with the help of a mixture of MEA amine and water injected into the absorber from the top of the column. The dissolved CO_2 then, exit the absorber column from the bottom and is sent to the rich amine pump to pressurize for further process in the desorber unit. Meanwhile, the clean gas will exit the column from the top.

In order to regenerate the MEA amine and separate the CO_2 captured from the flue gas in the absorber unit, the gas should carry out into the desorber unit where this is done with the help of a reboiler and a condenser. But before sending the gas into the stripper, its temperature should be increased to fulfill the reboiler requirement. The required energy to heat up the gas is supplied by the outlet of the reboiler itself as can be seen in figure 3.2.

As the regeneration process in the desorber column is done, CO_2 removed from the gas is obtained from the top of the stripper to send for further storage processes. The regenerated amine, however, will exit the stripper from the bottom and will be prepared to send it back to the cycle where the MEA injects into the absorber column. For this purpose, the assigned cooler will cool it down and reduce its pressure to proper temperature and pressure for reusing again at the first stage injecting it into the absorber. A mixer also compensates for the amine and water losses during the process.

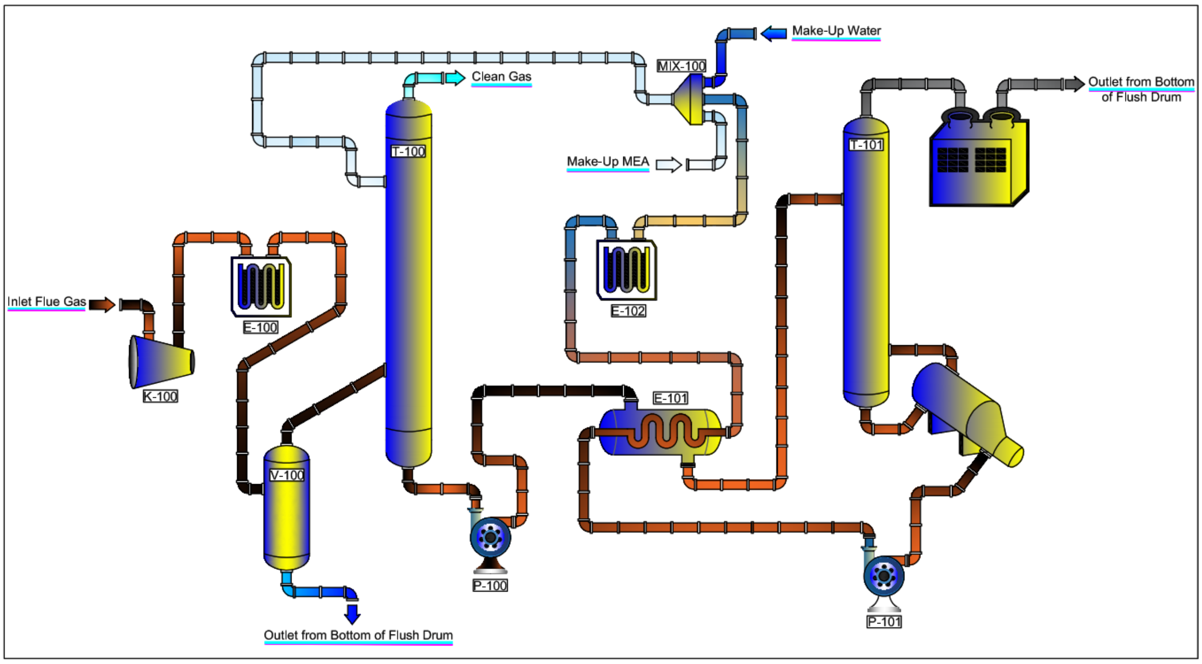

Figure 3.2: CO_2 capture process.

3.3 Problem description

In order to fulfill the Moreld Apply AS demand on the new plant described in section 1.1, two sequential processes are considered. First process is including the new WHRU unit works in the combined cycle as described in section 3.1. Due to the limitation in space and also weight (since WHRU is huge in size and weight), two parallel WHRU units are considered. So then, two parallel gas turbines are assigned to serve the two WHRU units. After traversing the whole process as explained, the high temperature exhaust gas will send to the assigned heat exchanger to cool down the gas before sending to the CCUS process in order to prevent damage to the compressor (K-100). Then after, it will send into the CO_2 capture process as described in section 3.2, to remove CO_2 from the gas and send the capture CO_2 for further action.

3.4 Piping flow diagram (PFD)

The piping flow diagram (PFD) of the described processes is shown in figure 3.3.

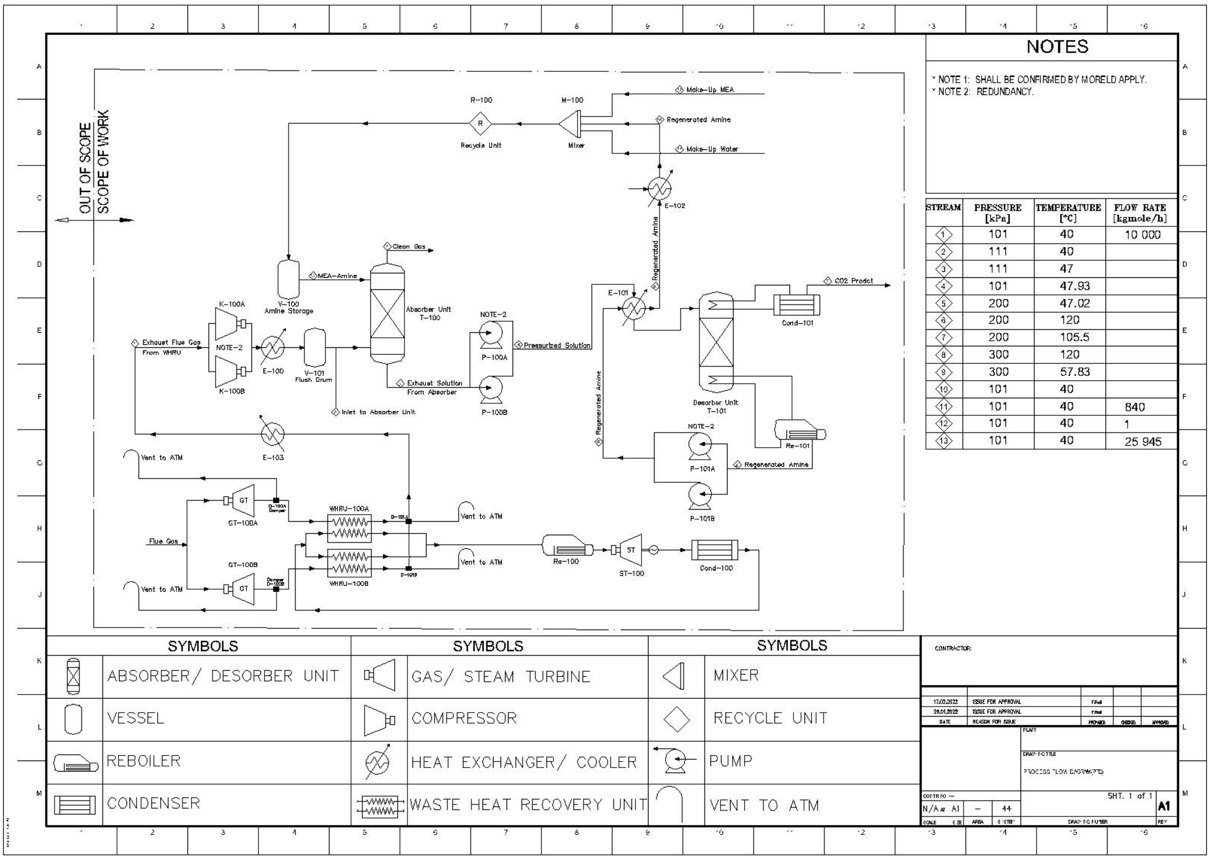

Figure 3.3: Piping flow diagram (PFD) for the up-stream and the CO_2 capture process.

4 Simulation of base case

To design and simulate a full-scale process described in section 3, the well-known Aspen HYSYS software is one of the best choices to solve equilibrium models through mass and energy balances.

In this study, a standard process of amine-based CO_2 capture is developed in Aspen HYSYS version 12 in which, an acid gas property package is used as the equilibrium model fluid package and the solver in both absorber and desorber units is considered to be the Modified HYSIM Inside-Out with adaptive damping [6], [10], [11].

4.1 Base case

Figure 4.1 illustrates the complete cycle of CO_2 capturing process described in section 3, except the recycle block which is considered to deliver recycling cycle for the regenerated amine, which has been designed and simulated in Aspen HYSYS.

In this project for the base case, the required CO_2 removal efficiency is defined to be 90 % and ΔT_{min} is assumed to be 10 °C. So then, to achieve this goal, a base case study has performed and resulted in requirement of an absorber column with 16 stages inside, a stripper with 6 stages and the required energy to remove one kg CO_2 is calculated to be 3577 kJ for the reboiler in regeneration process.

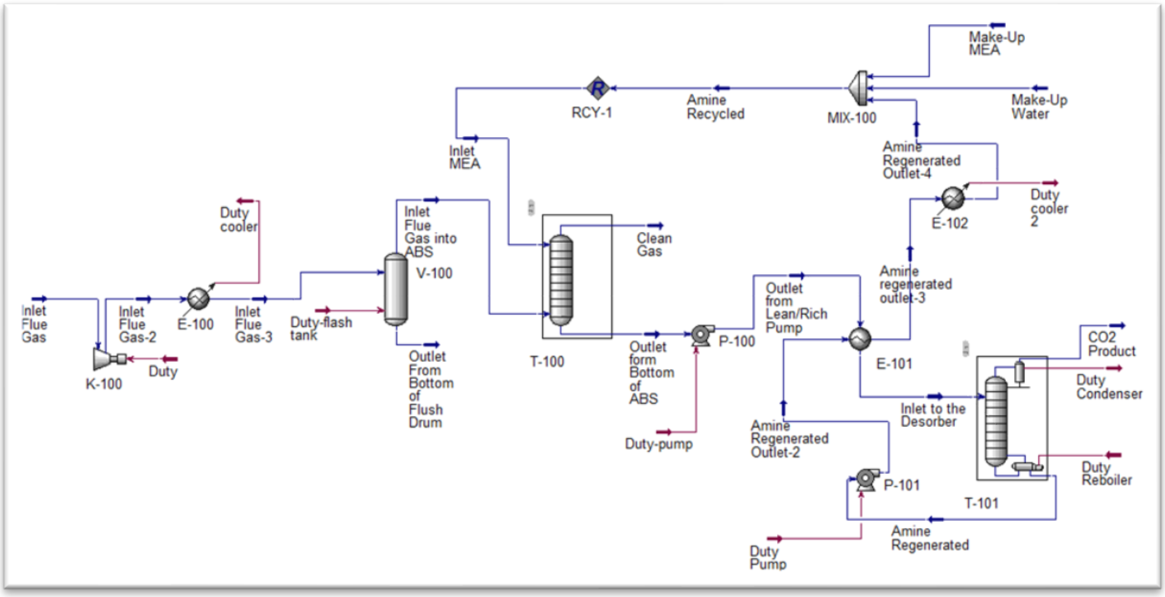

Figure 4.1: CO_2 capture process model in Aspen HYSYS.

4.1.1 Stream specification

In this subchapter, specification of feed streams including exhaust flue gas from WHRU, MEA amine stream injecting into the absorber, and also make-up streams for water and MEA amine will be introduced. It should be mentioned that since the actual stream specifications and composition were not determined at the time that the project started, the data used in this report comes from the work done by Lars Erik Øi [6] in 2007.

4.1.1.1 Feed flue gas specification

The specification of feed flue gas from WHRU unit is defined as reported in table 4.1 [6].

Table 4.1: Feed flue gas specification and some results.

FLUE GAS FROM WHRU:			
Properties:	**Value:**	**Unit:**	**Reference**
CO_2 (Mole Fraction)	0.0373	-	[6]
H_2O (Mole Fraction)	0.0671	-	[6]
MEAmine (Mole Fraction)	0	-	[6]
Nitrogen (Mole Fraction)	0.8956	-	[6]
H_2S (Mole Fraction)	0	-	[6]
Pressure	101	[kPa]	[6]
Temperature	40	[°C]	[6]
Molar Flow Rate	10000	[kgmole/h]	Assumed

4.1.1.2 Amine feed stream specification

Table 4.2 includes the amine feed stream specification [6].

Table 4.2: Amine feed stream specification and some results.

AMINE FEED:			
Properties:	**Value:**	**Unit:**	**Reference**
CO_2 (Mass Fraction)	0.055	-	[6]
H_2O (Mass Fraction)	0.6509	-	[6]
MEAmine (Mass Fraction)	0.2942	-	[6]
Nitrogen (Mass Fraction)	0	-	[6]
H_2S (Mass Fraction)	0	-	[6]
Pressure	101	[kPa]	[6]
Temperature	40	[°C]	[6]
Molar Flow Rate	25945	[kgmole/h]	Calculated

4.1.1.3 Make-up stream specification

Due to the losses in CO_2 capturing process, make-up water, and MEA amine are considered to compensate those losses with the help of a mixture unit. Table 4.3 and 4.4 show the specification of these make-up streams in which the obtained values are the difference between the value before entering to the mixer unit and in the feed stream.

Table 4.3: Make-Up water stream specification.

MAKE-UP WATER:			
Component:	**Value:**	**Unit:**	**Reference**
H_2O in Amine Feed	22216	[kgmole/h]	Calculated
H_2O in stream before Mixer	21376	[kgmole/h]	Calculated
Required Make-Up Water	840	[kgmole/h]	Calculated

Table 4.4: Make-Up amine stream specification.

MAKE-UP AMINE:			
Component:	**Value:**	**Unit:**	**Reference**
Amine in Amine Feed	2961	[kgmole/h]	Calculated
Amine in stream before Mixer	2960	[kgmole/h]	Calculated
Required Make-Up Amine	1	[kgmole/h]	Calculated

4.1.2 Process unit simulation

4.1.2.1 Fan (K-100)

To get ready the inlet flue gas for absorption process, where the CO_2 will remove from the gas through reaction with the MEA amine solution, a regular fan is designed to increase pressure flue gas from 101 kPa to a proper working pressure of 111 kPa as Lars Erik Øi [6] considered in his study. The specification prior and after the fan is reported in table 4.5.

Table 4.5: Fan (K-100) specification and some results.

FAN:			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flue gas temperature	40	[°C]	[6], [26]
Outlet flue gas temperature	51.15	[°C]	Calculated
Inlet flue gas pressure	101	[kPa]	[6], [26]
Outlet gas pressure	111	[kPa]	[6], [26]
Fan Duty	921.75	[kW]	Calculated

4.1.2.2 Heat exchanger (E-100)

To adjust the gas temperature for absorption process in absorber column, a heat exchanger is assigned to reduce the temperature increased during pressurization in fan, from 51.15 °C to 40 °C. Table 4.6 shows the gas specification before and after the heat exchanger unit.

Table 4.6: Heat exchanger (E-100) specification and some results.

HEAT EXCHANGER:			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flue gas temperature	51.15	[°C]	Calculated
Outlet flue gas temperature	40	[°C]	[6], [26]
Inlet flue gas pressure	111	[kPa]	[6], [26]
Outlet gas pressure	111	[kPa]	[6], [26]

4.1.2.3 Flash drum (V-100)

The inlet gas into the absorber unit first sent to a flash drum which is designated to separate water liquid from the gas at the bottom of the vessel. Table 4.7 shows the gas specification before and after the vessel.

Table 4.7: Flash Drum (V-100) specification.

FLASH DRUM:			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flue gas temperature	40	[°C]	[6], [26]
Outlet flue gas temperature	40	[°C]	[6], [26]
Inlet flue gas pressure	111	[kPa]	[6], [26]
Outlet gas pressure	111	[kPa]	[6], [26]

4.1.2.4 Absorber unit (T-100)

The process of CO_2 separation from the gas will occur in the absorption column in a way that MEA amine injects from the top of the column meanwhile gas travel from the bottom. During their journey they contact each other and carbon dioxide in the gas reacts with the injected amine and the absorption process will accomplish. To have done this important, the absorber column works between 101 kPa and 111 kPa and 16 packing stages with 1 meter height each with 0.15 efficiency, is calculated. More detailed specification regarding the streams before and after the column and also the absorber itself can be found in table 4.8.

Table 4.8: Absorber column (T-100) specification and some results.

ABSORBER:			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flow temperature	40	[°C]	[6], [26]
Inlet flow pressure	111	[kPa]	[6], [26]
P_1 pressure absorber	101	[kPa]	[6], [26]
P_n pressure absorber	111	[kPa]	[6], [26]
Number of stages	16	[-]	Calculated
Efficiency for each stage	0.15	[-]	[7], [11]
Packing height (Each)	1	[m]	[7], [11]
Outlet flow temperature at top (Clean Gas)	47.93	[°C]	Calculated
Outlet flow pressure at top (Clean Gas)	101	[kPa]	[6], [26]
CO_2 mole fraction in clean gas (Top product)	0.0085	[-]	Calculated
Outlet flow temperature in bottom	47	[°C]	Calculated
Outlet flow pressure in bottom	111	[kPa]	[6], [26]

4.1.2.5 Rich amine pump (P-100)

After the absorption process is done, the bottom mixture should be sent to the desorber unit for the desorption process and regeneration of amine. But before sending it into the stripper, the temperature and pressure should be adjusted in a way that suits for the working condition of the desorber unit. So then, the outlet mixture of rich amine and dissolved CO_2 from the bottom of absorber will send to the rich amine pump to increase the mixture pressure from 111 kPa to 200 kPa which is the desorber working pressure. The stream specification before and after the pump are tabulated in table 4.9.

Table 4.9: Rich amine pump (P-100) specification and some results.

RICH AMINE PUMP (P-100):			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flow temperature	47	[°C]	Calculated
Inlet flow pressure	111	[kPa]	[6], [26]
Efficiency	75	[%]	[7], [11]
Duty	19.05	[kW]	Calculated
Outlet flow temperature in bottom	47.02	[°C]	Calculated
Outlet flow pressure in bottom	200	[kPa]	[6], [11]

4.1.2.6 Desorber unit (T-101)

To proceed the desorption process, the solution is sent to the desorber unit where the reboiler in this unit works in a temperature of 120 °C and pressure of 200 kPa. Both reboiler and condenser in this unit designed to work with efficiency of 1 and packing stage efficiencies designed to be 0.5. More detailed stream specification before and after the stripper and also stripper itself can be found in table 4.10.

Table 4.10: Desorber column (T-101) specification.

DESORBER:			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flow temperature	109.5	[°C]	Calculated
Inlet flow pressure	200	[kPa]	[6], [11]
Pressure of condenser	200	[kPa]	[6], [11]
Pressure of reboiler	200	[kPa]	[6], [11]
Number of stages	6	[-]	Calculated
Packing height (Each)	1	[m]	[7], [11]
Reflux ratio in stripper	0.3	[-]	[6], [11]
Efficiency for condenser and reboiler	1	[-]	[6], [11]
Efficiency for each stage	0.5	[-]	[7]
Condenser duty	1.073e+007	[kJ/h]	Calculated
Reboiler duty	8.696e+007	[kJ/h]	Calculated
Outlet flow temperature at top (CO_2 product)	105.5	[°C]	Calculated
Outlet flow pressure at top (CO_2 product)	200	[kPa]	[6], [11]
CO_2 mole fraction at top (CO_2 product)	0.3828	[Wt%]	Calculated
Outlet flow temperature in bottom	120	[°C]	[6], [11]
Outlet flow pressure in bottom	200	[kPa]	[6], [11]

4.1.2.7 Lean amine pump (P-101)

A pump after stripper is designed to pressurized the lean amine solvent to 300 kPa and the stream specifications are as defined in table 4.11.

Table 4.11: Lean amine pump (P-101) specification and some results.

LEAN AMINE PUMP (P-101):			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flow temperature	120	[°C]	[6], [11]
Inlet flow pressure	200	[kPa]	[6], [11]
Efficiency	75	[%]	[7], [11]
Duty	22.75	[kW]	Calculated
Outlet flow temperature in bottom	120	[°C]	[6], [11]
Outlet flow pressure in bottom	300	[kPa]	Assumed

4.1.2.8 Lean / Rich heat exchanger (E-101)

An ideal heat exchanger is assigned to heat up the inlet lean-rich amine solvent into the stripper up to the proper temperature for the desorption process, through the hot bottom outlet of desorber unit. For the base case, it is assumed that Min approach temperature (ΔT_{min}) is 10 °C. Table 4.12 shows all specification for both cold and hot streams and some main results.

Table 4.12: Lean/rich heat exchanger (E-101) specification some main results.

LEAN/RICH HEAT EXCHANGER:				
Component:		**Value:**	**Unit:**	**Reference:**
Cold	Cold stream inlet temperature	47.02	[°C]	Calculated
	Cold stream outlet temperature	109.5	[°C]	Calculated
	Cold stream inlet pressure	200	[kPa]	[6], [11]
	Cold stream Outlet pressure	200	[kPa]	[6], [11]
Hot	Hot stream inlet temperature	120	[°C]	[6], [11]
	Hot stream outlet temperature	57.83	[°C]	Calculated
	Hot stream inlet pressure	300	[kPa]	Assumed
	Hot stream Outlet pressure	300	[kPa]	Assumed
Min approach temperature (ΔT_{min})		10.51	[°C]	Calculated
LMTD		10.66	[°C]	Calculated

4.1.2.9 Lean amine cooler (E-103)

To complete the MEA amine recycling process and send it back to the absorber column, its pressure and temperature should be adjusted to the initial condition of 101 kPa and 40 °C which is done by a cooler. Table 4.13 illustrates the specification of streams before and after the cooler.

Table 4.13: Lean amine cooler (E-103) specification and some results.

LEAN AMINE COOLER (P-101):			
Component:	**Value:**	**Unit:**	**Reference:**
Inlet flow temperature	57.83	[°C]	Calculated
Outlet flow temperature	40	[°C]	[6], [26]
Inlet flow pressure	300	[kPa]	Assumed
Outlet flow pressure	101	[kPa]	[6], [26]

5 Dimensioning

Since space is a super important issue in an offshore platform project, and there is usually limited area on offshore platforms to install facilities and equipment, a simplified dimensioning for main equipment and facilities is performed to have a better overview of the occupied space and also for further use in cost estimating. Having this purpose, this chapter covers dimensioning on main and most expensive parts including absorber column, main heat exchanger, steam consumption, and the electricity demand.

5.1 Dimensioning for base case

5.1.1 Absorber column

In order to estimate the required space needed for the absorption packing and shell column, absorber volume [m^3] is calculated as a dimensioning factor in this equipment. To calculate the volume, equations of 5.1 and 5.2 should be followed [13].

$$V_{shell} = \frac{\pi \cdot D^2}{4} \cdot h_{shell} \tag{5.1}$$

$$V_{packing} = \frac{\pi \cdot D^2}{4} \cdot h_{packing} \tag{5.2}$$

where:

- h_{shell} [m] is shell height in the absorber column.
- $h_{packing}$ [m] is packing height in the absorber column.
- D [m] is absorber column diameter.

Since the column volume is a function of its diameter and height, first these parameters need to be specified. By calculating column sectional area [m^2] based on equation 5.3, the absorber diameter [m] then, can be obtained by following equation 5.4.

$$A = \frac{\dot{V}}{v_{gas}} \tag{5.3}$$

where:

- v_{gas} is gas velocity $\left[\frac{m}{s}\right]$ and in this study it is assumed to be 2 $\frac{m}{s}$.
- $\dot{V}$ $\left[\frac{m^3}{s}\right]$ is the actual volumetric gas flow rate which is obtained from the Aspen HYSYS simulation.

$$D = \sqrt{\frac{4 \cdot A}{\pi}} \tag{5.4}$$

Where A is the cross-sectional area [m^2] calculated from equation 5.3.

Considering one meter height for each packing stage [7], the total packing heigh then can be calculated by multiple number of stages to the height of each stage as shown in equation 5.5.

$$h_{packing} = h_{stage} \cdot N_{stage} \tag{5.5}$$

where:

- h_{stage} [m] is packing height for each stage and in this study, it is assumed to be one meter.
- N_{stage} $[-]$ is the number of absorber column stages in simulation.

In base case, 16 number of stages is calculated to remove 90 % of the CO_2 content so then, the total packing height for the base case will be 16 m. However, absorber column height is not limited to the packing height, but also there are other parts, which is including inlet and outlet of the gas, water washer, demister, liquid distribution and sump, that should be take into account when calculating total absorber column height [10]. Having this approach, in base case, for 16 packing stages 34 m will added as extra space as Ove Braut Kallevik [7] is considered in his work. Therefore, the total absorber height for the base case as shown in equation 5.6 is:

$$h_{shell} = h_{packing} + 34 \tag{5.6}$$

Having calculated the absorber diameter and height, column volume can be then obtained for further application. In overall, the absorber column dimensioning calculation for base case is presented in table 5.1.

Table 5.1: Absorber column dimensioning specification and results.

Component:	**Symbol**	**Value:**	**Unit:**	**Reference:**
Data:				
Actual vapor volume flow rate	$\dot{V}$	283089	$\left[\frac{m^3}{h}\right]$	Calculated
Gas velocity	v_{gas}	2	$\left[\frac{m}{s}\right]$	Assumed
Number of stages	N_{stage}	300	[-]	Assumed
Packing height (each)	h_{stage}	1	[m]	[7]

Calculation:				
Cross sectional area	A	39.32	$[m^2]$	Calculated
Absorber diameter	D	7.08	[m]	Calculated
Total packing height	H	16	[m]	Calculated
Total absorber height	H	50	[m]	[7]
Absorber volume	V	~2150	$[m^3]$	Calculated
Absorber estimated weight[1]	W	173300	[kg]	Calculated

5.1.2 Lean/Rich heat exchanger

The dimensioning factor in heat exchangers is heat transfer area which determines the cost for further use in cost estimation section. To calculate the required heat transfer area [m^2] for the main heat exchanger, equation 5.7 can be used.

$$A = \frac{\dot{Q}}{U \cdot \Delta T_{lm}} \tag{5.7}$$

where:

- $\dot{Q}$ [KW] represents the heat exchanger duty which is obtained from the Aspen HYSYS simulation.
- $U\left[\frac{KW}{m^2.\ K}\right]$ is overall heat transfer coefficient and in this study, it is assumed to be 500 $\frac{KW}{m^2.\ K}$.
- ΔT_{lm} [K] is logarithmic mean temperature difference and is obtained from the Aspen HYSYS simulation. For this study it is assumed that there is an ideal countercurrent flow and equation 5.8 represents how ΔT_{lm} is calculated.

$$\Delta T_{lm} = \frac{\Delta T_{out} - \Delta T_{in}}{ln\left(\frac{\Delta T_{out}}{\Delta T_{in}}\right)} \tag{5.8}$$

[1] Less packing.

where:

$$\Delta T_{out} = T_{hot,in} - T_{cold,out} \tag{5.9}$$

$$\Delta T_{in} = T_{hot,out} - T_{cold,in} \tag{5.10}$$

Considering 10 K temperature difference between the cold stream inlet, outlet from P-100 in figure 4.1, and hot stream outlet, amine regenerated outlet-3 in figure 4.1, table 5.2 illustrates the area calculation for base case [7], [11], [13].

Table 5.2: Lean/rich heat exchanger dimensioning result.

Component:	**Symbol**	**Value:**	**Unit:**	**Reference:**
Data:				
HEX duty (Q)	$\dot{Q}$	1.35E+08	$\left[\frac{KJ}{h}\right]$	Calculated
LMTD	ΔT_{lm}	10.66	$[K]$	Calculated
Calculation:				
HEX area	A	7012.7	$[m^2]$	Calculated
HEX estimated weight	W	15500	[kg]	Calculated

5.1.3 Steam consumption

Overall steam consumed in the reboiler for the desorption process in desorber column, can be obtained from reboiler duty from Aspen HYSYS. For the base case condition, a 8.70E+07 $\frac{KJ}{h}$ or 24154 kW steam is consumed during the desorption process.

5.1.4 Electricity consumption

To get pumps and fan working, electricity is needed and since electricity is one the expensive item in OPEX, the electricity consumption should be estimated at evaluation and design stage. In this study for the base case a 75 % efficiency for pumps and fan is assumed, so then, a 963 kW in total is needed for 2 pumps (P-100 and P-101) and one fan (K-100) which is obtained from adding the equipment duty calculated by Aspen HYSYS.

6 Cost estimation

CO_2 capture process has known as an expensive and costly process. All equipment has their own role in such a costly process, however, absorber column and the main heat exchanger, take the first and second most expensive parts of the process respectively [8]. So, in order to have an overview on the final cost of the defined project, in this chapter a cost estimation on main and most expensive parts including absorber column, main heat exchanger, steam consumption, and the electricity demand will be covered. The presented cost estimation in this chapter will cover both capital cost estimation (CAPEX) and operational cost estimation (OPEX).

It should be mentioned that higher efficiency and resulting lower pressure loss of structured packing is the reason that structured packing is selected in this project [8].

6.1 CAPEX

To estimate the CAPEX in this study, Aspen In-Plant cost estimator software version 12 has been used. Also, Enhanced detail factor method, known as EDF method, is the method the has been used to perform the capital cost estimation (CAPEX). In this study the implemented detailed installation factors have been developed over the years by Nils Henrik Eldrup who is one of the most experienced people in oil and gas industry and was involved many projects in this area. It is really important to choose a reliable and accurate method for cost estimation in early stage as well as preparing the opportunity for performing an optimization on individual equipment. Enhanced detail factor method (EDF) has provided these opportunities for users. In addition, this method has provided possibility of conducting a techno-economic analysis for both new technologies as well as developing existed process plants [27].

All factors presented in Nils detailed installation factor table, attached in appendix A, are based on Carbon Steel material (C.S.). So, since all equipment except the fan including the CO_2 capture process plant are Stainless Steel, in order to use the Nils detailed installation factor, material should be converted to the Carbon Steel and then after will corrected to the Stainless Steel again. Equation 6.1 has presented for this purpose.

$$Equipment\ Cost_{CS} = {Equipment\ Cost_{ss}}/{f_{mat}} \tag{6.1}$$

where:

- $Equipment\ Cost_{SS}\ [NOK]$ is cost of an equipment with Stainless Steel material.
- $Equipment\ Cost_{CS}\ [NOK]$ is cost of an equipment with Carbon Steel material.
- $f_{mat}\ [-]$ is material factor which converts SS to CS and vice versa [27].

According to the most recent Nils detailed installation factor, which can be found in the appendix A, released in 2020 the material factors are presented as tabulated in table 6.1.

Table 6.1: Material factors.

Material Adjustment:			
Material:	**Type of equipment:**	f_{mat}**:**	**Reference:**
Carbon steel (C.S.)	-	1	Appendix A
Stainless steel (S.S.)	Welded	1.75	Appendix A
Stainless steel (S.S.)	Rotating	1.30	Appendix A

6.1.1 Calculation of total installation cost

In order to calculate the total cost of installation for each specific equipment, the total installation factor needs to be calculated which itself is consist of several sub installation factors. Direct cost factor, engineering cost factor, administration, commissioning and contingency cost factors are these sub installation factors which can be obtained from the Nils detailed installation factor table shown in Equation 6.2. Having calculated the total installation factor, the total equipment installation cost (EIC) can be obtained through multiplying the total calculated cost factor by equipment cost with Carbon Steel material as presented in equation 6.3.

$$F_{T,C.S.} = f_{direct} + f_{engineering} + f_{administration} + f_{commissioning} + f_{contingency} \qquad (6.2)$$

where:

- $F_{T,C.S.}[-]$ is total installation factor for an equipment with Carbon Steel material.
- $f_{direct}\ [-]$ is direct installation factor for an equipment with Carbon Steel material.
- $f_{engineering}\ [-]$ is engineering installation factor for an equipment with Carbon Steel material.
- $f_{administration}\ [-]$ is engineering installation factor for an equipment with Carbon Steel material.
- $f_{commissioning}\ [-]$ is engineering installation factor for an equipment with Carbon Steel material.
- $f_{xontingency}\ [-]$ is engineering installation factor for an equipment with Carbon Steel material.

$$EIC_{C.S.} = F_{T,C.S.} \times\ Equipment\ Cost_{C.S.} \qquad (6.3)$$

where:

- $EIC_{C.S.}[NOK]$ is equipment installed cost for an equipment with Carbon Steel material.
- $F_{T,C.S.}[-]$ is total installation factor for an equipment with Carbon Steel material.
- $Equipment\ Cost_{C.S.}[NOK]$ is cost of an equipment with Carbon Steel material.

To calculate total equipment installation cost for equipment with Stainless Steel material equation 6.4 should be followed.

$$EIC_{S.S.} = Equipment\ cost_{C.S.} \times F_{T,S.S.} \tag{6.4}$$

where:

- $EIC_{S.S.}[NOK]$ is equipment installed cost for an equipment with Stainless Steel material.
- $F_{T,S.S.}[-]$ is total installation factor for an equipment with Stainless Steel material.
- $Equipment\ Cost_{C.S.}[NOK]$ is cost of an equipment with Carbon Steel material.

To obtain the total installation factor for Stainless steel material, equation 6.5 should be followed [27].

$$F_{T,S.S.} = \left[F_{T,C.S.} + \left\{(f_{mat} - 1)(f_{equipment} + f_{piping})\right\}\right] \tag{6.5}$$

where:

- $f_{equipment}$ $[-]$ is equipment material installation factor for an equipment with Carbon Steel material.
- f_{piping} $[-]$ is piping installation factor for an equipment with Carbon Steel material.

6.1.2 Currency and inflation adjustment

Since the present project is located in Norway, all currency should be presented in the Norwegian currency [NOK]. The currency of calculated equipment cost from Aspen In-Plant cost estimator is in Euro [€] which is needed to convert to the Norwegian currency. The Nils detailed installation factor table also is based on Euro [€] currency. So, to adjust the currency, it is enough to multiply the cost in Euro [€] by the exchange rate of [Euro] to [NOK]. According to the Norges bank data base, the exchange rate in early October 2021 was 9.8 NOK per Euro [€]. Equation 6.6 is an illustration on this subject [28].

$$Total\ installed\ cost\ [€] = total\ installed\ cost\ [\mathrm{NOK}] \times exchange\ rate\ \left[\frac{€}{NOK}\right] \tag{6.6}$$

As well as currency, inflation needs to be considered. Since the Nils detailed installation factor table is from 2020 and output of Aspen In-Plant cost estimator software is also based on year 2020, to include inflation and adjust based on the time, equation 6.7 can be used [29].

$$Cost_{year\ 2020} = Cost_{year\ 2019}\left(\frac{Cost\ index_{year\ 2020}}{Cost\ index_{year\ 2019}}\right) \tag{6.7}$$

Where cost index is reported in table 6.2.

Table 6.2: Cost inflation index.

year:	Value:	Reference:
2021	317	[30]
2020	301	[30]

6.1.3 Annualized CAPEX

To have better understanding of the installation cost of equipment involving the CO_2 capture plant overall its operational life time, annualized CAPEX should be calculated as described in equation 6.8.

$$Annualized\ CAPEX\ \left[\frac{€}{yr}\right] = \frac{CAPEX}{Annualized\ factor} \tag{6.8}$$

In which, total installed cost will be calculated based description in section 6.1 and the annualized factor can be obtained from equation 6.9.

$$Annualized\ factor = \sum_{n=1}^{n}\left[\frac{1}{(1+r)^n}\right] \tag{6.9}$$

where:

- n represents operational life time.
- r represents interest rate.

In this project, it is assumed that the operational life time is 24 years with one year as the construction period (In total 25 years) [27]. So, having these assumptions, the annualized factor for the present project will be 10.53.

6.1.4 Base case CAPEX

Having theory and assumptions described in section 6.1, the cost of most expensive equipment including absorber column unit, main heat exchanger, pumps, and the fan for the base case are calculated in Aspen In-Plant cost estimator and the assumptions and the outputs can be followed in table 6.3 and 6.4.

Table 6.3: Details from CAPEX calculation.

Items		Absorber Shell	Absorber Packing	HEX	Pump (P-100)	Pump (P-101)	Fan
Material / Type		S.S 316	S.S 316L	S.S 316LW	S.S 316	S.S 316	C.S
Quantity		1	1	10	1	1	1
Welded/Machined		Welded	Welded	Welded	Machined	Machined	Machined
Material Factor		1.75	1.75	1.75	1.30	1.30	-
Fluid/Solid		Fluid	Fluid	Fluid	Fluid	Fluid	Fluid
Original Value in [kEuro] in 2020 S.S		2929	2294	236	66	69	-
Original Value in [kEuro] in 2020 C.S		1673	1310	135	51	53	230
Total Installation Factor for C.S		3.19	3.19	5.89	7.22	7.22	4.92
New Total Installation Factor for S.S		4.16	4.16	7.21	8.69	8.69	-
Total Installation Cost [kEuro] in S.S		6971	5459	973	442	463	1133
Total Installation Cost S.S.	Exchange rate in 2020	9.78	9.78	9.78	9.78	9.78	9.78
	Value in [kNOK] in 2020	68183	53396	9521	4327	4530	11081
	Index Value at 2020	301	301	301	301	301	301
	Index Value at 2021	317	317	317	317	317	317
	Value in [kNOK] in 2021 (S.S)	71807	56234	100274	4557	4771	11670
	Value in [kEuro] in 2021 (S.S)	7342	5749	10253	466	487	1193

Table 6.4: Total and annualized CAPEX result.

Equipment	Total CAPEX [MEuro]	Total CAPEX [MNOK]	Anualized CAPEX [MEuro/Yr]	Anualized CAPEX [MNOK/Yr]
Absorber	13.1	128.1	1.2	12
HEX	10.3	100.3	0.9	9.4
Pump (P-100)	0.5	4.6	0.05	0.5
Pump (P-101)	0.5	4.8	0.05	0.5
Fan	1.2	11.6	0.11	1.1
SUM	**25.5**	**249.4**	**2.4**	**23.4**

6.2 OPEX

As well as capital expenditure, operational costs (OPEX) also need to be calculated to have a realistic cost estimation. The OPEX is consist of electricity cost, steam consumption, process and cooling water, solvent MEA, maintenance cost, and operators and engineers cost involved the project during the operation. Since the present project is an offshore platform, it is assumed the sea water will be used as the cooling water. Table 6.5 shown the assumptions in OPEX calculation.

Table 6.5: Assumptions.

Item	Value	Unit	Reference
Operating Hour	8000	$\left[\frac{Hour}{Yr}\right]$	[12]
Electricity Cost Factor	0.078	$[\frac{€}{kWh}]$	[12]
Steam Cost Factor	0.032	$[\frac{€}{kWh}]$	[12]
Cooling Water	0	$[\frac{€}{m^3}]$	Assumed sea water will be used
Process Water	0.203	$[\frac{€}{m^3}]$	[12]
Solvent MEA	1516	$[\frac{€}{m^3}]$	[12]
Maintenance	3% of CAPEX	[kNOK]	[12]
Operator (6 operators)	482484	[€]	[12]
Engineering (1 engineer)	156650	[€]	[12]
Exchange rate	9.78	[-]	[28]

As mentioned before, a 25 years in total is assumed for this project which is consist of 1 year construction and 24 years of operation [27].

6.2.1 Base case OPEX

Having assumptions described in section 6.2, the calculated OPEX for the base case is described in table 6.6.

Table 6.6: OPEX calculation base case result.

Equipment	Annual OPEX	
	[MEuro/Yr]	[MNOK/Yr]
Electricity Cost	0.6	5.9
Steam Cost	6.2	60.5
Maintenance Cost	0.8	7.5
Cooling Water Cost	0	0
Process Water Cost	0.03	0.3
Operator Cost	0.5	4.7
Engineering Cost	0.2	1.5
Solvent Cost	1.6	15.8
SUM	**9.8**	**96.2**

6.2.2 CO_2 Capture Cost

Having calculated annualized CAPEX and annual OPEX, the cost per tons of CO_2 captured can be estimated through equation 6.10 [27].

$$CO_2\ Capture\ Cost = \left(\frac{Annualized\ CAPEX + Annual\ OPEX}{CO_2\ captured\ amount\ per\ year}\right) \tag{6.10}$$

6.2.2.1 Base case CO_2 capture cost estimation

In this project, for defined operational period, for the base case, the amount of CO_2 result that will be captured is as reported in table 6.7.

Table 6.7: Amount of CO_2 captured result.

Item	Value	Unit	Reference
CO_2 In	16415	$\left[\frac{kg}{h}\right]$	Calculated
CO_2 Out	1719	$\left[\frac{kg}{h}\right]$	Calculated
CO_2 Captured	14696	$\left[\frac{kg}{h}\right]$	Calculated
CO_2 Captured	117568	$\left[\frac{ton}{yr}\right]$	Calculated

And table 6.8 shown the cost per tons of CO_2 capture.

Table 6.8: CO_2 captured cost result.

Item	Value	Unit	Reference
CO_2 captured Cost	0.1	$\left[\frac{k€}{t\ CO_2}\right]$	Calculated
CO_2 captured Cost	1.01	$\left[\frac{kNOK}{t\ CO_2}\right]$	Calculated

Generally, in a practical offshore application, the total cost including the CAPEX is much higher than the costs calculated in this section. As Nils Henrik Eldrup recommended (personal communication, March 21, 2022), to have a more realistic and practical cost estimation that can be used as a reference, all calculated cost, the CAPEX should be multiplied by coefficient 3.

7 Sensitivity analysis

This chapter covers the sensitivity analysis on the CO_2 capture process plant. In this study the sensitivity analysis is performed through changing some parameters including changes in number of stages in the absorber column and changes in minimum approach temperature (ΔT_{min}).

7.1 Number of stages in absorber

As mentioned in chapter 4, for the defined base case, 16 stages in absorber column have been conducted to catch the efficiency of 90 %. In order to investigate impact of absorber height parameter in the outputs including energy consumption, MEA rate, and the efficiency of the process, seven case studies have been considered. The case studies are consisting of range of changes in number of stages in absorber column from 10, to have 80 % plant efficiency, to 16 by one.

As assumed in the base case, stage efficiency for all case studies assumed to be 0.15. In order to keep efficiency as close as possible to the base case efficiency of 90 %, during the cases, the MEA flow rate has been adjusted accordingly. As can be seen from outputs reported in chapter 8, for lower packing heights it was not possible to reach 90 % efficiency.

7.2 Minimum approach temperature (ΔT_{min})

The minimum approach temperature (ΔT_{min}) for the defined base case has been considered to be 10 °C. However, in real project especially in an offshore application having kept this temperature difference is fairly optimistic in terms of cost optimization. So, in this regard, two other cases with minimum approach temperature (ΔT_{min}) of 15 °C and 20 °C have been investigated.

It should be mentioned that during this analysis, all parameters and assumptions have been kept constant as the one in the base case.

8 Results

The output and results of the presented case studies in sensitivity analysis nominated in chapter 7, will be covered in this chapter.

8.1 Absorber column height

As mentioned in chapter 7, seven case studies are investigated to check out the sensitivity by change of number of stages in absorber column (or change of packing height). Minimum approach temperature (ΔT_{min}) also, has been varied between 10 °C to 20 °C for all the seven defined case studies and their results will be presented in this sub-chapter. In all three case studies of ΔT_{min} of 10, 15 and 20 °C, efficiency have been kept almost the same.

8.1.1 Minimum approach temperature of 10 degree

Considering the assumptions of the minimum approach temperature (ΔT_{min}) of 10 °C, stage efficiency as 0.15, the same value as base case, and same inlet flue gas flow rate of 10 000 $\frac{kgmol}{h}$ as the base case, table 8.1 shows the process responses for all seven case studies. These responses are including flow rate of MEA amine, the amount of energy required per kilogram of CO_2 captured during the process, CO_2 and amine content in the outlet from bottom of the absorber column and accordingly the coefficient of α which is the ratio of CO_2 content to the amine content from the same outlet.

Regarding the MEA amine requirement, with 16 stages in the absorber, 26 000 $\frac{kgmol}{h}$ MEA is demanded, while reducing number of stages, followed by higher MEA demand per hour. This increasing trend in MEA demand for case-7 is violated since the efficiency for this case is quite low even less than 80 %.

The energy demand in reboiler per one kilogram of CO_2 that captured, is another important and costly parameter that should take care of it. Decreasing of packing stages in the absorber, will increase energy demand significantly. For the base case with 16 stages, the required reboiler energy is 3577.32 $\frac{kJ}{Kg}$. But by switching to the one less stage, energy demand will grow to 4003.93 $\frac{kJ}{Kg}$. As can be followed in table 8.1, Case 7 is trend breaker since the efficiency is too less than the target. Equation 8.1 has shown how the energy demand is calculated.

$$Q_{reboiler} = \frac{Reboiler\ duty\ \left[\frac{kJ}{hr}\right]}{Mass\ flow\ of\ CO_2\ product\ \left[\frac{kg}{hr}\right]} \tag{8.1}$$

The next parameter which has been evaluated within the case studies is ratio of CO_2 content to amine content in the bottom outlet of the column, $\alpha\ [-]$, that can be calculated from equation

8.2. Higher α, represents higher CO_2 content in the bottom output which leads to higher CO_2 removal.

$$\alpha = \frac{Bottom\ outlet\ CO_2\ content}{Bottom\ outlet\ Amine\ content} \tag{8.2}$$

During these cases, it is tried to keep total efficiency as close as possible to the base case efficiency. However, as can be seen from the result reported in table 8.1, by decreasing the number of stages in the absorber column, the efficiency will also reduce specially for less than 13 stages, the efficiency will go less than 85 %.

Table 8.1:Sensitivity analysis for varying number of stages for minimum approach temperature of 10 °C.

Cases	**No. of stage**	**Stage Efficiency**	**Inlet flue gas flow rate**	**MEA flow rate**	**CO_2 content of bottom outlet**	**Amine content of bottom outlet**	$\boldsymbol{\alpha}$	$\boldsymbol{Q_{Reboiler}}$	**CO_2 removal efficiency**
Unit	[-]	[-]	$\left[\frac{kgmol}{h}\right]$	$\left[\frac{kgmol}{h}\right]$	[-]	[-]	[-]	$\left[\frac{kJ}{kg}\right]$	[%]
Case-1	16	0.15	10000	26000	0.04	0.11	0.37	3577	90.00
Case-2	15	0.15	10000	40298	0.03	0.11	0.33	4003	89.97
Case-3	14	0.15	10000	41505	0.04	0.11	0.32	4060	88.74
Case-4	13	0.15	10000	51000	0.03	0.11	0.31	4332	86.86
Case-5	12	0.15	10000	59000	0.03	0.11	0.30	4609	84.72
Case-6	11	0.15	10000	81500	0.03	0.11	0.28	5417	82.31
Case-7	10	0.15	10000	60000	0.03	0.11	0.29	4761	79.09

Figure 8.1 to 8.4, are the illustration of CO_2 removal efficiency, energy demand by reboiler, and MEA demand dependency on the number of stages in the absorber column respectively.

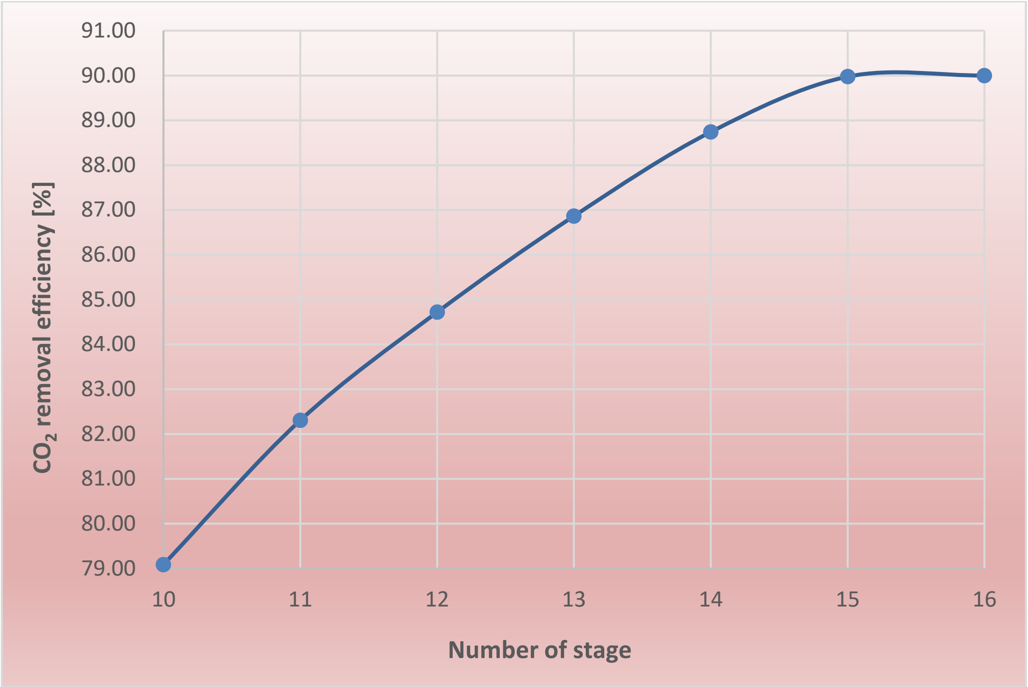

Figure 8.1: CO_2 removal efficiency dependency on the number of stages in the absorber column.

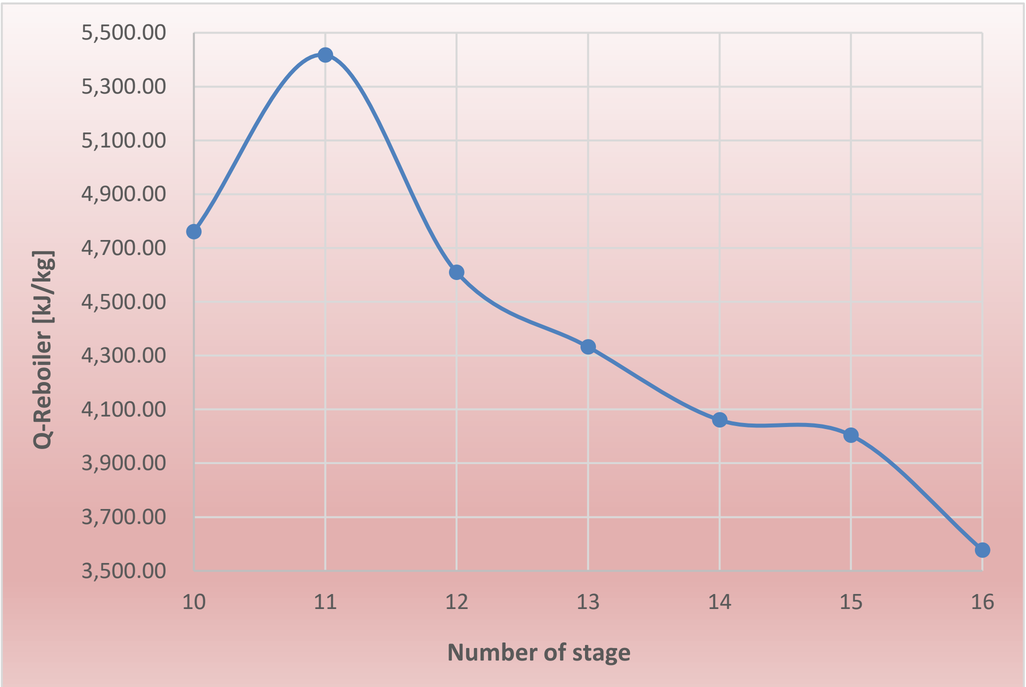

Figure 8.2: Energy demand dependency in the reboiler on the number of stages in the absorber column.

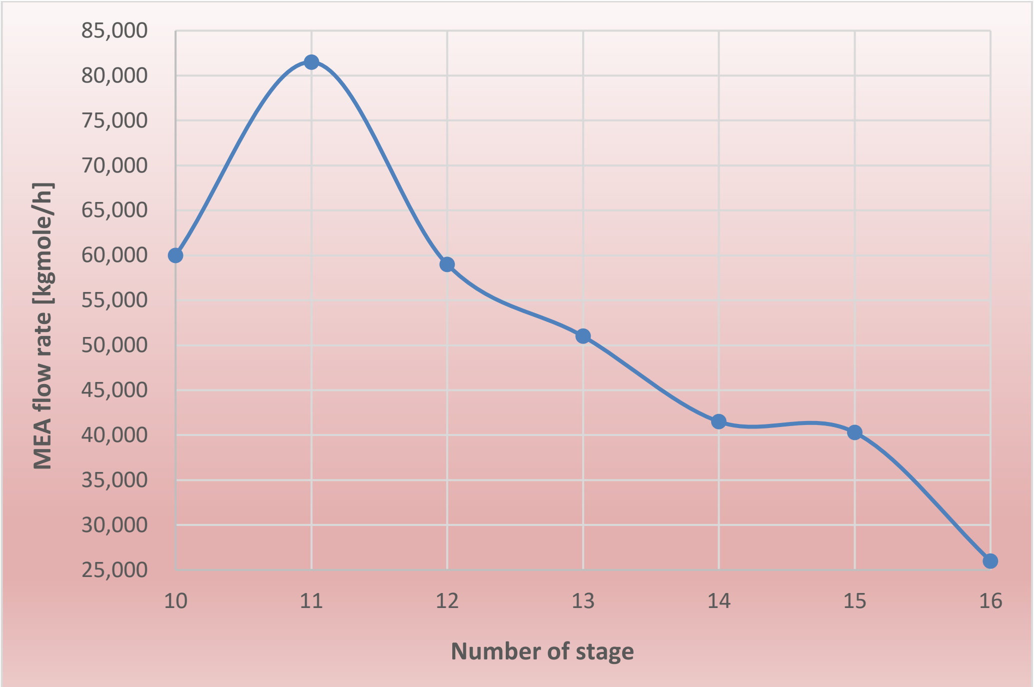

Figure 8.3: MEA requirement dependency on the number of stages in the absorber column.

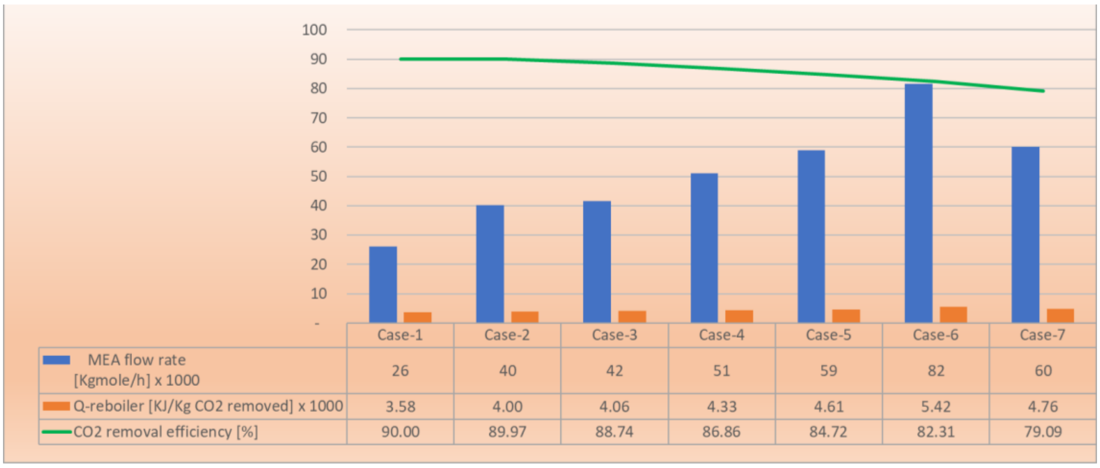

Figure 8.4: CO_2 removal efficiency, energy and MEA demand dependency on number of packing stages in the absorber column.

8.1.2 Minimum approach temperature of 15 degree

Same approach and assumptions as the one for minimum approach temperature (ΔT_{min}) of 10 °C has applied for the minimum approach temperature (ΔT_{min}) of 15 °C. Table 8.2 shows the MEA and energy demand, and also α ratio for all seven case studies.

Table 8.2:Sensitivity analysis for varying number of stages for minimum approach temperature of 15 °C.

Cases	**No. of stage**	**Stage Efficiency**	**Inlet flue gas flow rate**	**MEA flow rate**	**CO_2 content of bottom outlet**	**Amine content of bottom outlet**	$\boldsymbol{\alpha}$	$\boldsymbol{Q_{Reboiler}}$	**CO_2 removal efficiency**
Unit	[-]	[-]	$\left[\frac{kgmol}{h}\right]$	$\left[\frac{kgmol}{h}\right]$	[-]	[-]	[-]	$\left[\frac{kJ}{kg}\right]$	[%]
Case-1	16	0.15	10000	26500	0.03	0.08	0.37	4014	89.96
Case-2	15	0.15	10000	51319	0.02	0.08	0.29	5048	89.97
Case-3	14	0.15	10000	56191	0.03	0.10	0.30	5444	88.74
Case-4	13	0.15	10000	56556	0.03	0.09	0.29	5442	86.86
Case-5	12	0.15	10000	60148	0.03	0.11	0.30	5865	84.72
Case-6	11	0.15	10000	81750	0.03	0.11	0.28	7160	82.31
Case-7	10	0.15	10000	63500	0.02	0.06	0.26	5751	79.09

Figure 8.5 to 8.8, illustrate the CO_2 removal efficiency, energy demand by reboiler, and MEA demand dependency on the number of stages in the absorber column respectively.

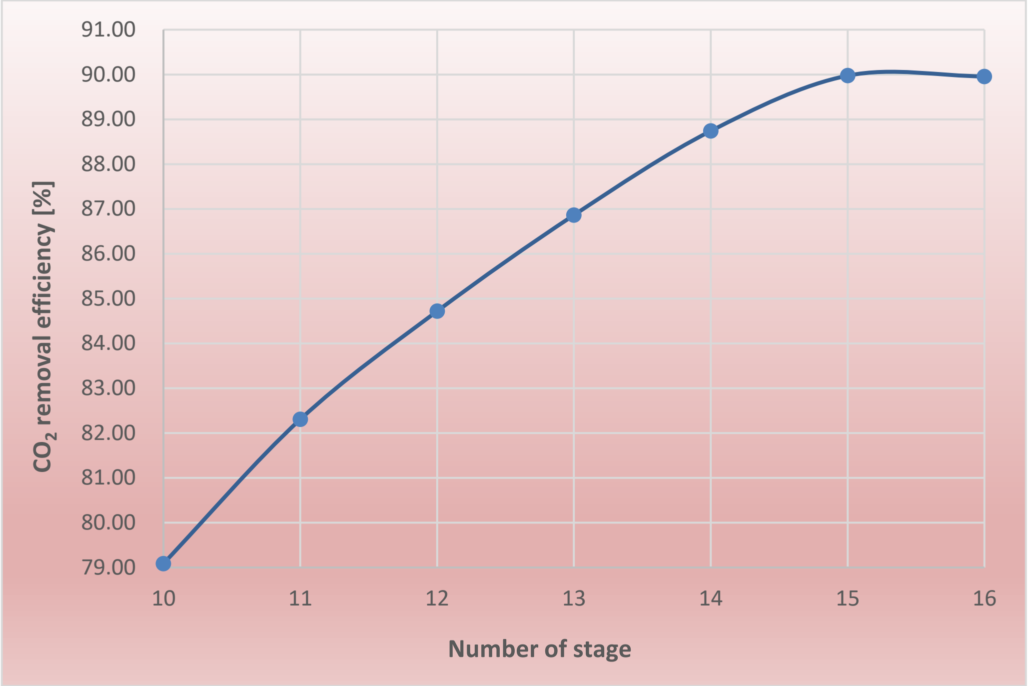

Figure 8.5: CO_2 removal efficiency dependency on the number of stages in the absorber column.

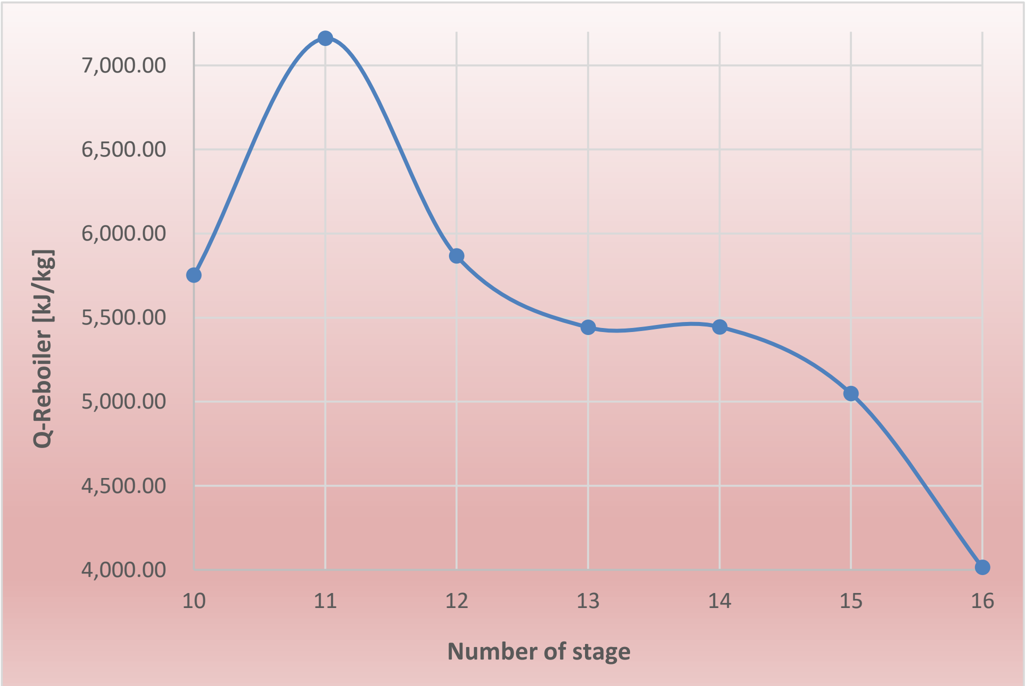

Figure 8.6: Energy demand dependency in the reboiler on the number of stages in the absorber column.

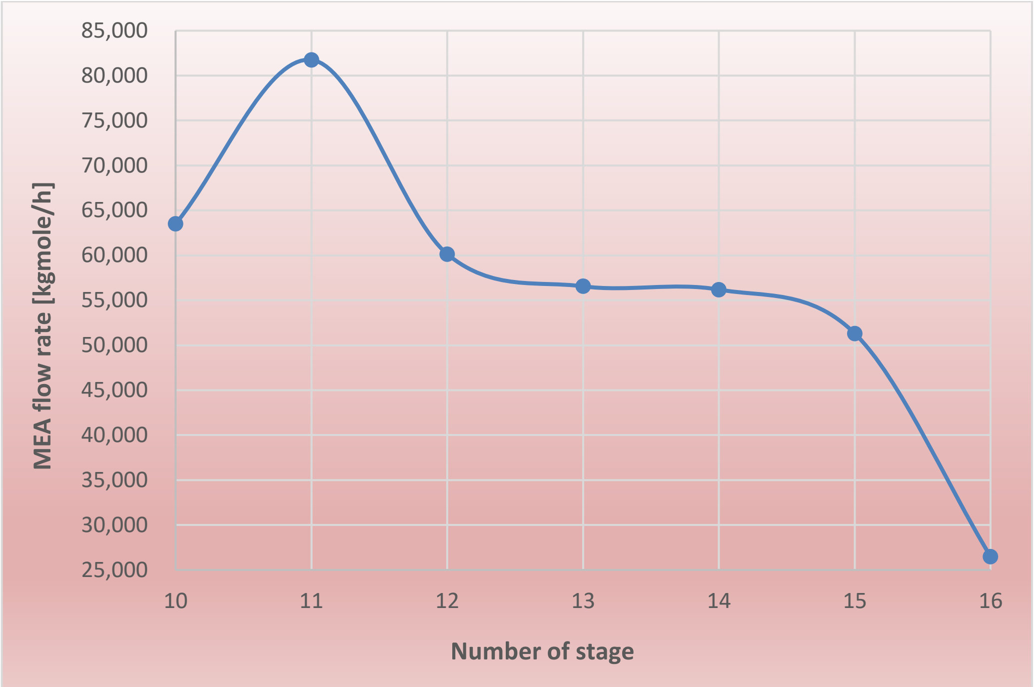

Figure 8.7: MEA requirement dependency on the number of stages in the absorber column.

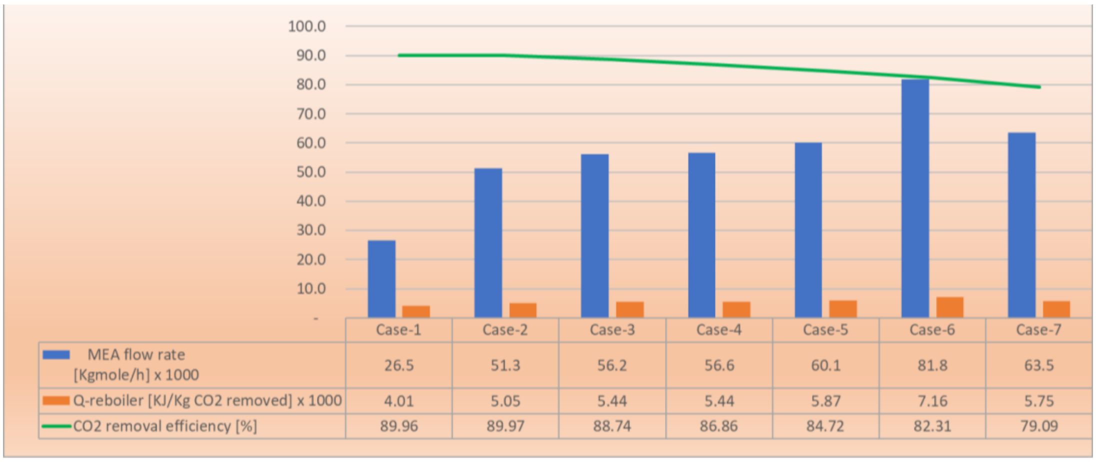

Figure 8.8: CO_2 removal efficiency, energy and MEA demand dependency on number of packing stages in the absorber column.

8.1.3 Minimum approach temperature of 20 degree

Table 8.3 shows the MEA and energy demand, and also α ratio for all seven case studies in case of minimum approach temperature (ΔT_{min}) of 20 °C. Like the first case (ΔT_{min} = 10 °C), all assumptions and approaches are the same.

Table 8.3: Sensitivity analysis for varying number of stages for minimum approach temperature of 20 °C.

Cases	No. of stage	Stage Efficiency	Inlet flue gas flow rate	MEA flow rate	CO_2 content of bottom outlet	Amine content of bottom outlet	α	$Q_{Reboiler}$	CO_2 removal efficiency
Unit	[-]	[-]	$\left[\frac{kgmol}{h}\right]$	$\left[\frac{kgmol}{h}\right]$	[-]	[-]	[-]	$\left[\frac{kJ}{kg}\right]$	[%]
Case-1	16	0.15	10000	46283	0.01	0.05	0.31	5851	90.13
Case-2	15	0.15	10000	52202	0.02	0.05	0.29	6187	89.97
Case-3	14	0.15	10000	56412	0.02	0.06	0.28	6504	88.74
Case-4	13	0.15	10000	57774	0.02	0.06	0.28	6677	86.86
Case-5	12	0.15	10000	69107	0.02	0.07	0.26	7660	84.72
Case-6	11	0.15	10000	94064	0.02	0.07	0.24	9793	82.31
Case-7	10	0.15	10000	60000	0.02	0.06	0.26	7153	79.09

Figure 8.9 to 8.12, illustrate the CO_2 removal efficiency, energy demand by reboiler, and MEA demand dependency on the number of stages in the absorber column respectively.

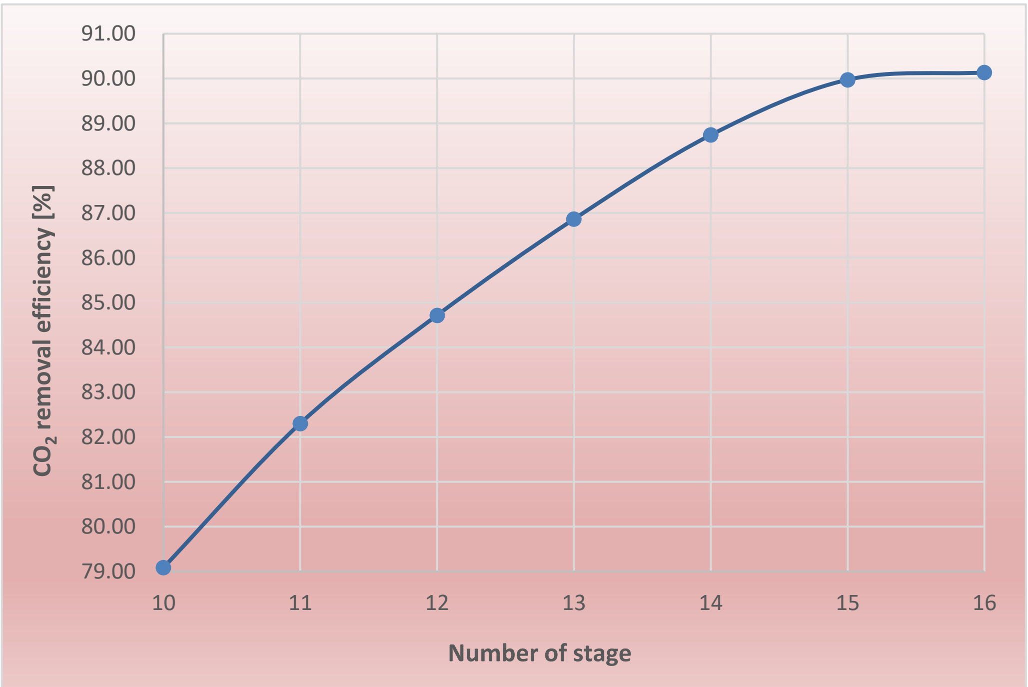

Figure 8.9: CO_2 removal efficiency dependency on the number of stages in the absorber column.

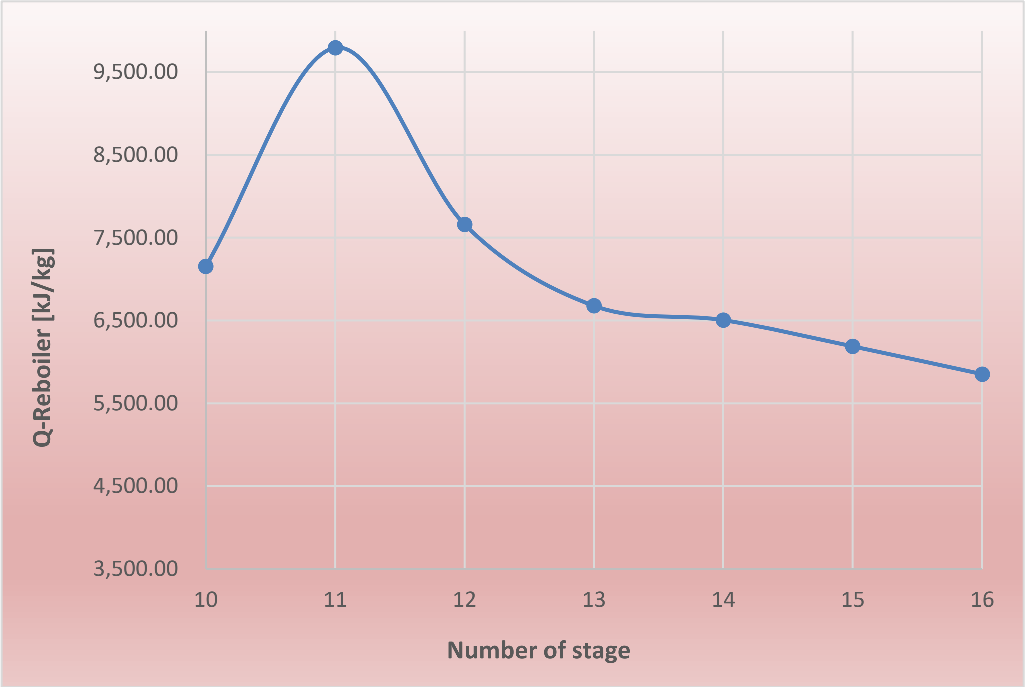

Figure 8.10: Energy demand dependency in the reboiler on the number of stages in the absorber column.

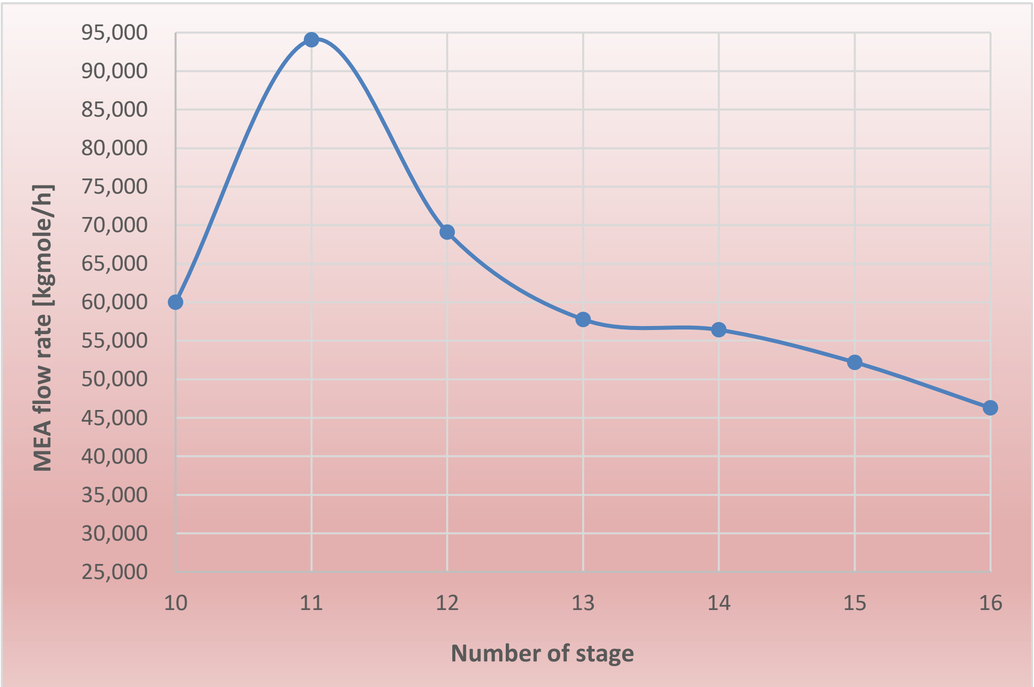

Figure 8.11: MEA requirement dependency on the number of stages in the absorber column.

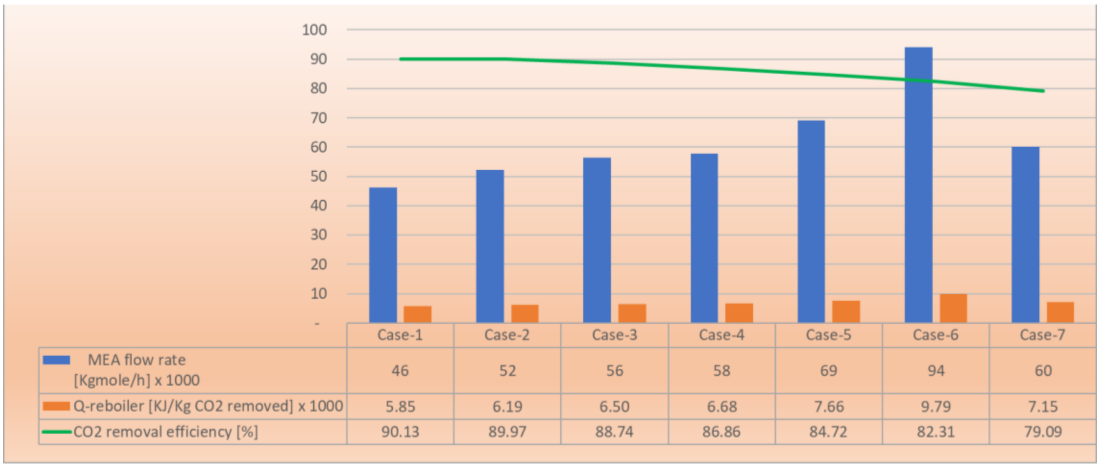

Figure 8.12: CO_2 removal efficiency, energy and MEA demand dependency on number of packing stages in the absorber column.

8.2 Dimensioning sensitivity analysis for case studies

To have a better overview on all case studies, a dimensioning has been performed on main and the most expensive equipment and items, including absorber column, main heat exchanger, steam consumption, and the electricity demand.

8.2.1 Absorber

Table 8.4, shows the dimensioning performed for all seven case studies when minimum approach temperature is considered to be 10 °C. Obviously, by decreasing the number of stages in the absorber column, absorber column size will reduce.

Table 8.4:Absorber column dimensioning of case studies for minimum approach temperature of 10 °C.

ΔT_{min} = 10 °C:									
Cases	**Vapor Volume flow rate [m^3/h]**	**Velocity [m/s]**	**Number of stages**	**Packing Height [m]**	**Cross section Area [m^2]**	**Diameter of Absorber [m]**	**Total Packing Height [m]**	**Absorber Height**	
								Value [m]	**Reference**
Case-1	283089	2	16	1	39.32	7.07	16	50	[7]
Case-2	255420	2	15	1	35.48	6.72	15	40	[27], [31]
Case-3	254520	2	14	1	35.35	6.70	14	40	Assumed
Case-4	245622	2	13	1	34.11	6.59	13	40	Assumed
Case-5	246857	2	12	1	34.29	6.60	12	40	Assumed
Case-6	244102	2	11	1	33.90	6.57	11	40	Assumed
Case-7	247507	2	10	1	34.38	6.61	10	40	[11]

Figure 8.13, is an illustration of absorber cross sectional area dependency on the number of stages in the absorber column for minimum approach temperature of 10 °C.

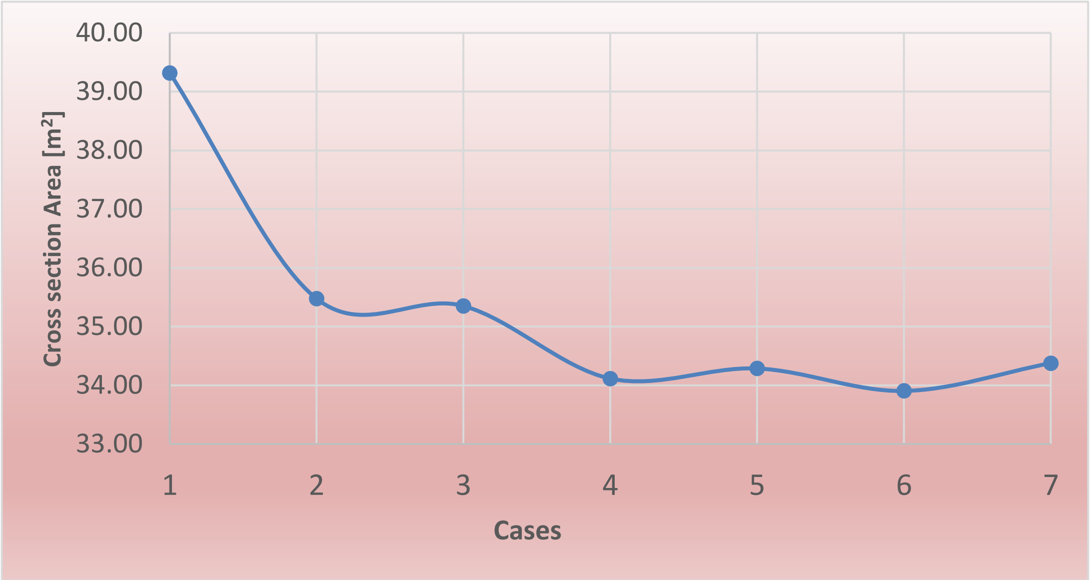

Figure 8.13: Absorber cross sectional area dependency on number of packing stages for ΔT_{min} of 10 °C.

Dimensioning performed for all seven case studies for minimum approach temperature of 15 °C has been reported in table 8.5. Same trend as previous case, ΔT_{min} of 10 °C, is observed.

Table 8.5:Absorber column dimensioning of case studies for minimum approach temperature of 15 °C.

ΔT_{min} = 15 °C:									
Cases	**Vapor Volume flow rate [m^3/h]**	**Velocity [m/s]**	**Number of stages**	**Packing Height [m]**	**Cross section Area [m^2]**	**Diameter of Absorber [m]**	**Packing Height [m]**	**Absorber Height**	
								Value [m]	**Reference**
Case-1	282449	2	16	1	39.23	7.06	16	50	[7]
Case-2	264247	2	15	1	36.70	6.83	15	40	[27], [31]
Case-3	246305	2	14	1	34.21	6.59	14	40	Assumed
Case-4	253849	2	13	1	35.26	6.70	13	40	Assumed
Case-5	247967	2	12	1	34.44	6.62	12	40	Assumed
Case-6	247208	2	11	1	34.33	6.61	11	40	Assumed
Case-7	251301	2	10	1	34.90	6.66	10	40	[11]

For the minimum approach temperature of 15 °C, dependency of absorber cross sectional area on the number of packing stages has been shown in Figure 8.14.

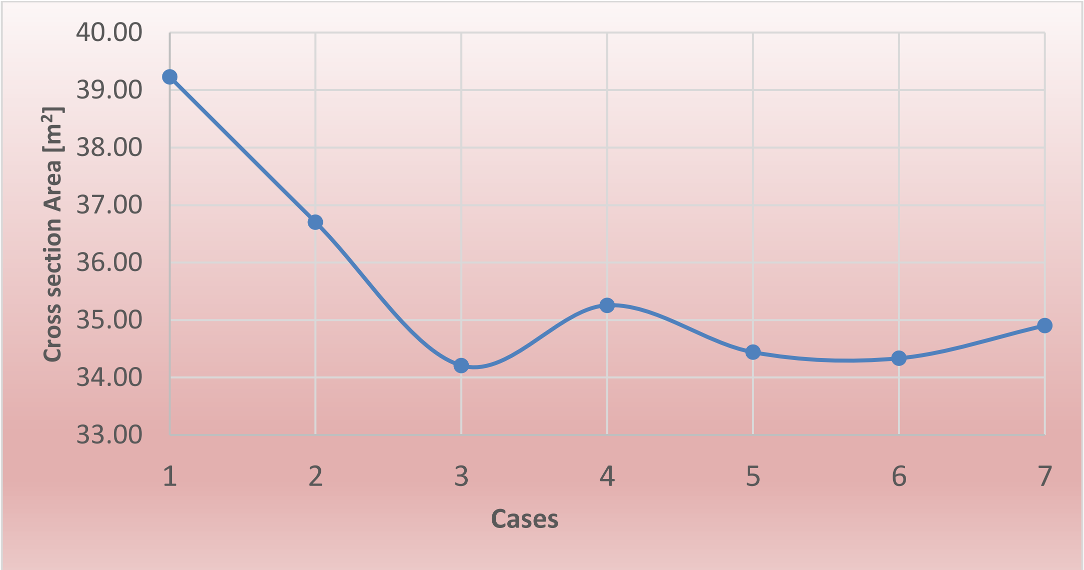

Figure 8.14: Absorber cross sectional area dependency on number of packing stages for ΔT_{min} of 15 °C.

Finally for the minimum approach temperature of 20 °C, performed dimensioning for case studies have been reported in table 8.6. Also, afterward, figure 8.15 displays the dependency of absorber cross sectional area on the number of packing stages.

Table 8.6:Absorber column dimensioning of case studies for minimum approach temperature of 20 °C.

ΔT_{min} = 15 [°C]:									
Cases	**Vapor Volume flow rate [m^3/h]**	**Velocity [m/s]**	**Number of stages**	**Packing Height [m]**	**Cross section Area [m^2]**	**Diameter of Absorber [m]**	**Packing Height [m]**	**Absorber Height**	
								Value [m]	**Reference**
Case-1	293930	2	16	1	40.82	7.20	16	50	[7]
Case-2	269961	2	15	1	37.49	6.90	15	40	[27], [31]
Case-3	258616	2	14	1	35.92	6.76	14	40	Assumed
Case-4	256937	2	13	1	35.69	6.74	13	40	Assumed
Case-5	252672	2	12	1	35.09	6.68	12	40	Assumed
Case-6	246820	2	11	1	34.28	6.60	11	40	Assumed
Case-7	247863	2	10	1	34.43	6.62	10	40	[11]

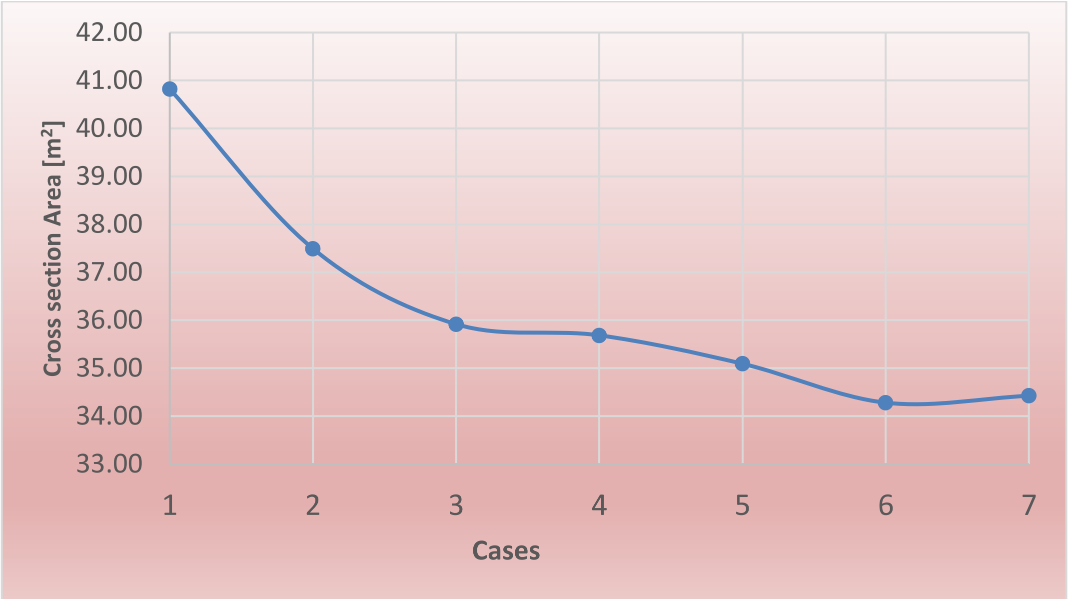

Figure 8.15: Absorber cross sectional area dependency on number of packing stages for ΔT_{min} of 20 °C.

8.2.2 Main heat exchanger

The main heat exchanger is the second most expensive equipment in the CO_2 capture process plant. Table 8.7, have shown the result of dimensioning for defined seven cases, in case of the minimum approach temperature of 10, 15 and 20 °C respectively. In all cases, the heat transfer coefficient (U), has assumed to be 500 $\frac{W}{m^2.K}$. As can be seen, lower packing stages in the absorber, requires considerably higher area in the main heat exchanger.

Table 8.7:Lean/rich heat exchanger dimensioning of case studies for minimum approach temperature of 10 °C.

Cases	**Heat transfer coefficient (U)**	**Q × 10⁸**	**LMTD**	**HEX Area**
Unit	$\left[\frac{W}{m^2.K}\right]$	$\left[\frac{kJ}{h}\right]$	$[K]$	$[m^2]$
Case-1	500	1.35	10.66	7012
Case-2	500	2.11	10.76	10872
Case-3	500	2.13	10.77	10968
Case-4	500	2.76	10.85	14152
Case-5	500	3.14	10.88	16027
Case-6	500	4.53	10.99	22909
Case-7	500	3.20	10.89	16308

Figure 8.16 shows this increasing trend in area demand for the main heat exchanger per reducing number of stages in the absorber.

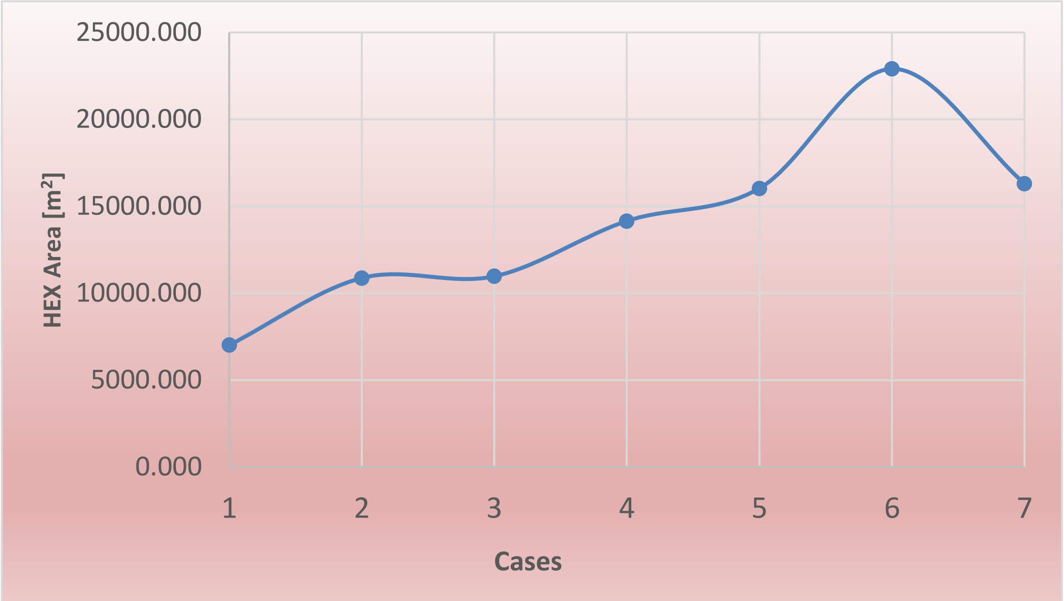

Figure 8.16: HEX area demand dependency on number of stages in the absorber for ΔT_{min} of 10 °C.

Table 8.8 is related to the dimensioning of main heat exchanger for case studies when the minimum approach temperature is selected to be 15 °C. In this case also, same increasing trend has followed in the area demand in the main heat exchanger per deduction of stages in the absorber and figure 8.17 is an illustration of it.

Table 8.8: Lean/rich heat exchanger dimensioning of case studies for minimum approach temperature of 15 °C.

Cases	**Heat transfer coefficient (U)**	$\mathbf{Q \times 10^8}$	**LMTD**	**HEX Area**
Unit	$\left[\frac{W}{m^2.K}\right]$	$\left[\frac{kJ}{h}\right]$	$[K]$	$[m^2]$
Case-1	500	1.25	15.48	4498
Case-2	500	2.36	15.57	8422
Case-3	500	2.73	15.66	9674
Case-4	500	2.62	15.62	9312
Case-5	500	2.93	15.70	10354
Case-6	500	4.12	15.79	14487
Case-7	500	3.02	15.61	10755

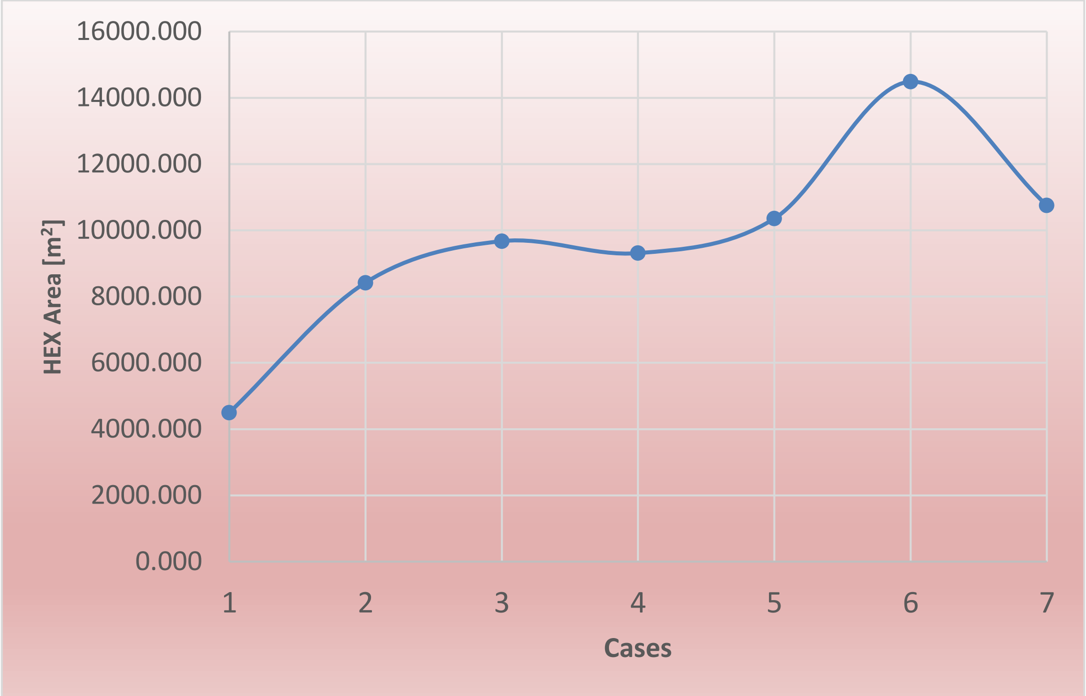

Figure 8.17: HEX area demand dependency on number of stages in the absorber for ΔT_{min} of 15 °C.

In case of minimum approach temperature of 20 °C, performed dimensioning for case studies have been reported in table 8.9.

Table 8.9: Lean/rich heat exchanger dimensioning of case studies for minimum approach temperature of 20 °C.

Cases	**Heat transfer coefficient (U)**	**Q × 10⁸**	**LMTD**	**HEX Area**
Unit	$\left[\frac{W}{m^2.K}\right]$	$\left[\frac{kJ}{h}\right]$	$[K]$	$[m^2]$
Case-1	500	1.80	20.74	4823
Case-2	500	2.05	20.77	5475
Case-3	500	2.29	20.81	6109
Case-4	500	2.36	20.83	6293
Case-5	500	2.89	20.89	7693
Case-6	500	4.31	21.01	11393
Case-7	500	2.65	20.91	7047

Considering ΔT_{min} of 20 °C, dependency of required heat exchanger area on the number of packing stages has shown in figure 8.18.

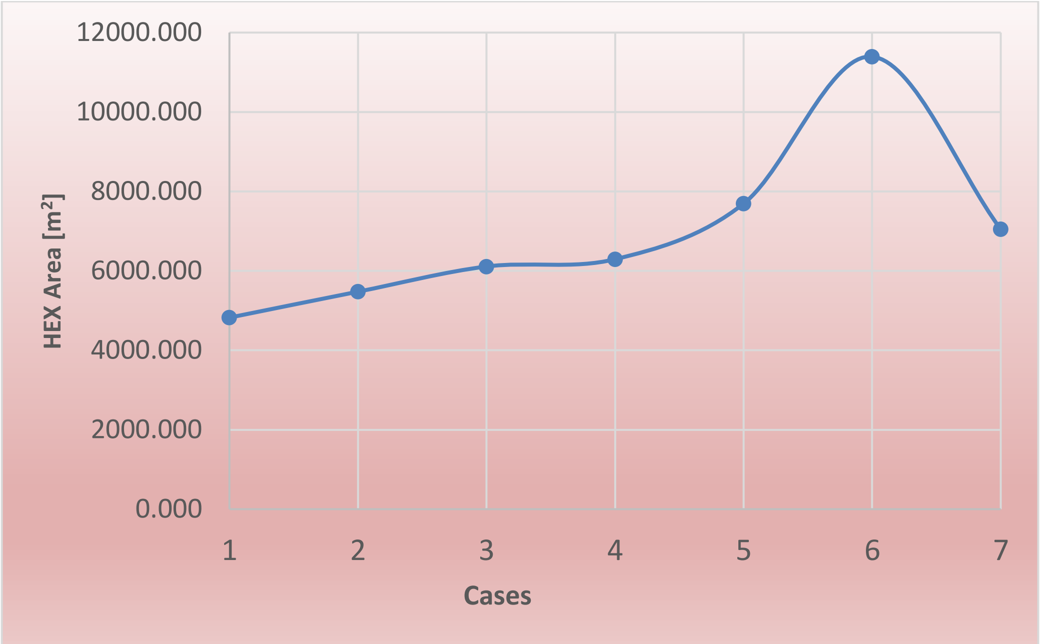

Figure 8.18: HEX area demand dependency on number of stages in the absorber for ΔT_{min} of 20 °C.

8.2.3 Steam consumption

Steam is the most expensive parameter in the OPEX, so then, the steam consumption for all case studies have been evaluated. Table 8.10 contains the reboiler steam consumption for 7 cases while the minimum approach temperature is set to be 10 °C.

As can be understood from the result, decreasing number of packing stages in the column, leads to considerably higher steam demand by the reboiler unit except for case-7, that its low efficiency makes this case as an exception.

Table 8.10: Steam consumption of case studies for minimum approach temperature of 10 °C.

Cases	**Reboiler Consumption**	**Reboiler Consumption**
Unit	$\left[\frac{kJ}{h}\right]$	$[kW]$
Case-1	8.70E+07	24154
Case-2	1.11E+08	30783
Case-3	1.12E+08	31033
Case-4	1.24E+08	34397
Case-5	1.31E+08	36492
Case-6	1.52E+08	42142
Case-7	1.28E+08	35652

Figure 8.19 illustrates the trend of listed steam demand by the reboiler in the table 8.10.

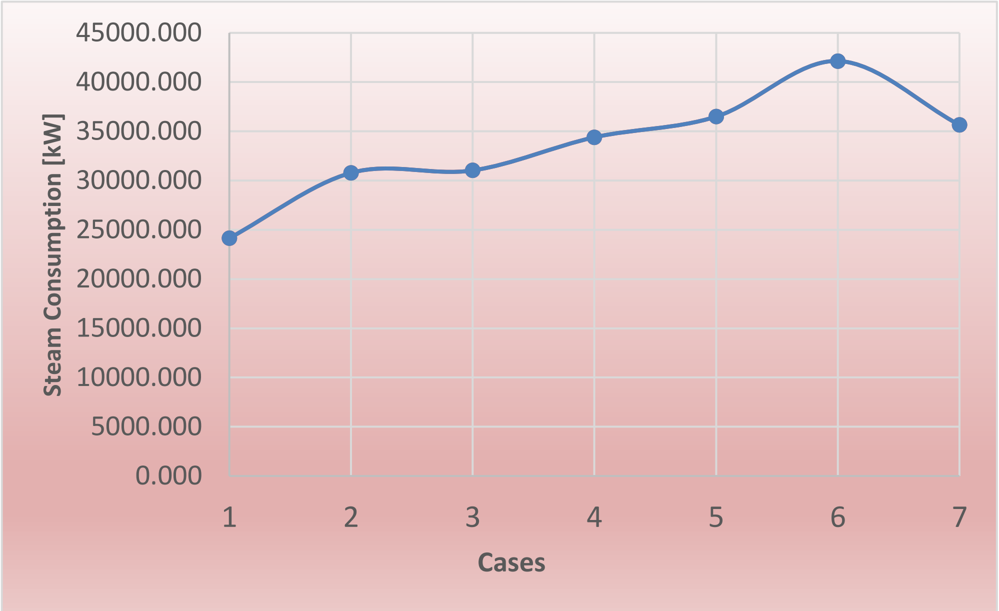

Figure 8.19: Steam consumption of case studies for minimum approach temperature of 10 °C.

The next case is related to when minimum approach temperature is assumed to be 15 °C. In this approach also, higher steam demand will result if the number of packing stages will reduce which has been listed in table 8.11. Also, to have a better overview, figure 8.20 illustrates the steam demand for different case studies.

Table 8.11: Steam consumption of case studies for minimum approach temperature of 15 °C.

Cases	**Reboiler Consumption**	**Reboiler Consumption**
Unit	$\left[\frac{kJ}{h}\right]$	$[kW]$
Case-1	9.27E+07	25756
Case-2	1.40E+08	38876
Case-3	1.45E+08	40291
Case-4	1.45E+08	40327
Case-5	1.46E+08	40678
Case-6	1.74E+08	48281
Case-7	1.52E+08	42166

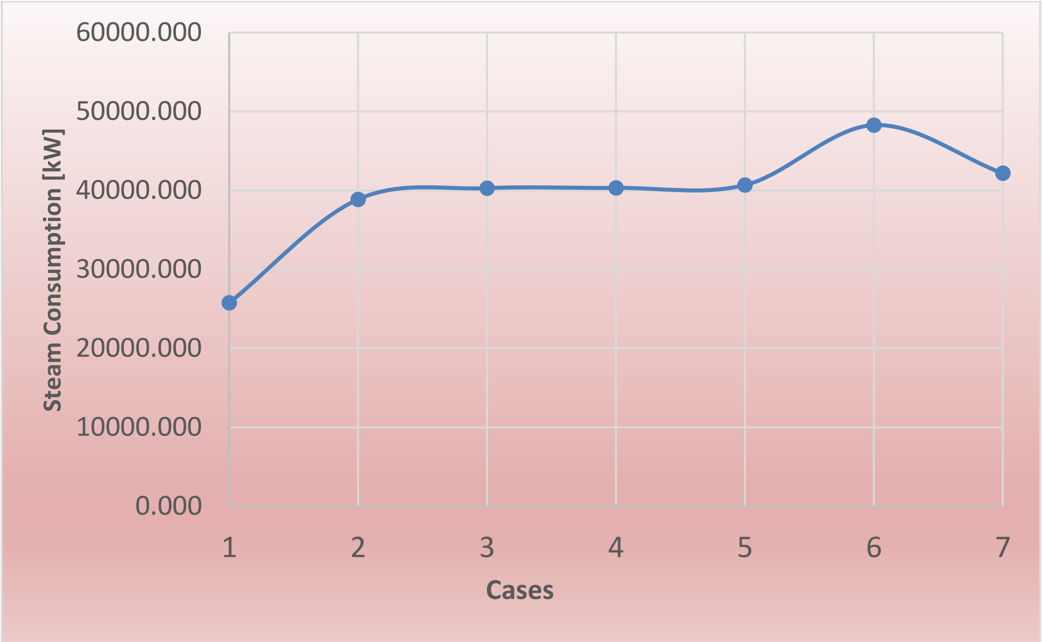

Figure 8.20: Steam consumption of case studies for minimum approach temperature of 15 °C.

Same trend also has been followed when the minimum approach temperature is set to be 20 °C as reported in table 8.12 and an illustration presented in figure 8.21.

Table 8.12: Steam consumption of case studies for minimum approach temperature of 20 °C.

Cases	**Reboiler Consumption**	**Reboiler Consumption**
Unit	$\left[\frac{kJ}{h}\right]$	$[KW]$
Case-1	1.37E+08	38153
Case-2	1.52E+08	42128
Case-3	1.60E+08	44498
Case-4	1.62E+08	45005
Case-5	1.81E+08	50322
Case-6	2.22E+08	61571
Case-7	1.61E+08	44772

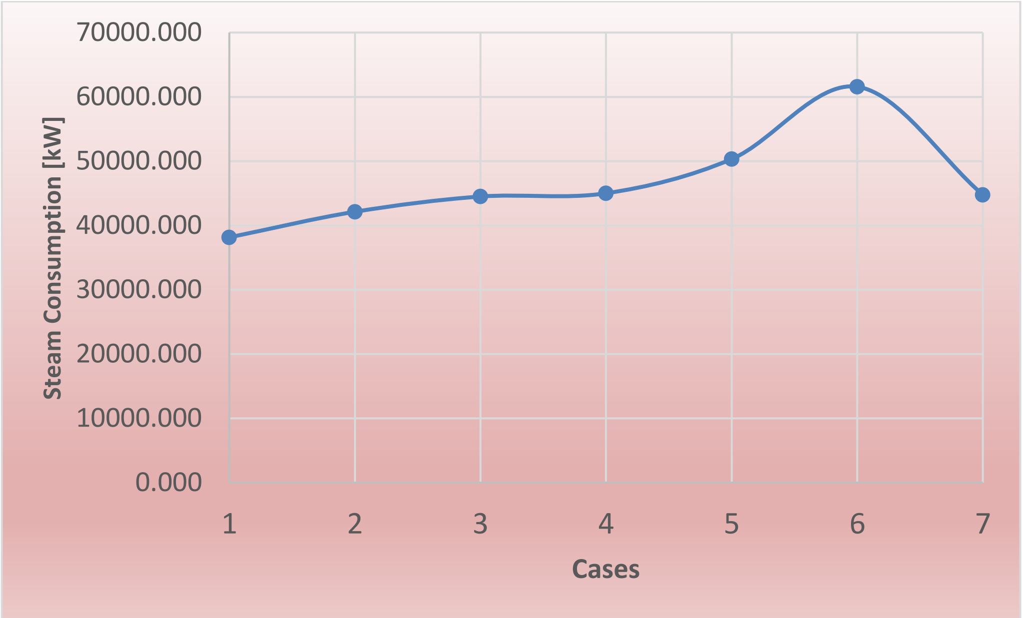

Figure 8.21: Steam consumption of case studies for minimum approach temperature of 20 °C.

8.2.4 Electricity consumption

The second most expensive parameter in the OPEX is the electricity. So, in this regard, the electricity consumption for the defined process for all case studies have been investigated. In the defined CO_2 capture process plant, consumers are including pumps (P-100 and P-101), and also the fan. Similar to other parameters mentioned in previous parts, the evaluation for electricity demand will be conducted for all three cases of minimum approach temperature of 10, 15, and 20 °C. The responses are shown in table 8.13 to 8.15 and also figure 8.22 to 8.24.

Based on the resulted trends for all 3 cases, generally reduction in number of stages in the absorber column led to higher electricity demand for pumps and fan.

Table 8.13: Energy demand in reboiler of case studies for minimum approach temperature of 10 °C.

Cases	Fan	Pump P-100	Pump P-101	Total [kW]
	Q [kW]	Q [kW]	Q [kW]	
Case-1	921	19	22	963
Case-2	921	29	35	986
Case-3	921	30	36	988
Case-4	921	37	43	1002
Case-5	921	43	50	1015
Case-6	921	59	70	1052
Case-7	921	43	52	1018

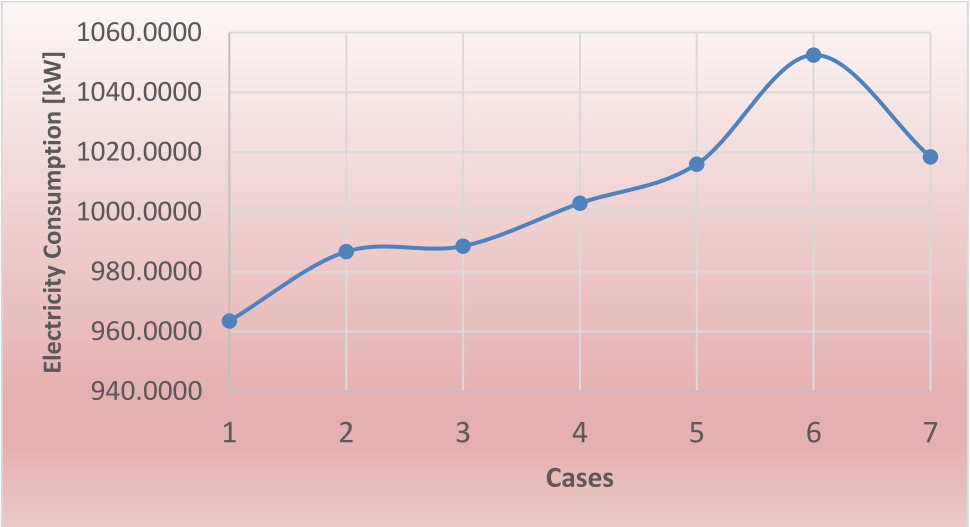

Figure 8.22: Energy demand in reboiler of case studies for minimum approach temperature of 10 °C.

Table 8.14: Energy demand in reboiler of case studies for minimum approach temperature of 15 °C.

Cases	Fan Q [kW]	Pump P-100 Q [kW]	Pump P-101 Q [kW]	Total [kW]
Case-1	921	18	21	961
Case-2	921	35	41	999
Case-3	921	40	48	1010
Case-4	921	40	47	1009
Case-5	921	44	52	1018
Case-6	921	59	71	1052
Case-7	921	43	50	1015

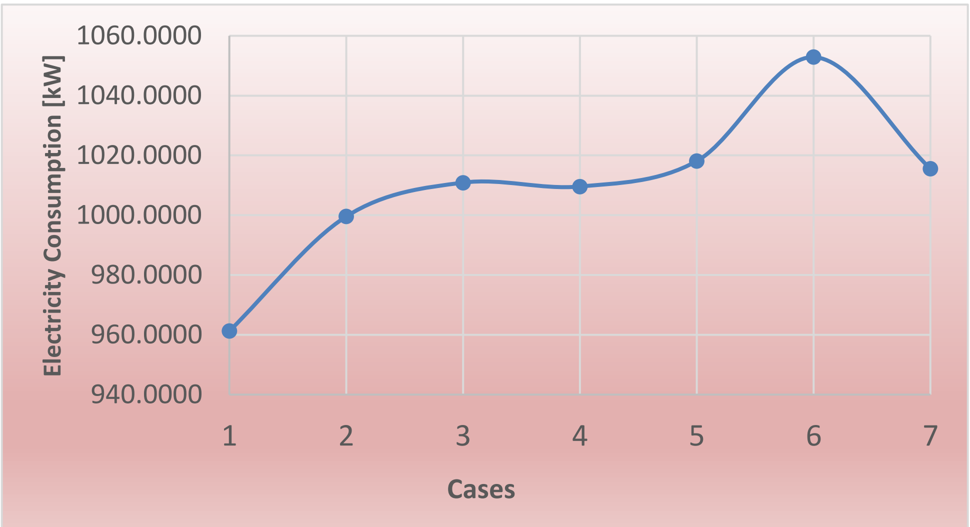

Figure 8.23: Energy demand in reboiler of case studies for minimum approach temperature of 15 °C.

Table 8.15: Energy demand in reboiler of case studies for minimum approach temperature of 20 °C.

Cases	Fan	Pump P-100	Pump P-101	Total [kW]
	Q [kW]	Q [kW]	Q [kW]	
Case-1	921	30	35	987
Case-2	921	34	40	997
Case-3	921	38	44	1004
Case-4	921	39	45	1007
Case-5	921	47	56	1025
Case-6	921	64	76	1062
Case-7	921	40	47	1009

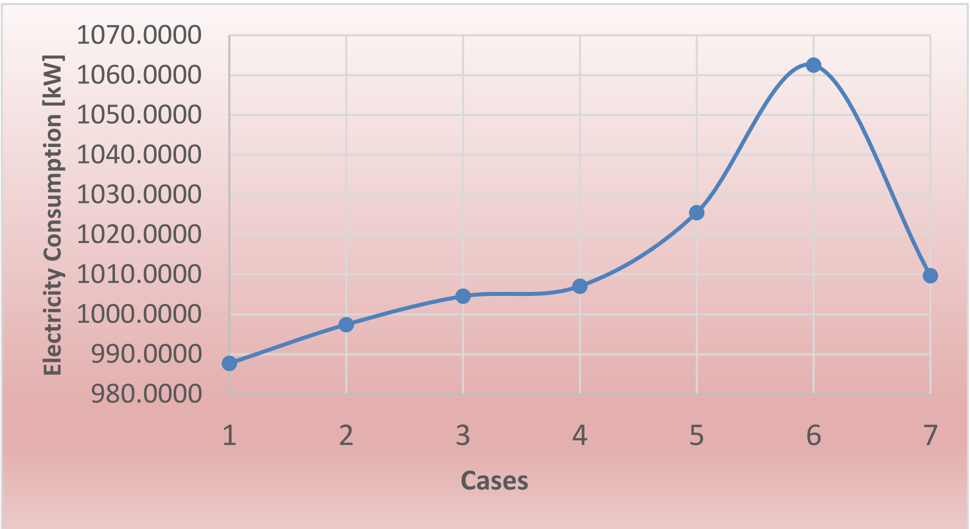

Figure 8.24: Energy demand in reboiler of case studies for minimum approach temperature of 20 °C.

9 Discussion

9.1 Comparison of case studies for minimum approach temperature 10, 15 and 20 degrees

In this chapter a comparison analysis on energy demand in the reboiler unit for removing one kilogram CO_2, the required MEA amine flow rate, and also the ratio of CO_2 content to amine content in the bottom outlet of absorber column (α coefficient), for 3 cases of minimum approach temperature (ΔT_{min}) of 10, 15, and 20 degrees Celsius have been conducted.

9.2 Energy demand

Figure 9.1 illustrates the dependency of the energy requirement in the reboiler unit per one kilogram CO_2 captured during the process, for all cases of the minimum approach temperature (ΔT_{min}), 10, 15, and 20 °C. As can be seen from the graph, overall, higher temperature difference resulted in considerably higher energy demand in the reboiler unit.

The lowest and the most economical energy demand is related to the base case with 16 packing stages in the absorber unit, in all 3 cases of minimum approach temperature (ΔT_{min}). The required energy in reboiler for the case-1 with 16 stages for minimum approach temperature of 10 °C is 3577 $\frac{kJ}{kg\ of\ CO_2\ removed}$, for ΔT_{min} of 15 and 20 °C are 4014 and 5851 $\frac{KJ}{Kg\ of\ CO_2\ removed}$ respectively. The highest energy demand is related to the case-6 with 11 packing stages in the absorber unit, which results 5417, 7160, and 9793 $\frac{kJ}{kg\ of\ CO_2\ removed}$ for ΔT_{min} of 10, 15 and 20 °C respectively.

Case-7 is an exception in following the trend since the CO_2 removal efficiency is considerably low.

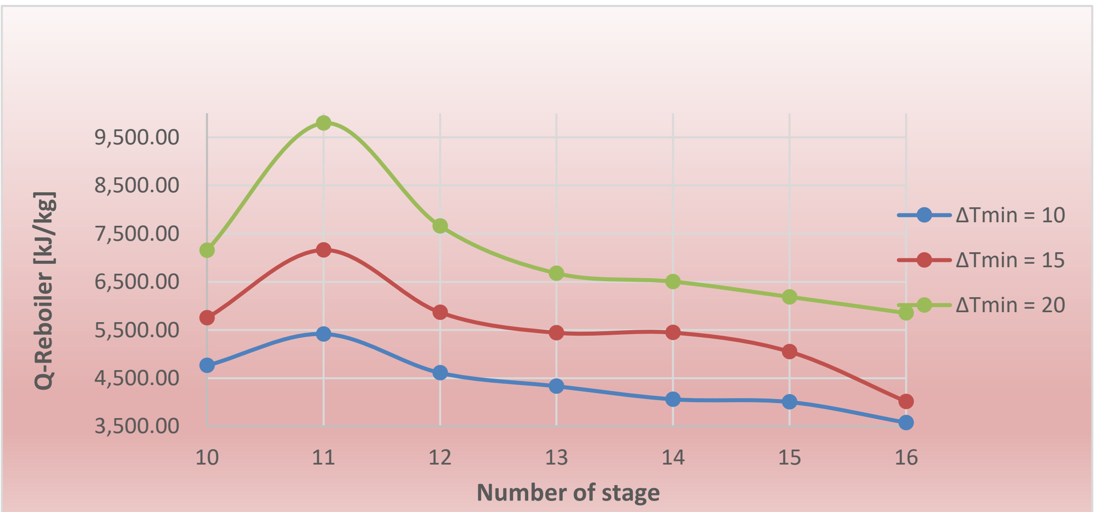

Figure 9.1: Energy requirement per CO_2 captured for ΔT_{min} of 10, 15, and 20 °C.

9.3 MEA requirement

Figure 9.2 shows the trend of MEA amine requirement dependency on packing stages in the absorber column for three cases of minimum approach temperature of 10, 15, and 20 °C.

As can be seen in the diagram, the lowest MEA requirement assigned to the case-1 with 16 packing stages which resulted in 26000, 26500, and 46283 kgmole of MEA amine per hour for ΔT_{min} of 10, 15 and 20 °C respectively. While the highest MEA requirement is related to the case-6 with 11 stages in the absorber column that requires 81500, 81750, 94064 $\frac{kgmole}{h}$ for minimum approach temperature of 10, 15, and 20 °C respectively.

Likewise of the energy demand, case-7 because of its low removal efficiency is an exception in following the trends. Also, in the graph related to the minimum approach temperature of 15 °C, there are two jumps in case-2 with 15 stages and case-3 with 14 stages. The reason is not clear. One possible explanation is inaccuracy in the simulation.

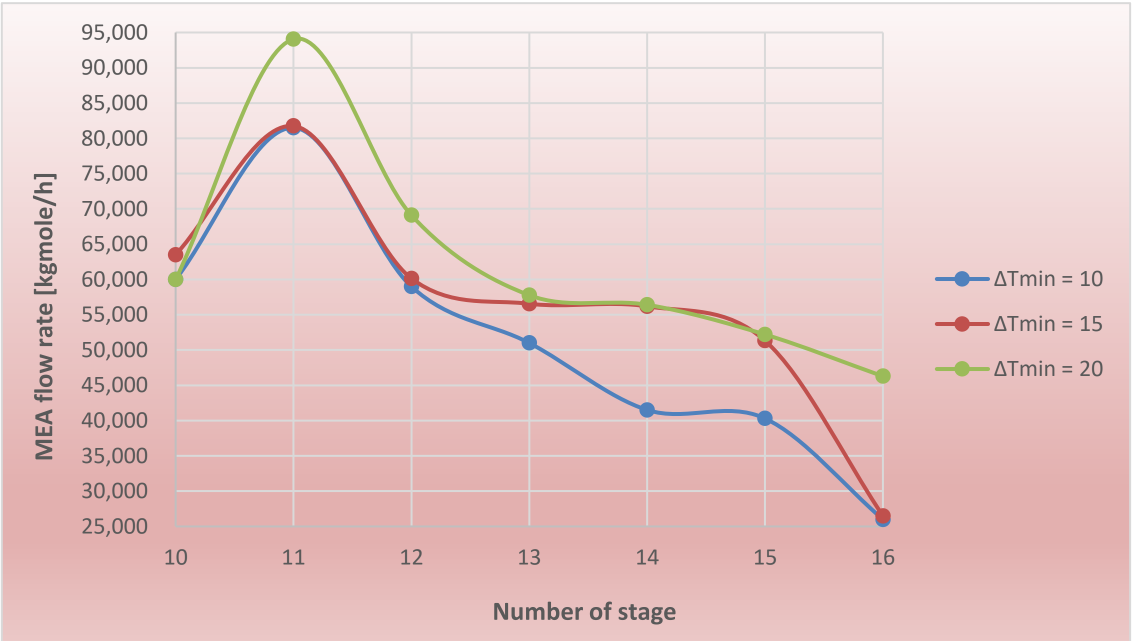

Figure 9.2: MEA requirement for ΔT_{min} of 10, 15, and 20 °C.

9.4 α coefficient

The ratio of CO_2 to amine content of bottom outlet of the absorber column (α) is shown in figure 9.3 for the cases of minimum approach temperature of 10, 15, and 20 °C in lean/rich heat exchanger. The less the minimum approach temperature difference, the higher ratio of α will obtain so that the highest ratio is belonged to the minimum approach temperature of 10 °C with amount of 0.371. The second and third highest ratio is followed by minimum approach temperature of 15 and 20 °C with the value of 0.372, and 0.314.

Generally, the lowest ratio of CO_2 to amine is related to the case-6 with 11 stages for all cases of minimum approach temperature except for minimum approach temperature of 15 °C which instead happened in case-7 with 10 stages. This inconsistency is maybe because of inaccuracy in the simulation. For minimum approach temperature of 10, 15, and 20 °C the lowest ratio is 0.283, 0.260, and 0.242 respectively.

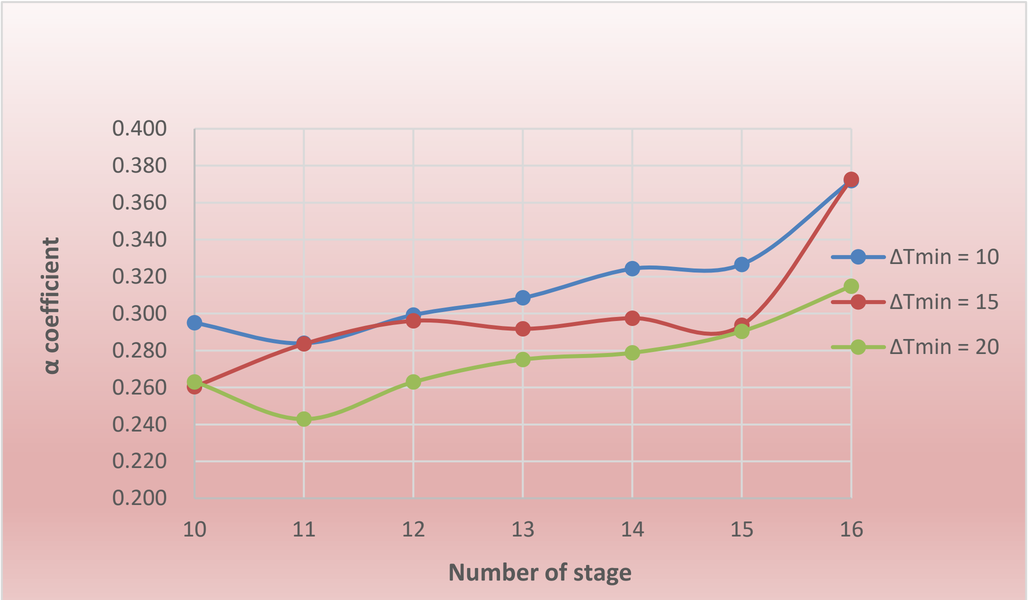

Figure 9.3: Ratio of CO_2 to amine content of absorber bottom outlet for ΔT_{min} of 10, 15, and 20 °C.

9.5 Dimensioning comparison

A comparative analysis on dimensioning of main and the most expensive items in CAPEX and OPEX for 3 cases of minimum approach temperature (ΔT_{min}) of 10, 15, and 20 °C will be covered in this chapter.

9.5.1 Absorber column

The calculated cross-sectional area of the absorber column for number of packing stages from 10 to 16 for three cases of minimum approach temperature (ΔT_{min}) of 10, 15, and 20 °C have been demonstrated in figure 9.4. The higher stages in the absorber results wider cross-sectional area. So, case-1 with 16 stages hit the highest cross area and the case-6 with 11 stages has the lowest cross-sectional area in this trend. Although case-7 with 10 stages has lower number of packing stages, its low efficiency makes this case an exception in these trends.

Regarding the minimum approach temperature (ΔT_{min}), the differences between ΔT_{min} of 10, 15, and 20 °C are not that much significant, however, the case of ΔT_{min} of 10 and 15 °C has eventuated almost the same value of 39.3 m^2 which is the lower cross sectional area demand for the absorber column, and instead, the case of ΔT_{min} of 20 °C requires higher 40.8 m^2 cross-sectional area of in this trend.

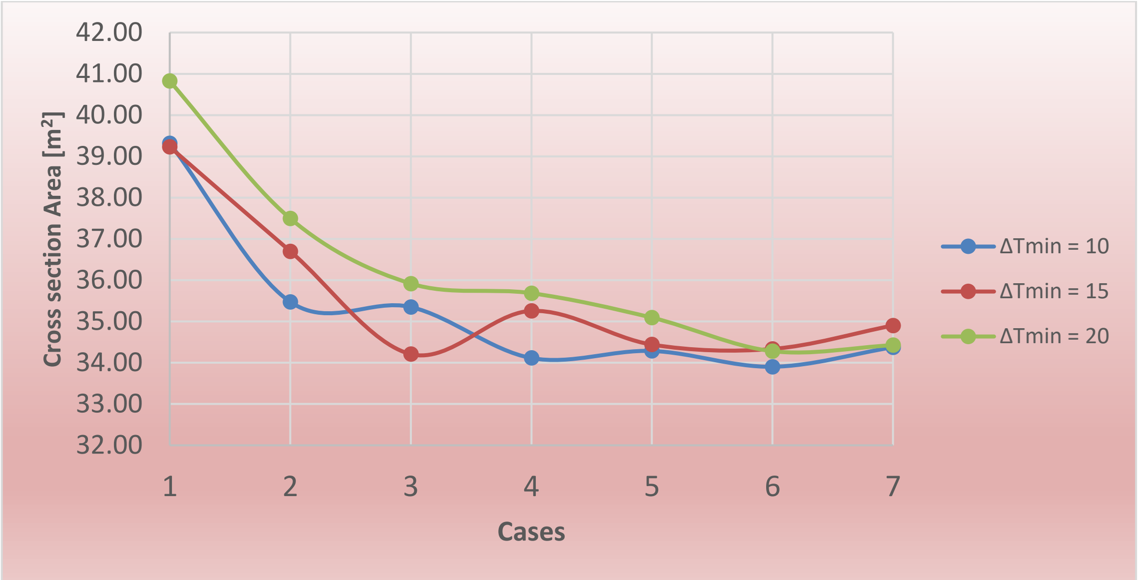

Figure 9.4: Cross-sectional area of the absorber column for ΔT_{min} of 10, 15, and 20 °C.

9.5.2 Lean/Rich heat exchanger area demand

The next most expensive component in CAPEX to evaluate, is lean/rich heat exchanger area which its trend is shown in figure 9.5 for the minimum approach temperature (ΔT_{min}) of 10, 15, and 20 °C.

Obviously, higher temperature difference in the heat exchanger (higher ΔT_{min}) will result in lower area requirement for the main heat exchanger so that the lowest area demand respectively is belonged to the minimum approach temperature (ΔT_{min}) of 20, 15, and 10 °C.

Regarding the dependency of heat exchanger area on the change of the number of stages in the absorber column, the higher stages is ended up with lower area demand. So, case-1 with 16 packing stages resulted in lower area demand in the main heat exchanger with the value of 4823, 4498, and 7012 m^2 for minimum approach temperature (ΔT_{min}) of 20, 15, and 10 °C respectively. Likewise, the highest area required in the main heat exchanger in the defined CO_2 removal process, is related to the case-6 with the value of 22909, 14487, and 11393 m^2 for minimum approach temperature of 10, 15, and 20 °C respectively.

Similar argument on case-7 with 10 stages in the absorber column is valid in this section.

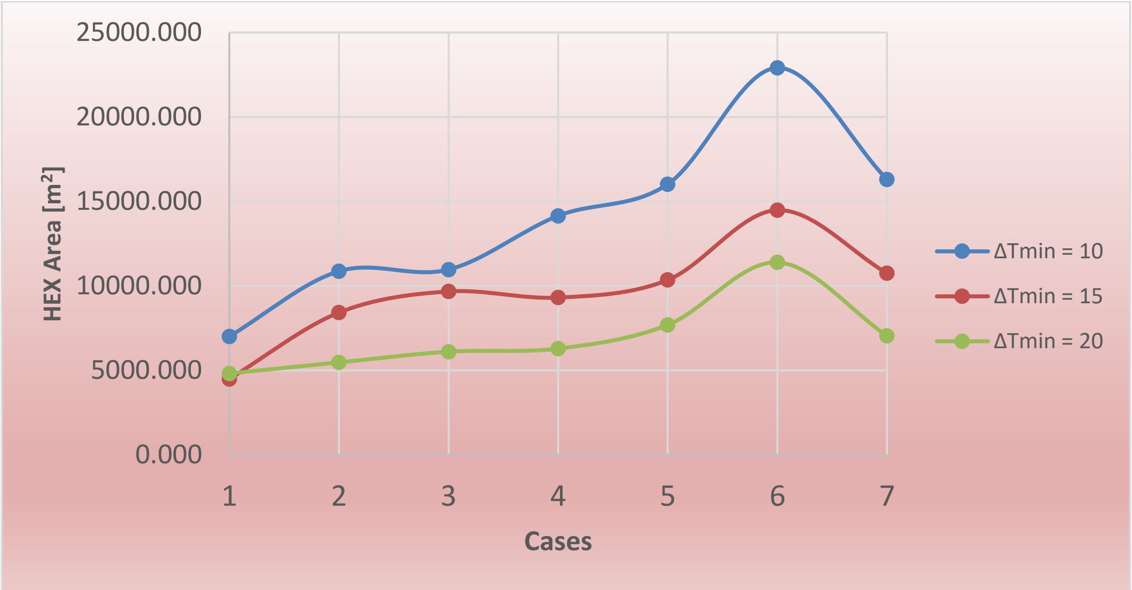

Figure 9.5: HEX area demand for ΔT_{min} of 10, 15, and 20 °C.

9.5.3 Steam consumption

Regarding the steam consumption in the CO_2 removal process, the higher temperature difference in the lean/rich heat exchanger will result in a higher steam demand in the reboiler unit as demonstrated in figure 9.6. So then in terms of steam consumption, the case of minimum approach temperature of 20, 15, and 10 are respectively the more costly in the process.

In terms of dependency on packing stages quantity in the absorber unit, the higher quantity of packing stages reduces the steam demand in the reboiler. Therefore, case-1 with 16 stages has registered the lowest target with the steam requirement value of 24154, 25756, and 38153 kW for the minimum approach temperature of 10, 15, and 20 °C respectively.

Ignoring the case-7, case-6 with 11 stages has hit the highest steam consumption in the reboiler in this trend and the values for three cases of ΔT_{min} of 20, 15, and 10 °C are 61571, 48280, and 42142 kW respectively.

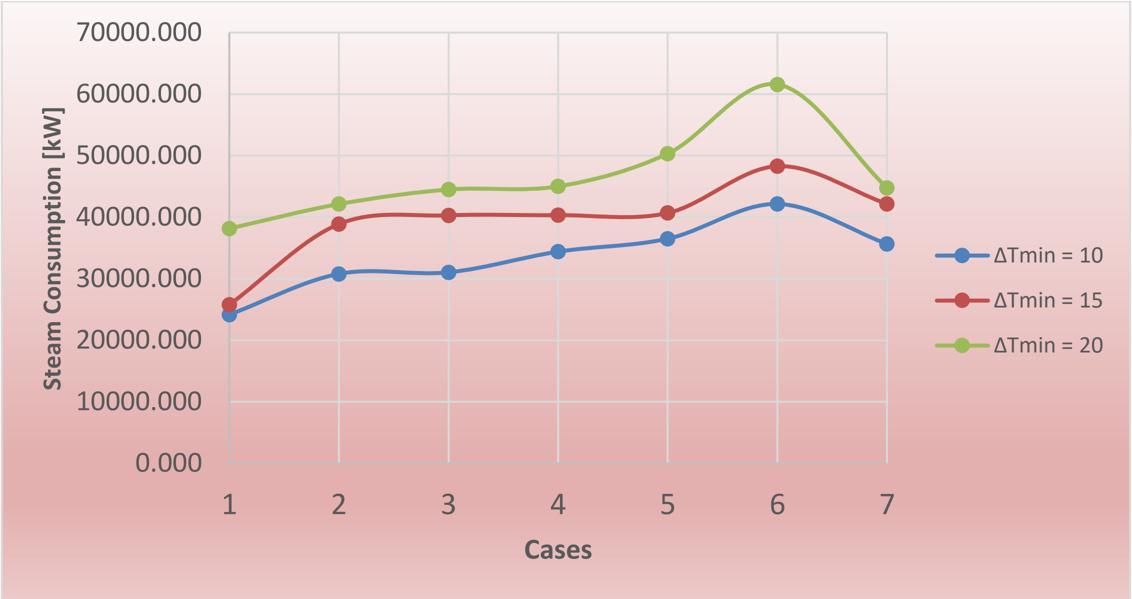

Figure 9.6: Steam consumption for ΔT_{min} of 10, 15, and 20 °C.

9.5.4 Electricity consumption

Electricity consumption in pumps and fan is another important and costly parameter in the OPEX that should be investigated. Figure 9.7 has demonstrated the electricity consumption for 3 cases of ΔT_{min} of 10, 15, and 20 °C. The results are almost the same and just some small differences are separated 3 cases from each other. In overall, electricity is more consumed in case of ΔT_{min} of 20 °C and lower consumption recorded in case of minimum approach temperature of 10 °C.

Higher electricity demand is calculated for higher packing stages in the absorber and lower consumption expected in less stages in the column. The lowest electricity consumed in pumps and fan is belonged to the case-1 with 16 stages having temperature difference of 10 and 15 °C, is almost the same amount of 962 kW, and ΔT_{min} of 20 °C is 987 kW. And the highest consumption is hit by case-6 with 11 stages, with the value of 1062 kW in case of temperature difference of 20 °C, and 1052 kW for both case of ΔT_{min} of 10 and 15 °C.

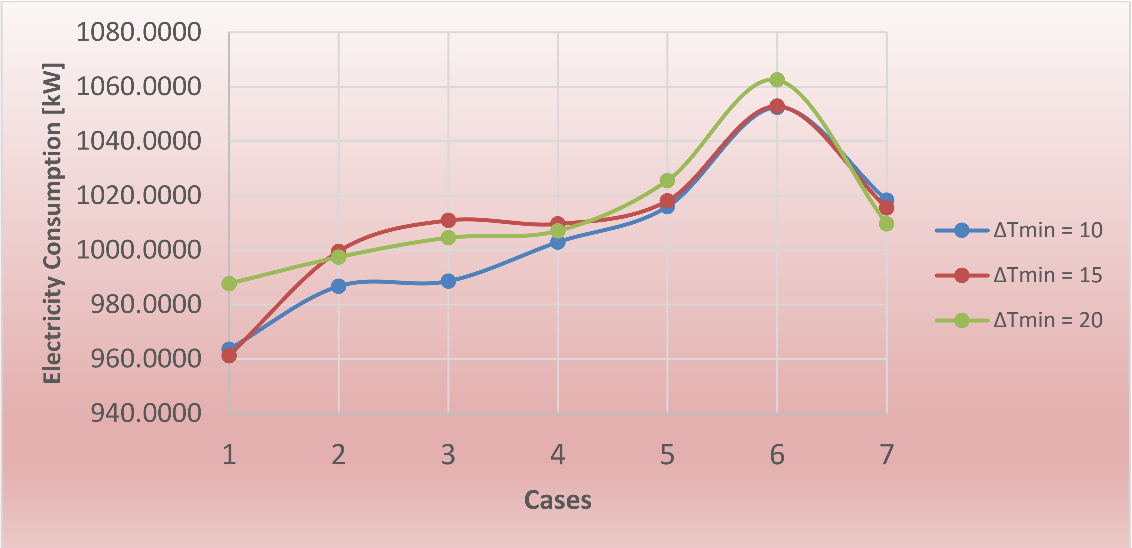

Figure 9.7: Electricity consumption for ΔT_{min} of 10, 15, and 20 °C.

9.6 Challenges for future work

9.6.1 Optimum case

Considering sensitivity analysis for all case studies, for an offshore application there is a challenge to select a case as the most optimum case study since there are a lot of parameters that should be taken into account. In addition to the entire CO_2 capture process plant itself, the space and weight requirements are quite costly and caused very large cost impact [22]. Especially for the current project that consist of a combined cycle including a WHRU, which is considerably huge in size and weight [2], in the upstream process. In this project, to decrease the size and weight of WHRU unit, 2 parallel WHRU unit are used instead of one. Having lower packing stages in the absorber column is an option to reduce the area requirement in the plant [4]. However, the penalty is a deduction in the plant efficiency.

Besides, as shown in the sensitivity analysis of case studies, technically it seems that the minimum approach temperature of 10 °C is the most optimum alternative, however, practically this selection is significantly expensive especially in an offshore application. So in this regard, higher temperature differences are may a better and more economical choice, but the penalty is higher energy demand in the reboiler.

On this basis, selecting 15 °C temperature difference in the heat exchanger and put the 87 % as the plant efficiency target, as Shirvan Shirdel et al [14]. Suggested as the most optimum solution, and 13-meter packing height in the absorber column to meet the efficiency target, could be a good suggestion as an optimum solution. For this setting, the energy and steam demand in the reboiler unit will be 5.5 MJ per one kilogram of CO_2 removed and 40.4 MW respectively and 1MW electricity in needed to run pumps and fan in this case.

9.6.2 Lack of similar work for an offshore application

The next challenge, as mentioned in section 2, in this study was that there were a few available open resources in this area, which contains both the combined cycle setting in the upstream and the CO_2 capture process for the offshore installation platform and it was a big challenge to carry out the project.

9.6.3 Lack of data for the real project

The current study was a real project assigned from Moreld Apply with the offshore application which was in the feasibility study stage. So, no realistic information was available in this stage and all basic required information for the base case and other case studies design and simulation needed to be assumed and suggesting the optimum solution is based on the sensitivity analysis reported in sections 7 and 8. So, in order to have a more reliable and optimized design in terms of size, weight, and cost, having real data from the project would be a great option.

10 Conclusion

An amine-based absorption carbon capture process is one of the best options that has been implemented in this master thesis in request of Moreld Apply for an offshore platform in North Sea in Norway.

The project was involved two process packages of upstream and downstream process. The upstream process was not a scope of this maser thesis and the upstream package was including a combined cycle system working with gas turbines, WHRU units, steam turbine, reboiler, condenser. and a heat exchanger. The scope of this project was to capture the carbon dioxide from the flue gas coming from upstream process through designing, simulating, and then after optimizing the downstream package of a standard absorption CO_2 capture process plant working with amine MEA as the solvent in the absorber column.

At the first stage, a base case model with setting of 90 % efficiency and a minimum approach temperature of 10 °C in the heat exchanger has been conducted and resulted in 16 number of packing stages in the absorber column, and an energy demand of 3.6 MJ in the reboiler to capture one kilogram CO_2 within the process. Then after, to suggest an optimum solution, a sensitivity analysis based on the absorber packing height and temperature difference in the lean/rich heat exchanger has been developed. Seven case studies for the packing height changing from 10 to 16, and three cases for the minimum approach temperature changing between 10, 15, and 20 °C, have been considered to investigate the process behavior in terms of energy and MEA solvent demand, steam and electricity consumption, and the ratio of CO_2 to amine content of absorber bottom outlet. In addition to the parameters and equipment including into the CO_2 capture process package, size and weight are the other important parameters that should be taken into account in selecting the optimum solution. Considering all parameters, a CO_2 capture plant working with minimum approach temperature of 15 °C requiring energy of 5.5 $\frac{MJ}{kg_{CO_2\ Captured}}$, and an absorption unit containing 13-meter packing height with 87 % efficiency, has been suggested as an optimum solution.

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Appendices

Appendix A Nils Henrik Eldrup 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 B Signed task description

University of South-Eastern Norway

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

FMH606 Master's Thesis

Title: Process design of CO_2 capture from the gas turbine at an oil platform

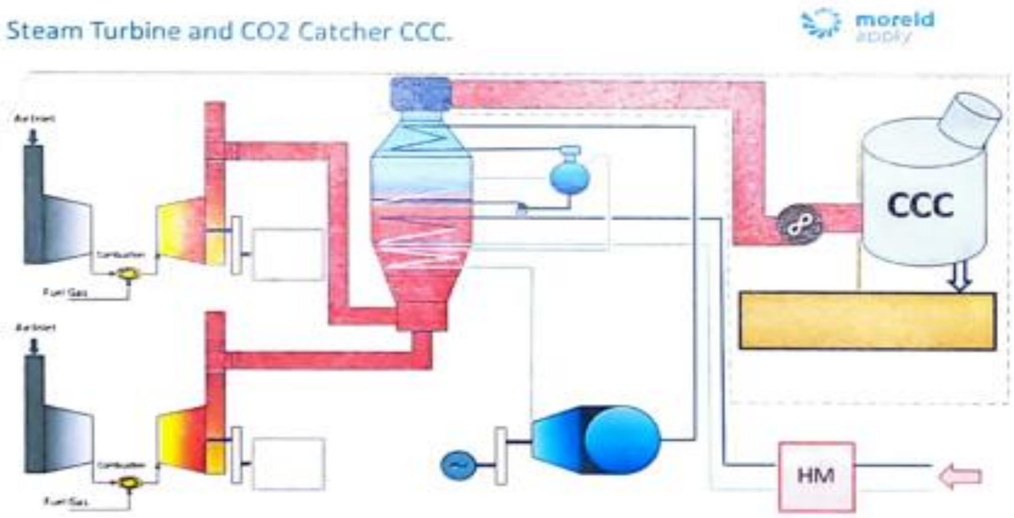
USN supervisors: Main supervisor: Rajan K. Thapa, co-supervisor: Lars Erik Øi

External partner: Moreld Apply

Task background:
Master projects from 2007 at the University of South-Eastern Norway and Telemark University College have evaluated CO_2 capture typically using Aspen HYSYS simulation. USN (HSN and TUC) has collaborated with different companies (SINTEF Tel-Tek, Equinor), working on CO_2 capture.

On one of the platforms in the North Sea, it is identified there will be a shortage of the heating medium of the platform due to demolishing of the old WHRU (Waste Heat Recovery Unit).

A new source for the heating medium will be required and one of the sustainability proposals is to install a new WHRU on the exhaust from a gas turbine.



Task description:
The general aim is to evaluate challenges in a CO_2 capture process from the outlet of a gas turbine.

1. Literature study on absorption-based CO_2 capture from a gas turbine on a platform.

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

2. Work out a process description for a process from the outlet of a gas turbine through a heat exchanger and CO2 capture. The process may be simulated in Aspen HYSYS.

3. Evaluate process challenges in collaboration with a Bachelor project and Moreld Apply.

4. Evaluation of advantages and drawbacks with the different process alternatives.
The detailed and more specific task description can be defined by close co-operation with Moreld Apply.

Student category: EET or PT

Is the task suitable for online students (not present at the campus)? Yes

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).

Signatures:

Supervisor (date and signature):

Student (write clearly in all capitalized letters): **FATEMEH FAZLI**

Student (date and signature): **01/02/2022** Fatemeh Fazli