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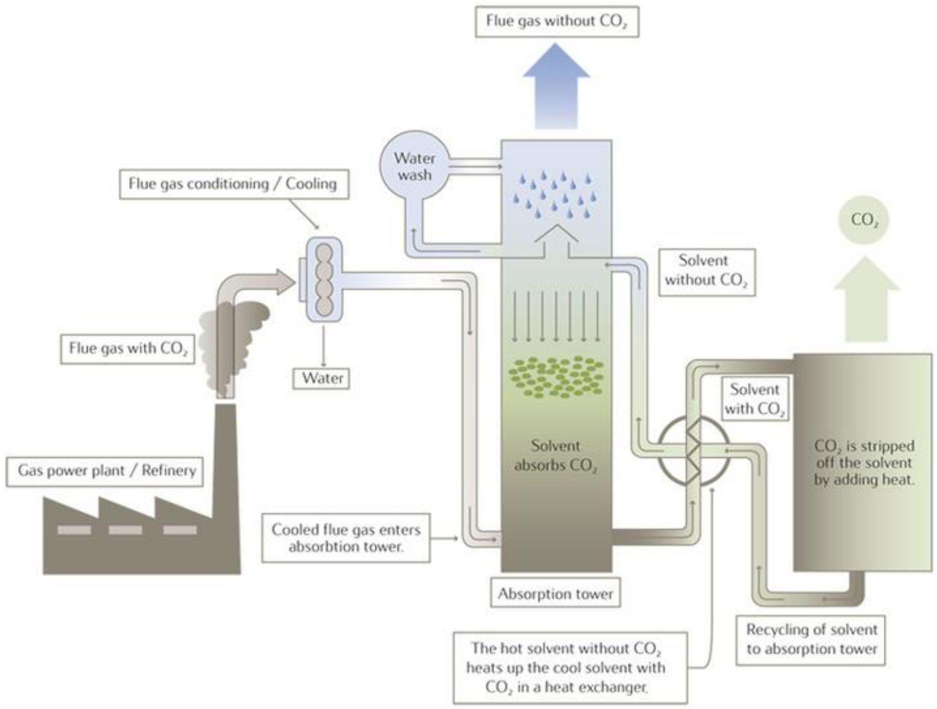
Process simulation and cost optimization of gas-based power plant integrated with amine-based CO_2 capture



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Summary:

Integration of gas-based power plants with CO_2 capture technologies is an efficient and cost-effective technique to remove carbon dioxide from their flue gas emissions and reduce greenhouse gas emissions to reduce global warming and climate change. Also, the removal of CO_2 by utilizing an amine-based solvent from the exhaust gas is an effective and well-proven process.

The base case scenario was modeled in the Aspen HYSYS by using the input data parameters from earlier works on a natural gas-fired power plant. In the design of the base case simulation in order to cost evaluate the turbine inlet temperatures, the power generation of the combined cycle, the CO_2 removal efficiency, the minimum temperature approach in the lean-rich heat exchanger and the flue gas inlet to absorber temperature were all set at 1500 °C, 400 MW, 85%, 10 °C, and 40 °C, respectively. And Aspen In-Plant Cost Estimator tool, Enhanced Detailed Factor (EDF) technique, and net present value (NPV) method were used to estimate the overall cost of the base case model. The CAPEX, OPEX, and revenue from selling power were considered in the cost evaluation. The cost calculation outcome indicates a net present value of €1570 million over a 25-year plant lifetime and a payback period of 7 years after project implementation.

Sensitivity analysis was applied to perform cost optimization. The Power-Law method was used to estimate the new cost when the size of equipment is altered. In sensitivity analysis, ambient temperature, inlet air/gas pressure into the power plant, flue gas outlet temperature from the power plant (off-gas), the lean-rich heat exchanger's minimum temperature approach, number of stages in the absorption column, inlet flue gas temperature to the absorber column were all changed to reach the project's highest net present value.

The cost-optimized process parameters, according to the results of the sensitivity analysis, will be achieved at lower ambient temperatures (in this case, 10 °C), inlet air/gas pressure of 25 bar, off-gas temperature of 50 °C, minimum temperature approaching 13 °C in lean-rich heat exchangers, 10 stages of absorber columns, and inlet flue gas temperatures in the range of 30 to 35.

The results of the cost optimization process parameter analysis obtained from the sensitivity study were used to modify the base case. The calculated NPV results provide an average 15 % increase in profit compared to the initial base case.

In this study, the main objective is to use the Aspen HYSYS package to estimate and optimize the cost of a gas-based power plant integrated with an MEA-based CO_2 capture plant. A cost optimization calculation of the model's process parameters, which include a CO_2 removal plant and a gas-based power plant simultaneously has not been found in earlier work and in this aspect, this work is innovative.

Preface

This thesis was prepared as part of the master's degree in Energy and Environmental Technology at the University of South-Eastern Norway in the Autumn of 2022.

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

I would like to express my special thanks to my supervisor, Lars Erik Øi, Professor at the University of South-Eastern Norway, for his valuable and helpful guidance and assistance throughout my thesis composition. Undoubtedly, this could not have been accomplished without his motivation and support.

I want to also express my gratitude to Solomon Aforkoghene Aromada and Shirvan Shirdel, for answering my questions and assisting me in this thesis.

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

Porsgrunn, January 2023

Esmaeil Aboukazempour Amiri

Contents

Nomenclature

Abbreviation	Description
CAPEX	Capital expenditure
OPEX	Operational expenditures
CCUS	capture utilization and storage
CO_2	Carbon Dioxide
e	exponential size factor
EIC_{CS}	total installed cost for each equipment
$F_{T,CS}$	total installation factor
f_x	sub-factors for the component x
f_{mat}	Material Factor for Stainless Steel welded/rotating
FG	Flue gas
H_2O	Water
IEA	International Energy Agency
MEA	Monoethanolamine
n	Operating lifetime
Ns	Number of stages
N_2	Nitrogen
O_2	Oxygen
r	discount rate
€	Euro

Abbreviation	**Description**
PCC	Post-combustion carbon capture
EGR	Exhaust gas recirculation
CCS	Carbon capture and storage
LHV	Low heating value
T&S	CO_2 transport and storage
LCOE	Levelised cost of electricity
DOC	Degrees of capture
SPECCA	Specific primary energy consumption per unit of CO_2 avoided
NGCC	Natural Gas Combined Cycle
PFD	Process flow diagram
NPV	Net present value
PBP	Payback period

1 Introduction

Global warming is now one of the most significant issues of the world community and one of the main causes of climate change. Greenhouse gases which are made by human activity are part of this problem and countries put some limitations on the emission of these pollutants. Among greenhouse gases, carbon dioxide is the main anthropogenic greenhouse gas that contributes to global warming with about 80% of its share [1].

Using fossil fuels such as Coal, Natural gas for generating electricity is one of the main sources of producing CO_2 [2]. The global energy-related CO_2 emission based on the sector is depicted in figure 1.1 [3].

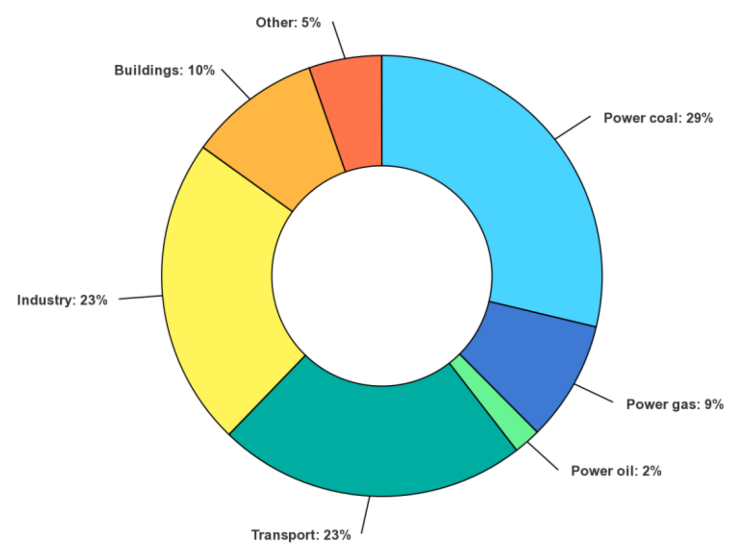

Figure 1.1: Global energy-related CO_2 emissions by sector [3]

CO_2 capture, utilization, and storage (CCUS) are considered as a possible solution to stop carbon dioxide emissions to the atmosphere. In the CCUS method, CO_2 is removed from flue gas and from the atmosphere, which is then recycled for use, and a safe and permanent storage method is determined. CCUS relates to the collection of CO_{2s} from significant onsite sources, such as industrial or power plants that burn primarily fossil fuels or biomass as fuel. The acquired CO_2 is compressed and either injected into deep geological structures (such as depleted oil and gas reservoirs or saline layers) that trap the CO_2 for long-term storage, or it is delivered by pipeline, ship, rail, or truck to be used in a variety of uses [4], [5].

1.1 Background

Based on the IEA report fossil fuels are the main source of CO_2 emission in the world, while also they are an important part of the world's energy resources and their role may be increased in the future [5].

To prevent and reduce the impact of carbon dioxide emissions from fossil fuel power plants, carbon capture, and storage (CCUS) is one of the most efficient solutions, along with other approaches such as increasing the efficiency of power plants and using green energy technology [6].

Between fossil fuels, natural gas is one of the most usable and useful sources of energy production. Using natural gas at combined power plants is one of the modern technologies and gives the highest efficiency compared to other fossil fuels. for example, Coal power plants emit twice as much carbon dioxide as natural gas power plants [7].

Despite the fact that the share of production and emission of carbon dioxide by natural gas power plants is less than that of other fossil fuel power plants, they are still considered one of the main sources of production and emission of carbon dioxide [6].

There are three main CO_2 capture methods that are used in combustion operation, pre-combustion capture, oxygen combustion, and post-combustion capture as shown in figure 1.2. Among these three methods, post-combustions are the more common and useful method for fossil fuel power plants and other industries as well [8].

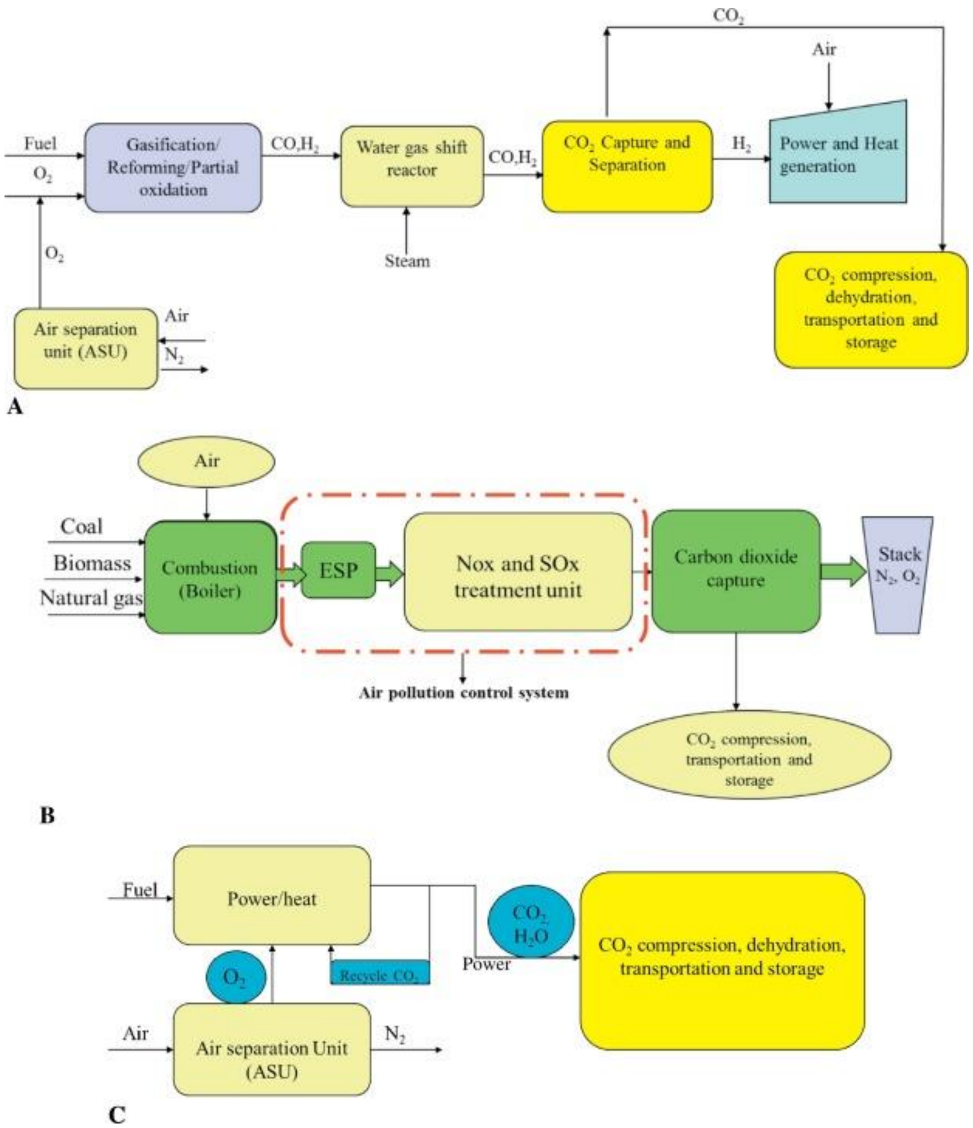

Figure 1.2: (A) Pre-combustion, (B) post-combustion, (C) oxy-fuel CO_2 capture technology [10]

The important point about post-combustion CO_2 capture is that upwards of 70% of the overall cost of the carbon capture and storage (CCS) process is spent on CO_2 capture, making it the most expensive component [9].

For post-combustion CO_2 capture, there are several common methods such as chemical absorption by aqueous amine solution [11], adsorption [12], cryogenic separation [11], membrane separation [11], and also microalgal systems which are shown in figure 1.3 below in more details.

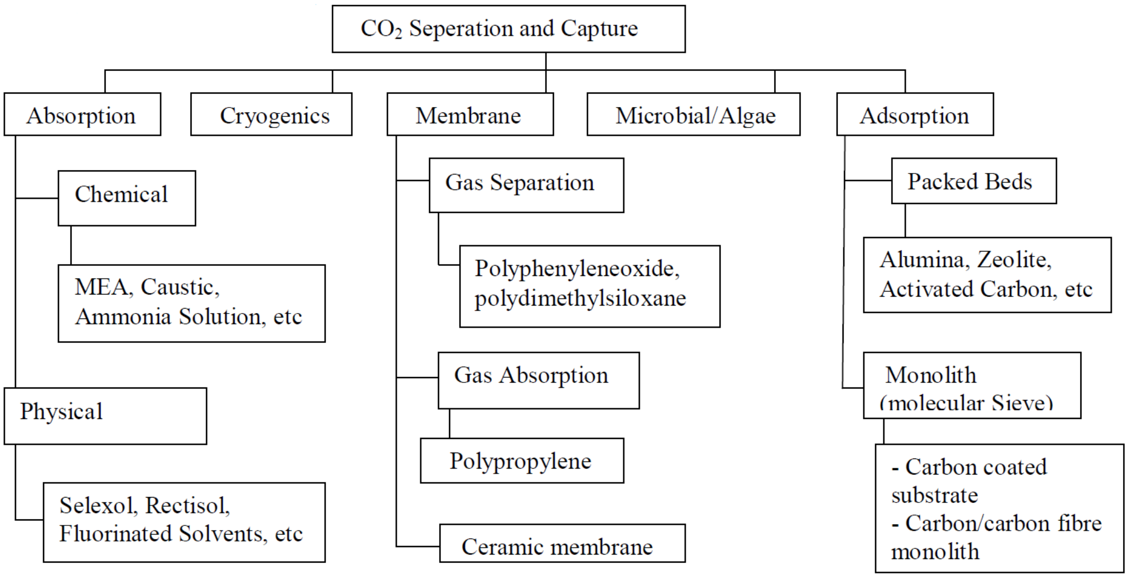

Figure 1.3: Post-combustion CO_2 capture technologies [13]

Among the mentioned methods, the chemical absorption/stripping process followed by desorption based on the use of MEA is the most suitable and applicable technology and it can be integrated with the NGCC power plant to provide efficient operation [14].

1.2 Literature review

The objective of this chapter is to discuss the most relevant literature concerning the design, simulation, dimensioning, and cost optimization of an integrated natural gas combined power plant with CO_2-capturing units. Many studies have been conducted on stand-alone CO_2 capture plants, but very few studies have been conducted on integrated capture and power plants. However, the most pertinent literature in this field is listed in table 1.1 according to the year it was released.

Table 1.1: A literature overview

Row	Literature Title	Year	Location	Reference
1	Aspen HYSYS Simulation of CO_2 Removal by Amine Absorption from a Gas Based Power Plant	2007	Norway	[14]

2	Combining bioenergy and CO_2 capture from gas fired power plant	2011	Norway	[16]
3	Removal of CO_2 from exhaust gas	2012	Norway	[15]
4	Natural Gas Combined Cycle Power Plant Integrated to Capture Plant.	2012	Norway	[17]
5	Natural gas combined cycle power plants with CO_2 capture – Opportunities to reduce cost	2012	Norway	[18]
6	Heat integration of natural gas combined cycle power plant integrated with post-combustion CO_2 capture and compression	2015	China	[19]
7	Optimal operation of MEA-based post combustion carbon capture for natural gas combined cycle power plants under different market conditions	2016	United Kingdom	[20]
8	Techno-economic process design of a commercial-scale amine-based CO_2 capture system for natural gas combined cycle power plant with exhaust gas recirculation.	2016	United Kingdom	[21]
9	A techno-economic analysis of post-combustion CO_2 capture and compression applied to a combined cycle gas turbine: Part I. A parametric study of the key technical performance indicators.	2016	United Kingdom	[22]
10	A techno-economic analysis of post-combustion CO_2 capture and compression applied to a combined cycle gas turbine: Part II. Identifying the cost-optimal control and design variables	2016	United Kingdom	[23]
11	Thermodynamic analysis and techno-economic evaluation of an integrated natural gas combined cycle (NGCC) power plant with post-combustion CO_2 capture.	2017	China	[24]
12	Selection and design of post-combustion CO_2 capture process for 600 MW natural gas fueled thermal power plant based on operability.	2017	Norway	[25]
13	A new integration system for natural gas combined cycle power plants with CO_2 capture and heat supply.	2018	China	[26]
14	Optimization of Post Combustion CO_2 Capture from a Combined-Cycle Gas Turbine Power Plant via Taguchi Design of Experiment	2019	United Kingdom	[27]

15	Preliminary performance and cost evaluation of four alternative technologies for post-combustion CO_2 capture in natural gas-fired power plants.	2020	Italy	[28]
16	A simulation study of the effect of post combustion amine-based carbon-capturing integrated with solar thermal collectors for combined cycle gas power plant.	2021	Egypt	[29]

Lars Erik Øi [14] simulated a basic natural gas combined power plant and a CO_2 removal process based on MEA with Aspen HYSYS using Peng Robinson and Amines property package models. Based on this model, gas power plane efficiency is 58% before and 50% after integrating the CO_2 removal process. The amine circulation rate, the height of the absorption column, the temperature of the absorption, and the temperature of steam were all set to achieve 85% CO_2 removal. In the case of 85% CO_2 removal, heat consumption for each kilogram of removed CO_2 was calculated to be 3.7 MJ/kg [14].

Lars Erik Øi's Ph.D. thesis [15] dealt with the calculation of CO_2 removal from the atmospheric exhaust gas from a natural gas combined power plant. This study aimed to calculate the optimum parameter values based on cost. In addition, he considered the split stream process with a heat consumption of 3 GJ/ton CO_2 removed compared to a standard process with a heat consumption of 4 GJ/ton CO_2. According to the optimum calculation results, the gas inlet temperature value is between 33 and 35°C, the minimum temperature approach value in the lean/rich amine heat exchanger is between 12 and 19°C, and the rich amine loading value is 0.47 mol CO_2 /mol MEA. Optimal automation calculations were also evaluated [15].

Eldrup et al. [16] studied a system that integrated natural gas power with CO_2 capture, powered by biomass-based external energy plants. A comparison is made between the concept and other options based on estimated capital and operational costs. The operating cost estimates are constrained by the requirement to purchase a CO_2 share for each tonne of non-bio-based CO_2 released to the atmosphere and to award a CO_2 quota value for each tonne of bio-based CO_2 recovered. The approach is economically viable when biomass costs are low and CO2 quota costs are high. In a sensitivity analysis, the prices for natural gas, CO_2 quotas, and biomass are altered. The suggested strategy benefits from scenarios in which CO_2 quota prices are raised [16].

A natural gas combined cycle (NGCC) power plant with a capacity of about 430 MW was evaluated by Karimi et al. [17] in conjunction with a solvent absorber/stripping capture plant. A 90% CO_2 removal ratio is achieved, and the captured CO_2 is compressed to 75 bars, liquefied, and then pumped up to 110 bars. CO_2 capture results in an energy penalty of 398.4 kWhel/ton, which reduces the plant's net efficiency by 7.5%. Steam extraction and electricity consumption contribute 4.6% and 2.9%, respectively, to this energy penalty [17].

The thermodynamics and economic analysis of a 440 MWe CO_2 removal plant integrated into a natural gas combined cycle power plant was done by Sipöcza and Tobiesen [18] using an aqueous solution of monoethanolamine (MEA). The absorber inter-cooling and lean vapor recompression are included in the flow sheet that represents the CO_2 capture plant and is optimized with regard to process parameters. Additionally, the gas turbine makes use of exhaust gas recirculation (EGR) at a level of 40% in order to further minimize the costs of CO_2 capture, increasing the CO_2 content in the exhaust gas to approximately twice that of conventionally functioning gas turbines. The combination of a lower particular reboiler duty with EGR is shown to have a considerable positive impact on both capital and operating expenses. Furthermore, it is indicated that accurate cost estimates are greatly influenced by fuel and currency exchange rates [18].

The integration of a 453 MWe natural gas combined cycle (NGCC) power plant with a CO_2 compression train and MEA-based post-combustion carbon capture (PCC) technology was the focus of a study by Luo et al. [19]. Aspen Plus was used to develop and validate the steady-state models for the NGCC power plant, the PCC process, and the compression train, using documented and experimental data. The analysis found a significant size decrease of the absorber and the stripper was accomplished with exhaust gas recirculation (EGR). When the NGCC power plant was combined with the PCC process and compression, the case study demonstrates that net efficiency based on low heating value (LHV) declines from 58.74% to 49.76%. The overall efficiency is raised to 49.93% with the use of EGR and to 50.25% and 50.47%, respectively, with the aid of two alternatives for integrating compression heat [19].

The study by Luo and Wang [20] evaluated how a natural gas combined cycle (NGCC) power plant integrated with post-combustion carbon capture (PCC) technology should operate in a variety of market conditions. In cost optimization studies, the objective function is defined as the levelized cost of electricity (LCOE). For the integrated system's basic scenario, which includes CO_2 transit and storage, an economic evaluation was done. The analysis demonstrates that for the cost of carbon capture from the NGCC power plant to be justified, the carbon price must be over €100/ton CO_2 and must be above €120/ton CO_2 for the carbon capture level to be driven at 90% [20].

Recirculating exhaust gas is a technique for rising the CO_2 concentration in the lean flue gas for systems that burn natural gas to generate electricity. According to the Ali et al. [21] study, four specific amine-based CO_2 capture devices were developed and scaled up, with a 90% capture rate and a 30% aqueous solution of amine. The design outcomes for a natural gas-fired combined cycle power plant with a net output power of 650 MWe are documented, along with the EGR percentages at 20%, 35%, and 50%. 0.96 is an ideal liquid-to-gas ratio of is predicted for an amine-based CO_2 capture plant with a gas-based combustion engine without EGR. The proper liquid-to-gas ratios are 1.22, 1.46, and 1.90 when EGR is applied at 20%, 35%, and 50%, respectively. These findings imply that lower energy demand and financial expenses for the amine-based CO_2 capture facility will come from the use of a natural gas-fired power generation with recirculation of exhaust gas [21].

Using monoethanolamine as the basis, Alhajaja et al [22] designed and analyzed a CO_2 capture plant and compression train model. As a result of the validation of this model, key operating parameters were assessed on the basis of selected non-monetized key economic and environmental performance indicators to determine how key operating parameters affect the performance of the CO_2 capture and compression process for a combined cycle gas turbine (CCGT). The findings show increased compression power and a sharp rise in the amount of cooling water needed by coolers and washing water systems. This paper explains the difficult trade-off between reducing environmental consequences and capital and operational expenditure measures [22].

Alhajaj et al. [23] noted in the second part of their study that the effect of the bypass option for CO_2 capture is the most affordable alternative when the total degree of capture [DOC] is less than 60%. Carbon costs were shown to significantly affect the cost-optimal DOC, changing it from 70%–80% to 85%–90% for carbon prices of \$4/tCO_2 to \$23/tCO_2, respectively [23].

The work by Xu et al. [24] proposes a novel integrated system with electricity generation and CO_2 capture with the goal of reducing the de-carbonization penalty. Four strategies are used in the new system: recycling some of the flue gases from the gas turbine to raise the Carbon dioxide levels in the combustion gases; incorporating some of the condensate water from the boiler with the exported steam to take advantage of the extracted steam's superheat degree; compressing the Carbon dioxide flow at the upper side of the stripper to recover the latent heat for sorbent recovery; and presenting a transcritical CO_2 cycle to make use of the sensible heat in the exhaust gas. Because of this, the new integrated system's power production is 26.15 MW more than that of the NGCC power plant without integration for decarbonization. It is anticipated that the CO_2 capture efficiency penalty would drop by 2.63%-points. However, the redesigned system's investment growth of 60.17 M\$ is just 4.66% higher compared to the decarbonization power generation plant without integration [24].

For a natural gas power plant, Dutta et al. [25] looked at the option and design of a post-combustion CO_2 capture (PCC) facility. It was decided to use two improved PCC system designs, each of them with a slight penalty of efficiency. When developing PCC plants, design restrictions based on operability and the construction of absorbers were taken into consideration. This was utilized to calculate the size of the plant's equipment. Two absorber designs were studied using two variables: at a full load of flue gas flow rate and the time average of the predicted load fluctuation of a flexible operating power plant. The cost of purchasing absorbers dropped by 4% as a result of the time-average load-based design. In full-load operation, the absorber reduced reboiler duty in order to keep a comparable capture rate to that of the other absorber under part-load performance [25].

In the Hu et al. [26] research, an integrated model including power production, CO_2 capture, and heat supply is suggested. This system uses three techniques to recycle the waste heat emitted during the CO_2 capture process: extraction steam recirculation, a CO_2 Rankine cycle, and a radiant floor heating system. The radiant floor heating subsystem, among other techniques, may effectively recycle the relatively low-temperature waste energy in the absorbent cooler. By using thermodynamic investigation, it is found that the new integrated system's power production is 19.48 MW more than that of the decarbonization Natural Gas Combined Cycle (NGCC) power plant without heat system integration. The radiant floor heat subsystem can reclaim 247.59 MW of heat, resulting in an increased total energy efficiency of 73.6%. When compared to a decarbonizing NGCC power plant without system integration, the integration only needs 2.6% more capital investment and earns an additional 3.40 \$/MWh from the simultaneous heat delivery, cutting the cost of CO_2 saved by 22.3% [26].

Soltani et al. [27] simulated 90% CO_2 capture of a 600 MW natural gas combined-cycle gas turbine power plant using an equilibrium-based technique in Aspen Plus. Signal to noise ratios and analysis of variance were used to assess the impacts of changing the inlet exhaust gas temperature, absorber column operating pressure, the volume of exhaust gas recycling, and amine level of concentration. Exhaust gas temperature of 50°C, absorber pressure of 1 bar, flue gas recirculation of 20%, and amine concentration of 35 wt.% were the ideal values that reduced the amount of specific energy needed, with the priority being amine concentration above absorber column pressure, followed by exhaust gas recirculation, then flue gas temperature. According to the analysis, the total energy needed for the capture unit is 5.05 GJ/ton CO_2, and the energy required for the boiler is 3.94 GJ/ton CO_2 [27].

The goal of Gatti's [28] study is to evaluate 4 different techniques for post-combustion CO_2 capture from natural gas-fired power plants in terms of their operational and economical possibilities. Which include molten carbonate fuel cells (MCFCs), CO_2 permeable membranes, pressured CO_2 absorption combined with a multi-shaft gas turbine and the steam cycle of heat recovery, and supersonic stream CO_2 anti-sublimation and inertial detachment. They considered modern natural gas combined cycle (NGCC) without CO_2 capture as the reference scenario, while the same NGCC developed with CO_2 capture (using chemical absorption with aqueous monoethanolamine solvent) is employed as the base case. The comparison reveals that a combined cycle using MCFCs seems to be an interesting technology, both economically and in terms of energy penalty, with a CO_2 averted cost of 49 \$/tCO_2 saved and a specific primary energy consumption per unit of CO_2 avoided (SPECCA) of 0.31 MJLHV/kgCO_2 saved. In terms of CO_2 capture technology, PZ scrubbing comes in second (SPECCA = 2.73 MJLHV/kgCO_2 avoided and CO_2 cost saved= 68 \$/tCO_2), then monoethanolamine (MEA) base case (SPECCA = 3.34 MJLHV/kgCO2 avoided and CO2 cost saved= 75 \$/tCO2), and supersonic flow driven CO_2 anti-sublimation, inertial separation, and CO_2 permeable membranes [28].

An economic analysis was conducted by Ayyad et al. [29] of the 495 MW West Damietta power plant in Egypt to reduce re-boiler duty and lost power due to re-boiler duty. The first strategy involves recycling some flue gas back into the combustor at various ratio percentages (0%-35%), while the other strategy involves using parabolic-trough solar collectors to manage the load on the boiler in place of low-pressure steam taken from the power plant. The results showed that raising carbon content lead to a notable drop in the re-boiler duty of up to 20%. The levelized cost of energy (LCOE) was also impacted by the rise in carbon emissions; it was reduced by 1.39% and the cost of avoiding carbon emissions by 6% using a 35% recirculation ratio. Additionally, it was discovered that the integration of a solar system and a thermal energy storage system greatly increased the facility's capacity to produce at its highest [29].

1.3 Scope of the study

This study focuses on the integration of natural gas combined cycle plant (NGCC) with an amine absorption CO_2 capture method using the input parameters of past studies on a gas-fired power plant system [14]. Firstly, based on the supplied data, two simulations for the natural gas power plant and CO_2 capture plant were designed separately in Aspen HYSYS, and in the continue, these two-simulation combined and included in a flow sheet, and then based on a new flowsheet dimensioning and cost estimation were executed. The Aspen In-Plant Cost estimator is used with the Enhanced Detailed Factor (EDF) to evaluate and calculate the total cost for the basic scenario. Cost optimization is carried out by using sensitivity analysis to optimize and reduce costs. To observe the dependence of operation parameters on the total price, a sensitivity analysis was performed. The Power-Law approach is employed to update the cost of equipment and NPV value is used for cost optimization evaluation. As part of this investigation, the inlet pressure and temperature into the combustion chamber, the output exhaust gas temperature (Off gas) from natural gas power plant, the lean/rich amine heat exchanger’s minimum temperature approach, the height of absorber packing, and the temperature of the entering flue gas into the absorber column were all modified in present study.

This study's primary objective is to employ the Aspen HYSYS to simulate, compute and optimize the cost of the entire process with the aim to reach 85% CO_2 removal and generating of 400 Mw electricity simultaneously, and also automation of the output exhaust gas temperature (Off gas) from the natural gas power plant, the lean/rich amine heat exchanger's minimum temperature approach, and the temperature of the entering flue gas into the absorber column as a sensitivity analysis for optimal cost calculation are other goals of this work.

2 Process overview and base case simulation

This chapter gives a summary of the integrated natural gas combined cycle with the CO_2 capture plant, as well as the major components and their roles. In the following section, the simulation of a base case in aspen HYSYS for cost optimization is presented.

2.1 Process Overview

To remove the CO_2 from natural combined gas power plant the post-combustion CO_2 capture system is the best approach among other options.

2.1.1 Natural gas combined power plant

Gas turbine–steam turbine combined-cycle power plants produce power by utilizing the exhaust heat from a gas turbine to run a boiler, which produces steam that is fed into a steam turbine for electricity production. Compared to simple-cycle plants, these plants produce 50% more electricity and have efficiencies as high as 50%–60% with low emissions [30].

The gas turbine is driven by the fuel's combustion byproducts (Brayton cycle), whereas the steam turbine (Rankine cycle) is driven by the steam produced by HRSG from the latent heat of the exhaust gases exiting the gas turbine [31].

The main components of a combined cycle power plant are:

- A Gas Turbine (GT)
- Heat recovery steam generators
- A Steam Turbine (ST)
- Condensers

The main components and the schematic of the natural gas combined power plant are illustrated in figure 2.1.

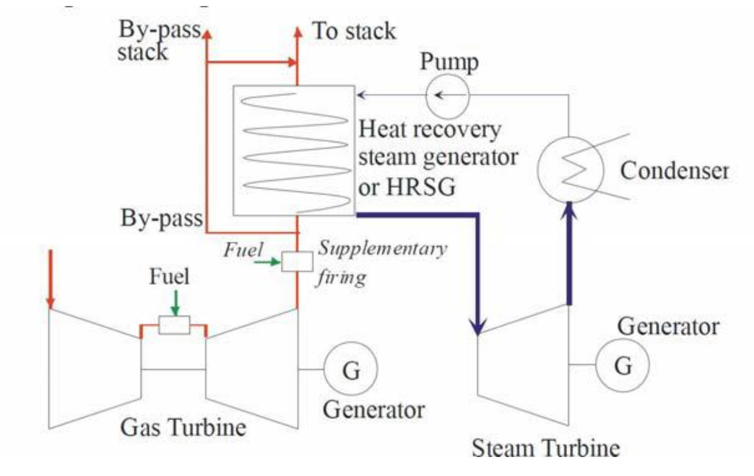

Figure 2.1: Schematic of combined-cycle power generation [32]

2.1.1.1 Gas Turbine

Gas turbines convert natural gas or fluid into mechanical energy in combined cycle power plants. Gas turbines are composed of three parts: compressor, combustor, and generator. As the compressed air mixes with the fuel, constant pressure is applied to the combustion process. In order to produce work, the hot gas is passed through the turbine [31].

Gas turbines work on the Brayton cycle, which consists of two isobaric and two isentropic processes in their ideal form. Both the gas turbine combustor system and the heat rejection side of the gas turbine as well as the HRSG heat gas side are isobaric processes. In a gas turbine, the compressor (Compressor) and the expander (Turbine Expander) represent the two isentropic processes [31].

A typical NGCC can produce 350–500MW of power with a thermal efficiency of 57–60%. The temperature at the combustion chamber outlet can reach up to 1500°C. Commonly known as the turbine inlet temperature (TIT) [32].

The gas and steam turbines' respective power outputs as well as the work required to operate auxiliary devices like pumps and compressors make up the power balance.

$$W_{NGCC} = W_{G} + W_{ST} - W_{C} - W_{P} \tag{2.1}$$

$$\eta_{th\,eff\,NGCC} = \frac{W_{NGCC}}{\dot{m}_{fuel}(LHV_{fuel})} \tag{2.2}$$

2.1.1.2 Heat recovery steam generators (HRSGs)

In recent years, heat recovery steam generators (HRSGs) have become increasingly popular. As a result, they are typically combined with a gas turbine, and the steam generates additional electricity. Combined-cycle units are highly efficient at generating electricity as well as being able to operate at partial loads [33].

Gas Turbine exhaust gases are received by the HRSG. Steam/water coils cool the exhaust gas by transferring heat to steam/water in the counterflow direction. About 110°C is the temperature of the flue gas at the stack. For fuel gases that are very clean and sulfur-free, temperatures as low as 93°C can be used [31].

Similar to a heat exchanger, the HRSG transports steam or water via tubes while flue gas is supplied on the shell side. Due to the presence of steam drums, where the produced steam is separated from boiling water before feeding the superheaters, it also possesses the features of a boiler [31].

Based on plant size, the HRSG could have one, two, or three levels of pressure. The pressure levels employed are HP, IP, and LP for plants with a capacity of 200–400 MW. Plants under 60 MW typically have two pressure levels (HP and LP), whereas smaller plants only have one. On occasion, the LP section will only provide the steam required for deaeration when three pressure levels are present [33].

2.1.1.3 Steam Turbine

In a combined cycle power plant, steam turbines use the energy in the steam to produce work that turns the turbine's shaft. The enthalpy drop throughout the device determines how much energy the steam turbine obtains from the steam. Based on its temperature and pressure, steam has a different enthalpy. A Mollier diagram may be used to calculate the amount of energy available if the input and output temperature and pressure are given [33].

There are three control modes used by the steam turbine: [31]

- Fixed pressure mode

The steam turbine will run in a fixed pressure mode when less than 50% of the load is present, which equates to around 50% of the live steam pressure. The principal control is in charge of maintaining a steady steam generator pressure in this mode of operation. A steam generator's pressure is regulated by its bypass valves in the situation that the steam turbine is not utilizing all of the output steam.

- Sliding pressure mode

The primary control valve is completely open after the load has reached 50%. As gas turbine loads increase, the steam turbine will run in sliding pressure mode. In this situation, the live steam pressure changes in direct proportion to the steam flow.

- Load control

The grid controls the frequency of the generator after it is synced with it. The steam flow is adjusted by the turbine controller to preserve the baseload.

2.1.1.4 Condensers

Condensing steam turbines are the most common form of large steam turbines. Condensers come in both air-cooled and water-cooled varieties. Condensers that are cooled by water are more widespread and more efficient. For the LP steam turbine to operate well, the condenser's cooling capacity is crucial. If the efficiency of the cooling is decreased, the back pressure of the LP steam turbine will grow and the steam turbine's power output will decrease [31].

2.1.2 CO_2 capture plant

A CO_2 capture method comprised of two steps: (i) separating CO_2 from a gas mixture using a selective reaction, and (ii) regenerating the material that was used to separate the CO_2 using a reverse reaction. Steps (i) and (ii) can be repeated in order to reuse the material for CO_2 capture. Reversibility is one important practical condition for CO_2 removal. The majority of CO_2 capture materials, including chemical solvents, porous sorbents, gels, and membranes, include amines as one of their primary chemical components. This is mostly due to the fact that moderate contact enables a successful segregation of CO_2 and amine through a reversible reaction. Due to the wide variability in amine chemical composition, it is possible to further modify the CO_2 capture material within the area of moderate reaction by changing the amine structure and/or mixing amines [34].

Figure 2.2 depicts the process' essential elements, the steps involved in removing CO_2, and the overall flowchart for the CO_2 removal power plant. The procedure may be categorized into three steps.

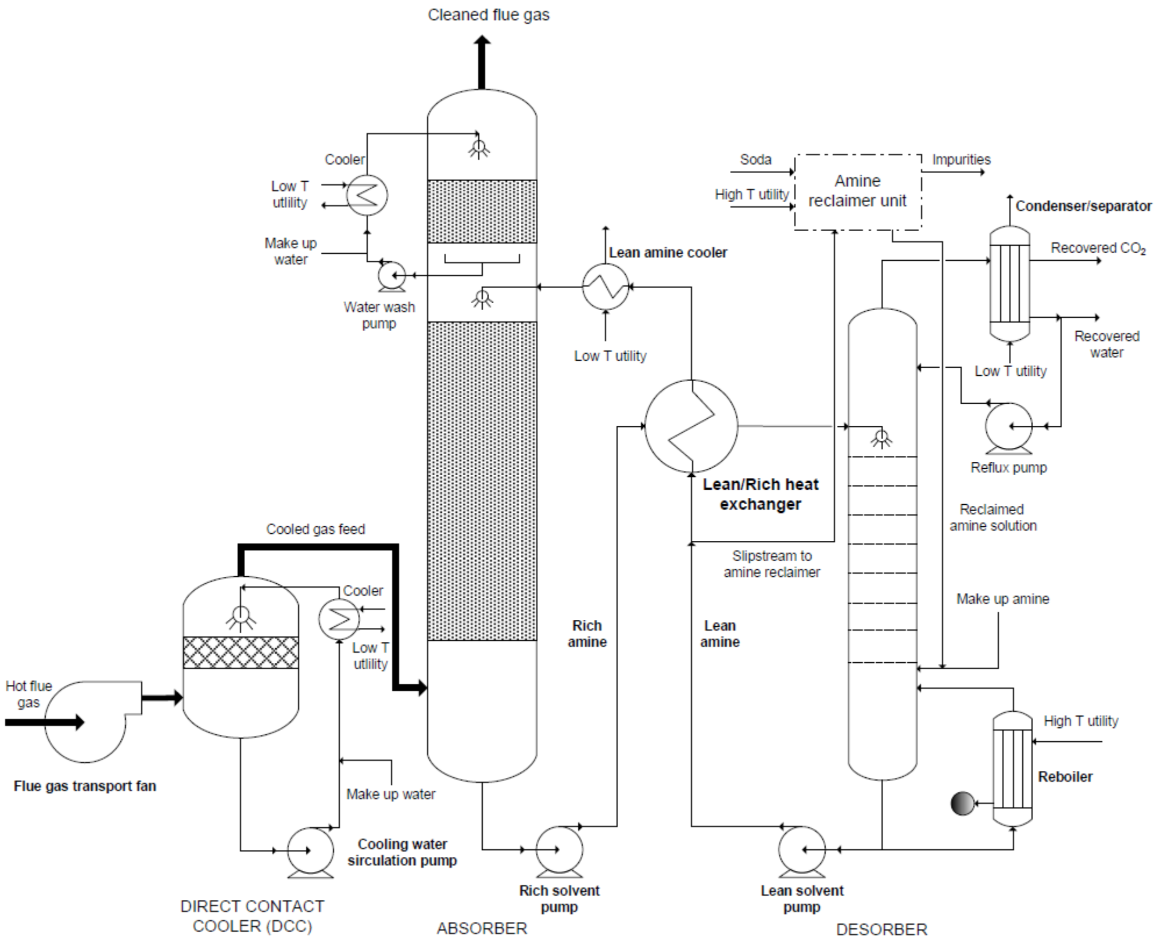

Figure 2.2: The typical amine-based CO_2 capture plant's flow diagram (PFD) [42]

- Pre-capture process and simulation: This phase of the process occurs before the absorber's primary CO_2 absorption. The flue gas fan, which transports flue gas to the absorber through a direct contact cooler (DCC) unit where the flue gas temperature is dropped to 40 C, is the only piece of equipment taken into consideration in this first stage of the process [35].

- Capture process: The related system consisted of a basic absorption column, a desorption column with a condenser and a reboiler, a main lean/rich heat exchanger (LRHX), two pumps (a lean pump and a rich pump), and a lean amine cooler. Flue gas from the DCC unit feeds the absorber at the bottom of the column, where the CO_2 is absorbed into a monoethanolamine (MEA)-based counter-current-flowing amine solvent. The absorber bottom releases an amine solution that is CO_2 rich. After being heated in the heat exchanger by the rich pump, it goes into the desorber for revival. As the CO_2 is extracted from the amine solution, it exits via the condenser and the head of the column. Lean amine, the solvent that has been recovered, is pushed by the lean pump and returned to the absorber, but before passing through into the LRHX to heat the rich amine stream. Before entering the absorber at the top to begin the next absorption cycle, it is first cooled down by a cooler to 40 C [35].

- Post-capture process: The compression of CO_2 to the necessary usage pressure occurs at this step of the process. The investigation does not take into consideration the transit or storage of the compressed CO_2.

2.1.3 The Best Integrated Technology (BIT) concept

A power plant configuration established by the CCP consortium involves three integration-related approaches that are believed to considerably lower NGCC power plants' power consumption. Configurations include exhaust gas recirculation (EGR), including an amine reboiler into the HRSG, and a low-cost CO_2 capture system that absorbs 90% of the CO_2 using a 30-wt.% MEA solvent (Figure 2.3). There has also been a techno-economic analysis performed to estimate the steam extraction point in the steam turbine that is best [43].

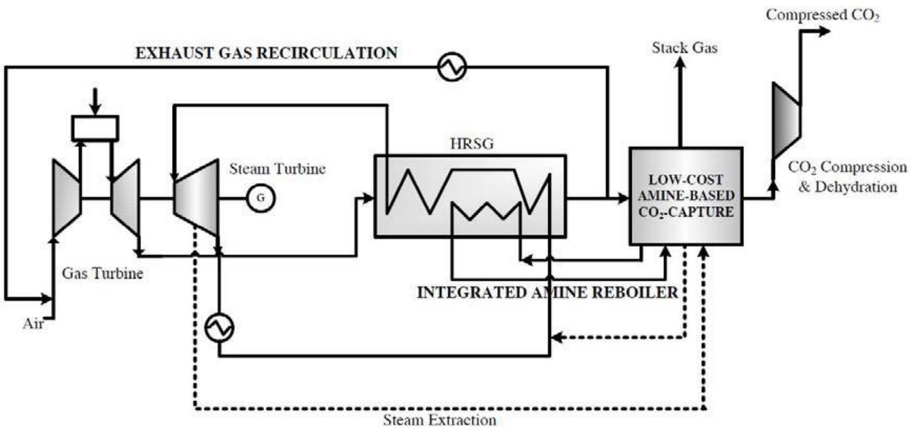

Figure 2.3 – Diagram of the BIT [43]

2.2 Base Case Aspen HYSYS Simulation

In this work, the power plant and the CO_2 capture plant were simulated using Aspen HYSYS version 12, separately based on the work of Lars Erik Øi [14], and then these two simulations were merged together and the final case was created which is the base for cost estimation and optimization.

2.2.1 Natural gas power plant simulation

Aspen HYSYS has been used to model a 400 MW gas-based, combined-cycle power plant (NGCC). Natural gas is utilized only as pure methane, air contains 79% nitrogen and 21% oxygen, 100% combustion is assumed, and conventional pressures and temperatures are employed during the process. For the thermodynamic characteristics of the power plant, the Peng Robinson model has been employed. The needed details for the combined gas power plant base case simulation using Aspen HYSYS are summarized in Table 2.1 [14].

Table 2.1: Specification and assumption for simulation of the base case model

Parameter	Value
Inlet air temperatures	25 ºC
Inlet natural gas pressure	30 bar
Combustion temperature	1500 ºC
Steam high pressure	120 bar
Steam medium pressure	3.5 bar
Steam low pressure	0.07 bar
Pressure to stack	1.01 bar
Stack temperature	100 ºC

Figure 2.3 illustrates the process flow diagram (PFD) for the Aspen HYSYS simulation of the natural gas combined cycle (NGCC).

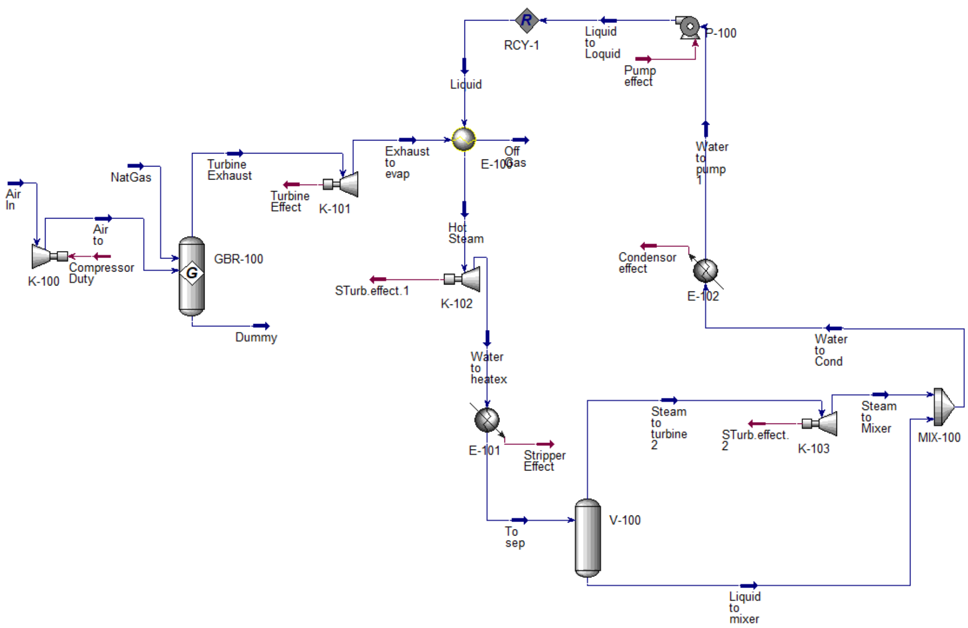

Figure 2.3: A simplified combined natural gas power plant modeled using Aspen HYSYS

The exhaust gas temperature (698 °C from the gas turbine to the exhaust at 100 °C) needs to be greater than the steam temperature throughout the whole steam heat exchanger in order for the process to be physically feasible.

Compressor, gas turbine, and steam turbine efficiencies have been put together to form a natural gas combined power plant with an overall thermal efficiency of 58%, which is the industry standard. The overall efficiency is estimated by dividing the turbine effects (minus compressor and pump effects) by the lower heating value of natural gas. The steam turbines, gas turbine's expander, and compressor efficiency were all modified to 85%, 90%, and 85%, respectively (also adiabatically). The steam delivery (for CO_2 removal) is kept at zero in this section of the power plant simulation [14].

2.2.2 CO_2 removal Simulation

A typical amine-based CO_2 capture process was simulated using Aspen HYSYS, and the results of the simulation were used for equipment sizing and cost estimation using the same approach as in the previous works [36], [37]. The input for this simulation is the flue gas from the power plant design. An Acid Gas property package for chemical solvents, which is provided in Aspen HYSYS, describes the thermodynamics for this combination. Table 2.2 displays the reactions for amine (solvent) reacting with CO_2 that are included in the Acid Gas property package [38].

Table 2.2: Reactions for the MEA solvent interacting with CO2 in the Acid Gas property package [38]

Category	No.	Reaction	Type
Water related	(1)	$2H_2O \leftrightarrow H_3O^+ + OH^-$	Equilibrium
CO_2 related	(2)	$H_2O + HCO_3^- \leftrightarrow H_3O^+ + CO_3^{2-}$	Equilibrium
	(3)	$CO_2 + OH^- \rightarrow HCO_3^-$	Kinetic
	(4)	$HCO_3^- \rightarrow CO_2 + OH^-$	Kinetic
MEA related	(5)	$HO(CH_2)_2H^+NH_2 + H_2O \leftrightarrow HO(CH_2)_2NH_2 + H_3O^+$	Equilibrium
$HO(CH_2)_2NH_2$	(6)	$HO(CH_2)_2NH_2 + H_2O + CO_2 \rightarrow HO(CH_2)_2NHCOO^- + H_3O^+$	Kinetic
	(7)	$HO(CH_2)_2NHCOO^- + H_3O^+ \rightarrow HO(CH_2)_2NH_2 + H_2O + CO_2$	Kinetic

Applying equilibrium stages with user-defined stage (Murphree) efficiency, the absorber and desorber were modelled. Constant Murphree efficiency of 0.25 and 1.00 was determined to be equivalent to one meter of structured packing for the absorber and desorber, respectively. These Murphree efficiencies are computed by dividing the variation in CO_2 mole ratio across stages by the variation in equilibrium assumption [39].

The specification of calculation which corresponds to an 85 % CO_2 removal efficiency and a minimum approach temperature of 10 °C in the lean/rich amine heat exchanger is listed in Table 2.3. The computing method is equal to that used in earlier investigations [36], [40], [41].

Table 2.3: Model parameters and specifications for Aspen HYSYS [14]

Items	Specifications [Unit]	Value
Inlet Flue Gas	Temperature [°C]	40
	Pressure [bar]	1.1
	Molar flow rate [kmol/h]	85000
	CO_2 content [mole %]	3.73
	H_2O content [mole %]	6.71
Lean MEA	Temperature [°C]	40
	Pressure [bar]	1.1
	Molar flow rate [kmol/h]	120000 *
	MEA content [W %]	29
	CO_2 content [W %]	5.5
Absorber	Number of stages	10
	Murphree efficiency	0.25
	Rich amine pump pressure [bar]	2
	Rich amine temp. out of Lean/Rich amine HEx [°C]	104.5 *
Desorber	Number of stages in stripper	6
	Murphree efficiency	1.00
	Reflux ratio in the desorber	0.3
	Reboiler temperature [°C]	120
	Pressure [bar]	2
	Lean amine pump pressure [bar]	5
*) In the first iteration		

Figure 2.4 illustrates the process flow diagram (PFD) for the Aspen HYSYS simulation of CO_2 removal. The absorption column is initially calculated (first assumed) from the input data of flue gas and lean amine [41]. Rich amine is transferred by the pump from the bottom of the

absorption column to the lean/rich amine heat exchanger. After the absorber and rich pump, the temperature is increased to some degree and then in the lean/rich amine heat exchanger the duty of the heat exchanger based on the assumed output temperature of lean/rich amine heat exchanger is determined. As the heated rich amine reaches the desorption column, then the CO_2 product and the hot lean amine are calculated at the outlet. In the lean amine pump, the heated lean amine is pumped to a higher pressure before passing through the lean/rich heat exchanger and after that being cooled in the lean amine cooler. The lean amine is then entered into a recycling block. It is assessed whether the flow and condition of the recycled lean amine are the same as that of the previously calculated lean amine flow, which may then be modified by iteration. The recycling block in Aspen HYSYS evaluates the block's in-stream to the block's out-stream in the last iteration to solve the flowsheet [42].

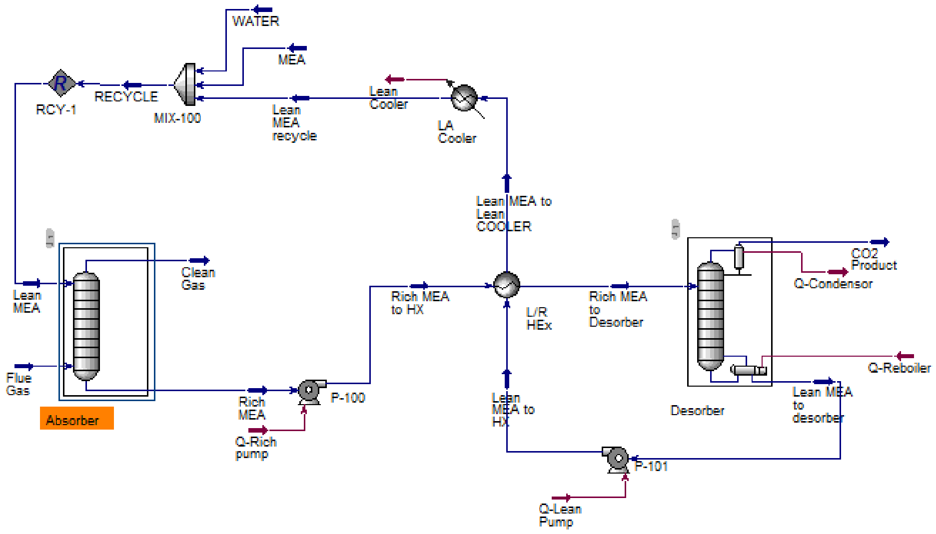

Figure 2.4: A simplified Aspen HYSYS flow sheet model for CO_2 removal

2.2.3 Base case simulation

The natural gas power plant and CO_2 capture plant which are simulated in the above subchapters are connected together and the final base case simulation is modelled. The Flue gas of the natural gas power plant is transferred to the CO_2 capture plant. The 400 MW net electricity output and 85% CO_2 removal is the goal in the Aspen HYSYS simulation. The concept of the new simulation is the same as previous simulations, the only difference is that the amount of input flue gas to the CO_2 removal plant is defined based on the output of the combined cycle power plant.

For the new simulation, some of the previous specification based on the exhaust gas of the cycle power plant is changed. Table 2.4 is shown the new specification data for the base case simulation which corresponds to generating 400 MW, 1500 °C combustion chamber exhaust temperature, 85% CO_2 removal, minimum approach temperature of 10 °C in the lean/rich amine heat exchanger and 40 °C temperature of input flue gas to the absorber.

Table 2.4: Aspen HYSYS specifications for Base Case model

Items	**Specifications [Unit]**	**Value**
Inlet Flue Gas	Temperature [°C]	40
	Pressure [bar]	1.1
	Molar flow rate [kmol/h]	71345
	CO_2 content [mole %]	4.61
	H_2O content [mole %]	6.71
Lean MEA	Temperature [°C]	40
	Pressure [bar]	1.1
	Molar flow rate [kmol/h]	99496
	MEA content [W %]	28.92
	CO_2 content [W %]	5.39
Absorber	Number of stages	10
	Murphree efficiency	0.25
	Rich amine pump pressure [bar]	2
	Rich amine temp. out of Lean/Rich amine HEx [°C]	102.7
Desorber	Number of stages in stripper	6
	Murphree efficiency	1.00
	Reflux ratio in the desorber	0.3
	Reboiler temperature [°C]	120
	Pressure [bar]	2
	Lean amine pump pressure [bar]	5

Figure 2.5 illustrates the process flow diagram (PFD) for the Aspen HYSYS base case simulation. Each stream's name has been selected as a label in the PFD for the fluid it carries to the next device.

Five adjust operations were added to the process in order to fulfill the needs of the simulation, doing the analysis and designing an automated simulation model. The Combustion temperature is set according to the flow rate of input air by using ADJ-1, the net electricity output is set based on the natural gas flow rate by using ADJ-2, and the CO_2 removal efficiency is set according to the flow rate of lean amine by using ADJ-3, the minimum approach temperature in the lean/rich heat exchanger can be used to adjust based on the rich amine outlet temperature of the lean/rich heat exchanger by using ADJ-4, and the ADJ-5 modifies the cooling water demand in the inlet cooler to adjust the temperature of the flue gas to the absorber.

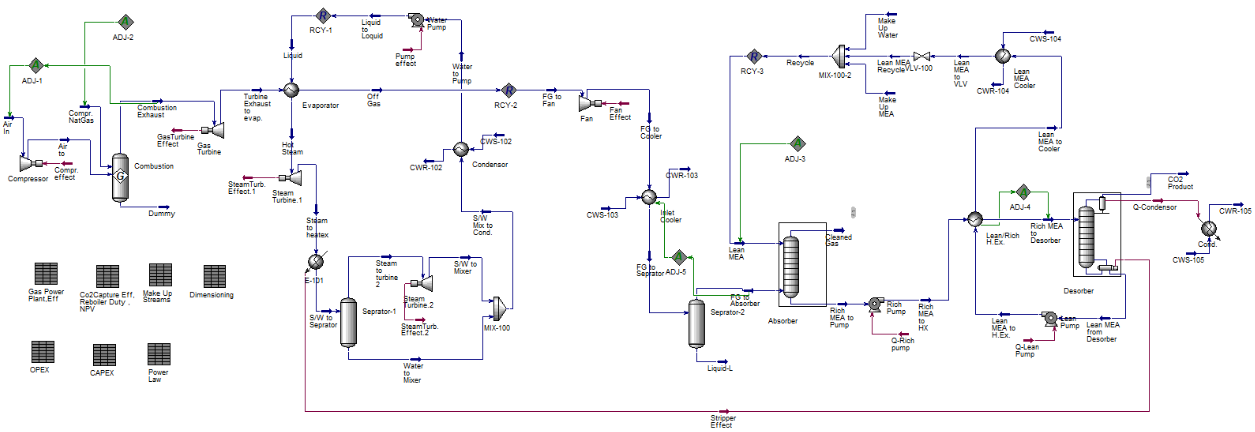

Figure 2.5: Aspen HYSYS flow sheet model for Base case

For better performance of the simulation, two other recycling blocks are also considered. One of them is for the recycling water process of the steam turbine and the other one is in the place the two simulations are connected to each other, before the CO_2 capture unit.

In the process, there is a lack of amine solution and water. So, makeup streams are injected into the main streams to recover it. In order to perform this, a spreadsheet for makeup streams was developed, and the lack of water and amine solution was calculated using mass balance and sent as mass flow to the makeup amine solution and makeup water flowing. Before the absorber, a fan unit, an inlet cooler, and a separator are considered in order to reach the target temperature and pressure of the input flue gas to the absorber and separate the water from the flue gas.

3 Dimensioning of equipment

In this section, the process equipment that is covered by the process simulation of the base case is basically dimensioned. The estimations are provided using the results of the process flowsheet of the base case simulation such as temperatures, flow rates, and heat and power duties. The following tables show output values from Aspen HYSYS while the other values are calculated or based on assumptions. The computations and assumptions used for dimensioning are only partially displayed in the tables. Only the main pieces of equipment, such as the gas and steam turbines, compressors, absorption and desorption columns, evaporator, condenser, heat exchangers, fans, pumps, and separators, are dimensioned. This section does not consider pre-treatments like inlet gas purification or post-treatments like CO_2 compression, transport, or storage.

Aspen Icarus Reference Guide is considered as the reference for dimensioning of mentioned equipment [44].

3.1 Compressor

The actual gas flow rate inlet is the main parameter in the design of the compressor. The dimensioning of the compressor based on the defined design parameter is shown in Table 3.1.

Table 3.1: Compressor dimensioning

Parameter	Value
Flow rate [m^3/h]	1705624
Max flow rate [m^3/h]	509700
Calculated No. of units	3.34
Actual no. of units	4.00
Calculated Flow rate [m^3/h]	426406

The centrifugal gas compressor with driver (motor, turbine, or gasoline-driven engine) with a maximum actual gas flow rate Inlet of 509,700 M^3/H is defined for base case simulation dimensioning [44]. Based on the required actual gas flow rate four compressors are considered in the dimensioning.

3.2 Gas turbine with the combustion chamber

Power output is the key parameter design of a gas turbine. Based on the defined power output design in Aspen Icarus Reference Guide [44] and the Aspen HYSYS power output data the following dimensioning is done.

Table 3.2 is shown the gas turbine dimensioning based on the base case simulation demand.

The maximum power output of a gas turbine which is possible to consider in this scenario is 375,000 KW. Based on the required power output two gas turbines are needed.

Table 3.2: Gas turbine with combustion chamber dimensioning

Parameter	Value
Power output (KW)	589716
Max Power output (KW)	375000
Calculated the Number of Unit	1.57
Actual Number of Unit	2.00
Power output per unit (KW)	294858

3.3 Steam Turbine

Power output is the major parameter design of a steam turbine. The performance and dimensioning of the steam turbines can be seen in Table 3.3.

Table 3.3: Steam turbines dimensioning

Parameter	Steam turbine 1 (HP)	Steam turbine 2 (LP)
Power output (KW)	84,196	29303
Max Power output (KW)	22,300	22300
Calculated the Number of Unit	3.77	1.31
Actual Number of Unit	4.00	2.00
Power output per unit (KW)	21,049	14651

The maximum power output of a non-condensing type steam turbine which is possible to consider in this scenario is 22,300 KW [44]. Based on the required power output four steam turbines for high-pressure load and two steam turbines for low-pressure load are needed.

3.4 Evaporator

The common dimensioning factor in the design of heat exchanger cost is the area of the heat transfer. To calculate the level of the required heat transfer in the evaporator, the overall heat transfer coefficient of 0.25 $KW/m^2 °K$ is specified [45]. The output exhaust from the expander of the gas turbine with the temperature 698.6 ℃ is the input parameter and the source of heat to the evaporation of water which is used in the steam turbine. The output flue gas from the evaporator with 100 ℃ temperature is assumed for the calculation [14].

The Long tube vertical evaporator with the maximum 4640 m^2 heat transfer area is the one that is specified in this dimensioning [44]. based on the total required heat transfer in the simulation four evaporator is considered.

Table 3.4 shows the major dimensioning parameters and the computed duty from Aspen HYSYS.

Table 3.4: Evaporator dimensioning

Parameter	Value
Q [KJ/h]	1418604568
Heat transfer coefficient [KW/m^2°K]	0.25
LMTD	111.7
Total Heat Transfer Area [m^2]	14117
Max. Area per Unit [m^2]	4640
Calculated the Number of Unit	3.04
Actual Number of Unit	4.00
Actual Area per unit [m^2]	3529

3.5 Condenser

The process water in the steam turbine cycle must change from the vapour phase to the liquid phase. In order to change it for reuse before the pump, the condenser was considered. The actual volume flow rate of cooling water is the key parameter in the design of the condenser. The dimensioning of the compressor based on the defined design parameter is shown in Table 3.5.

Table 3.5: Condenser dimensioning

Parameter	Value
Water flow rate [m^3/h]	13014
Water flow rate [L/S]	3615
Max water flow rate [L/S]	0.25
Calculated the Number of Unit	11.48
Actual Number of Unit	12
Actual water flow rate [L/S]	301.3

The barometric condenser with a maximum actual water flow rate Inlet of 315 L/S is defined for base case simulation dimensioning [44]. Based on the required actual water flow rate twelve condensers are considered in the dimensioning.

3.6 Absorber Column

Table 3.6 shows the absorber column's performance and dimensioning.

The absorber's overall CO_2 removal rate of 85%, which is originally modified by the flow rate of lean amine under the configuration of the base scenario, is regarded as a setpoint.

Table 3.6: Absorber column dimensioning

Parameter	Value
FG volume flow [m^3/h]	1773211
FG volume flow [m^3/s]	492.6
FG velocity [m/s]	2.5
Inner Diameter [m]	15.84
Packing height [m]	10
Column height [m]	25
Column volume [m^3]	4926
Packing volume [m^3]	1970
Number of units	3
Column Volume per unit [m^3]	1642
Packing volume per unit [m^3]	656.7
Diameter per unit [m]	9.144
SHELL MAT.	SS316
PACKING TYPE	MellaPak 250Y

Vertical flue gas velocity via the packed column is designed at 2.5 m/s, this is based on 75% of the Mellapak 250Y structured packing types' flooding velocity [46]. Three absorbers were taken into account due to the considered gas velocity and high flue gas volume flow rate, resulting in an absorber column diameter of 9.14 m per unit. A total height of 25 meters is expected for the absorption columns. In calculating absorber height, packaging, water wash, gas inflow and outflow, liquid distributors, sump, and demister are all considered.

3.7 Desorber Column

Table 3.7 displays the desorber column's performance and design characteristics.

Table 3.7: Desorber column dimensioning

Parameter	Value
FG volume flow [m^3/h]	108455
FG volume flow [m^3/s]	30.13
FG velocity [m/s]	1.00
Inner Diameter [m]	6.2
Packing height [m]	6
Column height [m]	15

Column volume [m^3]	451.9
Packing volume [m^3]	180.8
Number of units	1
SHELL MAT.	SS316
PACKING TYPE	MellaPak 250Y

From Kallevik's master thesis, the computation of vertical gas velocity in the column was applied. [42]. The output is a column with a 6.2 m diameter. It is assumed that the column will be 15 meters high overall, with 6 meters of structured packing and a Murphree stage efficiency of 100% per meter.

3.8 Heat Exchangers

The operation and dimensioning information for the lean/rich heat exchangers, lean MEA cooler, inlet cooler, reboiler, and condenser are provided in this chapter. Similar to other sections, a calculating scheme has been used, and it has been considered that all heat exchangers are shell and tube heat exchangers.

3.8.1 Lean/Rich heat exchanger

Table 3.8 displays the determined duty from Aspen HYSYS as well as the key dimensioning variables for a lean/rich heat exchanger.

The overall heat transfer coefficients for the lean/rich heat exchanger were specified to 732 W/(m^2. K) [47]. LMTD is calculated based on ΔT_{min} set to 10˚C. For the heat duty obtained, the total heat transfer area is determined. Based on the total heat transfer area and the maximum area of each heat exchanger unit, 16 heat exchanger units are required and the actual area of each unit is calculated as 952.8.

Table 3.8: Lean/rich heat exchanger dimensioning

Parameter	**Value**
Q [KJ/h]	535704681
Heat transfer coefficient [kW/m^2. K]	0.732
LMDT	13.33
Total Heat Transfer Area [m^2]	15245
Max. Area per Unit [m^2]	1000
Calculated the Number of Unit	15.25
Actual Number of Unit	16
Actual Area per Unit [m^2]	952.8

3.8.2 Lean MEA Cooler

Table 3.9 displays the determined duty from Aspen HYSYS as well as the key dimensioning variables for the lean MEA cooler.

Table 3.9: Lean MEA Cooler dimensioning

Parameter	**Value**
Q [KJ/h]	106752207
Heat transfer coefficient [kW/m^2. K]	0.8
LMDT	27.03
Total Heat Transfer Area [m^2]	1371
Max. Area per Unit [m^2]	1000
Calculated the Number of Unit	1.37
Actual Number of Unit	2
Actual Area per Unit [m^2]	685.7

The overall heat transfer coefficients for the lean MEA cooler were specified at 800 W/ (m^2. K) [47]. A lean MEA cooler's cold and hot sides' minimum temperatures are used to determine the LMTD. For the heat duty obtained, the total heat transfer area is determined. Based on the total heat transfer area and the maximum area of each lean MEA cooler unit, two lean MEA cooler units are required and the actual area of each unit is calculated as 685.7.

3.8.3 Inlet Cooler

Table 3.10 displays the inlet cooler's performance and calculations.

Table 3.10: Inlet cooler dimensioning

Parameter	**Value**
Q [KJ/h]	236564059
Heat transfer coefficient [kW/m^2. K]	0.8
LMDT	49.47
Total Heat Transfer Area [m^2]	1660
Max. Area per Unit [m^2]	1000
Calculated the Number of Unit	1.66
Actual Number of Unit	2
Actual Area per Unit [m^2]	830.2

Inlet coolers were designed with an overall heat transfer coefficient of 800 W/ (m^2. K) [47]. The LMTD is calculated from the minimum temperatures of the cold and hot sides of the inlet cooler. For the heat duty obtained, the total heat transfer area is determined. Based on the total heat transfer area and the maximum area of each inlet cooler unit, two inlet cooler units are required and the actual area of each unit is calculated as 685.7.

3.8.4 Reboiler (desorber)

The calculations and performance of the boiler used in the desorber are shown in Table 3.11.

Table 3.11: Reboiler dimensioning

Parameter	**Value**
Q [KJ/h]	447016231
Heat transfer coefficient [Kw/m^2. K]	1.20
LMDT	21.35
Total Heat Transfer Area [m^2]	4848
T(out,cold)	120
T(in,cold)	114.8
T(in,hot)	138.8
T(out,hot)	138.8
Max. Area per Unit [m^2]	1000
Calculated the Number of Unit	4.84
Actual Number of Unit	5
Actual Area per Unit [m^2]	969.5

The reboiler duty is calculated by Aspen HYSYS. The calculated LMDT is calculated based on the temperatures of the cold and hot sides of the reboiler. The overall heat transfer coefficient U is found in literature and is assumed to be constant at 1200 W/ (m^2. K) [47]. By using the heat exchanger equation, the total required heat exchanger area is calculated to be 4848 m^2.

Based on the total heat transfer area and the maximum area of each inlet cooler unit, five reboiler units are required and the actual area of each unit is calculated as 969.5.

3.8.5 Condenser (desorber)

Table 3.12 displays the calculations and results of the condenser employed in the desorber.

Table 3.12: Condenser-D dimensioning

Parameter	**Value**
Q [KJ/h]	57434353

Heat transfer coefficient [kW/m^2. K]	1.00
LMDT	76.42
Total Heat Transfer Area [m^2]	208.8
T(out,cold)	25
T(in,cold)	15
T(in,hot)	100.8
T(out,hot)	92.04
Max. Area per Unit [m^2]	1000
Calculated the Number of Unit	0.2088
Actual Number of Unit	1
Actual Area per Unit [m^2]	208.8

The condenser duty is calculated by Aspen HYSYS. The calculated LMDT is calculated based on the temperatures of the cold and hot sides of the condenser. The overall heat transfer coefficient U is found in literature and is assumed to be constant at 1000 W/ (m2. K) [47]. By using the heat exchanger equation, the total required heat exchanger area is calculated to be 208.8 m^2 and one condenser unit is required.

3.9 Fan and pumps

This section deals with the dimensions considered for the flue gas fan, water pump, lean pump, and rich pump. The fan and pumps in Aspen HYSYS are designed to achieve an adiabatic efficiency of 75%. The duty is used as a sizing criterion for the fans and pumps that were acquired from Aspen HYSYS, but the Aspen In-Plant cost calculator uses volumetric flow to determine equipment costs.

3.9.1 Flue gas fan

The flue gas fan's output parameters from Aspen HYSYS are displayed in Table 3.13.

Table 3.13: Flue gas fan dimensioning

Parameter	Value
Duty [kW]	6978
Flow rate [m^3/h]	2237966
Max flow rate [m^3/h]	1529000
Calculated No. of units	1.46
Actual no. of units	2.00
Actual Flow rate [m^3/h]	1118983
Actual Duty [kW]	3489

As mentioned before, the actual flow rate is the key parameter design for a flue gas fan. A centrifugal fan with a maximum actual flow rate of 1,529,000 m^3/h is specified in the dimensioning [44]. Based on obtained actual flue gas flow rate two fans are required in the dimensioning.

3.9.2 Rich amine pump

It is necessary to pump the rich amine solvent to the desorber. The operational pressure of the desorber which is 2 bars, is provided after the rich amine pump. Table 3.14 displays the Aspen HYSYS values for the rich amine pump.

Table 3.14: Rich amine pump dimensioning

Parameter	**Value**
Duty [kW]	72.28
Flow rate [m^3/h]	2168
Flow rate [L/s]	602.3
Actual no. of units	1.00

Based on the calculated flow rate by Aspen HYSYS, a centrifugal pump is defined for this process [44] and according to the specification of the pump, one unit is enough for this design.

3.9.3 Lean amine pump

A pump is needed after the desorber. It is connected to height of the absorber, which will require more consumed power of the pump to accomplish the necessary lifting height. The installation of a lean amine pump increased the lean amine's pressure to 5 bar. The details of the Aspen HYSYS calculation for the rich amine pump are shown in Table 3.15.

A centrifugal pump is specified for this process [44] based on the flow rate that Aspen HYSYS estimated, and one unit of the pump is sufficient for this design.

Table 3.15: Lean amine pump dimensioning

Parameter	**Value**
Duty [kW]	249.9
Flow rate [m^3/h]	2249
Flow rate [L/s]	624.8
Actual no. of units	1.00

3.9.4 Water pump

To circulate the condensed water for reuse in the steam turbine, a water pump is considered after the condenser and before the evaporator. In table 3.16, the specifics of the Aspen HYSYS calculation for the water pump are displayed.

Table 3.16: Water pump dimensioning

Parameter	Value
Duty [kW]	1863
Flow rate [m^3/h]	420.5
Flow rate [L/s]	116.8
Actual no. of units	1.00

Based on the projected flow rate by Aspen HYSYS, a centrifugal pump is specified for this procedure [44], and one unit of the pump is adequate for this design.

3.10 Separators

To provide the heat flow for the reboiler in the desorber, steam outlet from the first steam turbine is used for this process. Due to this heat transfer some of the steam turns into water and it should be separated before entering into the second steam turbine.

Also, before the flue gas flows through the absorber, a very little amount of water may be present due to the temperature decrease in the inlet cooler. The flue gas can enter into the absorption column at about 40 °C if the water is extracted from the flue gas in the separator.

3.10.1 Separator-1

Table 3.17 displays the water separator used to separate liquid water from steam.

The Souders–Brown equation was used to calculate the separator's diameter, which was calculated as a vertical vessel with a k-factor of 0.107 m/s and a tangent-to-tangent to diameter ratio of 1 [42]. Also, for the purpose of the cost estimate, vessel volume is used in this scenario.

Table 3.17: Separator-1 dimensioning

Parameter	Value
Actual Gas Flow Rate [m^3/h]	113224
Actual Gas Flow Rate [m^3/s]	31.45
Liquid Phase Mass Density [kg/m^3]	914
Gas Phase Mass density [kg/m^3]	1.88
K Factor, Sounder-Brown Velocity	0.107
Allowable Vapour Velocity [m/s]	2.354

Vessel Cross-Sectional Area [m^2]	13.36
Vessel Inner-Diameter (Di) [m]	4.124
Vessel Height, 1D [m]	4.124
Vessel Volume [m^3]	55.10

3.10.2 Separator-2

The water separator that is used to separate liquid water from flue gas is shown in Table 3.18.

Table 3.18: Separator-2 dimensioning

Parameter	**Value**
Actual Gas Flow Rate [m^3/h]	1686469
Actual Gas Flow Rate [m^3/s]	468.5
Liquid Phase Mass Density [kg/m^3]	996
Gas Phase Mass density [kg/m^3]	1.207
K Factor, Sounder-Brown Velocity	0.15
Allowable Vapour Velocity [m/s]	4.306
Vessel Cross-Sectional Area [m^2]	108.8
Vessel Inner-Diameter (Di) [m]	11.77
Vessel Height, 1D [m]	11.77
Vessel Volume [m^3]	1280

The diameter of the separator was determined using the Souders-Brown equation as a vertical vessel with a k-factor of 0.15 m/s and a diameter-to-tangent ratio of 1 [42]. Additionally, in this case, vessel volume is employed to evaluate costs.

4 Cost estimation method

Cost estimation's primary goal in this chapter is to determine the project's overall cost, which includes the combined cycle power plant and CO_2 capture plant. The dimensioning from the Aspen HYSYS results is the basis for the cost estimation calculations. Using the Aspen In-Plant Cost Estimator, prices for each component are calculated in the base case model.

The method described below is used to estimate the overall cost of the plant based on the simulation model:

i. Using equipment dimensioning data for the Base Case, Aspen In-Plant Cost Estimator (v.12) calculates the cost of the equipment.
ii. Using the Enhanced Detailed Factor (EDF) to specify the overall installation cost
iii. Using cost index correction to estimate present value (convert to the present year)
iv. Calculation of yearly operational expenses (OPEX)
v. Net present value (NPV) calculation using a specified discount rate and plant lifetime
vi. Scaling the cost during parameter modification using the Power Law method

4.1 CAPEX (Capital expenditure)

The version 12th of the Aspen In-Plant cost estimator program, was utilized to calculate the CAPEX in this study. Additionally, to achieve the capital cost calculation, the Enhanced Detail Factor Method, often known as the EDF Method, was employed. This approach is based on variables that have an impact on how each piece of process equipment is installed. The potential to optimize a specific piece of equipment has been made possible via the enhanced detail factor approach (EDF). Additionally, this approach has made it possible to do a techno-economic analysis for developing existing process plants as well as new technologies [48].

4.1.1 Cost of equipment

The Aspen In-Plant, a cost estimation software that combines process information, dimensioning factors, and material to provide accurate estimates for the overall equipment expense, was used to determine the price of each item of equipment. The other specifications were given default values by Aspen In-Plant, with the exception of the dimensioning factors covered in the previous chapter. Aspen In-Plant provides the pricing in Euro (€) for the year 2019, with Rotterdam, Netherlands, as the default location.

In the appendix B attachment, Nils Henrik Eldrup' detailed installation factor table, all of the parameters are for carbon steel (C.S.). Most equipment is made of stainless steel, but some equipment is also made of carbon steel. To apply the Nils detailed installation factor, the cost of stainless steel must be changed to the cost of carbon steel (CS) by the use of material factor based on the EDF method [48].

This has been accomplished using Equation (4.1).

$$\text{Equipment Cost}_{CS} = \frac{\text{Equipment Cost}_{SS}}{f_{mat}} \tag{4.1}$$

where:

- Equipment $Cost_{CS}$ is the cost of an item made of Carbon Steel.
- Equipment $Cost_{SS}$ is the cost of an item made of stainless steel.
- f_{mat} is the material factor that changes SS into CS.

The material factor to change expenses for rotating and welded equipment from SS316 to CS is 1.30 and 1.75, respectively [36].

4.1.2 Total installation cost calculation

The cost of the equipment can be used to determine the overall cost of the plant using the table of installation factor 2020 in Appendix B. This table contains the engineering, administration, commissioning, and contingency costs in addition to the direct costs.

Applying Equation (4.2) provides the total installation cost for each item of the equipment purchase cost.

$$C_i = C_p . [f_{TC} - f_p - f_E + f_m . (f_p + f_E)] \tag{4.2}$$

Where:

- Ci = Total installed cost, carbon steel [€]
- C_P = Equipment purchase cost, carbon steel [€]
- f_{TC} = Cost factor of the total installation
- f_P = Cost factor of equipment piping
- f_E = Cost factor of equipment
- f_m = Cost factor of material

4.1.3 Currency and inflation Settings

In this study, all cost calculations are made in Euros (€). The Aspen In-Plant cost calculator is used to evaluate the equipment's cost in euros. The factor table for the EDF approach also provides the equipment cost's currency in euros [36].

Version 12th of the Aspen In-Plant cost estimator evaluates equipment expenses through obtained data in 2019. It indicates that the expense must be indexed to inflation in order to obtain a current and accurate cost estimate. The information used to calculate the installed cost factors in the detail factor table is for 2020. As a result, the equipment cost first needs to be updated to reflect cost information as of 2020. After that, the cost of the total installation will be estimated by the use of the EDF method. As a final step, inflation must be applied to the overall installed cost from 2020 to the year after.

The cost has been converted from year a to year b using equation (4.3):

$$Cost_a = Cost_b \left(\frac{\text{Cost index}_a}{\text{Cost index}_b}\right) \tag{4.3}$$

The cost indexes for the present work are written in Table 4.1 [49]:

Table 4.1: Cost inflation indexes

Year	Value
2019	110.1
2020	112.2
2021	116.1

In the Aspen HYSYS simulation, a spreadsheet with the label CAPEX contained all the procedures related to calculating the CAPEX for the Base Case.

4.1.4 Power Law

The new facility's cost is determined using a comparable facility with a different capacity. Equation (4.4) illustrates the correlation of the power law capacity [50].

$$C_E = C_B\left(\frac{Q}{Q_B}\right)^M \tag{4.4}$$

Where:

- C_E = Equipment with Q capacity cost
- C_B = Equipment with Q_B capacity cost (calculated)
- M = Cost exponent or scaling constant according to the type of equipment

The valve for a scaling constant is typically 0.4 < M < 0.9, but it is typically considered to be 0.65.

4.2 OPEX (Operating expenditure)

For a comprehensive cost estimate, operational costs (OPEX) must be assessed in addition to capital costs (CAPEX). The OPEX includes the cost of electricity, natural gas cost, process and cooling water costs, solvents cost, and maintenance cost, as well as the salaries of the engineers and operators that work on the project.

Equation (4.5) can be used to compute the annual cost of the supplied utilities [48].

$$\text{Anual utility Cost}\left(\frac{euro}{year}\right) = \text{Consumption}\left(\frac{Unit}{hour}\right) \times \frac{\text{Operating hours}}{\text{year}} \times \text{Utility price}\left(\frac{euro}{Unit}\right) \tag{4.5}$$

Table 4.2 presents the OPEX specifications and assumptions [36], [47], [51].

Table 4.2: OPEX requirements and underlying presumptions

Item	Value	Unit
Operating Lifetime	25	[year]
Construction Lifetime	3	[year]
Operation Lifetime	22	[year]
Discount rate	7.5 %	-
Operating Hours	8000	[h/year]
Electricity Price	0.136	[€/kWh]
Natural gas Price	1.29	[€/m^3]
Cooling water Price	0.022	[€/m^3]
Water process Price	0.203	[€/m^3]
Solvent MEA Price	1450	[€/ton]
Maintenance Price	4% of CAPEX	[€/year]
Operator Price	80414 × (12 Operators)	[€/year]
Engineer Price	156650 × (2 Engineer)	[€/year]

All the steps involved in computing the OPEX for the Base Case were contained in a spreadsheet with the label OPEX in the Aspen HYSYS simulation.

4.3 Income from electricity sales

In this project, selling the electricity produced in the combined power plant by gas turbine and steam turbine is profitable and this income is included in the cost estimate.

The produced electricity is 400 MW and it was maintained constant during the evaluation of each scenario. The price of electricity is considered to be 0.136 [€/kWh] based on the average price in the months of 2022. Although there is a possibility that the price of electricity will increase in the coming years, the price of electricity is considered constant during the years of project implementation.

Norway's average monthly wholesale power price from January 2019 to November 2022 is shown in figure 4.1[51].

4.4 NPV (Net Present Value)

It is possible to evaluate the entire cost over time by taking both into account once the capital expenses associated with installing the required equipment and the operational cost of utilities are understood. The total cost of a project is determined using the net present value (NPV) technique, which considers both capital and operating expenses, income for the specified period of time, and discount rate. All costs of installation for the key pieces of equipment used in the combined power plant and the CO_2 capture plant are included in the capital cost in this calculation.

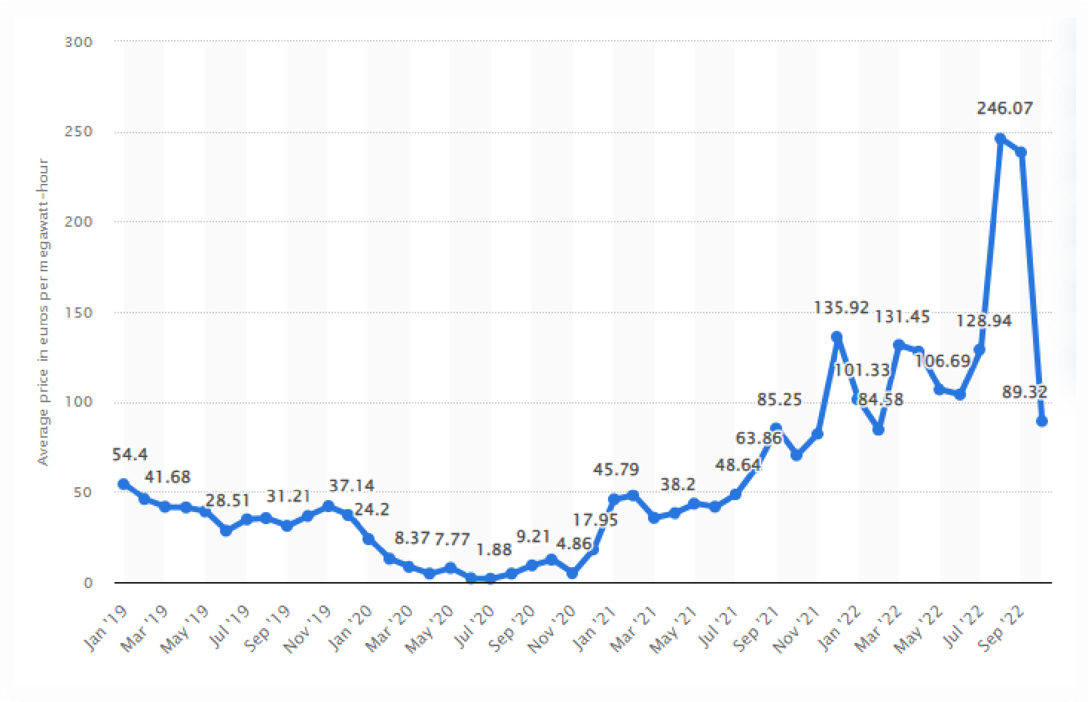

Figure 4.1: Norway's wholesale electricity prices for 2019-2022 [51]

The CAPEX is anticipated to happen in years 0 through 2 while the operational costs and incomes take place after this time.

Equation (4.7) depicts the expression for calculating the NPV of the total future operational costs [42].

$$\text{NPV}_{\,OPEX} = \sum_{N=3}^{End} \left\{ (a) \times \frac{1}{(1+i)^N} \right\} \tag{4.7}$$

Where:

- NPV_{OPEX} = The total cost of OPEX for the calculation period [€]
- i = annual interest rate
- a = annual operation cost [€]
- N = number of years

Equation (4.8) is used to determine the NPV for the total process in this work.

$$NPV = CAPEX + NPV_{OPEX} \tag{4.8}$$

Where:

- NPV= Net present value for the total costs [€]
- CAPEX = Installation expenses for equipment [€]

The calculated NPV in equation (4.8) considered all the incomes and costs related to the utilities and also the CAPEX. The NPV of the early years is negative, but it will be positive after this period due to the income related to the sale of electricity. A higher NPV indicates that the project is more profitable.

As was previously stated, the project's evaluation period is 25 years, and the discount rate is 7.5%.

5 Sensitivity analysis

The present chapter investigates the impact of varying different key process parameters to obtain the optimal trade-off for a gas-based power plant integrated with a CO_2 capture plant. This process optimization is based on the economic evaluation that gives us the highest profit achieved during the defined period.

There are several parameters that change in a gas-based power plant, including inlet air /gas temperature and pressure into the power plant, and the outlet temperature of flue gas (off the gas). In the CO_2 capture plant, however, there are a number of parameters that change, including the gas inlet temperature, minimum temperature approach in the lean/rich heat exchanger, and packing height in the absorption column.

The primary objective of this study is to use the Aspen HYSYS to estimate and optimize mentioned parameters according to the highest NPV of the combined gas power plant integrated with the CO_2 capture plant.

The steps to calculate the NPV are as follows: first, CAPEX and OPEX are calculated for the base case, and after changing the parameters, the new dimensioning parameters are transferred from dimensioning spreadsheet to the power law spreadsheet to calculate the new CAPEX and also the new operational cost is updated in the OPEX spreadsheet automatically based on the utility usage. The income from selling electricity is also added to the NPV evaluation.

In the NPV evaluation for cost optimization of process parameters, CAPEX and OPEX are considered negative values and income from electricity sales is considered a positive value.

5.1 Inlet temperature into the power plant (Ambient temperature)

To check the optimal inlet temperature (ambient temperature) based on the highest NPV, the inlet air temperature and the natural gas inlet temperature of the power plant must be set manually at the same time. The default temperature for the base case simulation is 25°C.

During each scenario, the inlet temperature was varied while the exhaust combustion temperature was maintained at 1500°C, the net electricity produced was maintained at 400 MW, the CO_2 removal efficiency remained at 85%, the minimum approach temperature was maintained at (ΔT_{min}) 10 °C and the inlet flue gas temperature to the absorber was maintained at 40 °C by ADJ-1, ADJ-2, ADJ-3, ADJ-4, ADJ-5, respectively.

In this evaluation, the inlet temperature varied from 10°C to 50°C with step 5°C. It should be noted that all factors and presumptions used in this evaluation were maintained at their default values for the base case scenario.

5.2 Inlet pressure into the power plant

In order to get the most benefit of the project based on the change of the inlet pressure to the power plant, the inlet air pressure and the compressed natural gas pressure to the combustion chamber should be adjusted manually at the same time. It should be noted that the compressor is used to change atmospheric air to a defined pressure before entering into the combustion chamber.

The default pressure for the base case simulation is 30 bar. For this investigation, the power plant's input pressure ranged from 15 bar to 40 bar with a step of 5 bar.

Each scenario involved varying the inlet pressure while maintaining the following parameters: the exhaust combustion temperature was kept constant at 1500°C, the net electricity produced was kept constant at 400 MW, the CO_2 removal efficiency remained at 85%, the minimum approach temperature was kept constant at (ΔT_{min}) 10°C, and the inlet flue gas temperature to the absorber was kept constant at 40°C.

5.3 Flue gas outlet temperature (Off-gas)

The temperature of output flue gas from the gas turbine after passing through the heat exchanger which is used to produce steam will drop to a default temperature of 100°C. This flue gas was totally transferred to the CO_2 capture plant for CO_2 removal.

In this investigation, the outlet temperature of flue gas was varied from 50°C to 150°C with a step of 10°C to obtain the highest NPV during the implementation of the project. A case study can be defined in the Aspen HYSYS to automatically calculate NPV for varied flue gas outlet temperature values.

Like other previous analyses, the case study involved varying the flue gas outlet temperature while keeping the exhaust combustion temperature constant at 1500°C, the net electricity produced constantly at 400 MW, the CO_2 removal efficiency constant at 85%, the minimum approach temperature constant at (ΔT_{min}) 10 °C, and the inlet flue gas temperature to the absorber constant at 40 °C.

5.4 Minimum temperature approach in the lean-rich heat exchanger (ΔT_{min})

The objective of this part is to establish the minimum approach temperature for the lean-rich heat exchanger with the maximum NPV. NPV can be calculated manually in Aspen HYSYS or by creating a case study for varied ΔT_{min} values. In order to accomplish this, a case study was conducted to evaluate the lean/rich amine heat exchanger's economic performance using the minimum approach temperature (ΔT_{min}).

In this case study, ΔT_{min} was varied by adjusting the outlet temperature of the rich amine from the lean/rich amine heat exchanger, while the boiler outlet temperature was fixed at 120°C. This will occur in the ADJ-2 while ADJ-1, ADJ-2, ADJ-3, and ADJ-5, respectively, maintained the exhaust combustion temperature at 1500°C, the net power produced at 400 MW, the CO_2 removal efficiency at 85%, and the incoming flue gas temperature to the absorber at 40 °C.

For the specified base case, 10 °C has been taken into consideration as the minimum approach temperature (ΔT_{min}). In this analysis, the ΔT_{min} was varied between 5°C and 20°C. The absorber's number of stages, the temperature of the flue gas after the pre-cooler, and other important factors have all kept unchanged.

5.5 Number of stages in an absorber column

The number of stages in the absorber column was changed manually in aspen HYSYS to obtain the optimal number of stages based on the highest NPV. The default number of stages for the base case simulation is 10, while for this investigation, it was changed from 7 to 12 stages.

The packing height of one absorber stage is assumed to be one meter. The bottom stage pressure of the absorber column is 110 kPa and the top stage pressure of the absorber column is 101 kPa. It is assumed that the pressure drop in the absorber is regulated by the number of stages. As a function of the stage number in the absorption column, the pressure drop of each stage was adjusted to correspond to a 1 kPa per stage factor in the base case. For the other cases, it was adjusted according to the number of stages in the absorber column.

The lean amine flow rate was adjusted for each case by ADJ-3 to achieve the defined 85% CO_2 removal efficiency; however, the composition of the lean amine remained constant. It should be noted that like previous analyses the exhaust combustion temperature was kept constant at 1500°C, the net electricity produced was kept constant at 400 MW, the minimum approach temperature was kept constant at (ΔT_{min}) 10°C, and the inlet flue gas temperature to the absorber was kept constant at 40°C by ADJ-1, ADJ-2, ADJ-4, ADJ-5, respectively.

5.6 The inlet flue gas temperature to the absorber

In this subchapter, the temperature of the inlet flue gas to the absorber column was changed in order to achieve the highest NPV of the project. It can be done both manually and automatically.

By changing the cooling water flow rate in the inlet cooler by ADJ-5, the temperature of the inlet flue gas was adjusted. The assessment range for adjusting the inlet flue gas temperature is 30 °C to 50 °C with a 5 °C step.

In this investigation, the stages number and the composition of the lean amine remained constant with the help of make-up stream spreadsheet calculations. Additionally, ADJ-1, ADJ-2, ADJ-3, and ADJ-4 kept the other parameters constant, such as the exhaust combustion temperature, which was kept at 1500°C, the net electricity produced, which was kept at 400 MW, the CO_2 removal efficiency, which stayed at 85%, and the minimum approach temperature, which was kept at 10 °C.

6 Results

In this chapter, the cost estimation results of the base case and sensitivity analysis for the defined parameter in chapter 5 will be shown and discussed. After that, the result of a modified case based on the optimum parameter of sensitivity analysis will be presented and discussed.

6.1 Base Case results

In this section, the details of CAPEX, OPEX, and revenue from the sale of electricity generated by the combined gas power plant are discussed and then the payback period is calculated based on the net present value of the project.

6.1.1 CAPEX results

The Base Case study's equipment costs for the gas-based power plant integrated with the CO_2 capture plant are shown in Table 6.1. The equipment has a total estimated cost of 1594 million euros, and the compressor and gas turbine are clearly the most expensive equipment, making up roughly 52% and 36% of the total cost, respectively.

Other expensive items include steam turbines, absorbers, evaporators, and lean/rich amine heat exchangers, each of which contributes 4.6%, 2.3%, 1.4%, and 1.4% of the total cost, respectively.

Table 6.1: Equipment installed costs for the gas-based power plant integrated with the CO_2 capture plant

Equipment	**Installed cost [MEUR]**	**Relative CAPEX [%]**
Gas Turbine with C-C	567.26	35.58
Steam Turbine 1	53.59	3.36
Steam Turbine 2	20.40	1.28
Compressor	827.52	51.91
Water Pump	2.08	0.13
Evaporator	22.67	1.42
Condenser	6.76	0.42
Separator-1	0.89	0.06
Absorber (Shell)	18.29	1.15
Absorber (Packing)	18.03	1.13
Desorber (Shell)	3.22	0.20
Desorber (Packing)	2.10	0.13
Lean/Rich H.Ex.	22.12	1.39
Lean MEA Cooler	2.04	0.13
Reboiler-D	7.79	0.49

Inlet Cooler	2.48	0.16
Condenser-D	0.78	0.05
Fan	8.89	0.56
Lean Pump	0.96	0.06
Rich Pump	0.86	0.05
Separator-2	5.41	0.34
Total CAPEX	1594.14	100.00

6.1.2 OPEX results

The overall operational expenditure (OPEX) is approximately 109 million euros per year for the base case, as shown in Table 6.2. The two most expensive costs for this plant are maintenance and natural gas, which respectively cost 64 and 26 million euros annually. This accounts for roughly 59% and 24% of total OPEX, respectively.

Electricity consumption in the CO_2 capture plant, solvent MEA, and cooling water are the other high-cost cost items, each of which accounts for 7.3%, 4.4%, and 3.6% of the overall cost.

Table 6.2: Operational costs for the gas-based power plant integrated with the CO_2 capture plant

Utilities and maintenance	**Operation cost [MEUR/yr]**	**Relative OPEX [%]**
Electricity	7.94	7.31
Natural gas	26.15	24.07
Cooling water	3.89	3.58
Water process	0.83	0.76
Solvent MEA	4.77	4.39
Operator	0.96	0.89
Engineer	0.31	0.28
Maintenance	63.77	58.71
Total OPEX	108.62	100.00

6.1.3 Income results

The net gas-based power plant electricity generation is around 400 MW and operating hours are considered 8000 hours per year. The price of selling electricity is 0.136 [€/kWh] and as a result, the income from electricity sales is 435.877 [MEUR/year].

Table 6.3 is shown the details of the generation and consumption of electricity in the gas-based power plant equipment.

Table 6.3: Equipment electricity generation and consumption in the power plant

Equipment	Electricity generation and consumption [MW]
Gas Turbine	589.72
Steam Turbine 1	84.20
Steam Turbine 2	29.30
Compressor	300.73
Water Pump	1.862
Net Power Plant Output	400.622

6.1.4 Payback period (PBP)

Table 6.7 is shown the net present value (NPV) of the project during the construction and operation time. It is estimated that construction will take three years and the operation period is regarded as twenty-two years. In seven years of operation after construction (the ninth year in the table), the NPV of the project becomes positive, which indicates a seven-year payback period.

Table 6.4: Net present value of the project

Year	NPV (Net present value) [MEUR]
0	-531,39
1	-1025,71
2	-1485,54
3	-1222,11
4	-977,06
5	-749,11
6	-537,06
7	-339,81
8	-156,31
9	14,38
10	173,16
11	320,86
12	458,26
13	586,08

14	704,97
15	815,57
16	918,46
17	1014,16
18	1103,19
19	1186,01
20	1263,05
21	1334,72
22	1401,38
23	1463,40
24	1521,08
25	1574,75

Additionally, Figure 6.1 displays the project's payback period based on NPV for each of its years.

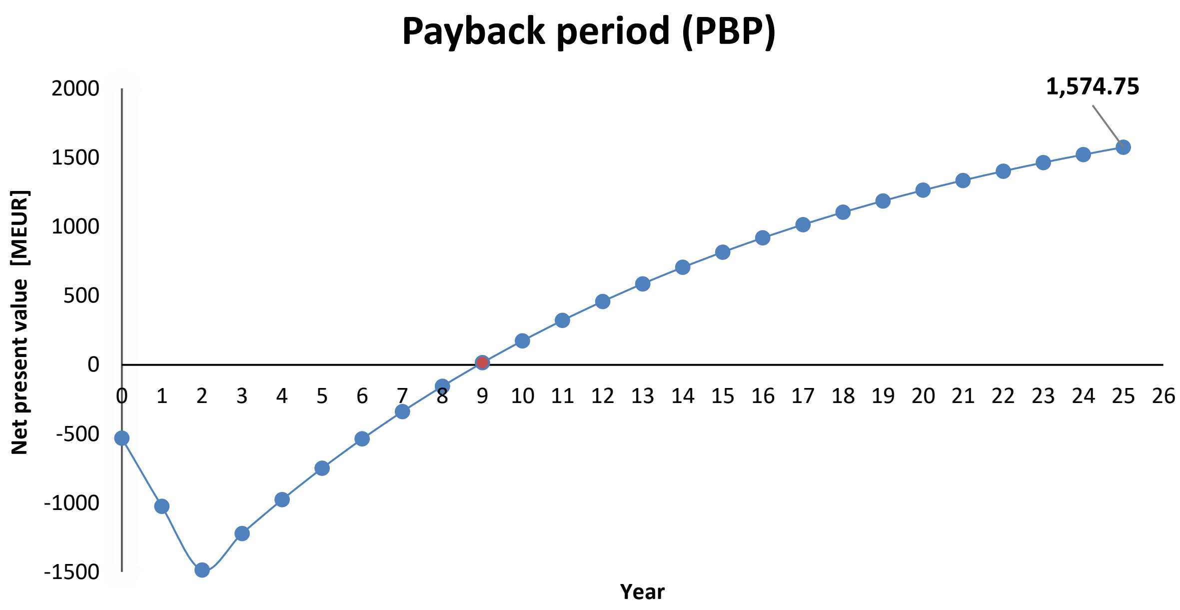

Figure 6.1: Payback period based on the Net present value of the project

6.2 Cost optimization

The results of the changing process parameters given in the sensitivity analysis that were described in the previous chapter will be discussed in this section. The method is based on the cost optimization of these process parameters. The optimum selected parameter in each case study is the one that gives the highest NPV for the gas-based power plant integrated with the CO_2 capture plant.

6.2.1 Inlet air/gas temperature into the power plant (Ambient temperature)

Figure 6.2 display the cost optimization of the inlet temperature in the power plant (ambient temperature) based on the total NPV of the project. The default inlet temperature into the power plant for the base case simulation is 25 °C and it is manually varied from 10 °C to 50 °C with a step change of 5 °C in this analysis.

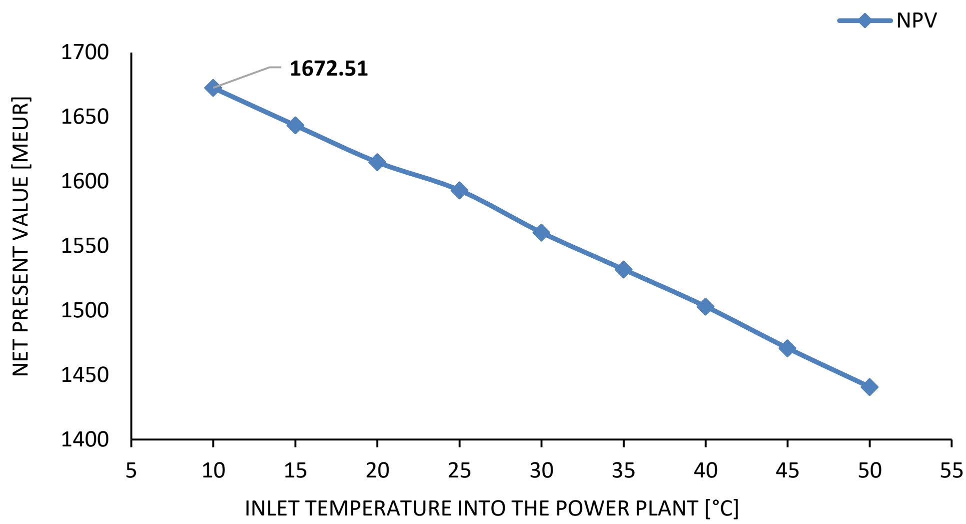

Figure 6.2: NPV calculation results of the inlet temperature in the power plant

According to the calculation results, the lower the temperature, the higher the project's net present value.

As a result of this analysis, the optimal inlet temperature is 10 °C, with a net present value of 1672 million euros. This demonstrates how well-suited this project is to the climate in Norway.

6.2.2 Inlet air/gas pressure into the power plant

The cost optimization of the power plant's inlet pressure (the pressure of input air and natural gas) based on the project's overall NPV is shown in Figure 6.3. In this analysis, the base case simulation's default inlet pressure into the power plant is 3000 kPa, and it is manually adjusted in this analysis from 1500 kPa to 4000 kPa with a step change of 500 kPa.

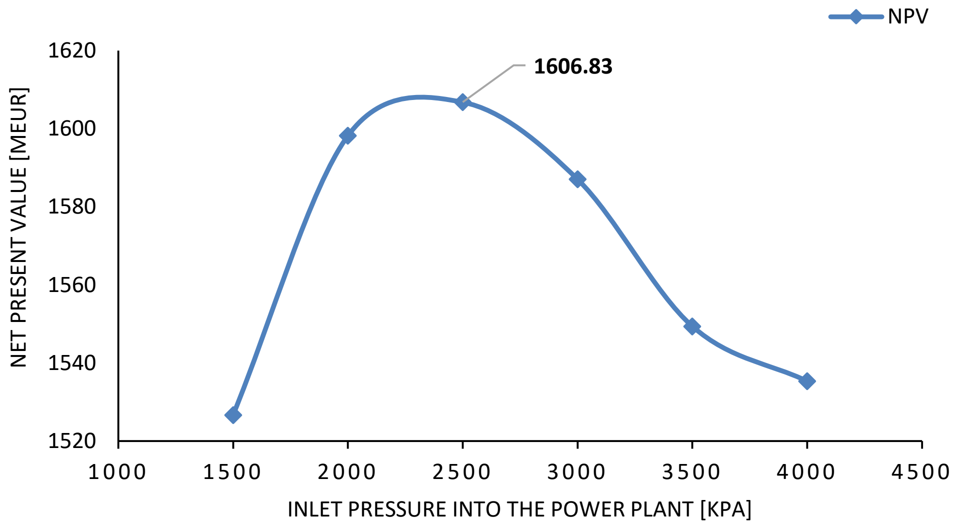

Figure 6.3: NPV calculation results of the inlet pressure into the power plant

It is estimated that the project's net present value will be approximately 1607 million euros at its optimal inlet pressure of 2500 kPa.

6.2.3 Flue gas outlet temperature (Off-gas)

Figure 6.4 depicts the flue gas outlet temperature automatic cost optimization based on the project's overall NPV. For the base case simulation, the flue gas outlet temperature from the power plant is set at 100 °C by default and in this case study, it is automatically altered from 50 °C to 150 °C with a step change of 10 °C.

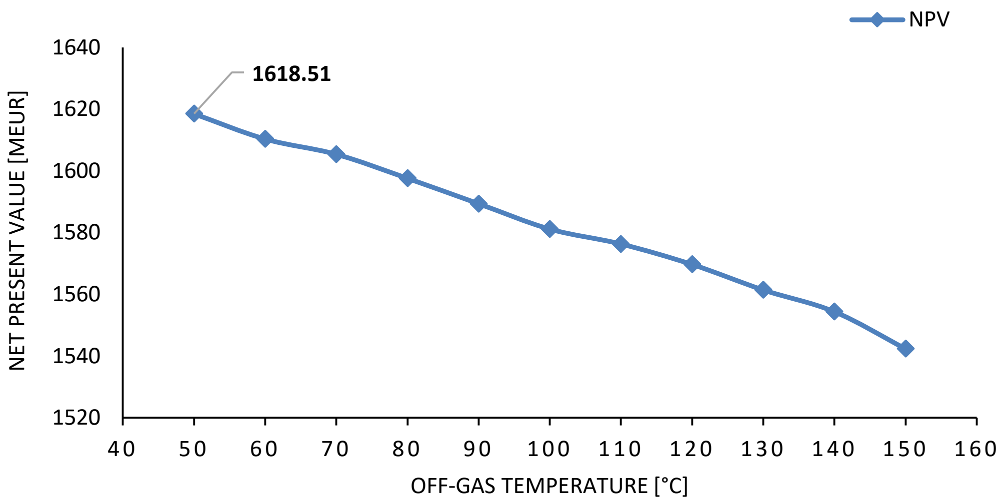

Figure 6.4: Automatic NPV results for the off-gas temperature

(Set point: T = 100 °C)

The calculation's findings show that the project's net present value increases as temperature decreases. The case study's results indicate that 50 °C is the ideal outlet temperature, with a net present value of 1618 million euros. It should be noted that the simulation does not work for temperatures lower than 50 degrees Celsius.

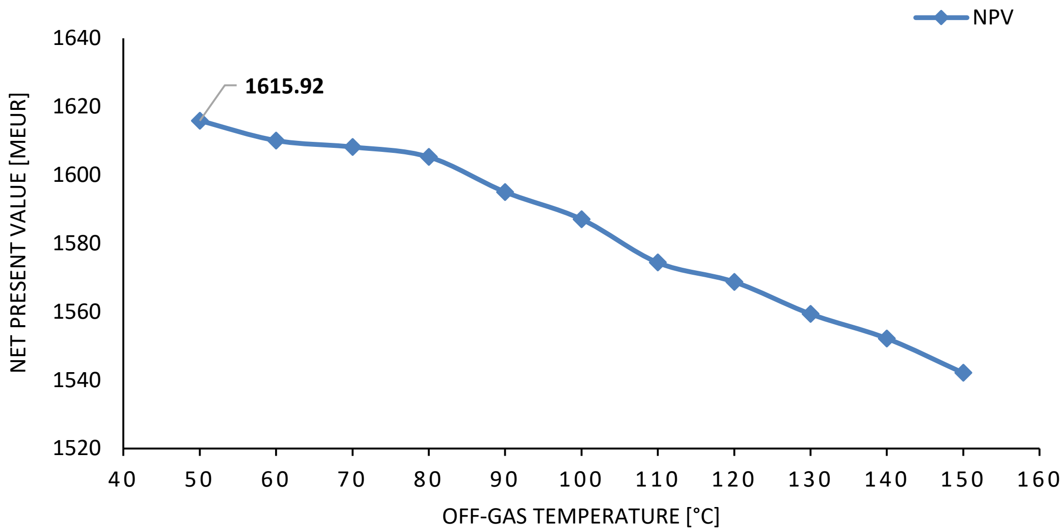

Figure 6.5: Manual NPV results for the off-gas temperature

Figure 6.5 depicts roughly the same results for the manual calculation results. It's important to note that throughout the manual calculation, just the off-gas temperature was changed; all Adjust and Recycle blocks stayed active, letting the computation for each step to be performed automatically. The optimal off-gas for the manual calculation was estimated at 50 °C with a net present value of 1616 million euros.

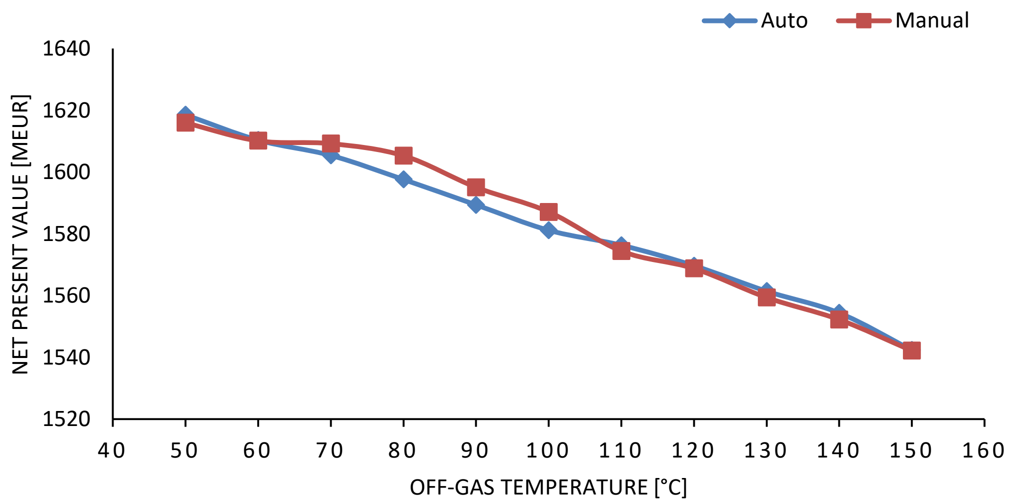

Figure 6.6: Comparison of manual and automatic NPV results for off-gas temperature

(Set point: T = 100 °C)

Figure 6.6 illustrates the difference between the computed values for the manual and automated analyses, where the automated case's starting point was 100 °C. As can be seen, the results of both calculations are almost identical.

6.2.4 Minimum approach temperature (ΔT_{min})

This section's objective is the cost optimization of the minimum approach temperature (ΔT_{min}) parameter in the lean-rich heat exchanger based on the project's highest total NPV.

Figure 6.7 illustrates the NPV results for changing the ΔT_{min} from 5 to 20 degrees Celsius, with a case study-specific set point of 5 degrees Celsius.

The result indicates that the highest computed NPV is practically consistent whereas the ΔT_{min} fluctuates in the range of 13 °C to 18 °C. This is equivalent to about 1587 million euros.

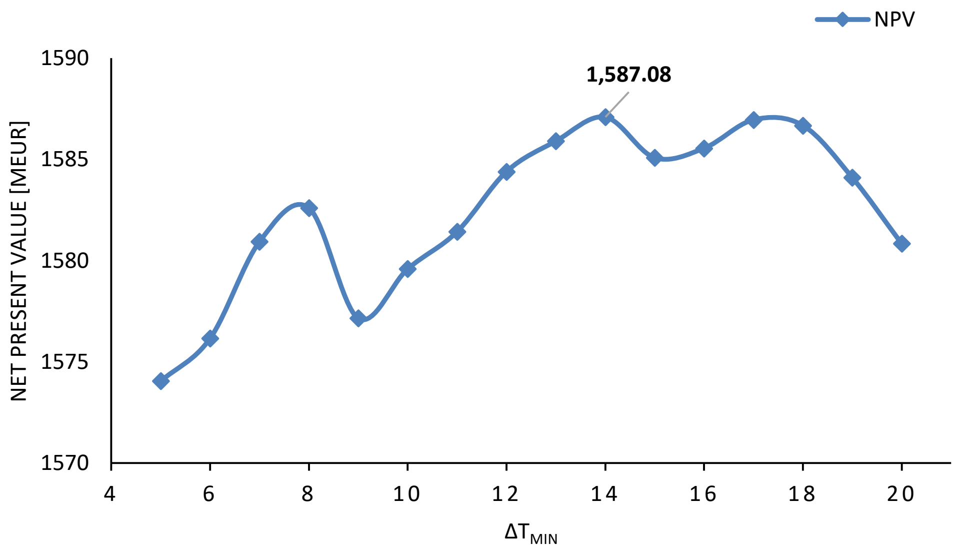

Figure 6.7: Automatic NPV calculation results as a function of the minimum temperature approach (Set Point: ΔT_{min} = 10 °C)

The results of the manual NPV calculation for altering the ΔT_{min} from 5°C to 20 °C are shown in Figure 6.8.

It should be mentioned that throughout the manual calculation method, only the target value in the ADJ-4 was adjusted; all the other Adjust and Recycle blocks continued working, letting the simulation for each stage to be performed automatically. As an outcome, the maximum NPV, with a net present value of around 1592 million euros, will be achieved for the temperature range between 13 and 18.

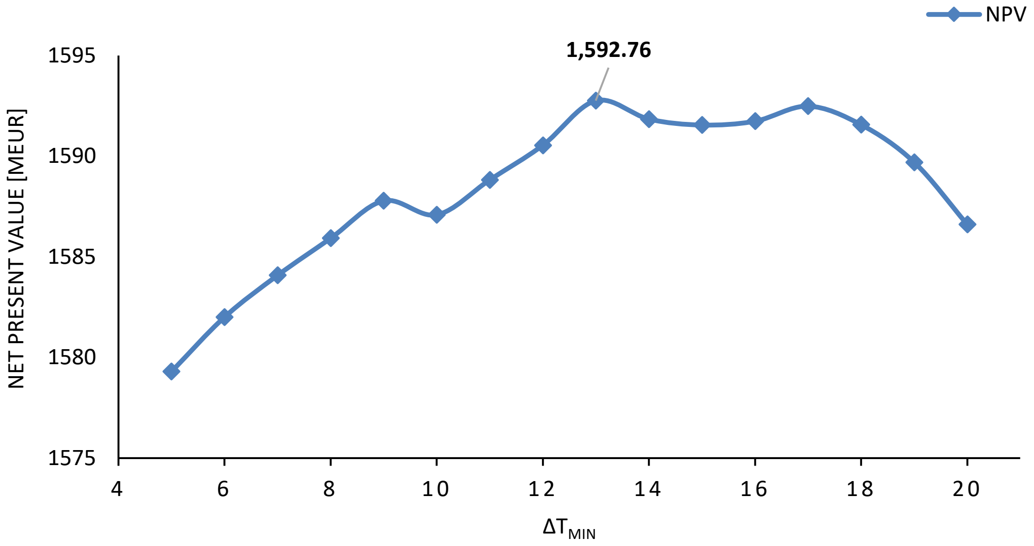

Figure 6.8: Manual NPV calculation results as a function of the minimum temperature approach

Figure 6.9 depicts a mismatch between the output for the automatic and manual methods when the set point is considered 10°C in the case study. The NPV calculated manually is around 0.4% higher than the case study scenario.

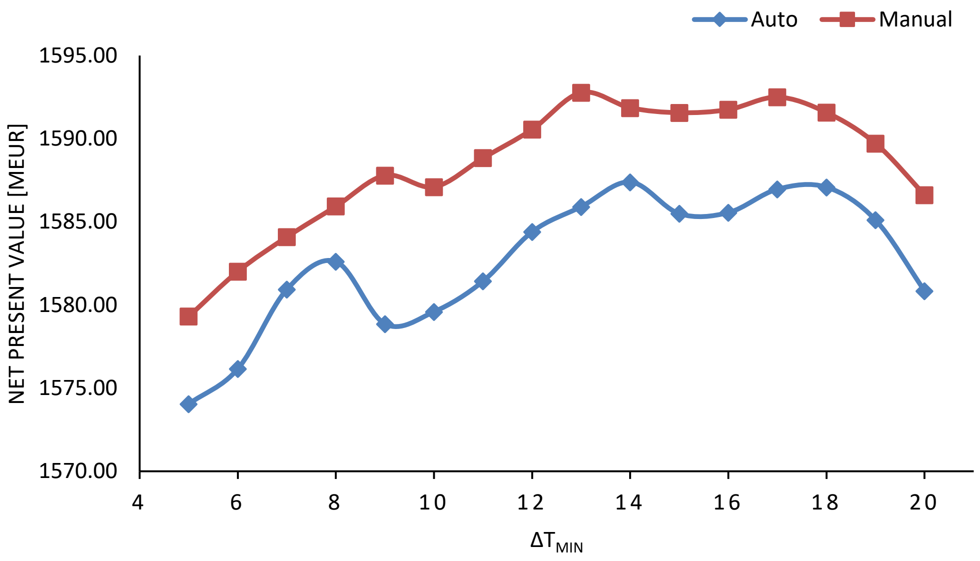

Figure 6.9: Comparison of manual and automatic NPV results for the minimum temperature approach

(Set Point: ΔT_{min} = 10 °C)

6.2.5 Number of stages in the absorption column

Based on the project's overall computed NPV, Figure 6.10 shows the cost optimization of the number of stages in the absorption column. For the base case simulation, 10 stages are considered in the absorption column by default, and in this evaluation, the number of stages was manually changed from 7 to 12.

To achieve the specified 85% CO_2 removal efficiency, the flow rate of the lean amine was updated for every scenario by ADJ-3. All other adjustments and Recycle blocks stayed active throughout the simulation to carry out calculations automatically.

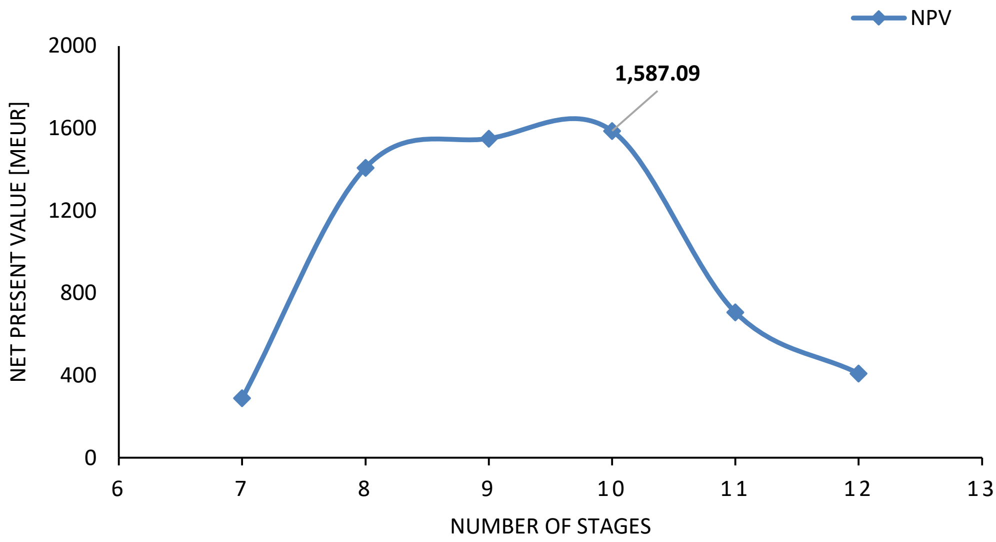

Figure 6.10: Manual NPV calculation results for the number of stages in the absorber column

The manual evaluation indicates that the 10^{th} stage of the absorber column, with a net present value of approximately 1587 million euros, yields the largest return.

6.2.6 The inlet flue gas temperature to the absorber column

This section aims to optimize the inlet flue gas temperature to the absorber column according to the project's highest total net present value.

NPV results are shown in figure 6.11 based on a five-degree Celsius change in the absorber's inlet temperature from 30 to 45 degrees Celsius with a 40-degree Celsius set point for the particular case study.

The calculation's findings show that the project's net present value decrease as temperature increase. The case study's results show that 30°C, with a net present value of 1599 million euros, is the best input flue gas temperature. It should be mentioned that simulation does not function at temperatures higher than 45 degrees Celsius in this case study evaluation. This is due to the high temperature of the flue gas, which causes no liquid in the flue gas, resulting in no flow in separator-2. Another case study simulation should be taken into consideration in order to assess the inlet flue gas temperature at temperatures higher than 45 degrees Celsius.

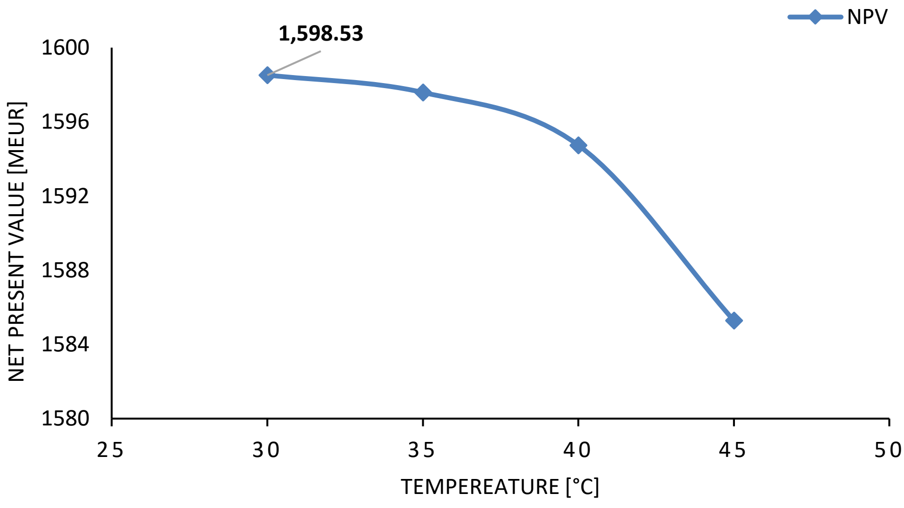

Figure 6.11: Automatic NPV calculation results for the inlet flue gas temperature to the absorber

(Set point: T = 40 °C)

The results of manual NPV calculation for inlet flue gas temperature are shown in Figure 6.12. It was found that 30 °C was the optimum inlet flue gas temperature with an estimated net present value of 1597 million euros.

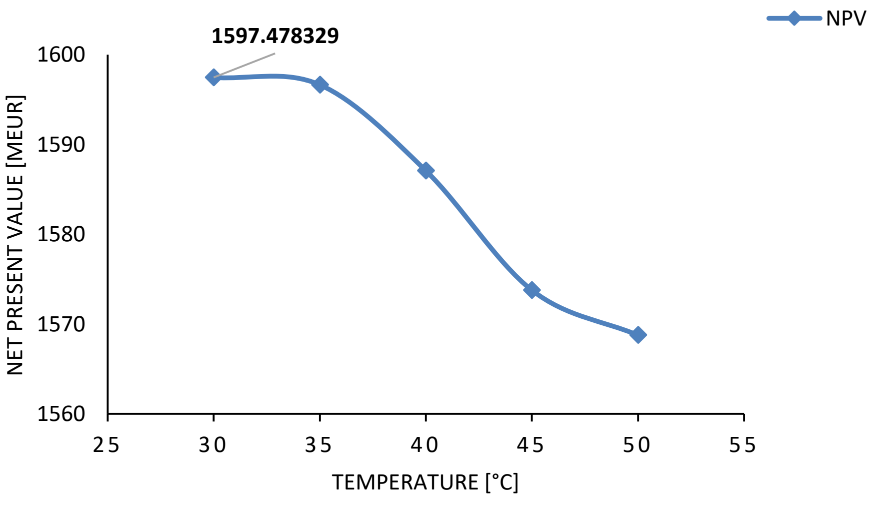

Figure 6.12: Manual NPV calculation results for the inlet flue gas temperature to the absorber

According to Figure 6.13, there is a discrepancy between the manual and automatic NPV results of the inlet flue gas temperature. A manual calculation of NPV shows a lower value of about 0.35% than the case study.

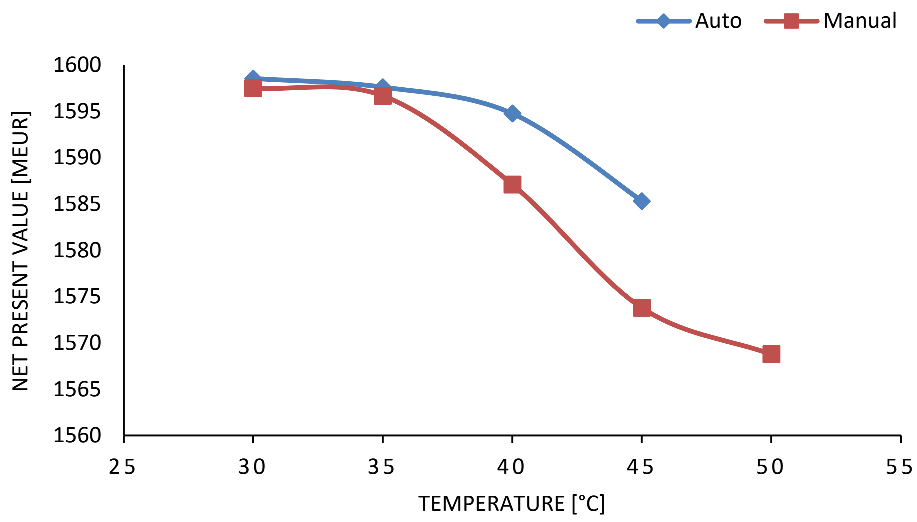

Figure 6.13: Comparison of manual and automatic NPV results for the inlet flue gas temperature to the absorber (Set Point: T = 40 °C)

Another NPV calculation is shown in Figure 6.14 based on a one-degree Celsius change in the absorber's inlet temperature from 30 to 45 degrees Celsius. It follows a pattern that is quite similar to that of Figure 6.20's findings, where the step size of the analysis in the Case Study was 5 °C. As it is clear the inlet flue gas temperatures between 30 and 35 °C provide the highest net present value with a negligible difference.

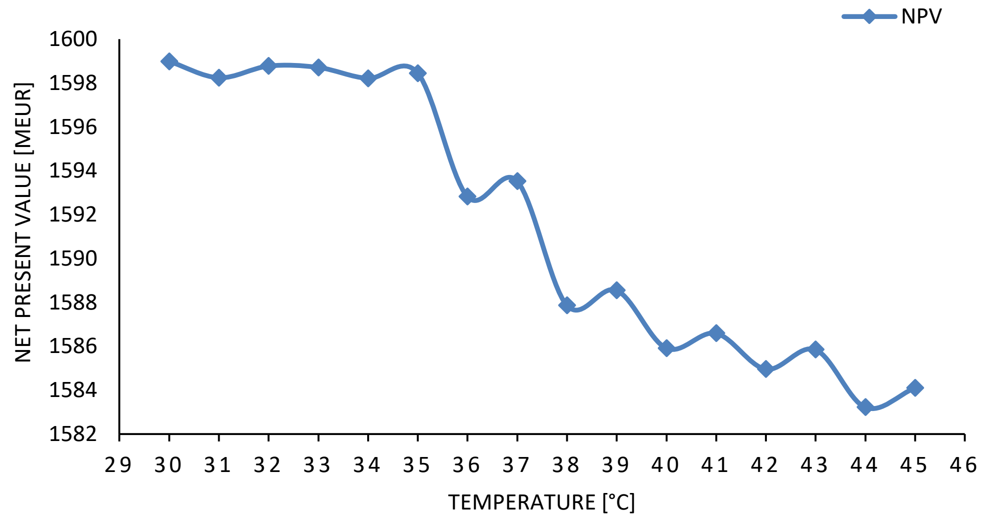

Figure 6.14: Automatic NPV calculation results for the inlet flue gas temperature to the absorber case in the case of a step size of 1 °C (Set point: T = 40 °C)

6.3 Modified base case

Based on the optimal parameters that were determined from the sensitivity analysis results, the base case was modified. The optimum parameters used in this modified base case simulation are shown in Table 6.5.

Table 6.5: Aspen HYSYS optimum parameters for modified base Case model

Modified parameter	**Value**
Inlet temperature into the power plant (Ambient temperature)	10 °C
Inlet pressure into the power plant	2500 kPa
Flue gas outlet temperature (off-gas)	50 °C
Minimum approach temperature (ΔT_{min})	13 °C
Number of stages in the absorption column	10
The inlet flue gas temperature to the absorber column	30 °C

The modified case's estimated net present value of 1855 million euros results in a profit of about 280 million euros more than the base case. It should be noted that the payback period is also shortened from 7 years to 6 years as a result of this modification. According to the project's net present value, Figure 6.15 displays the payback period of the modified case based on net present value. It is important to note that after modifications, the overall efficiency of gas-based power plants increased marginally to around 60%.

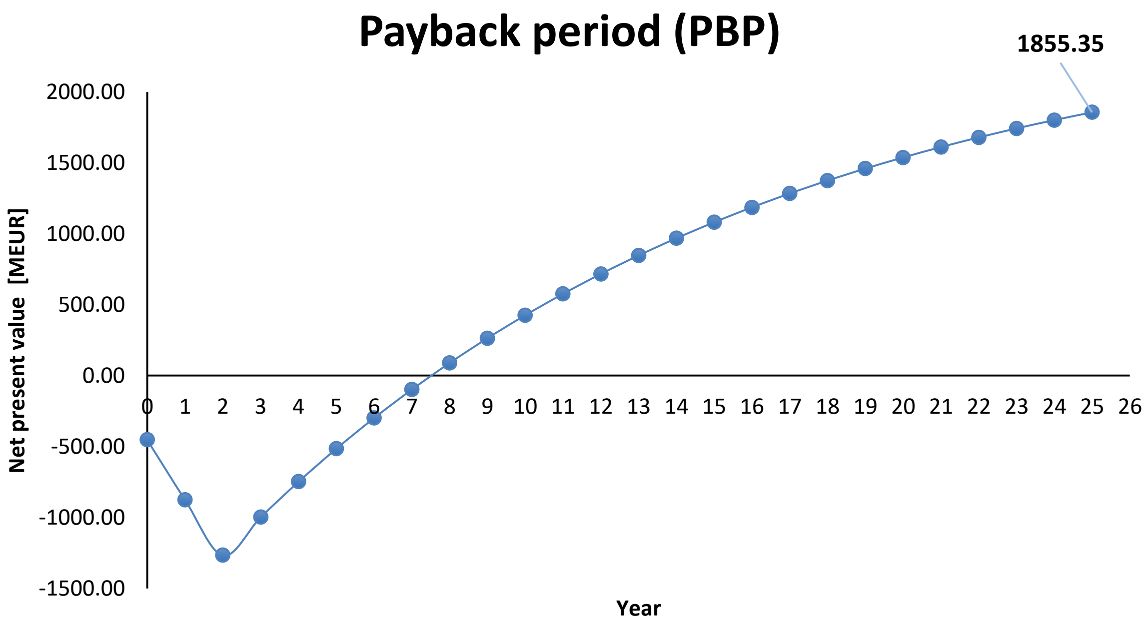

Figure 6.15: Payback period based on the Net present value of the modified base case

7 Discussion

In this section, the uncertainty and validity of the obtained results will be evaluated and then the results of the previous chapter will be compared with earlier similar works. Finally, some suggestions for future work will be presented. Considering these points will definitely improve future work.

7.1 Evaluation of uncertainty

Uncertainties in the study are mostly attributed to a number of factors, including assumptions considered for the simulation, dimensioning, cost, and income estimation:

- Cost estimation's objective is to find the optimum process variables, not to as exactly determine the absolute values as possible. The potential discrepancies in the different configurations have an impact on the uncertainty of the assessed costs. The anticipated project overall investment values, involving utility infrastructure, land, and other factors, show much greater differences. The purpose of this research does not include evaluating these modifications.

- Large uncertainties are anticipated when estimating the process equipment cost in relation to main equipment installation costs. Especially for large-cost items in the power plant, such as compressors and gas turbines, large sources of uncertainty are considered. These two items together make up 86% of the total CAPEX with 1395 million euros. Due to the high price of these items, more accuracy and attention to detail should be considered in the actual project, especially the dimensioning, equipment cost estimation, and definitely the value of the installation factor. The installation factor leads to an increase of more than two times the cost of this equipment. For more accurate results, it is advised to consult vendors and comparable components in other projects. The installed capital cost for a combined cycle gas turbine appears to be around 700 €/kW to 1200 €/kW based on websites and vendors [62].

- The price of electricity, which is used to calculate the income from the sale of electricity, is one of the main factors of uncertainty. Electricity price was assumed to be fixed based on monthly averages in 2020, while it will certainly fluctuate over the 25-year life of the project.

- The project's lifetime was also estimated to be 25 years. It was assumed that all equipment would have the same life extension. In practice, each item's working lifespan will vary depending on the type. The lifetime of each piece of equipment should therefore be investigated independently in order to get a more accurate cost calculation.

- In choosing the type of equipment in the dimensioning part, only the basic and essential factors related to the equipment are considered, while by considering other factors in the design of the equipment, a more precise option will be available and as a result, accuracy will be improved. For this purpose, it is useful to use the Aspen Icarus reference guide [44].

- A reasonable equipment cost would most likely be obtained by utilizing a cost estimation software such as Aspen In-Plan Cost Estimator. The cost estimation's accuracy, however, is influenced by the number of factors utilized as inputs. The least necessary input factors are considered in this study while estimating the cost of equipment components. The accuracy of the cost estimation will increase with the addition of the optional input factors.

- Another source of uncertainty is the scale-up factors' validity within the designated range and the applied installation factor. The scaling constant utilized for equipment was 0.65 (1.1 for absorber and desorber), and the installation factor for each piece of equipment was kept unchanged during sensitivity analysis, which added uncertainty to the cost scaling and impacted the comparison of various process parameters.

- Maintenance cost and natural gas price are the main sources of uncertainty among operating costs. The maintenance cost was considered 4% of the total CAPEX and based on the high amount of CAPEX, the major part (around 60 %) of the operating cost is included. The uncertainty in the cost estimation of the compressor and gas turbine items in CAPEX is nearly equivalent to the uncertainty in maintenance cost. The price of natural gas was assumed to be constant during the project's lifespan, but in reality, this price is expected to shift over the next years and will have an impact on the total cost estimate. Based on this reasoning, there will be uncertainty caused by the change in the price of natural gas. For more precision, it is recommended that different utility prices as well as changes in costs over one year be considered.

- The calculated cost and, hence, the optimal values may be affected by the assumptions and specifications which are selected. For instance, the type of packing selected has an impact on the cost and height of the absorption and desorption columns. The ΔT_{min} calculations are dependent on a variety of factors, such as the overall heat transfer coefficient of the lean-rich heat exchanger.

- The calculations may be impacted by the desorber's and absorber's constant Murphree efficiency simplification, which could lead to errors in the amine circulation flow and, as a result, in the estimated cost of the plant.

- Some uncertainties in the study's results are most likely related to the key output parameters that should be remained constant during the base case simulation and parametric studies. The net power output and CO_2 removal efficiency either need manual or iterative adjustments by (ADJ-1) and (ADJ-3) in the input air flow rate and in the lean amine flow rate to obtain the defined level.

- Other uncertainties in the results of the simulation are probably related to sensitivity analysis, especially for the power plant's inlet temperature and pressure, which require manual or iterative adjustments (ADJ-1 and ADJ-2) in the flow rate of the input air and compressed natural gas to reach the desired net power output. Additionally, the stages number and inlet flue gas temperature in the absorber must be adjusted manually or iteratively (ADJ-3) to achieve the desired CO_2 removal efficiency.

- This study considered only the main items in the power plant and capture plant to estimate costs. Including more items will improve the accuracy of the estimation and make it more precise.

7.2 Comparison of present work with earlier works

This work presents the base case cost estimation and cost optimization of the main parameters in a gas-based power plant integrated with amine-based CO_2 capture. Many previous studies have only considered the cost optimization of a CO_2 capture plant, and there are no existing similar works that considered both the cost optimization of a gas-based power plant and an amine-based CO_2 capture plant simultaneously.

However, among the parameters used in the sensitivity analysis for cost optimization, there are some common parameters in the cost optimization of the CO_2 capture section that can be used for comparison. For the gas-based power plant section, there are few sources to provide data on the relevant parameters discussed, it should also be noted that these data are not based on the cost optimization method.

The simulation results of the basic case scenario for the present work in the CO_2 capture plant sector are compared with those from some earlier works in Table 7.1.

Table 7.1: Comparison of base case specification with previous work in the CO_2 capture plant section

Year	Study	CO_2 Removal [%]	CO_2 Concentration (mol%)	ΔT_{min} [°C]	No. Absorber stages	E_M [%]
2007	Øi et al., [14]	85	3.75	10	10	25
2010	Kallevik [42]	85	5.9	14	15	15
2011	Amrollahi et al., [52]	90	3.8	8.5	13	-
2011	Sipöcz et al., [53]	90	4.2	10	26.9 (Height)	-
2016	Ricardez-Sandoval [65]	90	4.3	-	25	-
2018	Nwaoha et al., [54]	90	11.5	10	36	-
2019	Ali et al., [48]	90	22 - 28	10	15	11- 21

2021	Aromada et al., [36]	85	3.73	10	20	11- 21
2021	Øi et al., [41]	90	17.8	10	12	15
2022	Shirdel et al., [55]	90	7.5	10	20	15
2023	Present work	85	5.39	10	10	25

7.2.1 Inlet air/gas temperature into the power plant (Ambient temperature)

Based on the results of the present work the lower inlet air/gas temperature provides a higher net present value. The defined inlet temperature for the modified base case was considered 10 °C as a result of sensitivity analysis. It is common to consider ambient temperature as the inlet air/gas temperature in the power plant and most of the previous studies considered the ambient temperature.

Air density is influenced by ambient temperature, altitude, and humidity. Compared to dry, colder air, hot, humid air has less density. Power output in gas turbines is controlled by the mass flow through the compressor. As air density drops, higher power is needed to compress the same amount of air. As a result, the gas turbine's output power is decreased and its efficiency is also decreased [66].

Considering Carnot's efficiency, gas-based power plants will operate most efficiently at low compressor inlet temperatures and high turbine inlet temperatures (Combustion exhaust temperature) [56].

A combined cycle power plant will function more effectively and efficiently at lower ambient temperatures, according to a study by Ibrahim et al. [57].

The Brun et al. study's findings also show that the highest performance can be achieved at the coldest inlet air temperature by changing the compressor inlet temperature while maintaining the compressor inlet pressure and the turbine inlet temperature at the fixed values [56].

According to Barreto et al., the compressor's input air temperature should be around 10 and 12 °C in order to achieve maximum power output value in a power plant based on the exergetic analysis [58].

7.2.2 Inlet air/gas pressure into the power plant

According to the current study, the pressure ratio of 25 gives the highest net present value based on the cost optimization of inlet air/gas pressure in the power plant. The pressure ratio has a direct impact on the cost of compressors and gas turbines.

The most common pressure ratio in most previous studies has been around 18, but this selection was not based on overall efficiency or cost optimization as was done in the present work [17], [18], [19], [20], [21], [24], [26], [28].

Horlok [59] generally states that the optimum pressure ratio for a combined plant is equal to 18, whereas for a gas turbine operating alone, it is equal to 30. In this range, the results indicate higher overall efficacy, however, it should be emphasized that the optimization was not assessed in view of cost optimization.

Ibrahim et al [60] found that the highest total efficiency of combined cycle gas turbines takes place at a higher compression pressure ratio with low ambient temperature and high turbine inlet temperature.

According to a study by Soares [61], the optimum pressure ratio depends on the turbine inlet temperature, being equal to 12 at 1100 °K and exceeding 40 at 1800 °K.

The operating pressure ratio of the modern and cutting-edge heavy-duty gas turbines of Siemens Energy Company is 24, which is very close to the optimum pressure ratio of the current study [63].

7.2.3 Flue gas outlet temperature (Off-gas)

The default off-gas base case simulation temperature considered in the present work was 100°C, after cost optimization, the results show that a lower temperature yields a higher net present value. The lowest possible gas temperature for this simulation was 50 °C, due to the constraints related to the inlet gas temperature entering the absorber, which is defined as 40 °C.

The reason why the lowest temperature gives better results is related to the performance of the evaporator. By decreasing the temperature of off-gas from 100 °C to 50 °C, the minimum approach temperature of the evaporator decreased from 75°C to 25°C. More steam will be produced and more power will be generated in steam turbines as a result of the evaporator's performance being enhanced.

According to the study of Øi [14], the temperature of the stack after passing through the evaporator was considered to be 100 °C. In most of the relevant studies, the temperature of flue gas from HRSG is not mentioned, but in some of them, this value is considered to be around 120 °C [17], [25], [26].

The minimum temperature approach of the HRSG is the key parameter in this regard and most previous studies argued about this value. The ΔT_{min} which is considered for the water liquid-flue gas in these studied is 10 °C. [17], [19], [20].

By optimizing the minimum temperature approach of the evaporator and increasing the heat recovery the performance of steam turbines will increase and as a result, the net present value of the project will rise.

7.2.4 Minimum temperature approach in the lean-rich heat exchanger

Based on the result of manual and automatic cost optimization of ΔT_{min} in the present work, the value in the range of 13-18 degrees Celsius provides the highest net present value. In this range, the best optimal ΔT_{min} in manual and automatic evaluation is 13 °C and 14 °C, respectively. The estimated net present value varies from 13 to 18 degrees Celsius, although in all of these cases, the difference is practically minor.

The optimum minimum temperature approach will vary moderately from 10 °C to 15 °C, according to Øi et al. [41], depending on the circumstance. In most relevant evaluations, 13 °C, in particular, has been noted as the ideal scenario [39], [41], [47], [55], [64].

The energy consumption in the boiler is minimized by decreasing ΔT_{min}, and as a result of the reboiler's lower steam demand, more electricity is generated in the steam turbines, increasing the net present value. On the other hand, a drop in ΔT_{min} results in a larger heat exchanger surface area, increasing capital costs and lowering net present value. The required optimum ΔT_{min} is discovered by a cost trade-off between these two factors.

7.2.5 Number of stages in the absorption column

The lean amine flow rate required to achieve the specified CO_2 removal efficiency will differ depending on the number of steps in the absorption column. They are inversely related to one another. Increasing the number of steps leads to higher absorber packing height and decreased amine flow rate. As a result, the CAPEX linked with the absorber will increase, while the CAPEX connected with the lean/rich heat exchanger and the OPEX related to the amine flow rate will decrease, and vice versa. A balance between the above costs provides an optimum number of stages, which enhances net present value.

The default 10 stages in the absorption column provide the best net present value, according to cost optimization of the optimal number of stages in the current work. As is obvious in table 7.1, the majority of earlier studies considered a greater number of stages (packing height). It can be explained by two key specifications of these models, Murphree efficiency and CO_2 capture level of the absorption column. The mentioned previous studies took into account a higher CO_2 removal, which resulted in a higher number of stages, as well as a lower Murphree efficiency, which increases the number of stages.

Another important factor in the optimization of the number of stages is the CO_2 concentration level. A greater CO_2 concentration improves CO_2 transfer, which demands less amine flow rate and leads to a reduction in the number of stages.

7.2.6 The inlet flue gas temperature to the absorber column

The temperature of the inlet flue gas affects the performance of the inlet cooler and separator, the diameter of the absorption column, as well as the amine flow rate. Increasing the inlet temperature decreases the duty and size of the inlet cooler and separator, whereas rising the absorption column's size and the rate of amine flow. A cost trade-off between these items helps achieve a higher net present value. The amine flow rate for a specific number of stages will grow as the temperature of the inlet flue gas goes up due to decreasing CO_2 solubility at equilibrium and increasing amine evaporation tendency.

As the temperature of the flue gas entering the absorber increases, so does the actual flue gas flow rate. As a result, the size of the absorption column and the cost of the installation will increase as the vertical velocity is supposed to be unchanged [42].

Higher CO_2 solubility at equilibrium and a decreased tendency for amine evaporation are the main explanations in favour of a low temperature, whereas cheaper inlet cooler and separator expense, more reaction rate, and less viscosity are the major reasons for a high temperature [15].

According to the present study's calculated cost optimization results, inlet flue gas temperatures between 30 and 35 °C provide the highest net present value, which is followed by a sharp drop to 50 °C.

The majority of earlier research considered the inlet flue gas temperature of about 40 °C in their analyses, but they did not take into account cost calculation [14], [22], [23], [29], [53].

According to Kallevik [42], temperatures between 35 °C and 40 °C will achieve based on the net present value of the CO_2 capture plant. The Øi [39] study based on the cost estimation determined that the best temperature of the gas inlet to the absorber should be between 33 °C and 35 °C, which is lower than previously stated ranges.

7.3 Suggestion for future work

The following are some recommendations for further investigation in present field to enhance simulation and cost optimization's reliability and accuracy:

- Manual and automatic cost optimization of the turbine inlet temperatures (Combustion exhaust temperature) as a process parameter in the sensitivity analysis to obtain the optimal cost trade-off.
- Manual and automatic cost optimization of minimum temperature approach in the evaporator as a process parameter in the sensitivity analysis to achieve the optimum cost estimation.
- Increasing the pressure level steam cycle from two to three cycles (the high-pressure (HP), intermediate-pressure (IP), and low-pressure (LP) steam turbines) in order to use the steam content more effectively and thus improve the effectiveness of power generation.
- Automatic cost optimization of the inlet temperature/ pressure into the power plant. In order to do this calculation, the inlet temperature/ pressure of air and natural gas should change simultaneously. The employed approach in the present work was to improve one parameter while assuming other parameters fixed because of the regular difficulty of achieving convergence. A suggestion to determine the optimal configuration could be to perform a cost optimization for two or more process parameters to determine the least costly alternative by changing multiple parameters at once.
- Considering the Murphree efficiency correlation in the cost optimization of the inlet flue gas temperature. Depending on the temperature of the flue gas input, the temperature profile of the absorber column modifies. It is possible to adjust the Murphree efficiency for any new inlet flue gas temperature to improve analysis performance.
- Automatic cost optimization of the number of stages in the absorption column based on altering the Murphree efficiency of stages.
- Considering CO_2 transport and storage (T&S) when performing the simulation and cost estimation.
- Inclusion of carbon emission penalty and carbon price in order to optimize the techno-economic assessment.
- Taking into account the possibility of electricity and fuel price changes during the years of project implementation.

8 Conclusion

This study focuses on the process simulation and cost optimization of a gas-based power plant integrated with an amine-based CO_2 capture plant by Aspen HYSYS. The input data specification for the gas-based power plant model is based on input data from prior research on a natural gas-based power plant.

The model simulation consists of two main parts, a natural gas-based power plant, and a CO_2 capture plant. The output flue gas from the power plant after power generation in the gas turbines and heat transfer in the evaporators will directly transfer to the capture plant for CO_2 removal. It was used five adjustments and three recycle blocks in the design of the model to meet the requirements. In the base case simulation model, the purpose of ADJ-1, AD-2, ADJ-3, ADJ-4, and ADJ-5 is to set the turbine inlet temperatures to 1500 °C by altering the molar flow of input air to the compressor, the power generation of the combined cycle to 400 MW by modifying the molar flow of compressed natural gas, the CO_2 removal efficiency to 85% with changing the amine molar flow rate, the minimum temperature approach to 10 °C by adjusting the temperature of rich amine to the desorber, and the flue gas inlet to absorber temperature up to 40 °C by adapting the cooling water demand.

The CAPEX, OPEX, and revenue from selling power are used to estimate the net present value of the base case simulation. Aspen In-Plant Cost Estimator and Enhanced Detailed Factor (EDF) were used to determine the capital costs based on the dimensioning of the main equipment. The net present value method was considered as a criterion to evaluate the overall cost of the model. The base case simulation cost calculation findings show a net present value of 1574.75 million euros over a 25-year plant lifetime and a payback period of 7 years following project implementation.

Sensitivity analysis was used to perform cost optimization in order to minimize costs. When the size of equipment is altered via the sensitivity analysis, the equipment cost is modified by using the Power-Law method. In sensitivity analysis, ambient temperature, inlet air/gas pressure into the power plant, flue gas outlet temperature from the power plant (Off-gas), minimum temperature approach in the lean-rich heat exchanger, number of stages in the absorption column, inlet flue gas temperature to the absorber column were all altered to reach the project's highest net present value.

The ambient temperature sensitivity analysis shows lower temperature will result in a higher net present value. At lower temperatures, the air density will increase, causing a higher mass flow rate, which will improve power output and efficiency. In this investigation, the ambient temperature of 10°C results in a calculated NPV cost that is, on average, 5.8% higher than the default temperature. The power plant's inlet pressure was changed from 15 bar to 40 bar in the sensitivity analysis while other process parameters maintained constant and cost optimization calculation shows a 25 bar provides the highest net present value than other values. Cost estimation shows a calculated NPV result is, on average, 2% higher than the base case. According to the results, a lower off-gas temperature provides a profitable return. The off-gas temperature directly impacts the evaporator's minimum temperature approach. A lower temperature enhances the evaporator's performance, increases steam production and gas turbine power generation, and boosts project revenues. Due to limitations associated with the inlet gas temperature entering the absorber, which is set at 40 °C, the lowest allowable gas temperature for this simulation was 50 °C. Cost optimization shows a calculated NPV that is, on average, 2.7 % higher than the base case.

The sensitivity analysis for the minimum temperature approach in the lean-rich heat exchanger is a cost trade-off between the lean-rich heat exchanger area and the external utility consumption. By decreasing ΔT_{min}, the boiler's energy consumption will be minimized, and as a result of the reboiler's lower steam demand, more electricity can be produced in the steam turbines, raising the net present value. The value in the range of 13–18 degrees Celsius provides the highest net present value, according to manual and automatic cost optimization of ΔT_{min}. Cost optimization of the number of stages in the absorption column is a trade-off between the capital cost of the absorption column and the lean/rich amine heat exchanger, and the operational cost of the amine flow rate. the project's overall manual calculated NPV evaluation indicates that the 10th stage of the absorber column, yields the largest return. The temperature of the inlet flue gas has an effect on the functionality of the inlet cooler and separator, the diameter of the absorption column, and the flow rate of amine. The results of the calculated NPV cost optimization demonstrate that the highest net present value is achieved by inlet flue gas temperatures between 30 and 35 °C, with a significant decline to 50 °C occurring after.

Based on the outcomes of the sensitivity analysis's cost-optimized process parameter analysis, a base case was modified and a new model was simulated. In comparison to the initial base case scenario, the NPV computed results show an average 15 % increase in profit. This demonstrates how crucial cost optimization of the process parameters is and how it influences the project's profitability.

There are uncertainties in some areas that should be considered in future work, these considerations help to improve the model to reduce the total cost and increase the profitability of the project.

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Appendices

Appendix A - Project description

Appendix B – Submitted Abstract for SIMS 2023 conference

Appendix C – Aspen HYSYS base case PFD

Appendix D - Make up specifications

Appendix E - Base Case dimensioning

Appendix F - Detailed factor table

Appendix G - CAPEX calculation for the Base Case

Appendix H - OPEX calculation for the Base Case

Appendix I - Income of electricity sale calculation for the Base Case

Appendix J - Net present value calculation for the Base Case

Appendix A - Project Description

FMH606 Master's Thesis

Title: Process simulation and cost optimization of gas-based power plant integrated with amine-based CO_2 capture

USN supervisor: Lars Erik Øi

Task background:
Master projects from 2007 at the University of South-Eastern Norway and Telemark University College have included cost estimation in a spreadsheet connected to an Aspen HYSYS simulation. In 2007, process simulations of both a gas-based power plant and amine-based CO2 capture were simulated independently in Aspen HYSYS.

Task description:
The general aim is to develop a model in Aspen HYSYS combining a natural gas-based power plant and CO_2 capture by amine absorption. A special aim is to use this for energy and cost optimizing of the process.

1. Literature search on process simulation of natural gas-based power plants combined with absorption based CO_2 capture.

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

3. Process optimization of process parameters. Typical parameters are temperatures and pressures in the power plant and gas inlet temperature, temperature approach in the main heat exchanger and packing height in the absorption column.

4. Evaluation of limitations for the optimization, especially when using the process simulation program Aspen HYSYS.

Student category: reserved for Esmaeil Aboukazempour Amiri

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

Practical arrangements:

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

Supervision:

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

Appendix B – Submitted Abstract for SIMS 2023 conference

Process Simulation and Cost Optimization of a Gas based Power Plant including amine based CO_2 Capture

Lars Erik Øi*,[1], Esmaeil Aboukazempour Amiri[1]
[1]University of South-Eastern Norway, N-3901 Porsgrunn, Norway

lars.oi@usn.no

Keywords: Carbon capture, Power Plant, Aspen HYSYS, Simulation, Cost Estimation

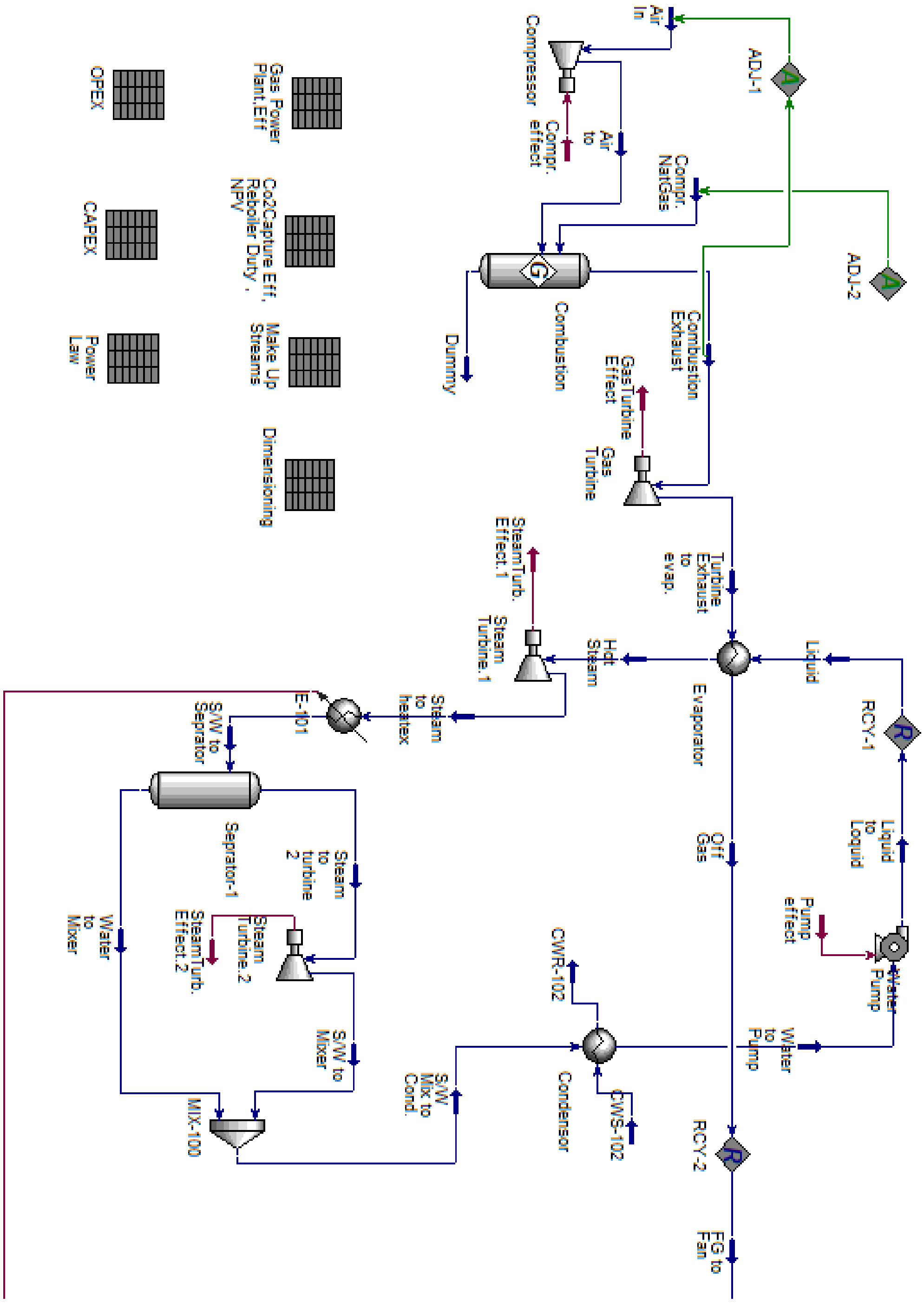
A standard process of a Gas based Power plant including CO_2 capture based on monoethanol amine (MEA) has been simulated and cost estimated with an equilibrium-based model in Aspen HYSYS™ V10.0. The gas based power plant exhaust is the input to the CO2 capture simulation, and the steam delivery from the power plant is the input to the power plant simulation. The aim has been to calculate cost optimum process parameters where both the power plant and the CO_2 capture section are involved in the cost optimization using a spreadsheet facility.

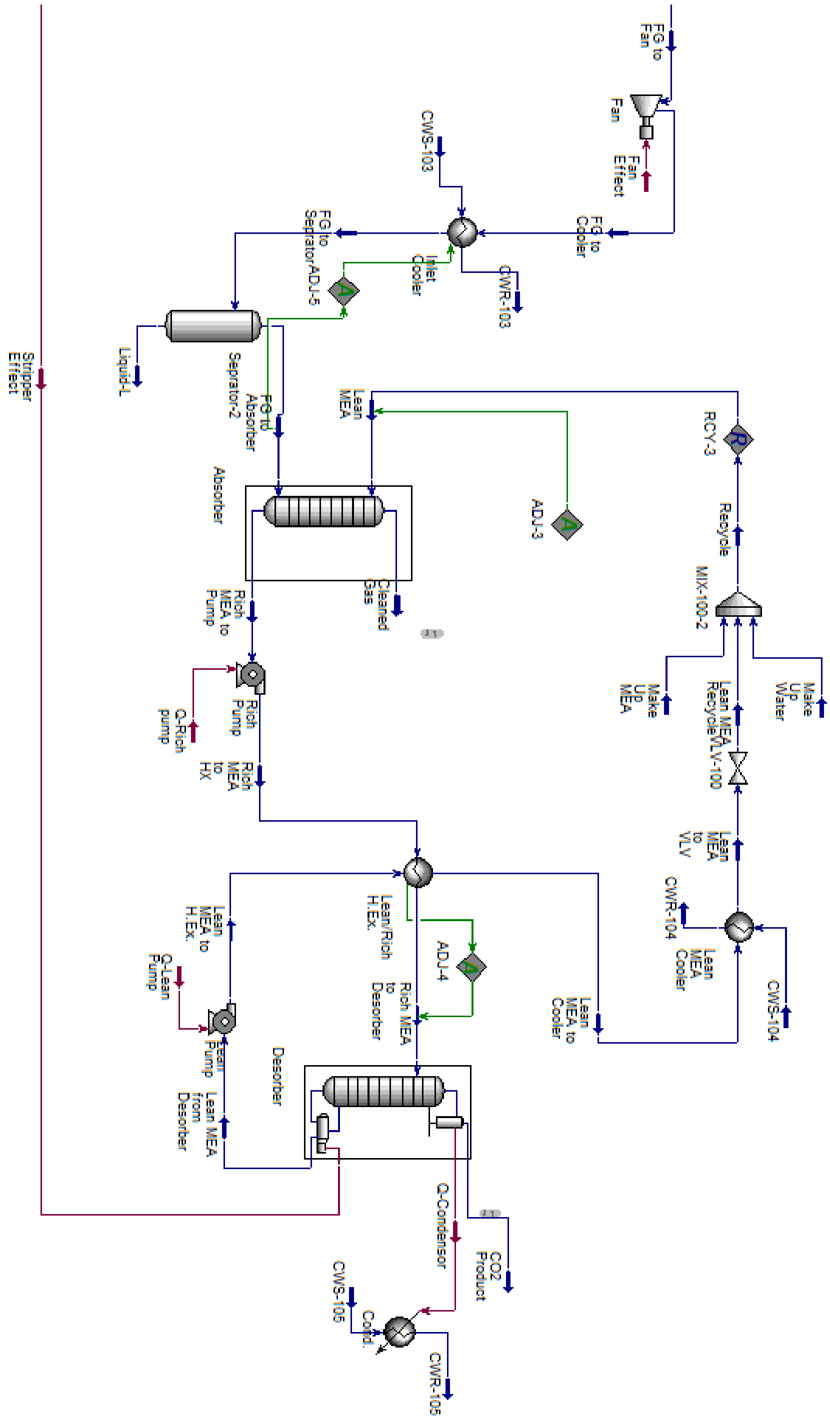
A base case with combustion 30 bar, minimum delta T 10 K, and 10 stages (meters of packing in absorber) was simulated. The power plant was calculated with a compressor, a combustion chamber, a turbine, a steam circuit with steam heater, high pressure steam turbine, low pressure steam turbine, a steam condenser and a circulating pump. The CO2 capture plant was simulated with an absorption column, a rich amine pump, a lean/rich amine heat exchanger, a desorber with reboiler and condenser, a lean pump and an amine cooler. The equipment cost was obtained from Aspen In-plant Cost Estimator™ V10.0, and an enhanced detailed factor (EDF) method was used to estimate the total investment cost.

In the parametric studies of absorber packing height and minimum approach temperature in the main heat exchanger were performed at 85 % capture efficiency. The optimizations are performed by varying one parameter at a time, minimizing the total cost calculated in the Aspen HYSYS spreadsheet. Optimized parameters are the minimum temperature in the rich amine/ lean amine heat exchanger, the number of absorption column stages, and the combustion pressure in the power plant. The calculated values for base case conditions are 13 K, 9 stages and 25 bar. Such optimization calculations including both the gas power plant and a CO_2 removal plant has not been found in literature.

The calculated cost optimum process parameters for the standard absorption and desorption process were compared to values found in literature and the combustion pressure was compared to standard pressures in industry. When both the power plant and the CO_2 capture plan are included In the optimization, the optimums can be more accurate.

Appendix C – Aspen HYSYS base case PFD




FG to Fan
Fan
Fan Effect
FG to Cooler
CWS-103
Inlet Cooler
CWR-103
FG to Separator
ADJ-5
Liquid-L
Separator-2
FG to Absorber
Lean MEA
RCY-3
Recycle
ADJ-3
Absorber
Cleaned Gas
Rich MEA to Pump
MIX-100-2
Make Up MEA
Lean MEA Recycle
VLV-100
Make Up Water
Rich Pump
Q-Rich pump
Rich MEA to HX
Lean MEA to VLV
CWR-104
Lean MEA Cooler
CWS-104
Lean/Rich H.Ex.
ADJ-4
Rich MEA to Desorber
Lean MEA to Cooler
Lean MEA to H.Ex.
Q-Lean Pump
Lean Pump
Desorber
Lean MEA from Desorber
Stripper Effect
Q-Condensor
CO2 Product
CWS-105
Cond.
CWR-105

Appendix D - Make up specifications

Make up MEA:

Lean MEA [kg/h]	680047,1
MEA loss in Cleaned Gas [kg/h]	325,8
Total MEA loss [kg/h]	325,8

Make up water:

Water Inlet in MEA Solution [kg/h]	1544394,69
Water Inlet in FG stream [kg/h]	86271,19
Water loss in Cleaned Gas [kg/h]	147061,55
Water loss in CO_2 Captured stream [kg/h]	31005,89
Total Water Loss [kg/h]	178067,44
Required Make up water [kg/h]	91796,25

Appendix E: Base Case dimensioning

Turbines:

	Gas turbine with combustion chamber	Steam turbine 1	Steam turbine 2
Power output (KW)	589716,17	84196,16	29303,11
Max power output (KW)	375000,00	22300,00	22300,00
Calculated Number of Unit	1,57	3,78	1,31
Actual Number of Unit	2,00	4,00	2,00
Power output per unit (KW)	294858,08	21049,04	14651,55

Compresor, Fan, and Pumps:

	Compressor	Water Pump	Fan	Lean Pump	Rich Pump
Duty [kW]	300729,85	1862,88	6978,03	249,92	72,28
Flow rate [m3/h]	1705624,20	420,54	2237966,26	2249,32	2168,33
Flow rate [L/s]	473784,50	116,82	621657,29	624,81	602,31
Max flow rate [m3/h]	509700,00	-	1529000,00	-	-
Calculated No. of units	3,35	-	1,46	-	-
Actual no. of units	4,00	1,00	2,00	1,00	1,00
Actual Flow rate [m3/h]	426406,05	-	1118983,13	-	-
Actual Duty [kW]	75182,46	-	3489,01	-	-

Heat Exchangers:

	Lean/Rich H.Ex.	Lean MEA Cooler	Inlet Cooler
Q [kJ/h]	535704681,14	106752207,68	236564059,61
Heat transfer coefficient [kw/m2K]	0,73	0,80	0,80
LMTD	13,33	27,03	49,47
Total Heat Transfer Area [m2]	15245,26	1371,35	1660,38
Max. Area per of Unit [m2]	1000,00	1000,00	1000,00
Calculated Number of Unit	15,25	1,37	1,66
Actual Number of Unit	16,00	2,00	2,00
Actual Area per unit [m2]	952,83	685,68	830,19

Columns:

	Absorber	Desorber
FG volume flow [m3/h]	1773211,74	108455,76
FG volume flow [m3/s]	492,56	30,13
FG velocity [m/s]	2,50	1,00
Inner Diameter [m]	15,84	6,19
Number of stages	10,00	6,00
Height of each stage [m]	1,00	1,00
Packing height [m]	10,00	6,00
Column height [m]	25,00	15,00
Column volume [m3]	4925,59	451,90
Packing volume [m3]	1970,24	180,76
Number of units	3,00	1,00
Volume per unit [m3]	1641,86	451,90
Packing volume per unit [m3]	656,75	180,76
Diameter per unit [m]	9,14	6,19
SHELL MAT.	SS316	SS316
PACKING TYPE	MellaPak 250Y	MellaPak 250Y
No. PACKED SECTIONS	3,00	2,00

Reboiler:

	Reboiler-D
Q [kJ/h]	447016230,81
Heat transfer coefficient [kw/m2K]	1,20
LMTD	21,35
Total Heat Transfer Area [m2]	4847,60
T(out,cold)	120,00
T(in,cold)	114,77
T(in,Hot)	138,84
T(out,Hot)	138,84
Max. Area per of Unit [m2]	1000,00
Calculated Number of Unit	4,85
Actual Number of Unit	5,00
Actual Area per unit [m2]	969,52

Condensers:

	Condenser	Condenser-D
CWS [m3/h]	13014,93	1312,09
CWS [L/S]	3615,26	364,47
Max CWS (L/S)	315,00	315,00
Actual Number of units	12,00	2,00
Actual CWS (L/S)	301,27	182,23

Evaporator:

	Evaporator
Q [kJ/h]	1418604569
Heat transfer coefficient [kw/m2K]	0,25
LMTD	111,6523415
Total Heat Transfer Area [m2]	14117,27938
Max. Area per of Unit [m2]	4640
Calculated Number of Unit	3,042517107
Actual Number of Unit	4
Actual Area per unit [m2]	3529,319844

Separators:

	Separator-1	Separator-2
Actual Gas Flow Rate [m3/h]	113224,47	1686469,10
Actual Gas Flow Rate [m3/s]	31,45	468,46
Liquid Phase Mass Density (dL) [kg/m3]	913,99	995,99
Gas Phase Mass density [kg/m3]	1,88	1,21
K Factor, Sounder-Brown Velocity	0,11	0,15
Allowable Vapour Velocity [m/s]	2,35	4,31
Vessel Cross-Sectional Area [m2]	13,36	108,79
Vessel Inner-Diameter (Di) [m]	4,12	11,77
Vessel Height , 1D, [m]	4,12	11,77
Vessel Volume [m3]	55,10	1280,34

Appendix F - Detailed factor table:

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 G - CAPEX calculation for the Base Case

Turbines:

		Gas Turbine with C-C	Steam Turbine 1	Steam Turbine 2
Power output (KW)		589716,17	84196,16	29303,11
Max power output (KW)		375000,00	22300,00	22300,00
Calculated Number of Unit		1,57	3,78	1,31
Actual Number of Unit		2,00	4,00	2,00
Power output per unit (KW)		294858,08	21049,04	14651,55
Material		CS	CS	CS
Equipment cost per unit [kEUR]	Aspen In Plant 2019	113490,00	4473,50	3405,80
Equipment cost (Total) [EUR]	One unit	-	-	-
Equipmen Cost per unit [kEUR] SS	Convert to 2020	115654,66	4558,83	3470,76
Equipmen Cost per unit [kEUR] CS	Convert to CS	-	-	-
Installation factor CS 2020 Total		2,37	2,84	2,84
Installation factor SS 2020 total		-	-	-
Total Equipment cost 2020 [kEUR]		274101,54	12947,06	9856,96
equipment cost 2021 [kEUR]	Convert to 2021	283629,14	13397,10	10199,58
Total equipment cost 2021 [kEUR]		567258,27	26794,19	20399,16

Compressor & Water pump:

		Compressor	Water Pump
Duty [kW]		300729,85	1862,88
Flow rate [m3/h]		1705624,20	420,54
Flow rate [L/s]		473784,50	116,82
Max flow rate [m3/h]		509700,00	-
Calculated No. of units		3,35	-
Actual no. of units		4,00	1,00
Actual Flow rate [m3/h]		426406,05	-
Actual Duty [kW]		75182,46	1862,88
Material		SS316	SS316
Equipment cost per unit [kEUR]	Aspen In Plant 2019	91843,00	383,40
Equipmen Cost per unit [kEUR] SS	Convert to 2020	93594,77	390,71
Equipmen Cost per unit [kEUR] CS	Convert to CS	71995,98	300,55
Installation factor CS 2020 Total		2,42	4,92
Installation factor SS 2020 total		2,78	6,70
Total Equipment cost 2020 [kEUR]		199932,84	2012,47
equipment cost 2021 [kEUR]	Convert to 2021	206882,37	2082,42
Total equipment cost 2021 [kEUR]		827529,49	2082,42

Evaporator & Condenser:

		Evaporator		Condenser
Total Heat Transfer Area [m2]		14117,28	CWS [m3/h]	13014,93
Max. Area per of Unit [m2]		4640,00	CWS [L/S]	3615,26
Calculated Number of Unit		3,04	Max CWS (L/S)	315,00
Actual Number of Unit		4,00	Actual Number of units	12,00
Actual Area per unit [m2]		3529,32	Actual CWS (L/S)	301,27
Material		SS316		CS
Equipmen Cost per unit [kEUR]	Aspen In Plant 2019	2016,00		74,00
Equipment cost (Total) [EUR]	One unit	-		-
Equipmen Cost per unit [kEUR] SS	Convert to 2020	2054,45		75,41
Equipmen Cost per unit [kEUR] CS	Convert to CS	1173,97		75,41
Installation factor CS 2020 Total		3,63		7,22
Installation factor SS 2020 total		4,67	-	-
Equipment cost 2020 [kEUR]		5476,58		544,47
Equipment cost 2021 [kEUR]	Convert to 2021	5666,95		563,40
Total Equipment cost 2021 [kEUR]		22667,78		6760,75

Separator-1:

		Separator-1
Vessel Inner-Diameter (Di) [m]		4,124411356
Vessel Height, 3D [m]		4,124411356
Vessel Volume [m3]		55,10307094
Actual no. of units		1
Actual Volume [m3]		55,10307094
Material		SS316
Equipment cost per unit [kEUR]	Aspen In Plant 2019	205,5
Equipment cost per unit [kEUR]	Convert to 2020	209,4196185
Equipmen Cost per unit [kEUR] CS	Convert to CS	119,6683534
Installation factor CS 2020 Total		5,89
Installation factor SS 2020 total		7,21
Equipment cost 2020 [kEUR]		862,8088283
Equipment cost 2021 [kEUR]	Convert to 2021	892,7995095
Total equipment cost 2021 [kEUR]		892,7995095

Columns:

		Absorber	Desorber
Packing height [m]		10,00	6,00
Column height [m]		25,00	15,00
Column volume [m3]		4925,59	451,90
Packing volume [m3]		1970,24	180,76
Diameter [m]		15,84	6,19
Number of units		3,00	1,00
Volume per unit [m3]		1641,86	451,90
Packing volume per unit [m3]		656,75	180,76
Diameter per unit [m]		9,14	6,19
SHELL MAT.		SS316	SS316
No. Packing section		3,00	2,00
Equipmen Cost per unit [kEUR]	Aspen In Plant 2019	4563,50	1668,40
Equipment cost (Total) [kEUR]	One unit	-	-
Packing cost per unit [kEUR]	Aspen In Plant 2019	2394,66	659,06
Total packing cost [kEUR]	One unit	-	-
Total Shell Volume [m3]		2955,35	271,14
Shell volume per unit [m3]		985,12	
Total Shell cost [kEUR]	One unit	-	-
Shell cost per unit [kEUR]	Aspen In Plant 2019	2168,84	1009,34
Equipmen Cost per unit [kEUR] SS	Convert to 2020	4650,54	1700,22
Shell cost per unit [kEUR]	Convert to 2020	2210,20	1028,59
Packing cost per unit [kEUR] SS	Convert to 2020	2440,34	671,63
Equipmen Cost per unit [kEUR] CS	Convert to CS	2657,45	971,56
Shell cost per unit [kEUR] CS	Convert to CS	1262,97	587,77
Packing cost per unit [kEUR] CS	Convert to CS	1394,48	383,79
Installation factor CS 2020 Total		2,84	3,63
Installation factor CS 2020 Shell		3,63	4,19
Installation factor CS 2020 Packing		3,19	4,19
Installation factor SS 2020 total		3,76	4,67
Installation factor SS 2020 Shell		4,67	5,30
Installation factor SS 2020 Packing		4,17	5,30
Shell cost 2020 [kEUR]		5891,77	3115,16
Packing cost 2020 [kEUR]		5808,01	2034,08
Shell cost 2021 [kEUR]	Convert to 2021	6096,57	3223,44
Packing cost 2021 [kEUR]	Convert to 2021	6009,89	2104,79
Total Shell cost 2021 [kEUR]		18289,70	3223,44
Total Packing cost 2021 [kEUR]		18029,67	2104,79
Total Shell & Packing cost 2021 [kEUR]		36319,36	5328,23

Heat Exchangers:

		Lean/Rich H.Ex.	Lean MEA Cooler	Reboiler-D	Inlet Cooler
Total Heat Transfer Area [m2]		15245,26	1371,35	4847,60	1660,38
Max. Area per of Unit [m2]		1000,00	1000,00	1000,00	1000,00
Calculated Number of Unit		15,25	1,37	4,85	1,66
Actual Number of Unit		16,00	2,00	5,00	2,00
Actual Area per unit [m2]		952,83	685,68	969,52	830,19
Material		SS316	SS316	SS316	SS316
Equipmen Cost per unit [kEUR]	Aspen In Plant 2019	374,90	276,60	422,60	336,80
Equipment cost (Total) [EUR]	One unit				
Equipmen Cost per unit [kEUR] SS	Convert to 2020	382,05	281,88	430,66	343,22
Equipmen Cost per unit [kEUR] CS	Convert to CS	218,31	161,07	246,09	196,13
Installation factor CS 2020 Total		4,92	4,92	4,92	4,92
Installation factor SS 2020 total		6,12	6,12	6,12	6,12
Total Equipment cost 2020 [kEUR]		1336,09	985,76	1506,08	1200,30
equipment cost 2021 [kEUR]	Convert to 2021	1382,53	1020,02	1558,43	1242,03
Total equipment cost 2021 [kEUR]		22120,44	2040,05	7792,16	2484,05

Condenser:

	Condenser-D
CWS [m3/h]	1312,09
CWS [L/S]	364,47
Max CWS (L/S)	315,00
Actual Number of units	2,00
Actual CWS (L/S)	182,23
Material	CS
Equipmen Cost per unit [kEUR]	51,40
Equipment cost (Total) [EUR]	
Equipmen Cost per unit [kEUR] SS	52,38
Equipmen Cost per unit [kEUR] CS	52,38
Installation factor CS 2020 Total	7,22
Installation factor SS 2020 total	-
Total Equipment cost 2020 [kEUR]	378,19
equipment cost 2021 [kEUR]	391,33
Total equipment cost 2021 [kEUR]	782,66

Fan & Pumps:

		Fan	Lean Pump	Rich Pump
Duty [kW]		6978,03	249,92	72,28
Flow rate [m3/h]		2237966,26	2249,32	2168,33
Flow rate [L/s]		621657,29	624,81	602,31
Max flow rate [m3/h]		1529000,00	-	-
Calculated No. of units		1,46	-	-
Actual no. of units		2,00	1,00	1,00
Actual Flow rate [m3/h]		1118983,13	-	-
Actual Duty [kW]		3489,01	249,92	72,28
Material		CS	SS316	SS316
Equipment cost per unit [kEUR]	Aspen In Plant 2019	1321,00	220,00	165,60
Equipment cost (Total) [EUR]	One unit	-	-	-
Equipmen Cost per unit [kEUR] SS	Convert to 2020	1346,20	224,20	168,76
Equipmen Cost per unit [kEUR] CS	Convert to CS		172,46	129,81
Installation factor CS 2020 Total		3,19	4,92	5,89
Installation factor SS 2020 total		-	5,40	6,42
Total Equipment cost 2020 [kEUR]		4294,37	931,28	833,15
equipment cost 2021 [kEUR]	Convert to 2021	4443,64	963,65	862,11
Total equipment cost 2021 [kEUR]		8887,27	963,65	862,11

Separator-2:

		Separator-2
Vessel Inner-Diameter (Di) [m]		11,77
Vessel Height, 3D [m]		11,77
Vessel Volume [m3]		1280,34
Actual no. of units		1,00
Actual Volume [m3]		1280,34
Material		SS316
Equipment cost per unit [kEUR]	Aspen In Plant 2019	1925,00
Equipment cost per unit [kEUR]	Convert to 2020	1961,72
Equipmen Cost per unit [kEUR] CS	Convert to CS	1120,98
Installation factor CS 2020 Total		3,63
Installation factor SS 2020 total		4,67
Total Equipment cost 2020 [kEUR]		5229,38
equipment cost 2021 [kEUR]	Convert to 2021	5411,15
Total equipment cost 2021 [kEUR]		5411,15

Total CAPEX:

Total CAPEX [kEUR]	1594170,19

Appendix H - OPEX calculation for the Base Case

Plant life time [year]	25
Construction time [year]	3
Operation time [year]	22
Operation time [h/year]	8000

Electricity:

Electricity price [EUR/kWh]	0,136	
	Consumption [kW]	Price [kEUR/year]
Fan	6978,03	7592,10
RA Pump	249,92	271,92
LA Pump	72,28	78,64
Total Price		7942,65

Natural Gas:

Natural Gas Price [EUR/m3]	1,29	
	Consumption	Price [kEUR/year]
Combustion Chamber [m3/h]	2533,78	26148,65

Cooling water:

Cooling Water Price [EUR/m3]	0,023	
	Consumption [m3]	Price [kEUR/year]
Lean MEA Cooler	2438,75	427,87
Condenser-D	1312,09	230,20
Inlet Cooler	5404,30	948,18
Condenser	13014,93	2283,45
Total Price		3889,70

Process water:

Process Water price [EUR/m3]	0,20	
	Consumption [m3]	Price [kEUR/year]
Water in MEA solution	1547,37	0,31
Make Up water	92,38	150,05
Total Price		150,36
	Consumption [m3]	Price [kEUR/year]
Liquid (Water)	419,39	681,09
Total Price		831,45

MEA:

MEA price [EUR/ton]	1450,00	
	Consumption	Price [kEUR/year]
MEA in MEA solution	680047,10	986,07
Make Up MEA	325,80	3779,31
Total price		4765,38

Operator:

Operator price [EUR/year]	80414,00	
	Number of Operator	Price [kEUR/year]
Operator	12,00	964,97

Engineer:

Engineer price [EUR/year]	156650	
	number of Engineer	Price [kEUR/year]
Engineer	2	313,3

Maintenance:

	4% of CAPEX	Price [kEUR/year]
Maintenance		63766,81

Total OPEX:

TOTAL OPEX [kEUR/year]	108622,92

Appendix I – Income of electricity sale calculation

Net power generation:

Compressor	300729,85
Pump	1862,88
Steam Turbine 1	84196,16
Steam Turbine 2	29303,11
Gas Turbine	589716,17
Power Plant Output	400622,71

Electricity:

Electricity price [EUR/kWh]	0,136	
	Generation [KW]	Price [kEUR/year]
Gas-based power Plant	400622,71	435877,51

Gas-based power plant efficiency:

LHV (NG) (KJ/mol)	755,78
Molar Flow (Kgmole/h)	3271,68
Power Plant Output (kw)	400622,71
Power Plant Output (MJ/h)	1442241,75
Natural Gas Heat Flow (MJ/h)	2472664,47
Power Plant Efficiency (%)	58,33

Appendix J - Net present value calculation

Year	Cost [KEUR]	Income [KEUR]	Nondiscounted Net Cash Flow [KEUR]	Discount factor	Discounted Net Cash Flow [KEUR]	NPV [KEUR]
0,00	531390,01	0,00	-531390,01	1,00	-531390,01	-531390,01
1,00	531390,01	0,00	-531390,01	0,93	-494316,29	-1025706,29
2,00	531390,01	0,00	-531390,01	0,87	-459829,10	-1485535,40
3,00	108622,92	435877,51	327254,59	0,80	263427,04	-1222108,35
4,00	108622,92	435877,51	327254,59	0,75	245048,41	-977059,94
5,00	108622,92	435877,51	327254,59	0,70	227952,01	-749107,93
6,00	108622,92	435877,51	327254,59	0,65	212048,38	-537059,55
7,00	108622,92	435877,51	327254,59	0,60	197254,31	-339805,24
8,00	108622,92	435877,51	327254,59	0,56	183492,38	-156312,86
9,00	108622,92	435877,51	327254,59	0,52	170690,59	14377,73
10,00	108622,92	435877,51	327254,59	0,49	158781,94	173159,67
11,00	108622,92	435877,51	327254,59	0,45	147704,13	320863,80
12,00	108622,92	435877,51	327254,59	0,42	137399,19	458262,99
13,00	108622,92	435877,51	327254,59	0,39	127813,20	586076,19
14,00	108622,92	435877,51	327254,59	0,36	118896,00	704972,20
15,00	108622,92	435877,51	327254,59	0,34	110600,93	815573,13
16,00	108622,92	435877,51	327254,59	0,31	102884,59	918457,71
17,00	108622,92	435877,51	327254,59	0,29	95706,59	1014164,31
18,00	108622,92	435877,51	327254,59	0,27	89029,39	1103193,70
19,00	108622,92	435877,51	327254,59	0,25	82818,04	1186011,73
20,00	108622,92	435877,51	327254,59	0,24	77040,03	1263051,77
21,00	108622,92	435877,51	327254,59	0,22	71665,15	1334716,91
22,00	108622,92	435877,51	327254,59	0,20	66665,25	1401382,17
23,00	108622,92	435877,51	327254,59	0,19	62014,19	1463396,36
24,00	108622,92	435877,51	327254,59	0,18	57687,62	1521083,98
25,00	108622,92	435877,51	327254,59	0,16	53662,90	1574746,88