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Process Simulation, Dimensioning and Automated Cost Optimization of CO_2 Capture

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Summary:

The use of an amine solvent to remove CO_2 from the exhaust gas is an established and well-studied technology. Available emission data from prior research on a natural gas-based power plant project at Mongstad, Norway, was used to model a typical CO_2 capture process in Aspen HYSYS. To undertake cost optimization, a Base Case was created with 15 m absorber packing height, 10 m desorber packing height, the removal efficiency of 85 %, and a minimum temperature approach (ΔT_{min}) in the lean/rich amine heat exchanger of 10 °C. To estimate and quantify the total cost for the basic scenario, the Enhanced Detailed Factor (EDF) was employed in combination with the Aspen In-Plant Cost estimator. The Base Case results showed a total cost of 42.9 EUR per ton of CO_2 removed and the reboiler energy use of 3750 kJ/kg.

In the sensitivity analysis, the absorber packing height, the minimum temperature approach (ΔT_{min}), and the entering flue gas temperature into the absorber column were all altered to find out how different factors affected pricing variations. When the sensitivity analysis changed the size, the Power-Law approach was applied to vary the equipment cost. Using the Adjust and Recycle blocks, as well as switching the calculated values between the simulation and spreadsheets, makes the analysis more automated.

When the ΔT_{min} was changed from 5 °C to 20 °C, in both automatic and manual scenarios, the variation in predicted cost from 11 °C to 15 °C was minimal.

Since changing the number of stages in an automated assessment is not possible, the stage's efficiency was changed from 0.15 to 0.9, which is equivalent to increasing the number of steps from 13 to 18. The optimum calculated packing height was 15 m, with a CO_2 collection cost of 42.6 EUR/t in the manual analysis. The automated calculated costs were on average 1.5 % and 0.9 % higher than the manual technique when the target stages for changing efficiency were the 13th stage, and the 10th stage, respectively.

A 15-stage absorber was employed to automatically assess the change in incoming flue gas temperature to the absorber from 30 to 50 degrees Celsius in 5 °C steps. The computed captured cost was around 2% lower than the Base Case research due to employing lower amine flow rate by enhancing average stages' efficiency. Similar research for a simulated case with a 13-stage absorber resulted in a cost reduction of more than 4% compared to the Base Case. When the step size was lowered to 1 °C, the best input temperature was determined to be 34 °C, with an estimated cost of 39.6 EUR per ton of CO_2 captured.

The major goal was to use the Aspen HYSYS software to automatically calculate and optimize the cost of an MEA-based CO_2 capture facility. This study states automated optimization of absorber packing height and gas inlet temperature using the Case Study tool in Aspen HYSYS, which has not been done before.

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 spring of 2020.

I want to take this opportunity to thank all the folks who assisted me in finishing my thesis.

I would like to thank my supervisor, Lars Erik Øi, a Professor at the University of South-Eastern Norway, for his guidance and assistance throughout my thesis composition.

I want to also express my gratitude to my family and friends for their encouragement and support throughout this time. Without their help, this thesis would not have been feasible.

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Shirvan Shirdel

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Nomenclature

Abbreviation	Description
CAPEX	Capital expenditure
CCUS	CO_2 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
fx	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
N_s	Number of stages
N_2	Nitrogen
O_2	Oxygen
OPEX	Operational expenditures
r	discount rate
€	Euro

Symbol	Description	Unit
A	Cross-sectional area	[m^2]
$h_{packing}$	Packing height	[m]
h_{shell}	Shell height	[m]
$V_{packing}$	Packing volume	[m^3]
V_{shell}	Shell volume	[m^3]
D	Diameter	[m]
$\dot{V}$	Volumetric flow rate	$[\frac{m^3}{s}]$
v_{gas}	gas velocity	$[\frac{m}{s}]$
$\dot{Q}$	Duty	$[kW]$
U	Overall heat transfer coefficient	$[\frac{kW}{m^2 \cdot k}]$
ΔT_{lm}	Logarithmic mean temperature difference	$[k]$

1 Introduction

The greenhouse effect is the primary cause of climate change. Some gases in the Earth's atmosphere function like greenhouse glass, trapping the sun's heat and preventing it from escaping into space, resulting in global warming. Although many greenhouse gases are produced naturally, human activity increases their amounts in the atmosphere. Human-caused CO_2 is the most significant contributor to global warming. Its concentration in the atmosphere has grown to 48% over pre-industrial levels (before 1750) by 2020 [1]. The burning of fossil fuels (coal, natural gas, and oil) for energy and transportation is the primary source of CO_2; however, specific industrial processes and land-use changes also generate CO_2 [2].

CO_2 capture, utilization, and storage (CCUS) has been proposed as a feasible strategy to deal with such emissions. The term "CCUS" refers to a method for capturing CO_2 from large-scale facilities that burn fossil or biomass fuels. The CO_2 gathered might be utilized on-site or compressed and stored in long-term storage facilities. Depleted oil and gas reservoirs, for example, can be used as storage sites [3], [4].

1.1 CO_2 capture in post-combustion power plants

According to the IEA report [5], emissions from the power industry (including electricity and heat generation) fell over 3%, or 0.4 Gt CO_2, in 2020, the largest-ever drop, while emissions intensity fell 2.8 %. Reduced power usage during the Covid-19 epidemic and a record percentage of renewables in overall generation (29 %) in 2020 have contributed to these changes. Because power generating accounts for 40% of energy-related CO_2 emissions and electricity is rapidly being utilized to fulfil end-use energy demand, this sector's transformation is crucial to clean energy transitions[5].

CO_2 capture and storage (CCS) is a three-step process that involves extracting CO_2 from emissions, transferring CO_2 to storage sites, and storing CO_2 in underground geological formations. CO_2 can be captured in both pre-combustion and post-combustion modes in combustion operations. Post-combustion CO_2 capture is the most widely used capture technology in fossil fuel power plants, as well as in the cement, steel, and iron sectors to some extent [6]. CO_2 capture is the most expensive part of the overall carbon capture and storage (CCS) operation, accounting for more than 70% of the total cost [7].

The absorption of CO_2 into solvents followed by desorption [8], the separation of CO_2 from the exhaust gas by membrane [8], the adsorption of CO_2 on solid adsorbents [9], the separation of CO_2 from flue gas by cryogenic means [8], and the direct injection of exhaust gas into naturally occurring gas hydrate reservoirs so CO_2 forms hydrate primarily with pore water [10] are all examples of carbon capture technologies and techniques.

The CO_2 removal process using monoethanolamine (MEA) solvent, in particular, is the most advanced CO_2 capture technology and is suitable for industrial implementation [11], [12]. The exorbitant cost of its industrial application remains the main obstacle. CO_2 capture and compression operations account for 80% of the total cost, whereas CO_2 transportation and storage processes contribute 10% [3], [13]. As a result, cost-cutting opportunities must be investigated, notably in the CO_2 capture process.

1.2 Literature review

The CO_2 capture process has the disadvantage of being associated with high capital costs (CAPEX) and an energy-intensive process that results in large operational expenses (OPEX) [12]. As a result, further cost-cutting efforts are critical. Several research projects have been undertaken to lower the cost of the CCS. This subchapter introduces and discusses some of the most relevant literature in this area.

1.2.1 Earlier work at HiT, HSN and USN

Lars Erik Øi utilized Aspen HYSYS to simulate a basic combined cycle gas power plant and a monoethanolamine (MEA)-based CO_2 removal process in 2007 [14]. The CO_2 removal in % and energy consumption in the CO_2 removal plant are calculated as a function of amine circulation rate, absorption column height, absorption temperature, and steam temperature. His PhD thesis [15]focused on MEA-based absorption and desorption calculation methods for CO_2 collection of ambient exhaust gas of a gas-based power plant. As a function of circulation rate, absorber temperature, and other factors, total CO_2 removal quality and heat consumption have been estimated. One of the project's goals was to determine the process's cost-optimal parameters.

Kallevik employed a novel make-up water and MEA calculation approach and modelling a direct contact cooler (DCC) unit in his master's thesis [16] at Telemark University College. For cost estimation purposes, the total heat transfer coefficient and the correction factor for heat exchangers were computed in this study. Cost fluctuations were recorded while changing process parameters such as the minimum approach temperature in the lean/rich heat exchanger, absorber packing height, and absorber gas supply temperature. CO_2 removal efficiencies of 80 %, 85 %, and 90 % were used in the parametric studies [16].

Using the Aspen HYSYS modelling tool, Park and Øi adjusted the gas velocity and pressure drop in a CO_2 absorption column for a typical amine-based CO_2 collection system in 2017 [17]. The six types of structured packings investigated in this study were Mellapak 250X, 250Y, 2X, 2Y, Mellapak Plus 252Y, and Flexipac 2Y. According to the modelling results, the cost-optimal gas velocity for all packings is in the range of 2.0 to 2.5 m/s, resulting in a pressure drop of 10 to 15 mbar across the absorber [17].

Øi et al. [18] modelled different absorption and desorption configurations for 85 % amine-based CO_2 removal from a natural gas-fired power plant using Aspen HYSYS. A standard process, split-stream, vapour recompression, and combinations were simulated. Simulations have been used to size equipment, predict costs, and optimize operations. The most cost-effective configuration of the investigated instances was judged to be a simple vapour recompression case [18]. Aromada and Øi [19] found that both vapour recompression and vapour recompression combined with split-stream operations can reduce energy consumption. The vapour recompression approach was found to be the most energy-efficient of the combinations evaluated [19]. In addition, energy optimization and economic analysis were carried out for the CO_2 capture parameters, which are based on a natural gas-based power plant project in Mongstad, Norway, and a calculation period of 20 years, including cost optimization using negative net present value (NPV). The vapour recompression option with 20 absorber stages, 9 desorber stages, and 1.2 bar flash pressure with a minimum approach temperature (ΔT_{min}) of 5 °C was anticipated to be the most energy efficient. According to a cost-benefit study, the vapour recompression technique with 15 absorber stages, 10 desorber stages, 1.3 bar flash pressure, and a ΔT_{min} of 13 °C was the most cost-effective, according to a cost-benefit study [20].

Ali et al. [21] explored the use of surplus heat to optimize amine-based CO_2 capture from a cement plant. Full-flow and part-flow flue gas were used to simulate traditional amine-based CO_2 removal. The Aspen In-Plant cost calculator was used to estimate the expenses of these case studies, and the cost of CO_2 collection was determined using a detailed factor approach and the Lang factor technique. The energy optimum was judged to be a full-flow alternative. When using the Lang factor technique, the cost-optimal condition was 60 per cent of the flue gas flow into the capture facility. When using the detailed factor approach, the scenario with 80 % of the flue gas flow is the most cost-effective choice [21]. Ali et al. [22] created a cost estimation tool that displays how different assumptions impact the overall cost of a capture plant and highlights the most relevant technical and economic factors. A simple process flow diagram and a list of equipment are supplied as input. Detailed installation criteria and equipment cost are the two critical components used to compute CAPEX, which are a basic component of the cost estimation approach[22].

In his PhD thesis [23], Hasan Ali sought to establish a system for conducting techno-economic analysis that identifies important components and illustrates the impact of various technological and economic assumptions on the overall cost of a capture plant. The proposed techno-economic analysis approach was used for a base scenario using amine-based post-combustion CO_2 collection (85 % capture rate) from a cement plant's flue gas. The cost of capture was 63 €/tCO_2. The steam cost, energy cost, and capital cost are the main contributors to the base case outputs. The Enhanced Detailed Factor (EDF) technique was used to determine the key cost factors, whereas the Lang factor method was not designed to provide such information. Although the projected steam cost is especially vulnerable to market variables such as fuel price, which changes considerably around the world, natural gas-based steam generation is anticipated to be more cost-effective than coal- or biomass-based steam generation in this study [23].

Øi et al. [24] attempted to calculate cost-optimal process parameters in a traditional amine-based approach for CO_2 collection from the cement industry in order to determine if automated cost estimation and optimization is feasible. The Aspen In-Plant cost estimator's equipment cost, and a detailed factor approach were used to assess the capital cost of CO_2 collecting. The number of absorption stages, the minimum temperature differential in the primary heat exchanger, and the percent CO_2 removed in the process were all factors. The ideal temperature differential in the primary heat exchanger was determined to be 10-15 °C depending on the parameters. The ideal column height was calculated using 12 stages (equivalent to 12 meters of structured packing) based on one simulation for each stage number [24].

Aromada et al. [25] investigated several CO_2 capture heat exchangers. Aspen HYSYS simulations of an 85 percent CO_2 absorption and desorption process for flue gas from the cement industry were used to calculate the costs. Every plant has its own type of lean/rich heat exchanger. The cost optimization of various heat exchangers is also part of this effort. Three different shell and tube heat exchangers were studied, as well as two plate and frame heat exchangers. Using a plate and frame heat exchanger instead of a typical shell and tube heat exchanger can result in significant capital and operational cost reductions [25]. Aromada et al. [26] investigated the installation factor and plant construction characteristic factor in another study. The influence of equipment installation parameters on the capital cost of an amine-based CO_2 collection system was evaluated using the EDF approach. The influence of installation parameters for seven approaches on capital cost were compared. Overestimation of high-cost equipment and underestimation of lower-cost equipment will almost surely result from a constant installation factor. Despite the fact that all methodologies determined the ideal ΔT_{min} in the cross-exchanger to be 15 °C. The results suggest that the EDF method may be used to estimate capital costs for new plants and modifications[26].

Aromada et al.[12] assessed the costs of using six different types of heat exchangers as the lean/rich heat exchanger in an amine-based CO_2 collection system. The gasketed-plate heat exchanger (G-PHE) saves a lot of space while also saving a lot of money. By replacing traditional shell and tube heat exchangers (STHXs) with the G-PHE, capture costs of €5–€6/tCO_2 can be lowered, and over €6/tCO_2 in the case of the finned double-pipe heat exchanger (FDP-HX). The cost of steam has the greatest impact on CO_2 collection costs in all situations [12]. In another research[27] a trade-off evaluation of energy cost and capital cost resulting from alternative temperature approaches in the cross-exchanger of a solvent-based CO_2 capture process was used to assess the efficacy of a plate heat exchanger (PHE) in contrast to traditional shell and tube types. The goal was to look at the cost-cutting and CO_2 emission-cutting potentials of various heat exchanger. For the PHE scenarios, the recommended minimum temperature approach based on CO_2 saved cost was 4 °C to 7 °C. The energy usage and indirect emissions are quite low in this area. The usage of plate heat exchangers for the cross-heat exchanger (at 4–7 °C), lean amine cooler, and DCC unit's circulation water cooler is recommended in this study [27].

1.2.2 Overall cost estimation of CO_2 capture

Roussanaly et al.[28] explore critical issues and elements that have a significant influence on cost evaluation outcomes yet are frequently ignored or inadequately handled. Cost indicators (particularly in the context of industrial facilities with numerous output products), energy supply concerns, retrofitting costs, CO_2 transit and storage, and capture technology maturity are among them. Because CCS retrofitting is so important for industrial plants, more thought is given to how to better account for the essential aspects that make up retrofitting costs [28].

Rubin et al. [29] proposed a standard costing approach and rules for CCS cost reporting to increase the clarity and uniformity of cost estimates for greenhouse gas mitigation methods. In 2014, Roussanaly et al. [30] presented a new systematic approach for designing and optimizing CO_2 capture membrane systems that integrate technological and economic principles. The technique was demonstrated by designing a post-combustion CO_2 capture membrane system placed on an Advanced Super Critical (ASC) power station and compared to an MEA capture unit.

In 2019, Van Der Spek et al. [31] investigated current advancements in CCS engineering and economic analysis. The design and size of equipment, cost indices and location considerations, process and project contingency costs, CO_2 transportation and storage costs, and uncertainty analysis and validation are all evaluated in this article. Xiaobo Luo presented modelling, simulation, and optimization research on the best design and operation of (MEA)-based post-combustion carbon capture (PCC) process and integrated system with natural gas combined cycle (NGCC) power plants in his PhD thesis [32], to lower the cost of PCC commercial deployment for NGCC power plants. The cost model in this study was built utilizing vendorprovided essential equipment costs from a benchmark report that contained a detailed technical design [32].

Simon Roussanaly [33] presents ways for assessing the CO_2 avoidance cost for the non-power production industry's Carbon Capture and Storage. In the case of CCS from industrial sources, unlike the power generation industry, three calculation approaches are frequently utilized to calculate the CO_2 avoidance cost. The connections between these three approaches are shown and confirmed using an example scenario to emphasize the requirements that must be satisfied for them to be used reliably, as well as their related defects. Finally, the foundation is offered to guarantee that the CO_2 avoidance cost calculation technique chosen is both valid and efficient for the applications examined by possible users [33]. Psarras et al.[34] assess the cost of CO_2 avoided for two scenarios: transport to and injection within reliable sequestration sites, and delivery and injection for the purpose of CO_2-enhanced oil recovery, utilizing site-specific emissions and regionally established cost parameters (EOR). Pieri and Angelis-Dimakis [35] looked at a significant number of research that assessed and reported CO_2 collection costs. The findings are categorized, homogenized, and standardized, and statistical models for each of the categories are created. Based on the amount of CO_2 collected and the type of source/separation principle of the capture system utilized, these models can estimate the capture costs.

In the context of plant-level multipollutant control needs, Rao and Rubin [36]created an integrated modeling framework (named IECM-cs) to assess the performance and cost of different CCS technologies and power systems. IECMcs is used to determine the optimum cost-effective degree of CO_2 management for PC plants utilizing currently available amine-based CO_2 collecting technologies. According to the results of this study, the most cost-effective degree of CO_2 reduction is determined by various plant design criteria, including plant size. The study also found that if low to moderate levels of CO_2 control are needed, the cost-effectiveness of CO_2 control may be enhanced by treating only a portion of the flue gas at high capture efficiency and bypassing the amine system with the remaining proportion of flue gas. In all cases, the maximum amine system train size and its impact on capital cost were shown to have a significant impact on CO_2 capture cost-effectiveness and the best (least expensive) degree of control [36].

Eldrup et al.[37] presented a Techno-economic analysis or early phase cost estimating approach (tool) that, when utilized correctly, may offer both an overall and individual cost indication. Such techniques may be utilized for both extremely young technologies in the early stages of development and applications with a greater Technological Readiness Level (TRL). A techno-economic analysis' process is as follows; first, determine the scope of the project. Second, create an equipment list. Third, determine the cost of the equipment. Fourth, determine the installation factors. Fifth, calculate the total cost, and finally in order to produce indications, combine CAPEX and OPEX. The fundamental benefit of this tool is that it achieves high accuracy with minimum effort, allowing for the identification of pieces with the greatest economic impact [37].

1.2.3 Cost estimation based on process simulation

The influence of amine type, energy penalty, and CO_2 capture efficiency (50 to 90 %) on capture costs (US$/ton CO_2 and US$/ton cement) was investigated by Nwaoha et al. [38]. A sensitivity analysis was conducted on the influence of CO_2 capture plants, carbon taxes (ranging from $20 to $40 per ton of CO_2), CO_2 sales prices (ranging from $10 to $40 per ton of CO_2), energy penalties, and CO_2 capture efficiency on cement prices. The capture costs of AMP-PZ-MEA are lower than MEA at 90 % CO_2 capture efficiency, according to the data. The MEA system also had a higher total equipment cost and capital expenditure (CAPEX) than the AMP-PZ-MEA mix [38].

Amir Ayyad et al. [39] modelled two configurations and assessed their economic feasibility in order to reduce reboiler duty and power loss at Egypt's 495 MW West Damietta power plant. The first approach is to recycle part of the exhaust gas back into the combustion chamber to increase carbon concentrations in the feed to the carbon capture plant; the second approach is to use parabolic-trough solar collectors to handle the reboiler load instead of low-pressure steam extracted from the power plant to handle the reboiler load. The findings demonstrated that increasing carbon content resulted in a significant reduction in reboiler duty of up to 20%. Carbon increases also had an impact on the levelized cost of energy, which decreased by 1.39 percent and the carbon cost of avoidance decreased by 6% when utilizing a 35 % recirculation ratio. It was also shown that combining a solar plant with a thermal storage system significantly enhanced optimal production when compared to a plant without thermal storage [39].

G. Manzolini et al. [40] evaluated the economic benefits of the CESAR-1 solvent, which is an aqueous solution of 2-amino-2-methylpropanol (AMP) and piperazine (PZ) used in advanced supercritical pulverized coal (ASC) and natural gas combined cycle (NGCC) power plants with CO_2 capture units. Because of the higher CO_2 content in the flue gas, the techno-economic benefit of CESAR-1 against MEA is greater for ASC than for NGCC, according to the findings. This is because switching from MEA to CESAR-1 solvents decreases the power cost by 4.16 €/MWh in the ASC plant against 0.67 €/MWh in the planned NGCC plant. In comparison to MEA, CESAR-1 lowers the cost of CO_2 avoided by 6 €/t CO_2 and 2 €/t CO_2 for the selected ASC and NGCC plants, respectively[40].

In the UniSim process simulator, Oh et al.[41] built a superstructure with the traditional amine-based CO_2 collection configuration and four alternative types of structural modifications. This superstructure's optimization exposes the configuration and operating circumstances that result in the lowest energy costs, taking into account all conceivable alterations in a systematic and simultaneous manner. The suggested modeling and optimization framework is applied to a CO_2 capture case study to demonstrate how it may successfully evaluate design possibilities for enhancing energy efficiency [41].

Dutta et al.[42] investigated the selection and design of a post-combustion CO_2 capture (PCC) facility for a natural gas power plant in another study. Two modified PCC plant layouts were chosen, each with a minor efficiency penalty. Design limitations based on operability and the building of absorbers were incorporated into the process for developing PCC plants. This was used to determine the plant's equipment size. Based on flue gas flow rate at full load and time-average of an estimated load fluctuation of a flexibly functioning power plant, two absorber layouts were investigated. The absorber built for time-average load resulted in a 4 percent decrease in absorber purchasing costs. In order to maintain a similar capture rate to that of the other absorber during part-load operation, the absorber designed for full-load operation resulted in lower reboiler duty [42].

Based on process and economic assessments, Agbonghae et al.[43] evaluated four MEA-based CO_2 capture units for both gas-fired and coal-fired power stations. The findings reveal that for absorber and stripper columns packed with Sulzer Mellapak 250YTM structured packing, the optimal lean CO_2 loading for MEA-based CO_2 capture systems that can serve commercial-scale power plants, whether natural gas-fired or coal-fired, is about 0.2 mol/mol. Furthermore, the optimum liquid/gas ratio for a natural gas combined cycle (NGCC) power plant with a flue gas composition of approximately 4 mol percent CO_2 is about 0.96, whereas the optimum liquid/gas ratio for a pulverized coal-fired (PC) power plant with a flue gas composition of 12.38 mol percent to 13.50 mol percent can range from 2.68 to 2.93 [43].

Luo and Wang [44] aimed to determine the best operation for an assumed existing natural gas combined cycle (NGCC) power plant with an MEA-based post-combustion carbon capture (PCC) process under various market situations. In optimization studies, the levelized cost of electricity (LCOE) is used as the objective function. The integrated system's basic scenario, which included CO_2 transport and storage, was evaluated economically. According to the analysis, a carbon price of above 100 EUR /ton CO_2 is required to justify the overall cost of carbon capture from the NGCC power plant, and a price of 120 EUR /ton CO_2 is required to achieve a carbon capture level of 90% [44].

Mathisen et al.[45] looked at a system that combined a natural gas power plant with CO_2 collection, with the energy coming from a biomass-based external energy plant. Estimates of capital and operational costs are used to compare the concept to other options. A constraint on the operating cost estimates is that for every tonne of non-biobased CO_2 emitted into the atmosphere, a CO_2 quota must be purchased, and for every tonne of biobased CO_2 recovered, the value of a CO_2 quota is awarded. Under specific conditions, such as low biomass costs and high CO_2 quota costs, the approach has been demonstrated to be economically viable. Different scenarios for the following are altered in a sensitivity analysis: CO_2 quota price, natural gas price, and biomass pricing. Scenarios in which the price of CO_2 quotas is raised benefit the suggested strategy [45].

Zhang et al. [46] looked at how employing MEA-MDEA-PZ as a post-combustion carbon dioxide capture (PCC) solution reduced energy costs. The heat of CO_2 absorption in MEA-MDEA-PZ was investigated at various mix ratios. The MEA-MDEA-PZ blend can cut CO_2 capture energy costs by 15.22–49.92 percent [46].

Gatti et al.[47] evaluated the technological and economic possibilities of four different techniques for capturing CO_2 from natural gas-fired power plants post combustion. These include CO_2 permeable membranes, molten carbonate fuel cells (MCFCs), pressured CO_2 absorption with a multi-shaft gas turbine and heat recovery steam cycle, and supersonic flow-driven CO_2 anti-sublimation and inertial separation. The reference example is a modern natural gas combined cycle (NGCC) without CO_2 capture, while the base case is the same NGCC constructed with CO_2 capture (using chemical absorption with an aqueous monoethanolamine solvent). In a separate benchmarking instance, the same NGCC is outfitted with aqueous piperazine (PZ) CO_2 absorption in order to examine the techno-economic potential of an advanced amine solvent. The analysis shows that a combined cycle with MCFCs appears to be the most appealing technology in terms of both energy penalty and economics. PZ scrubbing is the second-best capture method, followed by the monoethanolamine (MEA) base case [47].

Over a range of feed compositions, Hasan et al.[48] investigated the modeling, simulation, optimization, and energy integration of a monoethanolamine (MEA)-based chemical absorption process and a multistage membrane process. The minimal annualized cost of the MEA-absorption process is determined using a robust simulation-based optimization approach. The MEA-absorption process is made more energy efficient by optimizing the heat exchanger network. A unique mathematical model for the optimization of multistage and multicomponent

CO_2 separation using membranes is created, which may be used to a variety of membrane-based gas separation applications. The results, which indicate the best investment, operational, and total costs, give a quantitative method to technology comparison and scaling up CO_2 collection from diverse CO_2 generating sectors using absorption and membrane technology. As a function of feed flow rate and CO_2 composition, explicit formulations for the investment and operational costs of each possible post combustion CO_2 capture method are also generated [48].

Schmelz et al.[49] calculated the costs of carbon capture and storage (CCS) in subsurface geological formations for emissions from 138 electricity-generating power plants in the northeastern and midwestern United States. According to the calculations, coal-sourced CO_2 emissions can be stored in this location for $52–$60 per ton, but natural-gas-fired plant emissions may be kept for $80–$90 per ton.

Hasan et al. [50] compared the potential for CO_2 reduction of several amine-based solvent solutions (monoethanolamine (MEA), diethanolamine (DEA), and methyldiethanolamine (MDEA)) under various operating circumstances and costs. This was modeled as a basic absorber tower to collect CO_2 from flue gas using ProMax 5.0 software. According to the findings, MEA is a favorable solvent in terms of CO_2 capture when compared to DEA and MDEA; however, it is limited at the top outlet for clean flue gas, which contained 3.6295 percent CO_2 for 10 % MEA concentration and solvent circulation rate of 200 m^3/h, but this can be addressed by increasing the concentration to 15% or increasing the MEA circulation rate to 300 m^3/h [50].

1.2.4 Challenges in simulation and cost estimation

When employing MEA-based chemical absorption to capture CO_2 from an ambient gas stream, Husebye et al. [51] looked at the influence of CO_2 concentration and steam supply. An increase in CO_2 concentration decreases operating and investment costs due to lower energy consumption and reduced equipment capacity. Investment costs dominate the rapid reduction in the net present value of expenses when CO_2 concentrations are increased from 2.5 % to 10 %. Still, cost drops are more modest when CO_2 concentrations are increased from 10% to 20% [51].

Exhaust gas recirculation is a way of increasing CO_2 content in the lean flue gas for natural gas-fired power production systems. Ali et al.[52] reported on the design and scale-up of four separate scenarios of an amine-based CO_2 collection system with a 90% capture rate and a 30 wt.% MEA aqueous solution. Design findings for a natural gas-fired combined cycle system with a gross power output of 650 MWe without EGR and with EGR at 20 %, 35 %, and 50 % EGR percentages are presented. An optimum liquid to gas ratio of 0.96 is determined for an amine-based CO_2 capture plant with a natural gas-fired combined cycle without EGR. The ideal liquid to gas ratios are 1.22, 1.46, and 1.90, respectively, when EGR is used at 20 %, 35 %, and 50 %. These findings indicate that a natural gas-fired power plant with exhaust gas recirculation will result in lower energy consumption and costs than an amine-based CO_2 capture facility[52].

Using an aqueous solution of monoethanolamine (MEA), Sipöcza and Tobiesen [53] investigated the thermodynamics and economics of a natural gas combined cycle power plant with an integrated CO_2 removal facility. The CO_2 capture plant flowsheet has been adjusted for operation conditions and incorporates absorber intercooling and lean vapor recompression. In addition, to further minimize CO_2 capture costs, the gas turbine employs a 40 % level of exhaust gas recirculation (EGR), resulting in a CO_2 concentration in the gas turbine exhaust gas that is nearly double that of conventionally running gas turbines. It has been demonstrated that combining EGR with a lower specific reboiler duty reduces both operating and capital costs significantly. It is also demonstrated that fuel prices and currency rates play a significant impact in estimating expenditures with accuracy [53].

There is a lot of work being done in the areas of simulation and cost optimization. This study is a continuation of past CO_2 capture modeling and cost estimation research undertaken by the University of South-Eastern Norway (USN) for several years utilizing Aspen HYSYS. As a result, USN's earlier work play a significant role in this project. The goal of this project is to use Aspen HYSYS software to automate the cost optimization of an MEA-based CO_2 collecting process. As a consequence, it's worth looking at which elements influence plant costs.

The price of a plant is mostly determined by five things. The exhaust gas flow into the absorption column is the first. The size of the process equipment in the gas route are affected by this. The CO_2 level of the flue gas is the second factor to consider. Due to a larger driving force, a high concentration reduces energy consumption. Third, as the rate of CO_2 removal increases, so does energy consumption. Fourth, the size of the equipment and the amount of energy required are determined by the solvent flow rate. The fifth factor is the energy demand for hot water and electricity. With a high solvent flow rate, the amount of thermal energy required increases. With a high solvent flow rate, the amount of thermal energy required increases. The flue gas transit through the process accounts for the majority of the electricity consumption, which will increase as the pressure requirement and volume flow increase [16].

To save money on the plant, this research wants to run a sensitivity analysis to see how changing process parameters affects the overall cost. The flue gas temperature into the absorber, the pressure into the absorber, the minimum temperature difference in the lean/rich heat exchanger, the reboiler temperature, the condenser temperature or the reflux ratio, the solvent circulation rate, the pressure in the desorber, and the efficiency of the CO_2 removal rate of the process are the typical choices of process parameters, according to Øi [15]. The minimum temperature in the lean/rich amine heat exchanger (ΔT_{min}), the absorber packing height and the flue gas temperature into the absorber were all evaluated as process parameters in this study.

1.3 Scope of study

This study investigates an amine absorption CO_2 capture method using emission data from earlier research on a natural gas-based power plant project in Mongstad, Norway [20], [27]. Based on the supplied data, a base case was created in Aspen HYSYS, and then dimensioning and cost estimation were conducted. To estimate and compute the overall cost for the base scenario, the Enhanced Detailed Factor (EDF) is used in conjunction with the Aspen In-Plant Cost estimator. A sensitivity analysis is used to do cost optimization in order to minimize and lower costs. The approach was tested in a series of case studies to see how different variables influenced price fluctuations. When the sensitivity analysis alters the size, the Power-Law approach is used to adapt the equipment cost. The minimum temperature approach in the lean/rich amine heat exchanger, the absorber packing height, and the incoming flue gas temperature into the absorber column were all varied in this investigation.

The main objective of this study is to employ the Aspen HYSYS tool to calculate and optimize the cost of an MEA-based CO_2 collection plant automatically. Some research has been done to automatically compute the minimal cost by evaluating the change in the minimum temperature difference (ΔT_{min}) in the primary heat exchanger [27], [54], [55]. Still, this is the first time that automation is used in a sensitivity analysis to account for changes in the number of stages in the absorber as well as the temperature of the incoming flue gas.

2 Process description and base case simulation

This chapter includes a broad overview of a carbon capture process as well as an explanation of how to simulate the base case for cost evaluation and optimization in the sensitivity analysis.

2.1 CO_2 capture process

The CO_2 removal method may be classified into three major categories: post-combustion, oxy-combustion and pre-combustion. The CO_2 is removed from the combustion gases in a post-combustion process, according to the concept. Because CO_2 has a low partial pressure, absorption is a frequent and effective way to remove it from an exhaust gas. MEA and MDEA, for example, are aqueous amines that are employed in such absorption procedures [15]. N_2 is removed from air in an oxy-combustion process, and pure oxygen, together with fuel, enters the combustion chamber. Due to the pure oxygen, the combustion chamber will reach extremely high temperatures, which may cause design issues. CO_2 and water vapor will be the major products, with CO_2 being separated and then sent back to the combustion chamber to lower the combustion temperature. CO_2 is removed from the stream before it enters the combustion chamber in a pre-combustion process. This can be accomplished by using steam reforming, autothermal reforming, catalytic partial oxidation, or gasification to produce synthesis gas [16], [23].

The use of an amine solvent to remove CO_2 is the most widely used and well-studied approach for CO_2 removal. MEA is the solvent that has been studied the most, and it works well due to its quick interaction with CO_2. Another advantage of the MEA is its commercial availability as well as its high CO_2 capacity. MEA's disadvantages include a high proclivity for corrosion, toxicity, and deterioration[23]. Figure 2.1 is a typical process flow diagram for an amine-based CO_2 removal facility. Traditional absorption is done in a column using plates, random packing, or structured packing. The CO_2-containing gas rises, while the absorption liquid falls. The solvent (rich amine) is then fed to a desorption column through a heat exchanger. In a desorption (stripper) column, the CO_2 that has been absorbed is regenerated. The reboiler is heated, and a condenser provides reflux to the column. The regenerated solvent (lean amine) is recirculated to the absorption column after the desorber and cooled in a heat exchanger and cooler [15].

A direct contact cooler (DCC), a water wash section at the top of the absorber, and an amine reclaimer after the desorber are also included. The DCC uses circulating water that runs downhill in a column to cool the exhaust gas. The water is circulated by a pump and is cooled indirectly, for example, by cooling water. At the top of the absorption column is a water wash section. There are residues of the solvent near the top of the absorption part that should not be released into the environment. Water runs downhill in the wash stage, absorbing amines and other solvent components. A pump circulates the water, which is then supplemented with clean make-up water. A tiny portion of the wash water goes to the main absorption section of the column to avoid amine build-up [15].

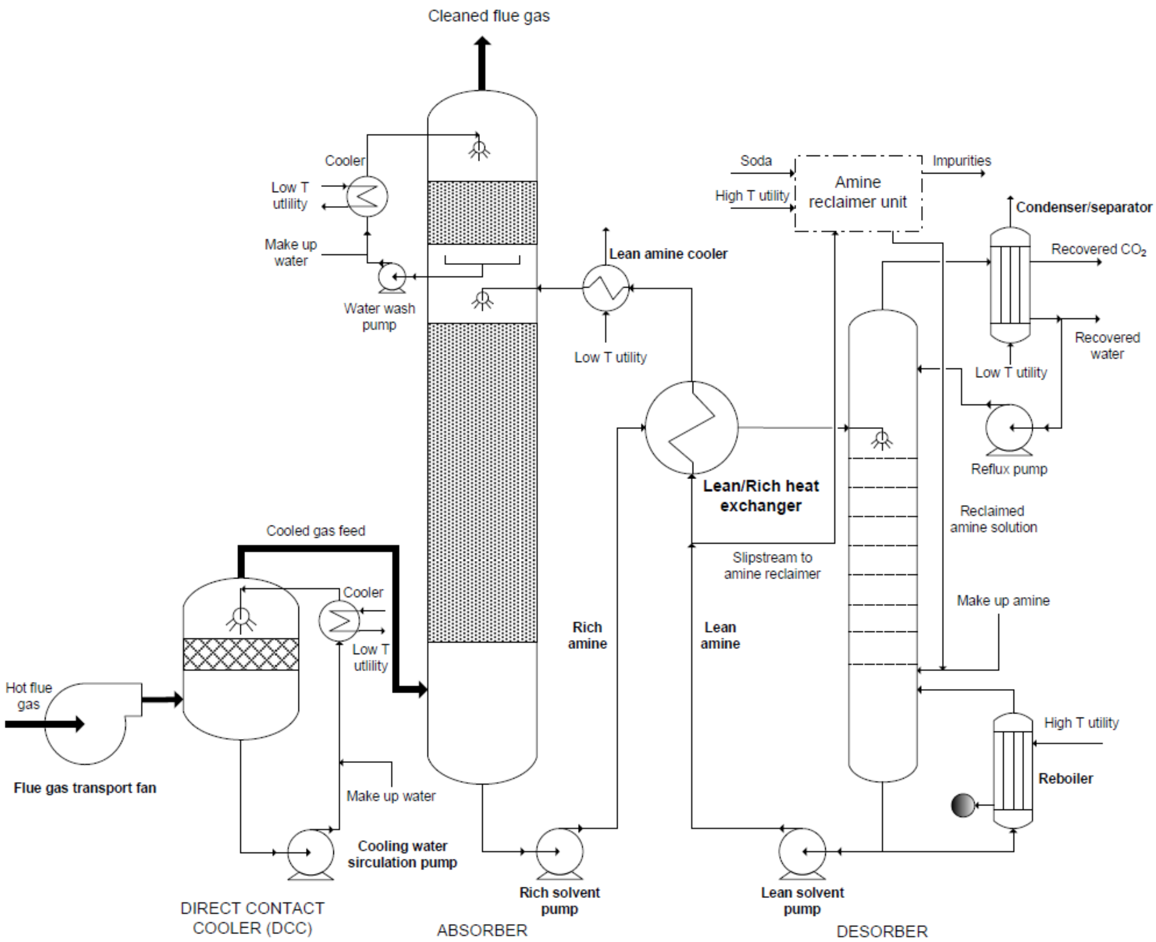

Figure 2.1: General process flow diagram (PFD) of a CO_2 removal process plant [16].

2.2 Base case simulation

In this investigation, Aspen HYSYS version 12 was used to model a conventional amine-based CO_2 capture process, and the simulated results were utilized to size equipment and estimate costs using the same calculation method as in the literature [56], [57]. In all simulations, the fluid package of the Acid Gas property package was employed, which included vapour and liquid equilibrium models for electrolyte. This package is intended to replace the Amine property package, which is widely used in literature. The electrolyte non-random two-liquid (e-NRTL) model for electrolyte thermodynamics and the Peng-Robinson equation of state for the vapor phase were used to create this property package. For numerous amine solvents and their mixes, the models provide rate-based and thermodynamic modeling of acid gas removal (H_2S and CO_2). The reactions provided in the Acid Gas property package for MEA (solvent) interacting with CO_2 are listed in Table 2.1[54].

Table 2.1: Reactions included in the Acid Gas property package for MEA solvent reacting with CO_2 [58]

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

The absorber and desorber were simulated using equilibrium stages containing user defined stage (Murphree) efficiency. For the absorber and desorber, a constant Murphree efficiency of 0.15 and 0.5 was assessed as one meter of structured packing, respectively. These Murphree efficiencies are calculated by dividing the change in CO_2 mole fraction from one stage to the next by the change in the assumption of equilibrium. Rather of assuming perfect equilibrium, this is a simple technique to attain a more realistic performance [15], [23].

Emission data from previous studies [20], [27] on a natural gas-based power plant project in Mongstad, Norway, were utilized to generate the base case for the simulations. The specifications in Table 2.2 correspond to an 85 per cent CO_2 removal efficiency and a minimum approach temperature of 10 °C in the lean/rich amine heat exchanger, which is considered the base case configuration. The technique of computation is similar to that employed in prior studies [19], [24], [56]. The absorption and desorption columns were modeled as equilibrium stages with stage efficiency. The absorber is modelled with 15 packing stages, while the desorber has eight. Equilibrium stages of 1 m height are examined for both columns. Murphree efficiencies of 15% were employed in the absorption column. A consistent Murphree efficiency of 50% was given for all stages of the desorption column. In the columns, the Modified HYSIM Inside-Out approach was adopted since it assists in convergence. The adiabatic efficiency of the pump and flue gas fan was stated to be 75%.

Table 2.2: Aspen HYSYS model parameters and specifications for the base case configuration

Items	Specifications [Unit]	Value
Inlet Flue Gas	Temperature [°C]	80
	Pressure [bar]	1.01
	Molar flow rate [kmol/h]	85,000
	CO_2 content [mole %]	3.75

	H_2O content [mole %]	6.71
	N_2 content [mole %]	89.54
Flue gas to absorber	Temperature [°C]	40
	Pressure [bar]	1.15
Lean MEA	Temperature [°C]	40
	Pressure [bar]	1.01
	Molar flow rate [kmol/h]	103,500
	MEA content [W %]	29
	CO_2 content [W %]	5.5
Absorber	Number of stages	15
	Murphree efficiency [%]	15
	Rich amine pump pressure [bar]	2
	Rich amine temp. out of Lean/Rich amine HEx [°C]	103.7
Desorber	Number of stages in stripper	10
	Murphree efficiency [%]	50
	Reflux ratio in the desorber	0.3
	Reboiler temperature [°C]	120
	Pressure [bar]	2
	Lean amine pump pressure [bar]	5

The Aspen HYSYS simulation process flow diagram (PFD) is shown in Figure 2.2. The computation procedure is similar to that of previous research [19], [24], [56]. Despite the fact that Aspen HYSYS is an equation-based tool, the computation process is sequential and modular. From the input gas and the lean amine, the absorption column is first determined (which is first guessed) [24]. The rich amine pump transports the rich amine from the bottom of the absorption column through the lean/rich amine heat exchanger. After the heat exchanger, the temperature is defined. The CO_2 product and the hot lean amine are calculated as the heated rich amine enters the desorption column. The heated lean amine is passed via the lean/rich heat exchanger then pushed to a greater pressure in the lean amine pump, before being cooled further in the lean cooler. The lean amine is then placed in a recycle block. It is determined whether the recycled lean amine's flow and condition are sufficiently similar to the previously estimated lean amine stream, which may be adjusted through iteration. In Aspen HYSYS, recycling blocks are required to solve the flowsheet, this block compares the block's in-stream to the block's out-stream in the previous iteration [16].

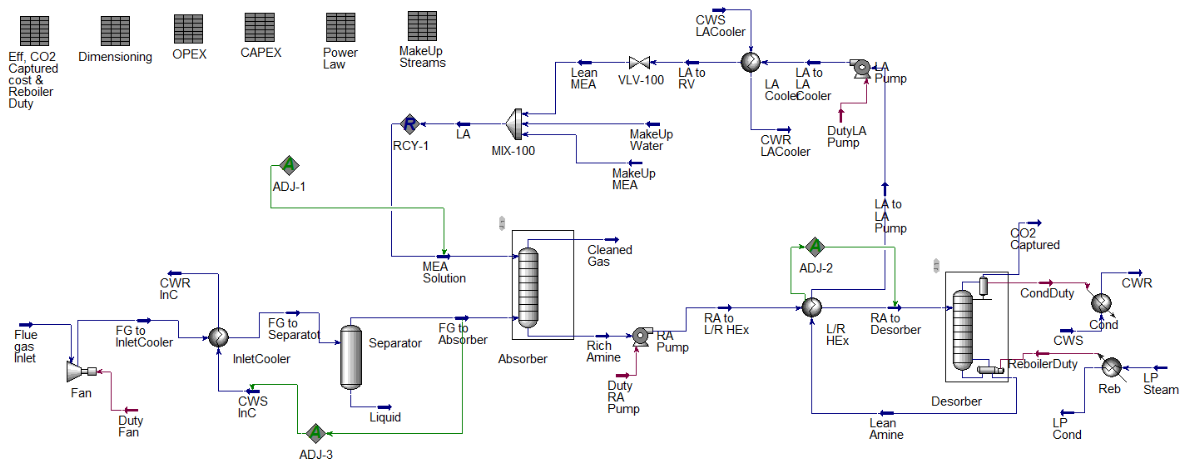

Figure 2.2: Aspen HYSYS flow sheet for the Base Case simulation

In order to create an automated simulation model, three adjust operations were added to the flowsheet. The removal efficiency can be adjusted based on the lean amine flow rate by ADJ-1, the minimum approach temperature in the lean/rich heat exchanger may be adjusted based on the rich amine outlet temperature of the lean/rich heat exchanger by ADJ-2, and for adjusting the flue gas temperature to the absorber, ADJ-3 changes the cooling water supply in the inlet cooler.

The simulation does not include the water wash and amine reclaimer equipment shown in Figure 2.1. To compensate for the loss of water and amine, makeup streams were added to the flowsheet. To do this, a makeup streams spreadsheet was created, and the loss of MEA solution and water was computed using mass balance and exported to the makeup water and makeup MEA streams as mass flow. This eliminates the need to manually compute the amount of MEA and water for each iteration in the recycle block (RCY-1) and instead automates the process, which may aid convergence.

Also, instead of direct contact cooler (DCC) unit a simplified inlet cooler and separator are simulated. The flue gas is drawn into a fan and the inlet cooler, both of which are required to provide the desired pressure and temperature at the absorber input. Other trace components were not included in the flue gas, which was only supposed to include CO_2, N_2, and H_2O. Furthermore, because of the temperature drop in the inlet cooler, there may be a tiny quantity of water in the flue gas before it passes through the absorber. Water can be separated from gas in the separator. MEA has not been subjected to thermal or chemical degradation [59]. The scope also excludes further processing of the desorber overhead product, such as water separation, drying, and compression.

3 Dimensioning of equipment

The process equipment included in the process simulation scope is simply dimensioned in this chapter. The calculations are based on the process flowsheet results from Aspen HYSYS with equations of state, and energy and material balances. The information in this part serves as the foundation for cost estimating computations. All of the equipment sizing calculations were conducted in a spreadsheet called dimensioning, which was then utilized to estimate costs.

3.1 Scope of analysis

Only the key components, such as the absorption and desorption column, heat exchangers, fan, pumps, and separator, will be dimensioned. This research does not include any pre-treatment, such as inlet gas purification, or post-treatment, such as CO_2 compression, transport, or storage. Only the cost of the specified equipment installed is included in the cost estimate. Land purchase, preparation, service buildings, and ownership expenditures are not included.

3.2 Absorber and desorber (stripper) columns

In this scenario, the Murphree efficiency was set at 15% per meter of absorber packing for each stage, with one stage equaling one meter in the absorber [24]. The pressure drop in the absorber was estimated to be 0.010 bar. The Modified HYSIM Inside-Out solver with adaptive damping was chosen due to concerns with the absorber's convergence. When using this solution, according to Øi [14], the convergence results are more stable. A full reflux condenser was simulated within the desorber. This means that the overhead product will only be gas in the vapour phase. Reflux ratio equal to 0.3 in condenser and constant temperature of 120 °C in reboiler were considered. A constant pressure of 2 bar was assumed for desorber. The efficiencies of the reboiler and condenser were both set to one. This means that the procedure is completely efficient. The Murphree efficiencies for the packing phases were set at 0.5 [16].

The volume of packing and shell are employed as dimensioning parameters for further cost assessment. The packing is the most expensive part of a column, and in this study, structured packing (Mellapack 250Y) was chosen because of its high efficiency, high capacity and minimal pressure drop [15].

Equations (3.1) and (3.2) are used to compute the packing/shell volume, which is a function of column diameter (D [m]) and packing/shell height ($h_{packing}$/h_{shell} [m]) [60].

$$V_{packing} = \frac{\pi \cdot D^2}{4} \cdot h_{packing} \; [m^3] \tag{3.1}$$

$$V_{shell} = \frac{\pi \cdot D^2}{4} \cdot h_{shell} \; [m^3] \tag{3.2}$$

The area is a function of the actual volumetric gas flow rate ($\dot{V}$ [m^3/s]), and the diameter is determined from the area, assuming a circle. At the stage with the highest flowrate, the real volumetric gas flow rate is derived from the simulation. In this scenario, the maximum flowrate was found in stage two for absorber, counting down from the top and stage 10 for desorber. The gas velocity (v_{gas} [m/s]) is a parameter that is assumed. Using a gas velocity of 2.5 m/s in the absorber and 1 m/s in the desorption column [17], the cross sectional area can be estimated using equation (3.3), and the diameter can be obtained using equation (3.4).

$$A = \frac{\dot{V}}{v_{gas}} \; [m^2] \tag{3.3}$$

$$D = \sqrt{\frac{4 \cdot A}{\pi}} \; [m] \tag{3.4}$$

The packing height was determined by assuming the 1-meter height of each stage (h_s [m]) and the number of stages (N_s [-]) collected from the simulation. Equation (4.5) presents the packing height.

$$h_{packing} = h_s \cdot N_s \tag{3.5}$$

The overall heights of the absorption and desorption columns are expected to be 35 and 25 meters, respectively. The packing, liquid distributors, water wash, demister, gas inflow and outflow, and sump are all taken into account when calculating the absorber height. The desorber height calculation takes into account the condenser inlet, packing, liquid distributor, gas input, and sump [23], [24]. The height was calculated using data from a Sulzer catalog. The packing height was given on a design of a wash tower in the catalog [61]. The given height was utilized as a reference height in the sketch to calculate the total height of the tower. Figure 3.1 is a screen picture from "SketchUp". As a result, the additional increased height in the absorber has been adjusted to 20 meters, while the extra added height in the desorber has been specified to 15 meters.

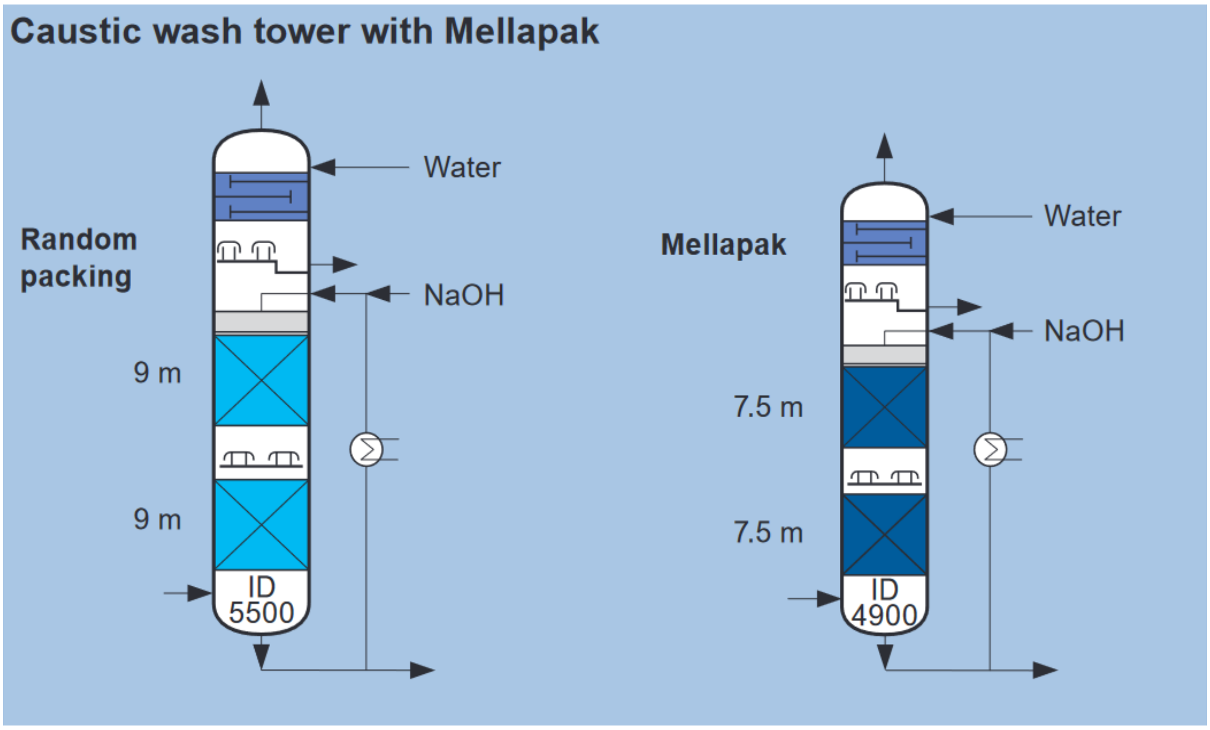

Figure 3.1: Dimensioned drawing from "SketchUp" [62].

3.3 Heat exchangers

The heat transfer area is a regularly used dimensioning factor for estimating the cost of heat exchangers. The duty ($\dot{Q}[kW]$), the overall heat transfer coefficient ($U\ [\frac{kW}{m^2 \cdot K}]$), and the logarithmic mean temperature differential ($\Delta T_{lm}\ [K]$) all affect the heat transfer area. Equation (3.6) is used to compute the heat transfer area (A [m2]) [63].

$$A = \frac{\dot{Q}}{U \cdot \Delta T_{lm}} [m^2] \tag{3.6}$$

The heat transfer areas of the heat exchangers are computed based on the duties and temperature conditions acquired from the simulations and an ideal countercurrent flow assumption. The logarithmic mean temperature (ΔT_{lm}) was calculated based on equation (3.7), whereas equations (3.8) and (3.9) provide the countercurrent flow configuration for inlet temperature.

$$\Delta T_{lm} \frac{\Delta T_1 - \Delta T_2}{ln\left(\frac{\Delta T_1}{\Delta T_2}\right)}\ [K] \tag{3.7}$$

$$\Delta T_1 = T_{hot\ in-} T_{cold\ out} [K] \tag{3.8}$$

$$\Delta T_2 = T_{hot\ out-} T_{cold\ in} [K] \tag{3.9}$$

The overall heat transfer coefficients were specified to be 732 W/ (m^2.K) for the lean/rich amine heat exchanger, 800 W/ (m^2.K) for the lean amine cooler and inlet cooler, 1200 W/ (m^2.K) the reboiler is, and the condenser is 1000 W/ (m^2.K)[23], [27], [38]. The heat exchangers used in this study are typical shell and tube heat exchangers (STHXs). The temperatures of the cooling water input and output were set at 15 and 25 degrees Celsius, respectively. Low pressure (LP) steam was fed to the reboiler at a temperature of 145 °C and a pressure of 3 bar, and the low pressure condensed left at 130 °C and 2.7 bar [27]. Furthermore, 1000 m^2 was specified as the maximum heat exchangers area per unit [31].

3.4 Fan and pumps

In Aspen HYSYS, the fan and pumps are supposed to have a 75 % adiabatic efficiency. The duty is utilized as the dimensioning parameter for the pumps and fans which obtained from the Aspen HYSYS, but the volumetric flow is also employed in the Aspen In-Plant cost estimator to calculate the equipment cost. For the fan, the maximum allowable flow is limited to 1.529 E+6 m^3/h, which was used to estimate the actual necessary units of this equipment. The fan's output pressure was supposed to be the same as the pressure of rich amine leaving the absorber's bottom.

The rich amine solvent must be pumped to the desorber. The pressure after the rich amine pump is 2 bar, which is the working pressure for the desorber. Pressure losses due to piping and friction has been neglected in this project. Following the desorber, a pump is required. This is related to the absorber height, which will necessitate more pump power in order to achieve the required lifting height. After the lean/rich heat exchanger, a lean amine pump was installed to raise the pressure of lean amine to 5 bar.

3.5 Separator

Because of the temperature drop in the inlet cooler, there may be a tiny quantity of water in the flue gas before it passes through the absorber. Water may be removed from the flue gas in the separator, allowing the flue gas to reach the absorption column saturated with water at around 40 °C. In this study instead of a DCC tower a separator is used. The separator was designed as a vertical vessel, with a diameter determined using the Souders–Brown equation (equation (3.10)) with a k-factor of 0.15 m/s and a tangent-to-tangent to diameter ratio of 1[16]. The wall thickness was neglected as it does not a big effect on the cost of separator. In this case the volume is utilized as a factor for cost estimation. Equations (3.3) and (3.4) are used to calculate the diameter.

$$v_{gas} = k * \sqrt{\frac{\rho_l - \rho_v}{\rho_v}} \ [m/s] \tag{3.10}$$

In this equation k is the sizing factor, ρ_l and ρ_v are the liquid and gas phase densities, respectively which are determined from simulation.

Appendices B and C present the parameters, specifications and the result for base case dimensioning.

4 Cost estimation

This chapter goes through some of the theories and methods for estimating the cost of a CO_2 collection operation. Based on the process simulation the following procedure is applied to estimate the total cost of the plant:

- Calculation of equipment cost in Aspen In-Plant Cost Estimator (v.12) based on equipment dimensioning parameters for the Base Case
- Calculation of the total installation cost applying the Enhanced Detailed Factor (EDF)
- Correction of cost index (conversion for year)
- Estimation of annual operational expenditure (OPEX)
- Calculation of total annual cost based on a given discount rate and plant lifetime
- Calculation of CO_2 captured cost
- Using the Power Law approach to scale the cost during parameter variation

The cost analysis is limited to the equipment shown in the Aspen HYSYS flowsheet in Figure 2.2. This study does not include any pretreatment, such as incoming gas purification, or posttreatment, such as CO2 compression, transport, or storage. Only the cost of the specified equipment installed is included in the cost estimate. Land purchase, preparation, service buildings, and ownership expenditures are not included.

4.1 Capital expenditure (CAPEX)

The Enhanced Detailed Factor (EDF) approach is used to estimate capital costs (CAPEX) in this study. This method is based on elements that affect the installation of each piece of process equipment. There are various benefits to employing the EDF approach, according to Ali et al. [22]. In the early stages of cost estimation, this approach works well and provides an accurate estimate. It also allows for techno-economic evaluations of both new and current technologies and plants.

4.1.1 Equipment cost

The cost of each piece of equipment was computed using the Aspen In-Plant Cost Estimator (v12.0) software, which is a cost estimating tool that uses material, dimensioning variables, and process data to get reasonable estimates for the overall equipment cost. The number of parameters given by the user determines the accuracy of the estimate. Except for the dimensioning parameters discussed in Chapter 3, Aspen In-Plant gave default values to the remaining specifications. Aspen In-Plant (v12.0) gives the price in Euro (€) for the year 2019 (1st Quarter), and the default location is Rotterdam in Netherland.

In this assessment, the actual location of Rotterdam is assumed. An installation factor sheet created by Nils Henrik Eldrup [56] was used to calculate the cost of each piece of equipment. This sheet covers all of the equipment's installation variables and specifies that the equipment's cost be estimated in carbon steel (CS). The component's preferred material can be chosen in the Aspen In-Plant Cost-estimator, however the cost should be converted to the cost of carbon steel (CS) using the material factor according to the EDF approach [22].

$$Equipment\ Cost_{CS} = \frac{Equipment\ Cost_{SS,\ exotic,...}}{f_{mat}} \quad (4.1)$$

f_{mat} in equation (4.1) is the material factor to convert the cost of one specific material to the cost of carbon steel. Apart from the flue gas fan, which is composed of carbon steel, all equipment is supposed to be stainless steel (SS316). The material factor to convert costs in SS316 to CS for welded and rotating equipment is 1.75 and 1.30, respectively [56].

4.1.2 Total installed cost

Each piece of equipment has a total installation factor ($F_{T,\ CS}$), which is the sum of the component's sub-factors (direct costs, engineering, administration, commissioning, and contingency) [22], [56]. This can be written as:

$$F_{T,CS} = f_{direct} + f_{engineering} + f_{administration} + f_{commissioning} + f_{contingency} \quad (4.2)$$

The total equipment installed cost (*EIC*) can be calculated from equation (4.3):

$$EIC_{CS} = F_{T,CS} * Equipment\ Cost_{CS} \quad (4.3)$$

Then the total installed cost will be the sum of installed cost for all equipment.

$$Total\ installed\ cost = \sum(EIC\ for\ all\ pieces\ of\ equipment) \quad (4.4)$$

If the equipment is to be made of a material other than CS, the installation factor must be adjusted accordingly. The following equation is used to make this correction:

$$F_{T,other\ mat} = \left[F_{T,CS} + \{(f_{mat} - 1)(f_{eq} + f_{pp})\}\right] \quad (4.5)$$

Where the f_{eq} is the equipment factor which is equal to 1, and the f_{pp} is the piping factor in the EDF table sheet.

4.1.3 Adjustments for currency and location

The cost calculations in this study are all done in Euro (€). The cost of the equipment was approximated in Euro using the Aspen In-Plant cost estimator. The currency of the equipment cost is also presented in Euros in the factor table for the EDF-method [56].

As the default location in the Aspen In-Plant cost estimator is Rotterdam city in Netherland the similar location of Rotterdam was assumed.

4.1.4 Cost inflation index

The Aspen In-Plant cost calculator version 12 analyzes equipment costs using figures from 2019. This means that in order to produce an updated and corrected cost estimate, the cost must be adjusted for inflation. The data in the detail factor table used to estimate the installed cost factors are for 2020. This means that the equipment cost must first be adjusted to 2020 cost data. The EDF approach will then be used to calculate the total installation cost. Finally, from 2020 to 2021, the total installed cost must be adjusted for inflation. Table 4.1 [64] lists the indices that were utilized in this project:

Table 4.1: Cost inflation indexes: 2019 - 2021

Year	Cost inflation index
2019	110.1
2020	112.2
2021	116.1

Equation (4.5) has been used to convert the cost from the year a to the year b [12]:

$$Cost_a = Cost_b \left(\frac{Cost\ index_a}{Cost\ index_b}\right) \tag{4.5}$$

All these steps related to calculating the CAPEX for the Base Case were defined in a spreadsheet named CAPEX in the Aspen HYSYS simulation.

4.2 Operating expenditure (OPEX)

The cost of operations and maintenance accounts for a major portion of overall costs. The division of OPEX into fixed and variable costs is a typical policy. Maintenance and operational labor expenses are two fixed expenditures. Maintenance costs are often estimated to be a proportion of the equipment installation cost (EIC) in the range of 2% to 6%, with 4% being used in this study. The cost of operational labor is determined by the number of employees and the number of hours worked during the year. Raw materials, electricity, cooling water, steam, solvents, and other consumables are examples of variable costs. The annual cost of the utilities specified could be calculated from equation (4.6) [22]. The OPEX assumptions and parameters are listed in Table 4.2 [27], [56].

$$Anual\ Utility\ Cost\ \left(\frac{euro}{yr}\right) = Consumption\ \left(\frac{unit}{hr}\right) \times \frac{Operating\ hours}{year} \times utility\ price\ \left(\frac{euro}{unit}\right) \tag{4.6}$$

Table 4.2: OPEX assumptions and specifications [27], [56]

Item	Symbol	Unit	Value
Operating lifetime	n	[Year]	25[1]
Operating Hours p	-	[h/year]	8000
Electricity cost	-	[€/kWh]	0.06
Steam cost	-	[€/kWh]	0.015
Cooling water cost	-	[€/m^3]	0.022

Water process cost	-	[€/m^3]	0.203
MEA cost	-	[€/ton]	1450
Maintenance cost	-	[€/year]	4% of CAPEX
Operator cost	-	[€/year]	80414 (× 6 operators)
Engineer cost	-	[€/year]	156650 (1 engineer)

[1] 2 years construction + 23 years operation

All the calculations and specifications related to OPEX have been done in the OPEX spreadsheet where the consumption of utilities was imported from the simulation results.

5 Method for automatic optimisation

This chapter contains information about methods for optimization and parameters that were varied during the sensitivity analysis to find the optimum cases. A sensitivity analysis of our configuration was performed for the economic evaluation. To explore the influence of various variables on cost, a series of case studies were conducted. The lean/rich amine heat exchanger minimum temperature difference (ΔT_{min}), absorber packing height and the inlet flue gas temperature are the variables.

The major goal of this research is to use the Aspen HYSYS tool to automatically calculate and optimize different parameters based on the cost of a MEA-based CO_2 collecting facility. The CAPEX and OPEX were estimated first for the Base Case in the spreadsheets of these names. The dimensions of the equipment were imported from the Dimensioning spreadsheet to the CAPEX spreadsheet, and relative equipment costs were transferred from the Aspen In-Plant cost estimator, and the total equipment installation cost was determined using the EDF technique. The achieved findings were then entered into the Powe Law spreadsheet to be utilized in the sensitivity analysis as a base for cost calculation. All utility usage was taken from the simulation and entered into the OPEX spreadsheet.

Aspen HYSYS's adjust and recycle blocks were used to automate the energy and material balance for a given configuration. In Aspen HYSYS, recycle blocks are utilized to solve the flowsheet. The in-stream is compared to the block with the stream from the previous iteration in recycle blocks. Adjust functions are used to change a parameter in order to get a certain outcome somewhere else in the simulation [65].

5.1 CO_2 captured cost

To evaluate the different project choices and choose the best one, it's required to calculate the annual CO_2 captured cost for each one, which may be calculated using equation (5.1) [22], [56].

$$CO_2\ Captured\ Cost\ =\ (Total\ Annual\ Cost)/(Mass\ of\ Captured\ CO_2\ \ per\ Year) \quad (5.1)$$

Total annual cost is the sum of the annualized CAPEX and the annualized OPEX.

$$Total\ annual\ cost\ =\ Annualized\ CAPEX\ +\ Annualized\ OPEX \quad (5.2)$$

The annualized OPEX could be computed from the sum of all fixed and variable costs calculated using equation (4.6). Equations (5.3) and (5.4) is used to calculate the annualized CAPEX, the operational lifetime and the discount rate must be determined [27], [56].

$$Annualized\ CAPEX\ \left(\frac{euro}{yr}\right) = \frac{CAPEX}{Annualized\ factor} \quad (5.3)$$

$$Annualized\ factor\ =\ \sum_{i=1}^{n}\left[\frac{1}{(1+r)^n}\right] \quad (5.4)$$

Where *n* is the operative lifetime (for one year construction), and *r* is the discount rate.

All these calculations were conducted in another spreadsheet called “Eff, CO_2 captured cost & reboiler duty”. In this case, the annualized CAPEX and annualized OPEX were imported from the Power Law and OPEX spreadsheets, respectively. The mass of captured CO_2 was extracted from the simulation and calculated for annual plant operation time.

When parameters are adjusted in order to run a sensitivity analysis on the Base Case, the size of the equipment will alter. The equipment cost will be compared to the beginning cost computed by the EDF technique for the Base Case using a spreadsheet titled Power Law and the cost-to-capacity methodology, often known as the Power Law. The Power Law states that changes in equipment size or performance are not always linearly related to costs, but instead costs are a function of capacity multiplied by an exponential ratio. Equation (5.5) can be used to represent this [66]:

$$\frac{Cost\ of\ B}{Cost\ of\ A} = \left(\frac{Capacity\ of\ B}{Capacity\ of\ A}\right)^{e} \tag{5.5}$$

Where e is an exponential size factor that normally ranges in the order of magnitude from 0.35 to 1.70 depending on the type of equipment [67], the exponential factor is assumed to be 1.0 for the absorber and desorber column in this study, and 0.65 for the other of the equipment.

5.2 Sensitivity analysis

Several case studies were done to investigate the impact of various variables on CO_2 captured cost. Using the Case Studies option in the Analysis section of the Home tab of the Aspen HYSYS, it has been attempted to automatically discover the best alternative for each scenario. When a parameter is changed in a process simulation tool like Aspen HYSYS, the usual method is to modify the parameter to a new value and run the simulation again. To reach a converged solution, it is frequently required to make changes to the flowsheet. Another option is to make advantage of Aspen HYSYS's Case Study tool. A set of computations can then be carried out automatically in that instance. Other modifications for each new parameter value are not feasible when utilizing the Case Study tool [65].

A specified Aspen HYSYS model serves as the basis for this technique, and for each parameter a new case should be specified to the model's case studies folder. The model's independent and dependent variables are defined for the new case study. In the setup tab the start point, end point, and step size of the independent variable should be determined. Sensitivity has been chosen as the case study type. The simulation has been completed for the chosen ranges, and the results is provided in the results tab and could be exported in Excel format for more assessments.

A comparison of the automatic and manual outcomes for three separate variable assessments has been given.

5.2.1 The lean/rich amine heat exchanger minimum temperature approach (ΔT_{min})

A case study was undertaken to look into the economic performance of the lean/rich amine heat exchanger when the degree of heat recovery was adjusted. This is measured using the minimum approach temperature (ΔT_{min}). The ΔT_{min} was changed for each scenario by assuming a constant temperature of 120 °C for the reboiler output and changing the temperature of the outlet rich amine from the lean/rich amine heat exchanger. This will happen in the ADJ-2, whereas the ADJ-1 and ADJ-3 will aim to maintain a constant CO_2 removal efficiency of

85% and a constant incoming flue gas temperature of 40 °C, respectively. Temperatures in this assessment ranged from 5 to 20 °C. All flue gas and absorption column parameters were held constant throughout the experiment for a certain total CO_2 removal efficiency, and the same was done for the rate and composition of lean amine flow.

5.2.2 Absorber packing height

To obtain the lowest CO_2 captured cost, the absorber stages were altered from 13 to 18 for this investigation. The pressure drop in the absorber is considered to be a function of the number of stages and associated with a 1 kPa per stage factor (the height of each stage is considered to be 1 meter). This will have an impact on the fan's required pressure increase [68]. This is the same method that Kallevik employed, however he used 0.94 kPa per step [16].

Because each change in the number of stages in the Design tab of the absorber requires running the tower again, the Case Study option cannot be utilized in the sensitivity analysis for altering absorber height (stages), which restricts the capability to create a Case Study for automated evaluation. In all stages, Murphree efficiency has been set to 0.15. In this simulation, Aspen HYSYS gives a value of 1 to new stages, which should be modified to 0.15. For each situation, the efficiency of new stages, the pressure of flue gas into the absorber, and the pressure in the absorber's last stage should all be updated. For this reason, a new spreadsheet was created, and the calculations for changing the absorber and fan outlet pressures based on the number of stages were completed.

In this study, a strategy was employed to define a Case Study by altering the efficiency of a stage. Because each stage is supposed to have a constant efficiency of 0.15, changing the efficiency from 0.15 to 0.9 for a configuration with 13 stages is like to increase the number of stages from 13 to 18. Throughout the case study, the absorber efficiency, lean/rich amine heat exchanger minimum temperature approach, all flue gas parameters and lean amine content were all kept constant. The lean amine feed in ADJ-1, the desorber's input temperature in ADJ-2, the flow rate of inlet cooling water in the inlet heat exchanger in ADJ-3 and the mass balance of makeup MEA and water in the MakeUp Streams spreadsheet had to be adjusted to maintain them consistent.

5.2.3 The inlet flue gas temperature

An analysis was conducted to determine the cost-effectiveness of adjusting the flue gas entrance temperature to the absorber column. This is accomplished in ADJ-3 by altering the cooling water input flow rate in the inlet heat exchanger while maintaining the number of stages in the absorber and, as a result, the absorber inlet and outlet pressures. Throughout the analyses, the lean amine composition was kept constant (by defining the MakeUp Streams spreadsheet), but the lean amine flow rate was modified in ADJ-1 for each sample to acquire the desired CO_2 removal efficiency. The ADJ-2 also regulates the rich amine input temperature to desorber in order to keep constant the ΔT_{min} in the lean /rich amine heat exchanger.

The Murphree efficiency must be adjusted for each new inlet temperature, which makes this analysis difficult. As a result, the research was carried out in combination with adjusting the Murphree stage efficiency to account for the impacts of varying temperature profiles in the absorber column at various input gas temperatures and lean amine flow rates. Øi [15] has created a computational approach for estimating the Murphree stage efficiency. Based on this calculation scheme, the Murphree efficiencies were computed only for the top-, bottom-, and maximum temperature stages and the intermediate stages have been obtained using a linearization between the three known locations, similar to Kallevik's computation technique [16].

The temperature profile of the absorber column changes based on the temperature of the feed gas input, the flow rate (and temperature) of lean amine, and the overall CO_2 removal efficiency [16]. In most cases, the temperature profile in the column peaks somewhere around the top of the column. After calculating the average Murphree efficiency for each inlet flue gas temperature, a relationship between the inlet temperature and efficiency was discovered. Then, another spreadsheet was created to export the calculated stages efficiency to the absorber after changing the incoming flue gas temperature from 30 to 50 degree Celsius in the Case Study.

6 Results

This chapter will provide the results for the Base Case and sensitivity analysis outcomes for each case study, as well as a comparison of the automated and manual calculation results for sensitivity analyses.

6.1 Base Case evaluation

Figure 6.1 depicts the CO_2 capture plant's equipment costs in the Base Case study. The overall equipment cost is around 110 MEUR, and the absorber is clearly the costliest equipment, accounting for about 54% of the total cost. It's worth noting that the packaging cost accounts for around 55 percent of the absorber's total cost.

The lean/rich amine heat exchanger, reboiler and fan are the other high-cost components, accounting for 22%, 8% and 8% of the overall cost, respectively.

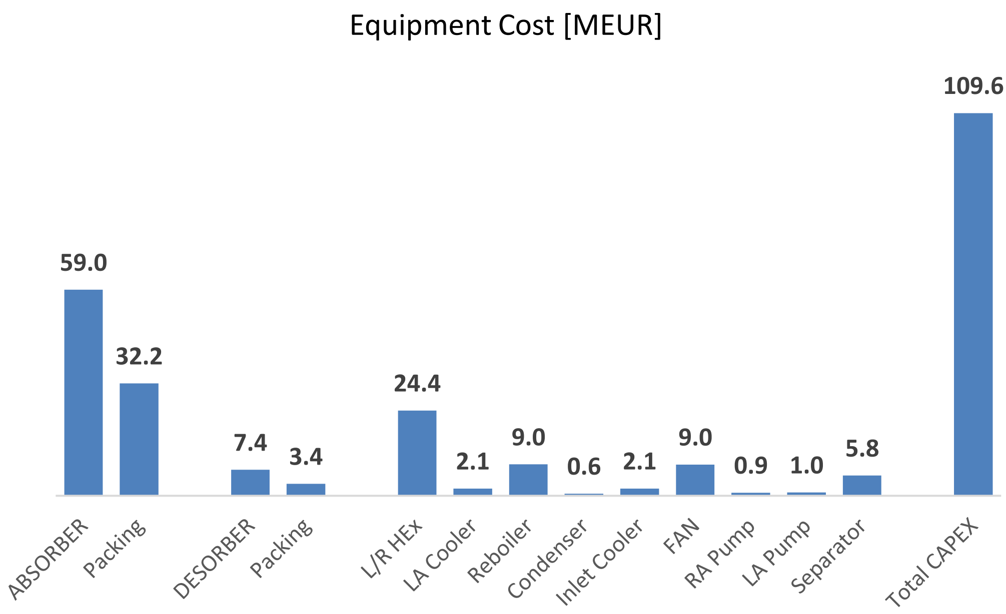

Figure 6.1: Equipment cost for CO_2 capture plant

According to figure 6.2 the total operational expenditure (OPEX) for the Base Case is about 29 MEUR/y. Steam is the costliest utility for this facility, costing approximately 15 MEUR each year. This accounts for around 52% of overall OPEX. The steam usage based on reboiler duty per each kilogram of captured CO_2 has been computed for each case of the sensitivity analysis.

Apart from steam, the other expensive cost components are MEA, maintenance and electricity as depicted in Figure 6.2.

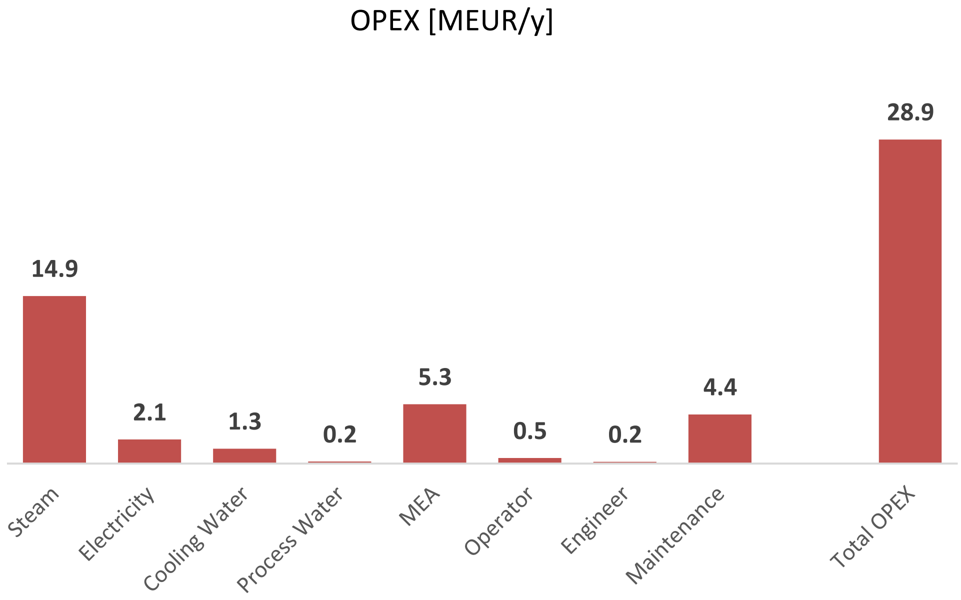

Figure 6.2: Operation cost for the Base Case study

The predicted cost of removing 1 ton of CO_2 from the flue gas in the Base Case analysis was 42.85 EUR, whereas the reboiler's duty was 3750 kJ for 1kg captured CO_2.

6.2 ΔT_{min} in the lean/rich amine heat exchanger

Figure 6.3 depicts the calculations performed during the sensitivity study for the lean/rich amine heat exchanger's minimum temperature approach (ΔT_{min}). The reboiler duty and captured CO_2 cost for altering the temperature from 5 to 20 degrees Celsius was evaluated in this study, with an initial point of 10 degrees Celsius in the Case Study.

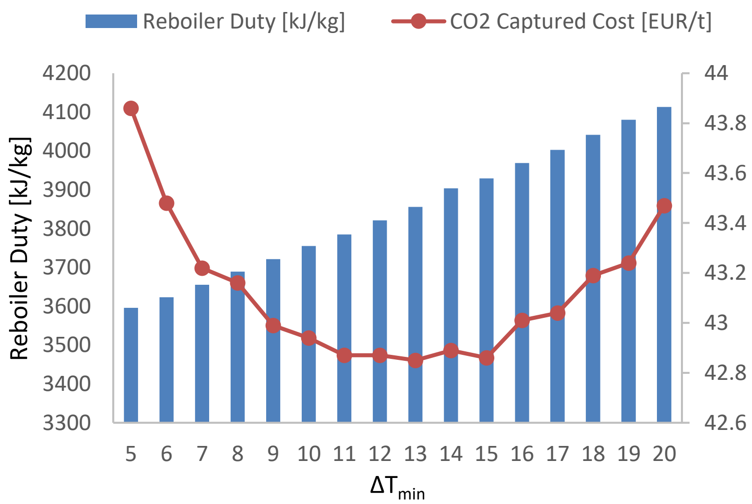

Figure 6.3: Automated calculation for the minimum temperature approach, ΔT_{min} (Initial point: ΔT_{min} = 10 °C)

From this automated calculation the optimum ΔT_{min} was 13 °C, where the captured cost is 42.85 EUR/t and reboiler's duty is 3856 kJ/kg. The figure shows that by increasing the ΔT_{min} the consumption of steam goes up steadily. For the manual calculation, the same result has been displayed in figure 6.4. It's worth noting that just the target value in the ADJ-2 was altered during the manual calculation, while all other Adjust and Recycle blocks remained active, allowing the calculation for each step to be completed automatically. For the manual assessment, the optimum ΔT_{min} was calculated at 12 °C with capturing cost of 42.73 EUR/t and reboiler's duty equal to 3856 kJ/kg. It is obvious that in both automated and manual cases the change in the cost from 11 to 15 degrees Celsius is negligible.

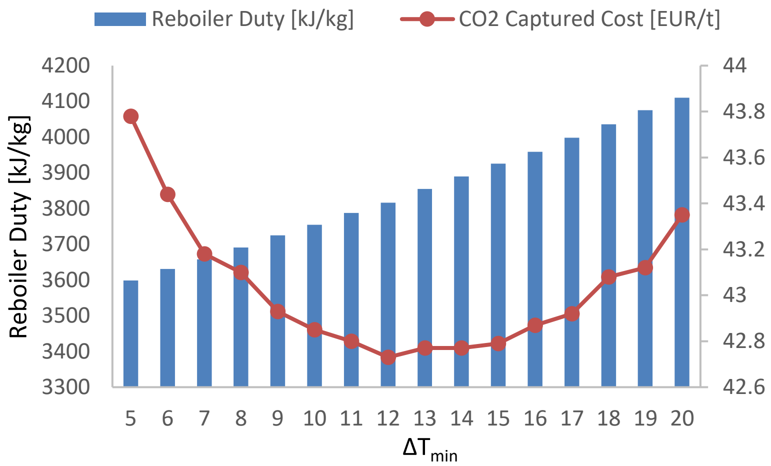

Figure 6.4: Manual calculation for the minimum temperature approach, ΔT_{min} (Initial point: ΔT_{min} = 10 °C)

Figure 6.5 illustrates a discrepancy between the obtained results for the manual and automated calculation while the starting point in the automated case has been 10 °C. The calculated cost in the automated case is about 0.3 % higher than the manual one.

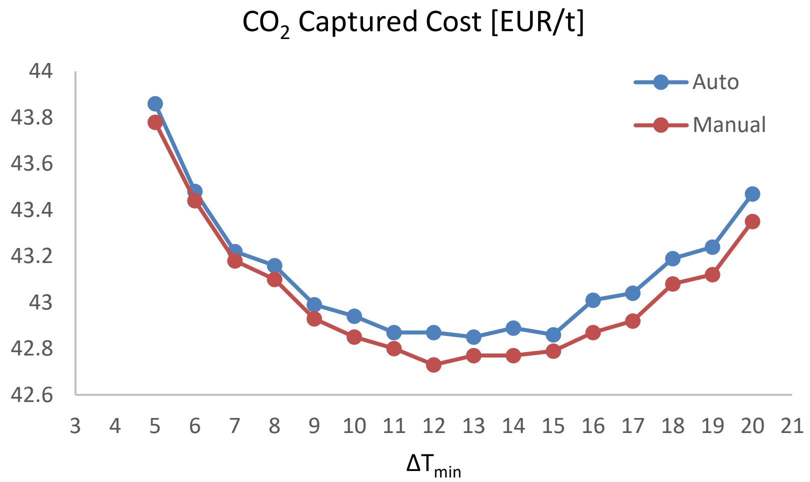

Figure 6.5: Comparison between manual and automated calculation for the minimum temperature approach, ΔT_{min} (Starting point: ΔT_{min} = 10 °C)

In another assessment for the optimum ΔT_{min}, the automated calculation has been done while the starting point was 5 °C, then the result has been compared to the manual evaluation. Figures 6.6 and 6.7 depicts the result for these cases.

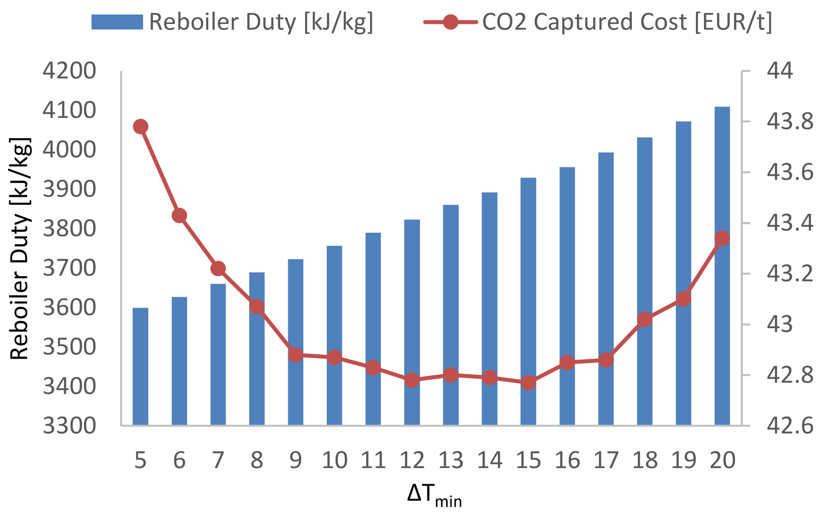

Figure 6.6: Automated calculation for the minimum temperature approach, ΔT_{min} (Starting point: ΔT_{min} = 5 °C)

Figure 6.6 represents that the minimum calculated captured CO_2 cost is almost constant while the ΔT_{min} varies between 12 and 15 degrees Celsius. This equates to around 42.78 EUR/ton.

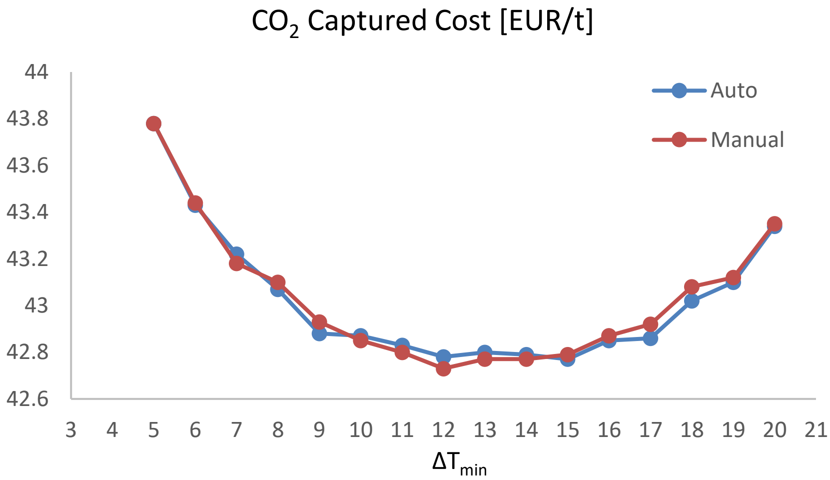

Figure 6.7: Comparison between manual and automated calculation for the minimum temperature approach, ΔT_{min} (Starting point: ΔT_{min} = 5 °C)

The gap between the automatic and manual outcomes in this study is lower than the automated assessment when the starting point is 10 °C, as shown in Figure 6.7.

6.3 Absorber packing height

Figure 6.8 depicts the change in reboiler duty and CO_2 captured cost when the absorber packing height is manually altered from 13 to 18 meters. In this analysis the number of stages in the absorber was modified, and for the added stages the Murphree efficiency for CO_2 was set to 0.15. The amine solution flow rate has been adjusted to reduce the convergence time. All the Adjust and Recycle blocks remained active during the simulation to execute computations automatically. According to the manual research, 15 meters of packing height is the best scenario, with a CO_2 capture cost of 42.64 EUR/t.

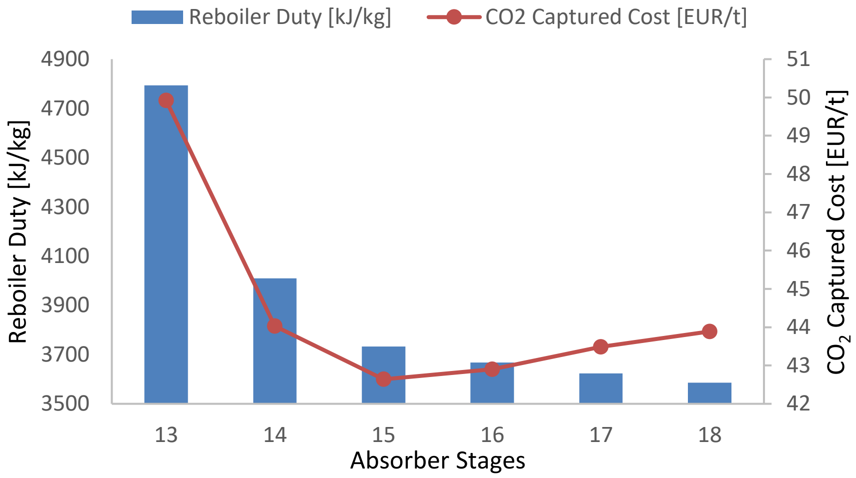

Figure 6.8: Manual calculation for the absorber packing height

In terms of automation, a 13-stage absorber was considered in the initial evaluation, and the efficiency of the last stage was changed from 0.15 to 0.9 throughout the Case Study. For the CAPEX computation, another spreadsheet was created to specify the changes in packing height and outlet pressure of the absorber and fan based on the chosen stage efficiency. Figure 6.9 illustrates the result for this analysis.

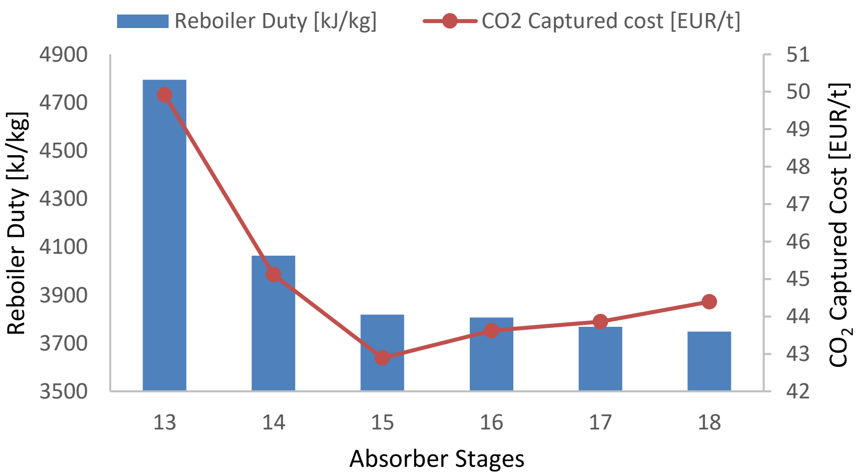

Figure 6.9: Automated calculation for the absorber packing height (change of efficiency in the 13^{th} stage)

The outcome of the manual and automatic calculations is shown in Figure 6.10. In this situation, the automated calculated costs are around 1.5 % more on average than the manual technique.

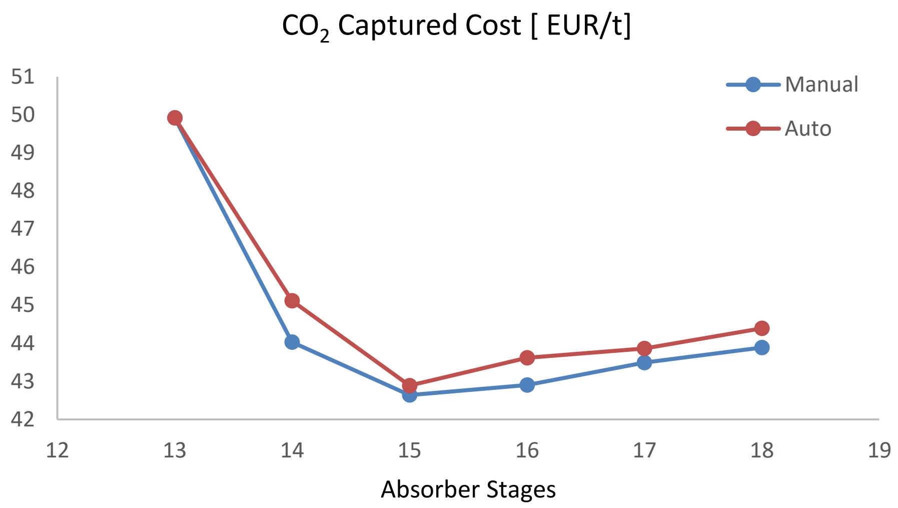

Figure 6.10: Comparison between automated and manual calculation for the absorber packing height (change of efficiency in the 13th stage)

Figure 6.11 demonstrates a similar result for automated computation, with the 10th stage of the absorber as the target stage for modifying the efficiency.

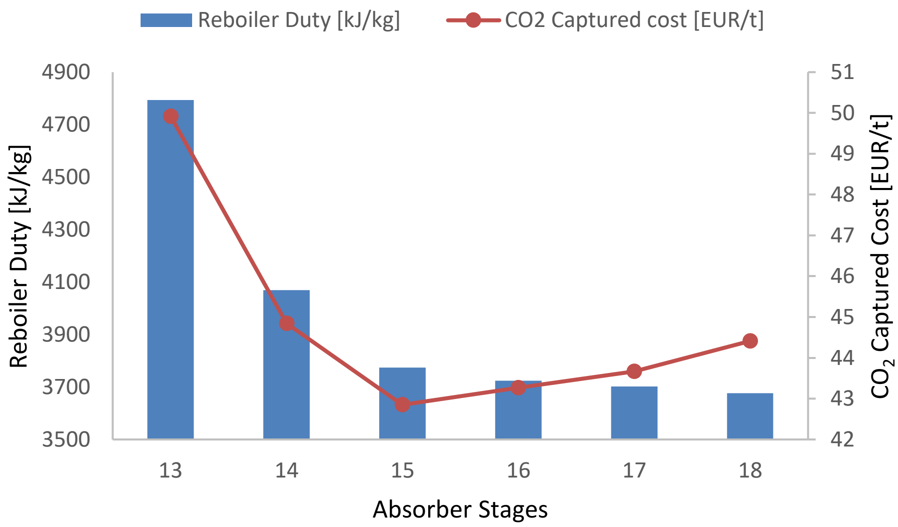

Figure 6.11: Automated calculation for the absorber packing height (change of efficiency in the 10th stage)

In this case like previous one, 15-stage was calculated as the optimum absorber packing height. Comparison between the automated and manual results in figure 6.12 indicates that the calculated costs for automatic approach are on average 0.9 % higher than the manual assessment.

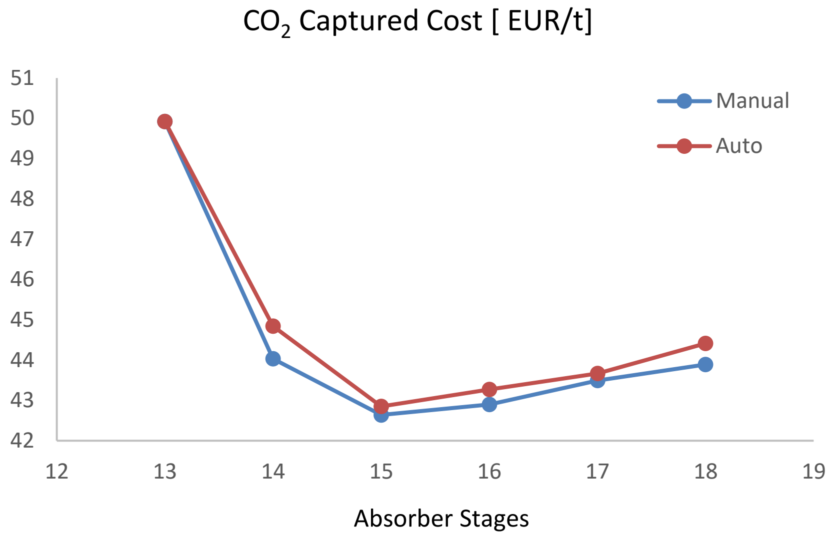

Figure 6.12: Comparison between automated and manual calculation for the absorber packing height (change of efficiency in the 10th stage)

6.4 The inlet flue gas temperature

The distribution of temperature and efficiency for the stages of the absorber is depicted in Figure 6.13. The dotted lines indicate the efficiency of the stages, which are based on three computed efficiencies.

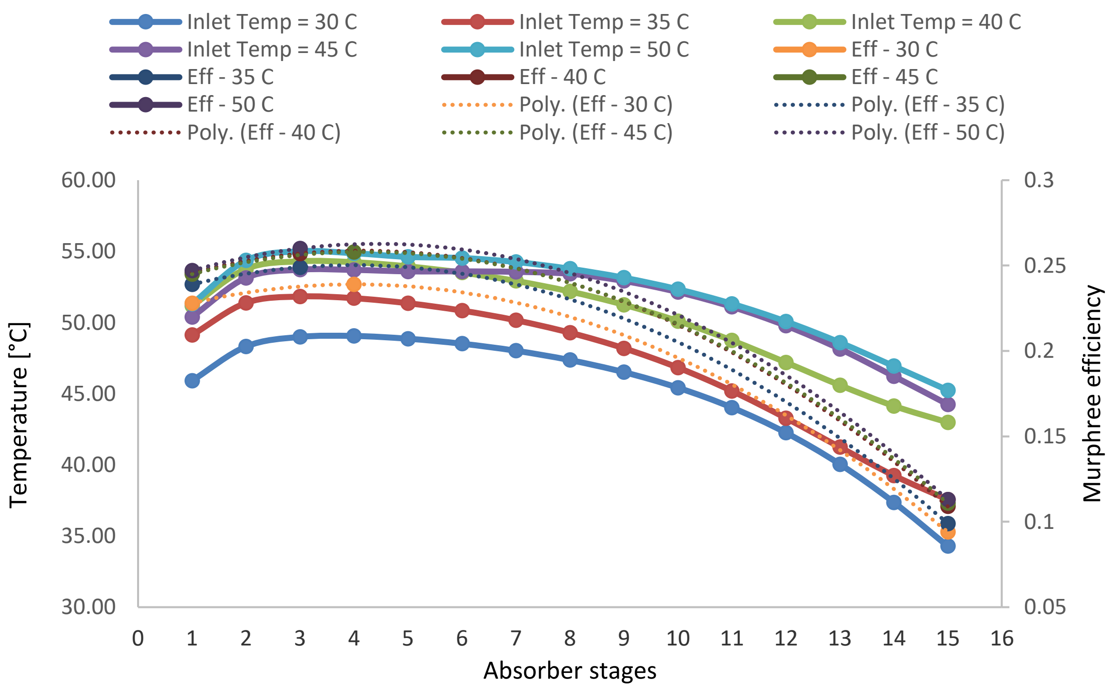

Figure 6.13: Distribution of temperature and efficiency in the absorber based on inlet temperature (First trial)

For this assessment, the efficiency must first be estimated for each inlet temperature. So that, based on Øi's calculation scheme [15] and Kallevik 's computing technique [16], the average efficiency for the inlet temperatures 30, 35, 40, 45, and 50 degrees Celsius was computed for the first trial according to stage temperatures for the Base Case simulation. Table 6.1 tabulated the average computed efficiencies for each inlet temperature.

Table 6.1: Average Murphree efficiency in the absorber based on inlet temperature (First trial)

Inlet Temperature [°C]	Average Murphree Efficiency
30	0.1665
35	0.1740
40	0.1830
45	0.1845
50	0.1865

A new computation was performed based on the initial trial data to adjust the anticipated efficiency. This implies that in the absorber calculation, instead of utilizing 0.15 as an average efficiency, the average efficiencies from Table 6.1 were used. Figure 6.14 and Table 6.2 display the results.

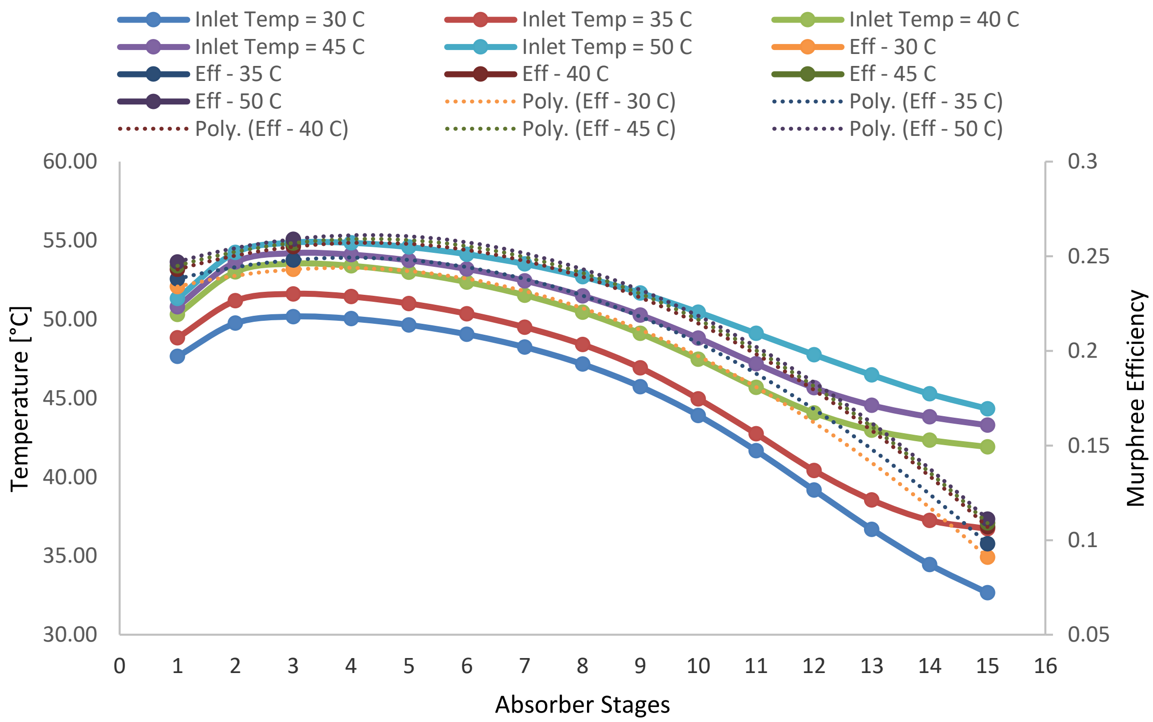

Figure 6.14: Distribution of temperature and efficiency in the absorber based on inlet temperature (Second trial)

Table 6.2: Average Murphree efficiency in the absorber based on inlet temperature (Second trial)

Inlet Temperature [°C]	Average Murphree Efficiency
30	0.167
35	0.173
40	0.181
45	0.183
50	0.185

A link between the inlet temperatures and their average Murphree efficiency must now be established for the automated computation. Figure 6.15 shows an equation that has been utilized in a separate Aspen HYSYS spreadsheet to change the stages efficiency after modifying the input temperature during the Case Study.

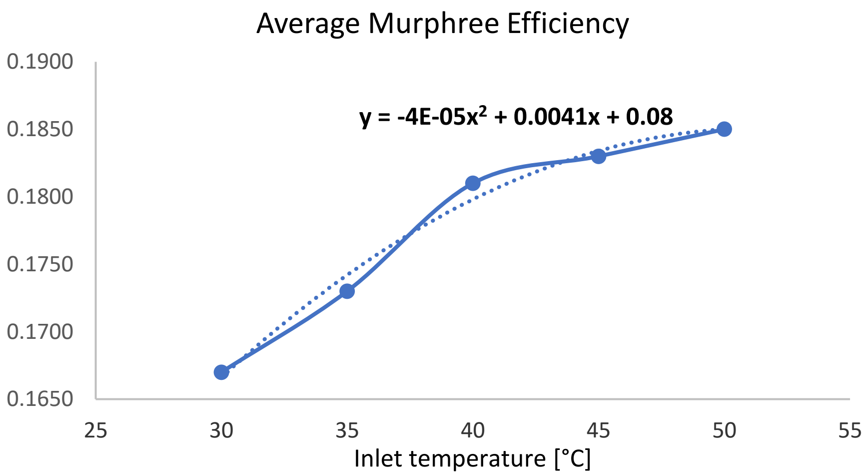

Figure 6.15: Average Murphree efficiency for different absorber' inlet temperature

The predicted CO_2 captured cost and reboiler's duty for the manual analysis are shown in Figure 6.16. For this evaluation, in a scenario identical to the base case except the stages' efficiency in the absorber, the target value in the ADJ-3 was altered to modify the flue gas input temperature to the absorber and the rest of the calculations were done automatically. The graph indicates a significant increase in both cost and duty as the inlet temperature rises from 30 to 40 degrees Celsius, followed by a modest increase till 45 °C, and then a nearly flat trend until 50 °C. Because the stages are more efficient in this research than in the Base Case, the flow rate of the lean amine solvent circulation is lower. As a result, both cost and duty are lower than in the Base Case.

A Case Study was established with a start point of 30 °C and an end point of 50 °C, with a step size of 5 °C, for automatic assessment in a similar simulated case with an initial flue gas input temperature of 40 °C. The outcome of this investigation is shown in Figure 6.17. Converging this study, especially for the 30 and 35 degrees Celsius, was too complex and demanding. It was contributing to converge this investigation by switching the boundary from 50 to 30 degrees Celsius.

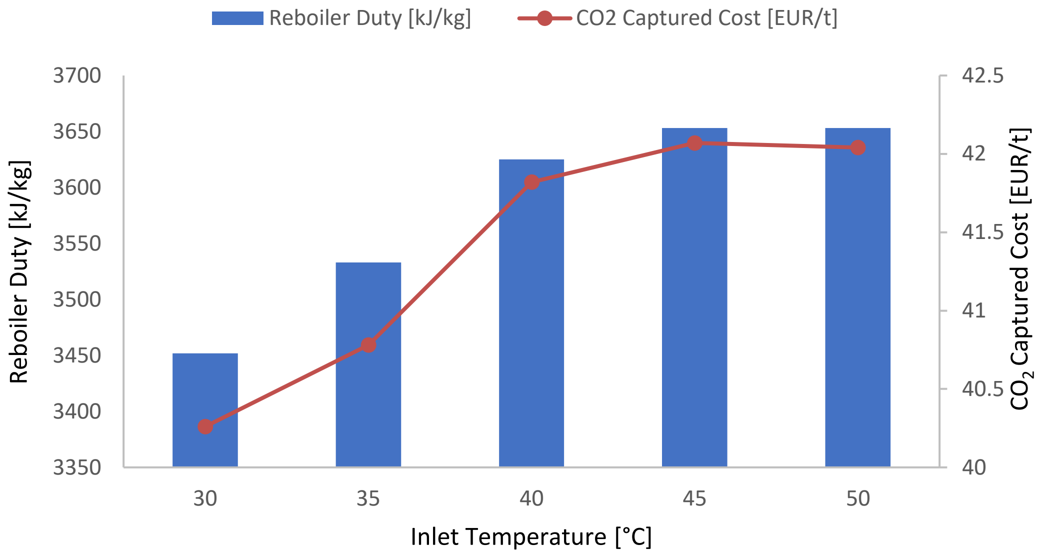

Figure 6.16: Manual calculation for the flue gas inlet temperature into the absorber (15 stages in the absorber)

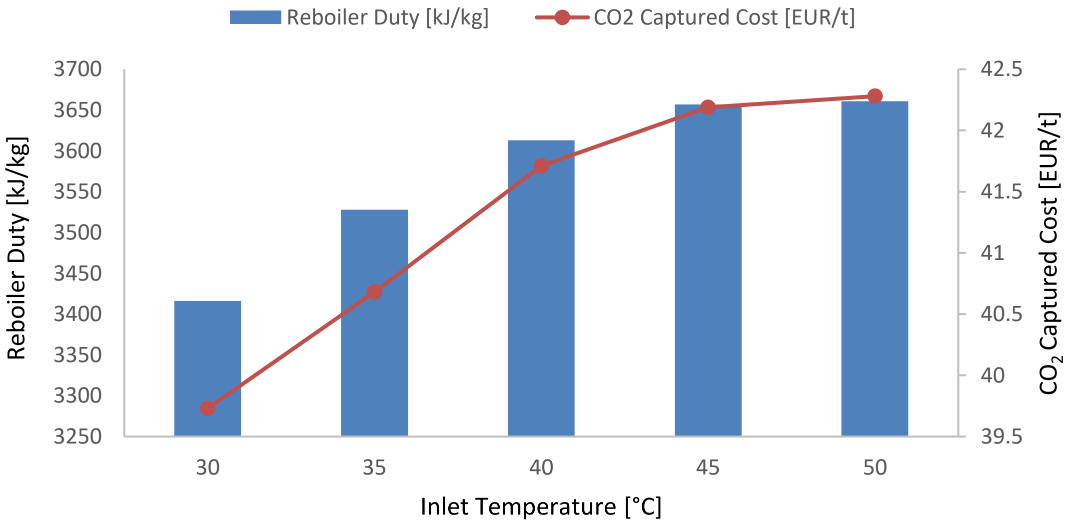

Figure 6.17: Automatic calculation for the flue gas inlet temperature into the absorber (15 stages in the absorber)

Figure 6.18 compares the cost of CO_2 captured for manual and automated calculations. In this scenario, both the auto and manual lines follow the same pattern, with the difference for the start and finish points between both techniques being greater than the difference between the midway points.

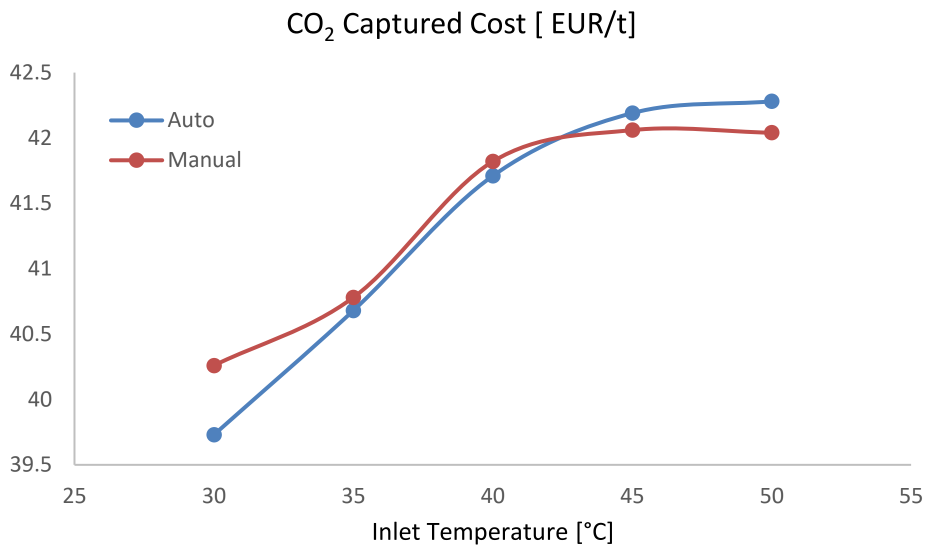

Figure 6.18: Comparison between Manual and Automatic calculation for the flue gas inlet temperature into the absorber (15 stages in the absorber)

It appears that by improving the effectiveness of the stages, a similar rate of CO_2 removal may be achieved with a lower-stage absorber. As a result, a comparable investigation was conducted for a simulated case by 13 stages in the absorber. The outcome of the manual computation is shown in Figure 6.19.

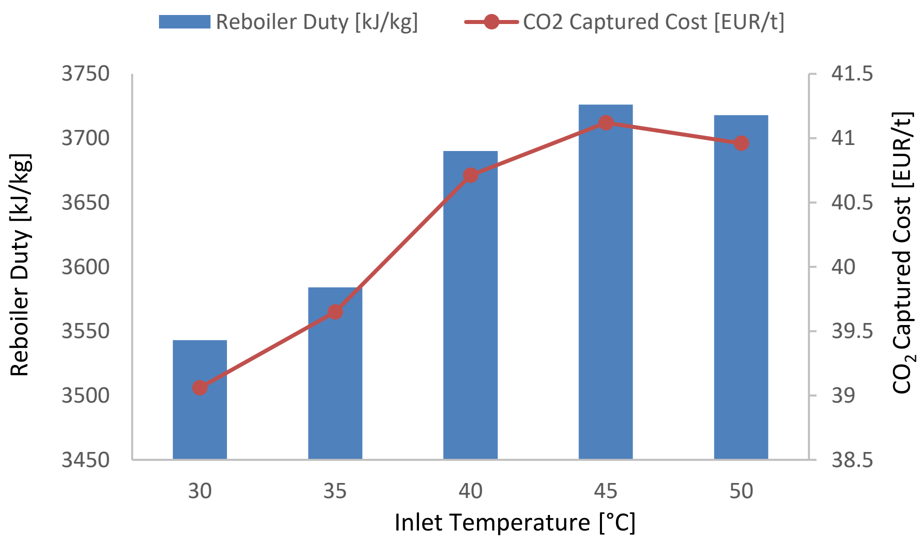

Figure 6.19: Manual calculation for the flue gas inlet temperature into the absorber (13 stages in the absorber)

Although Figure 6.19 and Figure 6.16 show the identical pattern, the computed cost for setting with 13 stages in the absorber is lower. For instance, the CO_2 collected cost in this scenario for an input temperature of 40°C is 1 EUR/t less than in the prior case with a packing height of 15 meter in the absorber. On the other hand, by reducing the number of stages, higher amine solvent circulation flow rate is required to remove the same amount of CO_2, therefore convergence in automated circumstances may be easier and faster. Figure 6.20 depicts the result for the automated simulation, while Figure 6.21 shows how it compares to the manual evaluation outcomes.

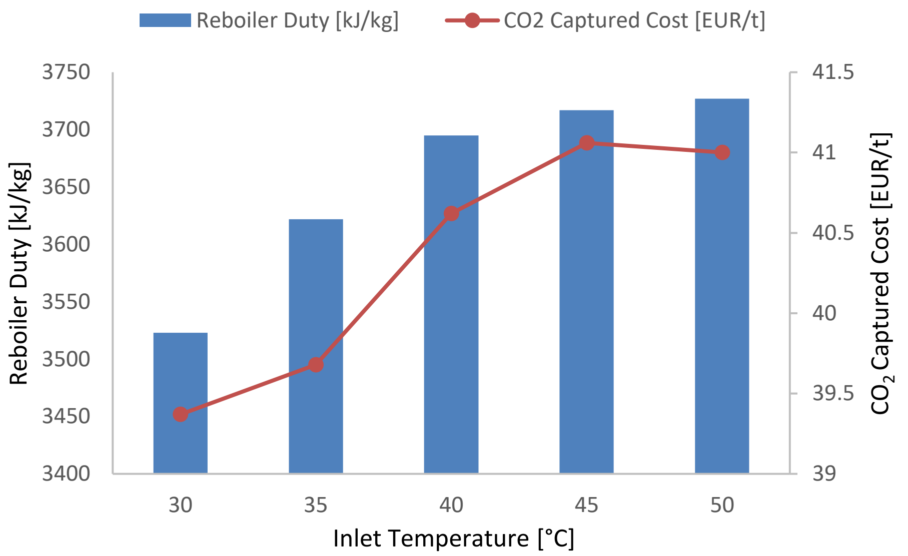

Figure 6.20: Automatic calculation for the flue gas inlet temperature into the absorber (13 stages in the absorber)

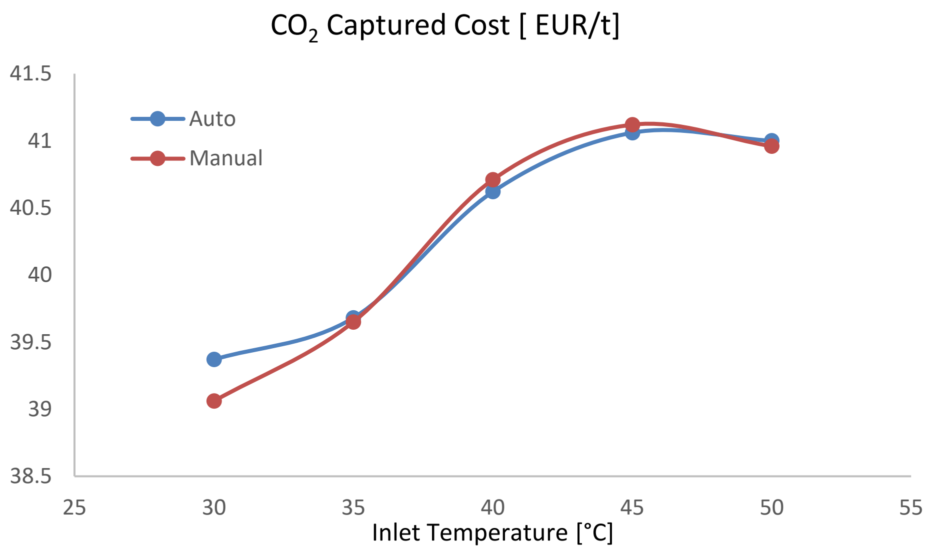

Figure 6.21: Comparison between Manual and Automatic calculation for the flue gas inlet temperature into the absorber (13 stages in the absorber)

When a 13-stage absorber was employed instead of a 15-stage absorber, the correspondence between the manual and automated calculation findings was considerably more obvious.

Another study attempted to automatically calculate the cost and duty for a one-degree Celsius change in the absorber's inlet temperature. The problem could not be addressed, and convergence did not occur in the 15-stage absorber scenario. However, in the 13-stage absorber situation, since a greater amine circulation rate was required, the simulation got converged, and the result is given in figure 6.22.

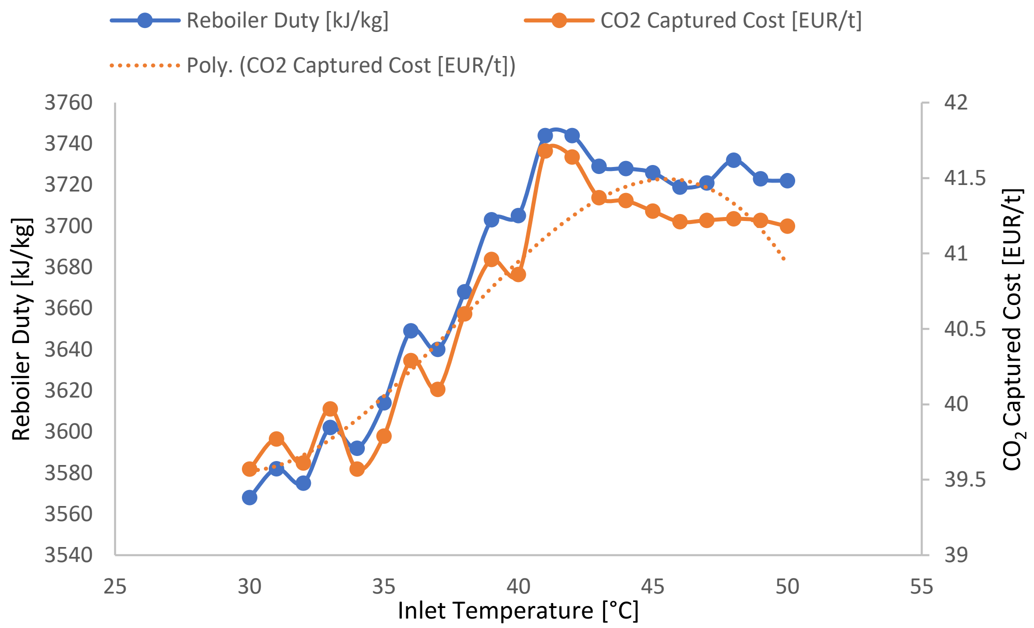

Figure 6.22: Automatic calculation for the flue gas inlet temperature into the 13-stage absorber case for the step size equal to 1 °C.

Both the reboiler's duty and the captured CO_2 cost follow a nearly identical trend in the graph. Although the computed results not being smooth. The dotted line indicates a linearization of the CO_2 captured cost curve, which has the same pattern as the findings in Figure 6.20 where the step size of evaluation in the Case Study was 5 °C.

7 Discussion

In this chapter, the results reported in the previous chapter will be discussed and compared to comparable research. The accuracy and uncertainty of the findings will be debated, and some recommendations for further research will be offered.

7.1 Comparison with previous studies

The overall cost of 40-43 EUR/t in this study is lower than the 50 EUR/t stated in certain publications such as Ali[22] and Aromada [27], [56]. However, the reduced calculated cost in this study might be owing to certain DCC unit simplification and the omission of CO_2 compression. Also, it is almost similar to the 39-40 EUR per ton CO_2 captured reported by Øi[65].

7.1.1 Base Case outcomes

Table 7.1 shows a comparison of the Base Case findings of this investigation with some of the previous studies. In their simulations, all of these researches employed a 29-30 (W %) MEA solvent. The concentration of CO_2 inflow changes due to variations in the flue gases of the plant (NGCC power plant, Cement and Waste incineration).

Table 7.1: Comparison of simulation results with literature

Study	CO_2 Capture Rate [%]	CO_2 Concentration [mol %]	ΔT_{min} [°C]	Absorber packing Height [m]	Reboiler Duty [kJ/kg]
This work (Base Case)	85	3.75	10	15	3750
Ali et al.,[22]	90	22 - 28	10	15	3970
Aromada et al., [20]	85	3.73	10	20	3600
Øi et al., [24]	90	17.8	10	12	3500
Amrollahi et al., [69]	90	3.8	8.5	13	3740
Sipöcz et al., [70]	90	4.2	10	26.9*	3930
Nwaoha et al., [38]	90	11.5	10	22 (36 Stages)	3860
Shirdel et al., [68]	90	7.5	10	20	3654
Øi et al.,[14]	85	3,75	10	10	3650

* Not defined whether it is packing height or total column height

7.1.2 Minimum temperature approach (ΔT_{min}) in the lean/rich heat exchanger

This research is a trade-off between the area of the lean/rich heat exchanger and the exterior utility needs. The area of the heat exchanger changes as ΔT_{min} is altered. The most important trade-off is between the capital cost of changing the size of the lean/rich heat exchanger and the operational cost of the steam consumption variation. This, however, has an impact on the temperature entering the desorber as well as the cooling required in the lean amine heat exchanger. According to Øi[15], a low ΔT_{min} will minimize steam and reboiler duty. A high ΔT_{min}, on the other hand, will lower the heat exchanger cost.

The study's major focus was on automated calculating. In the lean/rich amine heat exchanger, it appears that when the current point in the Case Study evaluation is similar to the start point (5 °C), the findings are better aligned with the manual assessment than when the current point is equivalent to 10 °C (as the Base Case).

During the several assessments in this study, the optimal determined ΔT_{min} in the lean/rich amine heat exchanger varied between 11 and 15 degrees Celsius. The point at which the manual assessment's minimal cost was computed was 12 °C, while the optimum case was computed at 13 °C for the automated assessment with a beginning point of 10 °C, when the current value in the Case Study was modified to 5 °C, 15 °C was estimated as the best case. It is worth noting that the difference in predicted CO_2 removal cost from 11 to 15 degrees Celsius is practically insignificant in all these situations.

According to Øi et al. [24], by changing the condition the optimum temperature approach will changed slightly between 10 to 15 degrees Celsius. Especially 13 °C has been reported as the optimal case in several related studies [16], [24], [27], [65], [68].

7.1.3 Absorber packing height

This research looks at the relationship between amine flow rate and absorber packing height, with the solvent flow rate increasing as the number of absorber stages reduces and vice versa. The size of the absorber and the heat transfer area in the lean/rich heat exchanger are two key capital expenses that are influenced by this shift. Because of the change in amine flow, the cost of a lean/rich heat exchanger is influenced. Fan power consumption, steam consumption, and amine consumption are the three main operating costs influenced by this adjustment. Since the pressure should be modified to indicate pressure loss across the packing, the fan power consumption fluctuates. The amine flow change affects the steam consumption, and the amine consumption is likely impacted by varied temperatures, pressures, and amine flow rate.

In this investigation, the calculated optimum packing height of the absorber was 15 stages, which corresponds to Kallevik [16], Aromada [20] and Øi[65]. With a CO_2 level of 17.8 mole%, Øi et al., [24] determined the minimum at 12 meters absorber packing height for 90% removal effectiveness. When compared to the outcome in this study, there is a great disparity. According to Husebye et al., [51] a rise in CO_2 concentration particularly between 2.5 and 10% leads to a decrease in cost. This is due to increased CO_2 transfer, which requires less solvent, and so a reduced number of steps would be predicted.

In terms of steam consumption, manual and automatic assessments show that as the number of stages grows, reboiler duty falls. This is due to a reduction in amine circulation rate, which drops sharply from 13 to 14 stages before gradually declining in the absorber with additional stages. On the other hand, in the manual case, the average calculated reboiler duty per removing 1 kilogram of CO_2 is lower than in the first automated assessment where the efficiency of the 13^{th} stage was changed, whereas these figures are closer when the efficiency of the 10^{th} stage was changed during the Case Study in automatic evaluation for the same reason as the reduction of the amine solvent circulation in the second case.

7.1.4 The inlet flue gas temperature

The amine circulation on the one hand, and the cost of the DCC unit (in this study inlet cooler and separator) and columns diameter on the other, are the two key cost drivers in these case studies. Increased inlet temperature reduces the size and duty of the DCC unit while increasing the size of the absorber column and increasing the amine flow rate. The flow of amine for a certain number of stages must increase as the flue gas temperature rises, despite the fact that the temperature rises, improving the stage efficiency. The real gas flow rate increases as the absorber input gas temperature rises. So that, the column size and installed cost will grow as the vertical velocity in the absorber column is assumed to be constant [16].

The major justifications for a low temperature are higher CO_2 solubility at equilibrium and reduced amine evaporation propensity, whereas the key arguments for a high temperature are lower DCC cost, higher reaction rate, and lower viscosity [15].

The absorber's maximum temperatures were usually reached in stages 2 or 3 (Figures 6.13 and 6.14). The first test was carried out on a 15-stage absorber. The computed result for captured cost is around 2% cheaper owing to utilizing lower amine flow rate by raising the average efficiency rather than the Base Case study, while the second assessment by a 13-stage absorber is twice lower (about 4% cheaper) due to the reduction in absorber size.

The obtained result shows that raising the input flue gas temperature from 30 to 40 degrees Celsius causes a significant increase in both reboiler duty and cost, followed by a slight increase to 45 °C and finally a negligible reduction to 50 degrees Celsius. Øi [14] obtained an almost identical pattern for heat consumption. While Kallevik [16] determined that with an 85% capture rate, 40 °C was the best situation, however, for a 90 % removal rate, temperature rises increased both cost and reboiler duty.

Although the convergence was too difficult for Case Study with 1 °C step size (Figure 6.22) and the obtained results are not smooth, the optimum inlet temperature in this case calculated as 34 °C which correspond to what is reported by Øi[15] who stated that the optimum case could be between 33 and 35 degrees Celsius.

7.2 Accuracy and uncertainties

Several aspects in the study, primarily linked to the simulation, dimensioning, and cost estimating assumptions, contribute to uncertainties:

- The goal of this work's cost estimates is to determine the best process parameters, not to estimate the absolute values of these estimates as precisely as possible. The uncertainty of the computed costs is influenced by probable discrepancies in various specifications. Even larger disparities exist in the estimated values for entire project investments, including land, utility systems, and other considerations. The evaluation of these variations is outside the scope of this project.
- Large uncertainties are predicted in the cost assessment of process equipment linked to the base equipment cost and the validity of the scale up factors in the stated range. Using a cost estimating tool like Aspen In-Plan Cost Estimator will most likely result in an acceptable equipment cost. However, the number of parameters used as inputs has an impact on the estimate's accuracy. The size exponent used for equipment scaling was

0.65 (1.0 for columns), and the installation factor for each piece of equipment was maintained constant, introducing uncertainty in cost scaling and affecting the comparability of different process parameters.

- Energy expenditure, particularly heat consumption for CO_2 regeneration, has the greatest impact on operational costs. The uncertainty in overall operating costs is nearly equivalent to the uncertainty in this heat's value.
- Choosing certain assumptions and specifications may have an impact on the estimated cost and, as a result, the optimal values. For example, the cost and height of the absorber and desorber are affected by the type of packing used. The ΔT_{min} calculations are influenced by specified characteristics such as the overall heat transfer coefficient of the heat exchangers, notably the lean/rich amine HEx. In addition, the fan outlet pressure and price were impacted by the pressure drop assumption in the absorber.
- The simplification of the constant Murphree efficiency in the absorber and desorber may have an impact on the calculations, causing inaccuracies in the amine circulation flow and, therefore, the plant's predicted cost. Likewise, the parametric analysis for the change in inlet flue gas temperature, raising the average efficiency of stages in the absorber reduced amine flow rate and overall cost.
- The most significant uncertainties in the simulation findings are most likely connected to parametric studies, particularly for absorber packing height and inlet flue gas temperature, as they need manual or iterative adjustments (ADJ-1) in solvent flow to obtain a certain capture rate.
- Convergence is difficult to obtain, particularly in the absorber and desorber; in this situation, using the modified HYSIM Inside-Out solver may be beneficial for easier and faster convergence; nevertheless, this made it impossible to fully automate the case studies, particularly for the optimal packing height. Automatically adjusting the appropriate amine circulation flow rate to achieve a certain capture rate, especially for a 13-stage absorber that requires greater circulation flow, is too difficult. In the research, default convergence criteria were first employed, and subsequently tolerance and sensitivity in the adjust and recycle blocks were reduced to achieve higher accuracy and smoother results.

7.3 Limitations in the models

The technique taken in this thesis was to optimize one parameter at a time while maintaining the other values constant. Automatic assessment demands the use of Adjusts and Recycle blocks. When a tight convergence tolerance is used, it frequently necessitates more iterations than the default limit of 10 (maximum iteration was set to 100 and for some cases 500). When the number of iterations in one operation in the flowsheet is increased, the number of iterations in the other operations must likewise be increased. As a result, combining numerous adjust and recycle operators with precise tolerances and sensitivity extends the computation time and increases the likelihood of divergence. On the other hand, it might help create a smoother outcome, such as in the Case Study for determining the best ΔT_{min} in a lean/rich amine heat exchanger (ADJ-2, RCY-1).

Convergence was difficult to accomplish specially in the adjust block for capture efficiency (ADJ-1) when the beginning values were away from the intended values. Reducing the number of stages from 14 to 13 resulted in a significant drop in capture efficiency. In addition, compared to 14 steps, the needed lean MEA flow rate to boost capture efficiency was substantially higher. This makes using adjust operators with constant values to achieve convergence for a large range of parameters difficult. Manual iteration of lean MEA flow will likely take less time since the flow may be gradually reduced until it reaches the target capture Efficiency. A mix of manual iteration and the usage of an adjust operator is a possibility.

Another issue during automated evaluation is to solve all adjust and recycle blocks at the same time. If there are several adjust or recycle unit operations, you may need to designate which adjust operation should converge first. If you discover that adjustments are competing with each other and the flowsheet is not converging (or converging very slowly), this might be helpful [71]. On the "Calculation Order" window of the "Solver" option in the "Home" tab, the calculation levels could be altered. For the input flue gas temperature assessment, the calculation level of ADJ-3 was changed to 3000, allowing the inlet temperature to be computed first and then the other blocks to be converged.

7.4 Future work

Some suggestions for future studies in this area to improve the robustness and accuracy of simulation and cost estimation are as follows:

- Because the lean/rich HEx is one of the study's major CAPEX contributors, a plate heat exchanger (PHE) arrangement might be a more cost-effective option because it is smaller and has a higher heat transfer coefficient [23], [27]. By using this type of heat exchanger instead of a shell and tube heat exchanger (STHE), cost estimates might be more exact.
- The efficiency of the stages was utilized to automatically optimize the packing height (number of stages) in the absorption column in this study. Providing a method to change the number of stages directly for this analysis might be beneficial. However, it must be possible to update the number of stages during the simulations without pressing the "run" button on the column faceplate. There are other challenges in terms of column and flowsheet convergence issues that must be carefully addressed in order to produce reliable and accurate results.
- Evaluate automatic optimization using automatic return of values calculated in spreadsheets to the simulation. Using Excel and Visual Basic in conjunction with Aspen HYSYS spreadsheets to conduct computations that are not available directly in Aspen HYSYS, and to exchange results with the simulation for more advanced iteration or optimization.
- Evaluate automatic optimization using a rate-based approach
- Because of the frequent challenge of reaching convergence, the strategy used in this thesis was to optimize one parameter at a time while keeping the other parameters constant. An overall optimization for all process parameters in order to identify the least expensive option by modifying several process parameters at the same time might be a proposal to estimate the best arrangement.

8 Conclusion

Aspen HYSYS was used to simulate an amine-based CO_2 capture process, utilizing emission data from previous research on a natural gas-based power plant project at Mongstad, Norway. The cost assessment is based on the CO_2 removal process' CAPEX and OPEX estimations. A base scenario with 15m packing height in the absorber, 10m packing height in the desorber, removal efficiency of 85%, and a minimal temperature differential (ΔT_{min}) in the lean/rich amine heat exchanger of 10 °C was designed to perform cost optimization. The Enhanced Detailed Factor (EDF) was used in conjunction with the Aspen In-Plant Cost calculator to estimate and compute the overall cost for the base scenario. Findings for the Base Case show a total cost of 42.85 EUR per ton CO_2 captured and 3750 kJ/kg energy usage in the reboiler.

To reduce and lower expenses, a sensitivity analysis was utilized to do cost optimization. In a series of case studies, the method was used to evaluate how different variables caused price changes. The Power-Law technique was used to modify the equipment cost when the sensitivity analysis changes the size. The absorber packing height, the minimum temperature approach in the lean/rich amine heat exchanger, and the entering flue gas temperature into the absorber column were all changed in this study. The major goal of this research was to use the Aspen HYSYS tool (Case Study) to automatically calculate and optimize the cost of CO_2 capture.

The ΔT_{min} of the lean/rich heat exchanger sensitivity analysis is a trade-off between the area of the lean/rich heat exchanger and the external utility demands. The capital cost of adjusting the size of the lean/rich heat exchanger versus the operating cost of the steam consumption variation is the most critical trade-off. A Case Study option in Aspen HYSYS was used to do this assessment automatically. Temperatures in the ADJ-2 block varied from 5 °C to 20 °C during this examination. The ADJ-1 and ADJ-3 unit operations, on the other hand, were employed to maintain a consistent CO_2 removal efficiency of 85 % and a constant incoming flue gas temperature of 40 °C, respectively.

The cost difference between 11 to 15 degrees Celsius was insignificant in both automated and manual scenarios. The discrepancy between the automatic and manual results was lower when the current point in the automatic analysis corresponds to the start point in the Case Study (5 °C) than when the current point was 10 °C.

The sensitivity analysis for the absorber packing height is a trade-off between the amine flow rate and the number of absorber stages. In this analysis, the absorber size and the heat transfer area in the lean/rich amine heat exchanger are two important capital costs. The three primary operational expenses impacted by this change are fan power consumption, steam consumption, and amine use. The absorber stages were changed from 13 to 18 to achieve the lowest CO_2 collected cost. As it is not possible to automatically conduct a sensitivity analysis by directly changing the number of stages in the absorber, an approach was used to define a Case Study by adjusting the effectiveness of a stage. Changing the efficiency from 0.15 to 0.9 for an absorber with 13 stages is equivalent to increasing the number of steps from 13 to 18.

The best scenario for all manual and automated evaluations was 15 meters of packing height, with a CO_2 collection cost of 42.64 EUR/t in the manual analysis. When the target stage for changing efficiency was the 13th stage, the automated calculated costs were on average 1.5 % higher than the manual technique, whereas when the target stage for changing efficiency was the 10^{th} stage in the Case Study, the automated calculated costs were on average 0.9 % higher than the manual assessment.

Increased incoming flue gas temperature decreases the DCC unit's size and duty, while increasing the absorber column's size and amine flow rate. To have an automated evaluation, it was necessary to first compute the new efficiency for the absorber stages based on the temperature distribution along the column, and then create a connection between the input temperatures and their average Murphree efficiency. When the step size is set to 5 °C in the Case Study, the produced result follows the same pattern for all automated and manual investigations. Raising the input flue gas temperature from 30 to 40 degrees Celsius increases reboiler duty and cost significantly, followed by a minor rise to 45 °C and then a tiny drop to 50 °C.

A 15-stage absorber was used in the first test. The computed result for collected cost was around 2% lower than the Base Case study due to using a lower amine flow rate by increasing the average efficiency. A lower-stage absorber can accomplish a similar rate of CO_2 removal by enhancing the efficacy of the stages. Therefore, a comparable investigation for a simulated scenario with 13 stages in the absorber was performed. Because of the smaller absorber, the captured cost was more than 4% lower than the basic Case.

The convergence was too difficult, and the acquired results were not smooth when the step size was reduced to 1 °C in the Case Study. The optimal inlet temperature in this scenario was determined as 34 °C, with a calculated cost of 39.57 EUR per ton captured CO_2.

Although there are some uncertainties and limitations that must be considered when using Aspen HYSYS software to model and estimate the cost of an amine-based CO_2 capture plant, such optimization studies might help to minimize total costs.

References

[1] "Causes of climate change." https://ec.europa.eu/clima/climate-change/causes-climate-change_en (accessed Mar. 28, 2022).

[2] US EPA (United State Environmental Protection Agency), "Overview of Greenhouse Gases," Dec. 23, 2015. https://www.epa.gov/ghgemissions/overview-greenhouse-gases (accessed Mar. 28, 2022).

[3] A. Alhajaj, N. Mac Dowell, and N. Shah, "A techno-economic analysis of post-combustion CO2 capture and compression applied to a combined cycle gas turbine: Part II. Identifying the cost-optimal control and design variables," *Int. J. Greenh. Gas Control*, vol. 52, pp. 331–343, Sep. 2016, doi: 10.1016/j.ijggc.2016.07.008.

[4] "IEA – International Energy Agency," *IEA*. https://www.iea.org (accessed Oct. 26, 2021).

[5] "Tracking Power 2021 – Analysis," *IEA*. https://www.iea.org/reports/tracking-power-2021 (accessed Mar. 28, 2022).

[6] M. H. W. N. Jinadasa, K.-J. Jens, and M. Halstensen, "Process Analytical Technology for CO2 Capture," in *Carbon Dioxide Chemistry, Capture and Oil Recovery*, I. Karamé, J. Shaya, and H. Srour, Eds. InTech, 2018. doi: 10.5772/intechopen.76176.

[7] H. A. Patel, J. Byun, and C. T. Yavuz, "Carbon Dioxide Capture Adsorbents: Chemistry and Methods," *ChemSusChem*, vol. 10, no. 7, pp. 1303–1317, Apr. 2017, doi: 10.1002/cssc.201601545.

[8] J. Singh and D. W. Dhar, "Overview of Carbon Capture Technology: Microalgal Biorefinery Concept and State-of-the-Art," *Front. Mar. Sci.*, vol. 6, p. 29, Feb. 2019, doi: 10.3389/fmars.2019.00029.

[9] M. K. Lam, K. T. Lee, and A. R. Mohamed, "Current status and challenges on microalgae-based carbon capture," *Int. J. Greenh. Gas Control*, vol. 10, pp. 456–469, Sep. 2012, doi: 10.1016/j.ijggc.2012.07.010.

[10] A. Hassanpouryouzband *et al.*, "Gas hydrates in sustainable chemistry," *Chem. Soc. Rev.*, vol. 49, no. 15, pp. 5225–5309, 2020, doi: 10.1039/C8CS00989A.

[11] M. Karimi, M. Hillestad, and H. F. Svendsen, "Capital costs and energy considerations of different alternative stripper configurations for post combustion CO2 capture," *Chem. Eng. Res. Des.*, vol. 89, no. 8, pp. 1229–1236, Aug. 2011, doi: 10.1016/j.cherd.2011.03.005.

[12] S. A. Aromada, N. H. Eldrup, F. Normann, and L. E. Øi, "Techno-Economic Assessment of Different Heat Exchangers for CO2 Capture," *Energies*, vol. 13, no. 23, Art. no. 23, Jan. 2020, doi: 10.3390/en13236315.

[13] "Carbon Dioxide Capture and Storage — IPCC." https://www.ipcc.ch/report/carbon-dioxide-capture-and-storage/ (accessed Mar. 28, 2022).

[14] L. Øi, "Aspen HYSYS simulation of CO2 removal by amine absorption from a gas based power plant," *SIMS2007 Conf.*, Jan. 2007.

[15] L. E. Øi, “Removal of CO2 from exhaust gas,” May 2012, Accessed: Oct. 10, 2021. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2437805

[16] O. B. Kallevik, “Cost estimation of CO2 removal in HYSYS,” Oct. 2010, Accessed: Oct. 28, 2021. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2439023

[17] K. Park and L. E. Øi, “Optimization of gas velocity and pressure drop in CO2 absorption column,” *292-297*, 2017, doi: 10.3384/ecp17138292.

[18] L. E. Øi *et al.*, “Optimization of Configurations for Amine based CO2 Absorption Using Aspen HYSYS,” *Energy Procedia*, vol. 51, pp. 224–233, Jan. 2014, doi: 10.1016/j.egypro.2014.07.026.

[19] S. Aforkoghene Aromada and L. Øi, “Simulation of improved absorption configurations for CO2 capture,” Nov. 2015, pp. 21–29. doi: 10.3384/ecp1511921.

[20] S. A. Aromada and L. E. Øi, “Energy and Economic Analysis of Improved Absorption Configurations for CO2 Capture,” *Energy Procedia*, vol. 114, pp. 1342–1351, Jul. 2017, doi: 10.1016/j.egypro.2017.03.1900.

[21] H. Ali, L. E. Øi, and N. H. Eldrup, “Simulation and Economic Optimization of Amine-based CO2 Capture using Excess Heat at a Cement Plant,” *58-64*, 2018, doi: 10.3384/ecp1815358.

[22] H. Ali, N. H. Eldrup, F. Normann, R. Skagestad, and L. E. Øi, “Cost Estimation of CO2 Absorption Plants for CO2 Mitigation – Method and Assumptions,” *Int. J. Greenh. Gas Control*, vol. 88, pp. 10–23, Sep. 2019, doi: 10.1016/j.ijggc.2019.05.028.

[23] H. Ali, *Techno-economic analysis of CO2 capture concepts*. University of South-Eastern Norway, 2019. Accessed: Oct. 16, 2021. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2622802

[24] L. E. Øi, N. Eldrup, S. Aromada, A. Haukås, J. HelvigIda Hæstad, and A. M. Lande, “Process Simulation, Cost Estimation and Optimization of CO2 Capture using Aspen HYSYS,” Mar. 2021, Accessed: Nov. 08, 2021. [Online]. Available: https://ep.liu.se/konferensartikel.aspx?series=ecp&issue=176&Article_No=46

[25] S. Aforkoghene Aromada, N. H. Eldrup, F. Normann, and L. E. Øi, “Simulation and Cost Optimization of different Heat Exchangers for CO2 Capture,” Mar. 2021, Accessed: Nov. 08, 2021. [Online]. Available: https://ep.liu.se/en/conference-article.aspx?series=ecp&issue=176&Article_No=45

[26] S. A. Aromada, N. H. Eldrup, and L. Erik Øi, “Capital cost estimation of CO2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors,” *Int. J. Greenh. Gas Control*, vol. 110, p. 103394, Sep. 2021, doi: 10.1016/j.ijggc.2021.103394.

[27] S. Aromada, N. Eldrup, and L. Øi, “Cost and Emissions Reduction in CO2 Capture Plant Dependent on Heat Exchanger Type and Different Process Configurations: Optimum Temperature Approach Analysis,” *Energies*, vol. 15, p. 425, Jan. 2022, doi: 10.3390/en15020425.

[28] S. Roussanaly *et al.*, “Towards improved cost evaluation of Carbon Capture and Storage from industry,” *Int. J. Greenh. Gas Control*, vol. 106, p. 103263, Mar. 2021, doi: 10.1016/j.ijggc.2021.103263.

[29] E. S. Rubin *et al.*, “A proposed methodology for CO2 capture and storage cost estimates,” *Int. J. Greenh. Gas Control*, vol. 17, pp. 488–503, Sep. 2013, doi: 10.1016/j.ijggc.2013.06.004.

[30] S. Roussanaly, K. Lindqvist, R. Anantharaman, and J. Jakobsen, “A Systematic Method for Membrane CO2 Capture Modeling and Analysis,” *Energy Procedia*, vol. 63, pp. 217–224, 2014, doi: 10.1016/j.egypro.2014.11.023.

[31] M. van der Spek, S. Roussanaly, and E. S. Rubin, “Best practices and recent advances in CCS cost engineering and economic analysis,” *91-104*, 2019, doi: 10.1016/j.ijggc.2019.02.006.

[32] X. Luo, “Process modelling, simulation and optimisation of natural gas combined cycle power plant integrated with carbon capture, compression and transport,” School of Engineering, The University of Hull, 2016. Accessed: Nov. 08, 2021. [Online]. Available: https://hydra.hull.ac.uk/resources/hull:14005

[33] S. Roussanaly, “Calculating CO2 avoidance costs of Carbon Capture and Storage from industry,” *Carbon Manag.*, vol. 10, no. 1, pp. 105–112, Jan. 2019, doi: 10.1080/17583004.2018.1553435.

[34] P. Psarras *et al.*, “Cost Analysis of Carbon Capture and Sequestration from U.S. Natural Gas-Fired Power Plants,” *Environ. Sci. Technol.*, vol. 54, no. 10, pp. 6272–6280, May 2020, doi: 10.1021/acs.est.9b06147.

[35] T. Pieri and A. Angelis-Dimakis, “Model Development for Carbon Capture Cost Estimation,” *Clean Technol.*, vol. 3, no. 4, Art. no. 4, Dec. 2021, doi: 10.3390/cleantechnol3040046.

[36] A. B. Rao and E. S. Rubin, “Identifying Cost-Effective CO2 Control Levels for Amine-Based CO2 Capture Systems,” *Ind. Eng. Chem. Res.*, vol. 45, no. 8, pp. 2421–2429, Apr. 2006, doi: 10.1021/ie050603p.

[37] N. H. Eldrup, A. Mathisen, R. Skagestad, and H. A. Haugen, “A Cost Estimation Tool for CO2 Capture Technologies,” Social Science Research Network, Rochester, NY, SSRN Scholarly Paper ID 3366036, Apr. 2019. doi: 10.2139/ssrn.3366036.

[38] C. Nwaoha, M. Beaulieu, P. Tontiwachwuthikul, and M. D. Gibson, “Techno-economic analysis of CO2 capture from a 1.2 million MTPA cement plant using AMP-PZ-MEA blend,” *Int. J. Greenh. Gas Control*, vol. 78, pp. 400–412, Nov. 2018, doi: 10.1016/j.ijggc.2018.07.015.

[39] A. Ayyad, A. Abbas, and N. Elminshawy, “A simulation study of the effect of post-combustion amine-based carbon-capturing integrated with solar thermal collectors for combined cycle gas power plant,” *Discov. Sustain.*, vol. 2, no. 1, p. 9, Mar. 2021, doi: 10.1007/s43621-021-00018-x.

[40] G. Manzolini, E. Sanchez Fernandez, S. Rezvani, E. Macchi, E. L. V. Goetheer, and T. J. H. Vlugt, “Economic assessment of novel amine based CO2 capture technologies integrated in power plants based on European Benchmarking Task Force methodology,” *Appl. Energy*, vol. 138, pp. 546–558, Jan. 2015, doi: 10.1016/j.apenergy.2014.04.066.

[41] S.-Y. Oh, M. Binns, H. Cho, and J.-K. Kim, “Energy minimization of MEA-based CO2 capture process,” *Appl. Energy*, vol. 169, pp. 353–362, May 2016, doi: 10.1016/j.apenergy.2016.02.046.

[42] R. Dutta, L. O. Nord, and O. Bolland, “Selection and design of post-combustion CO2 capture process for 600 MW natural gas fueled thermal power plant based on operability,” *Energy*, vol. 121, pp. 643–656, Feb. 2017, doi: 10.1016/j.energy.2017.01.053.

[43] E. O. Agbonghae, K. J. Hughes, D. B. Ingham, L. Ma, and M. Pourkashanian, “Optimal Process Design of Commercial-Scale Amine-Based CO $_2$ Capture Plants,” *Ind. Eng. Chem. Res.*, vol. 53, no. 38, pp. 14815–14829, Sep. 2014, doi: 10.1021/ie5023767.

[44] X. Luo and M. Wang, “Optimal operation of MEA-based post-combustion carbon capture for natural gas combined cycle power plants under different market conditions,” *Int. J. Greenh. Gas Control*, vol. 48, pp. 312–320, May 2016, doi: 10.1016/j.ijggc.2015.11.014.

[45] A. Mathisen, G. Hegerland, N. Eldrup, R. Skagestad, and H. Haugen, “Combining bioenergy and CO 2 capture from gas fired power plant,” *Energy Procedia*, vol. 4, pp. 2918–2925, Dec. 2011, doi: 10.1016/j.egypro.2011.02.199.

[46] R. Zhang, X. Zhang, Q. Yang, H. Yu, Z. Liang, and X. Luo, “Analysis of the reduction of energy cost by using MEA-MDEA-PZ solvent for post-combustion carbon dioxide capture (PCC),” *Appl. Energy*, vol. 205, pp. 1002–1011, Nov. 2017, doi: 10.1016/j.apenergy.2017.08.130.

[47] M. Gatti *et al.*, “Preliminary Performance and Cost Evaluation of Four Alternative Technologies for Post-Combustion CO2 Capture in Natural Gas-Fired Power Plants,” *Energies*, vol. 13, no. 3, Art. no. 3, Jan. 2020, doi: 10.3390/en13030543.

[48] M. M. F. Hasan, R. C. Baliban, J. A. Elia, and C. A. Floudas, “Modeling, Simulation, and Optimization of Postcombustion CO2 Capture for Variable Feed Concentration and Flow Rate. 1. Chemical Absorption and Membrane Processes,” *Ind. Eng. Chem. Res.*, vol. 51, no. 48, pp. 15642–15664, Dec. 2012, doi: 10.1021/ie301571d.

[49] W. J. Schmelz, G. Hochman, and K. G. Miller, “Total cost of carbon capture and storage implemented at a regional scale: northeastern and midwestern United States,” *Interface Focus*, vol. 10, no. 5, p. 20190065, Oct. 2020, doi: 10.1098/rsfs.2019.0065.

[50] S. Hasan, A. J. Abbas, and G. G. Nasr, “Improving the Carbon Capture Efficiency for Gas Power Plants through Amine-Based Absorbents,” *Sustainability*, vol. 13, no. 1, Art. no. 1, Jan. 2021, doi: 10.3390/su13010072.

[51] J. Husebye, A. L. Brunsvold, S. Roussanaly, and X. Zhang, “Techno Economic Evaluation of Amine based CO2 Capture: Impact of CO2 Concentration and Steam Supply,” *Energy Procedia*, vol. 23, pp. 381–390, 2012, doi: 10.1016/j.egypro.2012.06.053.

[52] U. Ali, E. O. Agbonghae, K. J. Hughes, D. B. Ingham, L. Ma, and M. Pourkashanian, “Techno-economic process design of a commercial-scale amine-based CO2 capture system for natural gas combined cycle power plant with exhaust gas recirculation,” *Appl. Therm. Eng.*, vol. 103, pp. 747–758, Jun. 2016, doi: 10.1016/j.applthermaleng.2016.04.145.

[53] N. Sipöcz and F. A. Tobiesen, “Natural gas combined cycle power plants with CO2 capture – Opportunities to reduce cost,” *Int. J. Greenh. Gas Control*, vol. 7, pp. 98–106, Mar. 2012, doi: 10.1016/j.ijggc.2012.01.003.

[54] A. L. Haukås, “Process Simulation and Cost Optimization of CO2 Capture Using Aspen HYSYS,” Master thesis 2020, University of South-Eastern Norway

[55] P. Rahmani, “Process simulation and automated cost optimization of CO2 capture using Aspen HYSYS,” 2021, Accessed: Mar. 30, 2022. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2832844

[56] S. A. Aromada, N. H. Eldrup, and L. Erik Øi, “Capital cost estimation of CO2 capture plant using Enhanced Detailed Factor (EDF) method: Installation factors and plant construction characteristic factors,” *Int. J. Greenh. Gas Control*, vol. 110, p. 103394, Sep. 2021, doi: 10.1016/j.ijggc.2021.103394.

[57] S. Fagerheim, “Process simulation of CO2 absorption at TCM Mongstad,” *140*, 2019, Accessed: Nov. 08, 2021. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2644680

[58] S. Laribi, L. Dubois, G. De Weireld, and D. Thomas, “Study of the post-combustion CO2 capture process by absorption-regeneration using amine solvents applied to cement plant flue gases with high CO2 contents,” *Int. J. Greenh. Gas Control*, vol. 90, p. 102799, Nov. 2019, doi: 10.1016/j.ijggc.2019.102799.

[59] C. Gouedard, D. Picq, F. Launay, and P.-L. Carrette, “Amine degradation in CO2 capture. I. A review,” *Int. J. Greenh. Gas Control*, vol. 10, pp. 244–270, Sep. 2012, doi: 10.1016/j.ijggc.2012.06.015.

[60] S. Orangi, “Simulation and cost estimation of CO2 capture processes using different solvents/blends,” 2021, Accessed: Oct. 10, 2021. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/2774682

[61] “Sulzer Chemtech, Separation Technology for the Hydrocarbon Processing Industry.” Accessed: Oct. 28, 2021. [Online]. Available: https://www.sulzer.com/-/media/files/applications/oil-gas-chemicals/downstream/brochures/separation_technology_for_hpi.ashx

[62] “Sulzer Chemtech, Separation Technology for the Hydrocarbon Processing Industry.” Accessed: Oct. 28, 2021. [Online]. Available: https://www.sulzer.com/-/media/files/applications/oil-gas-chemicals/downstream/brochures/separation_technology_for_hpi.ashx

[63] T. L. Bergman and F. P. Incropera, Eds., *Fundamentals of heat and mass transfer*, 7th ed. Hoboken, NJ: Wiley, 2011.

[64] “Consumer price index,” *SSB*. https://www.ssb.no/en/priser-og-prisindekser/konsumpriser/statistikk/konsumprisindeksen (accessed Feb. 05, 2022).

[65] L. E. Øi, A. Haukås, S. Aromada, and N. Eldrup, “Automated Cost Optimization of CO2 Capture Using Aspen HYSYS,” pp. 293–300, Mar. 2022, doi: 10.3384/ecp21185293.

[66] C. Baumann, “Cost-to-Capacity Method: Applications and Considerations,” *MTS J.*, vol. 30, no. 1, p. 8, 2014.

[67] R. Smith and P. of P. R. Smith, *Chemical Process: Design and Integration*. Hoboken: John Wiley & Sons, Incorporated, 2005.

[68] S. Shirdel *et al.*, “Sensitivity Analysis and Cost Estimation of a CO2 Capture Plant in Aspen HYSYS,” *ChemEngineering*, vol. 6, no. 2, Art. no. 2, Apr. 2022, doi: 10.3390/chemengineering6020028.

[69] Z. Amrollahi, P. A. M. Ystad, I. S. Ertesvåg, and O. Bolland, “Optimized process configurations of post-combustion CO2 capture for natural-gas-fired power plant – Power plant efficiency analysis,” *Int. J. Greenh. Gas Control*, vol. 8, pp. 1–11, May 2012, doi: 10.1016/j.ijggc.2012.01.005.

[70] N. Sipöcz, A. Tobiesen, and M. Assadi, “Integrated modelling and simulation of a 400 MW NGCC power plant with CO2 capture,” *Energy Procedia*, vol. 4, pp. 1941–1948, Jan. 2011, doi: 10.1016/j.egypro.2011.02.074.

[71] P. Ecology, “Improving flowsheet convergence in HYSYS.” https://processecology.com/articles/improving-flowsheet-convergence-in-hysys (accessed Jan. 21, 2022).

Appendices

Appendix A: Project description

University of
South-Eastern Norway

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

FMH606 Master's Thesis

Title: Process simulation, dimensioning and automated cost optimization of CO_2 capture

USN supervisor: Lars Erik Øi, Cosupervisor: Solomon Aromada (PhD-student USN)

External partner: SINTEF Tel-Tek /Nils Eldrup

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

Task description:
The general aim is to develop further models in Aspen HYSYS especially for cost optimization of CO_2 capture by amine absorption. A special aim is to utilize possibilities like the spreadsheet facility in Aspen HYSYS, the Aspen simulation workbook or an Excel connection to optimize the process.

1. Literature search on cost estimation and optimization of amine based CO_2 capture with emphasis on optimization based on simulation, dimensioning and cost estimation.

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

3. Process optimization of process parameters and possibly automated optimization based on Aspen simulation workbook or an Excel connection. Typical parameters are gas inlet temperature, temperature approach in the main heat exchanger and packing height in the absorption column. Possible challenges are optimization of the gas velocity and the pressure drop through the absorber.

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

Student category: EET or PT

Is the task 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.

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

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

Signatures:

Supervisor (date and signature): 9/5-22

Student (write clearly in all capitalized letters): SHIRVAN SHIRDEL

Student (date and signature): 09.05.2022

Appendix B: Make up specifications

MEA:

MEA Inlet [kg/h]	707781.7
MEA loss in Cleaned Gas [kg/h]	367.0171
MEA loss in CO_2 Captured stream [kg/h]	8.75E-04
Total MEA loss [kg/h]	367.0179

Water:

Water Inlet in MEA Solution [kg/h]	1601731
Water Inlet in FG stream [kg/h]	98920.46
Water loss in Cleaned Gas [kg/h]	173079.5
Water loss in CO_2 Captured stream [kg/h]	31762.13
Total Water Loss [kg/h]	204841.7
Required MakeUp water [kg/h]	105921.2

Appendix C: Base Case dimensioning

Columns (Absorber & Desorber):

	Absorber	Desorber
FG volume flow [m^3/h]	2333769	119996.2
FG volume flow [m^3/s]	648.3751	33.33227
FG velocity [m/s]	2.5	1
Inner Diameter [m]	18.17181	6.514596
Number of stages	15	10
Height of each stage [m]	1	1
Packing height [m]	15	10
Column height [m]	35	25
Column volume [m3]	9077.252	833.3067
Packing volume [m3]	3890.251	333.3227
Number of units	3	1
Volume per unit [m3]	3025.751	833.3067
Packing volume per unit [m3]	1296.75	333.3227
Diameter per unit [m]	10.4915	6.514596
SHELL MAT.	SS316	SS316
PACKING TYPE	MellaPak 250Y	MellaPak 250Y
No. PACKED SECTIONS	3	2

Heat exchangers:

	Reboiler	Condenser	L/R HEx	LA Cooler	Inlet Cooler
Q [kJ/h]	4.48E+08	56992110	5.61E+08	1.02E+08	1.26E+08
Heat transfer coefficient [kw/m2K]	1.2	1	0.732	0.8	0.8
LMTD	18.77747	77.11269	12.83078	26.55822	40.30689
Total Heat Transfer Area [m2]	5518.385	205.2988	16584.01	1339.142	1083.159
T(out,cold) [°C]	120.0094	25	103.7	25	25
T(in,cold) [°C]	116.3091	15	43.21	15	15
T(in,Hot) [°C]	145	101.3735	120	53.18	85.86
T(out,Hot) [°C]	130	92.85668	53.13	40	40
Max. Area per of Unit [m2]	1000	1000	1000	1000	1000
Calculated Number of Unit	5.518385	0.205299	16.58401	1.339142	1.083159
Actual Number of Unit	6	1	17	2	2
Actual Area per unit [m2]	919.7309	205.2988	975.5298	669.5709	541.5795

Fan & Pumps:

	Fan	RA Pump	LA Pump
Duty [kW]	4132.559	70.51914	246.27
Flow rate [m3/h]	2267807	2240.02	2216.43
Flow rate [L/s]	629946.4	622.2277	615.6751
Max flow rate [m3/h]	1529000	-	-
Calculated No. of units	1.483196	-	-
Actual no. of units	2	1	1
Actual Flow rate [m3/h]	1133904	-	-
Actual Duty [kW]	2066.279	-	-

Separator:

	Separator
Actual Gas Flow Rate [m3/h]	1917787
Actual Gas Flow Rate [m3/s]	532.7185
Liquid Phase Mass Density [kg/m3]	992.2268
Gas Phase Mass density [kg/m3]	1.236469
K Factor, Sounder-Brown Velocity	0.15
Allowable Vapour Velocity [m/s]	4.246531
Vessel Cross-Sectional Area [m2]	125.4479
Vessel Inner-Diameter (Di) [m]	12.63825
Vessel Height, 1D [m]	12.63825
Vessel Volume [m3]	1585.442

Appendix D: 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 E: CAPEX calculation for the Base Case

Columns (Absorber & Desorber):

		ABSORBER	Desorber
Packing height [m]		15	10
Column height [m]		35	25
Column volume [m3]		9077.252	833.3067
Packing volume [m3]		3890.251	333.3227
Shell Volume [m3]		5187.001	499.984
Diameter [m]		18.17181	6.514596
Number of units		3	1
Volume per unit [m3]		3025.751	
Packing volume per unit [m3]		1296.75	
Shell volume per unit [m3]		1729	
Diameter per unit [m]		10.4915	
SHELL MAT.		SS316	SS316
No. Packing section		3	2
Equipmen Cost per unit [kEUR]	Aspen In Plant 2019	8291	2641.7
Packing cost per unit [kEUR]	Aspen In Plant 2019	4727.312	1215.629
Shell cost per unit [kEUR]	Aspen In Plant 2019	3563.688	1426.071
Equipmen Cost per unit [kEUR] SS	Convert to 2020	8449.139	2692.087
Shell cost per unit [kEUR] SS	Convert to 2020	3631.66	1453.271
Packing cost per unit [kEUR] SS	Convert to 2020	4817.479	1238.815
Equipmen Cost per unit [kEUR] CS	Convert to CS	4828.079	1538.335
Shell cost per unit [kEUR] CS	Convert to CS	2075.234	830.4407
Packing cost per unit [kEUR] CS	Convert to CS	2752.845	707.8945
Installation factor CS 2020 Shell		3.19	3.63
Installation factor CS 2020 Packing		2.84	3.63
Installation factor SS 2020 Shell		4.165	4.665
Installation factor SS 2020 Packing		3.7625	4.665
Shell cost 2020 [kEUR]		8643.351	3874.006
Packing cost 2020 [kEUR]		10357.58	3302.328
Shell cost 2021 [kEUR]	Convert to 2021	8943.789	4008.664
Packing cost 2021 [kEUR]	Convert to 2021	10717.6	3417.115
Annulised Shell cost [kEUR/year]		2669.062	398.7635
Annulised packing cost [kEUR/year]		3198.414	339.9189
Annulised Shell & Packing cost [kEUR/year]		5867.476	738.6825

Heat exchangers:

		L/R HEx	LA Cooler	Reboiler	Condenser	Inlet Cooler
Total Heat Transfer Area [m2]		16584.01	1339.142	5518.385	205.2988	1083.159
Max. Area per of Unit [m2]		1000	1000	1000	1000	1000
Calculated Number of Unit		16.58401	1.339142	5.518385	0.205299	1.083159
Actual Number of Unit		17	2	6	1	2
Actual Area per unit [m2]		975.5298	669.5709	919.7309	205.2988	541.5795
Material		SS316	SS316	SS316	SS316	SS316
Equipmen Cost per unit [kEUR]	Aspen In-Plant 2019	389.3	280.1	406.2	118	281.6
Equipment cost (Total) [EUR]	One unit	6629.9	565.2	2494	118	430.6
Equipmen Cost per unit [kEUR] SS	Convert to 2020	396.7253	285.4425	413.9477	120.2507	286.9711
Equipmen Cost per unit [kEUR] CS	Convert to CS	226.7002	163.11	236.5415	68.71467	163.9835
Installation factor CS 2020 Total		4.92	4.92	4.92	7.22	4.92
Installation factor SS 2020 total		6.12	6.12	6.12	8.69	6.12
Total Equipment cost 2020 [kEUR]		1387.405	998.2332	1447.634	597.1305	1003.579
equipment cost 2021 [kEUR]	Convert to 2021	1435.631	1032.931	1497.953	617.8864	1038.463
Annulised Equipment cost [kEUR/year]		2427.769	205.5025	894.0571	61.46451	206.603

Fan & Pumps:

		FAN	RA Pump	LA Pump
Duty [kW]		4132.559	70.51914	246.27
Flow rate [m3/h]		2267807	2240.02	2216.43
Flow rate [L/s]		629946.4	622.2277	615.6751
Max flow rate [m3/h]		1529000	-	-
Calculated No. of units		1.483196	-	-
Actual no. of units		2	1	1
Actual Flow rate [m3/h]		1133904	-	-
Actual Duty [kW]		2066.279	70.51914	246.27
Material		CS	SS316	SS316
Equipment cost per unit [kEUR]	Aspen In-Plant 2019	1169.9	171.9	217.1
Equipmen Cost per unit [kEUR] SS	Convert to 2020	1192.214	175.1787	221.2409
Equipmen Cost per unit [kEUR] CS	Convert to CS	1192.214	134.7529	170.1853
Installation factor CS 2020 Total		3.63	5.89	4.92
Installation factor SS 2020 total		-	6.418	5.4
Total Equipment cost 2020 [kEUR]		4327.737	864.844	919.0005
equipment cost 2021 [kEUR]	Convert to 2021	4478.167	894.9054	950.9444
Annualized Equipment cost [kEUR/year]		890.9351	89.0211	94.5956

Separator:

	Separator	
Vessel Inner-Diameter (Di) [m]	12.63825	
Vessel Height, 1D [m]	12.63825	
Vessel Volume [m3]	1585.442	
Actual no. of units	1	
Actual Volume [m3]	1585.442	
Material	SS316	
Equipment cost per unit [kEUR]	2302.1	Aspen In-Plant 2019
Equipment cost per unit [kEUR]	2346.009	Convert to 2020
Equipmen Cost per unit [kEUR] CS	1340.577	Convert to CS
Installation factor CS 2020 Total	3.19	
Installation factor SS 2020 total	4.165	
Total Equipment cost 2020 [kEUR]	5583.502	
equipment cost 2021 [kEUR]	5777.581	Convert to 2021
Annualized Equipment cost [kEUR/year]	574.7273	

Total CAPEX:

Total Annualized cost [kEUR/y]	12050.83	[kEUR/year]
Total CAPEX	121143.83	[kEUR]

Appendix F: OPEX calculation for the Base Case

Operating time per year: 8000 hours

Steam:

Steam price [EUR/kWh]	1.50E-02	
	Consumption [kW]	Price [kEUR/year]
Reboiler	124345.6	14921.47

Electricity:

Electricity price [EUR/kWh]	6.00E-02	
	Consumption [kW]	Price [kEUR/year]
Fan	4132.559	1983.628
RA Pump	70.51914	33.84919
LA Pump	246.27	118.2096
Total Price		2135.687

Cooling water:

Cooling Water Price [EUR/m3]	2.20E-02	
	Consumption [m3/h]	Price [kEUR/year]
LA Cooler	2704.232	475.9449
Condenser	1504.668	264.8216
Inlet Cooler	3319.632	584.2552
Total Price		1325.022

Process water:

Process Water price [EUR/m3]	0.203	
	Consumption [m3/h]	Price [kEUR/year]
Water in MEA solution	1604.818	0.325778
Make Up water	106.7433	173.3512
Total Price		173.6769

MEA:

MEA price [EUR/ton]	1450	
	Consumption [kg/h]	Price [kEUR/year]
MEA in MEA solution	707781.7	1026.283
Make Up MEA	367.0179	4257.408
Total price		5283.692

Operator:

Operator price [EUR/year]	80414	
	Number of Operator	Price [kEUR/year]
Operator	6	482.484

Engineer:

Engineer price [EUR/year]	156650	
	number of Engineer	Price [kEUR/year]
Engineer	1	156.65

Maintenance:

4% of CAPEX	Price [kEUR/year]
Maintenance	4383.547

Total OPEX:

TOTAL OPEX [kEUR/year]	28862.22

Appendix G: CO_2 captured cost & Reboiler duty

CO_2 Captured cost:

CO_2 Captured [t/year]	954880.2
Total Annualized CAPEX [kEUR/y]	12050.83
TOTAL OPEX [kEUR/year]	28862.22
CO_2 Captured cost [EUR/t]	42.85

Reboiler Duty:

CO_2 Captured [kg/h]	119360
Reboiler Duty [kJ/h]	4.48E+08
Required heat [kJ/kg]	3750.368

CO_2 removal efficiency:

Inlet CO_2 [kg/h]	140281.2
CO_2 in Cleaned gas [kg/h]	21055.19
CO_2 Removal Efficiency	84.99073