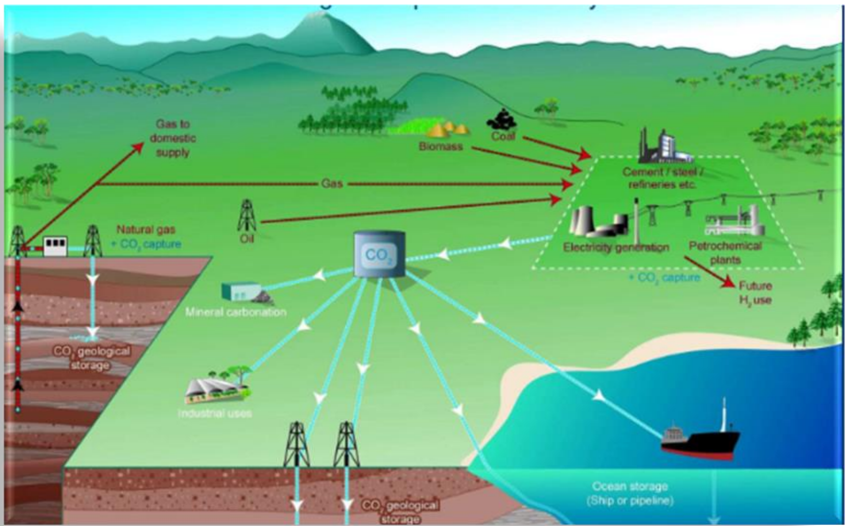
University of South-Eastern Norway

www.usn.no

FMH606 Master's Thesis 2024

Process Technology

Process simulation and automated cost optimization of CO_2 capture



Mohsen Gholizadefalah

Faculty of Technology, Natural sciences and Maritime Sciences
Campus Porsgrunn

University of South-Eastern Norway

www.usn.no

Course: FMH606 Master's Thesis, 2024

Title: Process simulation and automated cost optimization of CO_2 capture

Number of pages: 46

Keywords: CO_2 capture, Simulation, MEA, Amine solvent, Cost estimation, Cost optimization, Aspen HYSYS, Aspen In-Plant Cost Estimator, Absorption, Dimensioning

Student: Mohsen Gholizadefalah

Supervisor: Lars Erik Øi

Co-supervisor: Solomon Aromada

External partner: -

University of South-Eastern Norway

www.usn.no

Summary:

The growing concern over CO_2 emissions has led to the development of CO_2 capture technologies, with amine absorption, especially using monoethanolamine (MEA), being a widely adopted method for removing CO_2 from industrial gas streams.

This study aims to advance the automation of cost optimization for CO_2 capture via amine absorption, specifically using Aspen HYSYS models. The project focuses on simulating the CO_2 capture process, dimensioning equipment, and automating cost estimation using the Aspen In-Plant Cost Estimator and the Enhanced Detailed Factor (EDF) method. The primary goal is to create a robust and automated process that optimizes CO_2 capture costs, considering both capital expenditure (CAPEX) and operational expenditure (OPEX).

The study utilizes flue gas data from a natural gas-based power plant at Mongstad, Norway, to develop and simulate a process design in Aspen HYSYS. The base case scenario involves an absorber packing height of 15 meters, a desorber packing height of 4 meters, an 85% CO_2 removal efficiency, and a minimal temperature difference (ΔT_{min}) of 8°C in the lean/rich amine heat exchanger. For the base case, the overall cost was calculated at 44 EUR/ton of CO_2 captured, with an energy consumption of 4126 kJ/kg CO_2 in the reboiler.

Several case studies were conducted to identify cost-optimal scenarios. The first case study focused on optimizing the economic performance of the lean/rich amine heat exchanger by adjusting ΔT_{min} from 8 to 18°C. The optimal ΔT_{min} was found to be around 15°C, with a CO_2 capture cost of 42.5 EUR/ton and a reboiler duty of 4216 kJ/kg. This case study highlighted the trade-off between heat exchanger size and steam consumption, with automated calculations suggesting an optimal ΔT_{min} between 14 and 16°C.

The second case study examined the absorber packing height, varying it from 13 to 18 meters. The optimal packing height was determined to be 14 meters, resulting in a CO_2 capture cost of 43.9 EUR/ton and a reboiler duty of 4036 kJ/kg CO_2. This study revealed that increasing the number of absorber stages reduces the amine circulation rate and reboiler duty, influencing both capital and operational costs. It also suggested that higher CO_2 concentrations in flue gas could lower costs, warranting further research into this variable.

The third case study explored the impact of varying inlet gas velocity to the absorber, ranging from 1.5 to 3 m/s. The most cost-effective velocity was found to be 2.5 m/s, with a CO_2 capture cost of 53 EUR/ton and a minimum reboiler duty of 4183 kJ/kg CO_2. Higher velocities initially reduced costs by enhancing CO_2 absorption efficiency, but beyond 2.5 m/s, costs increased due to decreased contact time between the flue gas and amine solvent.

Preface

This thesis was completed as part of the master's degree in Process Technology at the University of South-Eastern Norway in the spring of 2024. I would like to take this opportunity to thank everyone who helped me finish my thesis.

My deepest gratitude goes to my supervisor, Professor Lars Erik Øi, and my co-supervisor, Solomon Aromada, for their guidance and support throughout the thesis process.

I also want to express my heartfelt appreciation to my family and friends for their encouragement and support during this time. Their help was invaluable in making this thesis possible.

Porsgrunn, June 2024

Mohsen Gholizadefalah

Contents

Nomenclature

A	Cross-sectional area [m^2]
CAPEX	Capital expenditure
CCUS	CO_2 capture utilization and storage
CO_2	Carbon Dioxide
D	Diameter [m]
e	exponential size factor
EIC_{CS}	total installed cost for each equipment
$F_{T,CS}$	total installation factor
f_{mat}	Material Factor
MEA	Monoethanolamine
OPEX	Operational expenditures
r	discount rate
$\dot{Q}$	Duty
U	overall heat transfer coefficient
$\dot{V}$	Volumetric flow rate
$\dot{v}_{gas}$	gas velocity
ΔT_{lm}	Logarithmic mean temperature difference [℃]
ΔT_{min}	Heat exchanger minimum temperature approach [℃]

1 Introduction

This chapter provides a brief background on greenhouse gases, particularly focusing on carbon dioxide (CO_2), and emphasizes the importance of its capture for environmental purposes. It then discusses various methods of capturing CO_2. Following this, the thesis's task description and objectives are outlined, along with an overview of the thesis structure.

1.1 Background

Human activities and needs, particularly in energy production, manufacturing, and transportation, have heavily relied on the combustion of fossil fuels. Carbon dioxide is a consequential by-product of these industrial processes. Regrettably, CO_2 is classified as a greenhouse gas, primarily responsible for exacerbating global warming and consequent climate change. Figure 1.1 illustrates the varying contributions of different sectors to global CO_2 emissions, highlighting a persistent upward trend in CO_2 emissions over time.[1]

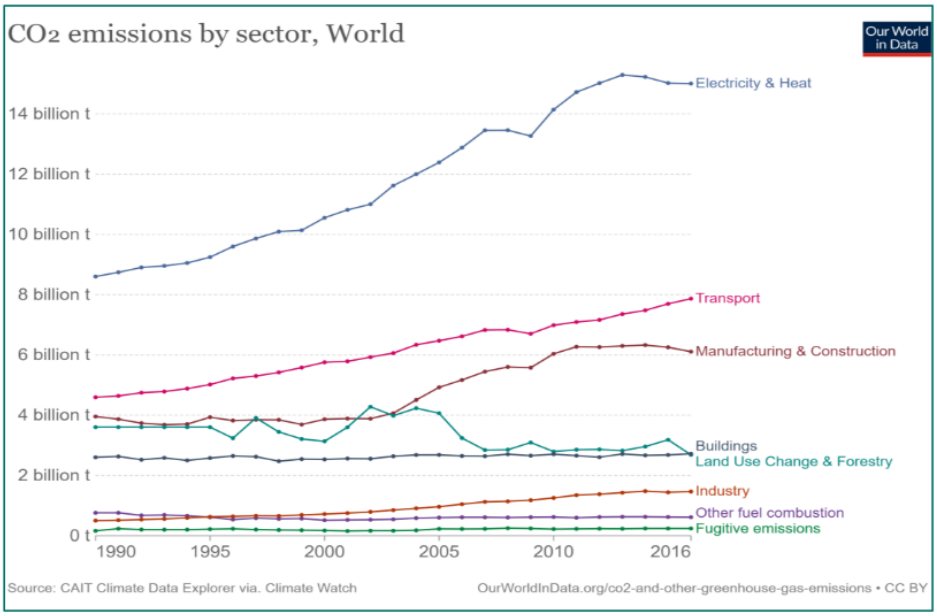


Figure 1.1 : Global annual CO_2 emissions by sector [1]

In recent years, the removal of CO_2 from exhaust gases, known as CO_2 capture, has emerged as a crucial subject. There is international agreement to limit greenhouse gas emissions to mitigate the issue of global warming, with CO_2 recognized as the most significant greenhouse gas.

The Kyoto Protocol stands out as a significant agreement, and subsequent United Nations Climate Change Conferences have seen high-level political discussions aimed at establishing future agreements on greenhouse gas emissions. Among the strategies proposed to reduce CO_2 emissions is to capture CO_2 from exhaust gases and then send it for storage. Both the IPCC (Intergovernmental Panel on Climate Change) and the IEA (International Energy Agency) underscore Carbon Capture and Storage (CCS) as a crucial measure for reducing global CO_2 emissions. Figure 1.2 illustrates potential CCS systems schematically. [2]

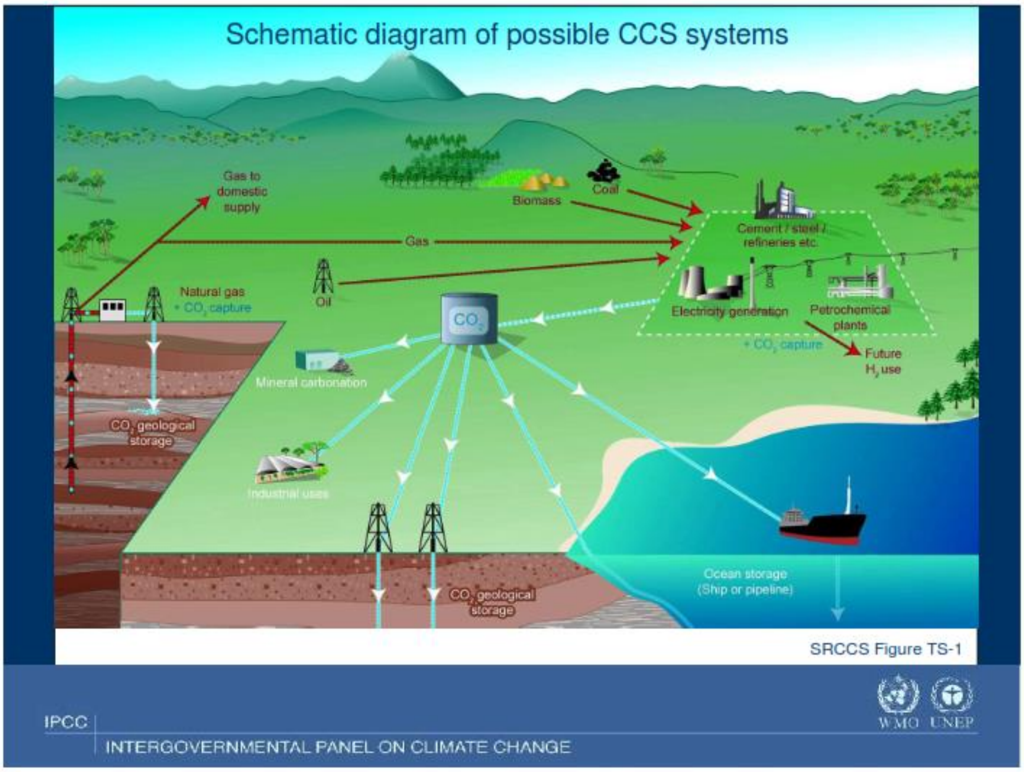


Figure 1.2 : A schematic diagram of possible CCS systems [3]

To date, large-scale CO_2 removal plants, handling over 1 million tons of CO_2 annually, typically remove CO_2 from industrial streams at pressures higher than atmospheric. CO_2 removal from atmospheric exhaust, limited to around 100,000 tons/yr, has mainly aimed for CO_2 production. Nevertheless, plans are underway for several large-scale CO_2 removal plants in the near future.

Most human-made CO_2 emissions are from fossil fuel combustion, with coal being the predominant fuel source. Consequently, most CO_2 capture projects have centered on coal-based power plants. However, in Norway, there's a specific focus on extracting CO_2 from natural gas power plant emissions. In 2007, the Norwegian Prime Minister announced plans for a natural gas power plant with CO_2 capture at Mongstad. Originally slated for operation in 2014, the CO_2 removal plant's timeline has since been delayed.

There are several suggested methods for removal or capture of CO_2. An overview of the different possibilities is presented in Figure 1.3. [2]

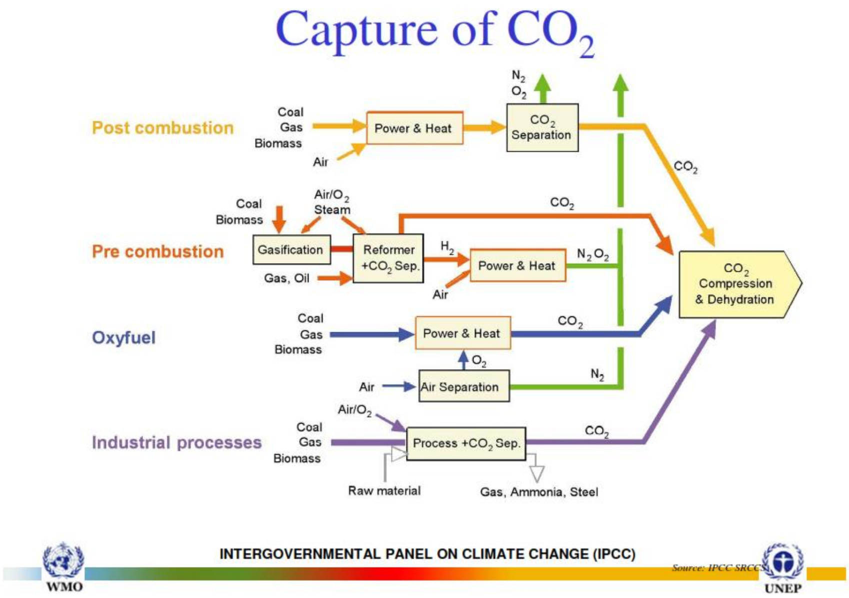


Figure 1.3 : Overview of different CO_2 capture principles [3]

One of the methods utilized for the removal of CO_2 from atmospheric gas on a large scale (100,000 tons/yr.), involves absorption using a combination of water and an amine. Monoethanolamine (MEA) stands out as a widely utilized solvent in carbon capture processes aimed at extracting CO_2 from industrial gas streams. Figure 1.4 shows a typical amine-based process for CO_2 removal, which serves as the foundational process design for this study. Detailed explanations of this process design will be provided in the subsequent chapters. [2]

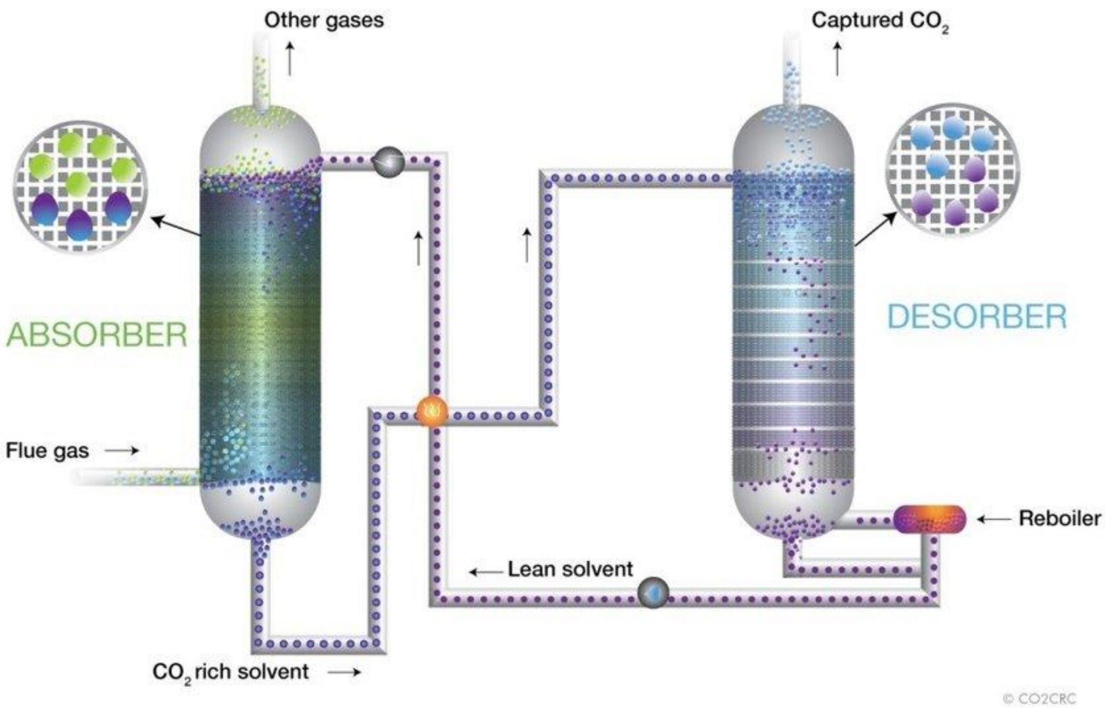


Figure 1.4 : Typical amine based CO_2 removal process [4]

1.2 Task description and objectives

Since 2007, master projects at the University of South-Eastern Norway and Telemark University College have integrated cost estimation into a spreadsheet linked to Aspen HYSYS simulation. Collaborating with various companies such as SINTEF Tel-Tek, Equinor, Aker Solutions, Norcem, Yara, Skagerak, and Gassnova.

The general objective of this work is to develop further Aspen HYSYS models, specifically focusing on automating cost optimization for CO_2 capture via amine absorption. A special objective is to explore options such as utilizing the spreadsheet feature within Aspen HYSYS, the Aspen simulation workbook, or establishing an Excel connection to enhance the process optimization.

Automated cost optimization is complicated due to model complexities, necessitating evaluation of model simplifications and iteration tolerances.

The objectives include:

1. Conducting a thorough review of literature regarding the estimation and optimization of costs associated with amine-based CO_2 capture, with a focus on optimizing through simulation, dimensioning, and automated cost estimation.
2. Utilizing Aspen HYSYS simulation, sizing, and cost estimation to explore various options, applying the spreadsheet functionality within the Aspen HYSYS software.
3. Optimizing process parameters and potentially automating cost optimization using the Aspen simulation workbook or an Excel link. Parameters include the minimum temperature approach in the main heat exchanger, packing height in the absorption column, as well as optimizing gas velocity and pressure drop through the absorber.
4. Assessing limitations associated with automated cost estimation and optimization for amine-based CO_2 absorption.

1.3 Scope of thesis

The scope of this study is to develop a robust and automated method for performing both process simulation and cost optimization for an amine absorption CO_2 capture method using flue gas data from a natural gas-based power plant project in Mongstad, Norway [5], [6]. To achieve this, a process design will be created based on the available data and simulated in Aspen HYSYS software. Following the simulation, dimensioning will be conducted within a spreadsheet in Aspen HYSYS. Using this data, along with the Aspen In-Plant Cost Estimator, a cost estimation will be performed. The Enhanced Detailed Factor (EDF) method will be utilized alongside the Aspen In-Plant Cost Estimator to estimate and compute the overall cost for the base case scenario. Capital expenditure (CAPEX) and operational expenditure (OPEX) will be calculated, and the cost per ton of CO_2 captured will be determined in EUR/ton.

Subsequently, a sensitivity analysis will be conducted to optimize costs and minimize expenses. This approach will be tested through a series of case studies to examine how different variables influence cost estimation. When the sensitivity analysis alters the equipment size, the Power-Law approach will be used to adjust the equipment cost. Key variables to be investigated include the minimum temperature difference in the lean/rich amine heat exchanger, the absorber packing height, and the inlet flue gas velocity into the absorber column. The CO_2 capture cost (EUR/ton) for different scenarios will be compared to identify the most cost-effective option.

2 Literature review

This study, conducted as part of ongoing research at the University of South-Eastern Norway (USN), aims to delve into CO_2 capture processes from various perspectives. It builds upon the foundation laid by Lars Erik Øi's PhD thesis at USN [2], as well as the extensive experience in cost estimation from Nils Henrik Eldrup through his teaching and industrial projects at both SINTEF Industry and USN. Additionally, insights from several student projects supervised by Øi. The methodology employed relies on the Enhanced Detailed Factor (EDF) method developed by Nils Henrik Eldrup. The review of relevant concepts and literature not only provides background information but also highlights areas where further research is needed.

2.1 Status of carbon capture

The utilization of amine-based solvents for CO_2 capture is not a new concept. Dating back to the 1920s-1930s, this method was initially employed for natural gas treatment to meet quality standards and in the production of syngas for methanol and ammonia. The use of CO_2 from natural gas treatment for enhanced oil recovery (EOR) was first established in the United States (Texas). However, it wasn't until 1977 that the idea of CO_2 separation, driven by concerns about climate change and Carbon Capture and Storage (CSS), was proposed. Throughout the 1990s, researchers began to focus on various aspects of CSS, leading to the implementation of several pilot plants and large-scale industrial projects worldwide. [7], [8]

The current status of CCS facilities globally in terms of number and capture capacity in metric tons per annum (Mtpa) is presented in Table 2.1. While their locations can be seen in Figure 2.1. [9], [10]

Table 2.1 provides an overview of the current worldwide status of CCS facilities, including the number of facilities and their respective capture capacities measured in metric tons per annum (Mtpa). The geographical distribution of these facilities can be visualized in Figure 2.1. [9], [10]

Table 2.1: Current commercial facilities globally as of September 2021 [10]

	Number of facilities	Capture capacity (Mtpa)
Operational	27	36.6
Under construction	4	3.1
Advanced development	58	46.7
Early development	44	60.9
Operation suspended	2	2.1
Total	135	149.3

Figure 2.1 : Locations of current commercial facilities globally as of September 2021 [10]

2.2 Carbon capture technologies

This section outlines the main categories of carbon dioxide removal technologies along with a short explanation of each. Carbon capture technologies cover processes or unit operations aimed at separating CO_2 from gas mixtures to yield a CO_2-rich stream for storage or utilization. These methods can be categorized into four main procedures: [11]

- Pre-combustion
  In pre-combustion systems, fossil fuels are transformed into synthesis gas (syngas) for subsequent combustion. Specifically, this process entails converting solid, liquid, or gaseous fuel into syngas without combustion. This allows for the removal of CO_2 from the mixture prior to the utilization of H2 for combustion.
- Oxy-combustion
  In oxy-combustion, the combustion process is carried out using pure oxygen instead of air. This oxygen-rich environment, nitrogen-free, leads to the production of flue gases mainly composed of CO_2 and H2O.
- Chemical looping
  The chemical looping process resembles oxy-fuel combustion, but it utilizes a metal oxide as an oxygen carrier for combustion instead of pure oxygen. In this process, the metal oxide undergoes reduction to metal while the fuel is oxidized to CO_2 and water.
- Post-combustion
  In post-combustion capture, CO_2 is extracted from the exhaust of a combustion process by absorbing it into a solvent. The captured CO_2 is then released from the solvent, compressed for transportation, and stored. Post-combustion technology is presently the most developed method for CO_2 capture.

2.3 Amine absorption method for CO_2 capture

As previously mentioned, various CO_2 removal technologies have been developed employing different physical and chemical processes. Among these, the absorption-based method utilizing amine solutions stands out as the most practical approach. In this process, CO_2 is absorbed into a mixture of amine and water. Monoethanolamine (MEA) is a commonly used amine due to its high reactivity with CO_2 and ease of regeneration. The absorption occurs in an absorber tower where the flue gas containing CO_2 comes into contact with the amine solution. CO_2 is selectively absorbed by the amine, forming a carbamate compound. Subsequently, the amine solution, now enriched with CO_2, is circulated to a desorber tower. In the desorber tower, heat is applied to release the captured CO_2 from the amine solution, regenerating the amine for reuse in the absorption process. This cyclic operation enables efficient capture of CO_2 from flue gas emissions while minimizing the environmental impact. [2] [12]

Figure 2.2 shows a more detailed depiction of the process, illustrating additional components such as a Direct Contact Cooler (DCC), a water wash section located at the top of the absorber, and an amine reclaimer following the desorber.

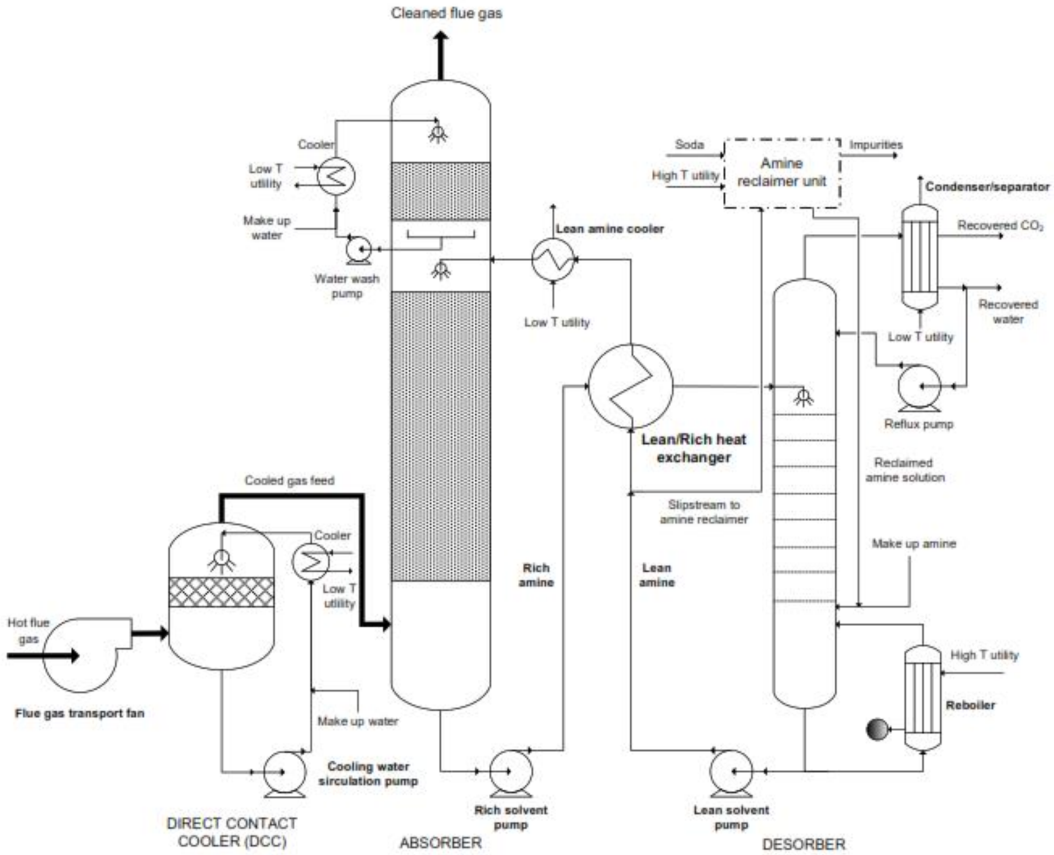


Figure 2.2 : General flow diagram of a CO_2 removal process plant [13]

2.4 Earlier works on process simulation and cost optimization in CO_2 capture

In 2007, Lars Erik Øi utilized Aspen HYSYS to model a basic combined cycle gas power plant along with a monoethanolamine (MEA)-based CO_2 removal process. His research focused on analyzing the CO_2 removal efficiency, energy consumption, and cost optimization parameters of the MEA-based absorption and desorption methods for capturing CO_2 from the exhaust gas of a gas-based power plant. By varying parameters such as amine circulation rate, absorption column height, absorption temperature, and steam temperature, Øi estimated the total CO_2 removal efficiency and heat consumption. The project aimed to identify the cost-optimal operating parameters of the CO_2 removal process. [2] [14]

In 2010, Kallevik in his master's thesis at Telemark University College, introduced an innovative approach for calculating make-up water and monoethanolamine (MEA) usage, as well as modeling a direct contact cooler (DCC) unit. For the purpose of cost estimation, the study computed the total heat transfer coefficient and correction factors for heat exchangers. By altering process parameters such as the minimum approach temperature in the lean/rich heat exchanger, absorber packing height, and absorber gas supply temperature, fluctuations in costs were observed. Parametric studies were conducted using CO_2 removal efficiencies of 80%, 85%, and 90%. [13]

In 2017, Park and Øi made adjustments to the gas velocity and pressure drop in a CO_2 absorption column for a typical amine-based CO_2 capture system by using the Aspen HYSYS modeling tool. They explored six types of structured packings, including Mellapak 250X, 250Y, 2X, 2Y, Mellapak Plus 252Y, and Flexipac 2Y. Modeling results indicated that the cost-optimal gas velocity for all packings fell within the range of 2.0 to 2.5 m/s, resulting in a pressure drop of 10 to 15 mbar across the absorber.[15]

Øi et al. (2014) utilized Aspen HYSYS to model various absorption and desorption setups for amine-based CO_2 removal (at 85%) from a natural gas-fired power plant. Among the configurations explored—standard processes, split-stream operations, vapor recompression, and their combinations—the most cost-effective one was found to be a simple vapor recompression setup. Further studies by Aromada and Øi (2015) revealed that vapor recompression, either alone or combined with split-stream operations, could significantly reduce energy consumption, with vapor recompression emerging as the most energy-efficient approach. Additionally, they conducted energy optimization and economic analyses for CO_2 capture parameters, focusing on a natural gas-based power plant project in Mongstad, Norway, over a 20-year period. The vapor recompression option, with specific operational parameters, was identified as the most energy efficient. Moreover, a cost-benefit study (2017) suggested that a vapor recompression technique with different operational parameters was the most cost-effective solution. [16] [17] [18]

Ali et al. (2018) examined how surplus heat could optimize amine-based CO_2 capture from a cement plant, comparing full-flow and part-flow flue gas scenarios to simulate conventional amine-based CO_2 removal. They employed the Aspen In-Plant cost calculator to estimate expenses for these scenarios, using both the detailed factor approach and the Lang factor technique to determine the cost of CO_2 collection. They found that the most energy-efficient solution was the full-flow alternative. According to their analyses, diverting 60% of the flue gas flow into the capture facility was the most cost-effective condition using the Lang factor technique, while diverting 80% of the flue gas flow was favored by the detailed factor approach. Additionally, Ali et al. (2019) developed a cost estimation tool that assesses how different assumptions affect the overall cost of a capture plant. This tool utilizes a simple process flow diagram and equipment list as input, focusing on detailed installation criteria and equipment costs to calculate CAPEX, which forms the core of the cost estimation approach. [19], [20]

Hasan Ali, in his PhD thesis (2019), aimed to establish a comprehensive system for conducting techno-economic analysis, which would highlight essential components and demonstrate the impact of various technological and economic assumptions on the overall cost of a capture plant. The proposed approach was applied to a base scenario involving amine-based post-combustion CO_2 capture (at an 85% capture rate) from a cement plant's flue gas, yielding a capture cost of 63 €/tCO_2. Key contributors to the base case outputs were identified as steam cost, energy cost, and capital cost. The Enhanced Detailed Factor (EDF) technique was utilized to identify significant cost factors, while the Lang factor method was not suitable for providing such detailed information. Although the projected steam cost is sensitive to market variables like fuel price fluctuations, the study suggests that natural gas-based steam generation is expected to be more cost-effective compared to coal- or biomass-based steam generation. [21]

Øi et al. (2021) aimed to determine the feasibility of automated cost estimation and optimization in a conventional amine-based approach for CO_2 capture from the cement industry. They employed the Aspen In-Plant cost estimator and a detailed factor approach to evaluate the capital cost of CO_2 collection. Factors such as the number of absorption stages, the minimum temperature differential in the primary heat exchanger, and the percentage of CO_2 removed were considered. An optimal temperature differential of 10-15 °C in the primary heat exchanger was identified based on various parameters. Furthermore, the ideal column height was determined to be 12 stages (equivalent to 12 meters of structured packing) through simulations for each stage number. [22]
Aromada et al. (2021) investigated various CO_2 capture heat exchangers, utilizing Aspen HYSYS simulations to analyze an 85 percent CO_2 absorption and desorption process for flue gas from the cement industry. Each plant was equipped with its own type of lean/rich heat exchanger, and the study included cost optimization of different heat exchanger types. Three types of shell and tube heat exchangers, as well as two plate and frame heat exchangers, were examined. The findings suggested that implementing a plate and frame heat exchanger instead of the typical shell and tube heat exchanger could lead to significant reductions in both capital and operational costs. [23]

In another study, Aromada et al. (2021) studied the impact of installation factors and plant construction characteristics on the capital cost of an amine-based CO_2 collection system. Using the Enhanced Detailed Factor (EDF) approach, they evaluated how equipment installation parameters influenced capital costs for seven different approaches. The study revealed that a constant installation factor could result in overestimation of high-cost equipment and underestimation of lower-cost equipment. Despite various methodologies determining the ideal minimum temperature differential (ΔT_{min}) in the cross-exchanger to be 15 °C, the results indicated that the EDF method could be effectively utilized for estimating capital costs for both new plants and modifications. [24]

Aromada et al. (2020) evaluated the cost implications of utilizing six different types of heat exchangers as the lean/rich heat exchanger in an amine-based CO_2 collection system. They found that the gasketed-plate heat exchanger (G-PHE) offered significant space and cost savings compared to traditional shell and tube heat exchangers (STHXs). By replacing STHXs with G-PHEs, capture costs could be reduced by €5–€6/tCO_2, and even more significantly, over €6/tCO_2 with the finned double-pipe heat exchanger (FDP-HX). The study highlighted that steam cost had the most substantial impact on CO_2 collection costs in all scenarios. [25]

In another study by Aromada et al. (2022), a trade-off evaluation between energy cost and capital cost resulting from alternative temperature approaches in the cross-exchanger of a solvent-based CO_2 capture process was conducted to assess the efficacy of plate heat exchangers (PHEs) compared to traditional shell and tube types. The goal was to identify cost-saving and CO_2 emission-cutting potentials of various heat exchangers. For PHE scenarios, the recommended minimum temperature approach based on cost-saving considerations ranged from 4°C to 7°C. This approach resulted in low energy usage and indirect emissions. Based on their findings, the study recommended the use of plate heat exchangers for the cross-heat exchanger (with a temperature approach of 4–7°C), lean amine cooler, and circulation water cooler for the direct contact cooler (DCC) unit. [26]

Øi et al (2021) aimed to explore CO_2 capture through absorption into monoethanolamine (MEA) followed by desorption, employing three configurations: standard, vapor recompression, and simple split-stream (rich split). Their study utilized equilibrium-based modeling in Aspen HYSYS V10.0, leveraging flue gas data from a natural gas-based power plant. They achieved automation of energy and material balance through adjust and recycle blocks in Aspen HYSYS, while optimization was conducted by minimizing the total cost calculated in an Aspen HYSYS spreadsheet. Equipment costs were sourced from Aspen In-plant Cost Estimator V10.0, and the total investment cost was estimated using the enhanced detailed factor (EDF) method. Parametric studies were conducted on absorber packing height, minimum approach temperature in the main heat exchanger, flash pressure, and split ratio at an 85% capture efficiency for the three configurations. The calculated optimum process parameters for the standard process were found to be a 15 m packing height and a 13°C minimum approach temperature. For the vapor recompression case, a flash pressure of 150 kPa resulted in the lowest total cost, while the calculated optimum rich split ratio was 12%. Automated calculations relied on stable convergence of the simulations, with a specific challenge being the adjustment of amine recirculation to achieve a specified total capture rate.[5]

In another study by Øi et al. (2023), they simulated a standard process for CO_2 capture using an equilibrium-based model in Aspen HYSYS. The simulation was integrated with equipment dimensioning and cost calculation in a spreadsheet facility. A novel aspect of this work was the variation of Murphree efficiencies to achieve automatic optimization of absorber height. The optimum process was determined as the one with the minimum calculated sum of capital and operational costs over 25 years. The cost-optimal process parameters for the standard process were determined to be a 15 m absorber packing height, a 13 K minimum approach temperature, and an inlet gas temperature of 34°C. This study illustrated the feasibility of automatically calculating the optimum packing height and inlet temperature by adjusting the Murphree efficiency in a case study function.[6]

2.5 Challenges in simulation and cost optimization

This study builds upon previous research conducted by the University of South-Eastern Norway (USN) in the fields of simulation and cost optimization, particularly focusing on CO_2 capture using Aspen HYSYS software. USN's prior investigations serve as a foundation for this project, which aims to automate the cost optimization of a CO_2 capture process employing MEA. Thus, it's essential to examine the factors that impact the overall cost of such a plant.

The cost of a plant is predominantly influenced by five key factors. Firstly, the volume of exhaust gas entering the absorption column impacts the size of the process equipment along the gas route. Secondly, the level of CO_2 in the flue gas is significant, as higher concentrations lead to greater energy efficiency due to increased driving force. Thirdly, as the rate of CO_2 removal rises, so does energy consumption. Fourthly, the size of equipment and energy requirements are dictated by the flow rate of the solvent. Lastly, the demand for hot water and electricity is influenced by the solvent flow rate, with higher rates necessitating more thermal energy. Additionally, electricity consumption, primarily from the flue gas transit through the process, increases with higher pressure requirements and volume flows.[13]

This study aims to conduct a sensitivity analysis to identify how alterations in various process parameters can impact the overall cost of the carbon capture plant, aiming to achieve cost savings. Parameters such as the flue gas temperature entering the absorber, absorber pressure, minimum temperature difference in the lean/rich heat exchanger, reboiler and condenser temperatures, reflux ratio, solvent circulation rate, desorber pressure, and the efficiency of CO_2 removal rate are typically considered in such analyses, as outlined by Øi (2007)[2].

3 Simulation and cost estimation

The primary goal of this study is to implement automated cost optimization through simulation for CO_2 capture, focusing on MEA absorption. A critical aspect involves simulating the CO_2 capturing process with the target of achieving an 85% removal efficiency. This chapter will detail the methodology and software employed, along with the base case model for cost evaluation and optimization in the sensitivity analysis.

3.1 Methodology and software

The utilization of an amine solvent is the predominant and extensively researched method for CO_2 removal, with MEA being the most extensively studied solvent due to its rapid CO_2 interaction and commercial availability, along with its high CO_2 capacity. Despite its advantages, MEA also presents drawbacks such as susceptibility to corrosion, toxicity, and degradation. A typical process flow diagram for an amine-based CO_2 removal facility is depicted in Figure 3.1. The conventional absorption process occurs within a column equipped with plates, random packing, or structured packing, where the CO_2-laden gas ascends while the absorption liquid descends. Subsequently, the rich amine solvent undergoes desorption in a separate column via heat exchange. Within the desorption column, the absorbed CO_2 is regenerated, facilitated by heating in a reboiler and reflux provided by a condenser. The regenerated lean amine solvent is then recirculated back to the absorption column after passing through a heat exchanger and cooler. [2]

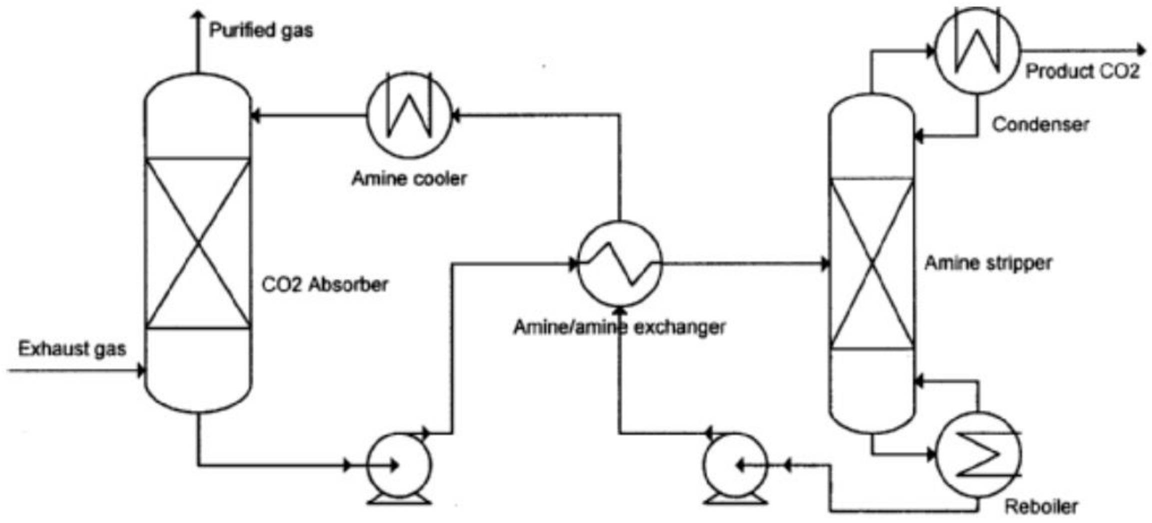


Figure 3.1 : Principle for CO removal process based on absorption in amine solution [2]

In this study, Aspen HYSYS version 12 was employed to simulate a conventional amine-based CO_2 capture process. The simulated outcomes were then utilized to determine equipment sizes and estimate costs using methodologies consistent with those described in existing literature [24], [27]. Throughout all simulations, the Acid Gas property package was utilized, which incorporates vapor and liquid equilibrium models for electrolytes. This package, designed to supersede the widely used Amine property package, employs the electrolyte non-random two-liquid (e-NRTL) model for electrolyte thermodynamics and the Peng-Robinson equation of state for the vapor phase.

3.2 Simulation Base case

The absorber and desorber were modeled using equilibrium stages with user-defined stage efficiencies, specifically Murphree efficiencies. A constant Murphree efficiency of 0.15 for the absorber and 0.5 for the desorber was implemented, corresponding to one meter of structured packing. As per the findings from the literature review, these Murphree efficiencies were derived by comparing the change in CO_2 mole fraction between consecutive stages with the change in the assumed equilibrium. This method, rather than assuming perfect equilibrium, offers a more accurate portrayal of performance. [2], [21]

Data from previous research [16], [24] on a natural gas-powered plant project in Mongstad, Norway, were used to create the base case scenario for the simulations. The parameters outlined in Table 3.1 represent an 85% CO_2 removal efficiency and a minimum temperature approach of 8°C in the lean/rich amine heat exchanger, which serves as the default configuration. The computational method mirrors that of earlier investigations [17], [22], [24]. The absorption and desorption columns were simulated as equilibrium stages with stage efficiency considered. The absorber was designed with 15 packing stages, while the desorber had four. Equilibrium stages with a height of 1 meter were employed for both columns. Murphree efficiencies of 15% were applied in the absorption column, while a uniform Murphree efficiency of 50% was assigned to all stages of the desorption column. The Modified HYSIM Inside-Out approach was utilized in the columns for facilitating convergence. The adiabatic efficiencies of the pump and flue gas fan were specified as 75%.

Table 3.1: Aspen HYSYS model parameters and specifications for the base case simulation

Stream/Equipment	Parameters [Unit]	Value
Inlet flue gas to absorber	Temperature [°C]	40
	Pressure [kPa]	121
	Molar flow rate [kgmol/h]	85036
	CO_2 content [mole %]	3.75
	H2O content [mole %]	6.16
	N2 content [mole %]	90.1
Lean Amine	Temperature [°C]	40
	Pressure [kPa]	101

Stream/Equipment	Parameters [Unit]	Value
	Mass flow rate [kmol/h]	2.5e+06
	MEA content [W %]	29
	CO_2 content [W %]	5.4
Absorber	Number of stages	15
	Murphree efficiency [m^{-1}]	0.15
	Rich amine pump pressure [kPa]	500
Heat Exchanger	Minimum temperature approach [°C]	8
	Rich amine temp. out of HEX. [°C]	104.3
Desorber	Number of stages	4
	Murphree efficiency [m^{-1}]	0.5
	Reboiler temperature [°C]	120
	Pressure [kPa]	200
	Lean amine pump pressure [kPa]	500
	Reflux ratio in stripper [-]	0.4

To develop an automated simulation model, robust adjustment and recycling manipulators are essential to ensure the convergence of the simulations. Traditionally, manual adjustments involve trial and error, especially when dealing with complex simulation models. The calculation sequence closely resembles that of previous simulation studies [13], [17], [23], [24].

The process begins with supplying the absorption column with the flow rate of input flue gas and making an initial estimate of the necessary amount of lean amine. The rich amine pump then transfers the rich amine from the bottom of the absorption column through the lean/rich amine heat exchanger. Following the heat exchange, the temperature is specified, and the rich amine is directed to the desorber. The CO_2 product and hot lean amine are then computed in the desorption column. The heated lean amine passes through the lean/rich heat exchanger, is pressurized in the lean amine pump, and undergoes further cooling in the lean cooler. Water is introduced into the process (water makeup), and its makeup is determined through a water material balance. The lean amine is subsequently routed to a recycle block (RCY-1), where it is assessed whether the recycled lean amine's flow and condition closely match the previously estimated lean amine stream, which may be adjusted iteratively.

To automate the simulation model, three adjustment blocks were defined into the flowsheet., ADJ-1 permits adjustment of the minimum approach temperature in the lean/rich heat exchanger based on the rich amine outlet temperature, ADJ-2 was specifically established to address makeup water, due to slight loss of water within the process commands the addition of additional water to maintain optimal process. and ADJ-3 allows for adjusting the removal efficiency based on the lean amine flow rate. The default tolerances provided in Aspen HYSYS were utilized in the simulations. The Aspen HYSYS simulation process flow diagram (PFD) is shown in Figure 3.2.

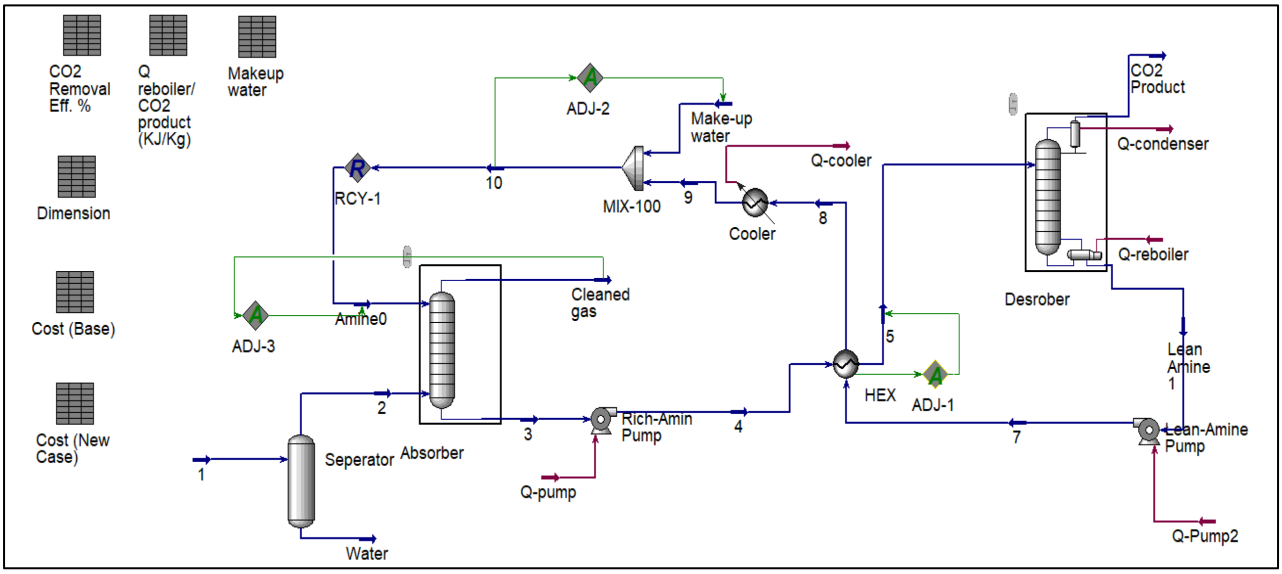


Figure 3.2: Process flow diagram (PFD) for the Aspen HYSYS base case simulation

The base case simulation presented in this chapter excludes the water wash and amine reclaimer processes, as well as the inlet flue gas fan and DCC (direct contact cooler) equipments depicted in Figure 2.2. Additionally, trace components beyond CO_2, N2, and H2O were not considered in the flue gas. The scope of the simulation also omits additional treatments of the desorber overhead product, such as water separation, drying, and compression.

3.3 Equipment dimensioning and assumptions

In this section, the process equipment outlined in the base case of the process simulation is described with basic dimensions. These dimensions are determined using the results of energy and material balances generated from Aspen HYSYS. This information is crucial for initiating cost estimations. The equipment sizing calculations were performed in a spreadsheet named "dimensioning," which was subsequently used for cost estimation purposes.

In the base case, only essential components like absorption and desorption columns, heat exchangers, pumps, cooler, condenser, and reboiler will be sized. This study excludes any pre-treatment, like purifying inlet gas, or post-treatment, such as compressing, transporting, or storing CO_2. The cost estimate solely encompasses the specified equipment installation costs. Additional expenses such as land acquisition, preparation, service buildings, and operational expenditures are not accounted for.

3.3.1 Absorber and desorber

Absorption, considered the primary process in CO_2 removal plants, involves the transfer of gaseous phase species into a liquid solvent. The area of absorber and desorber is determined by the actual volumetric gas flow rate ($\dot{\boldsymbol{V}}$ [$\boldsymbol{m^3/s}$]), and the diameter is calculated based on this area, assuming a circular shape. The highest flow rate stage dictates the actual volumetric gas flow rate, which is determined from the simulation. The gas velocity ($\dot{\boldsymbol{v}}$ [$\boldsymbol{m/s}$]) is a known parameter, assumed to be 2.5 m/s in the absorber and 1 m/s in the desorption column [15]. By using equation (33-1), the cross-sectional area can be estimated, subsequently by assuming a circular shape for column, can then be derived the diameter.

$$\boldsymbol{A} = \frac{\dot{\boldsymbol{V}}}{\dot{\boldsymbol{v}}_{\boldsymbol{gas}}} \ (\boldsymbol{m^2}) \tag{33-1}$$

3.3.2 Heat Exchangers

One of the most important and costly pieces of equipment is the main lean/rich amine heat exchanger. The purpose of using this heat exchanger in the removal process is to recover heat. Specifically, the rich amine from the bottom of the absorber is warmed by the hot lean amine leaving the stripper sump. This increase in the temperature of the rich amine reduces the reboiler duty. [13]

The heat transfer area is commonly used as a key factor in estimating the cost of heat exchangers. This area is influenced by the duty ($\dot{\boldsymbol{Q}}[kW]$), the overall heat transfer coefficient (U), and the logarithmic mean temperature difference ($\Delta \boldsymbol{T_{lm}}$ [°C]). To calculate the heat transfer area (A [m²]), Equation (3-2) is applied. [28]

$$\boldsymbol{A} = \frac{\dot{\boldsymbol{Q}}}{\boldsymbol{U}.\Delta \boldsymbol{T_{lm}}} \ (\boldsymbol{m^2}) \tag{3-2}$$

The heat transfer areas of the heat exchangers are determined using the duties and temperature conditions obtained from the simulations, assuming ideal countercurrent flow. The logarithmic mean temperature difference ($\Delta \boldsymbol{T_{lm}}$) is calculated using Equation (3-3), while Equations (3-4) and (3-5) describe the inlet temperature configuration for countercurrent flow.

$$\Delta \boldsymbol{T_{lm}} = \frac{\Delta \boldsymbol{T_1} - \Delta \boldsymbol{T_2}}{\mathbf{ln}\left(\frac{\Delta \boldsymbol{T_1}}{\Delta \boldsymbol{T_2}}\right)} \ (°C) \tag{3-3}$$

$$\Delta \boldsymbol{T_1} = \boldsymbol{T_{hot\ in}} - \boldsymbol{T_{cold\ out}} \ (°C) \tag{3-4}$$

$$\Delta \boldsymbol{T_2} = \boldsymbol{T_{hot\ out}} - \boldsymbol{T_{cold\ in}} \ (°C) \tag{3-5}$$

The overall heat transfer coefficients were defined as follows: 550 W/(m²·K) for the lean/rich amine heat exchanger, 800 W/(m²·K) for both the lean amine cooler, 1200 W/(m²·K) for the reboiler, and 1000 W/(m²·K) for the condenser. The heat exchangers utilized in this study are standard shell and tube heat exchangers (STHXs). [21], [26]

In this work, the calculated total area for the lean/rich heat exchanger is quite large. Therefore, it is divided into smaller units. The assumption for maximum heat exchanger area is 1000 m² per unit, as this is the most commonly available size in the market. Consequently, the number of smaller heat exchangers needed is calculated. [26]

3.3.3 Pumps and fan

The amine-based CO_2 capture process typically requires two pumps: the rich amine pump and the lean amine pump. The rich amine pump transfers the rich amine from the absorber's sump to the top of the stripper, while the lean amine pump provides the necessary energy to send the lean amine from the bottom of the desorption column to the top of the absorption column. Together, these centrifugal pumps must supply enough energy to overcome various losses in the pipes, heat exchangers, absorber, and desorber columns.

In this study, pumps and fan are assumed to have a 75% adiabatic efficiency. The duty obtained from Aspen HYSYS simulation is used as the dimensioning parameter for the pumps and fans, while the volumetric flow is also employed in the Aspen In-Plant cost estimator to determine the equipment cost. For the fan, the maximum allowable flow is limited to 1.529 E+6 m³/hr., which is used to estimate the required number of units. The fan's output pressure is assumed to match the pressure of the rich amine leaving the bottom of the absorber.

It is worth mentioning that, for the base case design of the carbon removal process, no fan or compressor was included before the absorber column. Only two pumps, for lean amine and rich amine, were used to compensate for pressure drops in the system. However, a fan was considered when examining the impact of inlet gas velocity. A comparison of these two different setups and their costs will be presented in the following chapters.

3.4 Cost estimation and assumptions

This chapter provides a comprehensive explanation of the methods used to estimate the cost of a CO_2-capture operation. The procedure, applied automatically to estimate the total cost of the plant for both base case and future case studies, involves the following steps:

- Equipment cost: Using Aspen In-Plant Cost Estimator (v.12) to calculate equipment costs based on their dimensions for the Base Case.
- Total installation cost: Applying the Enhanced Detailed Factor (EDF) to determine the total installation cost for each piece of equipment.
- CAPEX: Estimate the capital expenditure (CAPEX)
- Cost index correction: Adjusting the cost index to account for changes over time.
- OPEX: Estimating the annual operational expenditure (OPEX).
- Total annual cost: Calculating the total annual cost using a specified discount rate and plant lifetime.

- CO_2 capture cost: Determining the cost of CO_2 captured.
- Cost scaling: Utilizing the Power Law approach to scale costs for different case studies.

The cost estimation is restricted to the equipment detailed in the Aspen HYSYS flowsheet shown in Figure 3.2. The study excludes costs for pretreatment (e.g., incoming gas purification) and post-treatment processes (e.g., CO_2 compression, transport, or storage). Only the costs for the specified installed equipment are included.

3.4.1 Capital expenditure (CAPEX)

The Enhanced Detailed Factor (EDF) approach is utilized to estimate capital costs (CAPEX) in this study. This method takes into account various elements that influence the installation of each piece of process equipment. In the initial stages of cost estimation, this approach is effective and yields accurate estimates. Additionally, it facilitates techno-economic evaluations of both new and existing technologies and plants. [20]

3.4.1.1 Equipment material cost

Cost of the main equipment was determined using the Aspen In-Plant Cost Estimator (v12.0). This software estimates equipment costs by considering material, dimensioning variables, and process data, with the accuracy of the estimate depending on the number of parameters provided by the user. For unspecified parameters, Aspen In-Plant used default values. The cost estimates provided by Aspen In-Plant (v12.0) are in Euros (€) for the first quarter of 2019.

To calculate the installation cost of each piece of equipment, we utilized an installation factor sheet created by Nils Henrik Eldrup [24]. This sheet includes all installation variables and specifies that equipment costs should be estimated in carbon steel (CS). While Aspen In-Plant allows the selection of different materials, the costs need to be converted to carbon steel using a material factor ($\boldsymbol{f_{mat}}$) as per the EDF approach (Equation (3-6)). [20]

$$\boldsymbol{Equipment\ cost_{CS}} = \frac{\boldsymbol{Equipment\ cost_{SS}}}{\boldsymbol{f_{mat}}} \tag{3-6}$$

The material factor is used to convert the cost of equipment made from a specific material such as stainless steel to its equivalent in carbon steel. Except for the flue gas fan, which is made of carbon steel, all equipment is assumed to be made of stainless steel (SS304L). The material factor to convert the cost of SS304L to carbon steel is 1.75 for welded equipment and 1.30 for rotating equipment. [24]

3.4.1.2 Total installed cost

Each piece of equipment has a total installation factor ($\boldsymbol{F_{T,CS}}$), which encompasses various sub-factors including direct costs, engineering, administration, commissioning, and contingency [22, 56]. This is expressed in Equation (3-7).

$$\boldsymbol{F_{T,CS}} = \boldsymbol{f_{direct}} + \boldsymbol{f_{engineering}} + \boldsymbol{f_{administration}} + \boldsymbol{f_{comissioning}} + \boldsymbol{f_{contingency}} \tag{3-7}$$

The total equipment installed cost (EIC) is determined using Equation (3-8). The overall installed cost is then the cumulative sum of the installed costs for all equipment, as shown in Equation (3-9).

$$\boldsymbol{EIC_{CS} = F_{T,CS} * Equipment\ cost_{CS}} \tag{3-8}$$

$$\boldsymbol{Total\ installed\ cost = \sum (EIC\ for\ all\ pieces\ of\ equipment)} \tag{3-9}$$

For equipment made from materials other than carbon steel (CS), the installation factor must be adjusted. This adjustment is made using Equation (3-10):

$$\boldsymbol{F_{T,other\ mat.} = [F_{T,CS} + (f_{mat} - 1)(f_{eq} + f_{pp})]} \tag{3-10}$$

where $\boldsymbol{f_{eq}}$ is the equipment factor (equal to 1) and $\boldsymbol{f_{pp}}$ is the piping factor from the EDF table sheet. [26], [25], [24]

3.4.1.3 Cost inflation index

The Aspen In-Plant Cost Estimator uses equipment cost data from 2019. To provide an updated and accurate cost estimate, these figures need to be adjusted for inflation from 2019 to the end of 2023. Figure 3.3 present the indices used in this study. [27]

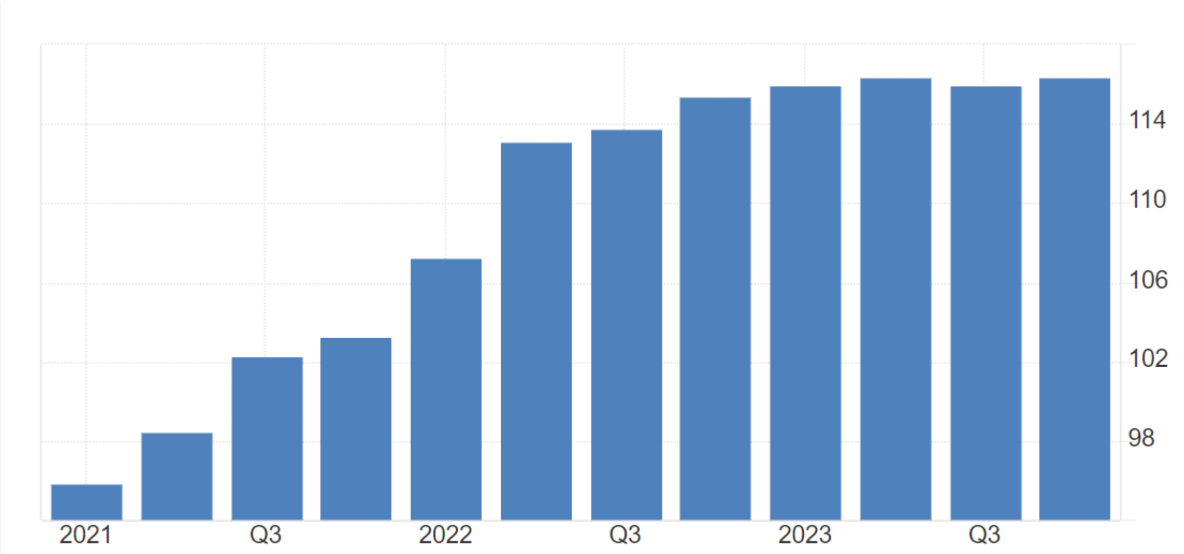


Figure 3.3: Euro Area - Construction cost index [29]

Equation (4.5) has been applied to update the CAPEX from 2019 to 2023. [25]

$$\boldsymbol{CAPEX_{2023} = CAPEX_{2019} \frac{cost\ index_{2023}}{cost\ index_{2019}}} \tag{3-11}$$

3.4.2 Operating expenditure (OPEX)

The cost of operations and maintenance represents a significant portion of the overall expenses. It is common practice to divide OPEX into fixed and variable costs. Fixed costs include maintenance and operational labor expenses. Maintenance costs are typically estimated as a percentage of the equipment installation cost (EIC), ranging from 2% to 6%, with this study using 4%. Operational labor costs are calculated based on the number of employees and their annual hours worked. Variable costs encompass raw materials, electricity, cooling water, steam, solvents, and other consumables. The annual cost for the specified utilities can be calculated using Equation (3-12). The OPEX assumptions and parameters are detailed in Table 3.2. [20], [6], [24], [26]

$$\textbf{Annual utility cost} = \textbf{Consumption}\left(\frac{\textbf{unit}}{\textbf{hr}}\right) * \frac{\textbf{Operating hours}}{\textbf{year}} * \textbf{utility price}\left(\frac{\textbf{Euro}}{\textbf{unit}}\right) \quad (3\text{-}12)$$

Table 3.2: Annual operating cost assumptions [6], [24], [26]

Item	Unit	Value
Operating lifetime	[Year]	25*
Annual hours of operation	[hr./year]	8000
Electricity cost	[euro/kWh]	0.06
Steam cost	[euro/kWh]	0.015
Cooling water cost	[euro/m^3]	0.022
Water process cost	[euro/m^3]	0.203
MEA cost	[euro/ton]	1450
Maintenance cost	[euro/year]	4% of CAPEX
Operator cost (6 person)	[euro/year]	80,414 (*6)
Engineer cost (1 person)	[euro/year]	156,650

* 2 years construction + 23 years operation

It is important to mention that all the procedures for calculating CAPEX, OPEX, and the cost of carbon capture per ton for the Base Case are recorded in a spreadsheet named “Base case” within the Aspen HYSYS flowsheet.

3.4.3 CO_2 captured cost

To evaluate different process design setups and select the optimal scenario from an economic and cost estimation perspective, it is essential to calculate and compare a unique economic value. In this study, the chosen metric is the annual CO_2 captured cost per ton, which can be calculated using equation (3-13). [20], [24]

$$\boldsymbol{CO2\ captured\ cost = \frac{Total\ annual\ cost}{Mass\ of\ CO2}\quad (\frac{euro}{ton})} \tag{3-13}$$

The total annual cost in equation (3-14) is the sum of the annualized CAPEX and OPEX.

$$\boldsymbol{Total\ annual\ cost = \ Annualized\ CAPEX + Annualized\ OPEX} \tag{3-14}$$

The annualized OPEX is derived from the sum of all fixed and variable costs. To calculate the annualized CAPEX, the operational lifetime and the discount rate must be determined. This can be done using Equations (3-15) and (3-16) as referenced in sources [24], [26].

$$\boldsymbol{Annualized\ CAPEX\left(\frac{euro}{yr}\right) = \frac{CAPEX}{Annualized\ factor}} \tag{3-15}$$

$$\boldsymbol{Annualized\ factor = \sum_{i=1}^{n}\left[\frac{1}{(1+r)^{n}}\right]} \tag{3-16}$$

In equation (3-16), $\boldsymbol{n}$ represents the operational lifetime of the CO_2 capture plant, and $\boldsymbol{r}$ denotes the discount rate.

3.5 Automated cost optimization

This section details the optimization methods and parameters explored during the case studies to determine the optimum scenario. A sensitivity analysis of our configuration was conducted for economic evaluation. To assess the impact of various variables on cost, a series of simulations were performed. These simulations evaluated the effects of the minimum temperature difference (ΔT_{min}) in the lean/rich amine heat exchanger, absorber packing height, and inlet flue gas velocity into absorber.

The primary aim of this research is to utilize the spreadsheet in Aspen HYSYS or Excel to automatically calculate and optimize various parameters based on the cost of an MEA-based CO_2 capture plant. Initially, CAPEX and OPEX were estimated for the Base Case in the corresponding spreadsheet. Equipment dimensions were imported from the Dimensioning spreadsheet to the Base Case (cost) spreadsheet, with relative equipment costs sourced from the Aspen In-Plant cost estimator. The total equipment installation cost was then determined using the EDF technique.

The equipment cost will be compared to the initial cost calculated using the EDF technique for the "Base Case" through the Power Law method. In this method, changes in equipment size or performance do not always have a linear relationship to costs; instead, costs are a function of capacity multiplied by an exponential ratio. This relationship can be represented by Equation (3-17) [30]:

$$\boldsymbol{\frac{Cost\ of\ B}{Cost\ of\ A} = (\frac{Capacity\ of\ B}{Capacity\ of\ A})^{e}} \tag{3-17}$$

In this equation, ***e*** is an exponential size factor typically ranging from 0.35 to 1.70, depending on the type of equipment [31]. For this study, the exponential factor is assumed to be 1.0 for the absorber and desorber columns and 0.65 for the other equipment. [6]

Utility usage data was automatically entered from the converged simulation case into the spreadsheet, enabling the calculation of the cost of carbon capture per ton for each scenario in a spreadsheet named “Cost (New case)”.

As previously mentioned in this report, Aspen HYSYS’s adjust and recycle blocks were employed to automate the energy and material balance for a given configuration, facilitating an automated simulation. In Aspen HYSYS, recycle blocks are used to solve the flowsheet by comparing the in-stream to the block with the stream from the previous iteration. Adjust functions are used to modify a parameter to achieve a desired outcome elsewhere in the simulation. [5], [6]

3.6 Case Studies

To investigate the impact of various variables on CO_2 capture cost, several case studies were conducted. The main goal of this study was to achieve automated cost optimization by integrating cost estimation procedures with process simulations for various scenarios. Initially, the approach involved using the case study option in Aspen HYSYS for specific variables, such as the minimum temperature approach. After obtaining converged solutions for each case, the necessary results were linked to Excel sheets to automatically perform dimensioning and cost estimation, allowing for comparison of the results.

However, this procedure encountered obstacles, prompting a change in the approach. The revised method involves first defining the base case and conducting a cost estimation within a spreadsheet in Aspen HYSYS. For each new scenario, the desired parameter is adjusted, and the cost estimation is performed directly within the model. The final results, such as the annual CO_2 capture cost per ton, reboiler duty, and CO_2 removal efficiency, are then connected to an Excel file to display and compare the outcomes effectively.

3.6.1 Heat exchanger minimum temperature approach (ΔT_{min})

The first case study was conducted to evaluate the economic performance of the lean/rich amine heat exchanger by adjusting the degree of heat recovery. This was measured using the minimum approach temperature (ΔT_{min}). In each scenario, ΔT_{min} was varied by keeping the reboiler output at a constant temperature of 120°C and changing the outlet temperature of the rich amine from the lean/rich amine heat exchanger. This adjustment was made in ADJ-1, while ADJ-3 aimed to maintain a constant CO_2 removal efficiency of 85%, and ADJ-2 aimed to keep the MEA flow rate at the desired value by adjusting the makeup water mass flow rate. The temperatures assessed ranged from 4 to 18°C. Throughout the experiment, all flue gas and absorption column parameters were kept constant to ensure a specific total CO_2 removal efficiency, as were the rate and composition of the lean amine flow.

3.6.2 Absorber packing height

To achieve the lowest CO_2 capture cost, the number of absorber stages varied from 13 to 18 in this study. Initially, the setup excluded the fan from the cost estimation at this stage, so the pressure drop was assumed to remain unchanged while investigating the effect of packing height in the absorber. This will be considered in the next case related to inlet gas velocity. Additionally, each stage height was assumed to be 1 meter.

Since changing the number of stages in the absorber's design tab requires rerunning the tower, the Case Study option could not be used for sensitivity analysis of absorber height (stages). This limited the ability to automate the evaluation. The Murphree efficiency was set to 0.15 for all stages. In Aspen HYSYS simulations, new stages are given an efficiency value of 1, which had to be manually adjusted to 0.15 for each case.

3.6.3 Absorber inlet gas velocity

Another scenario examined in this study is the effect of the inlet gas velocity to the absorber, which was varied between 1.5 and 3 m/s. By adjusting other parameters, such as the amount of amine injected to achieve an 85% carbon capture efficiency, the impact of this specific parameter on the carbon capture cost in the process was analyzed to determine the most economical condition.

It is important to note that the pressure drop in the absorber is a function of the inlet gas velocity. Therefore, the pressure drop had to be calculated for each velocity. The results from studies by Park and Øi’s study [15], presented in Figure 3.4, were used as the basis for setting the pressures at the bottom and top of the absorber column. Additionally, the desired outlet pressure from the fan was set for each scenario. Another important point is that the effect of a pressure drop on the total cost needs to be examined more precisely. Hence, a fan unit was added to the flowsheet before the absorber (Figure 3.5). As expected, it was observed that the total cost changed significantly, which will be discussed in the next chapter.

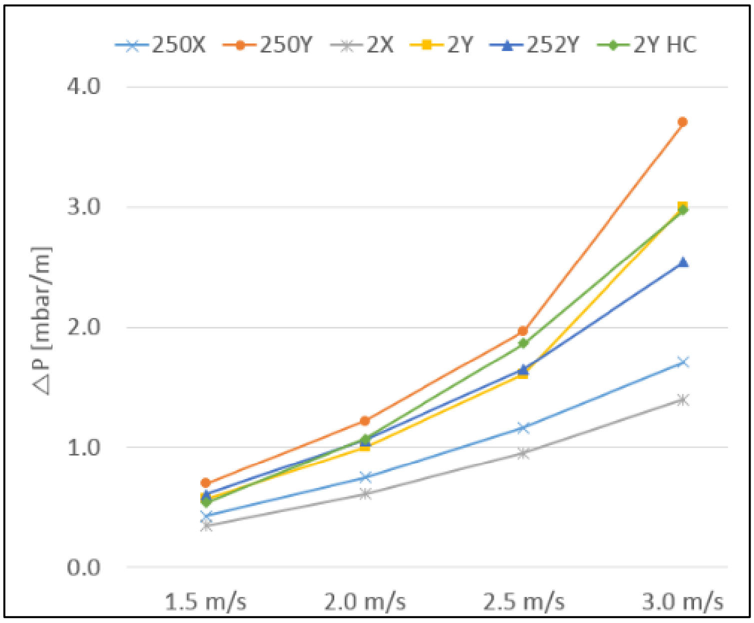


Figure 3.4: Pressure drop as a function of gas velocity for the different packings [15]

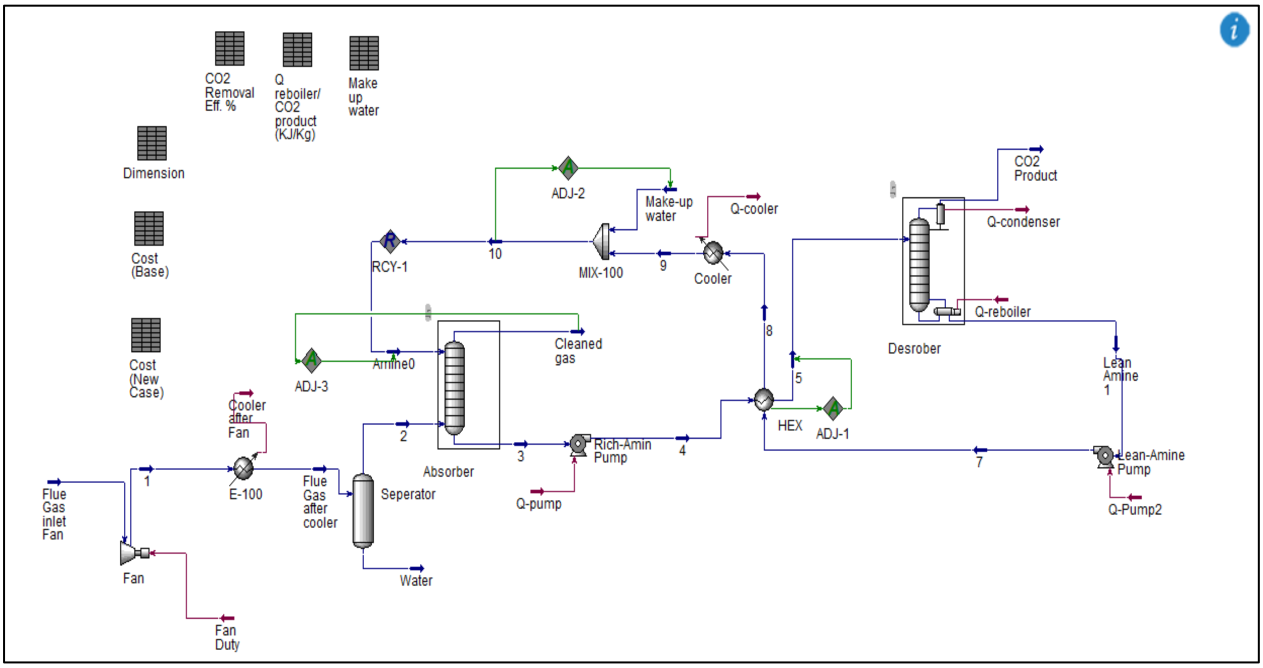


Figure 3.5: Process design flowsheet for inlet gas velocity case study

4 Results and Discussion

This chapter presents the results for the Base Case and performs a sensitivity analysis on various parameters that influence the outcomes for each case study. Initially, we will examine the simulation results and cost estimation for the base case scenario. Subsequently, we will evaluate the effects of specific parameters on the cost optimization of the amine-based CO_2 capture process, suggesting the most cost-effective scenario for each case and comparing it with previous results.

Following this, the heat and material balance for the base case will be detailed. We will then discuss the obstacles and challenges encountered in automating the process simulation and cost optimization of the amine-based CO_2 absorption process. Finally, we will propose directions for future work.

4.1 Base Case

Figure 4.1: Cost spreadsheet – Base case (CAPEX)Figure 4.1 shows the automated procedure for estimating CAPEX in the base case using an Aspen HYSYS spreadsheet. The total equipment cost (CAPEX) for the base case process design in 2023 is around 128 million euros (MEUR). The absorber and the lean/rich amine heat exchanger are the most costly components, each accounting for approximately 35% of the total CAPEX. Notably, the packing cost constitutes about 70% of the absorber's total expense. The reboiler with about 13 MEUR represents 10% of the overall cost. Figure 4.2 illustrates the cost distribution of the essential equipment in this process design in the Base Case study.

Spreadsheet: Cost (Base)

	A	B	C	D	E	F	G	H	I
1			Material Cost (SS) €	Material Cost (CS) €	Installation Factor	Installation Cost €	Number of Units	Total Equipmet Co...	
2			per Unit	per Unit		per Unit			
3	Hex		2.109e+005	1.205e+005	7.210	8.688e+005	41.00	3.562e+007	
4	Reboiler		2.085e+005	1.191e+005	7.210	8.590e+005	12.00	1.031e+007	
5	Cooler		1.103e+005	6.305e+004	8.690	5.479e+005	2.000	1.096e+006	
6	Condenser		6.078e+004	3.473e+004	10.20	3.544e+005	1.000	3.544e+005	
7	Absorber	Packing	1.276e+007	7.294e+006	3.453	2.518e+007	1.000	2.518e+007	
8		Material	5.095e+006	2.912e+006	3.453	1.005e+007	1.000	1.005e+007	
9	Desorber	Packing	3.091e+005	1.766e+005	5.300	9.360e+005	1.000	9.360e+005	
10		Material	5.203e+005	2.973e+005	5.300	1.576e+006	1.000	1.576e+006	
11	Rich Amine Pump		6.423e+004	4.941e+004	7.808	3.858e+005	1.000	3.858e+005	
12	Lean Amine Pump		8739	6722	15.95	1.072e+005	1.000	1.072e+005	
13	Unlisted equipment							1.712e+007	
14									
15							CAPEX @ 2019	1.027e+008	
16							CAPEX @ 2023	1.284e+008	
17									

Figure 4.1: Cost spreadsheet – Base case (CAPEX)

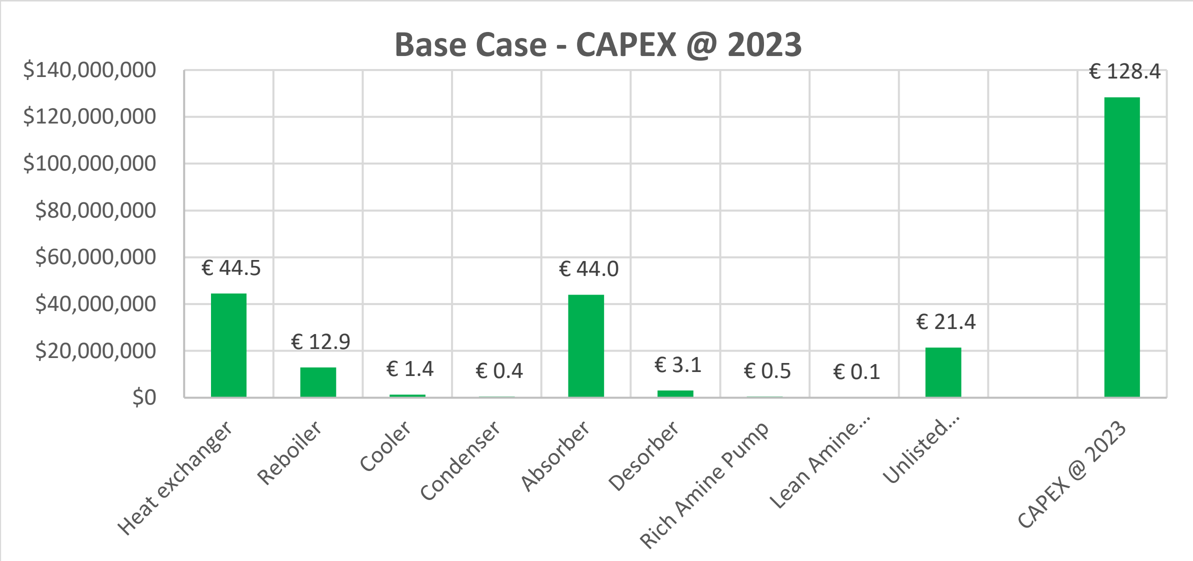


Figure 4.2: Equipment cost distribution – Base case (CAPEX)

To provide a comprehensive view of all costs, the calculation of operational expenditure (OPEX) was also conducted on the same spreadsheet used for CAPEX, as shown in Figure 4.3. The annual OPEX for the Base Case is approximately 30 MEUR. The most significant utility cost is steam (Reboiler), which amounts to about 20 MEUR annually, representing around 66% of the total OPEX.

In addition to steam, other substantial OPEX components include the cost of maintenance, and electricity for pumps, and monoethanolamine (MEA) as depicted in Figure 4.4. These costs highlight the importance of optimizing utility usage to reduce overall operational expenses. Furthermore, understanding the distribution of OPEX components aids in identifying potential areas for cost savings and efficiency improvements in the CO_2 capture process.

Spreadsheet: Cost (Base)

	A	B	C	D	E	F	G	H	I
17									
18	Operation Lifetime	23.00	yr					Annual Operation...	
19	Annual Hours of O...	8000	hr/yr			Reboiler	4.920e+008 kJ/h	1.640e+007	
20	Electricity Cost	0.6000	Euro/KWh			Condenser	8.576e+007 kJ/h	1.806e+005	
21	Steam Cost	1.500e-002	Euro/KWh			Cooler	8.811e+007 kJ/h	1.855e+005	
22	Cooling Water Cost	2.200e-002	Euro/m3			Rich Amine Pump	329.0 kW	1.579e+006	
23	Water Process Cost	0.2030	Euro/m3			Lean Amine Pump	91.06 kW	4.371e+005	
24	MEA Cost	1450	Euro/Ton			MEA	619982.5790 kg/h	8.990e+005	
25	Operator Cost (6 P...	8.041e+004	Euro/yr			Operator		4.825e+005	
26	Enginner Cost (1 P...	1.567e+005	Euro/yr			Engineer		1.567e+005	
27	Water Density	1000	Kg/m3			Maintenance		4.110e+006	
28	CO2 Product (Mas...	119236.1645 kg/h	Kg/h						
29	CO2 Removal eff.	85.01 %	%				OPEX @ 2019	2.443e+007	
30	Construction inde...	93.00	-				OPEX @ 2023	3.052e+007	
31	Construction inde...	116.2	-						
32	CCI (2019-2023)	1.249							
33									

Figure 4.3: Cost spreadsheet – Base case (OPEX)

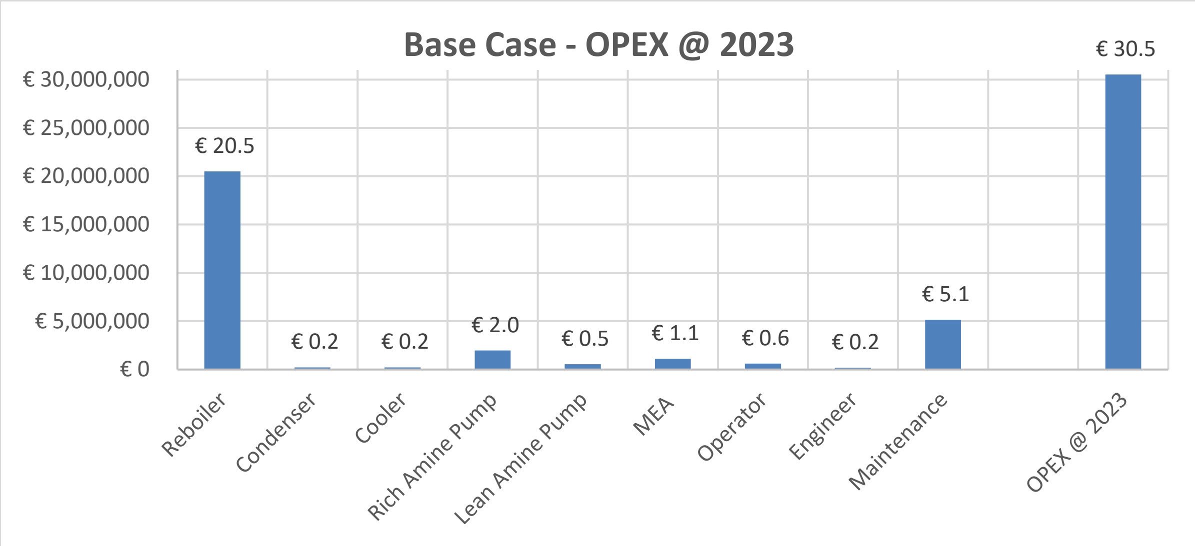


Figure 4.4: Operational cost distribution – Base case (OPEX)

In the Base Case analysis, the estimated cost of removing one ton of CO_2 from the flue gas was 44 EUR. The reboiler's duty was 4126 kJ per kilogram of captured CO_2.

34					
35		**2019**	**2023**		
36	**Discount rate**	**7.500**	**7.500**	**%**	
37	**Annualized factor**	**11.15**	**11.15**	**-**	
38	**Annualized CAPEX**	*9.214e+006*	*1.151e+007*	**Euro**	
39	**Total Annual Cost**	*3.364e+007*	*4.204e+007*	**Euro**	
40	**Mass of Capture C...**	*9.539e+005*	*9.539e+005*	**Tons**	
41	**CO2 Capture Cost...**	*35.27*	*44.07*	**Euro/Ton**	
42					

Figure 4.5: Estimated cost of removing one ton of CO_2 from the flue gas – Base case

To evaluate the accuracy of the convergence simulation model, material and energy balance calculations performed by Aspen HYSYS for the base case are shown in Figure 4.6 and Figure 4.7, respectively. The results indicate that the relative imbalance for both material and energy balances is below 5%, which is within an acceptable range.

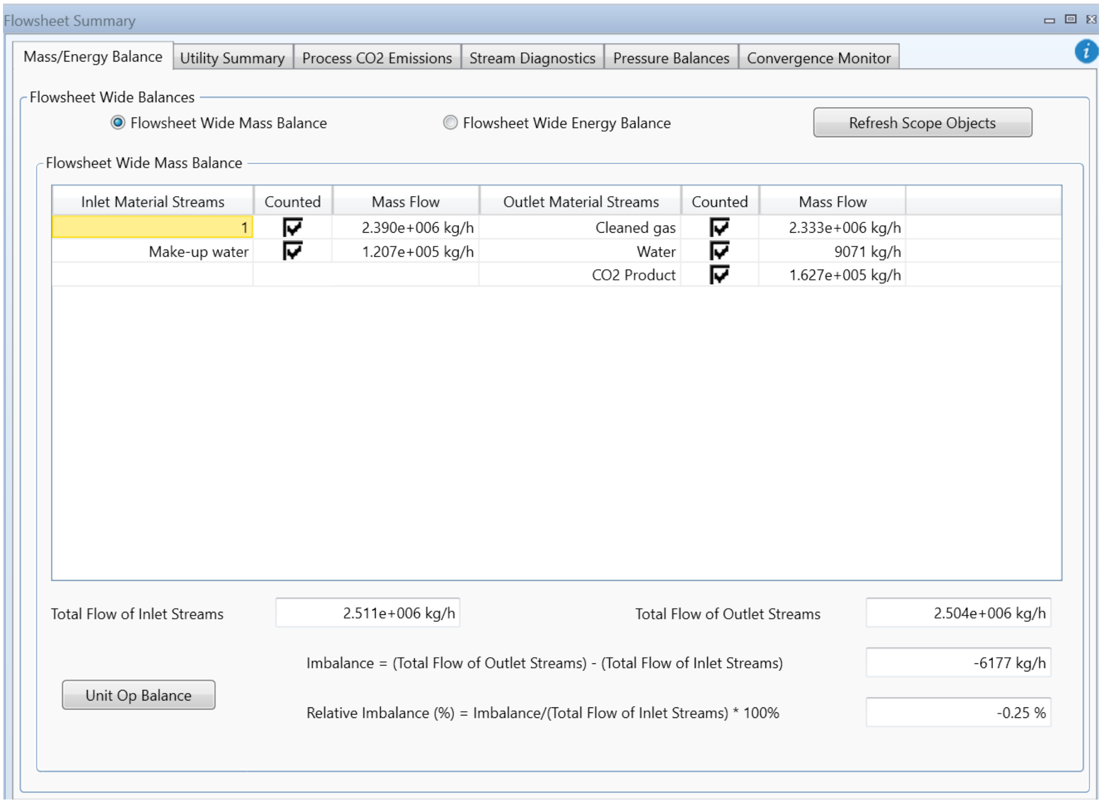


Figure 4.6: Material balance calculations, the base case

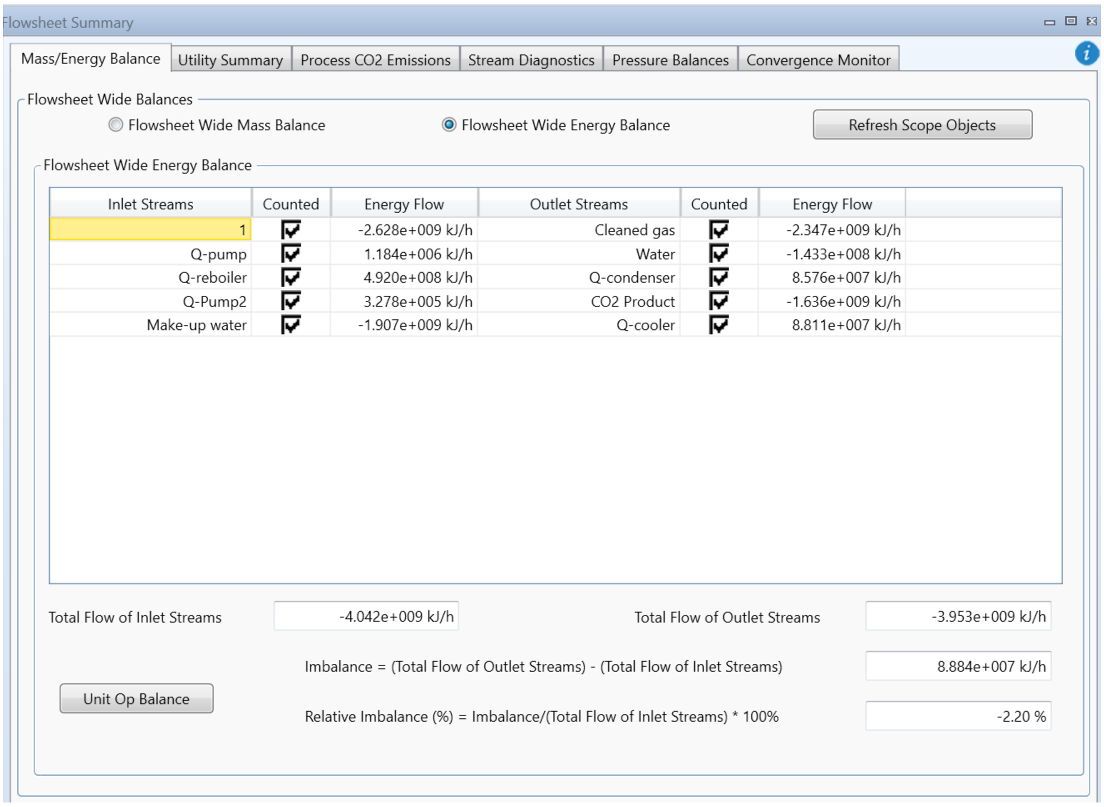


Figure 4.7: Energy balance calculations, the base case

4.2 Case study 1: ΔT_{min} in heat exchanger

In this case, the reboiler duty and CO_2 capture cost were assessed as the minimum temperature approach (ΔT_{min}) of the lean/rich amine heat exchanger varied from 4 to 18 ℃. For each simulation case, only the target value in ADJ-1, which is ΔT_{min}, was adjusted during the simulation. All other Adjust and Recycle blocks were left unchanged and remained active, allowing for automatic calculation for each case. Figure 4.8 illustrates the calculations conducted during the sensitivity analysis for the ΔT_{min}.

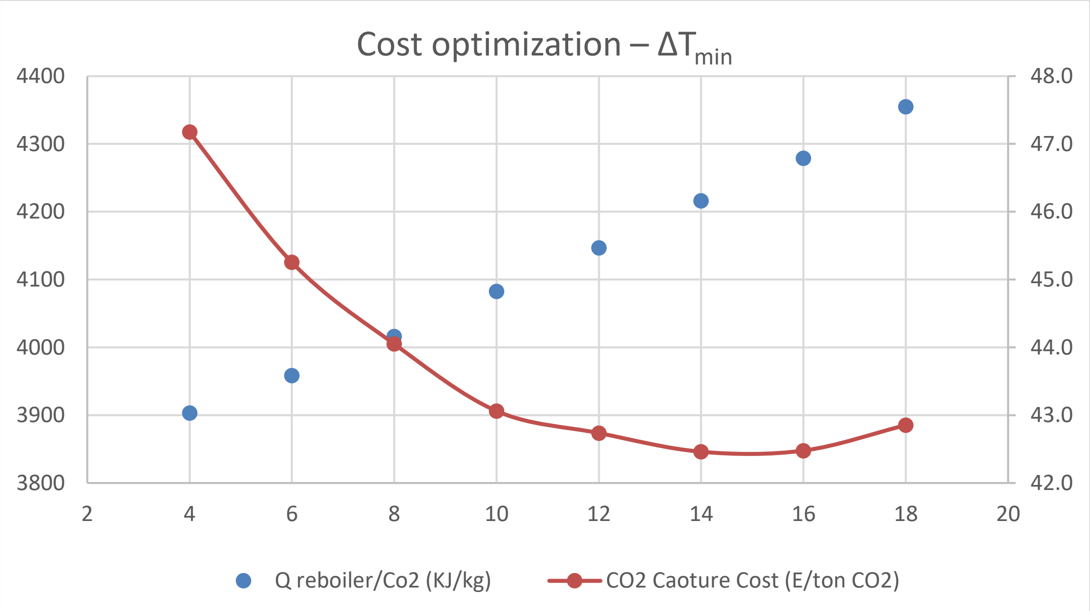


Figure 4.8: Cost optimization – minimum temperature approach case (ΔT_{min})

According to Figure 4.8, the optimal ΔT_{min} is approximately 15 °C, where the capture cost is around 42.5 EUR/t and the reboiler's duty is 4216 kJ/kg, and it indicates that increasing ΔT_{min} steadily raises steam consumption in the reboiler.

4.3 Case study 2: Packing height in absorber

In this scenario, the impact of absorber's packing height (number of stages) has been investigated. Figure 4.9 illustrates the variations in reboiler duty and CO_2 capture cost as the absorber packing height is changed from 13 to 18 meters. The analysis involved modifying the number of stages in the absorber, with the Murphree efficiency for CO_2 manually set to 15% for the additional stages. The automated adjustment was carried out by ADJ-3, which altered the amine flow rate to maintain an 85% CO_2 removal rate in the absorber. All adjust and recycle blocks were active during the simulation to ensure automatic and robust computations, leading to a converged solution.

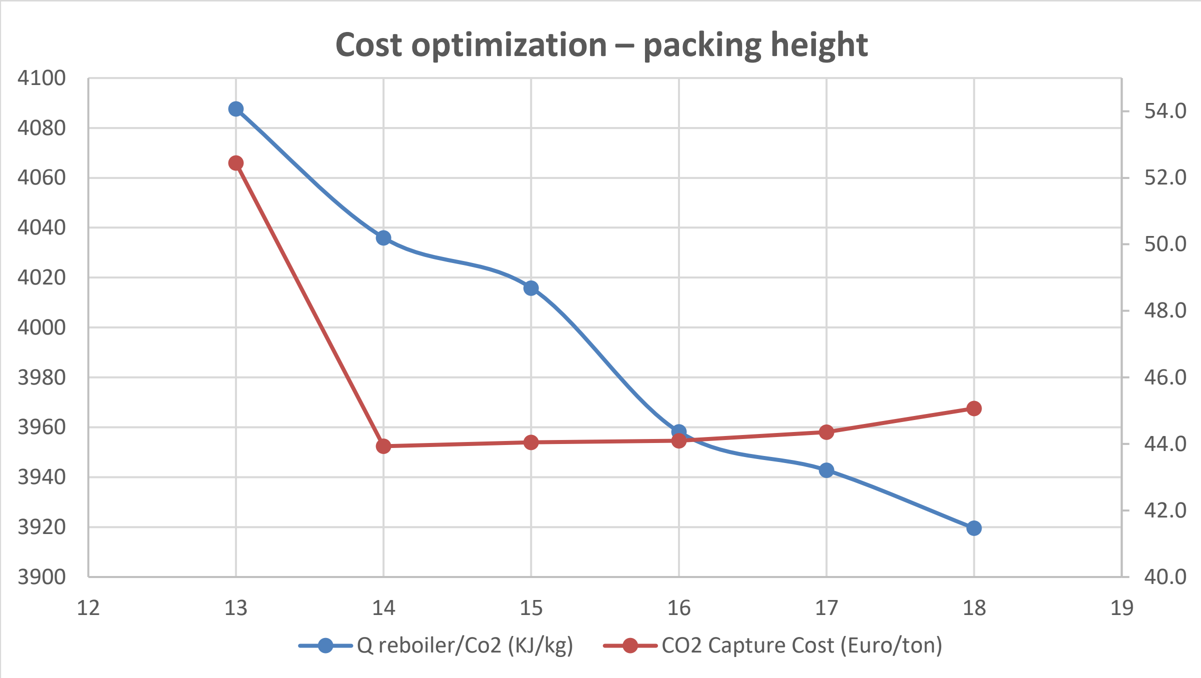


Figure 4.9: Cost optimization –packing height

According to Figure 4.9, a packing height of 14 meters proved to be the most optimal, with a CO_2 capture cost of 43.9 EUR/t and a reboiler duty of 4036 KJ/kg CO_2.

It is worth mentioning that with an increase in packing height, the pressure drop in the absorber column naturally increases. However, for this analysis, the pressure drop was considered constant for all cases, which reduces the accuracy of the calculations to some extent.

4.4 Case study 3: inlet gas velocity in absorber

For the sensitivity analysis of inlet gas velocity into the absorber, we first need to establish the relationship between gas velocity in the absorber and the resulting pressure drop. Referring to Figure 3.4 from Park and Øi's study [15], a pressure drop estimation for "Mellapak 250X" was conducted for each gas velocity. Table 4.1 lists the pressure drop for each inlet gas velocity that was simulated.

Table 4.1: The pressure drop corresponding to gas velocity (Mellapak 250X).

Gas Velocity (m/s)	Pressure drop (Kpa)
1.5	18.2
2	19.0
2.5	20.0
3	22.6

This estimated pressure drop was manually set in the simulation model to perform a sensitivity analysis and determine the cost-optimal inlet gas velocity for the absorber. Additionally, compensating for the pressure drop in the absorber involves significant power consumption by the fan, a costly piece of equipment that greatly impacts the final CO_2 capture cost. Therefore, it was decided to include the fan unit in the process design (Figure 3.5) and proceed with the simulation and cost estimation of the entire CO_2 removal plant. The estimated CO_2 capture cost and reboiler's duty for this analysis are shown in Figure 4.10.

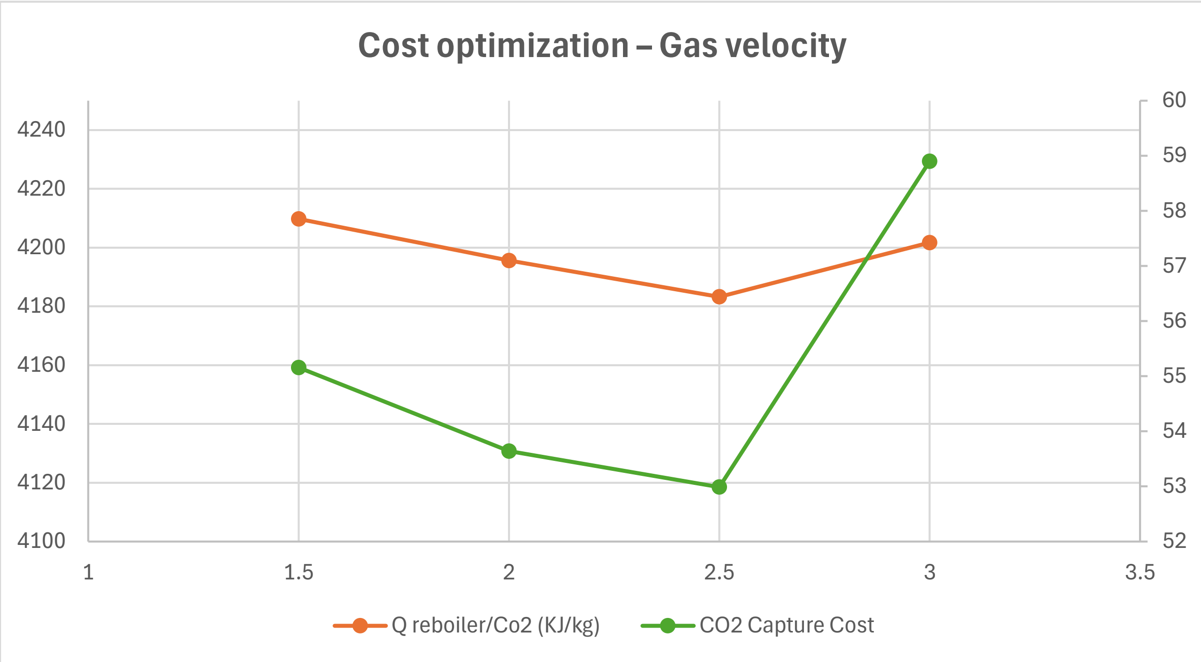


Figure 4.10: Cost optimization – gas velocity

Based on the results observed in Figure 4.10, an inlet gas velocity of 2.5 m/s for the absorber column can be suggested as the most cost-optimal case. The carbon capture cost at this velocity is 52.99 EUR/ton, which is the lowest among the other cases. Additionally, the reboiler duty is at its minimum possible value of 4183 KJ/kg CO_2.

4.5 Discussion

In this section, the results will be analyzed and compared with those of earlier studies. The precision and reliability of these results will be evaluated, and suggestions for future research will be proposed.

4.5.1 Results comparison

The carbon capture cost of 44 EUR/ton for the "Base case" (process design without a fan and separation unit before the absorber) is close to the 40-43 EUR/ton of CO_2 captured reported by Øi et al. [5], [6]. Additionally, the cost of around 53 EUR/ton for the gas velocity case that includes a fan aligns with previous results, such as the 50 EUR/t stated in publications by Ali et al [20] and Aromada et al. [24], [26]. The discrepancies between these calculated costs may be attributed to differences in process design boundaries and simplification assumptions used in the design of the CO_2 removal plant.

4.5.1.1 Base case

Table 4.2 presents a comparison of the Base Case results from this study with several recent studies that used the same amine composition, specifically 20-30 wt.% MEA solvent, and a CO_2 mole concentration of approximately 3.75% in the flue gas.

Table 4.2: Comparison of simulation results with previous studies

Study	CO_2 removal Eff. [%]	CO_2 mole fraction in flue gas [%]	ΔT_{min} [°C]	Absorber packing height [m]	Reboiler duty [KJ/kg CO_2]	Capture cost [EUR/ton]
This study (Base Case)	85	3.75	8	15	4126	44
Øi et al. [6]	85	3.75	10	15	3750	43
Aromada et al., [18]	85	3.73	10	20	3600	NA
Amrollahi et al., [32]	90	3.8	8.5	13	3740	NA
Øi et al. [12]	85	3,75	10	10	3650	NA

4.5.1.2 ΔT_{min} in heat exchanger

This section explores the balance between the size of the lean/rich heat exchanger and the external utility requirements, with the heat exchanger area changing based on the minimum temperature difference (ΔT_{min}). The main trade-off is between the capital cost of resizing the lean/rich heat exchanger and the operational cost due to variations in steam consumption. These adjustments impact the temperature entering the desorber and the cooling required in the lean amine heat exchanger. According to Øi [33], a lower ΔT_{min} reduces steam and reboiler duty, while a higher ΔT_{min} lowers heat exchanger costs. This analysis seeks to identify the cost-optimal minimum approach temperature for the lean/rich amine heat exchanger.

This case study concentrated on automated calculations to determine the optimal ΔT_{min} for the lean/rich amine heat exchanger. The findings indicated that the cost-optimal scenario lies between 14 and 16 degrees Celsius. According to Øi et al. [22], changing conditions can slightly adjust the optimal temperature approach to between 10 and 15 degrees Celsius, with 13 °C often cited as the optimal value in several related studies [7], [15], [19], [20], [6] , [22].

4.5.1.3 Packing height in absorber

This study calculated the optimal packing height of the absorber to be 14 stages, consistent with the findings of earlier studies [5], [6]. Husebye et al. [21] noted that low CO_2 concentrations (below 5%) in flue gas are more costly due to the need for more solvent and higher stages to achieve a high CO_2 removal efficiency of around 85% or more. Therefore, higher CO_2 concentrations should naturally reduce costs. This suggests a potential area for future research: evaluating the impact of varying CO_2 concentrations in flue gas entering the absorber on cost optimization, while maintaining constant CO_2 removal efficiency.

The relation between the amine flow rate and absorber packing height, noting that the solvent flow rate increases as the number of absorber stages decreases. The absorber size and the heat transfer area in the lean/rich heat exchanger are two major capital costs affected by this change. The amine flow rate adjustment impacts the cost of the lean/rich heat exchanger, while steam consumption, and amine consumption are the main operational costs influenced.

Regarding steam consumption, sensitivity analysis shows that reboiler duty decreases as the number of stages increases. This is due to a reduction in the amine circulation rate, which drops significantly from 13 to 14 stages and then gradually declines with additional stages.

4.5.1.4 inlet gas velocity in absorber

The simulation results of this study show a trend in the relationship between inlet flue gas velocity and carbon capture cost. As the inlet gas velocity increases from 1.5 to 2.5 m/s, a decline in carbon capture cost is observed. This trend indicates that higher gas velocities enhance the efficiency of CO_2 absorption processes, leading to lower overall costs associated with carbon capture. However, beyond a certain point, specifically at a gas velocity of 2.5 m/s, further increases in gas velocity result in diminishing returns. Beyond this optimal point, increasing the gas velocity leads to reduced absorption efficiency, subsequently increasing the carbon capture cost. This finding underscores the importance of carefully balancing operational parameters such as gas velocity to achieve optimal cost-effectiveness in CO_2 removal processes.

The observed phenomenon can be acknowledged to the relations between gas velocity and mass transfer kinetics within the absorber. At lower gas velocities, the contact time between the flue gas and the solvent within the absorber increases, facilitating more effective CO_2 absorption. Consequently, the carbon capture cost decreases because of improved absorption efficiency. However, as gas velocity continues to increase beyond the optimal point, the contact time decreases, limiting the extent of CO_2 absorption. This diminishing absorption efficiency leads to an increase in the overall carbon capture cost, as more energy and resources are required to achieve the desired level of CO_2 removal. Thus, optimizing the inlet flue gas velocity represents a crucial factor in achieving cost-effective CO_2 capture processes.

4.5.2 Accuracy, uncertainties and limitations

Several factors related to simulation, dimensioning, and cost estimating assumptions contribute to the accuracy and uncertainties in this study's results:

- Choosing the boundary of the CO_2 removal plant for simulation and cost optimization is a crucial parameter for assessing and achieving a cost-effective process design. Accurate boundary selection ensures that all relevant factors, including energy consumption, equipment costs, and process efficiencies, are thoroughly evaluated. In this study, the boundaries were defined to focus on key elements such as the amine flow rate, absorber and desorber performance, and heat exchanger efficiency. The goal was not to pinpoint absolute cost values but to identify the optimal process parameters that minimize overall costs. This approach acknowledges the inherent uncertainties in specifications and scaling factors, such as the Murphree efficiency and the size exponent for equipment scaling, which can significantly impact the cost estimates. By clearly defining the plant boundaries, the study could systematically explore the effects of varying parameters, like inlet flue gas velocity and pressure drops, on the overall cost. This meticulous boundary-setting allows for a more reliable comparison of different process designs and ensures that the results are relevant and applicable to real-world scenarios, facilitating better decision-making for future CO_2 removal projects.
- Energy expenditure, especially heat consumption for CO_2 regeneration, has the most significant impact on operational costs. The uncertainty in total operating costs is almost directly proportional to the uncertainty in the value of this heat.
- The assumptions and specifications chosen can influence the estimated cost and the resulting optimal values. For instance, the cost and height of the absorber and desorber are affected by the type of packing used. The ΔT_{min} calculations depend on specified characteristics such as the overall heat transfer coefficient of the heat exchangers, particularly the lean/rich amine heat exchanger. Additionally, the assumptions regarding pressure drop in the absorber influence the fan outlet pressure and cost.
- Achieving convergence is challenging, especially in the absorber and desorber. In such cases, using the modified HYSIM Inside-Out solver can facilitate easier and faster convergence. However, this approach hindered the full automation of case studies, particularly for determining the optimal packing height. Automatically adjusting the appropriate amine circulation flow rate to achieve a specific capture rate Initially, default convergence criteria were used, and then tolerance and sensitivity in the adjust and recycle blocks were reduced to achieve higher accuracy and smoother results.
- One of the issues encountered during this study and the automation of cost estimation was the integration and synchronization of Excel with Aspen HYSYS. Initially, efforts were made to create a simulation model, define a case study, and export the results directly and link them with Excel to automate the cost estimation and optimization process. However, it was observed that although an add-in for Aspen HYSYS exists in Excel, the results do not update automatically and tend to have bugs. Therefore, it was decided to conduct the entire dimensioning and cost estimation process for the base case, as well as use the power

law method for estimating the costs of simulation cases, within Aspen HYSYS's own spreadsheet. This approach has several limitations for writing formulas within the cells, increasing the likelihood of calculation errors, but it can be said that the results are generated semi-automatically.

- The approach adopted in this thesis involved optimizing one parameter at a time. This required the use of Adjusts and Recycle blocks for automatic assessment. In some cases, achieving convergence in the simulation model was challenging, necessitating adjustments to tolerance, step size, and iteration time to ensure the model's robustness. This issue was particularly evident during sensitivity analysis for inlet flue gas velocity and packing height.

4.5.3 Future work

Here are some suggestions for future studies to enhance the robustness and accuracy of simulation and cost estimation:

- As mentioned earlier, this study evaluated and optimized one parameter at a time. Future research could focus on the simultaneous optimization of multiple parameters, which would be a more challenging yet realistic approach. Simultaneous optimization introduces complexities and requires greater precision, consistency, and robustness in calculations.
- Developing a method to directly change the number of stages (packing height) for this analysis could be beneficial. It should be feasible to update the number of stages during simulations. However, this approach must address challenges related to column and flowsheet convergence to ensure reliable and accurate results.
- A simple and basic simulation model was used for this study. For future work, there is potential to optimize energy consumption, and consequently, to achieve cost optimization for the entire CO_2 capture process.
- It was observed that Aspen HYSYS has a built-in option for cost estimation of the designed process. However, when used, its estimates were significantly inaccurate compared to the final cost calculated using a combination of simulation, Aspen In-Plant Cost Estimator, and the EDF method. In the future, there may be a way to refine and adjust the Aspen HYSYS cost estimation module to achieve more accurate and robust cost optimization for each process design and case study.

5 Conclusion

The primary objective of this study was to develop a method for automating the cost optimization process simulation, specifically using Aspen HYSYS to model an amine-based CO_2 capture process. This was based on emission data from prior research on a natural gas-powered plant project at Mongstad, Norway. A base case scenario was created with an absorber packing height of 15 meters, a desorber packing height of 4 meters, a target CO_2 removal efficiency of 85%, and a minimal temperature difference (ΔT_{min}) of 8°C in the lean/rich amine heat exchanger. The Enhanced Detailed Factor (EDF) method, along with the Aspen In-Plant Cost estimator, was employed to calculate the overall cost for this base case scenario. The results for the base case indicated a total cost of 44 EUR/ton of CO_2 captured and an energy consumption of 4126 kJ/kgCO_2 in the reboiler.

The study conducted case studies to identify cost-optimal scenarios, examining parameters like minimum temperature approach, absorber packing height, and inlet flue gas velocity in amine absorption.

The first case study optimized the lean/rich amine heat exchanger's economic performance by adjusting the minimum approach temperature (ΔT_{min}) from 8 to 18°C. The optimal ΔT_{min} was around 15°C, with a CO_2 capture cost of 42.5 EUR/ton and a reboiler duty of 4216 kJ/kg. This study highlighted the balance between heat exchanger size and steam consumption, revealing that increasing ΔT_{min} raises steam usage. This result aligns with literature values, where varying conditions can shift the optimal ΔT_{min} to between 10 and 15°C, with 13°C frequently noted as ideal in related studies.

The second case study aimed to optimize the absorber packing height by varying the number of stages from 13 to 18 meters to achieve the lowest CO_2 capture cost. Initially, the fan was excluded from the cost estimation, assuming a constant pressure drop, with the fan considered in subsequent analysis on inlet gas velocity. The optimal packing height was found to be 14 meters, yielding a CO_2 capture cost of 43.9 EUR/ton and a reboiler duty of 4036 KJ/kg CO_2. The study manually set the Murphree efficiency for CO_2 at 15% for each stage and adjusted the amine flow rate to maintain an 85% CO_2 removal efficiency. Although the constant pressure drop assumption reduced calculation accuracy, the results align with previous studies.

Another case study investigated the impact of varying inlet gas velocity to the absorber, ranging from 1.5 to 3 m/s, on CO_2 capture cost. Pressure drop calculations were conducted for each velocity, considering data from Park and Øi and accounting for pressure at the absorber's top and bottom. A fan unit was integrated to address power consumption resulting from pressure drop, affecting overall costs. Results indicated that an inlet gas velocity of 2.5 m/s proved most cost-effective, with a CO_2 capture cost of 53 EUR/ton and a minimum reboiler duty of 4183 KJ/kg CO_2. Higher velocities initially reduced costs by enhancing CO_2 absorption efficiency, but beyond 2.5 m/s, reductions were observed due to decreased contact time between flue gas and amine solvent, leading to increased costs.

References

[1] "Cost estimation methods for CO2 capture processes." Accessed: May 05, 2024. [Online]. Available: https://openarchive.usn.no/usn-xmlui/handle/11250/3024523

[2] "Removal of CO2 from exhaust gas - PHD Thesis - Lars Erik Øi." Accessed: May 01, 2024. [Online]. Available: https://openarchive.usn.no/usn-xmlui/bitstream/handle/11250/2437805/PhDThesis.pdf?sequence=2

[3] "IPCC Special Report on Carbon Dioxide Capture and Storage." Accessed: May 29, 2024. [Online]. Available: https://archive.ipcc.ch/publications_and_data/_reports_carbon_dioxide_graphics.htm

[4] P. Choudhary, *Carbon Capture and Storage Program(CCSP) Final report 1.1.2011–31.10.2016*. 2016.

[5] L. E. Øi, A. Haukås, S. Aromada, and N. Eldrup, "Automated Cost Optimization of CO2 Capture Using Aspen HYSYS," presented at the The First SIMS EUROSIM Conference on Modelling and Simulation, SIMS EUROSIM 2021, and 62nd International Conference of Scandinavian Simulation Society, SIMS 2021, September 21-23, Virtual Conference, Finland, Mar. 2022, pp. 293–300. doi: 10.3384/ecp21185293.

[6] L. E. Øi, S. Shirdel, S. Karunarathne, and S. Aromada, "Process Simulation, Dimensioning and Automated Cost Optimization of CO2 Capture," presented at the 64th International Conference of Scandinavian Simulation Society, SIMS 2023 Västerås, Sweden, September 25-28, 2023, Oct. 2023, pp. 54–61. doi: 10.3384/ecp200008.

[7] E. S. Rubin *et al.*, "A proposed methodology for CO2 capture and storage cost estimates," *International Journal of Greenhouse Gas Control*, vol. 17, pp. 488–503, Sep. 2013, doi: 10.1016/j.ijggc.2013.06.004.

[8] 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.

[9] Aromada, Solomon Aforkoghene, "Cost estimation methods for CO2 capture processes - PHD Thesis - Aromada, Solomon Aforkoghene." 2022.

[10] 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," *International Journal of Greenhouse Gas Control*, vol. 78, pp. 400–412, Nov. 2018, doi: 10.1016/j.ijggc.2018.07.015.

[11] D. Y. C. Leung, G. Caramanna, and M. M. Maroto-Valer, "An overview of current status of carbon dioxide capture and storage technologies," *Renewable and Sustainable Energy Reviews*, vol. 39, pp. 426–443, Nov. 2014, doi: 10.1016/j.rser.2014.07.093.

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

[13] O. B. Kallevik, "Cost estimation of CO2 removal in HYSYS," 2010.

[14] "Full Text PDF." Accessed: May 09, 2024. [Online]. Available: https://www.researchgate.net/profile/Lars-Oi/publication/228402007_Aspen_HYSYS_simulation_of_CO2_removal_by_amine_ab

sorption_from_a_gas_based_power_plant/links/5a155d70a6fdccd697bc214e/Aspen-HYSYS-simulation-of-CO2-removal-by-amine-absorption-from-a-gas-based-power-plant.pdf

[15] K. Park and L. E. Øi, “Optimization of gas velocity and pressure drop in CO2 absorption column,” presented at the The 58th Conference on Simulation and Modelling (SIMS 58) Reykjavik, Iceland, September 25th – 27th, 2017, Sep. 2017, pp. 292–297. doi: 10.3384/ecp17138292.

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

[17] S. Aforkoghene Aromada and L. Øi, “Simulation of improved absorption configurations for CO2 capture,” presented at the The 56th Conference on Simulation and Modelling (SIMS 56), October, 7-9, 2015, Linköping University, Sweden, Nov. 2015, pp. 21–29. doi: 10.3384/ecp1511921.

[18] 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.

[19] H. Ali, L. E. Øi, and N. H. Eldrup, “Simulation and Economic Optimization of Amine-based CO_2 Capture using Excess Heat at a Cement Plant,” presented at the The 59th Conference on imulation and Modelling (SIMS 59), 26-28 September 2018, Oslo Metropolitan University, Norway, Nov. 2018, pp. 58–64. doi: 10.3384/ecp1815358.

[20] 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,” *International Journal of Greenhouse Gas Control*, vol. 88, pp. 10–23, Sep. 2019, doi: 10.1016/j.ijggc.2019.05.028.

[21] Ali, Hassan, “Techno-economic analysis of CO2 capture concepts,” PHD Thesis, 2019.

[22] 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,” presented at the SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland, Mar. 2021, pp. 326–331. doi: 10.3384/ecp20176326.

[23] S. Aforkoghene Aromada, N. H. Eldrup, F. Normann, and L. E. Øi, “Simulation and Cost Optimization of different Heat Exchangers for CO2 Capture,” presented at the SIMS Conference on Simulation and Modelling SIMS 2020, September 22-24, Virtual Conference, Finland, Mar. 2021, pp. 318–325. doi: 10.3384/ecp20176318.

[24] 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,” *International Journal of Greenhouse Gas Control*, vol. 110, p. 103394, Sep. 2021, doi: 10.1016/j.ijggc.2021.103394.

[25] 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, p. 6315, Nov. 2020, doi: 10.3390/en13236315.

[26] S. A. Aromada, N. H. Eldrup, and L. E. 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, no. 2, p. 425, Jan. 2022, doi: 10.3390/en15020425.

[27] S. Fagerheim, "Process simulation of CO2 absorption at TCM Mongstad," 2019.

[28] F. P. Incropera, D. P. DeWitt, T. L. Bergman, and A. S. Lavine, Eds., *Principles of heat and mass transfer*, 7. ed., International student version. Hoboken, NJ: Wiley, 2013.

[29] "Euro Area - Construction cost index - 2024 Data 2025 Forecast 2000-2023 Historical." Accessed: May 24, 2024. [Online]. Available: https://tradingeconomics.com/euro-area/construction-cost-idx-eurostat-data.html

[30] Clayton T.Baumann, "Cost-to-Capacity-Method_Applications-and-Considerations," 2014.

[31] J. W. Smith, R., *Smith, R. (2005). Chemical Process Design and Integration. John Wiley & Sons Ltd.* 2005.

[32] 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," *International Journal of Greenhouse Gas Control*, vol. 8, pp. 1–11, May 2012, doi: 10.1016/j.ijggc.2012.01.005.

[33] "Removal of CO2 from exhaust gas." Accessed: May 24, 2024. [Online]. Available: http://hdl.handle.net/11250/2437805