

# Impact of Implementing CCS Technology on a Biomass Power Plant

Master's Thesis

By

Azin Mohebbi

Department of Energy Sciences  
Faculty of Engineering, LTH, Lund University  
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Supervisor: **Hesam Fatehi, Lund University**

Co-supervisor: **Sadiq Al-Kaisy, Sweco Consultancy**

Examiner: Jens Klingmann



## **Abstract**

This study examines the implementation of carbon capture and storage (CCS) technology, specifically the monoethanolamine (MEA) chemical absorption method, in a biomass power plant. The aim is to assess the energy requirements and optimize the solvent processes to improve the overall efficiency of the carbon capture process. The use of bioenergy with carbon capture and storage (Bio-CCS) is proposed as a solution to mitigate global warming by reducing CO<sub>2</sub> concentrations in the atmosphere.

The shift in Sweden's heat and power generation infrastructure towards biomass and waste makes Bio-CCS a crucial element in achieving a sustainable and carbon-negative energy system. The study employs mass and energy balance calculations to model the implementation of CCS technology on the flue gas stream of the biomass power plant. This allows for the analysis of heating, cooling, and electricity requirements during the pre-cooling, carbon capture, and compression stages.

The findings highlight the significant influence of lean solvent loading on process performance, particularly in terms of thermal energy requirements. Optimizing the solvent processes requires careful consideration of lean solvent loading as a key factor. The conclusions drawn from this study have implications beyond biomass power plants, as they can be applied to CO<sub>2</sub> removal in various plant types. The research contributes to the understanding of energy requirements in MEA carbon capture technology and suggests potential avenues for enhancing efficiency and cost-effectiveness in the carbon capture process.



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## Abbreviations and symbols

### Abbreviations

BECCS	Bio Energy with Carbon Capture and Storage
CC	Carbon Capture
CCS	Carbon Capture and Storage
CCU	Carbon Capture and Utilization
CHP	Combined Heat and Power
CO	Carbon monoxide
CO <sub>2</sub>	Carbon dioxide
GHG	Green House Gas
H <sub>2</sub>	Hydrogen
HCl	Hydrogen Chloride
IGCC	Integrated Gasification Combined Cycle
MEA	Mono Ethanol Amine
NO <sub>x</sub>	Nitrogen oxide
N <sub>2</sub>	Nitrogen
SCR	Selective Catalytic Reduction
SO <sub>x</sub>	Sulfur oxide

### Symbols

$\alpha$	Loading	[mol CO <sub>2</sub> /mol MEA]
$C_p$	Specific heat capacity	[kJ/(kg.k)]
$E$	Electricity requirement for the system	[We]
$\Delta H$	Enthalpy difference	[J/mol]
$\Delta T$	Temperature difference	[°C]
$M$	Molar mass	[g/mol]
$\dot{m}$	Mass flow rate	[kg/s]

$\dot{n}$	Molar flow rate	[kmol/s]
$P$	Pressure	[Pa]
$P_A$	Partial pressure of A	[Pa]
$P_A^{sat}$	Saturated pressure of A	[Pa]
$Q$	Heat added to the system	[J/kg]
$R$	Molar gas constant	[J/(mol.k)]
$T$	Temperature	[°C]
$\dot{V}$	Volumetric flow rate	[m <sup>3</sup> /s]
$W$	Work performed on the system	[J/kg]
$x_A$	Molar fraction of A in liquid phase	[-]
$y_A$	Molar fraction of A in gas phase	[-]



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# 1. Introduction

There is no doubt that global warming is an urgent environmental issue. Since the industrial revolution, carbon dioxide, the primary greenhouse gas, has continuously been released into the atmosphere. As a result, there is a pressing need to continue to curb the increasing concentrations of CO<sub>2</sub> in the atmosphere (Quan et al., 2023).

In order to minimize the environmental impact of fossil fuels, several efforts have been undertaken to improve their efficiency, develop innovative energy conversion devices, and develop cost-effective renewable energy sources such as solar, biomass, wind, and geothermal energy that have little or no environmental impact (Wilberforce et al., 2021). Carbon capture and storage (CCS) is another technology that has emerged as a promising solution for reducing greenhouse gas emissions. In 1977, CCS was proposed to reduce CO<sub>2</sub> emissions using a three-stage process - the capture, transportation, and storage/utilization of CO<sub>2</sub> (Dubey & Arora, 2022). Alongside its impact on reducing the cost of energy, carbon capture and storage (CCS) holds the potential for rapid commercial viability. Most CCS technologies have the capacity to absorb approximately 85-95% of the CO<sub>2</sub> emissions produced by power plants. (Ramboll, 2022).

Three leading CO<sub>2</sub> capture technologies exist; pre-combustion, post-combustion, and oxyfuel. In pre-combustion CC, CO<sub>2</sub> is removed before the combustion is completed, for instance by gasifying the fuel with oxygen, air, or steam to create a synthesis gas (syngas) that consists of carbon monoxide and hydrogen. Post-combustion capture process involves the separation of CO<sub>2</sub> from flue gases before they are discharged into the atmosphere. In the oxy-fuel capture process, pure oxygen is used for combustion instead of air, resulting in flue gas mixtures mainly consisting of CO<sub>2</sub> and condensable water vapor that can be separated and cleaned during compression (Wilberforce et al., 2021). Among these methods, post-combustion CC is the most suitable to scale up existing power plants, due to its maturity, cost-effectiveness, and ability to handle large volumes of flue gases. Post-combustion carbon capture can be done using amine-based solutions to absorb CO<sub>2</sub> from flue gases. This method depends directly on selecting an appropriate absorption solution. It is common in the industry to use primary, secondary, and tertiary amine solutions. Among the best absorbents, monoethanolamine (MEA) is characterized by a high absorption rate, high mass transfer, ability to capture 90% of CO<sub>2</sub>, low chemical cost, and

biodegradability. It can also capture CO<sub>2</sub> under low-pressure flue gases (at atmospheric pressure) (Zhao et al., 2023).

The original idea of CCS was to use it on fossil-fuel-fired power plants, allowing fossil fuels to be used while reducing CO<sub>2</sub> emissions, however, this has expanded in recent years (Stockholm Exergi AB,2020). One of the available strategies is using biofuels in conjunction with carbon capture and storage (BECCS), which has been found to be a promising alternative for reducing atmospheric carbon emissions significantly. Ultimately, BECCS aims to keep the recent atmospheric CO<sub>2</sub> concentration below 421 ppmv and to keep global temperature rises under 2°C. In addition to reducing atmospheric CO<sub>2</sub> concentration, using BECCS also offers “negative emissions” potential. Negative emissions refer to the human removal of CO<sub>2</sub> from the atmosphere globally and are not considered natural processes (Mishra et al., 2020).

The Nordic countries, including Finland, Sweden, and Denmark, have emerged as leaders in the development of negative emissions through the implementation of BECCS technologies (Rydén et al., 2017). With favorable geological formations for carbon storage and abundant sustainable biomass resources, the Nordic region offers an ideal environment for deploying these technologies, which have the combined potential to achieve negative emissions in energy production. Recent studies indicate that bioenergy has the capacity to become the predominant energy carrier in the Nordic countries by 2050. This integration of bioenergy with carbon capture not only helps in reducing CO<sub>2</sub> emissions from industrial processes and electricity generation but also presents an opportunity to establish a sustainable energy system that actively removes CO<sub>2</sub> from the atmosphere. (Rydén et al., 2017, Ramboll, 2022).

## **1.1. Aim and Objectives**

This study is conducted at LTH, Lund University, in conjunction with Sweco, to analyze carbon capture technology mass and energy balance in a power plant that uses biomass as a fuel. In addition, the key parameters that affect the energy needed for the process are studied. Moreover, it investigates the impact of NO<sub>x</sub> and SO<sub>x</sub> in flue gas on carbon capture technology. Furthermore, it explores the requirements for installing a CC unit in a biomass plant and its optimum operational conditions. Energy integration, heat loss and electricity demand are investigated as the main challenges in integrating a CC unit to an existing power plant. The technology that is chosen for the investigations in this study is the commercially available MEA carbon capture

technology. All stages of the process, including pre-cooling, capturing, and compression, have been designed in a streamlined manner. This approach ensures a simplified and efficient process flow, where only the crucial elements are considered and emphasized. In addition to providing retrofitting recommendations, the results will provide an estimated energy demand.

The objectives of this report are:

- To create a general model of mass and energy balance that can be used for energy assessment of different CC systems based on MEA.
- Change the lean solvent loadings to determine how much energy is needed for the reboiler to capture a specific portion of CO<sub>2</sub>.
- A comparison of the performance of BECCS plants with different flue gas compositions and provide recommendations on how to improve the MEA's carbon capture system's efficiency.

## **1.2. Scope and Limitations**

The model was developed in Microsoft Excel based on analysis of different databases to provide the necessary inputs. These input data include:

- Flue gas data (such as flue gas composition, temperature, and flow rate)
- Carbon capture rate
- Efficiencies of different equipment
- Number of compression stages

In this thesis the focus is not on CO<sub>2</sub> storage and utilization after separation. Also, the design and cost estimation of equipment were beyond the scope of this study. However, the thesis discusses the energy required for compression.

The company provided information about the flue gases' composition. The calculations were performed utilizing Microsoft Excel as the preferred tool for obtaining the results, as directed by the company. This deliberate choice of utilizing Excel, instead of specialized modeling software, was made in accordance with the company's preferences and requirements.

## 2. Carbon Capture

### 2.1. Capturing routes

Carbon dioxide (CO<sub>2</sub>) is a highly significant anthropogenic greenhouse gas (GHG) that plays a substantial role in global warming. Accounting for approximately 76% of the total greenhouse gas emissions, CO<sub>2</sub> is released in large quantities into the atmosphere as a by-product of fossil fuel combustion in power plants. Its prominent presence and substantial contribution to the global GHG inventory make it a critical focus in efforts to mitigate climate change. Global warming can negatively affect both the environment and ecosystems. (Mukherjee et al., 2019).

CCS is one of the critical technologies that holds immense potential for the future of sustainability and can be regarded as one of the key technologies in the future. A CO<sub>2</sub> Capture and Storage technique, which is often referred to as CCS, is a technology for capturing atmospheric CO<sub>2</sub>, either directly from the air—hereafter called direct air capture (DAC)—or from one or more point sources—such as a giant fossil fuel-powered thermal power plant. The captured CO<sub>2</sub> is then injected into the subsurface for storage. Furthermore, the capture of CO<sub>2</sub> can be achieved through various methods, encompassing both biological and physicochemical approaches. Among the physicochemical techniques, absorption, adsorption, and cryogenic distillation are frequently employed for the carbon dioxide capture process. Since biological methods offer a cleaner and more environmentally friendly approach to CO<sub>2</sub> collection, they are gaining significant traction and recognition within the field. (Bajpai et al., 2022).

Several projects are currently underway worldwide that involve the research and development of carbon capture technology. There are currently three main technical routes in this field of study (Figure 1): post-combustion, oxyfuel combustion, and pre-combustion combustion.



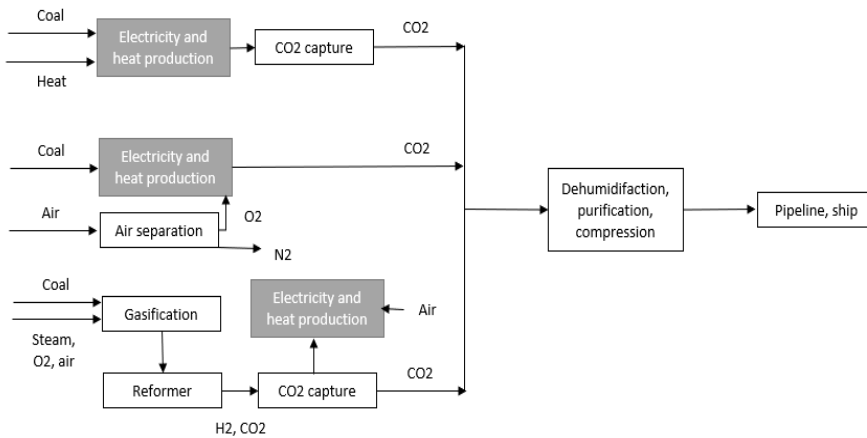


Figure 1- Carbon capturing routs (Adapted from Julich GmbH, 2015)

Numerous carbon capture processes exist, which necessitate extensive development before attaining implementation readiness. However, the mentioned processes also possess notable advantages, such as their potential for enhanced efficiency, compared to the current development strategies. These types of processes are often referred to as "second-generation" processes due to their efficiency (Julich GmbH, 2015). This section will provide a brief overview of the three technology lines that are currently being used in the industry.

### 2.1.1. Post-combustion capture

Using this approach, CO<sub>2</sub> is separated from the flue gases produced by large-scale fossil fuel combustion processes such as those in boilers, cement kilns, and industrial furnaces. In many power plants today, absorption is often done using chemical solvents like amine. A heat exchanger cools the flue gas to temperatures between 40 and 60°C, allowing it to be introduced to an absorber, where the gases bond with chemical solvents. Afterwards, the solvent rich in CO<sub>2</sub> is pumped into a stripper, which heats the solvent between 100 and 140°C to promote solvent regeneration (Wilberforce et al., 2021).

Post-combustion amine scrubbing has the advantage of being a mature technology due to its use in chemical industrial processes. Moreover, it is possible to improve efficiency in a significant way. A post-combustion process is perhaps the most effective CCS technology when it comes to capturing carbon dioxide since it achieves the highest degree of purity (over

99.99 %). Even though some modifications need to be made to the low-temperature steam part of the power plant process if the post-combustion process is to be integrated into one, it does not require fundamental changes in the design of the power plant. However, this solution has some significant disadvantages, such as the high investment costs (Julich GmbH, 2015).

### 2.1.2. Pre-combustion capture

The pre-combustion CC procedure involves the production of syngas (a mixture of hydrogen  $H_2$  and carbon monoxide  $CO$ ) from the fuel reforming process, followed by the separation of  $CO_2$  from it. The main processes leading to the formation of the synthesis gas are the reactions of fuel reforming and partial oxidation. The process involves removing sulfur and particulate matter to ensure that the catalyst stays active and operable (Wilberforce et al., 2021).

In the pre-combustion process, the solvent is an absorbent with a methanol basis. During absorption, pressures of 30 to 60 bars and temperatures of  $40^\circ C$  are needed (Kunze & Spliethoff, 2010). A result of the process is the capture of  $CO_2$  and hydrogen gas, which can then be burned as fuel, resulting in the release of water as the ultimate product of combustion (Wilberforce et al., 2021).

It is an advantage of the pre-combustion process that physical scrubbing is already commonly used on a large industrial scale for the purification of syngas and is, therefore, a direct analogy for Integrated Gasification Combined Cycle (IGCC) power plants. It is also possible to achieve high purities of  $CO_2$  using this process. Additionally, gasification, thermodynamic optimization, and the enhancement of gas turbine technology all contribute to the considerable efficiency potential of IGCC (BMW 2007). These plants, however, are highly complex when compared to conventional power plants. IGCC power plants also face very high investment costs, which also explains the low number of IGCC plants worldwide (Julich GmbH, 2015).

### 2.1.3. Oxyfuel combustion (OFC)

A CC based on oxy-fuel combustion produces nitrogen-free flue gases with only  $CO_2$  and  $H_2O$  as a result of burning fossil fuel in pure oxygen. A pure  $CO_2$  stream is produced by flue gas condensation, and  $NO_x$  gases are eliminated as well (Wilberforce et al., 2021).

It is estimated that oxyfuel power plants lose between 8% and 11% of efficiency. Furthermore, it is unclear whether existing plants can be retrofitted with oxyfuel processes. Since all necessary technical components are commercially available, no new development is required. Air separation units are already being constructed and operated as large plant units today, so supplying the oxygen volumes required for the operation of power plants is no longer a problem (Julich GmbH, 2015).

## **2.2. Separation techniques**

### **2.2.1. Absorption process**

This method involves the interaction of flue gas with a solvent, leading to the dissolution of CO<sub>2</sub> within the solvent. Streams of gases like H<sub>2</sub>S, NO<sub>x</sub>, and CO<sub>2</sub> have been treated using this process. Below are the various absorption processes used to capture CO<sub>2</sub>.

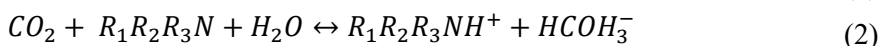
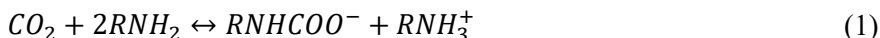
#### **2.2.1.1. Physical absorption**

In physical absorption process a physical absorbent allows a gas to pass through a solid or liquid under some conditions to absorb CO<sub>2</sub> and desorb it under conditions. Using flash tanks or mild thermal regeneration, CO<sub>2</sub> can be separated out at low pressure without any chemical reaction. In addition to being efficient at both high and low temperatures, the process is highly suited for pre-combustion due to the increased solubility of CO<sub>2</sub> above 3.5 bar. Despite the advantages of the low vapour pressure, low toxicity, and less corrosive solvent, some detriments are associated with physical absorption (Arora et al., 2020). In spite of the fact that physical absorption is a mature technology, there is still room for improvement in aspects such as energy consumption and operating costs through the discovery of new solvents (Dubey & Arora, 2022).

#### **2.2.1.2. Chemical absorption**

Using the chemical absorption technique, absorbents of higher selectivity for CO<sub>2</sub> than nitrogen are used to treat gas streams that contain H<sub>2</sub>S, NO<sub>x</sub>, and CO<sub>2</sub>. For capturing CO<sub>2</sub> from flue gases under low pressure, this method has been extensively used (Madejski et al., 2022). The following are some absorbents that can be used to capture CO<sub>2</sub>.

**Amines:** The most applicable CO<sub>2</sub> capture technology is aqueous alkanolamine. There is a higher absorption capacity for chemically-based absorbents at lower partial pressures than for physical absorbents due to the reactive nitrogen atoms in them. The following reactions take place during absorption and separation (Ochedi et al., 2020):



The structure of an amine is classified as primary, secondary, or tertiary based on the number of hydrogen atoms in the compound. The terms in the equations (1) and (2), RNH<sub>2</sub> and R<sub>1</sub>R<sub>2</sub>R<sub>3</sub>N, are the primary and secondary, and tertiary amine, respectively. The absorption capacity of primary and secondary amines is 0.5 mol CO<sub>2</sub>/mole amine, while the absorption capacity of tertiary amine is 1 mol CO<sub>2</sub>/mole amine. Standard practices include aqueous alkanolamine solutions. However, there are a number of disadvantages, including low CO<sub>2</sub> loading, corrosiveness, high energy consumption, solvent degradation, and evaporation loss (Lepaumier et al., 2010). Carbon sorbents based on polyethanolamine are the most common, and a Danish coal-fired pilot plant used MEA at a concentration of 20-30 wt% MEA. It was, however, reported that the high regeneration energy requirement of 3.7 GJ/tCO<sub>2</sub> was achieved by utilizing 1.4 kilograms of MEA/tCO<sub>2</sub> (Knudsen et al., 2009).

In primary and secondary amines, faster reactions, low solubility, and high reactive heat are observed, while in tertiary amines, absorption of CO<sub>2</sub> is higher. To improve performance, aqueous amine blends have been used. MEA and aminomethyl propanol (AMP) or other amines have been combined and shown to improve performance (Dey & Aroonwilas, 2009).

**Amino acids:** Salts of amino acids have higher CO<sub>2</sub> reactivity, are less volatile, less degradable, nontoxic, and maintain their surface tension in oxidizing blends (Zarei et al., 2020; Talkhan et al., 2020). The neutralized form of amino acids will have favourable characteristics regarding vapour pressure due to their ionic properties and resistance to oxidative degradation. The challenges lie in the limited solubility and precipitation (Lee et al., 2015). Nevertheless, a mixture of glycine, alanine, and sarcosine was found to have a positive effect on solvent stability during CO<sub>2</sub> separation (Li et al., 2021).

**Chilled ammonia process (CAP):** There are many multiphase absorption processes available, but aqueous ammonia (NH<sub>3</sub>) is the most advanced. The following reaction occurs in aqueous solutions between NH<sub>3</sub> and CO<sub>2</sub> at low temperatures (0-20 °C) (Darde et al., 2010).



Low degradation rates, low corrosion, ease of regeneration, high absorption capacity, and the ability to capture multiple pollutants are some of the advantages of using  $NH_3$ -based solvents. The formation of bicarbonate in a low pH solution is favoured at a low heat value of 27–92°C. A significantly lower amount of regeneration energy is required during this process (2–3 GJ/tCO<sub>2</sub>) compared to an amine process. Because  $NH_3$  escapes quickly from the environment, a decrease in its concentration occurs, requiring the makeup of ammonia during recirculation (Wang et al., 2011).

**Carbonation:** It has been found that aqueous potassium carbonate, calcium carbonate, and sodium carbonate are most suited for CO<sub>2</sub> absorption due to their low capturing cost, highly soluble, low toxicity, highly efficient, and low degradable compared to amines (Feng et al., 2019; Hornbostel et al., 2019). Carbonate solutions with a higher capture capacity, like potassium carbonate (K<sub>2</sub>CO<sub>3</sub>), are more commonly used. The process involves the following reaction:



Researchers have investigated the use of activators to increase the absorption rate and efficiency of absorbers, reducing both the size of the absorbers and the capital cost of the units (Hu et al., 2016).

It has been found that carbonate solutions have great potential, but more work is needed to understand how the process works, including the kinetics and thermodynamics of the process and the effectiveness of activators such as carbonic anhydrates and corrosion inhibitors (Feng et al., 2019; Hu et al., 2016).

### 2.2.2. Adsorption process

Adsorption involves binding CO<sub>2</sub> to surfaces without forming chemical bonds under certain conditions. Since the process requires less regeneration energy, existing plants can be retrofitted, and pre-and post-combustion routes can be used. In addition to high CO<sub>2</sub> sorption capacity, low energy requirements, and low corrosion probability, amine-based solid adsorbents have remarkably captured CO<sub>2</sub>. Despite its advantages, alkanolamine has certain disadvantages, such as corrosion, a high regeneration energy requirement, and a large absorber size (Wang & Song, 2019; Bui et al., 2018). The schematic of CO<sub>2</sub> adsorption is shown in Figure 2 below.

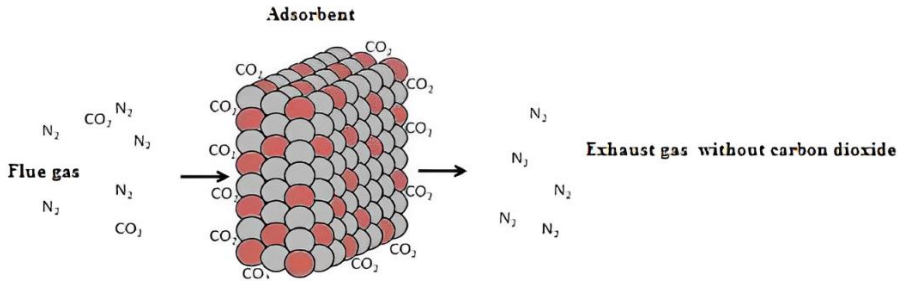


Figure 2- Schematic of physical adsorption (Dubey & Arora, 2022)

### 2.2.3. Membrane separation

The process involves physical or chemical interaction between gases and membranes and is thought to be more energy efficient and eco-friendly than other processes (Japip et al., 2014). Since the partial pressure of  $CO_2$  and the concentration of  $CO_2$  after post-combustion are very low, this process is challenging.

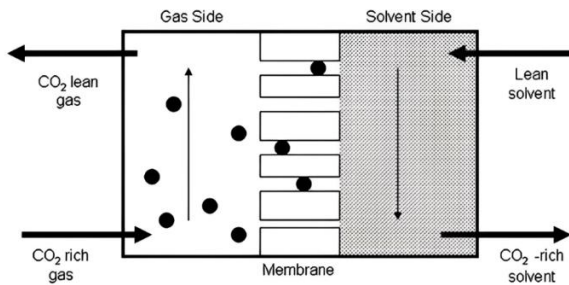


Figure 3- Schematic of membrane separation

A number of significant advantages have been identified when using amine-functionalized nanoporous materials to help separate  $CO_2$ , such as increased adsorption capacity, selectivity, increased kinetics, and decreased cost, compared to traditional membranes (Dubey & Arora, 2022).

### 3. Case study

As a case study a biomass-fired combined heat and power plant (CHP) situated in Sweden was selected. This plant illustrates a meritorious model with notable attributes and advantages. One of its key strengths is its district heating capacity, ranging from 75 to 95 MW. This capacity ensures a reliable and sustainable heat supply, including domestic hot water, space heating, and industrial processes. By providing heat to multiple sectors, the plant contributes to energy efficiency and meets the diverse heating needs of the region.

In addition to its district heating capabilities, the biomass power plant significantly boosts the regional power situation through its electricity production. By efficiently harnessing fuel resources, the plant generates a substantial amount of electricity, meeting the increasing energy demands of the region. This not only reduces reliance on conventional power sources but also contributes to a more sustainable and environmentally friendly power generation mix, promoting a greener energy landscape.

Another notable feature of the plant is its steam boiler, which enables the efficient delivery of the steam to another industrial facility. This versatility highlights the plant's ability to support local industries by providing a reliable source of process steam. By meeting the steam requirements of other industrial processes, the plant enhances its overall value proposition and strengthens its contribution to the regional economy, fostering industrial growth and productivity.

Furthermore, the biomass plant exhibits impressive fuel flexibility by utilizing forest residues and waste wood as feedstock. This strategic approach ensures the efficient utilization of available resources while promoting sustainable forest management practices. Using forest residues and wood waste as fuel, the plant reduces dependence on non-renewable resources and contributes to the circular economy by transforming waste into a valuable energy source. This sustainable fuel-sourcing approach aligns with environmental goals and supports the transition towards a more sustainable energy sector.

To provide further insight, Table 1 presents the boiler specifications used in the plant, highlighting its technical characteristics and capabilities.

*Table 1- Boiler specifications*

<b>Boiler capacity (MWth)</b>	110
<b>Steam pressure (bar(g))</b>	80
<b>Steam temperature (°C)</b>	500
<b>Nominal fuel</b>	30% waste wood 30% forest residues 25% bark 15% saw dust
<b>Maximum waste wood fraction</b>	50%
<b>Minimum boiler load</b>	30%

### **3.1. The process**

The study examines a biomass plant, where fluidized bed technology is employed in the boiler system. This approach takes advantage of the physics phenomenon observed when solid particles within a containment vessel exhibit fluid-like properties under specific operating conditions. A dynamic and fluidized bed environment is created by introducing pressurized fluid into the particle-filled chamber. This technology offers benefits, including improved combustion efficiency, enhanced fuel flexibility, and reduced emissions. By leveraging this cutting-edge fluidized bed technology, the biomass plant optimizes operational efficiency, ensures efficient fuel utilization, and improves overall sustainability and environmental performance.

In biomass electricity generation, the dominant method is direct combustion. In this process, steam generated in the boiler flows over turbine blades, causing them to rotate. The rotational motion of the turbine drives a generator, converting mechanical energy into electrical energy. The produced electricity can be consumed internally by the plant or supplied to the power grid for broader distribution and utilization. This integration of direct combustion, steam-driven turbines, and electricity generation represents a practical and effective approach for harnessing renewable energy from biomass sources.

The overall operation process of the plant is seen in Figure 4.



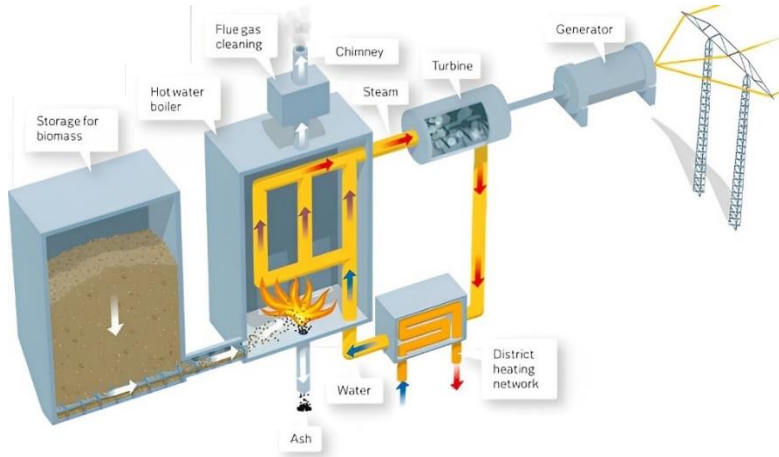


Figure 4- Operation process of the unit

### 3.2. Flue gas cleaning

In the context of solvent-based scrubbing technologies like the MEA carbon capture process, it is essential to attain a reduction in multiple pollutants in the incoming flue gas. Before the carbon capture process, thoroughly cleaning the flue gas is crucial to ensure the optimal efficiency and effectiveness of the carbon capture system. This pre-treatment step aims to remove impurities, contaminants, and pollutants from the flue gas stream, creating an environment conducive to efficient carbon capture. By achieving a high level of cleanliness in the pre-treatment, the overall performance and reliability of the carbon capture process are enhanced, leading to the successful capture and removal of carbon dioxide from the flue gas stream. Figure 5 illustrates the process diagram for the flue gas cleaning.

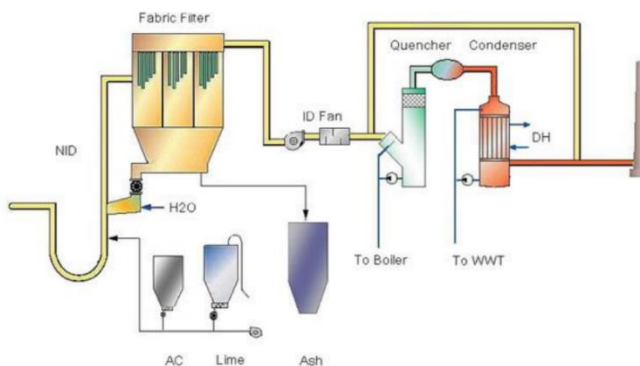


Figure 5 - Flue gas cleaning process

During the initial stage of the flue gas cleaning process, lime and active carbon are employed as effective agents to eliminate acidic gases, specifically sulfur dioxide (SO<sub>2</sub>) and hydrogen chloride (HCl), from the flue gas. These harmful gases are chemically reacted with the lime particles by utilizing lime, resulting in their removal and subsequent reduction of their environmental impact. In waste incineration plants, powdered activated carbon is also used to extract dioxins/furans and heavy metals effectively. This specialized carbon powder can absorb and capture these pollutants, mitigating their release into the atmosphere. Furthermore, the fly ash in the flue gas is efficiently collected using filter fabric collectors, commonly known as baghouses. This technology facilitates the separation of fly ash particles from the flue gas, preventing its dispersion into the environment. The collected fly ash is then transported to the silo for further management and utilization.

For the last step of flue gas cleaning, it is investigated an installation of a Selective Catalytic Reduction (SCR) system is needed. SCR is a technology used to reduce the level of nitrogen oxides (NO<sub>x</sub>) in the presence of a catalyst. Vaporized reagent (Ammonia) is injected into the flue gas and reacts with NO<sub>x</sub> on the SCR catalyst resulting in nitrogen (N<sub>2</sub>) and water (H<sub>2</sub>O).

Efficient elimination of NO<sub>x</sub>, SO<sub>x</sub>, and hazardous air pollutants from flue gas is crucial for the carbon capturing process and the environment. Amine-based solvents used in carbon capture tech can be negatively impacted by pollutants, creating harmful byproducts and reducing CO<sub>2</sub> absorption. Additionally, their emissions cause air pollution and acid rain formation, which endanger human health and ecosystems. Consequently, the removal of the substances mentioned above is imperative to ensure the utmost operation of carbon capture systems and mitigate environmental impacts such as respiratory issues, ecosystem imbalances, acid rain destruction, and the emission of toxic substances into the atmosphere.

## 4. Methods

This thesis focuses on conducting an energy assessment of the carbon capture (CC) unit implemented in a biomass plant. To achieve this, a model was developed, and its structure is shown in Figure 6. The research initiated a comprehensive literature study on various carbon capture technologies, enabling the collection of detailed information on each method and selecting the most suitable one considering industry-specific conditions.

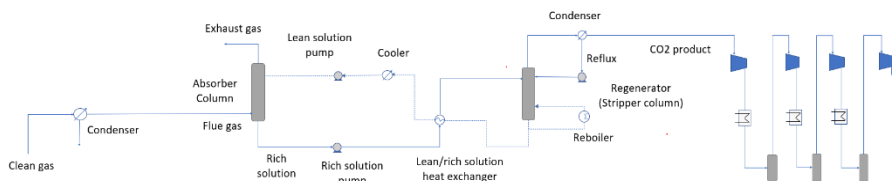


Figure 6 – Overall process diagram

For the purpose of calculations, a simplified carbon capture process was designed, necessitating data collection for mass and energy balance analyses. The initial model was based on the 30% MEA process, with assumed loading capacities of 0.32 for the rich solvent and 0.26 for the lean solvent. The term "loadings" refers to the amount of CO<sub>2</sub> in moles per mole of solvent. In this context, lean solution signifies a lower concentration of CO<sub>2</sub>, whereas rich solution indicates a higher concentration of CO<sub>2</sub>. The overall energy demand of the system, encompassing cooling, heating, and electricity requirements, was determined by considering factors such as pre-cooling (if needed), capture rate, flue gas composition, and the number of compression stages.

In the subsequent phase, different lean and rich solvent loadings were chosen to recalibrate the energy demand and identify the optimal composition based on the reboiler's energy requirements. In the final step, the obtained results were compared, and recommendations were formulated to enhance the unit's overall efficiency.

Throughout this thesis, efforts were made to thoroughly analyze the energy performance of the CC unit in the biomass plant. The findings and recommendations presented aim to contribute to the ongoing efforts to improve the efficiency and effectiveness of carbon capture processes in similar industrial settings.

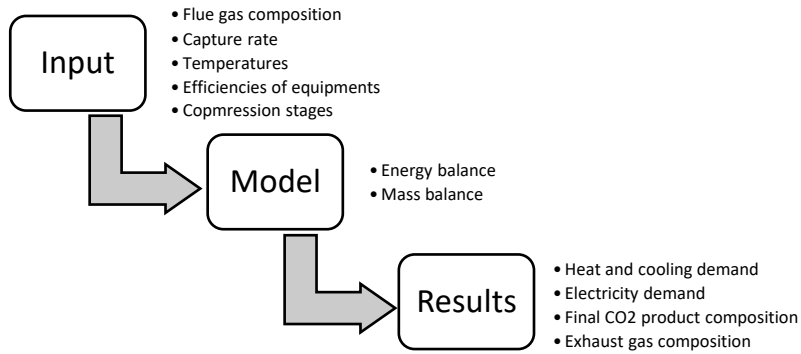


Figure 7- Project workflow

## 4.1. Process description

Different databases such as Google Scholar, Science Direct, Scopus, and LUB Search were used to find the suitable and the most common process used in the industry. The information provided from the literature review about technologies was used to design a simple carbon capture MEA-based process that can be used as a model for the calculations. In this study, an absorption post-combustion CO<sub>2</sub> capture unit, with pre-cooling system, followed by a CO<sub>2</sub> compression unit are modeled. This section provides a detailed description of the chosen strategy and its specifications.

### 4.1.1. Conventional MEA process

The carbon dioxide capture process involves using a chemical technology wherein the gas-liquid contact takes place in an absorber column in a counter-current manner. This allows for the absorption of carbon dioxide by the solvent. To compensate for solvent degradation, additional amine is introduced into the absorber as makeup solvent. The regeneration of the solvent takes place in a parallel unit known as the stripper, which enables the recycling of the solvent back to the absorber (Mazari et al., 2015). However, no consideration is given to the makeup amine in the current study.

The conventional method for carbon capture using MEA (monoethanolamine) is depicted in Figure 8. The key components of this process include the absorber, heat exchanger, and stripper. The flue gas, containing CO<sub>2</sub>, enters the bottom of the absorber, while the lean MEA solution is introduced from the top. Through an exothermic reaction with the MEA solvent, CO<sub>2</sub> is selectively absorbed. The CO<sub>2</sub> rich solution exits from the bottom of the absorber and undergoes pre-heating in the internal heat

exchanger. Upon entering the top of the stripper, the rich solution releases CO<sub>2</sub> under higher temperature and pressure conditions. The lean solvent is withdrawn from the bottom of the stripper, while the captured CO<sub>2</sub> is obtained at the top and proceeds to the compression stage. After passing through the heat exchanger for cooling, the lean solvent re-enters the absorber until it reaches the desired temperature (Jung et al., 2013).

A fundamental principle underlying this design is that the absorption process takes place within a temperature range of 40-50°C, while desorption occurs at approximately 120°C (Rochelle, 2016; Karlsson, 2021). The absorber operates at atmospheric pressure, whereas the desorption chamber operates at a pressure close to 2 bars. Aside from the electricity needed for pump operations involved in solution transportation, the reboiler stands out as the component with the highest energy consumption in the conventional MEA scrubbing process (Jung et al., 2013).

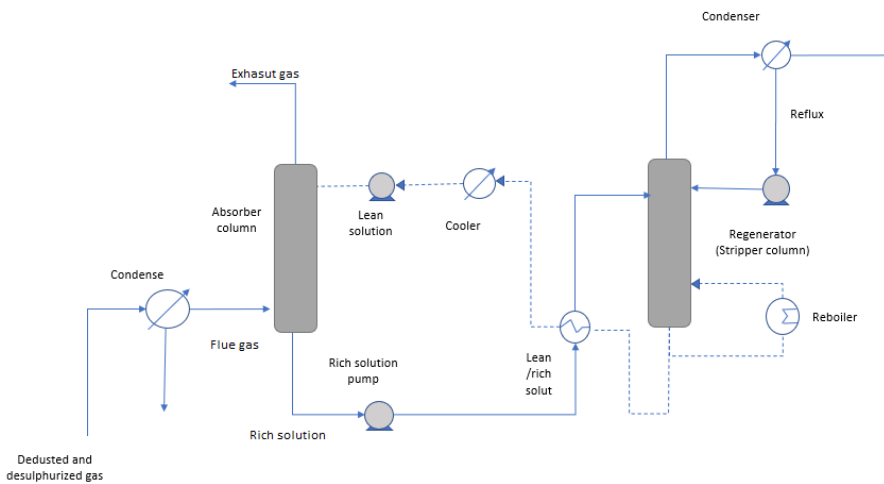


Figure 8- Flow diagram of simplified design for MEA method

## 4.2.Data collection

Sweco provided valuable data on the flue gases emitted by the biomass plant. The information supplied by Sweco is calculated but closely represents real-world conditions and is considered reliable. The dataset covers various parameters, including the inlet flue gas temperature, volumetric flow, and flue gas composition. Upon analyzing the flue gas's properties, the SO<sub>x</sub> and NO<sub>x</sub> levels were deemed insignificant for the process design. The composition of

the flue gas encompasses a wide range of gases, such as carbon dioxide (CO<sub>2</sub>), nitrogen (N<sub>2</sub>), oxygen (O<sub>2</sub>), and water vapor (H<sub>2</sub>O). Table 2 provides a summary of the relevant properties of the flue gases, which are utilized in the calculations of the unit's energy demand. This information plays a vital role in the design and optimization of the carbon capture process, ensuring the effectiveness of the capture technology in reducing greenhouse gas emissions.

*Table 2- Properties of the flue gas entering the CC unit.*

<b>Property</b>	<b>Unit</b>	<b>Value</b>
<b>Flue gas temperature</b>	°C	34
<b>Flue gas flow rate (wet)</b>	Nm <sup>3</sup> /h	166256
<b>H<sub>2</sub>O</b>	Vol%	5.22
<b>O<sub>2</sub></b>	Vol% (wet)	4.45
<b>CO<sub>2</sub></b>	Vol% (wet)	14.78
<b>N<sub>2</sub></b>	Vol% (wet)	75.55
<b>Capture rate</b>	Vol%	90

The specific heat capacities for different compositions of MEA solution, as well as the pressure and temperature of the reboiler and the partial pressure of CO<sub>2</sub> at the reboiler temperature, were collected from relevant literature sources. Furthermore, enthalpies of water and gas compositions were obtained using the FluidProp Excel function. The heat of the reaction between CO<sub>2</sub> and MEA was determined by referencing several papers. Sweco and the suppliers also provided additional information regarding the plant and other components such as the pumps and the compressors.

The pressure and temperature of the reboiler play crucial roles in determining the efficiency of the carbon capture process. In this study, the reboiler temperature was assumed to remain constant at 120°C for all cases.

Efficiencies and the physical conditions of the components were considered as inputs for the modeling approach.

### 4.3. Calculations

The subsequent section outlines the methodologies employed in the mass and energy balance calculations, as well as the outcomes obtained in terms of steam requirements for the reboiler. The same calculation approach can be applied to all solvent loadings and flue gas compositions. It is important to highlight that while the mass and energy balance method can be applied to any flue gas data, the process described in this study specifically pertains to the use of MEA as the solvent. During the energy assessment, three primary factors were taken into account: cooling, heating, and electrical demands.

#### 4.3.1. Flue gas pre-cooling

By assuming that the flue gas details provided by Sweco behave like ideal gases, equations (5) to (8) were used to estimate total CO<sub>2</sub> mass flow and flue gas content. These calculations relied on parameters such as the volumetric flow rate, CO<sub>2</sub> content, and water content of the flue gas.

$$\dot{m}_{CO_2} = \frac{\dot{V}_{fl} \cdot x_{CO_2} \cdot M_{CO_2}}{V_{tot}} \quad (5)$$

$$\dot{m}_{H_2O} = \frac{\dot{V}_{fl} \cdot x_{H_2O} \cdot M_{H_2O}}{V_{tot}} \quad (6)$$

$$\dot{m}_{O_2} = \frac{\dot{V}_{fl} \cdot x_{O_2} \cdot M_{O_2}}{V_{tot}} \quad (7)$$

$$\dot{m}_{N_2} = \frac{\dot{V}_{fl} \cdot x_{N_2} \cdot M_{N_2}}{V_{tot}} \quad (8)$$

Where  $\dot{m}$  shows the mass flow rate,  $\dot{V}_{fl}$  is the volumetric flow rate of the flue gas,  $x$  represents the molar fraction in the gaseous phase (which is equal to the volumetric fraction)  $M$  is the molar weight, and  $V_{tot}$  is the molar volume of the ideal gas.

As the temperature of the flue gas decreases in the pre-cooling process, some water will be condensed and is expelled the system. To find the amount of condensed water and the new flue gas composition, partial pressure of water was used based on Raoult's law. In this method, the partial pressure of water in the cooled gas equals the saturation pressure of water, as seen in equation (9).

The saturated pressure of the components ( $P_A^{sat}$ ) was obtained using FluidProp at the absorber temperature.  $P_A$  is the partial pressure of the component and  $y_A$  shows its molar fraction.

$$P_A = P_A^{sat} y_A \quad (9)$$

These calculations facilitated the determination of the new water content in the flue gas. Additionally, equation (10) was employed to calculate the new volumetric flow rate of the flue gas.

$$\dot{V}_{tot,new} = \frac{\sum x_i (dry)}{(1 - x_{H_2O,new})} \cdot \dot{V}_{tot,initial} \quad (10)$$

Where  $x_i$  is the dry molar fractions from the initial flue gas composition. The new flue gas composition was determined using equations (11)-(13) for  $O_2$ ,  $N_2$ , and  $CO_2$ , respectively, based on the total flow.

$$x_{O_2,fl,in} = \frac{x_{O_2,initial} \cdot \dot{V}_{tot,initial}}{\dot{V}_{tot,new}} \quad (11)$$

$$x_{N_2,fl,in} = \frac{x_{N_2,initial} \cdot \dot{V}_{tot,initial}}{\dot{V}_{tot,new}} \quad (12)$$

$$x_{CO_2,fl,in} = \frac{x_{CO_2,initial} \cdot \dot{V}_{tot,initial}}{\dot{V}_{tot,new}} \quad (13)$$

With the help of the ‘‘Gas Mix’’ function of FluidProp, the sensible enthalpy of cooling of the flue gas is calculated. First, the specific heat capacity of the flue gas was calculated, and then, equation (14) was employed to calculate the cooling effect. It should be noted that in this process, only elements other than water were assumed to be cooled, while the cooling of water content was calculated separately, considering the enthalpies. This approach was necessary as the software used did not account for the condensation of water. The temperature difference considered is between the initial flue gas temperature and the absorber temperature.

$$Q = \dot{m} \cdot C_p \cdot \Delta T \quad (14)$$

As for the water content, the difference in internal energies before and after cooling was utilized to determine the cooling effect using equation (15). In this equation,  $h_1$  and  $h_2$  represent the enthalpies before and after the cooling process, respectively.



$$Q_{H_2O} = (h_1 \cdot \dot{m}_1 - h_2 \cdot \dot{m}_2) \quad (15)$$

The total cooling effect for the condenser is the sum of the cooling effect from the flue gas itself as well as the cooling effect from the water present in it. Furthermore, the mass flow rate of condensed water is determined by calculating the difference between the water mass flow rate in the initial flue gas and the mass flow rate after cooling.

#### 4.3.2. Absorber

The amount of CO<sub>2</sub> captured is defined by the user, although in the initial model, it was assumed to be 90% of the CO<sub>2</sub> content in the flue gas. This allows for the calculation of the mass flow rates of the lean and rich solutions using equations (16) and (17).

$$\dot{m}_{lean} = \frac{\dot{x}_{CO_2} \cdot M_{MEA}}{(\alpha_{rich} - \alpha_{lean}) \cdot 30\%} \quad (16)$$

$$\dot{m}_{rich} = \dot{m}_{CO_2,captured} + \dot{m}_{lean} + (\dot{m}_{H_2O,fl,in} - \dot{m}_{H_2O,ex}) \quad (17)$$

$\dot{m}_{H_2O,cooledfl}$  represents the mass flow of the water content in the flue gas after the carbon capture process. As 90% of the inlet CO<sub>2</sub> is assumed to be captured, the remaining 10% leaves the system. The terms  $\alpha_{lean}$  and  $\alpha_{rich}$  refer to the CO<sub>2</sub> loading per mole of MEA in the lean and rich solutions, respectively.

$$\dot{m}_{CO_2,ex} = 10\% \cdot \dot{m}_{CO_2,fl,in} \quad (18)$$

The composition of the other exhaust gas components remains the same as the inlet flue gas, with the exception of the water content. The molar fraction of water can be determined by knowing its partial pressure at the desired temperature. In this case, it is assumed that the exhaust gas temperature is 48°C.

For calculating the energy requirements, it is necessary to know the mass flows of MEA, water in the rich and lean solution, and the amount of CO<sub>2</sub> in each stream. Equations (19) to (23) are utilized for this purpose. It is important to note that the initial case is designed to use a 30% MEA solution, and it is assumed that some CO<sub>2</sub> is recirculated throughout the entire process.

$$\dot{m}_{MEA} = \left( \dot{m}_{rich} - (\dot{m}_{CO_2,fl,in} - \dot{m}_{CO_2,ex}) \right) \cdot 30\% \quad (19)$$

$$\dot{m}_{CO_2,lean} = \alpha_{lean} \cdot \left( \frac{\dot{m}_{MEA}}{M_{MEA}} \right) \cdot M_{CO_2} \quad (20)$$

$$\dot{m}_{CO_2,rich} = \alpha_{rich} \cdot \left( \frac{\dot{m}_{MEA}}{M_{MEA}} \right) \cdot M_{CO_2} \quad (21)$$

$$\dot{m}_{H_2O,rich} = \dot{m}_{rich} - \dot{m}_{MEA} - \dot{m}_{CO_2,rich} \quad (22)$$

$$\dot{m}_{H_2O,lean} = \dot{m}_{lean} - \dot{m}_{MEA} - \dot{m}_{CO_2,lean} \quad (23)$$

The temperature of the rich solution exiting the absorber can be determined using the heat of the rich solution, mass flow of rich solution, and specific heat of it. The absorption of CO<sub>2</sub> in an MEA-based system is exothermic, releasing heat. The heat of the CO<sub>2</sub> reaction is assumed to be 84.60 kJ/mol CO<sub>2</sub> at the absorber temperature, based on references (Zhang et al., 2016; Oexmann & Kather, 2009).

$$Q_{absorption} = 84.6 \cdot \dot{x}_{CO_2} \quad (24)$$

Where  $\dot{x}_{CO_2}$  is the molar flow of the captured CO<sub>2</sub>.

Based on the amount of MEA in the solution and the loading factor for the lean and rich streams, the heat capacities of CO<sub>2</sub>-loaded MEA are obtained using data from various literature sources. For a solution with 30% MEA, the C<sub>p</sub> values for the lean and rich solutions are assumed to be 3.51 (J/g·K) and 3.41 (J/g·K), respectively (Weiland et al., 1997). The reference temperature is assumed to be the absorber temperature, which in this case is 40°C.

$$Q_{lean} = \dot{m}_{lean} \cdot C_{p,lean} \cdot T_{absorber} \quad (25)$$

$$Q_{H_2O,in} = \dot{m}_{H_2O,fl,in} \cdot h_{H_2O,fl,in} \quad (26)$$

$$Q_{H_2O,ex} = \dot{m}_{H_2O,ex} \cdot h_{H_2O,ex} \quad (27)$$

$$Q_{fl,in} = \dot{m}_{fl,in} \cdot C_{fl,in} \cdot T_{absorber} \quad (28)$$

$$Q_{fl,ex} = \dot{m}_{fl,ex} \cdot C_{fl,ex} \cdot T_{ex} \quad (29)$$

$$Q_{tot,in} = Q_{H_2O,in} + Q_{fl,in} \quad (30)$$

$$Q_{tot,ex} = Q_{H_2O,ex} + Q_{fl,ex} \quad (31)$$

In the equations presented above, the heat for the water content is calculated separately, indicating that the heat of the flue gas only includes N<sub>2</sub>, O<sub>2</sub>, and CO<sub>2</sub>.

The total energy assumed as the inlet energy is the sum of the heat released from the CO<sub>2</sub> absorption process, the heat of the lean solution, and the heat of the flue gas entering the carbon capture unit.

$$Q_{absorber,in} = Q_{absorption} + Q_{lean} + Q_{tot,in} \quad (32)$$

The only component exiting the system is the exhaust gas. To determine the energy required to heat up the rich solution, equation (33) is utilized. Once the heat capacity of the rich solution is known, the outlet temperature of the rich solution can be determined.

$$Q_{rich} = Q_{absorber,in} - Q_{tot,ex} \quad (33)$$

$$T_{rich} = \frac{Q_{rich}}{C_{p,rich} \cdot \dot{m}_{rich}} \quad (34)$$

#### 4.3.3. Regenerator (Stripper)

In this section, the pressure and temperature of the reboiler are assumed to be constant. The reboiler operates at a pressure of 1.7 bars and a temperature of 120°C. The pressure of the stripper depends on the partial pressure of water and CO<sub>2</sub> at the reboiler temperature. For the calculations, it is assumed that  $\alpha_{lean}=0.26$  at the mentioned temperature, and the partial pressure of CO<sub>2</sub> is close to 0.3 bars (Karlsson, 2021).

When the lean solution passes through the internal heat exchanger, its temperature decreases. However, a cooler before the absorber is required to cool the solution to the absorber temperature. The amount of heat exchanged between the two solutions is based on the mass flows, heat capacities, and assuming an 8°C temperature difference between the top and bottom of the absorber. The outputs from the heat exchanger are calculated using equations (35) to (38).

$$Q_{rich,HX} = Q_{lean,HX} = \dot{m}_{rich} \cdot C_{p,rich} \cdot (T_{st,top} - T_{rich}) \quad (35)$$

$$\Delta T_{lean} = \frac{Q_{lean,HX}}{(C_{p,lean} \cdot \dot{m}_{lean})} \quad (36)$$

$$T_{lean,HX} = T_{st,bot} - \Delta T_{lean} \quad (37)$$

$$Q_{lean,cooler} = \dot{m}_{lean} \cdot C_{p,lean} \cdot (T_{lean,HX} - T_{absorber}) \quad (38)$$

$T_{st,top}$  and  $T_{st,bot}$  are the temperatures at the top and the bottom of the stripper, respectively, while the  $T_{lean,HX}$  illustrates the outlet temperature of the lean solution from the heat exchanger.  $Q_{lean,cooler}$  indicates the heat from the lean solution cooler before entering the absorber.

As the reboiler is the most energy-intensive part, it is important to determine the heat requirements. The mass flow rate of the steam that provides the energy for the reboiler is calculated based on the steam's pressure and temperature. The composition of the CO<sub>2</sub> product leaving the stripper is found using the partial pressure of water vapor, while the rest of the mixture is CO<sub>2</sub>. The number of moles of captured CO<sub>2</sub> remains constant, allowing for the calculation of the total number of moles in the mixture and the number of moles of H<sub>2</sub>O.

$$n_{tot} = \frac{n_{CO2}}{x_{CO2}} \quad (39)$$

$$n_{H2O} = \frac{x_{H2O}}{n_{tot}} \quad (40)$$

The mass flow rate of each component and the total mass flow rate exiting the stripper are then known.

The energy entering the stripper is the heat of the rich solution and the reflux water that returns after being condensed. The heat of the lean solution, the mixture of CO<sub>2</sub> and H<sub>2</sub>O leaving the regenerator, and the heat of desorption exit the system. The heat required for the reboiler is determined based on the inlet and outlet energies using equation (41). The heat of desorption is equal to the absorption with the opposite sign for the calculations. The heat of the stream at the outlet of the stripper can be found by knowing the mass flow rate and the heat capacity, which is calculated using the FluidProp Gas Mix function. The temperature of the condenser is assumed to be 60°C. Thus, the amount of reflux water is the difference between the water content in the stripper outlet stream and the CO<sub>2</sub> product after the carbon capture unit.

Additionally, by knowing the heat of evaporation of water at the reboiler temperature, it is possible to calculate the mass flow rate of steam used to provide the energy for the reboiler.

$$Q_{reboiler} = Q_{lean} + Q_{st,top} + Q_{desorption} - Q_{rich} - Q_{H2O,reflux} \quad (41)$$

$$\dot{m}_{steam} = \frac{Q_{reboiler}}{h_{fg}} \quad (42)$$

Where  $h_{fg}$  is the heat of evaporation at the temperature of the steam, which in this case is assumed to be 150°C.

It is also possible to calculate the cooling effect of the condenser after the regeneration process. Since some water is condensed and FluidProp cannot account for it, the energies of CO<sub>2</sub> and H<sub>2</sub>O before and after the condenser are calculated separately. The cooling effect for the condenser is determined using the provided equations. Indexes 1 and 2 denote the situation before and after the condenser.

$$Q_{condenser} = Q_{H2O,1} - Q_{H2O,2} - Q_{H2O,reflux} + Q_{CO2,1} - Q_{CO2,2} \quad (43)$$

Finally, to balance the system, some makeup water needs to be added to the stripper.

$$\dot{m}_{makeup\ water} = \dot{m}_{H2O,lean} + \dot{m}_{H2O,2} - \dot{m}_{H2O,rich} \quad (44)$$

#### 4.3.4. Electricity requirements

For the electricity calculations, the two prominent cases are the lean and rich pumps in the carbon capture unit. The height of the absorber is assumed to be 30m, while the height of the stripper is 28m. The internal heat exchanger causes a pressure drop of 60 kPa, and the pump efficiency is considered to be 75% for both pumps. The densities are found using FluidProp. The calculations for both pumps follow the same process, utilizing equations (45) to (49). The only difference is that the rich pump needs to deliver the solution to the top of the stripper, so the stripper height is used for the static pressure drop, while the absorber height is used for the lean pump. The densities and volumetric flows vary between the two pumps. Presented below are the calculations for the rich pump as an example:

$$\Delta P_{static} = \rho \cdot g \cdot H_{st} \quad (45)$$

$$P_2 = \Delta P_{static} + \Delta P_{HX} \quad (46)$$

$$\Delta P = P_2 + (P_{st} \cdot P_{atm}) - P_{abs} \quad (47)$$

$$\dot{V} = \frac{\dot{m}_{rich}}{\rho_{rich}} \quad (48)$$

$$E_{rich\ pump} = \frac{\dot{V} \cdot \Delta P}{\eta_{pump}} \quad (49)$$

Where  $\rho$  represents the density of the solution,  $g$  is the acceleration due to gravity, equal to  $9.81 \text{ m/s}^2$ ,  $H$  is the height of the regenerator column,  $P_2$  is the final destination pressure, accounting for the pressure drop of the heat exchanger, and  $\Delta P$  is the pressure difference that the pump must overcome. By knowing the pump efficiency, volumetric flow rate, and pressure difference, the required electricity can be determined.

#### 4.3.5. Compression

The compression of the  $\text{CO}_2$  product involves multiple stages. In each stage, the gas stream is compressed and passed through a heat exchanger to reduce its temperature. The condensed water produced during the process is subsequently removed from the system. This process is repeated until the desired pressure is achieved. In this work, it is assumed that all compressors have the same efficiency, and the pressure ratio, which represents the increase in pressure after compression, remains constant throughout the process. Additionally, all heat exchangers cool the stream to  $60^\circ\text{C}$ , resulting in equal inlet temperatures for all compressors. Four compression stages are employed in this case, with a final pressure of 40 bars(a). Due to the difficulties in finding a straightforward correlation for non-isentropic processes and the ensuing complexity of calculations, the process is assumed to be isentropic. Nonetheless, it is feasible to utilize efficiency measurements for the compression stages.

The pressure ratio can be determined by knowing the initial pressure at the start of the compression stage, the final pressure, and the number of stages. The equation below represents the pressure ratio, where  $P_r$  is the pressure ratio,  $P_{\text{final}}$  is the pressure at the end of the compression process,  $P_{\text{st}}$  is the pressure from the stripper, and  $n$  denotes the number of stages.

$$P_r = \left(\frac{P_{\text{final}}}{P_{\text{st}}}\right)^{\frac{1}{n}} \quad (50)$$

The number of captured  $\text{CO}_2$  moles remains constant throughout the process. The composition and mass flow at each stage can be estimated based on the partial pressure of water vapor. Furthermore, since the temperatures before the compressors are the same, the enthalpy of the gas mixture can be determined using steam tables or FluidProp. As the process is assumed to be isentropic, there is no change in entropy. The work of the compressor can be calculated using equations (51) and (52), where  $h_{2r}$  represents the actual enthalpy,  $h_{2s}$  is the enthalpy for the isentropic process, and  $h_1$  denotes the

initial enthalpy at the entrance. The temperature after compression for each stage can be determined using the enthalpy at that point.

$$h_{2r} = \frac{h_{2s} - h_1}{\eta_{comp}} + h_1 \quad (51)$$

$$W_{comp} = (h_{2r} - h_1) \cdot \dot{m}_{tot} \quad (52)$$

The compositions at each stage are required for the heat exchangers' calculations. Subsequently, the enthalpies at each point can be determined. By using the mass flow of the stream, the amount of heat exchanged can be calculated.  $h_2$  represents the enthalpy after the heat exchanger, and  $h_1$  denotes the enthalpy before the heat exchanger. The mass flow corresponds to the total mass flow before water removal.

$$Q_{1,2,3,\dots} = (h_2 - h_1) \cdot \dot{m}_{tot} \quad (53)$$

By obtaining all the necessary data from the calculations mentioned above, it becomes possible to estimate the energy required to operate an MEA CC unit or evaluate the feasibility of retrofitting it into an existing plant.

## 5. Results

The case study assumes that the flue gas has undergone treatment before entering the CC unit, resulting in a significant decrease in emissions. Furthermore, the temperature of the flue gas is assumed to be lower than the temperature of the absorber. The results for each component of the system will be presented in the following sections.

### 5.1. Pre-cooling

The initial temperature of the flue gas is estimated to be 34°C. Since the temperature of the absorber is 40°C, there is no need for pre-cooling. The flue gas from the flue gas cleaning unit can enter the absorber directly. As a result, there is no condensation of water or cooling requirement for this part.

### 5.2. Absorber

It is assumed that 90%vol of the inlet CO<sub>2</sub> is captured from the flue gas, which has an exhaust gas temperature of 48°C and is at atmospheric pressure. The composition of the exhaust gas is presented in Table 3.

Table 3- Exhaust gas details

Elements	Units					
	kg/s	kmol/s	Nm <sup>3</sup> /s	Nm <sup>3</sup> /h	%wt	%dry
H <sub>2</sub> O	2.6	0.14	3.23	11,643	7.69	-
CO <sub>2</sub>	1.34	0.03	0.68	24,57	1.62	1.76
N <sub>2</sub>	45.17	1.61	36.11	129,999	85.51	92.95
O <sub>2</sub>	2.94	0.09	2.06	7,398	4.88	5.29
Mass flow, wet (kg/s)	52.05	1.88	42.08	151,498	100	-
Mass flow, dry (kg/s)	49.44	1.73	38.85	139,855	-	100



One important parameter obtained from the absorber calculations is the temperature of the rich solution as it exits the absorber. This temperature is highly dependent on the CO<sub>2</sub>-loading factors for both the lean and rich solutions. Based on the previously provided information regarding the loadings and the MEA solution, the mass flows and the temperature of the rich solution are explained in Table 4. It is crucial that the rich solution temperature falls within the range of 40-50°C, and the temperature increase should be within the allowed range.

*Table 4- Information from the absorber*

Captured CO <sub>2</sub> (kg/s)	12.07
$\dot{m}_{Lean}$ (kg/s)	930.5
$\dot{m}_{Rich}$ (kg/s)	942
$\dot{m}_{MEA}$ (kg/s)	279
$\dot{m}_{CO_2,Lean}$ (kg/s)	52.2
$\dot{m}_{CO_2,Rich}$ (kg/s)	64.3
$\dot{m}_{H_2O,Lean}$ (kg/s)	599.3
$\dot{m}_{H_2O,Rich}$ (kg/s)	598.6
T <sub>rich</sub> (°C)	46.47

A summary of the critical data from the absorber calculations is shown in the process flow diagram depicted in Figure 9.

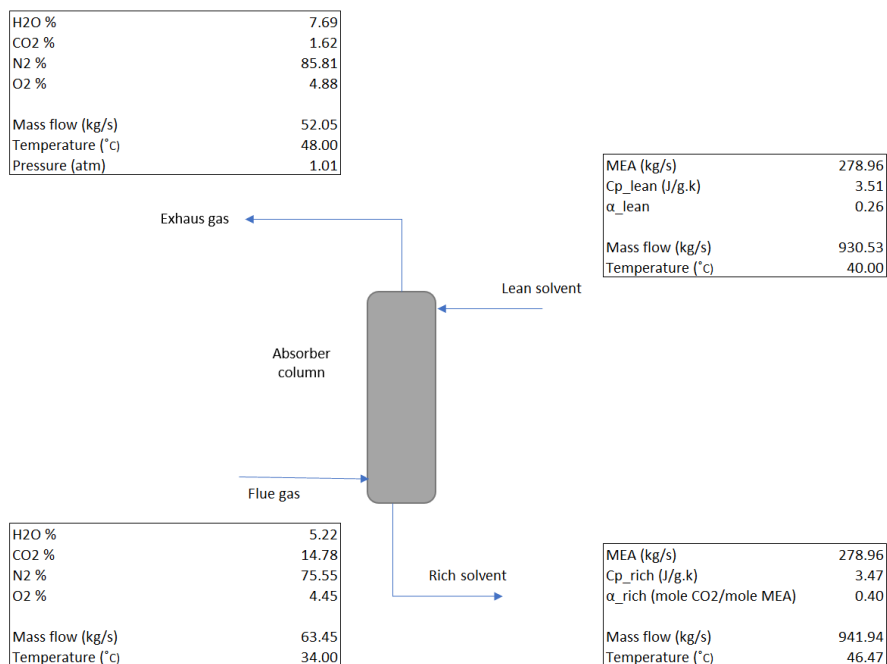


Figure 9- Absorber PFD

### 5.3. Regenerator

The most crucial equipment in this design is the reboiler, which requires the highest amount of heat. Additionally, the cooling requirement for the lean solution is also significant. Therefore, in the stripper column, the energies from the reboiler and the condenser play vital roles. The efficiency of the system and the energy demand are considerably affected by the steam (usually from high-pressure or low-pressure turbine) used to provide heat to the reboiler. The vapor is assumed to have a pressure and temperature of 4 bars and 150°C, respectively. The temperatures at the top and bottom of the stripper (reboiler) are 112°C and 120°C, while the temperature of the condenser is 60°C. The gas mixture composition is defined in Tables 5 and 6, representing the composition before and after the condenser, respectively.

Table 5- CO2 product composition before condenser

Elements	Units	
	%wet	kg/s
CO <sub>2</sub>	37.6	12.07
H <sub>2</sub> O	62.4	8.19
Total	100	20.25

After condensing the stream, some water is condensed and returned to the absorber. It is crucial to consider this water in the calculations.

Table 6- CO2 final product and the reflux water mass flows

Elements	Units (kg/s)
CO <sub>2</sub>	12.07
H <sub>2</sub> O	0.65
H <sub>2</sub> O, <i>reflux</i>	7.54

After condensing the stream, some water is condensed and returned to the absorber. It is crucial to consider this water in the calculations. It should be noted that the temperature before the lean cooler affects the cooling requirement. In this specific case, a temperature of 54.5°C was reached. Considering the system specifications, Table 7 presents the steam and energy demand for the lean cooler, reboiler, and condenser. The negative signs indicate that the equipment releases energy rather than consuming it.

Table 7- Heat and cool requirements for CC unit

Equipment	Energy required (MW)
Lean cooler	-47.35
Reboiler	50.57
Condenser	-19.1
Steam requirement (kg/s) @ 150°C, 4bar(a)	23.9

A summary of the critical data from the regenerator and other discussed parts are presented in the process flow diagram depicted in Figure 10.

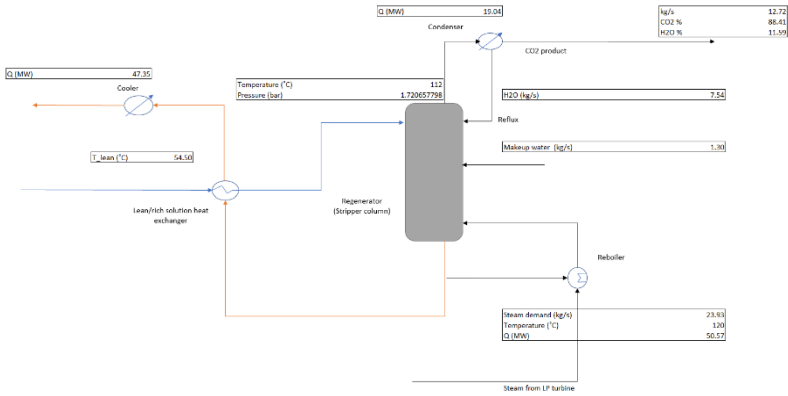


Figure 10- Stripper PFD

## 5.4. Electricity demand (CC)

The lean and rich pumps are the only equipment that consume electricity. Table 8 shows the demand for these two pumps.

Table 8- pumps requirements

Equipment	Electricity demand (kW)
Lean pump	the 348.9
Rich pump	520.9

## 5.5. Compression

In the compression process, the compressors require electricity to operate, while the heat exchangers play a crucial role in stream cooling. The energy from the heat exchangers is used for district heating purposes. This strategic cooling approach enables the collection of condensed water in tanks. The compression is carried out in four stages, with three heat exchangers interspersed between the stages. The stream from the last compressor is directed towards various components, including liquefaction. The pressure ratio, calculated regarding the final pressure (12 bars) and the initial pressure (1.7 bars), is 1.63. A process diagram illustrating the compressors can be found in Figure 11.

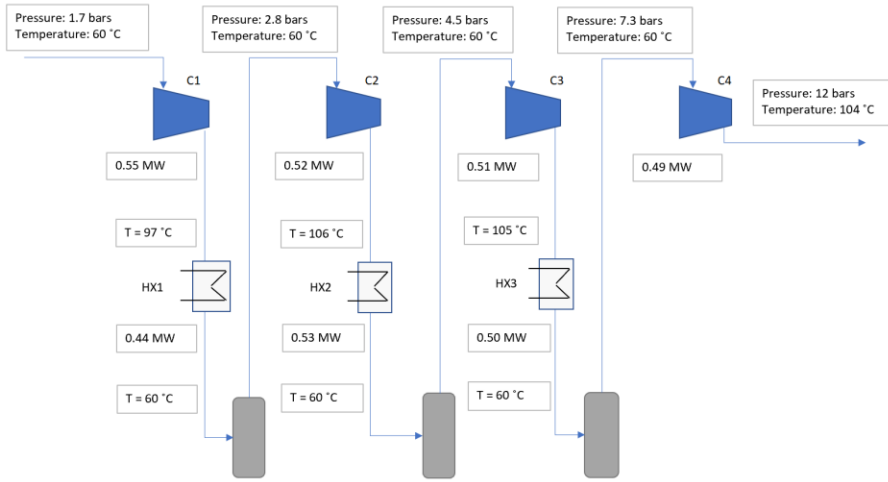


Figure 11- Compressor(s) PFD

Table 9 provides information on the electricity consumption for each compressor as well as the total energy required, while Table 10 presents the energy obtained from the heat exchangers.

Table 9- Compressor consumption

Equipment	Electricity required (MW)
Compressor 1	0.55
Compressor 2	0.52
Compressor 3	0.50
Compressor 4	0.49
Total	2

Table 10-HX energy requirement

Equipment	Energy required (MW)
Heat Exchanger 1	-0.44
Heat Exchanger 2	-0.52
Heat Exchanger 3	-0.50
Total	-1.47

The energy requirements of the CC (Carbon Capture) unit for the base case are summarized in Table 11.

*Table 11-Total energy demand of the CC and compression unit*

Captured CO <sub>2</sub> (kg/s)	12.07
Reboiler duty (MW)	50
Steam demand (kg/s) @ 150°C, 4bars	24
Condenser (MW)	-19
Lean cooler (MW)	-47
Heat exchanger(s) (MW)	-1.4
Compressor(s) (MW)	2
Lean and rich pumps (MW)	0.9

The energy demand of the CC unit can be influenced by several parameters, including the lean and rich loadings and the pressure of the stripper. The impact of each parameter will be discussed in the subsequent chapter.

## 6. Discussion

Based on the findings from the energy demand calculation and literature review, it was determined that the regeneration step constitutes the most energy-intensive phase within the aqueous MEA process. This phase exerts a substantial impact on the overall operating cost, accounting for approximately 70-80% of the total energy consumption (Aaron & Tsouris, 2005). This crucial stage involves the separation of CO<sub>2</sub> from the amine solvent, requiring substantial energy input to restore the amine's capturing capacity. Notably, commercial plants utilizing aqueous MEA technology to capture CO<sub>2</sub> consume an estimated 3.5-4.3 GJ/ ton CO<sub>2</sub>. This energy consumption serves as an indicator of the energy efficiency and resource demands associated with the utilization of aqueous MEA for large-scale CO<sub>2</sub> capture. The development of more efficient regeneration methods or alternative solvents holds the potential to reduce the energy consumption and operating costs associated with this process (Arachchige et al., 2013; Karlsson, 2021).

This chapter focuses on investigating the impact of various parameters on MEA carbon capture technology. The primary objective of the study was to identify an optimized process that exhibits a reduced thermal energy demand compared to the base case. Several key parameters were examined, including the CO<sub>2</sub> lean and rich solvent loading and the stripper pressure. These parameters were carefully varied to assess their influence on the performance of the carbon capture process. By analyzing these crucial parameters, the study aimed to uncover potential strategies for optimizing the efficiency and energy requirements of MEA carbon capture technology.

The partial pressure of CO<sub>2</sub> is a significant factor that influences both the absorption and desorption processes (Feron, 2016). It is through the partial pressure of CO<sub>2</sub> in the incoming gas that the maximum rich loading or absorption capacity can be estimated. This parameter provides valuable insights into the amount of CO<sub>2</sub> that can be effectively captured and absorbed in the system. Figure 12 presents literature data depicting the solubility of CO<sub>2</sub> in 30 wt% aqueous MEA at two different temperatures, namely 40 and 120°C (Karlsson, 2021).

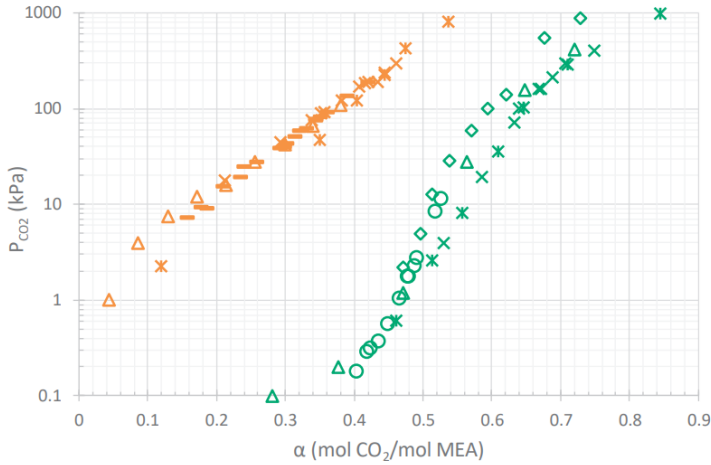


Figure 12- The provided graph presents an overview of solubility data obtained from literature sources for the absorption system utilizing 30 wt% aqueous MEA. The data specifically focuses on the solubility of CO<sub>2</sub> at two different temperatures: 40°C (depicted in green) and 120°C (depicted in orange).

The initial step of optimization of the CC unit involves determining the maximum rich loading at a temperature of 40°C by considering the partial pressure of CO<sub>2</sub> in the incoming flue gas. For the partial pressure of 15 kPa from the incoming flue gas, the rich loading equals to 0.54. Subsequently, the focus shifts to identifying the optimal lean loading that minimizes the energy consumption of the reboiler, while simultaneously ensuring that the temperature of the rich solution remains within the range of 40 to 60°C.

The findings from testing different lean loading for the base case indicate that, in order to achieve the maximum rich loading for the given flue gas, an optimal lean loading of approximately 0.37 is recommended. It is worth mentioning that the typical pressure in the stripper column is approximately 2 bars (Karimi et al., 2011). However, in the base case study discussed in this work, with a lean loading of 0.37, the pressure in the stripper is slightly higher, specifically 2.2 bars.

In an alternative approach, adjustments were made to both the rich and lean loadings in order to identify the optimal values and enhance energy efficiency. In this case, the priority was to maintain  $T_{rich}$  below and equal to 50°C, which aligns with common practices. Additionally, the stipulated objective was to keep the stripper pressure near or below 2 bars, further contributing to the overall optimization efforts.



Figure 13 illustrates the outcomes concerning the elevation of stripper pressure and the reboiler duty, which were derived from the variation in lean solvent loading.

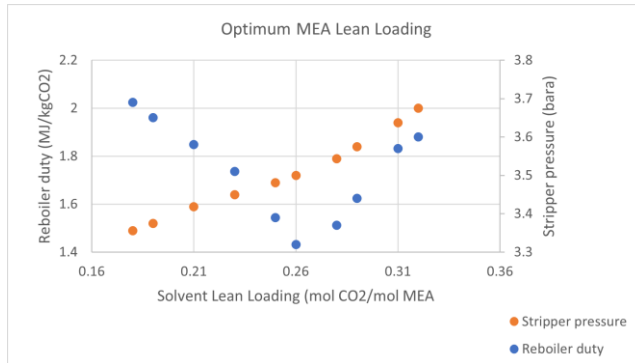
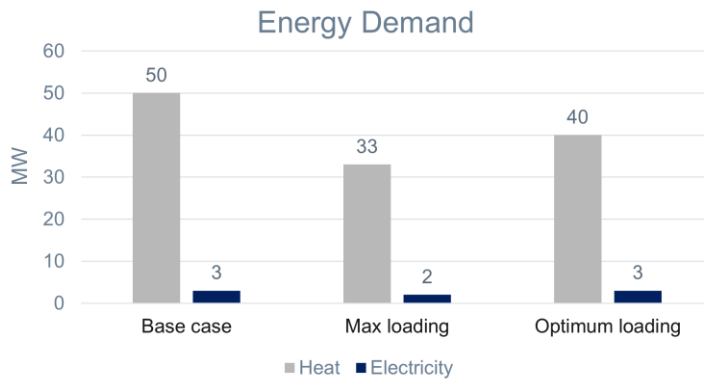


Figure 13- Optimum MEA lean loading depending on reboiler duty and stripper pressure.

At lower values of lean solvent loading, the primary factor influencing the thermal energy requirement is the amount of stripping steam necessary to achieve the desired low solvent loading. Conversely, at higher values of lean solvent loading, the dominant factor becomes the heating of the solvent at elevated circulation flow rates. Consequently, it is expected that a minimum thermal energy requirement will be reached. Figure 13 confirms this expectation by demonstrating a decrease in thermal energy requirement as lean solvent loading increases until reaching a minimum point. This minimum point, representing the optimal lean solvent loading, is defined as the point at which the energy requirement is lowest. For a 90% removal rate using a 30 wt.% MEA solution, the optimal lean solvent loading was found to be approximately 0.26-0.27 mol CO<sub>2</sub>/mol MEA, with a corresponding thermal energy requirement of 3.31 GJ/ton CO<sub>2</sub>. These findings align well with industry-reported data.

Through a comparison of energy demands between the two approaches, it is evident that both exhibit reduced energy requirements for the reboiler when compared to our base case (Figure 14). Notably, increasing the operating pressure of the stripper leads to a significant decrease in energy demand. However, it is crucial to acknowledge and carefully assess the potential challenges associated with higher pressure, such as amine degradation and corrosion risks. The implications of elevated pressure on the stripper design and construction must be thoroughly evaluated and considered. As an alternative, operating at lower pressures in a thermally integrated power plant can be cost-effective.



*Figure 14 - Energy demand comparison*

## 7. Conclusion

Carbon capture and sequestration (CCS) is a critical technology that plays a vital role in securing our future. Numerous methods have been proposed in the literature for capturing CO<sub>2</sub> from the atmosphere. This study focused on modeling the post-combustion MEA carbon capture technology to assess its energy requirements. While the research focused on CO<sub>2</sub> removal from a biomass power plant, the findings can also be extended to the study of CO<sub>2</sub> removal from other types of plants.

In this study, the optimum process utilizing a 30 wt.% MEA solution resulted in an energy requirement of 3.3 GJ/ton CO<sub>2</sub>. These findings hold promise for reducing costs and improving the overall efficiency of the carbon capture process for the specified flue gas data.

It was found that the lean solvent loading has a significant impact on the process performance, particularly the thermal energy requirement. Therefore, optimizing the solvent processes requires careful consideration of the lean solvent loading as a key factor.

Increasing the operating pressure in the stripper offers the potential for higher regeneration efficiency and reduced thermal energy requirements. Additionally, it can lead to cost savings and decreased energy consumption for CO<sub>2</sub> compression. However, addressing the challenges associated with higher operating pressures, such as amine degradation and corrosion, is important.

## 8. Recommendations and Future Work

The following recommendations and future work suggestions aim to enhance the understanding and implementation of carbon capture technologies, leading to improved efficiency and sustainability.

- It is crucial to conduct investigations into alternative carbon capture technologies and explore the feasibility of retrofitting them to the case study.
- More realistic designs based on current research and industry practices should be considered. This may involve incorporating a mixture of absorbents and implementing advanced process designs.
- More comprehensive calculations should be performed, taking into account factors such as absorbent and solvent losses, as well as MEA makeup requirements.
- The potential for Carbon Capture and Utilization (CCU) was not considered for the captured CO<sub>2</sub> at the plant, which could have implications for cost efficiency.
- From an environmental standpoint, monitoring the amine content and assessing the extent of degradation in the solvent is crucial. It is essential to address the risks associated with handling solvents and to consider the effects of local emissions, as well as comply with relevant legislation regarding solvent handling.
- Although the carbon capture system is positioned downstream of the flue gas cleaning system, further improvements and optimizations should be made to the flue gas conditioning processes. The aim should be to minimize the deterioration of the amine by significantly reducing the levels of SO<sub>x</sub>, NO<sub>x</sub>, oxygen, and particulate matter below the tolerance threshold of the amine process.

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