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Comparison of CLC with existing CCS technologies: based on Energy, Exergy and Techno-economic analysis

Master's thesis in Sustainable Energy Systems

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Cover: An image of flue gas from a power plant being emitted into the atmosphere

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Abstract

Climate change is an immediate global crisis that is driven by the accumulation of greenhouse gases, primarily carbon dioxide, in the Earth´s atmosphere. Carbon Capture and Storage (CCS) technologies play a pivotal role in reducing our carbon footprint as we move towards a more sustainable future. This study aims to compare four different carbon capture technologies and evaluate them based on techno-economic analysis, energy analysis and exergy analysis.

The four technologies under evaluation are chemical looping combustion (CLC), oxy-fuel combustion, post-combustion, and pre-combustion. The process modelling of the four systems for a selected combined heat and power (CHP) plant is used to carry out this comparison, which is next followed by energy, energy-related, and economic analysis. The modelling of these four plants was conducted using Aspen Plus simulation software and validated using the results given in the literature.

The study reveals that the CLC capturing technology outperforms other CCS processes regarding cost efficiency, CO_2 capture rate and electrical efficiency. The highest overall efficiency was observed for Oxyfuel combustion but the CLC plant has the best net electric efficiency. For the CLC model, An almost pure stream of CO_2 is extracted from the Fuel reactor which is then subsequently compressed after heat extraction. The CLC has the highest CO_2 capture rate among the four models for this reason. From the exergy analysis, it becomes clear that the most exergy destruction takes place in the combustion process for all the cases and that it needs to be optimized to improve overall efficiency. The comparison of the overall exergy destruction of the models shows us that CLC has the least exergy destroyed and thus has the highest exergy efficiency, despite incurring higher initial expenses, CLC displays the lowest levelized cost of electricity (LCOE) and payback period when compared with all the other technologies. This indicates that CLC has been estimated to be the most economical choice over the long term, despite necessitating a more substantial initial expenditure.

To sum up, this case study offers insightful information about the advancement of CLC technology and its advantage over competing carbon capture solutions. The results provide a useful manual for energy sector decision-makers, highlighting how important it is to select the appropriate carbon capture and combustion technology with the intent to optimise performance and economics .

Keywords: Carbon capture and storage; chemical looping combustion, Aspen plus simulation, thermodynamic and techno-economic analysis.

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Avinash Pramode, Gothenburg, November 2023

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Abhie Dhait, Gothenburg, December 2023

List of Acronyms

Below is the list of acronyms that have been used throughout this thesis listed in alphabetical order:

AGR	Acid Gas Removal
AR	Air Reactor
ASU	Air Separation Unit
CC	Carbon Capture
CCO_2	Captured Carbon Dioxide per Year
CCS	Carbon Capture and Storage
CE	Capture Efficiency
CH_4	Methane
CHP	Combined Heat and Power
CLC	Chemical-Looping Combustion
CO	Carbon monoxide
CO_2	Carbon Dioxide
CFB	Circulating Fluidized Bed
CRF	Capital Recovery Factor
Cp	Pump Cost
C_{DH}	District Heating Selling Cost
C_{ES}	Cost Of Electricity Sold
$C_{O\&M}$	Cost Of Operations and Management
C_{EU}	Cost Of Electricity Used
C_{TF}	the cost of total fuel used per year
DH	District Heating
ER_{CO_2}	Emission rate of Carbon Dioxide
FC	Fuel Cost
FCV	Fuel Calorific Value
FG	Fluegas
FR	Fuel Reactor
H2	Hydrogen
H_2O	Water
H_2S	Hydrogen Sulphide
HHV	Higher Heating Value
HRSG	Heat Recovery Steam Generators
IGCC	Integrated Gasification Combined Cycle

LF	Load Factor
LCOE	Levelized Cost Of Electricity
LHV	Lower Heating Value
MEA	Monoethanolamine
MW	Mega-Watts
N_2	Nitrogen
NO_x	Nitrogen Oxides
NPO	Net Power Output
OC	Operating Capacity
O&M	Operations and Management Cost
P_r	Annual Profits
PBP	Payback Period
ROI	Rate of Interest
RMEA	Rich-Monoethanolamine
SOE	Send Out Electricity
SO_x	Sulphur oxides
TCR	Total Capital Requirement
TEA	Techno-Economic Analysis
TFU	Total Fuel Used
TPO	Total Plant Output
WGS	Water Gas Shift
W_p	Pump Work

Nomenclature

Below is the nomenclature of indices, sets, parameters, and variables that have been used throughout this thesis.

E_{x_ph}	Physical exergy
h_0	Enthalpy at reference state
h	Enthaply at current state
T	Temperature at current state
T_0	Temperature at reference state
s_0	Entropy at reference state
s	Entropy at current state
$E_{x_{ch}}$	Chemical exergy
$E_{x_0ch,i}$	Chemical exergy at reference state
E_f, k	Exergy of fuel
E_p, k	Exergy of product
E_d, k	Exergy destroyed
ϵ	Exergy efficiency
$\Delta\ Hrxn$	Enthalpy of reaction
η	Efficiency
Ex	Exergy flow
Q	Heat
H	Hydrogen content
C	Carbon content
O	Oxygen content
N	Nitrogen content

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1

Introduction

In today's modern world, energy serves as the life force of economic growth, technological advancements, and overall human prosperity. However, the insatiable need for energy on a worldwide scale, which is predominantly provided by burning fossil fuels, has given rise to a grave concern: global warming and a surge in greenhouse gas emissions.

Global warming poses an imminent danger to our planet because of the rising levels of greenhouse gases like carbon dioxide (CO_2), and methane (CH_4)in our atmosphere. Large amounts of CO_2 have been released due to the widespread use of fossil fuels in transportation, industry, and the production of electricity, which has intensified the greenhouse effect. The trapping of solar heat because of these gases has had a profound effect on an increase in world temperatures and the disruption of climate patterns. The effects are substantial and include, among other worrisome occurrences, rising sea levels, harsher weather conditions, and loss of biodiversity.

It is now extremely necessary to address the various issues of modern energy use, global warming, and greenhouse gas emissions. To effectively mitigate the far-reaching effects of climate change[5], a decisive shift towards renewable and sustainable energy sources must be made in tandem with comprehensive initiatives to reduce emissions. All of these efforts are essential to keep our planet healthy and, also for ensuring a sustainable future for our future generations.

1.1 Background and motivation

The rising global awareness of reducing carbon emissions is mirrored via international agreements and initiatives that emphasize mitigating climate change. Numerous nations have signed the Paris Agreement, which promotes the importance of keeping global warming, far below 2 degrees Celsius, above pre-industrial levels.

In light of the growing worries surrounding global climate change and the pressing need for sustainable energy solutions, carbon capture and storage (CCS) technologies have emerged as an essential tool in our efforts to mitigate carbon dioxide (CO_2) emissions.

There are conventional ways such as Pre-Combustion, Post-Combustion, and Oxy-

Fuel, as well as the newly developed method dubbed Chemical Looping Combustion (CLC) among these technologies[6]. These techniques show promise in absorbing CO_2 emissions from a variety of industrial processes and power plants. These different CCS techniques operate at distinct stages in the combustion process, each with its particular advantages and disadvantages. To ensure widespread application and enhance efficiency, it is crucial to delve into the complexities of these systems through a comprehensive investigation.

The drive for this thesis stems from the desire to thoroughly analyze different carbon capture systems from both technical and economic standpoints, with a particular emphasis on the CLC system. As the entire globe grapples with the repercussions of greenhouse gas emissions, there has never been a greater need for practical, efficient, and economically viable solutions. By performing a thorough exergy analysis, which focuses not only on the energy quantity but also on the quality and usability of energy, and with a techno-economic assessment, we can gain valuable insights into these systems' performance, efficiency, and cost-effectiveness.

Additionally, this thesis tries to fill a research gap by thoroughly contrasting these various carbon capture methods. Although certain systems have been examined separately, an in-depth analysis that takes into account both the exergy and the techno-economic aspects is necessary for assisting in industrial and policy decisions. This thesis looks at all of these approaches to determine which is the most effective and financially sound strategy for particular industrial contexts, advancing the adoption of carbon capture systems globally.

The results of this study can potentially assist the readers make informed choices on the use of carbon capture systems. We have the potential to unlock a future in which carbon capture technologies play a crucial role in limiting climate change and ensuring a sustainable energy future for future generations as we delve deeper into the complexity of these systems through thorough exergy analysis and precise techno-economic evaluations.

1.2 Aim and scope

The primary aim of this thesis is to create and construct a Chemical Looping Combustion (CLC)-based carbon capture model and compare it to the three established Carbon Capture and Storage (CCS) procedures: post-combustion, pre-combustion, and oxyfuel combustion. A wide range of criteria, involving studies of the techno-economic, energy, and exergy analysis, will be used to conduct this comparison.

The scope of this thesis is to simulate a particular combined heat and power (CHP) system using Aspen Plus while maintaining a set of consistent key parameters that are the same across all simulation models to allow for fair and relevant comparisons.

Energy analysis will help us to determine the system's energy efficiency and perfor-

mance whereas, exergy analysis will reveal the thermodynamic losses and point out opportunities for improvement. The techno-economic analysis will help us assess whether the proposed CLC system is more economically viable than the traditional CCS technologies.

2

Literature Review

2.1 Carbon Capture and Storage

Carbon Capture and Storage (CCS) is a set of technologies and strategies that are designed to mitigate greenhouse gas emissions, principally carbon dioxide (CO_2), by capturing CO_2 emissions from processes effectively and securely storing them. CCS is regarded as a vital climate change response, particularly in industries where lowering emissions is difficult. The three basic steps of the CCS process are carbon collection, transportation, and storage.

2.1.1 Post-Combustion Carbon Capture

Post-combustion carbon capture is another critical carbon capture technology, which intends to absorb the carbon dioxide(CO_2) emissions, after burning of the fuel. This technology is noteworthy because it can be adapted in existing power plants thus offering a flexible approach to lowering CO_2 emissions.

The post-combustion capture in our thesis uses an amine aqueous solution in a reactive absorption-solvent process. The two main sections can be identified as the absorption section, where carbon dioxide is transferred from the gaseous state to the liquid via the amine solution and the stripping section, where the solvent containing carbon dioxide is regenerated. The two sections are interconnected by a cross-heat exchanger[7]. A simplified flowsheet of the system is reported in 2.1

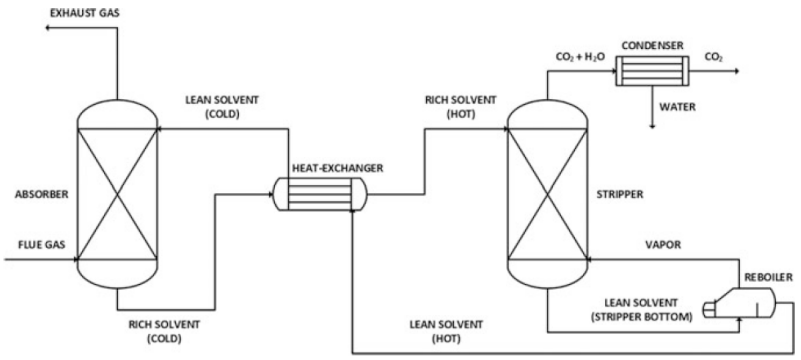


Figure 2.1: Post-combustion process[1]

The flue gas rich in CO_2 enters the absorber where it is exposed to the CO_2 lean solvent. Carbon dioxide-free flue gas exits the absorber and solvent rich in Carbon dioxide leaves the absorber and is pumped into a stripper through a heat exchanger where the solvent is regenerated via a reboiler. A train of high-purity CO_2 is obtained from the condenser and water is recovered from the same. The regenerated solvent is then led back into the absorber and the cycle continues.

The key aspects of post-combustion carbon capture technology include:

- Fuel combustion and exhaust flue gas: The process begins with the combustion of fuel and generates heat and/or, electricity. The exhaust gas, which typically contains carbon dioxide(CO_2), nitrogen(N_2), and sulphur oxides(SO_x), is then further sent to the carbon capture section.
- Flue gas treatment: The follow-up is the separation of carbon dioxide(CO_2) from the other components of the flue gas. This can be done by employing various methods including, the use of solid adsorbents, or utilizing amine-based solutions to absorb CO_2 from the flue gas.
- CO_2 retrieval: The CO_2 rich stream is further processed for purification. The absorbents are recycled and the retrieved CO_2 is compressed and stored.

Advantages of Post-Combustion Carbon Capture:

1. Retrofit Capability: One of the notable advantages of post-combustion carbon capture is its amazing adaptability. It enables lower levels of emissions without the need for major changes to the system and is readily integrated with power plants that are already in service.
2. Cost-Effective Deployment: Post-combustion carbon capture is a carbon capture technology that is both efficient and affordable. Unlike other carbon capture devices, it doesn't require major alterations to existing systems. Because of lower initial capital expenses, it's a more attractive choice for plenty of industries looking to cut back on their carbon footprint.
3. Reduced Downtime: Installing post-combustion carbon capture systems in existing buildings generally brings about little to no disruption. This is a huge benefit since it enables businesses to adopt carbon capture equipment with virtually no downtime.

Disadvantages of Post-Combustion Carbon Capture:

1. Energy Consumption: One of the main problems with post-combustion carbon capture is the amount of energy it consumes. Energy is needed in large amounts to isolate and capture carbon dioxide from exhaust gasses. This energy drain may cause power plants' net power output to decrease, which would increase operating costs.
2. Chemical Solvents: Chemical solvents are the main tool used in post-combustion carbon capture to absorb carbon dioxide from flue gases. Since these solvents are often toxic and corrosive, they must be handled and disposed of with caution.

3. Large Equipment and Space Requirements: Achieving significant carbon capture, especially when dealing with flue gases which contain a lower concentration of carbon dioxide(CO_2) often requires large and space-consuming equipment. Some sites may not be able to implement this because of space limitations.

2.1.2 Pre-Combustion Carbon Capture

Pre-combustion carbon capture, also known as pre-combustion CO_2 capture, in simple words, is a technology that focuses on capturing CO_2 emissions before the combustion stage of the process. It is frequently used in hydrogen generation systems and integrated gasification combined cycle (IGCC) power plants. The main goal of this technology is to transform carbon-based fuels like coal or biomass into synthesis gas (syngas), a mixture of hydrogen (H_2) and carbon monoxide (CO). Before burning, this syngas must go through a series of steps in order to extract CO_2.

The fuel is first gasified and used to produce syngas in a gasifier. This train of syngas along with some combustion products is then directed into Water Gas Shift reactors where a water gas shift reaction occurs.

The CO_2 produced is then removed from the syngas train via the Selexol process. The Selexol process is an absorption-based process that uses a solvent to remove H_2S from the syngas mixture along with CO_2 to produce nearly pure syngas. This thesis does not cover the specifics of the selector process. The syngas obtained can now be used as a fuel for combustion in a gas turbine and carbon dioxide is separated[8].

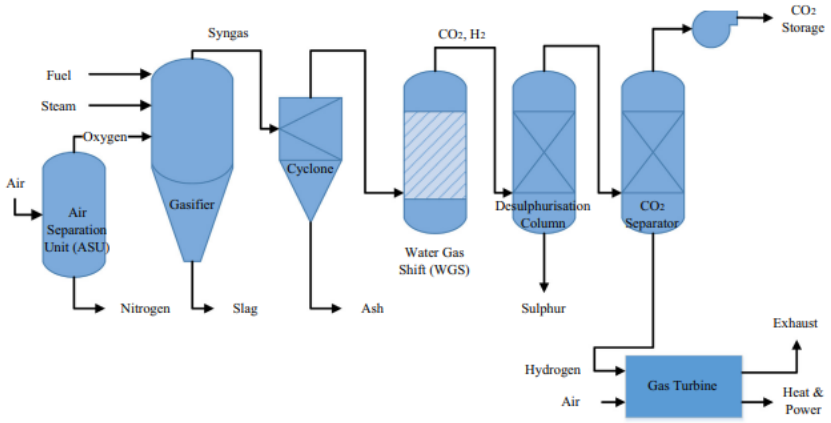


Figure 2.2: Pre-combustion process [2]

The basic elements in a pre-combustion carbon capture technology include:

- Fuel Gasification: The carbon-based fuel is first gasified, in a high-temperature process involving oxygen and/or steam.

- Syngas Cleanup: The syngas undergo a cleaning process to eliminate any impurities like the sulphur compounds. This is essential to ensure no poisonous gas emissions.
- Water-gas Shift Reaction: The CO-rich syngas undergoes the WGS reaction which is the conversion of CO to CO_2, along with H_2 production. This reaction leads to higher a higher hydrogen-to-carbon ratio in the syngas, thus preparing it for CO_2 capture.

Advantages of Pre-Combustion CCS:

1. Higher Efficiency: One of the most significant advantages of pre-combustion CCS is the ability to sustain a higher energy efficiency in power generation. This could lead to lower operating expenses and help the plant maintain its competitiveness.
2. Greater CO_2 Concentration: Pre-combustion CCS frequently has a higher concentration of CO_2. The capture process is made simpler and more economical by the higher CO_2 concentration, particularly for industrial applications where less energy and equipment are needed for separation.

Disadvantages of Pre-Combustion CCS:

1. Complex Integration: Pre-combustion CCS can be more challenging and costly to implement than post-combustion CCS since it needs extra steps to turn fuel into syngas or gasified form before to combustion. This complexity can lead to higher capital costs and increased maintenance requirements.
2. High Initial Investment: Pre-combustion CCS system construction and operation can involve large upfront costs. Some companies may be discouraged from implementing this technology due to the high expense of CO_2 separation and compression equipment, as well as the requirement for gasification or other fuel processing facilities.
3. Limited Applicability: Pre-combustion CCS is most suitable for facilities that use gasification processes or for applications like hydrogen production. Its broad adoption may be limited by the fact that it isn't suitable for many industries or power generation scenarios. Moreover, the technology depends on the infrastructure and availability of suitable feedstocks for fuel gasification.

2.1.3 Oxy-Combustion Carbon Capture

Oxy-combustion carbon capture, also known as oxy-fuel carbon capture, is a sophisticated method of carbon capture that differs from our previous two entries. Air, which has around 78% nitrogen and 21% oxygen, is used in the conventional combustion process. However, in the system of oxy-fuel, pure oxygen(or enriched oxygen) is used for the combustion process resulting in the generation of a substantially greater concentration of carbon dioxide. This distinctive feature makes oxy-combustion a promising carbon capture technology.

This carbon dioxide-rich flue gas is treated for other pollutants before some of it is separated for liquification and storage while the rest is mixed with fresh air with 95% pure oxygen and directed to the combustion chamber for fuel oxidation. It is worth noting that carbon dioxide does not take part in the oxidation of fuel and is simply a supporting material[9].

Oxy-combustion carbon capture technology operates in the following ways:

- Oxygen separation: Pure oxygen is separated from the air, using an air separation unit(ASU). This step consumes electricity, and the separated oxygen can then even be sent at the desired levels of pressure conditions.
- Combustion: The fuel is then burned, in a boiler or furnace, in the presence of this pure oxygen, creating a highly concentrated CO_2 exhaust stream.
- Flue gas re-circulation: This is a technique unique to oxy-fuel carbon capture, and is used to improve combustion stability, control flame temperature and enhance the overall efficiency of the process.

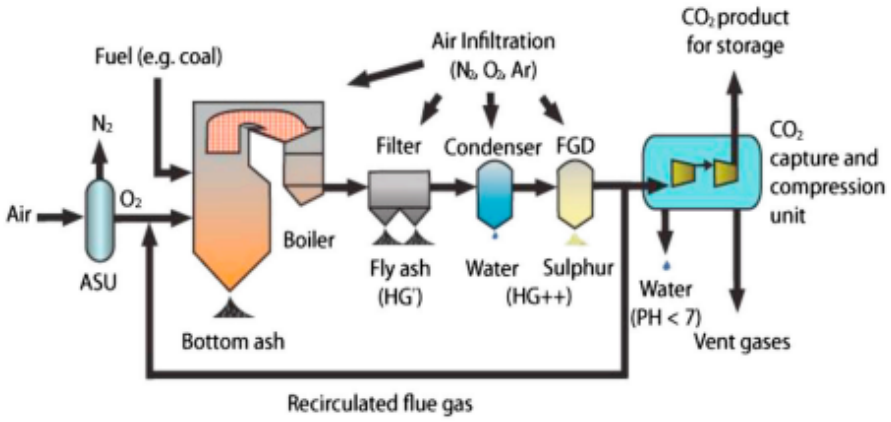


Figure 2.3: Oxyfuel process [3]

Advantages of Oxy-Combustion CCS:

1. High CO_2 Capture Efficiency: Oxy-combustion CCS has the potential to attain high capture rates—typically above 90%. The combustion process in an oxygen-rich environment produces a flue gas stream with a higher concentration of CO2, which makes it easier to capture as compared to conventional air combustion.
2. Reduction of Impurities: Oxy-combustion reduces the emissions of pollutants and impurities, such as nitrogen oxides (NO_x) and sulfur dioxide (SO_2), which usually appear in flue gases. Cleaner emissions can come from streamlining the capture and purification procedure.
3. Flexible Application: Oxy-combustion CCS can be retrofitted to existing power plants and industrial facilities, allowing for the capture of CO_2 from sources that are already in operation. This flexibility is essential for making the transition to a low-carbon energy system.

Disadvantages of Oxy-Combustion CCS:

1. Energy Intensive: Oxy-combustion requires the separation of oxygen, which is a process that consumes energy. This energy-intensive oxygen separation stage lowers the power plant's or industrial process's net energy efficiency and reduces the process's viability from an economic standpoint.
2. High Costs: In comparison to other CCS methods, oxy-combustion CCS has comparatively higher prices. These costs encompass the oxygen generation process, CO_2 capture and compression, and transportation and storage infrastructure. These costly initial and ongoing expenses may prevent this technology from being widely utilized.
3. Oxygen Supply and Infrastructure: Oxy-combustion CCS requires a steady consistent supply of pure oxygen. It can be difficult and expensive to set up the infrastructure necessary for the production, storage, and transportation of oxygen, especially in large-scale applications. Smaller facilities will find the technology less accessible as a result.

2.1.4 Chemical Looping Combustion Carbon Capture

Chemical Loop Combustion (CLC) has gained attention in academic circles since the 1950s as a groundbreaking alternative to traditional combustion methods. [10] Notably, it has emerged as an appealing choice for a more sustainable approach to biomass combustion due to its innate capability to effectively separate CO_2 from other flue gas components.[11] This distinguishing feature minimizes, or in some cases entirely negates, the energy penalties typically associated with carbon capture technologies.

CLC primarily operates through a sophisticated double-fluidized bed system, comprising an air reactor (AR) and a fuel reactor (FR). Within this system, a circulating oxygen carrier takes the form of metal oxide particles, and its dynamic journey involves successive oxidation and reduction processes during circulation.

In the fuel reactor (FR), the oxygen carrier facilitates the oxidation of the fuel, ultimately yielding a pure outlet stream primarily consisting of CO_2 and H_2O.[11] The regeneration of the oxygen carrier occurs in the air reactor (AR) as it is mixed with an air stream. This regeneration process enables the oxygen carrier's reuse in the fuel reactor, thereby completing the CLC cycle.

Solid fuels consist of moisture, ash, char, and volatiles. These solid fuels go through three distinct stages of mass loss: drying, devolatilization, and char combustion/-gasification.

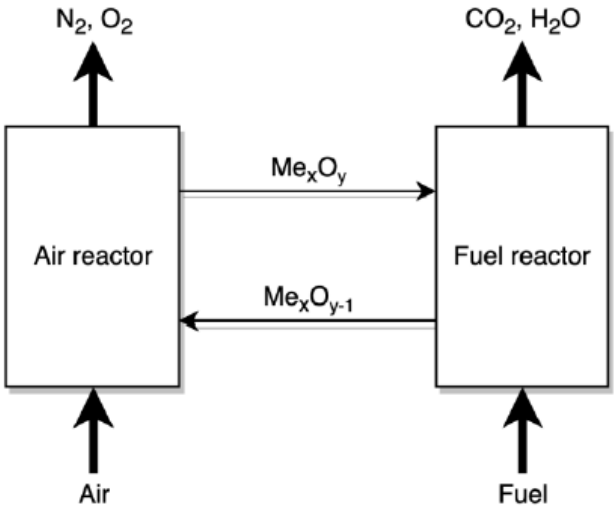


Figure 2.4: AR-FR cycle [4]

The initial two stages, drying and devolatilization, are fundamentally endothermic processes. Drying involves the release of moisture from the fuel, maintaining a constant temperature until all moisture is extracted. Subsequently, devolatilization commences as volatiles, including H_2, CO, CO_2, H_2O, and short-chained hydrocarbons, are released under the influence of sufficiently high pyrolysis temperatures. These volatiles undergo several chemical reactions, dependent on the surrounding gas environment. The residual char undergoes combustion. This phase typically lasts for a longer extended duration compared to the preceding drying and devolatilization stages. After the combustion of the fuel particle, ash remains due to the presence of minerals. [12]

The fuel reactor (FR) contains volatile gases and char from the solid fuel that go through various chemical reactions. Importantly, these reactions are endothermic in nature. Because the conversion of char is a slow process, a small fraction remains unreacted and exits the FR to the AR due to insufficient time for reaction. Additionally, a portion of volatiles accompanies the flue gas from the FR.

In an ideal scenario, the gases leaving the AR should primarily consist of nitrogen with minimal oxygen content. However, the previously mentioned unreacted char that enters the AR undergoes oxidation by the oxygen present in the air. This results in the formation of a small amount of CO_2, which cannot be effectively captured.

The reactions within the air reactor (AR) are significantly exothermic, generating heat that plays a crucial role in providing the necessary energy for the endothermic fuel reactor (FR). Achieving the right balance in the circulation of the oxygen carrier is vital for transferring sufficient heat to the FR.

One implication of the differing heat characteristics of the FR and AR is the need for the FR to be adiabatic to prevent heat losses, while the AR should be cooled

to prevent overheating. Overheating in the AR can result in issues with the oxygen carrier, such as the agglomeration of bed materials. (21)

Advantages of Chemical-Looping Combustion CCS:

1. Efficient Carbon Capture: CLC's excellent capture efficiency is well-known. It is capable of capturing a large amount of CO_2 emissions during combustion by using metal oxides to trap CO_2. Because of its great efficiency, power plants and industries have a lower overall carbon footprint.
2. Lower Energy Penalty: CLC usually has a lower energy penalty as compared to other carbon capture systems. This implies that it is a more energy-efficient choice for reducing emissions as it needs less extra energy to capture CO_2.
3. Fuel Flexibility: CLC can be applied to a variety of fuels, including natural gas, biomass, and solid fuels. Because of its versatility, this technology is a flexible one that can be incorporated into power plants and existing industrial processes without requiring serious changes to the fuel source.

Disadvantages of Chemical-Looping Combustion CCS:

1. Complex Design and Operation: CLC-CCS systems can be complex to design and operate. The chemical reactions require precise control and optimization, which can be challenging as well as costly.
2. High Initial Costs: CLC-CCS system installation might be costly. Many industrial enterprises and power plants may find it hard to adopt these systems owing to the initial capital costs associated with their construction.
3. Limited Proven Scale: CLC-CCS is not as commonly utilized or established at bigger scales as some other CCS technologies, although exhibiting promise in lab and small-scale experimental operations. Its short track record may raise questions about its reliability and long-term success.

2.2 Energy Analysis

Energy analysis plays a pivotal role in assessing the efficiency and practicality of carbon capture processes for reducing greenhouse gas emissions. It's vital to delve into the key concepts, methodologies, and the significance of energy analysis within various carbon capture technologies.

Thermodynamic Principle: Energy analysis in carbon capture hinges on the fundamental principles of thermodynamics. These principles provide the theoretical underpinning for evaluating how energy flows within a system and how it undergoes transformations. In the context of carbon capture, the central goal is to assess how efficiently energy is utilized and converted throughout the process.

Energy Flows and Transformations: Energy flows are meticulously tracked at every stage of carbon capture. This involves identifying energy inputs, such as the

energy needed for capturing and separating CO_2, as well as energy outputs, which may include the useful energy generated by the power plant. Furthermore, energy transformations are observed, such as the conversion of chemical energy from fossil fuels into thermal energy during combustion.

Significance of Energy Analysis:

1. Efficiency Assessment: Energy analysis quantifies how effectively a carbon capture process reduces CO_2 emissions while maintaining energy production. It provides a vital measure of operational efficiency.
2. Resource Allocation: By identifying energy losses and areas for improvement, energy analysis guides resource allocation, whether for retrofitting existing facilities or designing new ones, ensuring efficient use of resources.
3. Environmental Impact: Improved energy efficiency in carbon capture processes reduces CO_2 emissions and lessens the environmental footprint. Energy analysis helps estimate these reductions, aiding environmental impact assessments.
4. Economic Implications: Enhanced energy efficiency often lowers operational costs, making carbon capture technologies more economically viable. This economic benefit is crucial for the broader adoption of these technologies in the transition to a low-carbon economy.

2.3 Exergy Analysis

Exergy is defined as the amount of work a system can perform when it is brought into thermodynamic equilibrium with its environment. In other words, it is the amount of useful work that can be extracted from a stream when it goes from a state of energy to a reference state. Exergy analysis serves as a widely employed tool for identifying process irreversibility in the form of losses and quantifying their impact on overall process efficiency, providing insights into the quantitative aspects of available energy utilization, environmental considerations, and economic performance [48–50]. Unlike energy, exergy doesn't adhere to the conservation rule. The magnitude of exergy destruction is determined by calculating the exergy balance over a particular component, defined as the difference between the exergy input to that component (comprising the total exergy of streams and supplied work) and the exergy output from the component (comprising the total exergy of streams and the work output) [13]

The exergy of a process stream can be categorized into two components: physical exergy and chemical exergy. Physical exergy (E_{x_ph}) represents the maximum useful work attainable by transitioning a unit of mass of a substance from its current state (T, P) to the environmental/reference state (To, Po) solely through physical processes]. The formula for the physical exergy component is given by Eq. (2,1).

$$E_{x_{ph}} = (h - h_0) - T_0(s - s_0) \tag{2.1}$$

In contrast, chemical exergy embodies the maximum useful energy achievable by transitioning from the environmental state to the reference state through chemical processes. These processes involve reactants and products at environmental temperature and pressure, even when the stream composition is not in chemical equilibrium with the environment. [14]

$$E_{x_{ch}} = \sum_i x_i E_{x0_{ch,i}} + RT_0 \sum_i x_i \ln x_i \tag{2.2}$$

The exergy analysis in this project was conducted based on the fuel-product definition. Fuel exergy is defined as the input exergy or imported net exergy and the product is defined as the product or exported net exergy.[13] The exergy balance equation can be expressed as:

$$E_{f,k} = E_{p,k} + E_{d,k} \tag{2.3}$$

Where,
E_f, k = Exergy of fuel,
E_p, k = Exergy of product
E_d, k = Exergy destroyed

Exergy destroyed refers to the irreversibility or losses. It should be noted that for dissipative components (such as coolers, mixers and valves) no useful exergy is produced but exergy is destroyed. The exergy efficiency of a system component can be expressed as the ratio of product exergy to the fuel exergy.

$$\epsilon_K = \frac{E_{p,k}}{E_{f,k}} \tag{2.4}$$

In this study, the reference state for exergy calculation is set at 25°C (T_o) and 1 bar (P_o).

2.4 Techno-Economic Analysis

Techno-economic analysis plays a pivotal role in assessing the feasibility and economic viability of carbon capture processes. In this section, we delve into the key concepts and methodologies associated with techno-economic analysis within the context of evaluating carbon capture technologies.

Methodologies for Techno-economic Analysis:

1. Capital Cost Estimation: The initial investment required for constructing and commissioning a carbon capture facility is a critical factor. This includes costs for equipment, materials, labour, and engineering.
2. Operating Cost Assessment: Ongoing expenses associated with the operation of the carbon capture system must be estimated. This encompasses costs such as energy consumption, maintenance, labour, and raw material requirements.

3. Lifecycle Analysis: Taking a holistic approach, this analysis considers both the environmental and economic impact of the technology throughout its lifecycle. This involves evaluating factors like emissions, resource consumption, and end-of-life disposal.

4. Sensitivity Analysis: This involves examining how variations in input parameters, such as energy prices, carbon prices, and regulatory incentives, affect the economic viability of the carbon capture process.

5. Return on Investment (ROI) Analysis: Assessing the time it takes for the initial capital investment to be recouped through operational savings or revenue generated by the captured carbon is crucial. This metric is vital for understanding the economic attractiveness of the technology.

Carbon Capture and Storage (CCS) techno-economic analysis is critical for comprehending and assessing the viability and impact of CCS technology. Here are five key elements that emphasize its significance:

1. Making Investment Decisions: Techno-economic analysis helps assist investors and stakeholders in making informed decisions about the implementation of CCS technologies. It gives essential insights into the financial viability, hazards, and estimated returns on investment for CCS projects, which are required for getting funding and support.

2. Policy Development and Support: Policymakers frequently rely on techno-economic analysis to establish and strengthen carbon reduction measures. Governments may develop effective motivations, regulations, and subsidy programs that promote CCS use and combat climate change by recognizing its economic ramifications.

3. Carbon Emission Reduction Targets: CCS is critical to meeting carbon emission reduction targets. Techno-economic analysis assists in determining how CCS can contribute to these goals, as well as estimating the cost-effectiveness of various CCS systems in comparison to other mitigation approaches.

4. Market Viability: Understanding the economic aspects of CCS is critical for determining its market viability. The analysis aids in the identification of prospective markets for captured carbon products, such as enhanced oil recovery or carbon credits, as well as the assessment of CCS's competitiveness in these markets.

5. Risk Mitigation: CCS initiatives include considerable financial and technical risks. The techno-economic analysis aids in evaluating these risks and establishing mitigation methods.

2.5 Software

2.5.1 Aspen Plus

Aspen Plus® is the primary process simulation software exclusively used by the chemical industry and has many different applications. Chemical and process engineers use it to simulate reactors, separators, pressure changes, heat exchangers, and various other unit operations.

Using a graphical interface, process flow diagrams are constructed by connecting different unit operations. Aspen plus simulations involve the handling of solid materials, fluids, and electrolytes, drawing from an extensive database of physical properties, mathematical equations, and thermodynamic models.

3
Methodology

3.1 Biomass and Oxygen Carrier

The fuel used in Örtoftaverket mainly consists of forestry biomass and some recycled wood and peat. As the fuel mainly consists of forestry biomass, the composition of the fuel is assumed to be the same as that of forest residue in Sweden [15]. The calorific, proximate, and ultimate values of the fuel are listed in the table (3.1).

We assume that the fuel has already been dried to a moisture content of 10% [16] before we use it for combustion. The costs for drying the fuel are not taken into consideration either, as the cost of the fuel includes the drying costs.

Table 3.1: Fuel Properties Based on Proximate and Ultimate Analysis

Parameters	**Value**
Higher Heating Value (MJ/kg)	20.54
Proximate Analysis (wt%)	
Moisture	10.00
Fixed Carbon	21.85
Volatile Matter	74.10
Ash	4.05
Ultimate Analysis (wt%)	
Ash	4.05
Carbon	51.00
Hydrogen	5.80
Nitrogen	0.90
Chlorine	0.00
Sulphur	0.04
Oxygen	38.21

To ensure that the results of the process simulations could be compared fairly, some standard parameters had to be established. The first parameter set was the fuel flow for the combustion chamber. We chose to scale the simulation model for a fuel input of 100MW. The fuel flow rate was then determined using the Lower Heating Value of the fuel, as indicated in the table (3.1).

$$Energy\ input(MW) = LHV(MJ/Kg) * fuel\ flow\ rate(Kg/s) \tag{3.1}$$

Specifically, for the CLC model, Ilmenite was chosen as the oxygen carrier. The composition of the Ilmenite ore was extracted from Cheng et al[17]., where all the most important constituents of Ilmenite ore are mentioned. Out of the many constituents of Ilmenite, the major ones have been used in the simulation. The composition of the OC is tabulated in table (3.2):

Table 3.2: Composition of Ilmenite

Constituents	Composition (Wt%)
Fe_2TiO_5	54.7
Fe_2O_3	11.2
SiO_2	5.5
TiO_2	28.6

In the Aspen plus model, Fe_2TiO_5 is defined as Fe_2O_3 + TiO_2, where SiO_2 and TiO_2 are inert.

3.2 Simulation

3.2.1 Simulation set-up

An appropriate stream class must be set to begin an Aspen plus simulation. The stream class MIXCINC was used for all the simulations, which allowed the mixture of three sub-streams: MIXED, Conventional inert (CI) solid, and non-conventional solid. The substream "MIXED" accounts for all the components in the liquid-vapour phase, whereas the "CI solid" and "NC solid" consist of all the solid compounds with and without molecular weights. Ash and biomass are defined as "NC solids", while OC is defined as "CI solid".

Biomass is defined as a non-conventional component where its attributes are entered according to the data in Table (3.1), where the heat of combustion value is the same as the HHV. To consider the formation of ash in the combustion process, it should be defined in Aspen Plus. The ash is also defined as a non-conventional component; its attributes are defined as 100% ash. The enthalpy and density of the biomass are calculated via "HCOALGEN" and "DCOALIGT" models in Aspen Plus.

The property method chosen for the simulations is based on IDEAL but concerns other methods because the models require different property methods for certain simulation parts and, therefore, need to be referenced. The CCS processes were modelled separately and then moved to HIERARCHY blocks, which were connected to other parts of the CHP plant.[18],[19],[20],[1]

3.2.2 Fuel decomposition

In the Aspen Plus simulations, the fuel must be decomposed in a separate unit before it is used for combustion. The RYIELD reactor decomposes the fuel depending on the yield specified during the input and does not require stoichiometry or kinetics of the reaction. It is worth mentioning that the fuel decomposition is an endothermic process and the heat for the reaction is derived from the combustion chamber which is modelled using an RGIBBS reactor. The method of fuel decomposition is reproduced according to the method described in the Aspen plus - Getting started modelling processes with solids [21]. This is the same method that is used for all the simulation cases.

The fuel combustion parameters differ in each simulation even though they are simulated by the use of an RGIBBS reactor and will be discussed further in the following sections.

3.2.3 Combined Heat and Power Setup

The turbine and DH system were modelled referencing the CHEMCAD simulation modelled by [22]). A simplified version of the Heat Recovery Steam Generator (HRSG) (Figure 3.1) depicts how heat is extracted from the flue gas and is used for DH and the steam turbine. The heat from the flue gas is extracted via the superheater and FG condenser which are both modelled as heat exchangers in the simulation. The steam turbine is fed steam that is pressurized to 107 bars by pump 1 and is heated up to 540°C via the superheater.

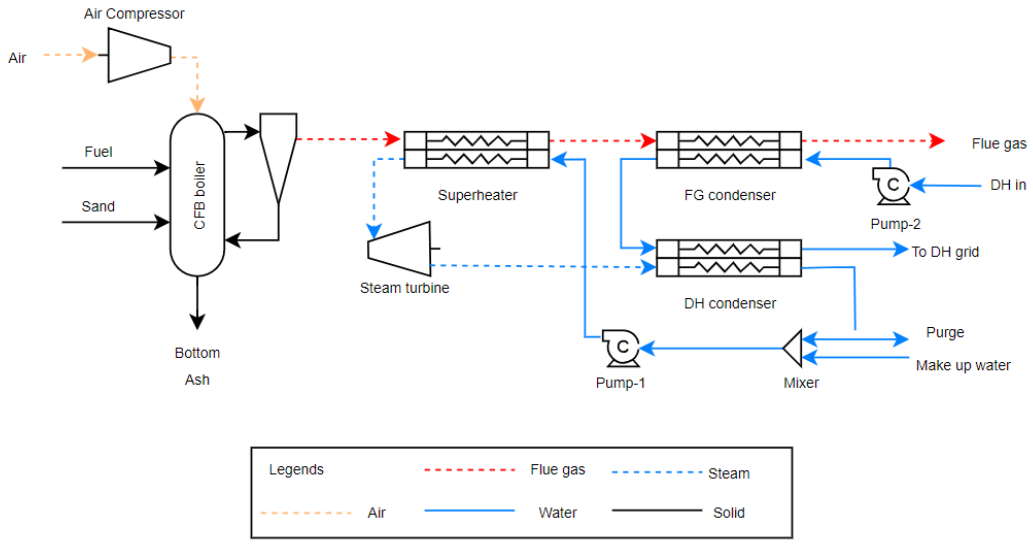


Figure 3.1: Flow sheet of the HRSG

The amount of heat extracted by the superheater and the Flue gas condenser depends on the flue gas characteristics, namely flow rate and temperature. The DH

water enters through pump 2 at 44°C and 1 bar pressure where it is pressurized to 16 bar pressure. It then goes through the FG condenser and DH condenser where it is heated to 93°C and is then sent for DH purposes.

A cooler is used to function as a heat consumer to denote the flow of heat from the stream to the DH system. The DH condenser, modelled using a heat exchanger, extracts heat from the stream exiting the Steam turbine and transfers it to the district heating water flow.

The mass flow into the DH system and steam turbine vary from system to system but the parameters listed in table (3.3) have been modelled to be constant for all of the simulation models. Depending on the working of the different CCS systems, the HRSG systems and their components have been subjected to change which will be further discussed in the forthcoming sections.

Table 3.3: Common parameters for HRSG system

Plant Data	**Value**
Admission Steam Temperature	540°C
Admission Steam Pressure	107 bar
Turbine Isentropic Efficiency	0.82
Pump Isentropic Efficiency	0.85
District Heating Supply Temperature	93.3°C
District Heating Return Temperature	44°C

3.3 Carbon Capture models

3.3.1 Post combustion

Dried biomass of the aforementioned characteristics enters the combustion chamber, which is modelled using an RYIELD reactor for decomposition and an RGIBBS reactor for combustion. An air compressor directs atmospheric air consisting of 21% O_2 and 79% N_2 into the RGibbs reactor. The amount of Air entering the RGibbs reactor is set to 3 Kg/s and is determined by the CO/CO_2 ratio (mole basis) in the flue gas exiting the combustion chamber. An acceptable ratio would be below 15% [1] and the ratio in our model is set to 5%. The RGibbs reactor is set to a pressure of 1 bar. The combustion products then enter a cyclone where the flue gas is separated from the ash and other solid particles. This hot flue gas then enters the HRSG where heat extraction occurs via the components mentioned in section HRSG.

The post-combustion system is retrofitted after the HRSG and therefore the HRSG system does not require any additional components or changes. A simplified diagram of the system is represented by figure 3.2. This system is installed after the Turbine

and HRSG of the CHP plant where all the possible heat and work from the process has already been extracted. The flue gas is also treated for Sulphur oxides before it enters the MEA absorption plant. The flue gas characteristics entering the post-combustion carbon capture plant are described in table (3.4).

Table 3.4: Flue gas characteristics (Post-combustion)

Mass Flow	35.06 Kg/s
Temperature	46°C
Pressure	1 bar
Component (mol%)	
CO	0.01
CO_2	0.16
O_2	0.01
N_2	0.69
H_2O	0.14
H_2	0.00

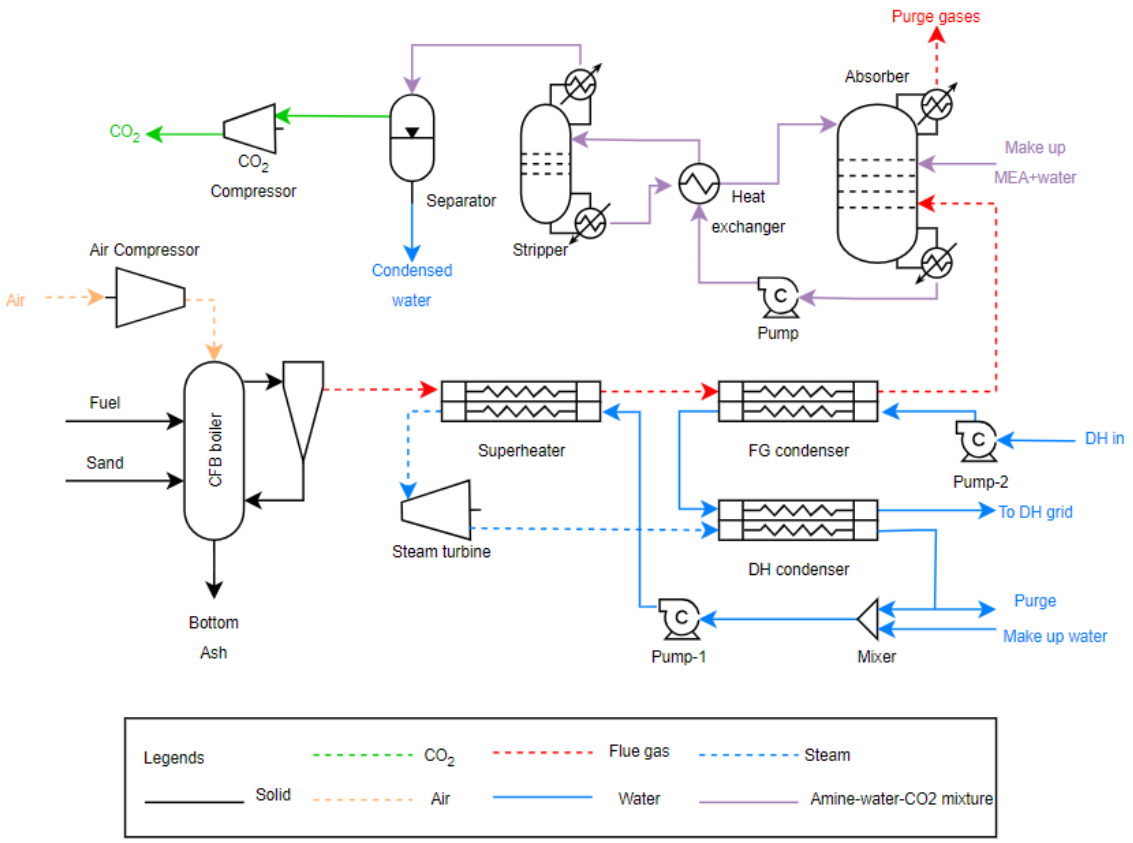


Figure 3.2: Post-combustion process flowsheet

The flue gas cleaned of impurities such as particulate matter and Sulphur dioxides is directed into the Absorber which is a packed column in the Aspen model where it

is exposed to the MEA solution or Lean Amine solution. For the MEA absorption simulation, the ENRTL – RK method is used with the electrolyte calculation options set to MEA-CHEM. A 30% loaded MEA solution is circulated from the top of the tower at a flow rate of 208Kg/s and the flue gas flows from the bottom with a flow rate of 35.06Kg/s, up the Absorber column facilitating the absorption of CO_2. The characteristics of the Amine solution are listed in the table (3.5)

Table 3.5: Lean Amine Characteristics

Mass Flow	208.4 Kg/s
Temperature	40°C
Pressure	1 bar
Component (mol%)	
MEA	0.295
H2O	0.635
CO2	0.065

The now clean flue gas flows out of the absorber and is led into a stripper another packed column, as the name suggests, the CO_2 is stripped from the amine solution. The regeneration of the MEA solution is facilitated using a reboiler where the Rich amine solution (RMEA) is now heated to 123°C. The separated stream of CO_2 can be obtained from the condenser where other solvents are removed. Before the stripper section, a heat exchanger is used to heat the regenerated Amine solution to 110°C. Additionally, the RMEA solution is pressurized to 1.8 bars before it enters the stripper [1]. The reaction of Absorption and desorption is written below and is the same reaction used in the Aspen Plus model.

$$2H_2O \rightarrow H_3O^+ + OH^- \tag{3.2}$$

$$MEA^{\equiv} + H_2O \rightarrow H_3O^+ + MEA \tag{3.3}$$

$$HCO_3{}^- \rightarrow H_2O + H_3O^+ + CO_2 \tag{3.4}$$

$$CO_2 + MEA + H_2O \rightarrow MEACOO^- + H_3O^+ \tag{3.5}$$

$$CO_2 + OH^- \rightarrow HCO_3{}^- \tag{3.6}$$

The vapour from the stripper is directed into a condenser of temperature 40°C. The regenerated Amine solution flows through the heat exchanger and is mixed with the condensate from the condenser and subsequently cooled to acceptable levels before the new solvent is added, and old solvent is purged, and the cycle continues. In this thesis, the amount of solvent required for the process is determined through the sizing of the model absorber described in (the MEA absorption booklet) and the same procedure is followed and the estimated value for the same was obtained from

the work of Dawid P. Hanak et. Al.[13]. The aim was to obtain a capture efficiency of around 90%. The reboiler duty, condenser duty and ultimately cooling duty were obtained from previous work performed by Dawid P Hanak as well. These values have been tabulated in Table (3.6). The CO_2 rich stream is then fed to the CO_2 compression unit at 45°C and 1 bar pressure.

Table 3.6: Components with their corresponding duty

Component	**Duty (MW)**
Stripper Pump	0.018
Reboiler duty	39.2
Condenser duty	16
Cooler duty	15

3.3.2 Pre-combustion

3.3.2.1 Validation of model

The first step in the ASPEN PLUS model is to determine which method is to be used and we decided on COMMON/IDEAL as our base scenario. It all begins with our standalone Air Separation Unit (ASU), which plays a crucial role in the process. The ASU's primary function is to separate air into its key components, primarily nitrogen and oxygen. In this validated model, the ASU section consists of an air compressor, the main ASU unit, and a multistage compressor for regulating the temperature and pressure conditions of oxygen. The separated oxygen, at a pressure of 36 bar, temperature of 25°C and purity of 95% by volume, is directed to the gasifier section, a pivotal component of our operation. The flow rate of oxygen is decided based on the research paper in use and the pressure is decided as a factor of 1.2 times [18] the gasifier pressure.

Within the gasifier, coal enters the RYIELD and RGIBBS components, where the gasification process happens at a high temperature of 1300°C and pressure of 30 bars. This stage initiates the conversion process and generates excess heat, which we carefully harness for subsequent steps in the HRSG section.

The resultant product of the gasifier is syngas, a mixture of gases with carbon monoxide (CO) and hydrogen (H_2) as the primary components. This syngas is initially hot and under pressure, setting the stage for the subsequent phases.

Moving on to the Water-Gas Shift (WGS) and Heat Recovery Steam Generation (HRSG) section, we have a complex setup. Here, the syngas undergoes a series of transformations. To begin, the syngas are efficiently cooled to 350°C in the heat recovery steam generation unit (HRSG) and then sent to the WGS-1 reactor, designed using an RSTOIC reactor, with preheated steam at the same conditions. Here, CO is partially converted to CO_2 by the water gas shift reaction. The exhaust

is then cooled to 178°C in the HRSG section and fed to the WGS-2 which operates at 178°C and 30 bars. The cumulative CO conversion efficiency in both WGS reactors is at 98%. The shift reactions inside the two WGS reactors are exothermic (ΔHrxn = 44.5 KJ/mol). At this point, the syngas is cooled to 40°C in the HRSG unit and fed into the Acid Gas Removal Unit (AGR), where 99.99% of H_2S and 94.8% CO_2 is removed.

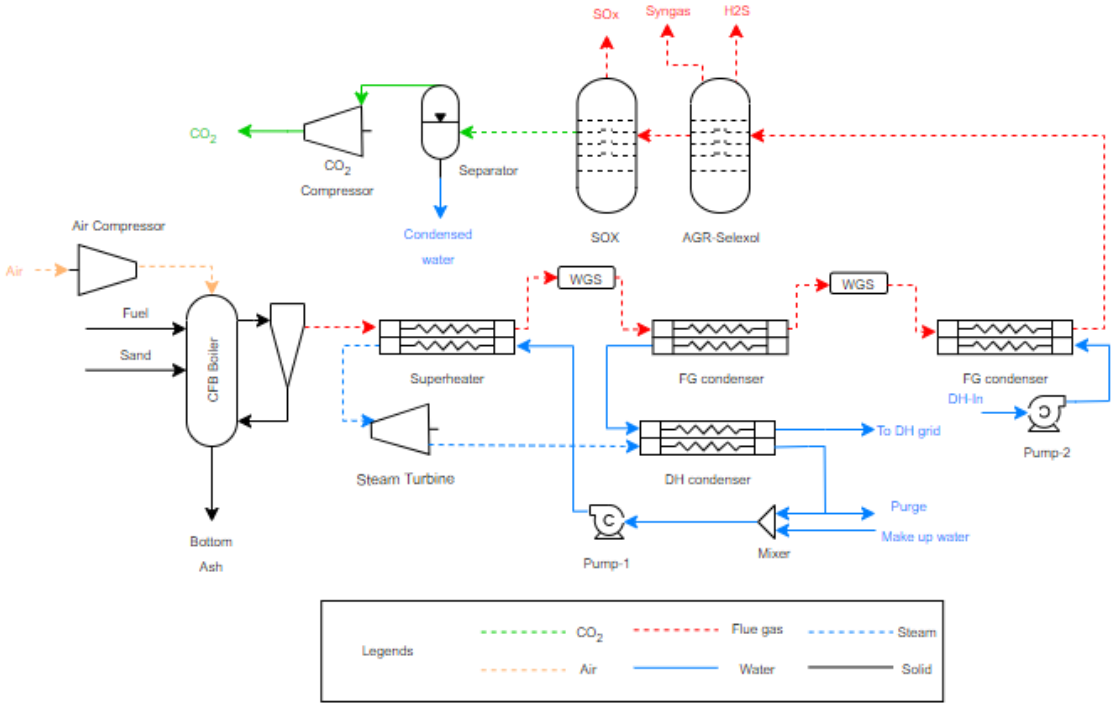


Figure 3.3: Pre-combustion process flowsheet

3.3.2.2 Model Modification and process integration

After the pre-combustion model had been validated, it was adjusted to fit the needs of our investigation. The flow rate was changed to 5.26Kg/s at 25°C, and the coal was swapped out for biomass. The ASU unit worked the same way however, the amount of oxygen separated and then used changed to 5.98 kg/sec, which was defined by using the DESIGN-SPEC manipulator and fixing the CO:CO_2 ratio of 1:3(based on moles).

In our model, we have decided to integrate the Water Gas Shift section with the Heart Recovery Steam Generation section. We have a superheater, a DH heater, an FG condenser and an FG condenser-WGS which play an active role in heat transfer and its recovery. A similar HRSG system was implemented in all our models, however, because of the presence of 2 WGS reactors and the need for cooling in the step between them, an extra heat exchanger, FG condenser-WGS, is introduced in this particular process. The superheater now cools down the syngas from 1300°C

to 350°C, which then enters the WGS-1. We decided that the water gas shift reaction would happen adiabatically so, the exhaust gas from the reactor has a higher temperature output of 463°C.

Another heat exchanger, the FG condenser 1, cools the exhaust that will enter the WGS-2. In the final step, of this section, the FG condenser cools down the exhaust of WGS-2 to 45°C and then sends it to the AGR. The syngas continues to the AGR unit now containing CO_2, H_2, and steam. Approximately 90% of the CO_2 and hydrogen sulfide (H_2S) 99.99% in the gas stream is removed from this unit using a Selexol-based technique. After this, the CO_2 is sent at the final compression and cooling stage, where it leaves at 40°C temperature and 110 bar pressure.

3.3.3 Oxyfuel combustion

3.3.3.1 Validation of model

The oxyfuel combustion process, unlike the traditional combustion process, uses a nearly pure stream of oxygen with carbon dioxide to facilitate the combustion of fuel. This removes the formation of Nitrogen oxides and facilitates the complete combustion and conversion of C to CO_2. The first step in modelling the oxyfuel process was to validate the model produced by Pei X. et. Al.[19] after reproducing the same.

Wet coal with a flow rate of 50Kg/s at 25°C with characteristics mentioned in Appendix 1.2 is dried by primary air (21% O2:79% N2), with a flow rate of 250Kg/s in an RSTOIC block. The moisture content is reduced from 10% to 2% during this process. A FLASH block is then used to separate the moisture from the dry coal. Now similar to the decomposition process described in section 3.2.2, the coal is decomposed into its constituent components by a RYIELD block and fed into the combustion chamber (RGIBBS reactor).

A mixture of O_2 and recycled flue gas is used as an oxidant entering the combustion chamber, the mixture enters at 350C and has a mass flow of 672.11Kg/s. The hot combustion products are then subsequently directed into a heater unit where they enter the heater at 1395°C and leave at 350°C, simulating heating consumption. The flue gas train is then fed into a solid separation unit, followed by a separator unit where the CO_2 and O_2 are separated and recycled. The recycle ratio for the flue gas is 83.6% and this results in 21%/79% O_2 to CO_2 ratio.The recycled gas is mixed with 100% pure oxygen with a flow rate of 93.77Kg/s at 350°C. The flue gas exiting the combustion chamber was validated against the results produced in Pei X. et al. work.[19]

3.3.3.2 Model Modification and process integration

Firstly, the fuel was changed to biomass and so was the fuel flow to 5.26 Kg/s. The drying using the RSTOIC unit was removed. The recycle ratio of flue gas has been maintained for the modified model. However, the flow rate and temperature of recycled flue gas as well as the amount of pure oxygen being mixed with the same has been changed. A 95% pure stream mixed with the recycled flue gas consisting of CO_2 and O2 with a flow rate of 6.88Kg/s. The mixture of air enters the combustion chamber at 45°C instead of 350°C. This is because all the possible heat from the flue gas is extracted via the HRSG before it is recycled.

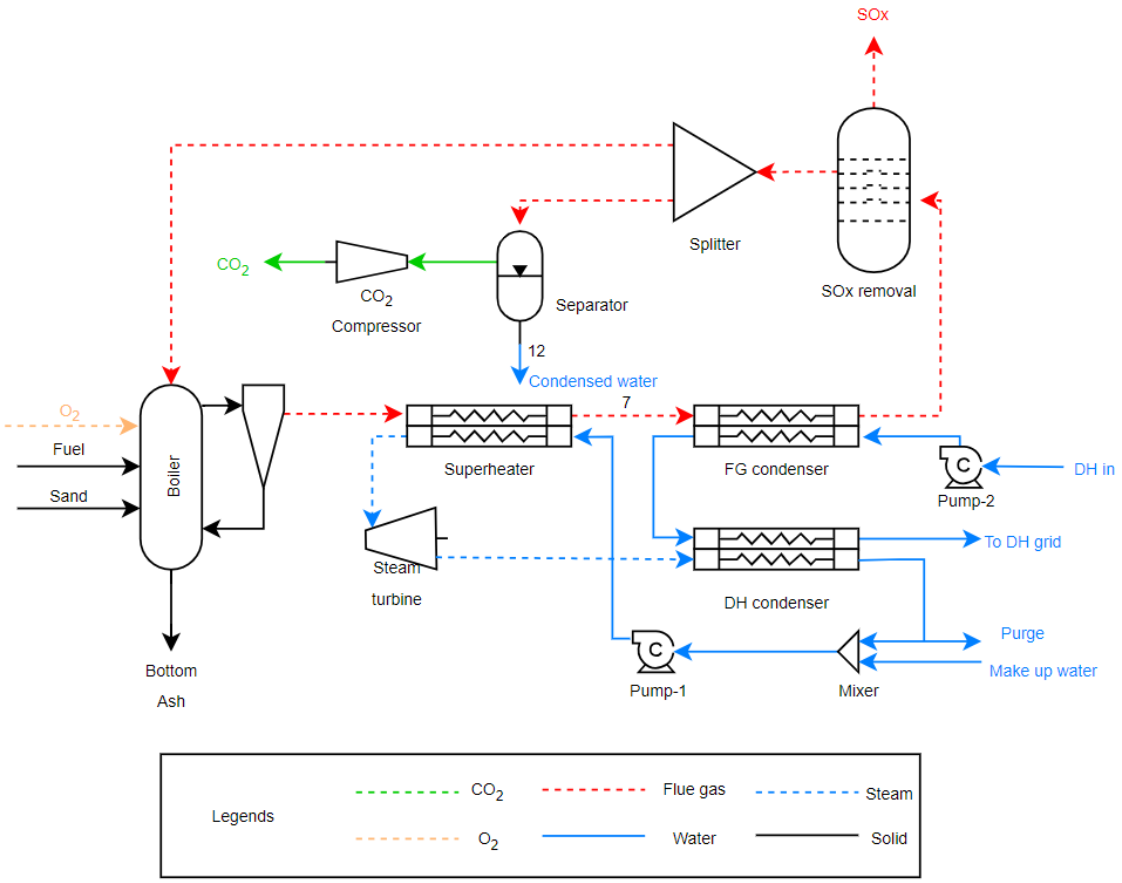


Figure 3.4: Oxyfuel combustion process flowsheet

After the combustion process, the flue gas is directed into the HRSG unit where useful heat is extracted from the flue gas train. The HRSG layout does not require any additional changes in terms of components and is similar to the system described in section HRSG. The flue gas enters the HRSG at 1316 °C and leaves it at 45°C. The flue gas characteristics entering the HRSG have been tabulated in Table (3.7). The flue gas is first treated for Sulphur oxides before it is recycled and compressed. The recycle ratio of 83.6% is maintained and is achieved by using a splitter.

The flue gas with a smaller flow rate is redirected for CO_2 compression at 45°C and 1 bar pressure. The rest of the flue gas is mixed with fresh oxygen of high purity

Table 3.7: Flue Gas Characteristics (Oxy-fuel combustion)

Mass Flow	56.94 Kg/s
Temperature	1316.66°C
Pressure	1 bar
Component (mol%)	
CO_2	0.88
O_2	0.00
H_2O	0.12

and is used for fuel oxidation. The 95% pure stream of Oxygen is obtained from an Air Separation Unit (ASU). The modelling of the ASU is considered out of scope; however, the energy requirements are taken into account, and the value for the same is taken from [18]. As shown in the simplified figure 3.4, all of the flue gas exiting the combustion chamber is sent into the HRSG for heat extraction.

3.3.4 Chemical Looping Combustion

3.3.4.1 Validation of results

The Chemical looping combustion process model is based on using two interconnected fluidized beds modelled using two RGIBBS reactors and a solid oxygen carrier that provides the oxygen for fuel combustion. The model described in the work of Olivier Authier et. Al. [20] was first reproduced and validated before being modified for our study. The referenced study aimed to investigate Chemical Looping heat integration for electricity generation with steam cycle in the case of a conceptual baseload 250 MWe coal power plant. The power plant being simulated in the work has four main sub-systems: The CLC system, The AR heat recovery, the FR heat recovery, the FR treatment, and CO_2 compression. The CLC system is the only part of the simulation we validated, as the rest was irrelevant to our study.

The fuel used in the model is pulverized coal. The properties and composition of the fuel are tabulated in Appendix 1.2. The oxygen carrier was Mn3O4 and supported by $MgAl_2O_4$ (70:30 mass %). The AR used the Soave-Redlich-Kwong model for gas-solid chemical reactions modelling, and FR and NBS tables were used for water properties. The components used have been tabulated in Appendix 1.2. The Fuel reactor was fed with pulverized coal at 81 t/h at 1.01 bar pressure and 20°C without drying. The pulverized coal was fed into an RYIELD reactor, decomposed into its components, and then directed into an RGIBBS reactor, where the fuel oxidation occurred. The RYIELD and RGIBBS reactors make the fuel reactor, but all the flows except fuel directly enter the RGIBBS unit. A pure stream of Oxygen at 14.4 t/h at 1.05 bar and steam at 60t/h at 550°C was directed into the FR. The oxygen carrier (Mn_3O_4 supported with $MgAl_2O_4$) was supplied to the fuel reactor at a rate of 11736 t/h at 1 bar pressure. The hot combustion products were then directed to a cyclone (FR cyclone), where the Solids (Ash and reduced OC) were separated from the flue gas. The reduced oxygen carrier (Mn_3O_4 and MnO mixture) was

re-oxidized in the AR with air (79.4 N_2 and 20.6 O_2 volume basis) at an 828 t/h flow rate at 186°C.

The AR operating temperature was set to 1000°C. The fresh oxygen carrier (Mn_3O_4) was then led into another cyclone (AR cyclone) to separate the freshly oxidized OC from the ash and oxygen-depleted air. The fresh oxygen carrier is then led into the Fuel reactor to complete the oxidation-reduction cycle. The results pertaining to the fraction of components present in the product streams were compared and were well within the range to say the model was validated.

3.3.4.2 Modifications and process integration

Once the validation was complete, the validated CLC model was then modified to meet our study requirements. The pulverized coal was replaced by biomass (table fuel) and the flow rate was changed to 5.26Kg/s at 25°C. The Oxygen carrier was then switched to Ilmenite (table 3.2) and the flow rate was set to 147Kg/s to maintain a CO/CO_2 conversion ratio below 5% mole basis[23]. The stream of high-purity oxygen and steam was removed as it was not relevant to our study. The reduced OC from the FR cyclone is then directed to the AR operating where pre-heated air flows in at 85Kg/s. The airflow is set to 85Kg/s to make the AR function adiabatic. Another stream of fresh Oxygen carriers, 5% of the mass of the existing OC, also enters the AR to replace some lost Oxygen carriers. Figure 3.5 represents a simplified version of the process. The flue gas and oxygen-depleted air characteristics exiting the CLC are tabulated below.

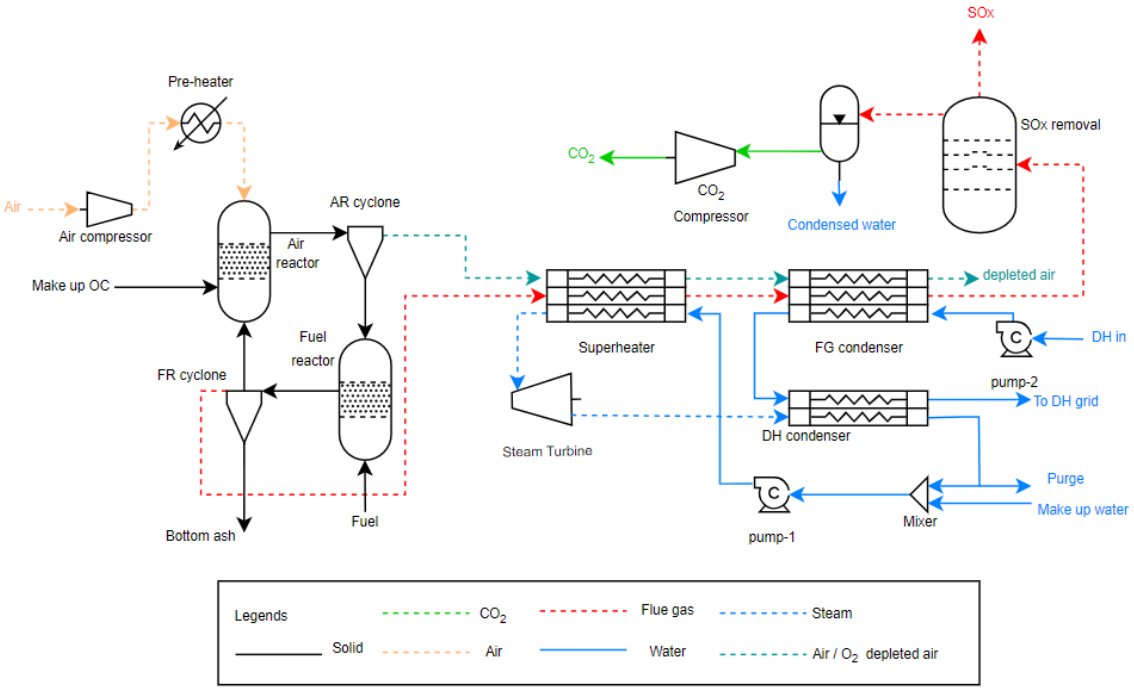


Figure 3.5: Chemical Looping Combustion flowsheet

Table 3.8: Exit Gas Characteristics for CLC

	Flue Gas	**Depleted Air**
Mass Flow (Kg/s)	11.65	77.54
Temperature (°C)	959	1000
Pressure (bar)	1.01	1
Component (mol%)		
H2O	0.436	0.000
N2	0.004	0.853
O2	0.000	0.147
H2	0.013	0.000
CO2	0.523	0.000
CO	0.023	0.000

The flue gas and depleted air enter the HRSG, where their heat is extracted via the Superheater and FG condenser. The flue gas and the oxygen-depleted air transfers heat to the Superheater and Flue gas condenser. In the simulation, since a heat exchanger can only have two input and output streams on the hot and cold side, heat transfer from the streams is performed by placing heater units on the Flue gas train, depleted air train, DH water stream and Steam turbine inlet stream and heat is transferred between them via heat streams. The depleted air, mainly consisting of Nitrogen, is re-introduced to the atmosphere as it is no longer hot, and the flue gas is further treated for Sulphur oxides and then subsequently compressed. The CO_2 rich flue gas stream enters the compressor unit at 60°C and 1 bar pressure

3.3.5 Flue gas treatment

In all of the simulations, a flue gas treatment for Sulphur oxides, Nitrogen oxides and other particulate matter is required in order to successfully sequestrate an almost pure stream of CO_2 from the condenser. This is modelled using a separator unit in the simulation and the working of the realistic unit is considered to be out of scope. However, the energy, heat and cost requirements of these flue gas treatment processes are considered when performing the analysis.

3.3.6 CO_2 compression

The CO_2 compression of the treated flue gas is simulated by the use of a multi-stage compressor. A flash is used to separate water from the CO_2 stream before compression. The temperature, pressure and composition of the flue gas entering the CO_2 compressor depends on the flue gas treatment used but the pressure at which the flue gas is compressed is fixed to 110 bars at 40°C in all the simulation models.

3.4 Energy Analysis

The energy analysis for each process simulation was performed after the simulation results were obtained. The three main parameters that were taken into account for the energy analysis were the net Electrical efficiency, District heating efficiency and overall efficiency.

$$\text{Electrical efficiency } (\eta) = \frac{\text{Net Work done}}{\text{Energy input}}$$

The net work done by the process is considered to be a summation of all work consumed subtracted from the work produced by the process simulation during the operation. The generator efficiency is assumed to be 0.99.

$$\text{District heating efficiency} = \frac{\text{Net District heat}}{\text{Energy input}}$$

The net District heat is the summation of heat released and consumed during the operation of the process simulation. Finally, the overall or net thermal efficiency was calculated considering the total amount of valuable work obtained from the system, both as Electrical power and heat.

$$\text{Net Thermal Efficiency} = \frac{\text{Net district heat + Net electrical work done}}{\text{Energy input}}$$

3.5 Exergy Analysis

The energy destruction and energy efficiency of each component in every process simulation were calculated and tabulated. From Aspen Plus, the energy flow of streams entering and exiting the components was obtained. The energy flow for a heat stream was calculated using the formula [13] below.

$$\text{Ex} = Q\left(1 - \frac{T_0}{T}\right)$$

Where,
Ex is Exergy flow
Q is the heat input/output
T_O is the reference temperature
T is the temperature of the heat stream

In Aspen Plus, you can view the exergy flow of a particular stream by adjusting the settings to display it in the stream properties. Using this, the exergy flow entering or exiting a component or system is tabulated. The exergy flow of a stream entering a component is considered the input exergy and the exergy flow of the stream exiting the component is considered the output exergy. For component-wise exergy calculations, The difference in the exergy flow of the input stream and output stream is the exergy destruction. Similarly, the exergy efficiency of the component is calculated by output exergy divided by the input exergy times 100 (for percentage).

The chemical energy of stream flows was calculated according to [14]. This is specifically used in the Water Gas Shift reactor as the energy flows extracted from Aspen Plus do not account for the energy of the reaction.

The exergy efficiency of the overall system was calculated by considering the net exergy input and the net energy output of the plant including stream exergy flows, work done, work consumed, heat produced, and heat consumed.

The Exergy of fuel was also calculated using the formula [24]

$$\text{Ex} = (\text{LHV}) \cdot \left[1.0064 + 0.1519\frac{H}{C} + 0.0616\frac{O}{C} + 0.0429\frac{N}{C}\right]$$

3.6 Techno-Economic Analysis

Techno-economic analysis (TEA) is the methodology used for assessing the economic feasibility of technology. It determines the financial viability of a given technology or project by combining technological and economic evaluations.
The technical assessment for each plant consisted of the plant's performance characteristics. The parameters considered for these were based on the chosen research paper and are as follows

- Plant thermal efficiency
- Plant electric efficiency
- Plant electricity penalty

On the other hand, in the economic assessment of technologies, we considered and evaluated the following parameters:

- Power-plant retrofit cost
- Total investment cost
- Operating costs
- Revenue
- Storage costs
- Levelized cost of electricity

3.6.1 Technical Assessment

For our procedure, a set of equations and parameters were used which are being determined below,

$$TPO = \frac{SOE \cdot 1000}{OC} \tag{3.7}$$

$$OC = LF \times 365 \times 24 \tag{3.8}$$

where, TPO is typical plant output(MW), SOE is the send out electricity(GWh), OC is operating capacity(hr/yr), and LF is the load factor(assumed to be 90% in our study)

$$CCO_2 = CE \cdot ER_{CO_2} \cdot SOE \tag{3.9}$$

where, CCO_2 is captured CO_2 per year(Mt/yr), CE is capture efficiency, ER_{CO_2} is emission rate(kg/MWh).
Further, data, including the district heat generated and the electricity generated and used in each scenario, was taken from the ASPEN PLUS model.

3.6.2 Economic Assessment

The cost assessment model was taken from the paper(ref). In the first step, we looked at the fuel cost calculations, which can be summarized using the following equations,

$$TFU = \frac{FF \cdot LF \cdot 3.6}{10^6} \tag{3.10}$$

$$C_{TF} = FCV \cdot C_F \cdot TFU \tag{3.11}$$

where, TFU is the total fuel used per year(Mt), C_{TF} is the cost of total fuel used per year(M€), FF is the fuel flow(5.26kg/sec in all our cases), FCV is the fuel calorific value(20.54MJ/kg in all our cases) and, C_F is the fuel cost (€/GJ).
For the calculation of Total Plant Investment, the primary cost data were obtained from refs (a-z) and are summarized in table (3.9).

Table 3.9: Parameters for Total plant cost evaluation

Component	**Base Cost (M€)**	**Scaling Factory**	**Scaling unit**	**Reference Scale**
Combust/CFB Boiler[25]	251.15	0.74	MW	300
Air Compressor[26]	4.52	0.67	MW	10
Oxy Compressor[26]	6.04	0.67	MW	10
Air Separation unit[15]	31.5	0.67	O_2 flow kg/s	45.5
Heat Exchanger[25]	12.27	0.6	MW	138
Steam Turbine[25]	55.574	0.67	MW	275
WGS[26]	3.31	0.67	kg/s	30.14
AGR[26]	21.83	0.65	kg/s	66.67
CO_2 Compressor[26]	5.187	0.94	MW	10
Civil Works[15]	12.8	0.85	MW	300
Biomass Handling[15]	5.3	0.31	MW	157

Pump costs were calculated based on the amount of work done by the unit. The equation for this is as follows;

$$C_p = 1.11 \cdot W_p + 0.07 \tag{3.12}$$

where Cp is the cost of equipment(M€), and Wp is the work done by the pump (MW).
In the course of our evaluation, the selection of energy utilities for the presence in the economic analysis was an intentional choice on our part. We strategically chose particular utilities and calculated the related expenses by taking personal considerations into account. Interestingly, the electricity procurement costs for essential machinery like compressors and pumps were determined using standard prices at the

time in Sweden. This methodology guaranteed a pragmatic and situation-specific basis for our financial evaluations.
Our calculations are also included in the expenses related to selling district heating. This thorough examination of the revenue side—district heating sales—and the expenditure side—electricity purchases—contributed to a comprehensive assessment of the entire investment.
The equations for calculating these costs are as follows;

$$C_{DH} = Q_{DH} \cdot P_{DH} \tag{3.13}$$

$$C_{EU} = EU \cdot PE \tag{3.14}$$

where C_{DH} is the cost of district heat per year M€/yr, Q_{DH} is the district heat produced per year(MW), P_{DH} is the price at which district heating is sold which is 80.3 €/MWh, C_{EU} is the cost of electricity used per year M€/yr, EU is the electricity consumed per year(MW), PE is the price at which electricity is bought from the market which is 9.32 €/MWh

Levelized Cost of Electricity(LCOE): The final parameter that was calculated was the LCOE. For simplicity of the calculations, we assumed the net output of the plant, O&M costs and fuel costs are constant over the life of the plant. The following equations based on ref([27]) were used to perform all LCOE calculations;

$$LCOE = \left(\frac{(TCR \cdot FCF + O\&M)}{NPO}\right) + \left(\frac{C_S}{NPO \cdot OC}\right) + HR \cdot FC \tag{3.15}$$

$$CRF = \frac{r(1+r)^t}{(1+r)^t - 1} \tag{3.16}$$

$$HR = \frac{AFC \cdot FCV}{NPO \cdot OC} \tag{3.17}$$

where LCOE is the Levelized Cost of Electricity (€/MWh), TCR is the total capital requirement, also called total plant investment(M€), O&M are the operations and management cost annually and assumed to be 5% of the TCR(M€), FCF is the fixed charge factor, C_S is the cost of storage of CO2 which is assumed to be 72$/t of CO2, HR is the r is the rate of capital discharge and assumed to be 7%, t is the plant life which is assumed to be 25 years.
Thereafter, LCOE is used to calculate the payback period. The payback period is an essential monetary measure used for evaluating investment risk and liquidity. By calculating it, firms with limited resources can make better decisions by understanding how quickly an investment may recover its initial cost. In general, people appreciate shorter payback periods as less risky because they provide a faster return on investment and guarantee consistent cash flow. Because of its simplicity, people and small businesses with immediate financial goals may think of utilizing it.

$$PBP = \frac{TCR}{Pr} \tag{3.18}$$

$$P_r = (C_{DH} + C_{ES}) - (C_{O\&M} + FC + C_{EU}) \tag{3.19}$$

where PBP is the payback period in years, TCR is the total capital requirement in M€, P_r is the annual profits in M€, C_{DH} is the cost earned by selling district heating in M€, C_{ES} is cost of sold electricity in M€, $C_{O\&M}$ is the operations and management cost in M€, C_F is the fuel cost per year in M€, C_{EU} is the cost of electricity used per year M€/yr.

4

Results and Discussion

4.1 CO_2 Capture efficiency

The CO_2 capture efficiency of each process is mentioned in the table (4.1). The highest Carbon capture efficiency was observed for the CLC process, which can be explained by the efficient conversion of C to CO_2 via the Oxygen Carrier and the absence of Nitrogen oxides in the flue gas. In the simulation model, the rate of Oxygen Carrier was fixed at 147Kg/s to maintain a CO/CO_2 conversion rate of 5%. By increasing the amount of OC, the amount of CO_2 converted would be increased, but it would be marginal.

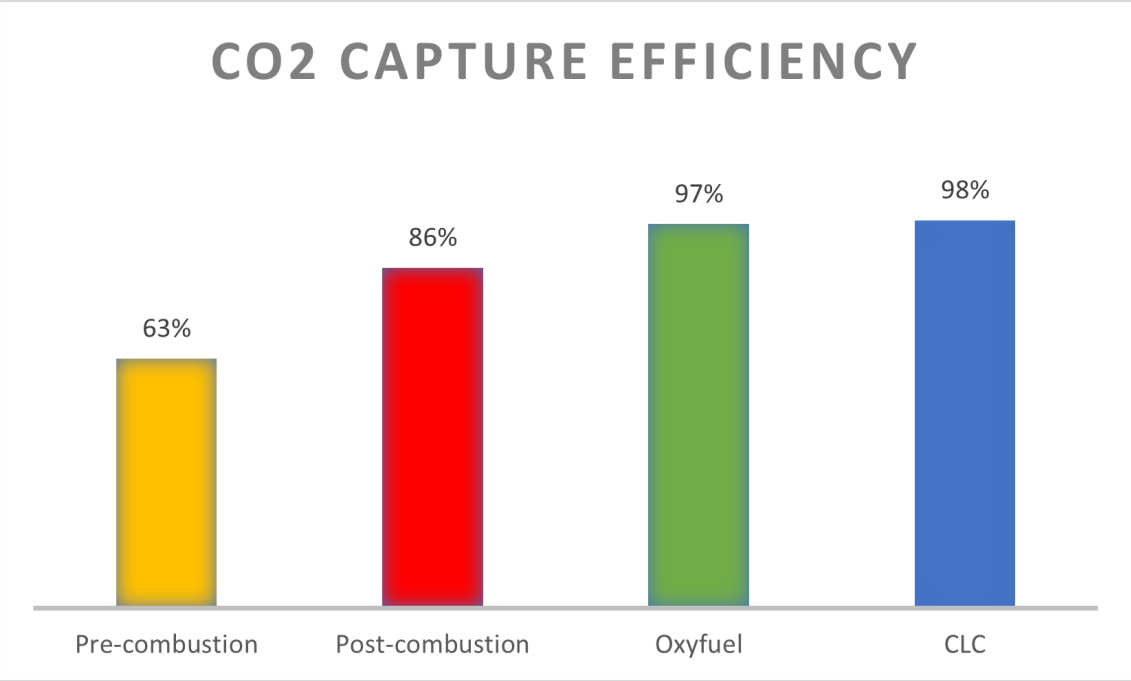


Figure 4.1: CO_2 capture efficiency

The Oxyfuel combustion process has the second highest CO_2 capture efficiency as the presence of Oxygen rich air facilitates fuel combustion without the formation of Nitrogen oxides. High combustion temperatures in an oxygen-rich atmosphere led to fewer incomplete combustion products. The CO_2 capture efficiency of the post-combustion plant is limited to the Absorption solvent used. The Lean solvent loading

is kept at 30% because (reference). In our simulation model, the lean solvent flow was set to 220Kg/s of LMEA to achieve a capture efficiency of 90% in the absorber. An efficiency of 95% can be achieved by changing the amount of Lean solvent flow. However, the amount of CO_2 obtained from the stripper section does not correspond to the amount of CO_2 absorbed by the Absorber. Complete regeneration of MEA in the stripper does not occur; therefore, some CO_2 remains in the solvent. Thus explaining the resultant capture efficiency of 86%.

The precombustion plant has a better Carbon Capture efficiency than the post-combustion plant. The WGS in the precombustion plant has a high conversion efficiency. However, the Selexol process used to remove CO_2 from the syngas mixture limits the amount of CO_2 captured.

4.2 Energy Analysis

The row titled "Auxiliary Power Consumption" is the total power consumed by those components that use electricity to function. Net electricity denotes the total amount of electricity that can be sold to the market and net heat denotes the amount of heat being sent for District heating purposes. The net energy denotes the total amount of energy produced by each Power plant (Net electricity + Net Heat) after subtracting corresponding energy penalties (Auxiliary Power consumption and Heat penalties). The key values being compared in this section are the overall efficiency, Net electrical efficiency, and Net energy efficiency. The flow rates of steam and water entering the district heating system and the steam turbine for the different models have been tabulated below. These flow rates determine the amount of district heat and power produced.

Table 4.1: Flowrates in HRSG and Steam turbine

	Flow rate of water (Kg/s)	
Models	**District heating**	**Steam turbine**
Pre-combustion	281.93	17.45
Post-combustion	300.15	22.4
Oxyfuel combustion	319.34	23.81
CLC	359.68	24

The overall efficiency of the oxyfuel plant is the highest when compared to Pre-combustion, Post-combustion and CLC plants. A high rate of CO_2 conversion facilitated by the use of a CO_2 and O_2 mixture results in a flue gas stream with high energy flow which is efficiently made use of in the HRSG. However, gas streams from the CLC plant have higher energy flows when compared to the other plants. The CLC plant generates the most net electricity at 19.20MW but supplies less net Heat compared to the Oxyfuel model. This can be attributed to the large amount of heat consumed by the Air Pre-heater to re-oxidize the reduced Oxygen carrier in

Table 4.2: Energy Distribution and Efficiency

Parameter/Plant	Different Simulation Models			
	Pre-combustion	Post-combustion	Oxyfuel	CLC
Energy input (MW)	100.00	100.00	100.00	100.00
Pump-1 (MW)	0.25	0.32	0.34	0.35
Pump-2 (MW)	0.51	0.54	0.57	0.65
Oxy compressor (MW)	2.10	-	-	-
Air compressor (MW)	-	0.16	-	0.33
CO_2 compressor (MW)	0.75	2.74	3.30	3.64
MEA System (MW)	-	8.22	-	-
AGR (MW)	1.63	-	-	-
SO_x removal (MW)	3.01	3.21	3.41	3.85
Air pre-heater (MW)	-	-	-	13.52
ASU (MW)	4.65	-	5.51	-
Auxiliary Power Consumption (MW)	9.89	3.76	9.72	4.97
Steam turbine work (MW)	17.45	22.40	23.81	24.17
District heat (MW)	60.29	64.18	68.29	76.91
Net Heat (MW)	57.28	52.75	64.88	59.54
Net electricity (MW)	7.56	18.63	14.09	19.20
Net energy (MW)	64.84	71.39	78.97	78.74
Net electric efficiency (%)	7%	18%	14%	19%
Net energy efficiency (%)	65%	71%	79%	79%
CO_2 capture efficiency	63%	86%	97%	98%

the Air reactor. By reducing the temperature of the Pre-heated air entering the Air reactor, the energy consumption of the Pre-heater can be reduced, thus increasing the overall efficiency of the cycle. Even though the air compressor's work would increase, the energy penalty would still be comparably smaller. Furthermore, this would increase the amount of heated depleted air entering the HRSG, thus increasing the heat that can be transferred to the Steam entering the Steam turbine and the DH system. The CLC plant is more attractive to investors when it is not a part

of a CHP setup, despite having a relatively smaller net energy production. This is because generating more net electricity leads to increased profits in the form of sellable electricity.

The Pre-combustion plant has the lowest overall efficiency of 65%, which can be attributed to the high net energy consumption. It is worth noting that the Syngas separated from the process can be used for energy production, which is not simulated in our Aspen Plus model. Thus, the results regarding the power generation and overall energy production for the Pre-combustion simulation are subject to change. Another process that is unique to the Pre-combustion process is the AGR process which is necessary for the separation of H_2 and CO_2 from the gas mixture. However, The CO_2 compressor for the Pre-combustion plant requires the least amount of power for operation when compared to the other models. This is because CO_2 enters the compressor at 30 bars. Whereas, The most amount of electricity consumed for CO_2 compression is for the CLC plant which is attributed to the higher CO_2 flow rate. The model with the second highest power consumption is Oxyfuel combustion, which can be attributed to the high amount of power consumed by the Air Separation unit. We assume that the Oxygen entering the combustion chamber for the Oxyfuel process is already pressurized to a sufficient extent and therefore does not require a separate compression step like Pre-combustion. The post-combustion process, as expected, has a significant energy penalty associated with the operation of the Reboiler in the Solvent-Stripping. The Reboiler duty was integrated with the condenser and cooler duty to reduce overall heat duty. The energy penalty for the operation of the Reboiler is derived from the District heat produced. For all models, an energy penalty of 5% of the District Heating produced was assumed for the Desulphurisation process.

4.3 Exergy Analysis

The Exergy Analysis was performed for the overall plant i.e. considering all input and output streams entering the system as a whole and for each of the components in each model. The overall exergy efficiency is tabulated in Table (4.7) and the component-wise exergy destruction for each plant can be seen in Tables (4.3 - 4.6). To quantify the exergy of a system, we must specify both the system and its surroundings. It is assumed that the intensive properties of the surroundings remain unaffected by the processes. The reference state is a state where the system is in equilibrium with its surroundings. The reference temperature and pressure are 298.15 K and 101.3 KPa.The component-wise exergy destruction of the units of the CHP plants accounts for the amount of exergy destroyed in each component. The exergy destroyed in each component provides an understanding as to what components need further improvements to fully utilize the exergy supplied.

In all the models, the most exergy is destroyed in the combustion process, where entropy is produced. This is attributed to the chemical reactions and the irreversibility associated with the oxidation of fuel. The fuel oxidation cycle in the CLC cycle is more efficient than the Oxyfuel and Pre-combustion systems and has the same

Table 4.3: Comparison of Energy efficiency and Exergy efficiency

Plant	**Overall Energy Efficiency (%)**	**Overall Exergy Efficiency (%)**
Pre-combustion	65	28
Post-combustion	71	31
Oxyfuel combustion	79	33
CLC	79	39

exergy efficiency as the post-combustion simulation. The difference in efficiencies for the combustion process can be attributed to how efficiently the fuel is converted into the combust chamber. In the CLC oxidation-reduction cycle, the conversion of fuel is more efficient due to the use of an oxygen carrier. In a real-time set-up, a fluidized bed set-up would be used for the AR and FR which enables the thorough mixing of fuel and OC. The combustion parameters, namely, the Temperature and composition of the inlet air stream and, the temperature and pressure of the combustion chamber, play a vital role in the efficiency of the combustion process. The exergy efficiency of the combustion process when compared to the energy efficiency of the combustion process gives us a better understanding of the irreversibilities of the process and a more qualitative measure of how energy is utilised.

On further inspection of results, we see that the second most place of exergy destruction is in the Superheater. During the transfer of exergy from the hot stream and cold streams, exergy is destroyed in the component via heat loss. When the hot stream and cold stream mass flow rates are slower, a larger amount of heat is transferred and thus less exergy is destroyed. This can be observed on the FG condenser in the post-combustion process model. The pumps that supply water for District heating and the steam turbine have similar efficiencies and so does the steam turbine due to similar input conditions. The isentropic efficiencies of the pumps and turbines were the same across all the models and so was the temperature of water entering and exiting these units. A rather large difference in efficiency can be observed in the CO2 compressor between the Pre-combustion model and the other models. This is because the stream of CO2 that is being compressed in the pre-combustion model enters at 30 bars compared to 1 bar for the others. This results in a decrease in the CO2 compressor work thus increasing its efficiency.

Due to similar working conditions and efficiencies of the pumps and turbines in the HRSG of all the models, the exergy efficiencies are rather similar. The same could be said for the District heating condenser as well. The rate of flow into these components may differ but the temperature and pressure influence the exergy efficiency more.

The CLC plant has the best overall exergy efficiency among the other cases at 39% and has the least amount of exergy destroyed. The most amount of exergy destroyed is in the Pre-combustion model. This corresponds to the large number of units that

Table 4.4: Exergy Destruction in CLC

Component	**Exergy in (MW)**	**Exergy out (MW)**	**Exergy Destruction (MW)**	**Exergy Efficiency**
Combustor	109.92	60.78	49.14	55%
Air compressor	0.36	0.32	0.04	89%
Superheater	56.69	44.03	12.66	78%
FG condenser	3.20	2.02	1.18	63%
DH condenser	15.27	11.49	3.78	75%
Pump-1	0.49	0.17	0.32	35%
Pump-2	1.63	1.05	0.58	64%
Turbine	41.86	37.64	4.22	90%
CO_2 compressor	3.81	2.35	1.46	62%
Heater-2	5.00	3.04	1.96	61%
Total Exergy destruction			75.34	

Table 4.5: Exergy Destruction in Pre-combustion

Component	**Exergy in (MW)**	**Exergy out (MW)**	**Exergy Destruction (MW)**	**Exergy Efficiency**
Combustor	108.55	56.08	52.47	52%
Air compressor	2.51	2.14	0.38	85%
Superheater	17.07	10.43	6.64	61%
FG condenser	7.96	6.09	1.87	77%
DH condenser	11.49	8.99	2.50	78%
Pump-1	0.35	0.13	0.22	36%
Pump-2	1.27	0.80	0.47	63%
Turbine	30.22	27.18	3.04	90%
CO_2 compressor	2.20	2.04	0.16	93%
Water Heater	1.53	0.57	0.96	37%
Heat exchanger 3	5.13	3.66	1.47	71%
Heater-2	35.93	30.22	5.71	84%
WGS-1	6.66	6.61	0.05	99%
WGS-2	4.34	4.20	0.14	97%
Oxy-comp	2.57	1.67	0.90	65%
Total Exergy destruction			76.99	

the Pre-combustion model requires for operation. The exergy destroyed and created in the Selexol process, and the combustion of syngas was not taken into consideration so these results are subject to change. The power output after the use of the syngas was not taken into consideration in the overall exergy efficiency calculation either

Table 4.6: Exergy Destruction in Post-combustion

Component	**Exergy in (MW)**	**Exergy out (MW)**	**Exergy Destruction (MW)**	**Exergy Efficiency**
Combustor	107.01	59.38	47.63	55%
Air compressor	0.17	0.13	0.04	76%
Superheater	59.54	39.12	20.42	66%
FG condenser	1.19	1.16	0.03	97%
DH condenser	13.50	9.59	3.91	71%
Pump-1	0.45	0.16	0.29	36%
Pump-2	1.36	0.85	0.51	63%
Turbine	38.85	34.90	3.95	90%
CO2 compressor	2.74	1.84	0.90	67%
Total Exergy destruction			77.68	

Table 4.7: Exergy Destruction in Oxyfuel

Component	**Exergy in (MW)**	**Exergy out (MW)**	**Exergy Destruction (MW)**	**Exergy Efficiency**
Combustor	106.89	57.92	48.97	54%
Air compressor		0.00		
Superheater	58.09	41.60	16.49	72%
FG condenser	1.27	1.24	0.03	98%
DH condenser	14.36	10.21	4.15	71%
Pump-1	0.48	0.17	0.31	36%
Pump-2	1.44	0.91	0.53	63%
Turbine	41.23	37.08	4.15	90%
CO2 compressor	3.30	2.24	1.06	68%
Total Exergy destruction			75.69	

Table 4.8: Exergic efficiencies of different models

Plant	**Exergy in (MW)**	**Exergy out(MW)**	**Exergy destruction(MW)**	**Exergy efficiency**
CLC	116.60	45.19	71.40	39%
Pre-combustion	119.28	32.89	86.38	28%
Post-combustion	120.57	37.10	83.47	31%
Oxyfuel combustion	116.60	38.79	77.82	33%

but we suspect that there will not be a drastic change in overall exergy efficiency. It

is worth noting that in Table 4.6, the exergy efficiencies are calculated considering all exergy flows, some of which are not reflected in the exergy destruction tables. Namely, the AGR process in the Pre-combustion model, Heat duty corresponding to the MEA cycle in Post-combustion, and the ASU in the Oxyfuel and Pre-combustion models. The unit-wise exergy destruction of these units was not calculated, but their values had to be taken into consideration during the calculation of the overall exergy efficiency of the cycle for acceptable results.

4.4 Techno-economic Analysis

The table below summarizes the equipment requirements for four carbon capture technologies: post-combustion, pre-combustion, oxyfuel combustion, and chemical looping combustion (CLC).

Each technology has its own set of equipment demands, and some equipment is incompatible with others. Post Combustion and Oxyfuel Combustion have comparable equipment requirements, including CFB Boiler, ASU, Superheater, FG condenser, DH heater, Steam turbine, and CO2 Compression unit. WGS and AGR are required for pre-combustion, however not necessary for the others. CLC is distinguished by its specific equipment needs.

Table 4.9: Equipment costs in M€

Equipment	**Pre-combustion**	**Post-Combustion**	**Oxyfuel Combustion**	**CLC**
CFB	157.71	157.71	157.71	-
AR-FR	-	-	-	205.46
ASU	-	11.47	12.85	-
SUPERHEATER	12.84	11.06	13.32	13.44
FG CONDENSER	1.52	2.28	1.6	3.65
DH HEATER	10.68	9.19	11.1	11.17
FG COND-WGS	-	3	-	-
STEAM TURBINE	13.68	11.5	14.25	14.39
MEA-SECTION	26.68	-	-	-
WGS	-	4.78	-	-
AGR	-	9.41	-	-
CO2 COMPRESSION UNIT	2.22	0.63	2.54	2.75

The cost of a CFB boiler remains identical for pre-combustion, post-combustion, and oxyfuel combustion technologies because they all involve the same fuel-burning process. The CFB boiler typically accounts for approximately 65% of the entire plant cost. The unique feature of Chemical Looping Combustion (CLC) is that the fuel enters the air reactor-fuel reactor vessel, which is the most expensive component, accounting for approximately 75% of the entire plant cost. An Air Separation Unit (ASU) is required to supply pure oxygen for the processes of pre-combustion and oxyfuel combustion. Components such as the superheater, DH heater, and FG

condenser are common to all situations and have constant pricing under the Heat Recovery Steam Generator (HRSG) segment. Additionally, there is an extra heat exchanger, FG-COND-WGS, which is necessary for transferring heat between two stages of water gas shift reactors. The steam turbine costs are similar across all models, reflecting the similarity in power production among them. The MEA section is exclusive to the post-combustion process, while WGS and AGR are specifically integrated into the pre-combustion setup, contributing to the overall cost. It's noteworthy that the CO2 compressor is particularly intriguing due to the substantial cost difference in the pre-combustion scenario. This discrepancy arises from the fact that the entire pre-combustion process takes place at a pressure of 30 bars, resulting in the CO2 obtained at the end already being at a higher pressure compared to the other cases.

Table 4.2 clearly shows that Chemical Looping Combustion (CLC) is the carbon capture system that excels in both power production and district heating, with Oxyfuel combustion trailing closely behind. Pre Combustion, on the other hand, looks to be less efficient, generating lower heat and electricity outputs.
When we look at net power output, there is an interesting sway. CLC continues to lead the pack, whereas Post Combustion gets second place, with Oxyfuel and Pre Combustion trailing. This variance is mostly due to the significant power needs of the air separation unit in Pre and Oxy combustion, which has a negative effect on overall power production. When considering the overall plant costs, Oxyfuel and Post Combustion come out as the most affordable technologies to implement, boasting lower initial investment requirements. CLC, on the other hand, is the most expensive of the bunch, demanding a substantially greater initial expenditure.
Let us now come to the heart of determining cost-effectiveness. Despite its higher starting expenses, CLC comes out as the most cost-effective technology overall as reflected in its lowest Levelized Cost of Electricity (LCOE). Its improved power and district heating performance more than makeup for the original outlay over its entire life. Pre Combustion, on the other hand, has the highest Levelized Cost of Electricity (LCOE) because of its low energy output, making it a less cost-effective option over the long run.

Table 4.10: Main results and parameters for the Techno-economic analysis

	PARAMETER		**Post combustion**	**Pre combustion**	**Oxyfuel**	**CLC**
W_{ST}	**Turbine Power**	**MW**	22.40	17.45	23.81	24.17
DH	**District Heat**	**MW**	64.18	60.29	68.29	76.91
NPO	**Net Power Output**	**MW**	18.63	11.35	14.09	19.20
TCR	**Total Capital Requirement**	**M€**	241.60	242.42	229.36	267.06
LCOE	**Levelized Cost Of Electricity**	**€/MWh**	142.68	237.58	182.35	139.94

5

Further Considerations

Exergoeconomic analysis is an evaluation method for energy conversion systems that focuses on the cost rates of exergy streams and exergy destructions. Its primary objective is to assess the inefficiencies and costs related to these inefficiencies, as well as the investment expenditures required to reduce them. This is accomplished by creating cost balance equations for each component of the system. The analysis provides an exergoeconomic factor that aids in determining whether it is more beneficial to enhance the component's efficiency or reduce investment costs.

Assessing and adjusting the cost-effectiveness of each component can improve processes by revealing major cost sources. This also provides an opportunity to reallocate resources and come up with new and improved configurations that decrease the costs associated with the exergy streams and their relevant components. Comparing the four simulated models based on exergoeconomic analysis would translate to comparing the cost equations for the components that are common between them as well as the processes unique to them. Comparing the cost equations and specific cost of the common components in terms of exergy would help us draw precise and quantitative conclusions regarding which of the four models best integrates the components used. By comparing the processes and components unique to each model, the specific costs of those processes and thus ways to improve them can be devised.

Carbon Capture and Storage (CCS) technologies are becoming increasingly important in addressing climate change. Chemical looping combustion (CLC) is a promising CCS technology that has the potential to play a significant role in reducing greenhouse gas emissions. An exergoeconomic analysis, which investigates the performance of CCS technologies based on their exergy and economics, could provide valuable insights for investing in and implementing viable solutions.

Additionally, we may broaden the analysis by including life cycle considerations, thus deepening the scope beyond the operational phases. This may also incorporate ecological factors and impacts throughout the plant life. Through in-depth analyses, we can identify inefficient hotspots and hidden environmental impacts, offering a broad grasp of various energy technologies. This all-encompassing method enhances decision-making by assuring that long-term sustainability is taken into account at all phases throughout the system's life.

Furthermore, there is an opportunity to investigate how renewable energy might be incorporated into our systems, thus providing a crucial path towards accelerating

the development of sustainable energy solutions. When optimizing these hybrid systems, important factors such as carbon footprint, financial viability, and energy storage could be taken into account. A detailed and thorough evaluation can yield important insights that will help drive the transition to a more sustainable and resilient energy sector.

6
Conclusion

This thesis evaluates the performance of a CHP plant of 100MW using Chemical looping combustion against three conventional CCS technologies: Pre-combustion, Post-combustion and Oxyfuel combustion. The comparison was done based on energy, exergy and techno-economics. The downstream flue gas cleaning compression and liquefaction have been simulated using Aspen Plus. The different CCS technologies would be integrated into the heat and power plant, producing DH and electricity.

The simulation and heat integration process was performed and applied for the specific input we calculated for an energy input of 100MW. The specifics for the HRSG were derived from the Örtoftaverket plant. A literature study was conducted to understand and simulate the different CCS technologies. The operating conditions for the different technologies were obtained, simulated and validated against the literature reviewed. The simulation models were then modified to fit our study and then compared. The comparison in terms of energy, exergy and techno-economics was performed according to the steps mentioned in the literature.
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The study shows that the CLC-capturing technology would be more cost-efficient than any of the other CCS processes. The CLC plant also has the best CO2 capture rate and overall energy and electrical efficiency. The oxyfuel process has the best overall efficiency, but there are very clear advantages that the CLC plant still possesses. From an investor's perspective, the CLC plant would reap more benefits in terms of profits due to the increased electricity production. The most exergy is destroyed in the Combustion process for all the models, and the CLC plant has the best exergy efficiency.

Finally, the choice of carbon capture and combustion technology is critical in deciding a power plant's cost, effectiveness, and overall performance. Among the options taken into account, Chemical Looping Combustion (CLC) stands out as the clear winner, excelling in both power generation and district heating. CLC, despite its greater starting expenses, is the most cost-effective in the long run. Following closely behind is oxyfuel combustion. Pre Combustion falls behind because of lower energy output. Post-combustion is the cheapest option when looking at the initial investment as it can be retrofitted in the already existing plant without any extreme modifications. This study emphasizes the importance of technology selection in obtaining the best mix of performance and affordability in carbon capture and

electricity-generating projects.

This case study helps to have an understanding of the upscaling of the CLC technology and the high competitiveness that CLC has as a carbon capture technology.

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