Development and analysis of two innovative physical and chemical CO_2 absorption systems to achieve a more effective CCU process

Mourad El Helou

A Thesis in

The Department

of

Chemical and Materials Engineering

Presented in Partial Fulfillment of the

Requirements for the Degree of Master of Applied Science

(Chemical Engineering) at Concordia University

Montréal, Québec, Canada

June 2024

Concordia University

School of Graduate Studies

This is to certify that the thesis prepared

By: **Mourad El Helou**

Entitled: **Development and analysis of two innovative physical and chemical CO_2 absorption systems to achieve a more effective CCU process**

and submitted in partial fulfillment of the requirements for the degree of

Master of Applied Science (Chemical Engineering)

complies with the regulations of this University and meets the accepted standards with respect to originality and quality.

Signed by the Final Examining Committee:

____________________________Chair
Dr. Patricia Comeau

____________________________Examiner
Dr. Ivan Kantor

____________________________Supervisor
Dr. Yaser Khojasteh-Salkuyeh

____________________________Co-Supervisor
Dr. Melanie Jane Hazlett

Approved by ____
Dr. Zhibin Ye, Graduate Program Director

June 2024 ____
Dr. Mourad Debbabi, Dean of the Gina Cody School of Engineering and Computer Science

Abstract

Development and analysis of two innovative physical and chemical CO_2 absorption systems to achieve a more effective CCU process

Mourad El Helou

Human activities have caused a continuous increase in global temperatures due to high global greenhouse gas emissions, especially CO_2. The Intergovernmental Panel on Climate Change stated that to reduce these emissions carbon capture and utilization (CCU) must be implemented, despite its energy-intensive nature. In this project, various innovative approaches have been developed to improve the performance of solvent based absorption systems and their integration with CCU. The reduction in energy is achieved by hydrogen injection into the CO_2 removal separators to act as a stripping agent. Nine configurations (4 physical, 5 chemical) are simulated and analyzed. Furthermore, the proposed approaches are tested for two scenarios: 1) CO_2 capture from the flue gas of a cement plant. 2) revamping the CO_2 capture unit of the synthesis gas stream of an existing ammonia plant. The results for physical absorption revealed that around 48%-77% reduction in the net energy consumption for flue gas and syngas are achievable respectively compared to the base case capture process and the traditional processes available in the market. As a result, the capture cost has decreased by 10%-25%, for flue gas and syngas respectively (\$80.92/\$29.43 from \$89.34/\$39 per tonne CO_2). In the chemical case, the proposed concept is integrated with the use of high-temperature heat pump. According to our results, a 13% improvement for flue gas and 5% for syngas in the coefficient of performance of the heat pump is achievable. The results of this project can be used to improve the overall performance of the CCU technologies and to be compared with other emerging technologies.

Acknowledgements

Throughout my academic endeavor, I have been incredibly fortunate to receive generous support and assistance from various individuals and institutions. I am particularly grateful to Professor Yaser Khojasteh-Salkuyeh, my supervisor, whose unwavering guidance and insightful feedback have been instrumental in shaping my thesis. Additionally, I extend my appreciation to Professor Melanie Jane Hazlett, my co-supervisor, for her consistent assistance throughout my research.

I am also deeply thankful for the collaboration and support of Marzieh Shokrollahi, Philippe Navarri, and my friends in CSTAR, whose teamwork and encouragement have fostered a conducive research environment.

On a personal note, I am indebted to my parents, Mr. Sobhi EL Helou and Mrs. Zeina Chaker, whose love and support have been my foundation throughout my Master's degree journey. My sister, Bouchra, has also been a source of constant inspiration, and I am grateful for her unwavering belief in me.

Finally, I express sincere gratitude to Concordia University and its staff for providing exceptional facilities, organizing enriching events, and offering impactful classes that have greatly contributed to my academic growth and the successful completion of this thesis.

Table of Contents

List of Figures

List of Tables

List of acronyms

GHG: Global greenhouse gas

IPCC: Intergovernmental Panel on Climate Change

CCS: Carbon capture and storage

CCU: Carbon capture and utilization

IGCC: Integrated gasification combined cycle

TSA: Temperature swing adsorption

ESA: Electrical swing adsorption

PSA: Pressure swing adsorption

VSA: Vacuum swing adsorption

MOF: Metal organic framework

CLC: Chemical looping combustion

CLP: Calcium looping process

MEA: Monoethanolamine

MDEA: Methyldiethanolamine

DEPG: Dimethyl ether of polyethylene glycol

DGA: Diglycolamine

HP: Heat pump

COP: Coefficient of performance

PC-SAFT: Perturbed Chain-Statistical Associating Fluid Theory

CEPCI: The Chemical Engineering Plant Cost Index

CAPEX: Capital expenditure

OPEX: Operational expenditure

TWSE: Total operating and maintenance wages, supervision and engineering expenses

CONV: Conventional

CHEM: Chemical

PHY: Physical

Chapter 1 Introduction

1.1 Background and motivation

Global greenhouse gas emissions (GHGs) caused by human activities ranging from unsustainable energy use, land use, and everyday human lifestyle, have caused a continuous increase in global temperatures across the decades, reaching on average 1.1°C above 1850-1900 temperatures in 2011-2020 [1]. The main contributor in the global mean temperature rise is CO_2 emissions represented by a 0.45 ± 0.18°C per 1000 $GtCO_2$ [2]. Approximately 78% of global GHG emissions come from the energy sector, industries, transport, and buildings sectors together and 22% from agriculture, forestry, and other land use [1].

In 2021 the Intergovernmental Panel on Climate Change (IPCC) stated that in order to meet our climate goal of limiting global warming to below 1.5°C, reaching net zero CO_2 emissions is required [1]. The IPCC defines global modelled pathways that provide data on limiting the warming temperatures to different levels. The most important one in this case is to limit warming to 1.5°C with no or limited overshoot or 2°C. Based on these pathways the year the temperature limit is reached is affected by the percentage of CO_2 emissions reductions. For example, reaching the 2°C limit by 2030 is estimated to require 22% reduction of CO_2 emissions; by 2050 this requirement is 73%. However, in order to reduce the temperature rise to the 1.5°C limit by 2050, 99% reduction in CO_2 emissions is required [1].

In these modelled pathways reaching the net zero CO_2 emissions require transitioning from fossil fuels without carbon capture and storage (CCS) to renewables or fossil fuels with CCS. However, CCS often faces major challenges. The cost of capturing, transporting and geological sequestration are the most financially challenging parts. Another key challenge is finding adequate underground storage spaces for large-scale and long-term CO_2 storage is the other challenge [3]. Therefore, alternatives to CCS have been proposed. CCU (carbon capture and utilization) is one alternative, where carbon capture is implemented in a process to produce other chemical products, such as the urea synthesis process where CO_2 capture is an integral part of the overall cycle [3]. Various CCU technologies can be implemented to achieve the optimal CO_2 reduction [4]. However, different challenges such as cost of capture and power demand can halt the implementation of these technologies in the industry.

1.2 Research Objective

The key objective of this study is to improve the performance and reduce the cost of the CO2 capture units, as one of the most expensive sections in the CCU processes. Deep analysis on the development and improvement of various chemical and physical absorption carbon capture processes is conducted by incorporating comprehensive process simulation results using Aspen Plus® software. The energy requirement of the processes in terms of CO_2 captured was determined for each design.

The economic analysis of the processes and its comparison with each other and available references is a crucial part of this study in order to identify the best design. The capture cost of CO_2 was evaluated with varying parameters, such as feedstock, solvent used, pressure and temperature. By comparing these prices, the optimal design in terms of power consumption and capture cost was determined.

1.3 Thesis Layout

Chapter 2 of this thesis presents a review of the current carbon capture technologies and the modifications that can improve the chemical and physical absorption processes. In Chapter 3, the assumptions and design data of the process simulation will be presented and a simulation case will be developed to determine the energy consumption. In Chapter 4, the results of the simulations will be presented and analyzed. In Chapter 5, the economics of each design will be evaluated against each other to determine the optimal one. Finally, a conclusion of the research will be provided in Chapter 6.

Chapter 2 Literature review

Chapter 2 of this thesis presents a literature review on carbon capture technologies, focusing on available and emerging technologies. The chapter begins by listing all the categories of carbon capture and the processes available in each one. The literature review investigates the principles, advantages, and disadvantages of each process. Furthermore, the chapter delves into the commercially dominant methods of chemical and physical absorption. Recent advancements are reviewed, including using different solvents, developing new ones, and modifying the process. By critically examining the literature, this chapter lays the groundwork for the subsequent chapters' analysis and evaluation by providing an understanding of the current state of carbon capture.

2.1 Carbon Capture Technologies

There are three categories for carbon capture which depend on the chemical composition of the CO_2-containing stream: Pre-Combustion, Post-combustion, and Oxy-Combustion. In these categories, there are various CO_2 separation methods that can be implemented: absorption, adsorption, membrane, chemical looping, hydrate-based separation, and cryogenic [5]. The choice of the carbon capture category and the separation method implemented depends on the fuel and the type of plant where the capture will be added [5].

2.1.1 Pre-Combustion

As the name suggests, in this method carbon capture happens before the complete combustion of the fuel. The fuel, such as coal, gas and biomass are sent into a reforming or gasification process where it is converted into a mixture of CO and H_2. Afterwards, CO_2 and H_2 are produced using water-gas shift and then separated. Pre-combustion is usually used in an integrated gasification combined cycle for power generation (IGCC). The process steps in this case include: CO_2 and H_2 rich synthesis gas (syngas) production from reacting steam with oxygen followed by water-gas shift, CO_2 separation, CO_2 compression and power generation [6]. Figure 1 shows the process flow diagram of the IGCC power plant [7].

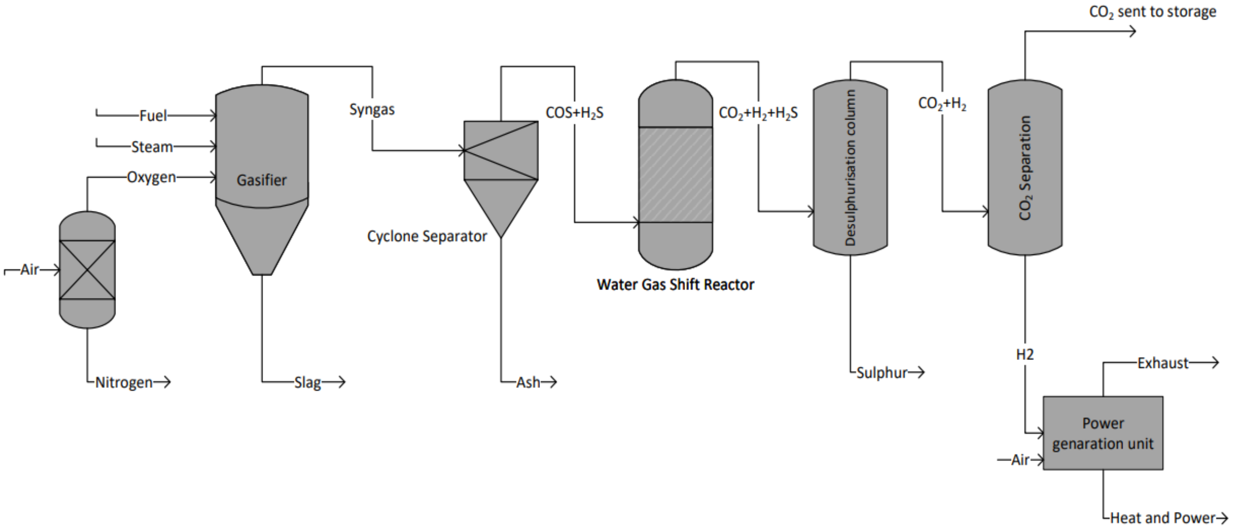

Figure 1: Pre-combustion capture system for gasification power plant process flow diagram. Adapted from Theo et al.[7]

The main advantages of pre-combustion are the production of carbon-free fuel (hydrogen under pressure) and the potential for high-pressure CO_2 production (Syngas from the gasifier is obtained at high pressures which allows the CO_2 separation process to operate at higher pressure) which reduces the power demand required for compression or liquefication of CO_2 before transportation and storage[8]. In addition, there's the added flexibility of switching between hydrogen production and power generation depending on the electricity demand. Furthermore, pre-combustion capture may increase the implementation of H_2 as an energy carrier in a low carbon energy system. However, from an economic point of view the need for multiple processes, especially gasification which is expensive, makes pre-combustion less attractive in comparison to some post combustion capture processes [8].

Pre-combustion process efficiency and feasibility, in terms of energy consumption and cost expenditure, depend on the utilities and CO_2 capture process [6]. A lot of work has been carried out to identify the best performing CO_2 separation for this process. Martin et al. evaluated the CO_2 adsorption capacity of hyper crosslinked organic polymers in comparison to zeolite-based materials and commercial activated carbon [9]. According to their results, these polymers provide superior CO_2 uptake and show good selectivity towards CO_2 and low heat of adsorption [9]. In addition, Dai et al. investigated the effect of high pressure and temperature on the performance of ionic based membrane contactors, the results showed that high pressure reduces the efficiency of the membrane since it makes the contactor become wetted, while the temperature had no effect on the process [10].

2.1.2 Post combustion

While pre-combustion carbon capture is applied before complete combustion of a fuel, post-combustion separates CO_2 from a fully combusted fuel such as flue gas. In this technique the carbon capture can be implemented in an existing process such as power plants without major modification to the plant itself. Therefore, this method is easier to retrofit compared to other technologies [4].

2.1.3 Oxy combustion:

Unlike the other methods where the combustion is done in air, in Oxy-Combustion the fuel is burnt in high-purity oxygen. Therefore, the flue gas produced in this process is mostly made of water and carbon dioxide. After the combustion CO_2 is captured by condensation which removes water from the product stream. Finally, the captured CO_2 is purified and compressed to the required pressure either for transportation or utilization in the cycle [11]. Figure 2 shows the block diagram of Oxy-combustion CO_2 capture [6].

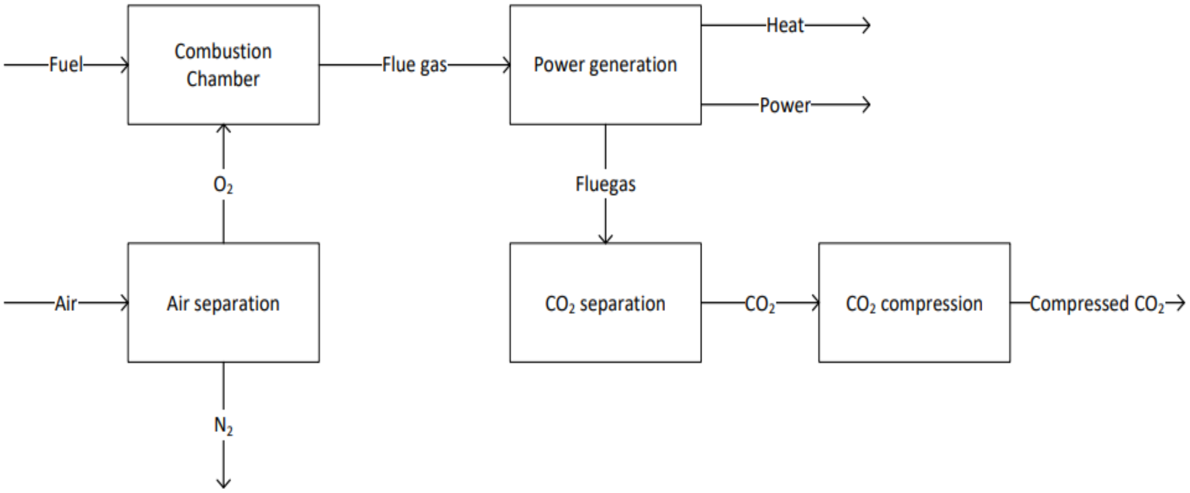

Figure 2: Oxy Combustion block diagram [6]

In conventional air firing systems, a large amount of N_2 remains, since in Oxy-Combustion this bulk amount of N_2 is absent, a much lower amount of NO_x is produced. Additionally, there are no other pollutants in the product stream. However, this method has a high operational cost mainly due to the need to produce high-purity oxygen [12,13]. Many studies are ongoing to enhance the understanding and refinement of Oxy-Combustion. Barba et al. studied nineteen different oxy-turbine cycles for CO_2 capture and identified the main operational parameters and the best cycle based on a Political, Environmental, Social, Technological, Legislative and Economic (PESTLE) risk analysis. [14]. F. Smith et al. conducted research on oxygen membrane reactors for oxy-fuel combustion and carbon capture and determined the O_2 permeability of the membrane reactors and the CO_2 selectivity. According to their results an 85% to 87% CO_2 selectivity could be achieved [15].

2.2 CO_2 separation methods

2.2.1 Membrane separation

In this method the gas is cooled to the operating temperature of the membrane in a wet scrubber before being sent to the membrane [16]. The CO_2 separation can be done using two types of membrane technology: gas separation membranes and gas absorption membranes. The difference between the two technologies is the mechanism by which CO_2 is separated. For gas separation membrane systems, the CO_2 is separated based on the pressure on both sides of the membrane. CO_2 is introduced into the membrane at high pressure and recovered on the other side of the membrane at low pressure. On the other hand, gas absorption membrane systems use a microporous solid membrane and an absorption liquid flow to separate CO_2 from the gas [6]. A schematic of a typical membrane CO_2 capture process is shown in Figure 3.

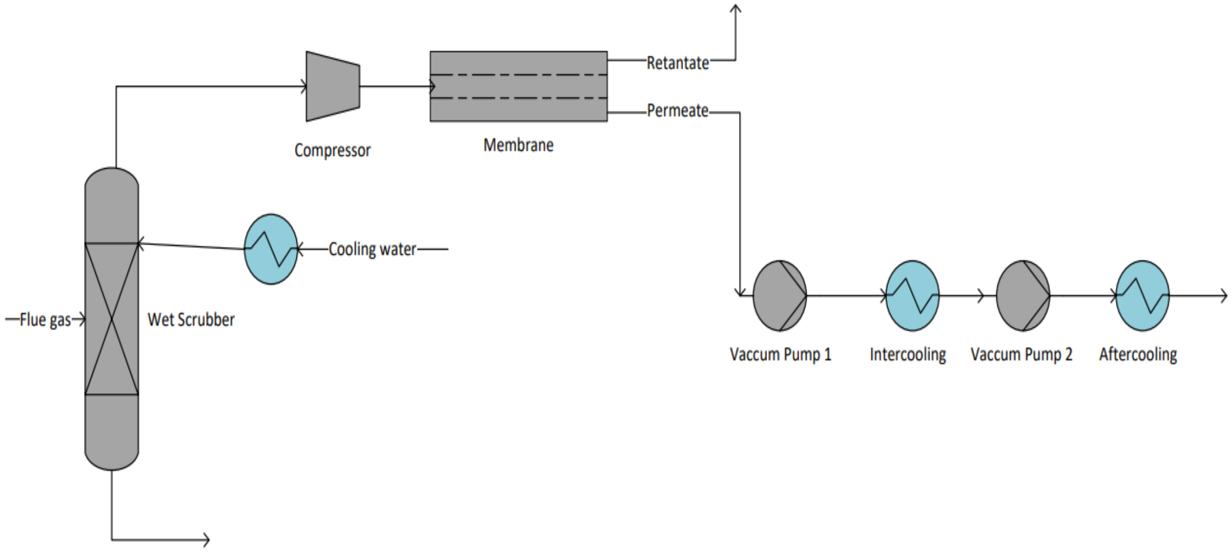

Figure 3: Membrane CO_2 capture process flowsheet. Adapted from Wang et.al [16]

The main advantages of membrane separation are operating without weeping, entrainment, foaming and flooding issues, which are commonly occurring problems when operating a packed column [4]. However, when CO_2 concentrations in the gas is low, the efficiency of membrane separation significantly reduces. In addition, membrane lifetime is limited and they require periodic replacement, which increases the operating cost [4]. Furthermore, the desired condition for the membrane to be efficient is to completely fill the membrane pores with gas. However, the liquid phase could replace the gas in the membrane pores which leads to a less efficient mass transfer through the membrane, this phenomenon is known as wetting [4]. Several studies have been conducted to investigate the wetting effect on the mass transfer through the membrane. Malek et al. modelled hollow fiber membrane modules operated under partial wetting conditions. The results showed that a high mass transfer coefficient is attainable with respect to the water velocity [17]. On the other hand, Zhang et al. studied the effect of membrane properties on CO_2 absorption. According to the authors, an increased length and the number of fibers improves the CO_2 capture, while a decreased thickness, inner fiber radius and inner module radius increase CO_2 removal but decrease the absorption performance [18].

2.2.2 Adsorption

In adsorption, a component is removed from a mixture using a solid surface. The gas stream containing CO_2 is passed through an adsorption column filled with adsorbent which the CO_2 then adheres to its surface leading to adsorbent saturation. Afterwards, the CO_2 saturated surface is regenerated and CO_2 is desorbed. Figure 4 illustrates the process flow diagram for a flue gas adsorption process. The flue gas enters a pretreatment stage that consists of a fan, a dryer, and a compressor at 1 atm, 25°C with 9.7% CO_2. In order to have a continuous process, the gas stream is sent to one of three parallel adsorption chambers; one chamber receives the feed, the second regenerates the adsorbent by desorbing the captured CO_2, and the third is on stand-by to receive more of the feed. The illustrated adsorption process can capture 97% CO_2 [19].

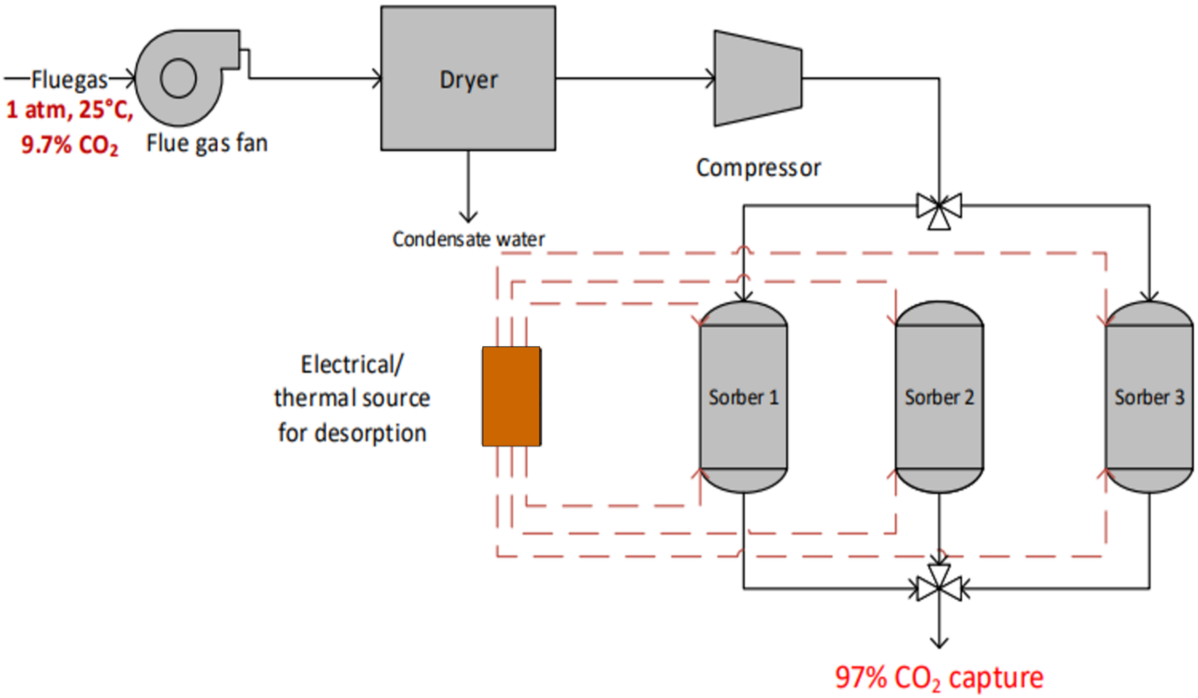

Figure 4: Adsorption process flow diagram; Adapted from Song et.al [19]

Various methods of adsorbent regeneration can be applied based on the driving force used, each with different advantages and disadvantages [4]. First off, in temperature swing adsorption (TSA) the bonds between CO_2 and the adsorbent are broken by increasing the temperature. This high increase in temperature required makes this method expensive, in addition to being time-consuming [20]. Instead of regular heating, this can be done using electrical swing adsorption (ESA) where the adsorbent is heated up using low-voltage electrical current by applying the joule effect. However, high grade electrical energy is required to achieve this in comparison to low grade heat energy used in TSA [21]. Pressure and vacuum swing adsorption (PSA and VSA) can be used to regenerate the adsorbent by reducing the pressure to separate the CO_2. Vacuum is used when the feed pressure is close to ambient pressure [20]. These two methods are affected by the CO_2 partial pressure and concentration. If the partial pressure is high, pressure swing adsorption (PSA) is not favorable, and if the concentration is low, a longer time is needed [22].

Adsorption has been studied extensively. The focus of research on this process is finding the optimal sorbent to separate CO_2. Li et.al conducted research on metal-organic frameworks (MOFs) in adsorption processes [23]. MOFs are a mix of metal ions or ion clusters linked by organic ligands and bridges which form strong bonds. The experimental results showed that these MOFs have high porosity, ease of design and tailored pore properties for CO_2 capture [23]. In addition, instead of changing the sorbents, Agarwal et al. worked on developing a pressure swing adsorption superstructure to design optimal PSA cycle configurations. The results showed that a PSA cycle with up to 98% purity and recovery of CO_2 could be predicted using the superstructure [24].

2.2.3 Chemical looping combustion (CLC) and calcium looping process (CLP)

In chemical looping combustion two reactors are used to capture CO_2, an air reactor, and a fuel reactor. The reactions that occur in the air and fuel reactor are shown in equation (1) and (2) as follows [6].

Fuel reactor: $(2n+m)M_yO_x + C_nH_{2m} \rightarrow (2n+1)M_yO_{x-1} + mH_2O + nCO_2$ (1)

Air reactor: $M_yO_{x-1} + \frac{1}{2}O_2 \rightarrow M_yO_x$ (2)

An oxygen carrier such as metal particles (iron, manganese or copper) oxidizes with the oxygen from the air in the air reactor (M_yO_x), as depicted in Equation (2). The metal oxides formed are sent to the second reactor containing the fuel where it consumes the oxygen content of the metal oxide and reduces that to M_yO_{x-1}, as shown in Equation (1). The reduced metal oxide is then sent back to the air reactor, completing the loop. Energy and a flue gas stream of CO_2 and H_2O are produced due to the reduction of the metal oxides during combustion. The flue gas is then condensed to receive pure CO_2 [25]. A schematic of the chemical loop is shown in Figure 5 [26].

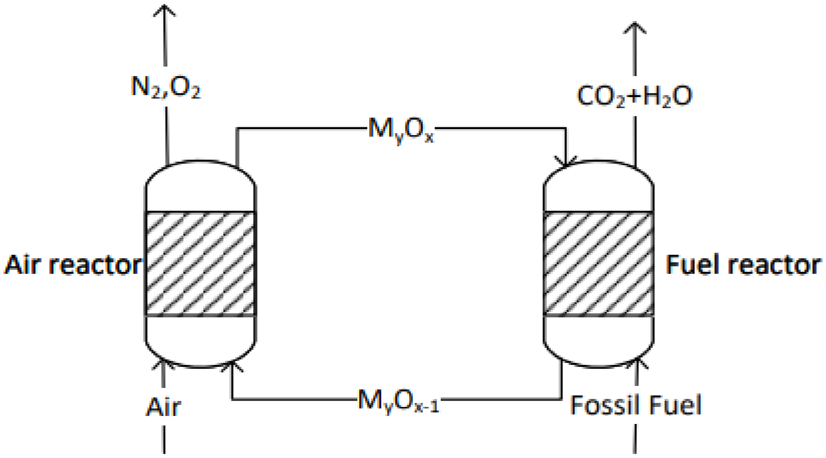

Figure 5: Chemical looping combustion schematic, adapted from Hossain et.al [26]

For the calcium looping process, instead of an oxidization reaction, a reversible carbonation reaction between calcium oxide and carbon dioxide occurs in the first reactor, shown as the forward reaction in equation (3). The formed carbonate is sent to the second reactor where the reverse reaction forms CaO and CO_2 and a high-purity CO_2 stream is produced [27].

$$CaO + CO_2 \leftrightarrows CaCO_3 \quad (3)$$

The CaO sorbent used is generally lime, which is relatively inexpensive. Additionally, the reaction to form the carbonate is exothermic and the heat produced can be used to generate power [28]. On the other hand, the sorbent loses its carrying capacity with repeated use, consequently, a huge amount of sorbent is required [29].

Chemical looping and calcium looping for CO_2 capture are emerging technologies. However, their main advantage is the high operating temperature and resulting high-grade heat produced by the reactions. Hanak et al. studied the integration of CLP for the decarbonization of a coal fired power plant [28]. Current state-of-the-art calcium looping integration scenarios involve recovering high-grade heat to produce steam at high pressure. This steam is subsequently utilized to generate power through the conventional steam cycle, which is based on the Rankine cycle [28]. Hanak et al. proposed the implementation of a supercritical CO_2 cycle (s-CO_2 is carbon dioxide held above its critical temperature and pressure [30]) instead of steam. According to their results, the proposed design is 1% to 2.1 % more efficient than using a steam cycle depending on the configuration. Moreover, the authors suggest that this design could also be implemented in chemical looping [28].

2.2.4 Cryogenic

The cryogenic method captures CO_2 by providing cold energy using refrigerants such as liquified natural gas. With this method, high-purity CO_2 can be produced. The only advantage of cryogenic is that a few processes are needed, compression, expansion, separation, and cooling. However, the operational cost of this method is very high [19,31]. A cryogenic CO_2 capture process flow diagram is shown in Figure 6 [32]. The feed gas is sent to a compression system to be cooled before being sent to the pre-cooler and liquefier for further cooling. The obtained liquified gas is sent to the distillation tower to separate the CO_2 from the other gases.

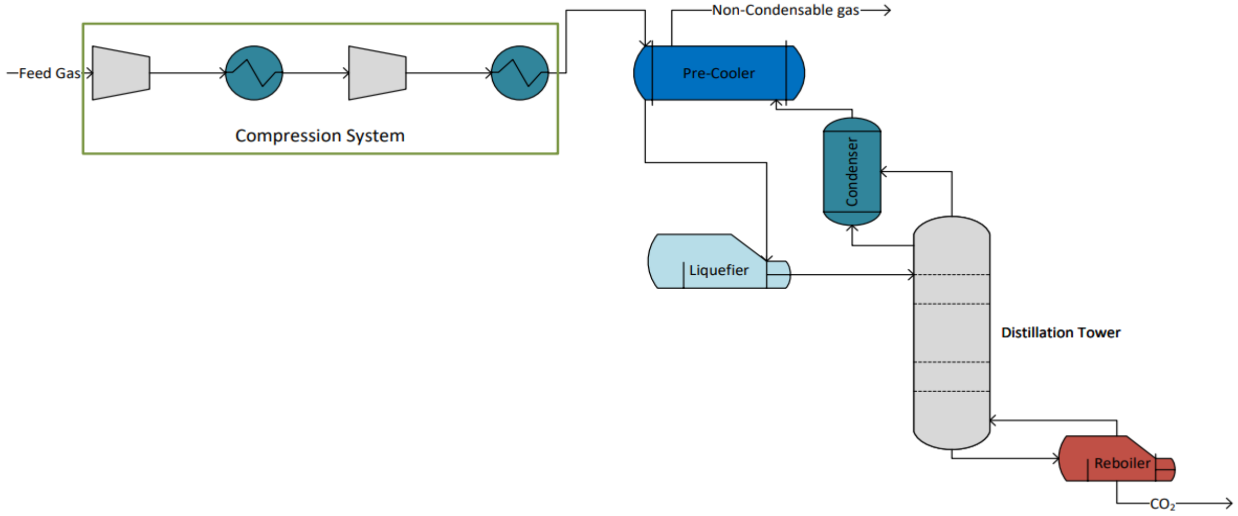

Figure 6: Cryogenic CO_2 capture process flow diagram, adapted from [32].

Due to the high energy consumption and cost of cryogenic CO_2 capture, the integration of cryogenic with other CO_2 capture technologies has been suggested and explored in some works [19]. For instance, a hybrid technology based on membrane-cryogenic separation was proposed by Scholes et al. According to their results, combining cryogenic with a three-stage membrane separation could achieve a high CO_2 capture rate with a competitive capture cost in comparison to conventional technologies [33].

2.2.5 Solvent based absorption

Separating CO_2 from a gas by absorption in a solvent has been developed and employed on an industrial scale for over 50 years [4]. Chemical absorption for example has been used to remove CO_2 from natural gas operations, while power plants and industrial plants have used chemical and physical solvents to recover CO_2 from flue gas to be used in the food processing and chemical industries [34]. Absorption uses either a chemical or physical solvent to separate CO_2 from a gas stream. For chemical absorption, a reaction between a chemical solvent and carbon dioxide occurs and a rich lean stream is obtained which then is separated to obtain pure CO_2. Physical absorption, however, uses a chemically inert solvent to physically absorb CO_2 [20]. Chemical absorption is a very energy-consuming process. Low selectivity and loss of efficiency are the major drawbacks of physical absorption [6,35]. Since this separation method has already been implemented in existing plants and industries, modifications could be applied to them faster than newly developed CO_2 separation methods that have not been implemented on the industrial scale [36]. Therefore, solvent-based absorption remains the most common and widely used method, particularly in CCU projects [36]. In this thesis, we also focused on the absorption processes, particularly since one of our objectives was to revamp the existing CO2 capture units and make them suitable for CCU processes. A deeper review of chemical and physical absorption will be discussed later in the thesis in Section 2.3.

2.2.6 Conclusion

Each technology has different performance depending on the characteristics of the feed gas, operating conditions, CO_2 capture capacity, and the required utilities and their technology readiness level. For instance, Voldsund et al. investigated various CO_2 capture technologies for the CO_2 capture of a cement plant [37]. According to their process assessment and techno-economic analysis results, while some technologies can achieve lower operating costs, no technology can be identified as a superior process when all metrics are included. Other metrics that affect the decision-making are retrofit ability in actual plants (space requirement, added load to local power grids) and overall emissions reductions [37]. The summary of the advantages and disadvantages is shown in Table 1.

Table 1: Advantages and Disadvantages Summary

Separation method	Advantages	Disadvantages
Membrane separation	- Operating without weeping, flooding and foaming	- Low membrane efficiency - Limited membrane lifetime
Adsorption	- Variety in driving forces to separate CO_2	- High cost for some driving forces
Chemical and calcium looping	- High operating temperature resulting in high grade heat production - Inexpensive sorbents	- Sorbent efficiency loss - A big amount of sorbent is required
Cryogenic separation	- Ability to implement it with other separation methods	- High energy consumption and cost
Solvent-based absorption (chemical and physical absorption)	- High process efficiency when using chemical solvents - High CO_2 purity for both chemical and physical solvents	- High energy demand in chemical absorption - Loss of efficiency in physical absorption

2.3 Chemical and physical absorption

As mentioned earlier, the most common capture process is solvent-based absorption which includes two main sub-categories: physical solvent absorption and chemical solvent absorption. While the chemical routes have been used more extensively, they have two major problems: the high energy required for CO_2 separation and recovery in the stripper and the high CO_2 capture cost [36,38]. A classic chemical absorption process uses 20-30 wt% aqueous monoethanolamine (or other amines) to separate CO_2 since it was proposed for acid-gas separation in an amine process in the original patent from 1930 [39]. For a 90% CO_2 removal from flue gas using 30wt% MEA, typical energy requirements for separation and recovery of CO_2 ranges between 3-4 MJ/Kg-CO_2 captured, as shown in Figure 7. The values shown in the figure have been validated in a number of studies for actual small to medium-pilot-scale plants [40–43]. Furthermore, according to S.M. Hassan the capture cost for a typical amine capture process ranges between \$48 to \$55\tonne$CO_2$ [44]. On the other hand, in the physical case, the main challenge is the efficiency loss of the CO_2 in the capture unit and low selectivity towards CO_2 [6,35].

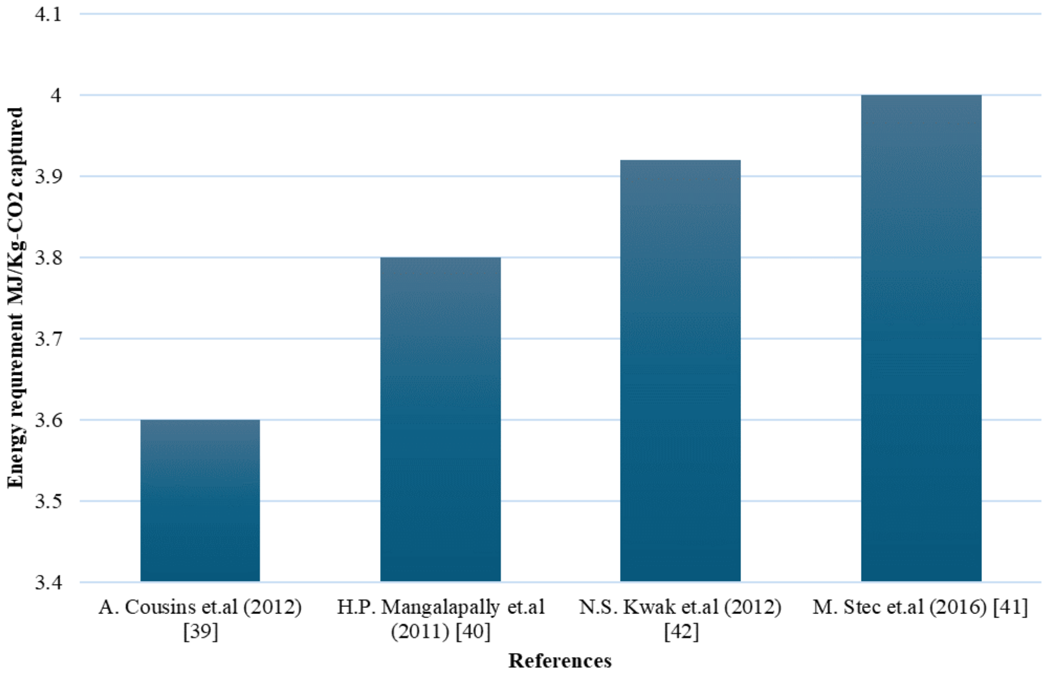

Figure 7: Energy requirement range for separation and recovery of CO_2 [40–43]

2.4 Current advancements in chemical and physical absorption

In order to address the high energy consumption, high capture cost and loss of efficiency, several aspects of the process must be studied [5]:

i. Utilization and development of alternative solvents instead of the widely used ones (MEA, MDEA, DPEG…)
ii. Modifying the process configurations
iii. Optimizing the already available process flowsheet
iv. Energy integration

Borhani et al. [5] conducted an extensive study on the available solvents and listed the advantages and disadvantages of each solvent. The author also conducted experiments to review the selection and design of solvents, in addition to using predictive models. Mumford et.al [45] discussed some of the common solvents used and briefly went into mixed solvents and their applications. Salkuyeh et.al [36] focused on the two most common chemical solvents MEA and DGA based on their advantages over the other amines. The authors discussed the effect of the CO_2 concentration in the feed on the efficiency of the solvent. In addition, they proposed different changes to the process to reduce energy consumption. The results showed that DGA can reduce the reboiler duty for minimum loading case by 50-70% in comparison to 13-20% for MEA for different CO_2 concentrations. Xin et al. [46] focused on testing different mass transfer correlations with experimental mixtures of physical solvents in order to reduce the capture cost. According to their results depending on the correlation used different physical properties affect the capture cost, for example when liquid viscosity is modified the conventional physical absorption process is more economical.

One method for reducing the cost and energy consumption of CO_2 capture processes is to implement innovative technologies such as industrial heat pumps and vacuum pumps (item ii). SINTEF Industry and Politecnico di Milano developed a novel Pressure Swing Adsorption method for CO_2 capture from the flue gases of a cement plant [47]. According to their design, by incorporating a vacuum heat pump in the CO_2 recovery step and an intermediate heat pump that links the adsorption and regeneration steps, a CO_2 capture demand, as low as 2.04 MJ/kg CO_2 capture is achievable for 90% CO_2 capture from the flue gas [47]. Furthermore, Jensen et al. investigated the electrification of the CO_2 capture unit of a biogas upgrading unit, using a 2-stage heat pump [48]. Their results showed that up to 68% reduction in the overall energy demand is achievable [48].

The other method for energy consumption reduction of the CO_2 absorption processes is the use of a stripping agent to facilitate the CO_2 recovery step. Machida et al. investigated the hydrogen injection impact on the heat demand of the CO_2 capture process based on Monoethanolamine (MEA) solvent [38]. According to their simulation results, the hydrogen injection can reduce the temperature profile of the amine regeneration column by around 15°C. However, the overall energy demand increases by increasing the hydrogen ratio in the recovered gas.

While there are several works on the design and analysis of various solvents, there are still challenges regarding the energy consumption and cost of the CO_2 capture process. On one hand, the conventional processes, even after innovative modifications require a high amount of energy, more than 3-4 MJ/kg CO_2 [36,38]. On the other hand, advanced technologies, such as membrane [49,50] and chemical looping [51] are still not ready for large-scale CO_2 capture systems. These challenges become even more important for the CCU pathways, as the CO_2 capture unit is one of the most expensive sections of such pathways [52].

In this thesis, we tried to address these challenges by investigating all four aspects listed previously. First, we investigated two different pathways and solvents for CO_2 capture: physical (DEPG) and chemical solvents (MEA and DGA). Second, by combining the efficient electrification concept using an industrial heat pump with the incorporation of the hydrogen injection at the CO_2 stripper we tried to reduce the total energy demand based on the work of Machida et.al [38]. Several process simulations are conducted to identify the best solvent and design.

Besides the CO_2 capture for flue gas, we investigated the proposed processes for the syngas of the ammonia production process. Ammonia plants are one of the most energy-intensive chemical processes, which account for around 2% of the global CO_2 emissions [53]. First, we conducted the process simulation of the existing syngas treatment using chemical absorption and MEA solvent. After the validation of the incorporated reaction chemistry and process model, we designed alternative approaches using a different solvent (DGA), and by incorporating the heat pump.

The use of industrial heat pumps that utilize the low-temperature heat of the condenser to supply the energy demand of the reboiler is an effective approach to reduce the energy demand of the separation and recovery steps. Emerging industrial heat pumps can satisfy energy demands at temperatures up to 200°C, making them appropriate for many industries and applications [54]. When renewable power is used, these modern industrial heat pumps may give a high coefficient of performance while emitting no greenhouse gases or depleting fossil fuels. Several works have been undertaken for the incorporation of such heat pumps for various processes [55,56]. For the heat pump design, we used the CoolTools version 1.0.3 [57] software to evaluate different refrigerants for the heat pump. The selected heat pump is then modelled in Aspen Plus® V12.1 and integrated with our amine simulation model. The model is then modified with the hydrogen injection alternative to simulate the impact of the use of hydrogen as the stripping agent. The physical solvent model using DEPG solvent, with and without hydrogen is also developed and results are compared with those of the chemical solvent routes. A similar approach is also used for the flue gas as the feedstock.

2.4.1 Heat pump description

A heat pump system consists of three steps [58]:

- Receiving heat from a waste-heat source
- Rasing the temperature of the waste heat
- Delivering useful heat at the elevated temperature

Industrial heat pumps are very energy efficient compared to traditional heating methods because they generate more thermal energy than the electrical energy needed to operate the compressor, resulting in significant energy savings. The efficiency of a heat pump is determined by calculating the coefficient of performance (COP) [59]. The COP is a measure of how efficiently a heat pump transfers thermal energy compared to the electrical power it consumes. For instance, if a heat pump uses 1 kW of electrical energy to transfer 3 kW of heat, the COP would be 3 [60]. The equation to calculate the COP is [61]:

$$COP = \frac{|Q|}{W} \qquad (4)$$

Where:

- Q is the heat output of the HP condenser
- W is the work required to operate the heat pump compressor

The key benefits of heat pumps are [59]:

- They reduce overall energy use
- They reduce carbon emissions since renewable energy sources could be used to provide electricity to drive the pump
- They allow more scalability in comparison to other available technologies such as bioenergy and geothermal
- They reduce reliance on imported fossil fuels which increases the security of the supply

2.5 Difference between conventional and novel design

As stated earlier, we introduced the novel hydrogen injection concept for both chemical and physical solvents and investigated various process configurations to reduce the overall energy consumption. The proposed processes are evaluated for two types of feedstocks, Flue gas and high-pressure syngas. Multiple designs were tested for each feedstock.

A block diagram of all the designs is shown in Figure 8, which includes two design configurations for physical absorption:

- Physical absorption without hydrogen injection
- Physical absorption with hydrogen injection

And also, four designs for the chemical absorption:

- Chemical absorption without hydrogen injection or heat pump
- Chemical absorption without hydrogen injection and with a heat pump
- Chemical absorption with hydrogen injection
- Chemical absorption with both hydrogen injection and heat pump

It should be noted that in the thermal and cost analysis of each route, the energy demand and cost of hydrogen production are not accounted for. This is due to the fact that the hydrogen is not consumed during the CO_2 capture process. Indeed, the hydrogen that was previously intended for the CO_2 conversion unit of the CCU plant (for instance, the CCU-methanol process), was re-directed to the CO_2 capture section and the CO_2+H_2 mixture is then delivered to the CO_2 conversion reactor.

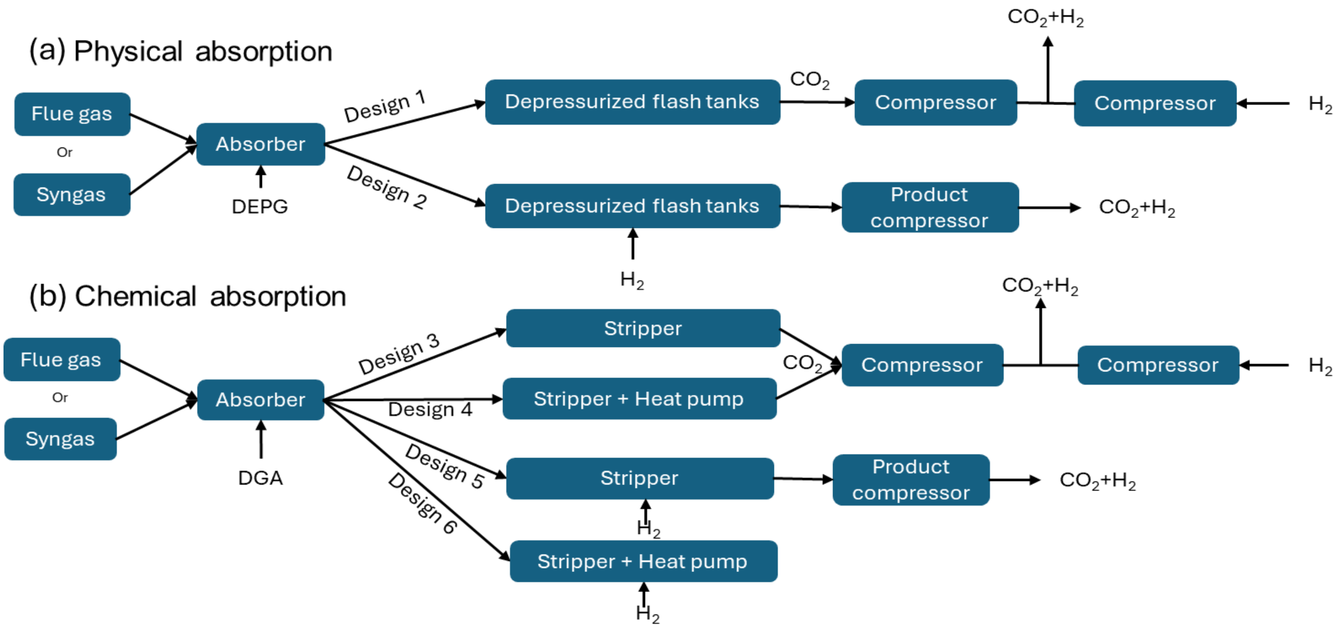

Figure 8: Block diagrams: (a) physical absorption, (b) chemical absorption

2.5.1 Conventional Process

As illustrated in Figure 8 in the conventional CO_2 capture process (Design 1, 3) the feed gas is sent into an absorber where it is mixed with the solvent and CO_2 is removed from the gas. The rich solvent is then sent to the regeneration unit. In the physical absorption route, the regeneration unit includes depressurization tanks (flash tanks), while it consists of a regeneration column (stripper) in the chemical absorption route. After removing CO_2, the solvent is recycled back to the absorber [46,62]. The recovered CO_2 is then compressed and mixed with H_2 before being sent to the CO_2 conversion unit (e.g. CCU-methanol).

2.5.2 Proposed novel process

In comparison to the conventional design, the proposed process consists of injecting H_2 directly into the regeneration section instead of mixing with the captured CO_2 recovery (Figure 8 design 2, 5, 6). The key advantage is that the partial pressure of CO_2 in the gas stream decreases. In the physical route, it means that the solvent depressurization section can happen at a higher pressure compared to the conventional process. Since the environment inside the flash tanks is decompressed which subsequently decreases the CO_2 partial pressure and increases the amount of CO_2 released from the solvent, the efficiency of CO_2 capture increases. Furthermore, due to the H_2 injection, the temperature in the flash tanks liquid output streams increases. As a result, it becomes possible to run the lean DEPG at a higher temperature into the absorber without the need to cool it down. However, a higher flow of DEPG is needed to achieve the required CO_2 capture rate. In addition, this increase in temperature eliminates the need for the lean DEPG recycle cooler which affects the cost since the cooling cost is higher than the increase in DEPG flow cost. (Details in Supplementary Document)

The same concept applies in the stripper for the chemical absorption, the H_2O partial pressure decreases, as a result, the bottom temperature decreases since the temperature is mainly controlled by the boiling point of the solution [38]. However, because of the increase of the H_2/CO_2 ratio due to H_2 stripping the required energy for the reboiler increases. Even though the energy consumption increases, the lower temperature at the stripper outputs improves the coefficient of performance of the heat pump.

Taking into consideration the effect of injecting H_2 into the regeneration section and the different designs proposed for both physical and chemical absorption, an economic analysis is done on all the designs to determine the best one.

Chapter 3 Process simulation, Description, and Economic assumptions

This chapter focuses on the process simulation of chemical and physical absorption. The chapter begins by presenting the initial conditions and composition of the feedstocks. Afterwards, a detailed process description of both chemical and physical absorption is discussed, followed by the economic assumptions and parameters used to conduct the economic analysis of the designs developed. The aim of this chapter is to provide a detailed representation of how to simulate the base case for physical and chemical absorption, in addition to the modified novel designs.

3.1 Feedstock composition and initial conditions

The process simulation of the physical and chemical absorption considered in this work is performed using Aspen Plus® V12.1. As mentioned earlier, the process modellings of all different scenarios are conducted for both flue gas and syngas feedstock. The composition of the flue gas and syngas fed into the absorbers is presented in Table 2. The flue gas originates from a cement kiln with a flowrate of 250 tonnes per hour [3]. The syngas flowrate (100 tonnes per hour) and composition are adopted from an ammonia plant. In this plant, ammonia synthesis is conducted by converting the natural gas to synthesis gas, in a steam methane reformer. The raw syngas is then sent to the shift reactor for further conversion of syngas to H_2. The product, before being sent to the ammonia synthesis unit, is sent to the CO_2 capture unit, where mono-ethanol amine (MEA) solvent is used for the chemical absorption of CO_2.

Table 2: Feed gas composition (%mole)

	CO_2	H_2O	N_2	O_2	H_2	CO	CH_4
Flue-gas	0.224	0.072	0.681	0.023	0	0	0
Syngas	0.175	0.009	0.198	0	0.61	0.004	0.004

3.2 Process simulation and description

3.2.1 Physical absorption process description

For the physical process, the CO_2 capture process is designed using dimethyl ether of polyethylene glycol (DEPG) solvent at the inlet temperature of -1.11°C. The Perturbed Chain-Statistical Associating Fluid Theory (PC-SAFT) model is used for the calculation of the vapor-liquid equilibrium of the DEPG solvent and gases, their density, volatility and heat capacity, and the simulation of this process [63]. The PC-SAFT has been extensively used for the modelling of different types of physical solvents [64]. A simplified schematic of the process is shown in Figure 9. The feed gas is contacted with the solvent in the absorber, after being cooled and compressed. CO_2 is stripped from the solvent in the regeneration unit which consists of a series of flash tanks where the rich solvent is depressurized gradually from 17 bar to 6.9 in the slump tank (Stream 1), 1.72 in the first flash tank (Stream 2, 3), 1.01 in the second tank (Stream 4, 5), and 0.3 in third (Stream 6, 7) for flue gas. The syngas is only cooled, since it is already at a high pressure of 27.7 bar and depressurized to 17 bar in the slump followed by the same depressurization in the other flash tanks. The flue gas is not compressed to the same pressure as the syngas in order to avoid increasing the energy demand and consequently the cost since the compression accounts for the highest percentage in the total equipment cost distribution, which will be discussed in the economic analysis in section 5.2. The recovered solvent is then pumped and cooled to the original pressure and temperature of -1.11°C, 17 bar (27.7 bar) (Stream 8). The parameters of the absorber and the other units are adjusted to achieve a 90% CO_2 capture rate, the final parameters for flue gas and syngas are shown in section 3.2.3.

In our integrated design, the hydrogen is injected into the first flash tank at 2 bar and 25°C after being run through a separator to remove any water present in the stream. The hydrogen flowrate is adjusted to have an outlet H_2/CO_2 ratio of 3, which is suitable for most CCU pathways, such as methanol and ethanol production [65]. Since H_2 allows a higher pressure in the flash tank, the pressure is adjusted to 2 bar (Stream 2, 3) the same pressure as the hydrogen stream.

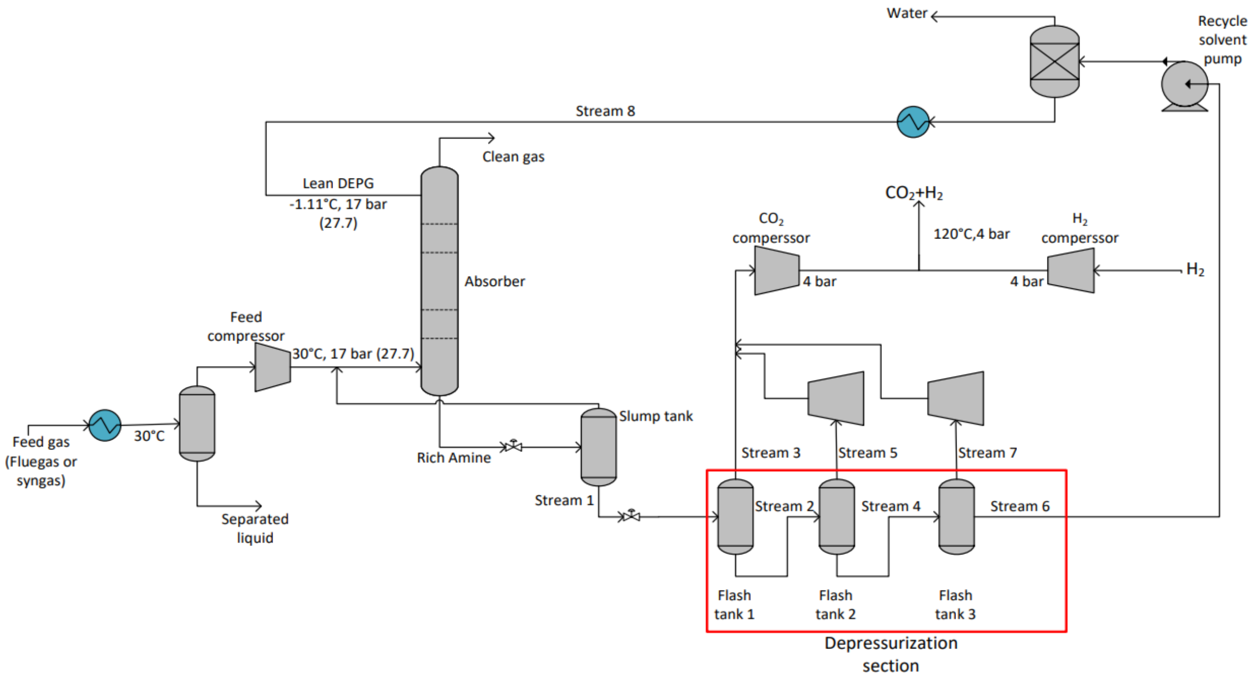

Figure 9: Physical absorption process flowsheet

3.2.2 Chemical absorption process description

For the chemical absorption case, after the simulation of the baseline process with syngas and validation of the results with the actual process data, the process model is modified with the Diglycol Amine (DGA). While both MEA and DGA are both primary amines, since DGA has lower corrosion issues, and can be used at higher concentrations with higher CO_2 loading in the CO_2 reach stream [42–44]; thus, DGA reduces the circulation rate and the heat-duty of reboiler compared to those of MEA [4,22]. As mentioned earlier, in addition to the syngas case, another process simulation is done using flue gas as the feedstock.

The modified chemical absorption process was based on the work of Salkuyeh et.al [36], and involves Diglycolamine (DGA) at 30°C with the ElecNRTL model. The reactions that occur between the amine and CO_2 in the absorber are as follows [36]:

Bicarbonate formation: $$CO_2 + 2H_2O \leftrightarrow HCO_3^- + H_3O^+ \quad (5)$$

Ionization of water: $$2H_2O \leftrightarrow H_3O^+ + OH^- \quad (6)$$

Protonation of alkanolamine: $$RNH_2 + H_3O^+ \leftrightarrow RNH_3^+ + H_2O \quad (7)$$

Carbamate formation: $$RNH_2 + CO_2 + H_2O \leftrightarrow RNHCOO^- + H_3O^+ \quad (8)$$

As shown in Figure 10, after the absorption section, the rich amine is heated in a process heat recovery exchanger and sent to the CO_2 stripper column. The design is conducted based on a 99% CO_2 recovery in the stripper and 90% overall CO_2 capture rate. Hence, it should be noted that two different temperatures (T_{rich}) for the rich stream going into the stripper are obtained for flue gas and syngas in the process heat exchanger, 95°C and 112°C respectively. This temperature is affected by the stripper reboiler output temperatures T_{reb}. In addition, two reflux, boil-up ratios (0.04-0.7, 0.025-0.51) are used to achieve the design recovery percentages specified. Details of the methodology for the process simulation can be found in our previous work [66], and not shown here for brevity.

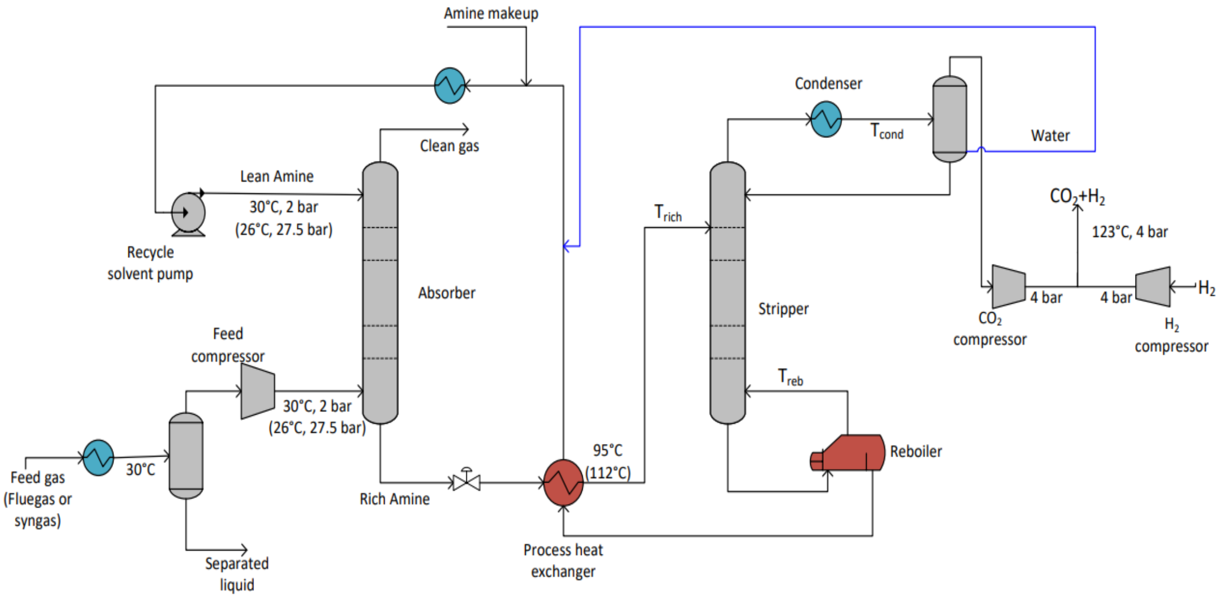

Figure 10: Chemical absorption process flowsheet

Besides modifying the process by using a more efficient solvent, the use of a heat pump is incorporated into our new design. One of the main challenges was to define the best heat pump that could effectively transfer the low-temperature heat. Using CoolTools software [67], we investigated 14 different heat pump fluids to define the system with the highest COP. Our results showed that the vacuum steam can achieve the highest coefficient of performance, with a COP of up to 6.3. A simplified schematic of the modified design is shown in Figure 11. The waste heat stream evaporates the working fluid at low temperature (T_2) before being compressed to a higher pressure which increases temperature (T_3, P_3). The heat pump working fluid condenses at this high temperature and pressure which consequently provides heat to the process, in this case the reboiler. The condensed fluid (T_4) is then expanded back to the evaporator (T_5, P_5). The temperatures and pressures obtained for all the designs are shown in Table 3.

Table 3: Heat pump temperatures and pressures

Design	T_1°C	T_2°C	T_3°C, P_3 bar	T_4°C	T_5°C, P_5 bar	T_{cond}°C	T_{reb}°C
Flue gas							
Conventional with HP	103	79	426, 4.8	150	73, 0.35	79	143
Design with HP/H_2	102	82	389, 4.3	146	76, 0.41	82	140
Syngas							
Conventional with HP	115	100	315, 4.3	147	95, 0.8	100	136
Design with HP/H_2	100	87	301, 2.7	130	81, 0.45	87	122

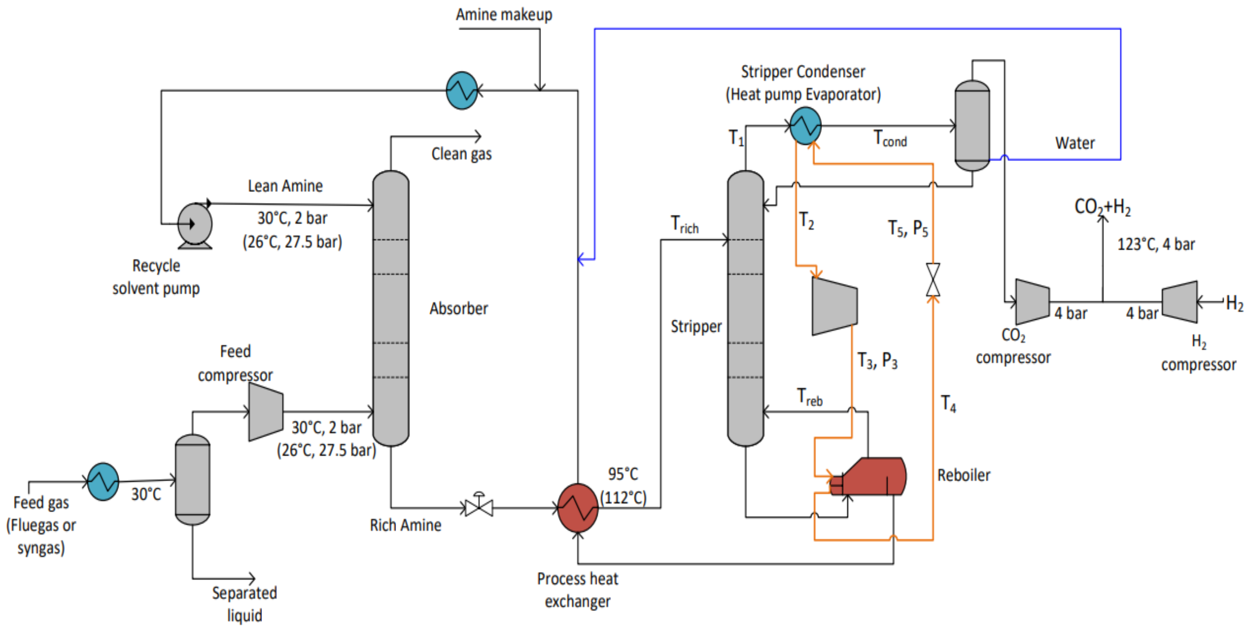

Figure 11: Chemical absorption with heat pump schematic

3.2.3 Design parameters and assumptions

The following table presents a summary of all the parameters and assumptions to develop a process simulation of chemical and physical absorption using Aspen Plus® software. As previously mentioned, these parameters are adjusted based on a 90% overall CO_2 capture rate.

Table 4: Design parameters and assumptions

Flue gas stream	
Temperature	160°C
Pressure	1 bar
Syngas stream	
Temperature	63°C
Pressure	27.7 bar
Physical absorption	
Absorber	
Calculation type	Equilibrium
Number of stages	12

Column pressure drop	1 bar
Lean DEPG temperature	-1.11°C
Lean DEPG pressure for Flue gas (Syngas)	17 bar (27.7 bar)
Gas-in temperature	30°C
Gas-in pressure for Flue gas (Syngas)	17 bar (27.7)
CO_2 capture rate	90%
Chemical absorption	
Absorber	
Calculation type	Equilibrium
Number of stages	20
CO_2 Murphree efficiencies	0.33
Column pressure drop for Flue gas and Syngas	0.3 bar
Lean DGA temperature for Flue gas (Syngas)	30°C (26°C)
Lean DGA pressure for Flue gas (Syngas)	2 bar (27.5 bar)
Inlet gas temperature for Flue gas (Syngas)	30°C (26°C)
Gas-in pressure for Flue gas (Syngas)	2 bar (27.5 bar)
CO_2 capture rate	90%
Stripper	
Calculation type	Equilibrium
Number of stages	20
CO_2 Murphree efficiencies (excluding the condenser and reboiler)	50%
Column pressure drop for Flue gas and Syngas	
Flue gas Reflux ratio	0.3 bar
Flue gas Boil-up ratio	0.04
Syngas Reflux ratio	0.7
Syngas Boil-up ratio	0.025
Rich solvent temperature for Flue gas (Syngas)	0.51
	95°C (112°C)

3.3 Economic assumptions and parameters

Aspen Process Economic Analyzer is used for the equipment sizing and capital cost estimation of all different units, except for the heat exchanger network. Using the Aspen Energy Analyzer, the heat exchanger network is designed for a minimum energy target. Aspen Energy Analyzer is also used to estimate the cost of the heat exchange network for all design cases. The capital estimation values are updated to 2022, using the CEPCI index. The key cost parameters and assumptions, as well as the prices, are listed in Table 5, respectively. The total capital cost and operating costs estimation, as well as the annualization of the capital costs, are conducted based on the procedure explained by Seider et al. [68], and not discussed here for brevity.

Table 5: Economic analysis parameters, assumptions, and prices

Prices ($US)			
Parameter	**Value**	**Parameter**	**Value**
Base year	2022	***CEPCI***[1]	808.7 (2022)
Electricity	$13.2 per kilowatt of billing demand plus 2.7 cents per kWh [69]		
Labor wage, $/hr	26.26 $/hr [70]	***DEPG***	7200 $/tonne [46]
DGA	2050 $/tonne [71]	***Cooling water***	0.031 $/tonne
Process water	1.231 $tonne	***Waste water disposal***	0.492 $/tonne
CAPEX[1] **assumptions**			

Parameter	**Value**	**Parameter**	**Value**
Delivery costs	8% of the equipment purchased cost	***Equipment erection***	25% of the equipment delivered cost
Piping	20% of the equipment delivered cost	***Instrumentation and control***	10% of the equipment delivered cost
Electrical	10% of the equipment delivered cost	***Utility cost***	10% of the equipment delivered cost
Off-sites	20% of the equipment delivered cost	***Buildings***	20% of the equipment delivered cost
Site preparation	10% of the equipment delivered cost	***Land***	4% of the equipment delivered cost
Engineering and supervision	15% of the total direct costs	***Construction overhead***	6% of the total direct costs
Project contingency	5% of the fixed capital investment	***Working Capital***	5% of the total capital investment
Start-up costs	8% of the fixed capital investments	***Heat pump installation cost***	20% of equipment cost
OPEX[1] assumptions			
Parameter	**Value**	**Parameter**	**Value**
Number of operators per shift	2 (based on 1 process unit)	***Number of shifts***	5
Supervision and engineering	15% of labor wages	***Operating supplies and services***	5% of labor wages
Laboratory expenses	10% of labor wages	***Payroll charges***	30% of total labor wages and supervision
Maintenance wages	3.5% of fixed capital investment (excluding half)	***Maintenance, supervision, and engineering***	25% of maintenance wages
Material supplies	100% of maintenance wages	***Maintenance overhead***	5% of maintenance wages
Plant overhead	7.1% of TWSE[1]	***Mechanical department services***	2.4% of TWSE
Employee relations department	5.9% of TWSE	***Business services***	7.4% of TWSE
Property insurance and taxes, $/yr	2% of fixed capital investment	***Sale expenses***	3% of sales
Research and development	5% of sales	***Administrative expenses***	3% of sales
Economic assumptions			
Parameter	**Value**	**Parameter**	**Value**
Plant lifetime, year	30	***Loan lifetime, year***	15
Capacity factor	90%	***Operation factor***	90%
Construction time, year	3	***Operation time, hr/year***	8760
Loan interest rate	5%	***Federal and provincial tax rate***	26%
Debt ratio	40%	***Internal rate of return***	10%

[1] Abbreviations: CEPCI, chemical engineering plant cost index; CAPEX, capital expenditure; OPEX, operational expenditure; TWSE, total operating and maintenance wages, supervision and engineering expenses.

Chapter 4 Results and discussion

Chapter 4 of this thesis presents the results of physical and chemical absorption and the power demand results of all the designs. This chapter critically examines and compares the base cases of each method with the novel designs developed. The effect of all the modifications implemented will be presented and discussed to investigate their feasibility. The chapter begins with the discussion of the results of physical and chemical designs separately based on material flows for physical and temperature and coefficient of performance (COP) for chemical. Furthermore, a comparison between the two methods based on power demand will be conducted to determine the best design.

4.1 Physical absorption results

A summary of the process simulation results of the depressurization unit for each type of design is provided in Table 6. It can be seen that adding H_2 increased the amount of CO_2 released in the gas stream of the first flash tank (Stream 3) by 51.6% and 41.7% for flue and syngas respectively. As a result, the amount of CO_2 going into the second tank (Stream 4) decreased by 47.8% for flue gas and 52.6% for syngas, which led to a 97% decrease in the amount of CO_2 needed to be released in the second tank gas stream (Stream 5) for both feedstocks. The same effect can be seen for the third tank, the amount of CO_2 released (Stream 7) decreases by 74% for both gases. As discussed in section 2.5.2, this change is due to the lower partial pressure of CO_2 in the tank caused by the H_2 injection.

Table 6: Simulation results of physical absorption

	CO_2 Mass flow rate kg/hr						
Design/Stream	**Stream 1**	**Stream 2**	**Stream 3**	**Stream 4**	**Stream 5**	**Stream 6**	**Stream 7**
Flue gas							
Conventional	78823	35986.5	42836.3	22235	13751.5	7885.2	14349.8
Design with H_2	78823	11997.7	66825.2	11610	387.7	7885.2	3724.8
Syngas							
Conventional	50338	18854.6	31483.9	12175.5	6679	3436.5	8739.1
Design with H_2	50337	5952.6	44384.3	5769.8	182.7	3434.9	2335

4.2 Chemical absorption results

As mentioned earlier, the key impact of adding H_2 to the stripper is reducing the temperature difference between the condenser and the reboiler [38]. As it can be seen in Table 7, by adding H_2 the temperature of the condenser output changes from 79°C to 82°C for flue gas and from 100°C to 87°C for syngas cases, while the reboiler output temperature is reduced from 143°C to 140°C for flue gas and 137°C to 122°C for syngas cases respectively. This leads to a decrease in temperature difference for flue gas from 64°C to 58°C and from 37°C to 35°C for syngas. Since the coefficient of performance of heat pumps increases with a decreasing temperature difference; electrifying the energy demand of the reboiler by using a heat pump that is connected to the condenser as the heat source becomes more efficient. The simulation results showed that the COP increased by 13% for flue gas and 5% for syngas. However, this increase in COP does not offset the increase in the required energy needed for the reboiler which will be discussed in the next section.

Table 7: Simulation results of chemical absorption

Design	Flue Gas Conventional with HP	 Design with HP/H_2	Syngas Conventional with HP	 Design with HP/H_2
Output Temperature (°C)				
T_{con}	79	82	100	87
T_{reb}	143	140	137	122
Diff	64	58	37	35
COP	3.9	4.4	6	6.3

4.3 Power demand comparison

The total power demand is divided into two categories: compression of feed and heating demand. The physical absorption only requires electricity to run the compressors, while the chemical absorption requires heat for the reboiler and electrical for the compressors and pumps. For consistency, in design cases where the heat pump is not used, it is assumed that an electric boiler with an efficiency of 95% [72] is used to supply the energy demand of the reboiler. Hence, as shown in Figure 12, two different electricity consumption numbers are reported: Total electrical power demand without a heat pump (Blue) for conventional chemical designs and modified designs with H_2 injection but without the heat pump (HP), and Total electrical power demand (Gray) for physical absorption designs and chemical absorption designs with HP added. It can be seen that, in the chemical solvent designs, adding H_2 increases the reboiler duty which leads to a higher total electrical power demand of 5.52 and 5.37 MJ/kgCO_2 for flue and syngas in comparison to 4.17 and 4.65 MJ/kgCO_2. Furthermore, adding the heat pump system significantly decreases the power demand for such designs. Nevertheless, it's more effective in the base cases where the power demand decreases by 67% for flue gas and 82% for syngas. It should be noted that the H_2 could be pre-heated before being injected using the excess heat from other sections of the CCU process (e.g. the excess heat of the CO2 conversion reactor), which could have a positive effect and help in reducing the reboiler duty. However, this effect was not considered in this study since it needs to integrate the simulation models of other sections to determine their excess heat in those sections and the impact of heat integration on the overall performance of the CCU process. On the other hand, the lowest power demand between all the designs (physical and chemical), goes for the physical case with H_2, with 1.36 MJ/kgCO_2 and 0.17 MJ/kgCO_2 required for flue and syngas respectively. This significant reduction in the physical absorption process is because the flash tanks can operate at a higher pressure due to the increase in CO_2 stripping after adding H_2 and the elimination of the need to compress H_2 before mixing it with the product. As a result, a lower product compression is needed after the flash tanks.

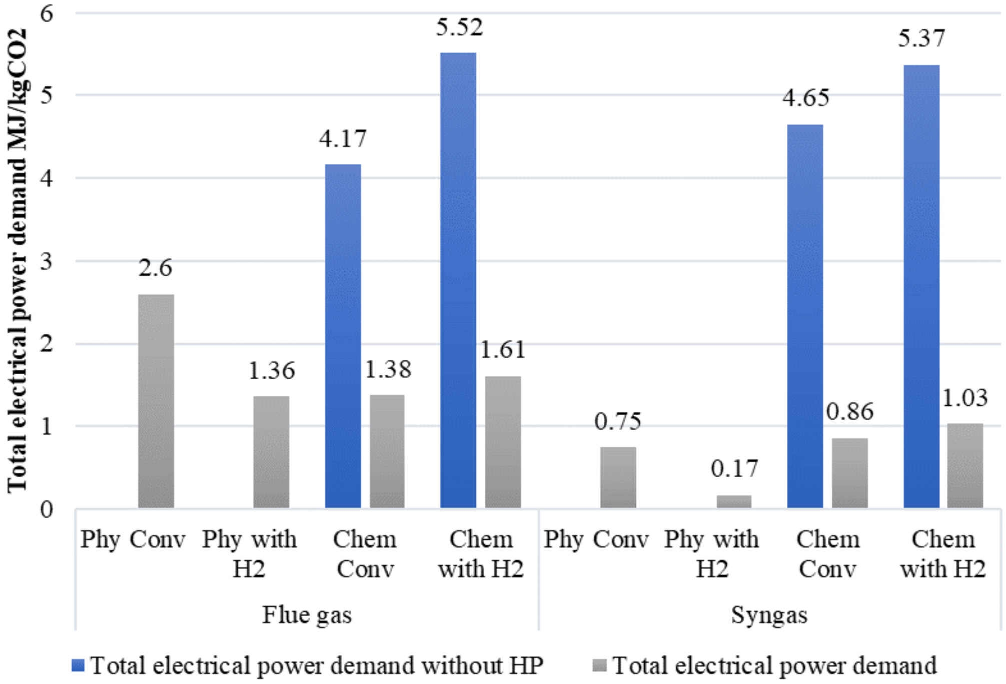

Figure 12: Total electrical power demand of each design

Chapter 5 Economic analysis

This chapter focuses on the financial viability of the novel designs developed in comparison to the conventional ones. In this chapter, detailed results for cost are presented. The comparison between the designs will be conducted based on capital costs, operating costs and equipment cost distribution. Eventually, the capture cost of CO_2 for each model is estimated based on the total annual cost (CAPEX and OPEX) in terms of tonne CO_2. By examining these economic aspects, this chapter aims to provide a better understanding of the financial feasibility and competitiveness of the novel designs introduced.

5.1 Economic results summary

Table 8 summarizes the simulation and economic analysis results of all different design scenarios based on the economic assumptions in Table 5. Details of the cost analysis results are provided in the Supplementary document. For the CO_2 capture cost, two prices are shown. One where the design includes the compression cost of the outlet H_2/CO_2 mixture, and a second which excludes the compression cost. The first price is introduced to have a consistent design basis for our analysis and to better understand the actual optimal process, since the final pressure of the gas mixture is different in each design. The second cost results, where the compressor's cost is excluded, can be used for comparison with other works. The cost of the product compression, which is around 27$/tonne CO_2, is calculated based on the compression of the captured CO_2 and hydrogen to 4 bar which is one bar above the highest final pressure obtained among the designs. In addition, having two feedstocks and different parameters the final CO_2 capture rate differs between the designs. Moreover, as mentioned earlier, the cost for the H_2 injected in the processes is not included in the final cost calculations since the H_2 ratio used is already needed in the production process. This H_2 stream is being repurposed and used in the carbon capture process to obtain the H_2/CO_2 stream before being sent to the CO_2 conversion unit.

Table 8: Techno-economic analysis results of each design using Aspen Economic Analyzer.

	Flue gas				**Syngas**				
	Physical absorption		**Chemical absorption**		**Physical absorption**		**Chemical absorption**		
Design	**Conv**	**Design with H_2**	**Conv with HP**	**Design with HP/H_2**	**Conv**	**Design with H_2**	**Conv with HP**	**Design with HP/H_2**	**Conv**
Inputs									
Flue gas, tonne/hr	250	250	250	250	0	0	0	0	0
Syngas, tonne/hr	0	0	0	0	100	100	100	100	100
DEPG, tonne/hr	2968	3332	0	0	1581	1784	0	0	0
DGA, tonne/hr	0	0	404.77	404.77	0	0	345.9	345.9	345.9
Electricity, MWhr/year	2.13×10^5	2.0×10^5	2.14×10^5	2.41×10^5	3.93×10^4	3.35×10^4	9.64×10^4	1.14×10^5	4.01×10^5
Outputs									
CO_2 capture rate, tonne/year	5.03×10^5	5.03×10^5	5.06×10^5	5.06×10^5	3.33×10^5	3.33×10^5	3.31×10^5	3.31×10^5	3.31×10^5
Waster water, tonne/year	7.43×10^4	8.17×10^4	5.41×10^4	5.31×10^4	7.79×10^3	3.03×10^4	6.63×10^3	6.63×10^3	6.63×10^3
CAPEX, $Million									
Solvent	21.4	23.9	0.83	0.83	11.4	12.8	0.7	0.7	0.7
Total equipment cost	94.6	81.4	66.9	83.9	26.1	16.8	45.3	60.7	34.2
Utility cost	8.8	7.5	5.4	6.4	2.2	1.3	3.3	4.7	2.6
Off-sites	17.6	15.1	10.9	12.8	4.5	2.7	6.7	9.5	5.1
Buildings	17.6	15.1	10.9	12.8	4.5	2.7	6.7	9.5	5.1
Site preparation	8.8	7.5	5.4	6.4	2.2	1.3	3.3	4.7	2.6
Land	3.5	3.0	2.2	2.5	0.9	0.5	1.3	1.9	1.0
Engineering and supervision	22.6	19.5	15.3	18.7	6.1	3.8	10.0	13.6	7.6
Construction overhead	9.0	7.8	6.1	7.5	2.4	1.5	4.0	5.5	3.0
Project contingency	9.6	8.3	6.5	7.9	2.6	1.6	4.2	5.8	3.2
Working capital	12.1	10.8	7.4	9.1	3.6	2.6	4.9	6.6	3.7
Start-up costs	17.1	15.1	10.4	12.8	5.0	3.6	6.9	9.3	5.2
Total capital investment, $Million	**242.7**	**215**	**148.23**	**181.63**	**71.5**	**51.2**	**97.3**	**132.5**	**74**

OPEX, $Million/year									
Raw materials	0.04	0.04	5.9	6.0	0.004	0.015	0.12	0.13	0.12
Electricity	9.6	9.4	9.6	10.8	1.8	1.5	4.3	5.1	18.0
Operating labour costs	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89	0.89
Maintenance costs	16.9	14.9	10.3	12.6	5.0	3.6	6.8	9.2	5.2
Overhead costs	2.2	2.0	1.4	1.7	0.8	0.6	0.9	1.3	0.8
Property insurance and taxes	4.3	3.8	2.6	3.2	1.3	0.9	1.7	2.3	1.3
Total annual expenses, $Million/year	**33.93**	**31.03**	**30.7**	**35.2**	**9.8**	**7.5**	**14.7**	**18.9**	**26.3**

5.2 Evaluation and comparison of capital and operating cost

As verified in the previous sections, in the chemical absorption the H_2 injection with the heat pump does not improve the power consumption of the process. However, the heat pump addition to the conventional design without H_2 improves the design by lowering the reboiler demand. This improvement is verified by the 18% and 12.7% reduction in CAPEX and OPEX respectively for flue gas, and 26.7%, 22.9% respectively for syngas (as shown in Figure 13). In addition, if the heat pump is removed from the conventional design, the OPEX increases by 74.2% due to the higher heat demand needed that is not supplied by the heat pump.

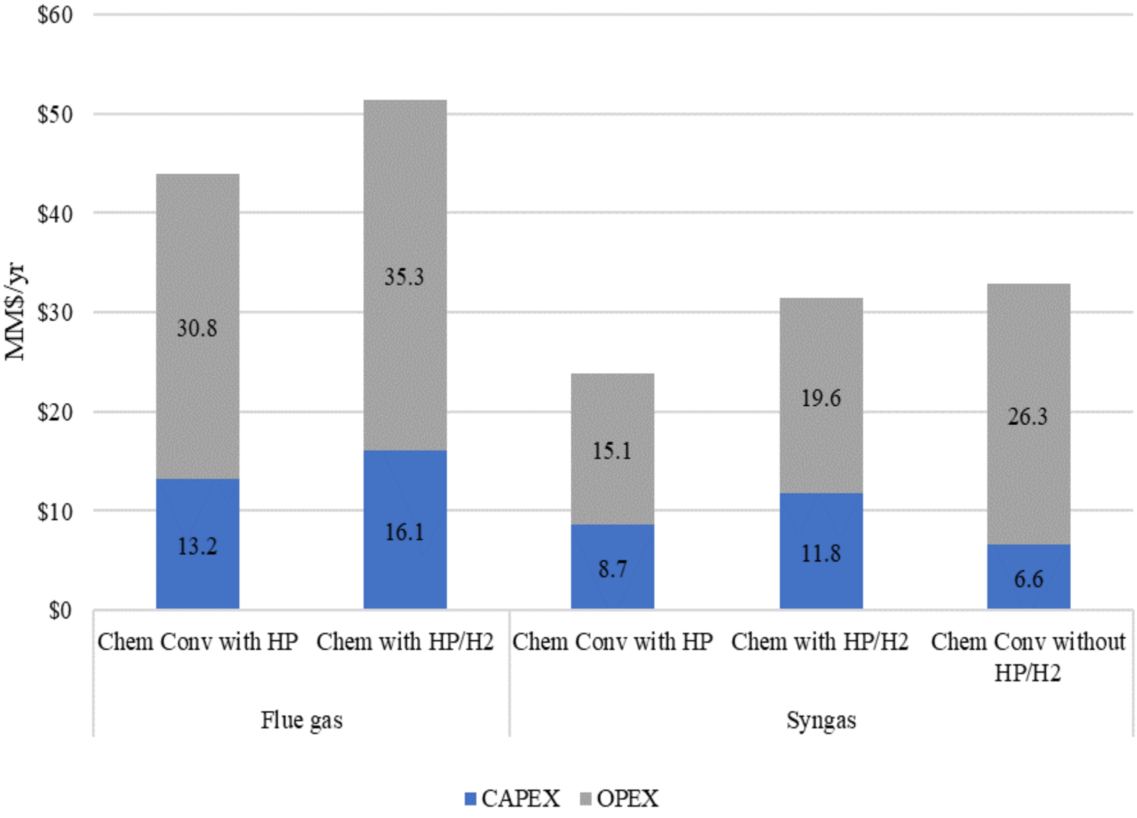

Figure 13: Total annualized CO2 capture costs of different chemical absorption design scenarios.

On the other hand, adding H_2 to the physical absorption process leads to the lowest power consumption. For an additional assessment, the equipment cost distribution is calculated for each physical absorption design. Figure 14 shows that the compressors cost accounts for the highest percentage of the total equipment cost. After adding H_2 this cost is decreased by 5% for flue gas and 18% for syngas due to the lower compression demand needed.

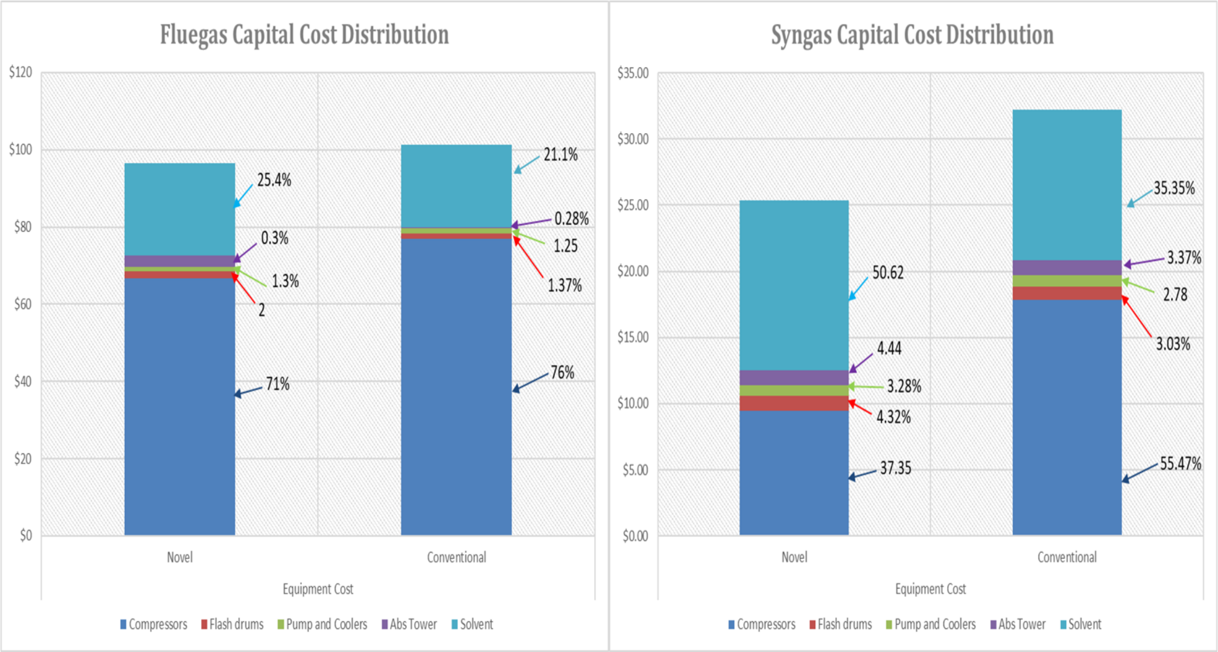

Figure 14: Physical absorption capital cost distribution

Overall, both assessments prove that adding H_2 is only beneficial in the physical absorption case, while for the chemical absorption, the heat pump addition is more helpful.

5.3 CO_2 capture cost

The total CO_2 capture costs of all pathways, with and without the compression cost are shown in Figure 15. As expected, routes with lower power consumption achieve lower capture costs. As seen in Figure 15, physical absorption with H_2 for syngas has the lowest capture cost, with 29.43 \$/tonne CO_2. However, this does not apply to the flue gas case, mainly due to the difference in the feed compression between physical and chemical designs, since in syngas there's no need for feed compression as mentioned in previous sections. As a result, chemical absorption is the cheapest for flue gas with 70.38 \$/tonneCO_2. Furthermore, if the outlet \$27 compression cost is deducted, the final cost is reduced to \$2.4 for syngas. In conclusion, the optimal design in terms of power consumption and capture cost is the physical absorption with H_2 injection for syngas, and chemical absorption with heat pump only for flue gas.

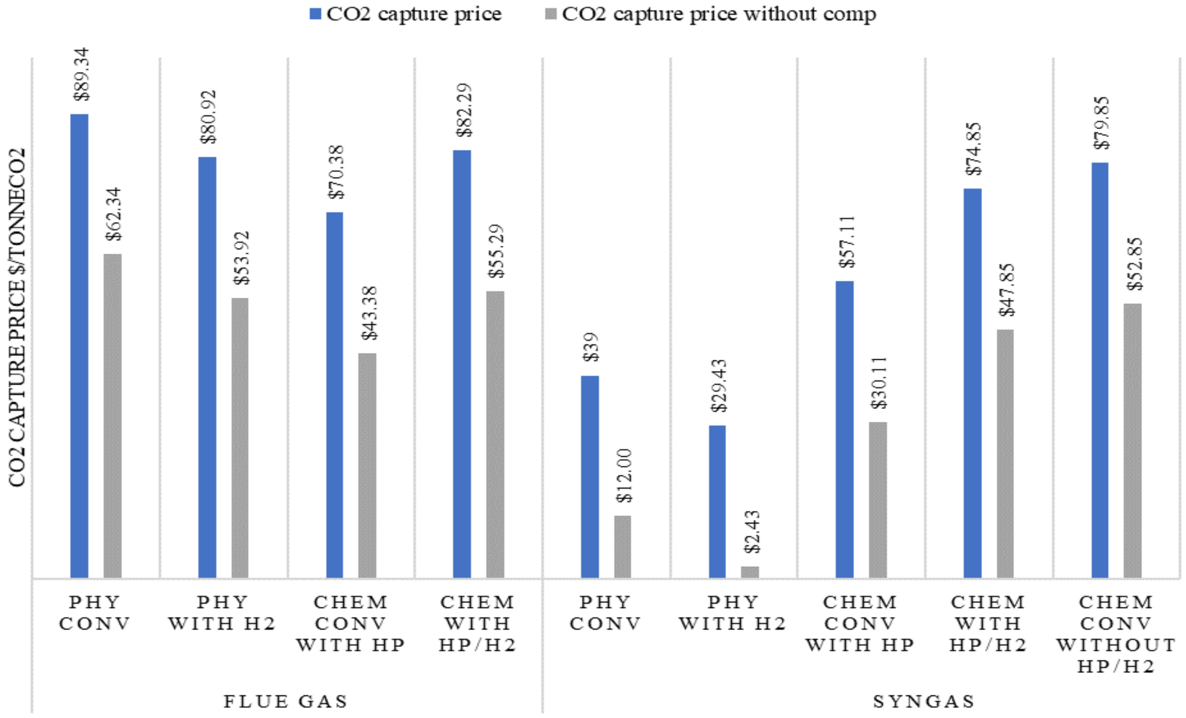

Figure 15: CO_2 capture cost of all pathways

5.4 Revamping the CO_2 capture process

All previous design cases were based on the electrification of the CO_2 capture unit, using an electric boiler, heat pump, heat pump with hydrogen injection, converting to physical absorption, or physical absorption with hydrogen injection. However, in the conventional CO_2 capture processes, using steam produced by a natural gas boiler is still the most common approach, which is mainly due to the low price of natural gas. Hence, in order to compare the energy consumption of each route, in this section, we compared the results of the syngas CO_2 capture process with a natural gas boiler, with those of various electrified approaches. Figure 16 illustrates a simplified schematic of the natural gas boiler that produces low-pressure steam for the reboiler of the syngas CO_2 capture unit (Figure 10).

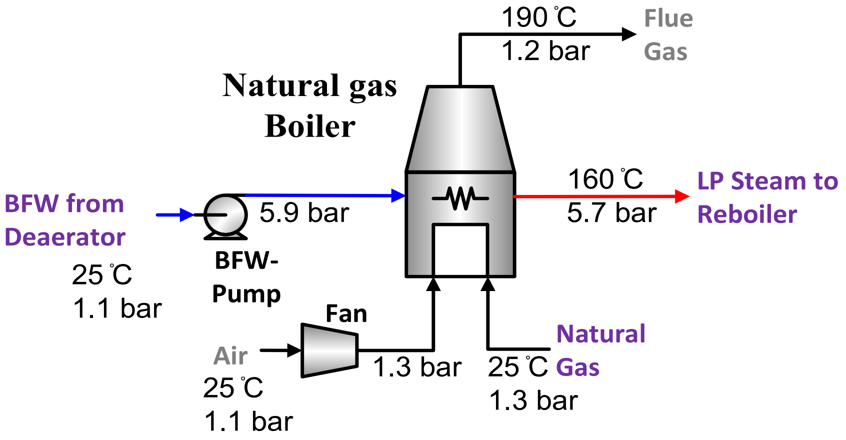

Figure 16: Natural gas boiler to supply the steam demand of the syngas CO2 capture process.

A summary of the simulation results is shown in Table 9. As previously shown (Figure 15), the chemical absorption with a heat pump and physical with hydrogen injection can provide a much lower CO_2 capture cost than the conventional process (with an electric boiler). Hence, for the sake of brevity, we only focus on the total energy demand and cost of each route and compare them with those of the conventional process (with a natural gas boiler). The natural gas and electricity prices are assumed to be \$2.4 per GJ [73] and 2.7 cents per kWh [69], accordingly. Furthermore, it is assumed that the GHG emissions for natural gas production and transportation is 9 kg $CO2_{eq.}$ per GJ [3]. The default value for the GHG emissions of electricity generation is assumed to be 21.2 g $CO2_{eq.}$ per kWh electricity, which is similar to the carbon intensity of the grid in Quebec, Canada [3]. However, the carbon intensity of the grid is varied as a sensitivity analysis parameter, as it will be discussed later.

Table 9: Summary of the energy consumption and utility cost of the syngas CO_2 capture unit based on natural gas boiler.

Captured CO_2	***46942.5***	***kg/h***
Required Reboiler Duty	***49.6***	***Gcal/h***
	4.4	MJ/kg CO2
Natural Gas Consumption	***5664.1***	***kg/h***
	273852	MJ/h, LHV
Electricity Demand of Boiler (Pump and Blower)	***2380.4***	***MJ/h***
Electricity Demand of CO_2 capture and compression	***15780.3***	***MJ/h***
Total Electricity Demand	**5044.6**	**kWh/h**
CO_2 emissions from Reboiler	15216.7	kg/h
CO_2 emissions for Natural Gas supply	2464.7	kg/h
CO_2 emissions for the electricity demand of the boiler	6.6	kg/h
Net CO_2 capture	***29254.6***	***kg/h***
Utility Cost before considering the GHG emissions of natural gas and electricity production	24.7	\$/tonnes CO_2
Utility Cost after considering the GHG emissions of natural gas and electricity production	39.6	\$/tonnes CO_2

As shown in Table 9, the utility (natural gas and electricity) cost for the conventional CO_2 capture process from syngas using a natural gas boiler can be as low as \$24.7 per tonne of CO_2. This utility cost includes the CO_2 and downstream hydrogen compression costs, as described previously. Although this is a common approach that is used in many works to report the CO_2 capture cost, it should be noted that the actual capture cost is higher when the upstream GHG emissions for electricity generation and natural gas production are included. In our syngas case, the CO_2 capture cost increases by 60%, to \$39.6 per tonne of CO_2.

Further investigation is conducted by varying the electricity price and also the carbon intensity of the grid. In Figure 17, the default carbon intensity of 21.2 g CO_2 per kWh is used, while in Figure 18, the carbon intensity is increased to 683.5 g CO_2 per kWh, which is similar to the GHG emissions of electricity generation in Alberta, Canada [3].

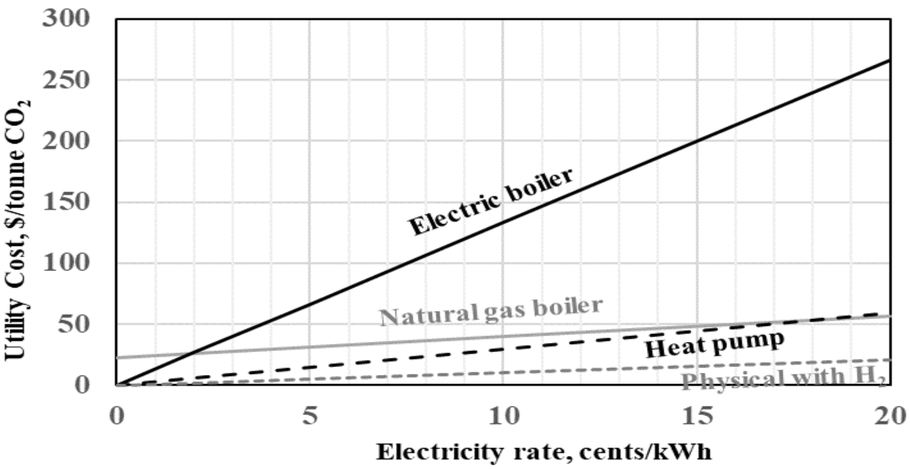

Figure 17: Impact of electricity price on the utility (natural gas and electricity) cost of different syngas CO_2 capture processes (carbon intensity of electricity=21.2 g CO_2 per kWh).

As it can be seen from Figure 17, when renewable electricity is used, the natural gas boiler route becomes a less expensive option than the electric boiler only when the electricity price is more than 2 cents per kWh. The chemical absorption with heat pump has a less expensive energy demand compared to the conventional process with natural gas boiler, unless the electricity price is more than 18 cents per kWh. Moreover, the physical absorption with hydrogen injection is always less expensive due to its low electricity demand.

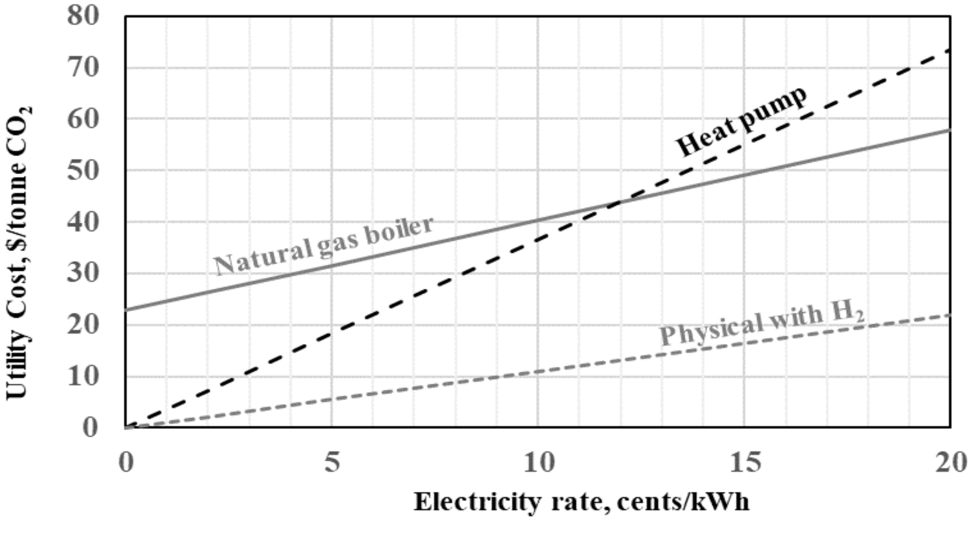

Figure 18: Impact of electricity price on the utility (natural gas and electricity) cost of different syngas CO_2 capture processes (carbon intensity of electricity=683.5 g CO_2 per kWh).

As shown in Figure 18, using electricity from non-renewable sources increases the cost of all options, with more impact on electrified routes. The electric boiler option is not shown as its cost was much higher than other routes. The maximum electricity prices to have the heat-pump route is still relatively high, around 12 cents per kWh.

The impact of natural gas price is the other element that is investigated in Figure 19 and Figure 20. Results show that both chemical with heat-pump and physical with hydrogen injection can provide lower utility cost, when the natural gas price is more than $1.4 to $2.0 per GJ, depending on the carbon intensity of grid. These values are much lower than the market price of natural gas as the trend is illustrated in a few different years.

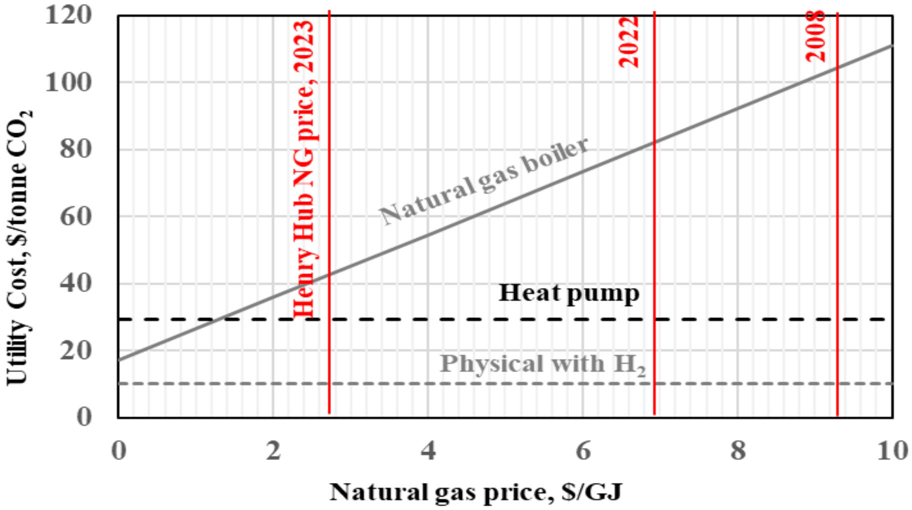

Figure 19: Impact of natural gas price on the utility (natural gas and electricity) cost of different syngas CO_2 capture processes (carbon intensity of electricity=21.2 g CO_2 per kWh).

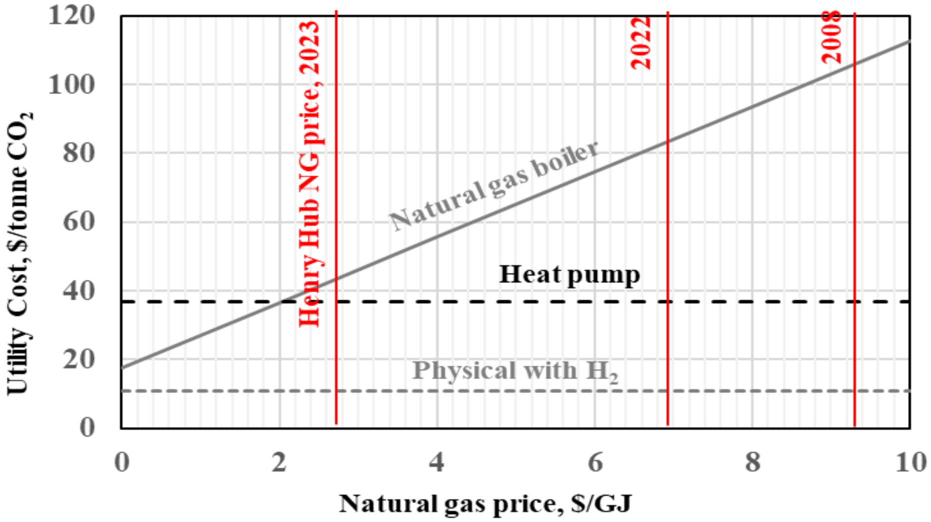

Figure 20: Impact of natural gas price on the utility (natural gas and electricity) cost of different syngas CO_2 capture processes (carbon intensity of electricity=683.5 g CO_2 per kWh).

Hence, it can be seen that revamping the conventional CO_2 capture unit which is typically used in various chemical plants such as natural gas treatment, ammonia and methanol plants, with effective electrification methods such as heat-pump or the proposed physical absorption with hydrogen injection can significantly reduce the utility costs, even when non-renewable electricity is incorporated.

Chapter 6 Conclusion

The main goal of this thesis was to design a novel CO_2 capture process to reduce the energy consumption and cost of existing and new CO_2 capture units for various applications. We investigated different CO_2 chemical and physical absorption processes, calculated their power consumption, and conducted an economic analysis to identify the optimal process with the lowest power consumption and CO_2 capture cost. Several designs were simulated for each process with two different feed gases, flue gas from a cement kiln and syngas from an ammonia synthesis plant. Results showed that the H_2 addition with the heat pump in the chemical absorption process could not be considered an optimal design since it increases the total electrical power demand by 17% and 20% for flue and syngas in comparison to the designs with only the heat pump modification (1.61 and 1.03 MJ/kgCO$_2$ in comparison to 1.38 and 0.86 MJ/kgCO$_2$) . In addition, the capture cost is 17% higher for flue gas and 31% for syngas (\$82.29 and \$74.85 in comparison to \$70.38 and \$57.11 per tonne CO_2). On the other hand, H_2 injection in the physical process achieves substantial power and cost savings by increasing the CO_2 stripping in the flash tanks and increasing the pressure needed to run them, thereby decreasing the power consumption by 48% (1.36 MJ/kgCO$_2$ from 2.6 MJ/kgCO$_2$) and the capture cost of CO_2 by 10% (\$80.92 from \$89.34 per tonne CO_2) for flue gas, and 77% (0.17 MJ/kgCO$_2$ from 0.75 MJ/kgCO$_2$), 25% (\$29.43 from \$39 per tonne CO_2) for syngas. However, in the flue gas case due the need for a higher feed compression in comparison to syngas where no compression is needed, the capture cost is lower in the conventional chemical design with the heat pump modification (\$70.38 per tonne CO_2) in comparison to the physical design with H_2 (\$80.92 per tonne CO_2).

6.1 Study Limitations:

The limitations encountered during this study are as follows:

- The minimal amount of references on physical absorption with a flue gas stream as a feed
- The limitation on the ability of simulating the chemical absorption process using a rate-based model instead of equilibrium model due to the lack of data and references.
- The difficulty of doing a deeper analysis on the H_2 injection in terms of temperature and location of injection due to the limitations of time and the simulation software.

6.2 Future works

As the next step of this work, a lifecycle assessment will be conducted on all the pathways presented previously, as well as their incorporation into the CCU pathways. These pathways must undergo thorough investigation, encompassing their environmental impacts and eco-efficiency, especially in terms of the source of electricity used to supply power for the processes since the main aspect of the modification implemented was the electrification of the capture process, in addition to the type of solvents used to achieve the carbon capture. These inquiries have the possibility to significantly enhance our understanding of and ability to implement these modifications, all while promoting sustainable and environmentally friendly methods for carbon capture. Furthermore, the H_2 injection in the chemical absorption case requires further investigation in terms of the effect of the temperature of the H_2 stream and its heat integration with other units of the CCU process, on the reboiler duty of the stripper, in addition to the effect of the stage at which the hydrogen is injected.

References

[1] K. Calvin, D. Dasgupta, G. Krinner, A. Mukherji, P.W. Thorne, C. Trisos, J. Romero, P. Aldunce, K. Barrett, G. Blanco, W.W.L. Cheung, S. Connors, F. Denton, A. Diongue-Niang, D. Dodman, M. Garschagen, O. Geden, B. Hayward, C. Jones, F. Jotzo, T. Krug, R. Lasco, Y.-Y. Lee, V. Masson-Delmotte, M. Meinshausen, K. Mintenbeck, A. Mokssit, F.E.L. Otto, M. Pathak, A. Pirani, E. Poloczanska, H.-O. Pörtner, A. Revi, D.C. Roberts, J. Roy, A.C. Ruane, J. Skea, P.R. Shukla, R. Slade, A. Slangen, Y. Sokona, A.A. Sörensson, M. Tignor, D. Van Vuuren, Y.-M. Wei, H. Winkler, P. Zhai, Z. Zommers, J.-C. Hourcade, F.X. Johnson, S. Pachauri, N.P. Simpson, C. Singh, A. Thomas, E. Totin, P. Arias, M. Bustamante, I. Elgizouli, G. Flato, M. Howden, C. Méndez-Vallejo, J.J. Pereira, R. Pichs-Madruga, S.K. Rose, Y. Saheb, R. Sánchez Rodríguez, D. Ürge-Vorsatz, C. Xiao, N. Yassaa, A. Alegría, K. Armour, B. Bednar-Friedl, K. Blok, G. Cissé, F. Dentener, S. Eriksen, E. Fischer, G. Garner, C. Guivarch, M. Haasnoot, G. Hansen, M. Hauser, E. Hawkins, T. Hermans, R. Kopp, N. Leprince-Ringuet, J. Lewis, D. Ley, C. Ludden, L. Niamir, Z. Nicholls, S. Some, S. Szopa, B. Trewin, K.-I. Van Der Wijst, G. Winter, M. Witting, A. Birt, M. Ha, J. Romero, J. Kim, E.F. Haites, Y. Jung, R. Stavins, A. Birt, M. Ha, D.J.A. Orendain, L. Ignon, S. Park, Y. Park, A. Reisinger, D. Cammaramo, A. Fischlin, J.S. Fuglestvedt, G. Hansen, C. Ludden, V. Masson-Delmotte, J.B.R. Matthews, K. Mintenbeck, A. Pirani, E. Poloczanska, N. Leprince-Ringuet, C. Péan, IPCC, 2023: Climate Change 2023: Synthesis Report. Contribution of Working Groups I, II and III to the Sixth Assessment Report of the Intergovernmental Panel on Climate Change [Core Writing Team, H. Lee and J. Romero (eds.)]. IPCC, Geneva, Switzerland., First, Intergovernmental Panel on Climate Change (IPCC), 2023. https://doi.org/10.59327/IPCC/AR6-9789291691647.

[2] J.G. Vitillo, M.D. Eisaman, E.S.P. Aradóttir, F. Passarini, T. Wang, S.W. Sheehan, The role of carbon capture, utilization, and storage for economic pathways that limit global warming to below 1.5°C, iScience 25 (2022) 104237. https://doi.org/10.1016/j.isci.2022.104237.

[3] Y. Khojasteh-Salkuyeh, O. Ashrafi, E. Mostafavi, P. Navarri, CO2 utilization for methanol production; Part I: Process design and life cycle GHG assessment of different pathways, Journal of CO2 Utilization 50 (2021) 101608. https://doi.org/10.1016/j.jcou.2021.101608.

[4] N.S. Sifat, Y. Haseli, A Critical Review of CO2 Capture Technologies and Prospects for Clean Power Generation, Energies 12 (2019) 4143. https://doi.org/10.3390/en12214143.

[5] T. N.Borhani, M. Wang, Role of solvents in CO2 capture processes: The review of selection and design methods, Renewable and Sustainable Energy Reviews 114 (2019) 109299. https://doi.org/10.1016/j.rser.2019.109299.

[6] P. Madejski, K. Chmiel, N. Subramanian, T. Kuś, Methods and Techniques for CO2 Capture: Review of Potential Solutions and Applications in Modern Energy Technologies, Energies 15 (2022) 887. https://doi.org/10.3390/en15030887.

[7] W.L. Theo, J.S. Lim, H. Hashim, A.A. Mustaffa, W.S. Ho, Review of pre-combustion capture and ionic liquid in carbon capture and storage, Applied Energy 183 (2016) 1633–1663. https://doi.org/10.1016/j.apenergy.2016.09.103.

[8] D. Jansen, M. Gazzani, G. Manzolini, E. van Dijk, M. Carbo, Pre-combustion CO2 capture, International Journal of Greenhouse Gas Control 40 (2015) 167–187. https://doi.org/10.1016/j.ijggc.2015.05.028.

[9] C.F. Martín, E. Stöckel, R. Clowes, D.J. Adams, A.I. Cooper, J.J. Pis, F. Rubiera, C. Pevida, Hypercrosslinked organic polymer networks as potential adsorbents for pre-combustion CO2 capture, J. Mater. Chem. 21 (2011) 5475–5483. https://doi.org/10.1039/C0JM03534C.

[10] Z. Dai, L. Deng, Membrane absorption using ionic liquid for pre-combustion CO2 capture at elevated pressure and temperature, International Journal of Greenhouse Gas Control 54 (2016) 59–69. https://doi.org/10.1016/j.ijggc.2016.09.001.

[11] S. Seepana, S. Jayanti, Steam-moderated oxy-fuel combustion, Energy Conversion and Management 51 (2010) 1981–1988. https://doi.org/10.1016/j.enconman.2010.02.031.

[12] P. Edge, M. Gharebaghi, R. Irons, R. Porter, R.T.J. Porter, M. Pourkashanian, D. Smith, P. Stephenson, A. Williams, Combustion modelling opportunities and challenges for oxy-coal carbon capture technology, Chemical Engineering Research and Design 89 (2011) 1470–1493. https://doi.org/10.1016/j.cherd.2010.11.010.

[13] F. Wu, M.D. Argyle, P.A. Dellenback, M. Fan, Progress in O2 separation for oxy-fuel combustion–A promising way for cost-effective CO2 capture: A review, Progress in Energy and Combustion Science 67 (2018) 188–205. https://doi.org/10.1016/j.pecs.2018.01.004.

[14] F. Climent Barba, G. Martínez-Denegri Sánchez, B. Soler Seguí, H. Gohari Darabkhani, E.J. Anthony, A technical evaluation, performance analysis and risk assessment of multiple novel oxy-turbine power cycles with complete CO2 capture, Journal of Cleaner Production 133 (2016) 971–985. https://doi.org/10.1016/j.jclepro.2016.05.189.

[15] R. Falkenstein-Smith, P. Zeng, J. Ahn, Investigation of oxygen transport membrane reactors for oxy-fuel combustion and carbon capture purposes, Proceedings of the Combustion Institute 36 (2017) 3969–3976. https://doi.org/10.1016/j.proci.2016.09.005.

[16] Y. Wang, L. Zhao, A. Otto, M. Robinius, D. Stolten, A Review of Post-combustion CO2 Capture Technologies from Coal-fired Power Plants, Energy Procedia 114 (2017) 650–665. https://doi.org/10.1016/j.egypro.2017.03.1209.

[17] A. Malek, K. Li, W.K. Teo, Modeling of Microporous Hollow Fiber Membrane Modules Operated under Partially Wetted Conditions, Ind. Eng. Chem. Res. 36 (1997) 784–793. https://doi.org/10.1021/ie960529y.

[18] Z. Zhang, Y. Yan, L. Zhang, Y. Chen, S. Ju, CFD investigation of CO2 capture by methyldiethanolamine and 2-(1-piperazinyl)-ethylamine in membranes: Part B. Effect of membrane properties, Journal of Natural Gas Science and Engineering 19 (2014) 311–316. https://doi.org/10.1016/j.jngse.2014.05.023.

[19] C. Song, Q. Liu, S. Deng, H. Li, Y. Kitamura, Cryogenic-based CO2 capture technologies: State-of-the-art developments and current challenges, Renewable and Sustainable Energy Reviews 101 (2019) 265–278. https://doi.org/10.1016/j.rser.2018.11.018.

[20] S.A. Rackley, Carbon Capture and Storage, Elsevier, 2010. https://app.knovel.com/hotlink/toc/id:kpCCS00001/carbon-capture-storage/carbon-capture-storage.

[21] S. Lillia, D. Bonalumi, C. Grande, G. Manzolini, A comprehensive modeling of the hybrid temperature electric swing adsorption process for CO2 capture, International Journal of Greenhouse Gas Control 74 (2018) 155–173. https://doi.org/10.1016/j.ijggc.2018.04.012.

[22] C.A. Grande, Advances in Pressure Swing Adsorption for Gas Separation, International Scholarly Research Notices 2012 (2012) e982934. https://doi.org/10.5402/2012/982934.

[23] J.-R. Li, Y. Ma, M.C. McCarthy, J. Sculley, J. Yu, H.-K. Jeong, P.B. Balbuena, H.-C. Zhou, Carbon dioxide capture-related gas adsorption and separation in metal-organic frameworks,

Coordination Chemistry Reviews 255 (2011) 1791–1823. https://doi.org/10.1016/j.ccr.2011.02.012.

[24] A. Agarwal, L.T. Biegler, S.E. Zitney, A superstructure-based optimal synthesis of PSA cycles for post-combustion CO2 capture, AIChE Journal 56 (2010) 1813–1828. https://doi.org/10.1002/aic.12107.

[25] S. Bhavsar, M. Najera, A. More, G. Veser, Chapter 7 - Chemical-looping processes for fuel-flexible combustion and fuel production, in: F. Shi (Ed.), Reactor and Process Design in Sustainable Energy Technology, Elsevier, Amsterdam, 2014: pp. 233–280. https://doi.org/10.1016/B978-0-444-59566-9.00007-7.

[26] M.M. Hossain, H.I. de Lasa, Chemical-looping combustion (CLC) for inherent CO2<math><msub is="true"><mrow is="true"><mi is="true">CO</mi></mrow><mrow is="true"><mn is="true">2</mn></mrow></msub></math> separations—a review, Chemical Engineering Science 63 (2008) 4433–4451. https://doi.org/10.1016/j.ces.2008.05.028.

[27] P. Tilak, M.M. El-Halwagi, Process integration of Calcium Looping with industrial plants for monetizing CO2 into value-added products, Carbon Resources Conversion 1 (2018) 191–199. https://doi.org/10.1016/j.crcon.2018.07.004.

[28] D.P. Hanak, V. Manovic, Calcium looping with supercritical CO2 cycle for decarbonisation of coal-fired power plant, Energy 102 (2016) 343–353. https://doi.org/10.1016/j.energy.2016.02.079.

[29] A. SILABAN, D.P. HARRISON, HIGH TEMPERATURE CAPTURE OF CARBON DIOXIDE: CHARACTERISTICS OF THE REVERSIBLE REACTION BETWEEN CaO(s) and CO2(g), Chemical Engineering Communications 137 (1995) 177–190. https://doi.org/10.1080/00986449508936375.

[30] SCO2 Power Cycles, Energy.Gov (n.d.). https://www.energy.gov/sco2-power-cycles (accessed April 21, 2024).

[31] E. Knapik, P. Kosowski, J. Stopa, Cryogenic liquefaction and separation of CO2 using nitrogen removal unit cold energy, Chemical Engineering Research and Design 131 (2018) 66–79. https://doi.org/10.1016/j.cherd.2017.12.027.

[32] B. Liu, X. Yang, P.-C. Chiang, T. Wang, Energy Consumption Analysis of Cryogenic-membrane Hybrid Process for CO2 Capture from CO2-EOR Extraction Gas, Aerosol and Air Quality Research 20 (2020) 820–832. https://doi.org/10.4209/aaqr.2020.02.0047.

[33] C.A. Scholes, M.T. Ho, D.E. Wiley, G.W. Stevens, S.E. Kentish, Cost competitive membrane—cryogenic post-combustion carbon capture, International Journal of Greenhouse Gas Control 17 (2013) 341–348. https://doi.org/10.1016/j.ijggc.2013.05.017.

[34] M. Gupta, I. Coyle, K. Thambimuthu, CO2 Capture Technologies and Opportunities in Canada, (n.d.).

[35] C.-H. Yu, C.-H. Huang, C.-S. Tan, A Review of CO2 Capture by Absorption and Adsorption, Aerosol Air Qual. Res. 12 (2012) 745–769. https://doi.org/10.4209/aaqr.2012.05.0132.

[36] Y.K. Salkuyeh, M. Mofarahi, Comparison of MEA and DGA performance for CO2 capture under different operational conditions, International Journal of Energy Research 36 (2012) 259–268. https://doi.org/10.1002/er.1812.

[37] M. Voldsund, S.O. Gardarsdottir, E. De Lena, J.-F. Pérez-Calvo, A. Jamali, D. Berstad, C. Fu, M. Romano, S. Roussanaly, R. Anantharaman, H. Hoppe, D. Sutter, M. Mazzotti, M. Gazzani, G. Cinti, K. Jordal, Comparison of Technologies for CO2 Capture from Cement

Production—Part 1: Technical Evaluation, Energies 12 (2019) 559. https://doi.org/10.3390/en12030559.

[38] H. Machida, T. Esaki, T. Yamaguchi, K. Norinaga, Energy-Saving CO2 Capture by H2 Gas Stripping for Integrating CO2 Separation and Conversion Processes, ACS Sustainable Chem. Eng. 8 (2020) 8732–8740. https://doi.org/10.1021/acssuschemeng.0c02459.

[39] A.L. Kohl, R. Nielsen, Gas purification, 5th ed., Gulf Pub., Houston, Tex., 1997. http://site.ebrary.com/id/10251221.

[40] A. Cousins, A. Cottrell, A. Lawson, S. Huang, P.H.M. Feron, Model verification and evaluation of the rich-split process modification at an Australian-based post combustion CO2 capture pilot plant, Greenhouse Gases: Science and Technology 2 (2012) 329–345. https://doi.org/10.1002/ghg.1295.

[41] H.P. Mangalapally, H. Hasse, Pilot plant study of post-combustion carbon dioxide capture by reactive absorption: Methodology, comparison of different structured packings, and comprehensive results for monoethanolamine, Chemical Engineering Research and Design 89 (2011) 1216–1228. https://doi.org/10.1016/j.cherd.2011.01.013.

[42] M. Stec, A. Tatarczuk, L. Więcław-Solny, A. Krótki, T. Spietz, A. Wilk, D. Śpiewak, Demonstration of a post-combustion carbon capture pilot plant using amine-based solvents at the Łaziska Power Plant in Poland, Clean Techn Environ Policy 18 (2016) 151–160. https://doi.org/10.1007/s10098-015-1001-2.

[43] N.-S. Kwak, J.H. Lee, I.Y. Lee, K.R. Jang, J.-G. Shim, A study of the CO2 capture pilot plant by amine absorption, Energy 47 (2012) 41–46. https://doi.org/10.1016/j.energy.2012.07.016.

[44] S.M.N. Hassan, Techno-Economic Study of CO2 Capture Process for Cement Plants, (n.d.).

[45] K.A. Mumford, Y. Wu, K.H. Smith, G.W. Stevens, Review of solvent based carbon-dioxide capture technologies, Front. Chem. Sci. Eng. 9 (2015) 125–141. https://doi.org/10.1007/s11705-015-1514-6.

[46] K. Xin, M. Hashish, I. Roghair, M. van Sint Annaland, Process Simulation and Economic Analysis of Pre-combustion CO2 Capture With Deep Eutectic Solvents, Frontiers in Energy Research 8 (2020). https://www.frontiersin.org/articles/10.3389/fenrg.2020.573267 (accessed November 25, 2023).

[47] S. Cloete, A. Giuffrida, M.C. Romano, A. Zaabout, The swing adsorption reactor cluster for post-combustion CO2 capture from cement plants, Journal of Cleaner Production 223 (2019) 692–703. https://doi.org/10.1016/j.jclepro.2019.03.109.

[48] E.H. Jensen, A. Andreasen, J.K. Jørsboe, M.P. Andersen, M. Hostrup, B. Elmegaard, C. Riber, P.L. Fosbøl, Electrification of amine-based CO2 capture utilizing heat pumps, Carbon Capture Science & Technology 10 (2024) 100154. https://doi.org/10.1016/j.ccst.2023.100154.

[49] A.G. Olabi, A.H. Alami, M. Ayoub, H. Aljaghoub, S. Alasad, A. Inayat, M.A. Abdelkareem, K.-J. Chae, E.T. Sayed, Membrane-based carbon capture: Recent progress, challenges, and their role in achieving the sustainable development goals, Chemosphere 320 (2023) 137996. https://doi.org/10.1016/j.chemosphere.2023.137996.

[50] C.A. Scholes, Chapter 16 - Challenges for CO2 capture by membranes, in: M.R. Rahimpour, M. Farsi, M.A. Makarem (Eds.), Advances in Carbon Capture, Woodhead Publishing, 2020: pp. 357–377. https://doi.org/10.1016/B978-0-12-819657-1.00016-5.

[51] A. Lyngfelt, Chemical Looping Combustion: Status and Development Challenges, Energy & Fuels 34 (2020) 9077–9093. https://doi.org/10.1021/acs.energyfuels.0c01454.

[52] Y. Khojasteh-Salkuyeh, O. Ashrafi, E. Mostafavi, P. Navarri, CO2 utilization for methanol production; Optimal pathways with minimum GHG reduction cost, The Canadian Journal of Chemical Engineering 101 (2023) 5446–5459. https://doi.org/10.1002/cjce.24975.
[53] W.I.F. David, Ammonia: Zero-carbon fertiliser, fuel and energy store, Policy Briefing (2020).
[54] R. de Boer, A. Marina, B. Zühlsdorf, White paper: Strengthening Industrial Heat Pump Innovation - Decarbonizing Industrial Heat, (2020). https://hthp-symposium.org/high-temperature-heat-pumps/white-paper-strengthening-industrial-heat-pump-innovation/ (accessed April 9, 2023).
[55] A.K. Jana, Advances in heat pump assisted distillation column: A review, Energy Conversion and Management 77 (2014) 287–297. https://doi.org/10.1016/j.enconman.2013.09.055.
[56] Z. Fonyo, N. Benkö, Comparison of Various Heat Pump Assisted Distillation Configurations, Chemical Engineering Research and Design 76 (1998) 348–360. https://doi.org/10.1205/026387698524776.
[57] CoolTools, IPU (n.d.). https://www.ipu.dk/products/cooltools/ (accessed July 9, 2024).
[58] Hydrogen, Fuel Cells & Infrastructure Technologies Program, 2005. https://doi.org/10.2172/15011474.
[59] Industrial Heat Pumps: it's time to go electric, World Business Council for Sustainable Development (WBCSD) (n.d.). https://www.wbcsd.org/wgthg (accessed April 14, 2024).
[60] N.R. Canada, Heating and Cooling With a Heat Pump, (2009). https://natural-resources.canada.ca/energy-efficiency/energy-star-canada/about/energy-star-announcements/publications/heating-and-cooling-heat-pump/6817 (accessed April 14, 2024).
[61] Coefficient of Performance (COP) heat pumps, (n.d.). https://www.grundfos.com/ca/learn/research-and-insights/coefficient-of-system-performance (accessed April 14, 2024).
[62] Enrtl-Rk Rate Based Dga Model | PDF | Chemical Equilibrium | Chemical Kinetics, Scribd (n.d.). https://www.scribd.com/document/387116356/Enrtl-rk-Rate-Based-Dga-Model (accessed November 25, 2023).
[63] Aspen Plus DEPG Model | PDF | Vapor | Pressure, Scribd (n.d.). https://www.scribd.com/document/190629661/Aspen-Plus-DEPG-Model (accessed December 10, 2023).
[64] H.E. Ashkanani, R. Wang, W. Shi, N.S. Siefert, R.L. Thompson, K. Smith, J.A. Steckel, I.K. Gamwo, D. Hopkinson, K. Resnik, B.I. Morsi, Levelized Cost of CO2 Captured Using Five Physical Solvents in Pre-combustion Applications, International Journal of Greenhouse Gas Control 101 (2020) 103135. https://doi.org/10.1016/j.ijggc.2020.103135.
[65] K. Atsonios, K.D. Panopoulos, E. Kakaras, Thermocatalytic CO2 hydrogenation for methanol and ethanol production: Process improvements, International Journal of Hydrogen Energy 41 (2016) 792–806. https://doi.org/10.1016/j.ijhydene.2015.12.001.
[66] T.A. Adams II, Y. Khojasteh Salkuyeh, J. Nease, Chapter 6 - Processes and simulations for solvent-based CO2 capture and syngas cleanup, in: F. Shi (Ed.), Reactor and Process Design in Sustainable Energy Technology, Elsevier, Amsterdam, 2014: pp. 163–231. https://doi.org/10.1016/B978-0-444-59566-9.00006-5.
[67] CoolPack, IPU (n.d.). https://www.ipu.dk/products/coolpack/ (accessed March 24, 2024).
[68] W.D. Seider, J.D. Seader, D.R. Lewin, Product & Process Design Principles: Synthesis, Analysis and Evaluation, John Wiley & Sons, 2003.

[69] Electricity Rates Effective April 1, 2023., (n.d.).
[70] Tables Created by BLS : U.S. Bureau of Labor Statistics, (n.d.). https://www.bls.gov/oes/tables.htm#51-0000 (accessed March 11, 2024).
[71] A. Nuchitprasittichai, S. Cremaschi, Sensitivity of amine-based CO2 capture cost: The influences of CO2 concentration in flue gas and utility cost fluctuations, International Journal of Greenhouse Gas Control 13 (2013) 34–43. https://doi.org/10.1016/j.ijggc.2012.12.012.
[72] Furnaces and Boilers, Energy.Gov (n.d.). https://www.energy.gov/energysaver/furnaces-and-boilers (accessed March 25, 2024).
[73] Henry Hub Natural Gas Spot Price (Dollars per Million Btu), (n.d.). https://www.eia.gov/dnav/ng/hist/rngwhhdA.htm (accessed May 2, 2024).

Supplementary Materials

1. Effect of H_2 injection on flash tank liquid outputs temperatures

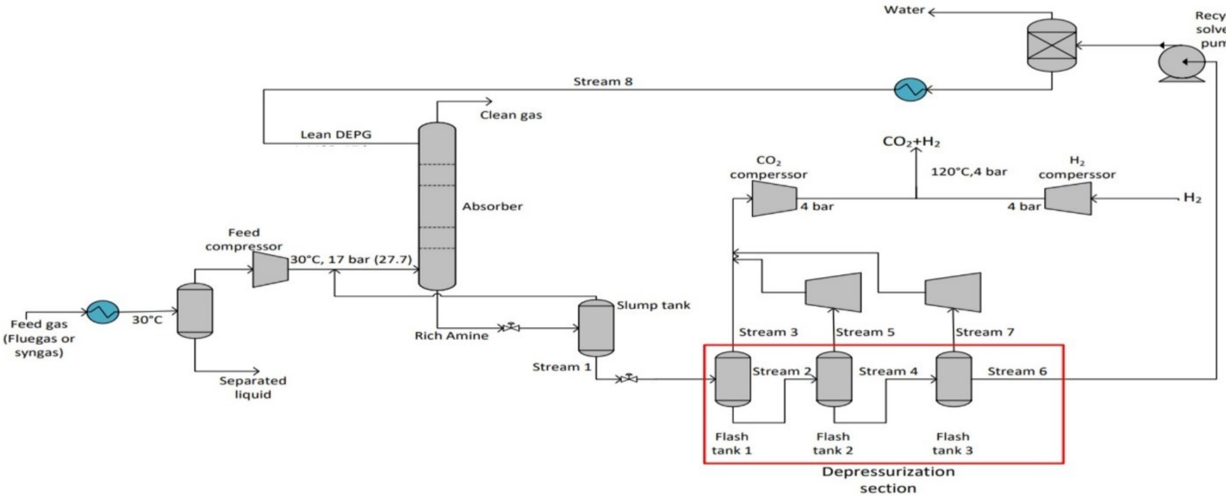

Figure 21: Physical absorption schematic

Table 10: Flash tanks liquid output temperatures

	Flash Tanks Liquid Output Temperature in °C			
Design/Stream	**Stream 2**	**Stream 4**	**Stream 6**	**Stream 8**
Flue gas				
Conventional	5.10	4.18	3.25	-1.11
Novel design with H_2	4.53	4.49	4.26	4.77
Syngas				
Conventional	3.93	3.11	2.03	-1.11
Novel design with H_2	3.70	3.66	3.38	4.21

Table 10 shows the temperatures for the liquid output streams for each flash tank. As the pressure decreases between each flash tank the temperature decreases from 5.10°C in stream 2 to 3.25°C in stream 6 for flue gas (3.93°C to 3.38°C for syngas) before being cooled to the required lean DEPG temperature of -1.11°C. However, as H_2 is injected into the first flash tank the temperature decrease between each tank becomes lower and the overall temperature of the liquid output increases (4.53°C to 4.59°C to 4.26°C for example), in addition pumping the recycle to the required pressure for each design slightly increases the temperature. As a result, it is possible to eliminate the cooling demand and run the lean DEPG at a higher temperature, however a higher DEPG flow is needed. The cost of the increase in flowrate is lower than the cost of the cooling.

2. Equipment costs for all designs

- **Flue gas physical conventional**

Total Equipment Cost	$81,732,880
Total Installed Cost	$88,026,496
Installation cost	$6,293,616

CO_2 ABSORPTION

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Feed gas compressor	$55,557,757	$57,208,03	$1,650,280
2	Absorber tower	$281,414	$574,144	$292,729
3	Feed gas cooler	$126,730	$208,997	$82,268
4	Feed gas flash tank	$147,762	$403,618	$255,855
5	Slump tank	$383,384	$851,564	$468,181
6	Gas recycle compressor	$1,141,232	$1,357,817	$216,585
Total		**$57,638,279**	**$60,604,17**	**$2,965,899**

CO_2 STRIPPING

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Flash Tank 1	$280,083	$748,530	$468,447
2	Flash Tank 2	$280,882	$750,127	$469,246
3	Flash Tank 3	$265,839	$730,825	$464,986
4	Compressor for stream 5	$157,880	$213,523	$55,644
5	Compressor for stream 7	$330,003	$453,538	$123,535
6	Product Compressor	$3,256,232	$3,698,188	$441,956
7	H_2 Compressor	$16,539,013	$17,041,13	$502,126
Total		**$21,109,932**	**$23,635,87**	**$2,525,940**

Lean DEPG recycle

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Recycle Pump	$976,164	$1,288,861	$312,697
2	Dehydrator	$27,156	$130,723	$103,567
3	Recycle Cooler	$1,981,348	$2,366,862	$385,514
Total		**$2,984,669**	**$3,786,447**	**$801,778**

- **Flue gas physical novel with H_2**

Total Equipment Cost	$69,991,221
Total Installed Cost	$75,802,413
Installation cost	$5,811,192

CO_2 ABSORPTION

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Feed gas compressor	$53,615,812	$55,201,92	$1,586,117
2	Absorber tower	$281,414	$574,144	$292,729
3	Feed gas cooler	$125,265	$207,533	$82,268
4	Feed gas flash tank	$257,053	$539,666	$282,612
5	Slump tank	$409,741	$923,316	$513,574
6	Gas recycle compressor	$1,133,378	$1,348,100	$214,721
Total		**$55,822,664**	**$58,794,68**	**$2,972,022**

CO_2 STRIPPING

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Flash Tank 1	$373,134	$803,109	$429,975
2	Flash Tank 2	$373,134	$898,822	$525,688
3	Flash Tank 3	$343,581	$863,146	$519,565
4	Compressor for stream 5	$34,611	$52,981	$18,370
5	Compressor for stream 7	$235,355	$327,607	$92,252
6	Product Compressor	$11,660,322	$12,355,07	$694,750
7	H_2 Flash Tank	$35,942	$173,987	$138,045
Total		**$13,056,079**	**$15,474,72**	**$2,418,645**

Lean DEPG recycle

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Recycle Pump	$1,087,851	$1,405,075	$317,223
2	Dehydrator	$24,627	$127,928	$103,301
Total		**$1,112,478**	**$1,533,002**	**$420,524**

- **Flue gas chemical Conventional with heat pump**

Total Equipment Cost	$50,385,709
Total Installed Cost	$57,310,771
Installation cost	$6,925,062

CO_2 ABSORPTION

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Feed gas compressor	$17,477,371	$18,623,662	$1,146,291
2	Absorber tower	$501,061	$1,289,926	$788,865
3	Feed gas cooler	$21,965	$97,044	$75,079
4	Feed gas flash tank	$143,636	$398,293	$254,657

Total		**$18,144,033**	**$20,408,926**	**$2,264,892**

CO₂ STRIPPING				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Stripper tower	$2,311,750	$4,005,694	$1,693,944
2	Product flash tank	$49,787	$208,731	$158,944
3	H2 Compressor	$16,539,013	$17,041,139	$502,126
4	Product Compressor	$3,278,463	$3,757,959	$479,496
Total		**$22,179,013**	**$25,013,524**	**$2,834,510**

Heat Pump				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Compressor	$9,924,751	$11,408,826	$1,484,075
2	Reboiler heat exchanger	$137,912	$479,496	$341,584
Total		**$10,062,662**	**$11,888,321**	**$1,825,659**

- **Flue gas chemical novel with H_2 and heat pump**

Total Equipment Cost	$59,064,601
Total Installed Cost	$73,581,048
Installation cost	$14,516,447

CO₂ ABSORPTION				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Feed Gas compressor	$17,477,371	$18,623,662	$1,146,291
2	Absorber tower	$483,356	$1,168,788	$685,431
3	Feed Gas cooler	$21,832	$96,112	$74,281
4	Feed gas flash tank	$143,636	$398,293	$254,657
Total		**$18,126,195**	**$20,286,855**	**$2,160,660**

CO₂ STRIPPING				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Stripper tower	$7,937,773	$13,907,244	$5,969,471
2	Product flash tank	$58,306	$240,813	$182,507
3	H2 flash tank	$36,475	$175,318	$138,843
8	Product Compressor	$19,272,619	$20,132,171	$859,552
Total		**$27,305,173**	**$34,455,545**	**$7,150,372**

Heat Pump				
Equipment	Description	Equipment cost	Installed cost	Installation cost

1	Compressor	$11,229,763	$12,908,980	$1,679,217
2	Reboiler heat exchanger	$2,403,470	$5,929,668	$3,526,198
Total		**$13,633,233**	**$18,838,648**	**$5,205,415**

- **Syngas physical conventional**

Total Equipment Cost	$20,814,141
Total Installed Cost	$24,430,727
Installation cost	$3,616,586

CO_2 ABSORPTION

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Absorber tower	$1,084,790	$1,489,339	$404,550
2	Feed gas cooler	$85,329	$244,274	$158,944
3	Slump tank	$347,575	$698,344	$350,769
4	Gas recycle compressor	$959,924	$1,131,115	$171,191
Total		**$2,477,617**	**$3,563,072**	**$1,085,455**

CO_2 STRIPPING

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Flash tank 1	$201,010	$508,250	$307,239
2	Flash tank 2	$201,010	$568,420	$367,409
3	Flash tank 3	$193,822	$559,501	$365,679
4	Compressor for stream 5	$23,828	$39,536	$15,708
5	Compressor for stream 7	$108,892	$147,097	$38,205
6	Product Compressor	$2,173,040	$2,556,956	$383,916
7	H_2 Compressor	$14,592,409	$15,064,717	$472,307
Total		**$17,494,011**	**$19,444,476**	**$1,950,465**

Lean DEPG Recycle

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Recycle pump	$713,520	$993,070	$279,551
2	Dehydrator	$32,215	$138,311	$106,096
3	Recycle cooler	$96,778	$291,798	$195,020
Total		**$842,512**	**$1,423,179**	**$580,667**

- **Syngas physical novel with H_2**

Total Equipment Cost	$12,533,585			
Total Installed Cost	$15,847,591			
Installation cost	$3,314,006			
CO_2 ABSORPTION				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Absorber tower	$1,127,388	$1,537,662	$410,274
2	Feed gas cooler	$85,329	$244,274	$158,944
3	Slump tank	$407,212	$709,792	$302,580
4	Gas recycle compressor	$912,932	$1,083,591	$170,659
Total		**$2,532,862**	**$3,575,319**	**$1,042,458**
CO_2 STRIPPING				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Flash tank 1	$208,598	$518,367	$309,769
2	Flash tank 2	$208,598	$581,066	$372,468
3	Flash tank 3	$201,410	$573,345	$371,935
4	Compressor for stream 5	$108,892	$147,097	$38,205
5	Compressor for stream 7	$23,828	$39,536	$15,708
6	Product Compressor	$8,434,175	$9,057,972	$623,797
7	H2 flash tank	$36,475	$175,318	$138,843
Total		**$9,221,976**	**$11,092,702**	**$1,870,726**
Lean DEPG Recycle				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Recycle pump	$746,533	$1,041,259	$294,726
2	Dehydrator	$32,215	$138,311	$106,096
Total		**$778,748**	**$1,179,570**	**$400,822**

- **Syngas chemical conventional with heat pump**

Total Equipment Cost	$30,984,426			
Total Installed Cost	$37,211,168			
Installation cost	$6,226,742			
CO_2 ABSORPTION				

Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Absorber tower	$189,695	$443,554	$253,859
2	Feed gas cooler	$314,162	$451,674	$137,512
3	Feed gas flash tank	$68,290	$192,757	$124,467
Total		**$572,147**	**$1,087,984**	**$515,837**
CO_2 STRIPPING				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Stripper tower	$1,647,485	$3,105,009	$1,457,524
2	Product flash tank	$39,536	$181,442	$141,905
3	H2 flash tank	$0	$0	$0
7	H2 Compressor	$16,539,013	$17,041,139	$502,126
8	Product Compressor	$4,888,542	$5,285,503	$396,962
Total		**$23,114,576**	**$25,613,093**	**$2,498,517**
Heat Pump				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Compressor	$6,047,047	$6,951,278	$904,231
2	Reboiler heat exchanger	$1,250,656	$3,558,812	$2,308,156
Total		**$7,297,703**	**$10,510,091**	**$3,212,388**

- **Syngas chemical novel with H_2 and heat pump**

Total Equipment Cost	$43,783,699			
Total Installed Cost	$51,637,622			
Installation cost	$7,853,923			
CO_2 ABSORPTION				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Absorber tower	$197,150	$453,804	$256,654
2	Feed gas cooler	$312,564	$439,826	$127,262
3	Feed gas flash	$67,225	$191,559	$124,333
Total		**$576,939**	**$1,085,189**	**$508,250**
CO_2 STRIPPING				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Stripper tower	$3,603,807	$8,585,399	$4,981,592
2	Product flash tank	$39,536	$181,442	$141,905
3	H2 flash tank	$36,475	$175,318	$138,843
8	Product compressor	$32,789,291	$33,716,201	$926,910

Total		**$36,469,109**	**$42,658,359**	**$6,189,251**
Heat Pump				
Equipment	Description	Equipment cost	Installed cost	Installation cost
1	Compressor	$6,632,354	$7,624,108	$991,754
2	Reboiler heat exchanger	$105,297	$269,966	$164,669
Total		**$6,737,651**	**$7,894,074**	**$1,156,423**