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Process Integration and Techno-Economic Assessment of CO_2 Capture Processes Based on Phase-Change Solvents

Scientific Study on a State-of-the-Art Natural Gas Fired Power Plant

Master's Thesis in Sustainable Energy Systems

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Department of Space, Earth and Environment
Chalmers University of Technology
Gothenburg, Sweden 2019

Master's thesis 2019

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Abstract

Carbon Capture and Storage (CCS) has been suggested as a promising technology to combat climate change and eventually reach negative CO_2 emissions. An existing carbon capture method is Post Combustion Carbon Capture (PCCC), which is a type of CCS where typically an aqueous solution of amines is used to absorb CO_2 from the exhaust gases of an industrial plant. PCCC is currently under research in an EU funded project called ROLINCAP, which investigates new types of absorption technologies, in hope of decreasing the cost of carbon capture. One of these technologies is the so-called phase-change solvents, which can reach a multiple phase equilibrium in the presence of CO_2, each phase with different content in terms of amine, CO_2 and water. Because of this quality, phase-change solvents require less regeneration energy compared to 30 wt% Monoethanolamine (MEA), which is the conventional solvent for absorption-based PCCC.

In this thesis, PCCC using the 35 wt% phase-change solvent N-methylcyclohexylamine (MCA) is compared with using the conventional solvent 30 wt% MEA. The capture processes are theoretically implemented on a 400 MW natural gas combined cycle power plant located in Thessaloniki, Greece. Process integration between the power plant and capture plant was performed to reduce the overall process system energy consumption. The most cost-effective heat exchanger network design among many was chosen for each solvent for further evaluation. Finally, a techno-economic assessment was performed on the proposed capture plant designs, to estimate the capital and operational cost for each solvent process. This was performed with two different cost estimation methods: one factorial method, the other used at the engineering consultant company COWI which is an actor in the ROLINCAP project.

Results from this thesis show that MCA is the better choice of solvent. It demands about half the heat for solvent regeneration compared to MEA and this difference is reflected in the operational costs. The MCA plant also has smaller flows and equipment, which resulted in lower electricity cost for machinery but also lower capital costs. Altogether, this makes the cost of capturing one tonne of CO_2 much lower for MCA than MEA: 18.7 €/tonne CO_2 compared to 47.3 €/tonne CO_2. Another conclusion is that the potential of heat recovery between the power plant and capture plant is low. The reason is mainly because this is a retrofitting project which limits the integration potential due to constraints.

These results should be used with care. The thesis is based on models and information received from other actors. Because of a misunderstanding and modelling irregularity, the capture plant models exclude 19% of the exhaust gas flow and use a different gas composition. Because of this, the capture plant dimensions are not representative for the natural gas power plant in question. However, the composition and flows are still reasonable, and could represent partial capture. Also, the modelled MEA absorption process operates at a higher pressure than MCA. Because of this, the exhaust gas needed to be pressurised which resulted in higher electricity consumption, which increased the operational cost drastically. The higher operating pressure was most likely not required, and the capture cost for the MEA process is therefore higher than it could be.

Keywords: CCS, Process integration, MEA, MCA, ROLINCAP, Phase-change solvents, Techno-economic assessment.

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First of all we want to show our greatest gratitude to the division of Energy Technology, especially to our examiner and supervisor Dr. Stavros Papadokonstantakis for navigating us through the project. Another thanks to Guillermo Martinez Castilla and Gulnara Shavalieva for valuable advise throughout the project. Thank you to our opponents Erik Lindén and Emil Thureborn for suggesting improvements for the report. We would also like to thank Isabella Herstad Norin and Tomas Sundström at COWI for making us feel like welcome in their office and providing us with all the information needed to make a techno-economic assessment. At last, thank you Geoguessr for providing us with quality lunchtime entertainment during the spring, and of course to our friends in the Master's thesis room for all the fun we've had.

Rasmus Edvardsson & Herman Quint, Gothenburg, May 2019

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Abbreviations & Nomenclature

CAPEX	Capital Expenditures
CCS	Carbon Capture and Storage
CERTH	Center for Research and Technology Hellas
CEPCI	Chemical Engineering Plant Cost Index
CO_2	Carbon Dioxide
COP21	The 21st Conference of the Parties
DMCA	N-dimetylcyclohexylamine
EOR	Enhanced Oil Recovery
EU	European Union
HEN	Heat Exchanger Network
HP	High Pressure
HX	Heat Exchanger
HRSG	Heat Recovery Steam Generator
IEA	International Energy Agency
ISO	International Organization for Standardization
IP	Intermediate Pressure
LP	Low Pressure
MCA	N-methylcyclohexylamin
MEA	Monoethanolamine
MER	Maximum Energy Recovery
NPV	Net Present Value
NGCC	Natural Gas fired Combined Cycle
OPEX	Operational Expenditures
PBC	Packed Bed Column
PCCC	Post Combustion Carbon Capture
PFD	Process Flow Diagram
ROLINCAP	Systematic Design and Testing of Advanced ROtating Packed Bed Processes and Phase-Change SoLvents for INtensified Post-Combustion CO_2 CAPture
RPB	Rotating Packed Bed
SO_2	Sulphur dioxide
SS316	Stainless steel grade 316
TAC	Total annualised cost
ΔT_{min}	Minimum temperature difference

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1

Introduction

1.1 Background

Ever since 1895 when the Swedish chemist Svante Arrenius discovered that carbon dioxide (CO_2) has the ability of capturing heat on earth causing the greenhouse effect, human kind has been aware of the consequences of CO_2 emissions [1]. Despite this, carbon dioxide emissions from human activities have since then globally increased with more than 2400% up until 2015 [2]. The current CO_2 concentration in the atmosphere is consequently 35% higher than any other time the last 800 000 years [3]. In the latest decades however, global warming has gotten much more attention from the public. In 2015 the 21st conference of the parties (COP21) was held in Paris leading to the Paris agreement. The central aim of the agreement is to keep the average global temperature rise under 2 degrees compared to pre-industrial levels. [4]

To prevent future temperature rise above this target there is a need for new, innovative technologies. Emissions need to decrease and many believe that the technology of capturing CO_2 from large emission sources, so called carbon capture and storage (CCS), will play an important role in the near future to solve the climate problems [5]. According to the International Energy Agency's (IEA) sustainable development scenario[1], CCS will account for 7% of the global cumulative emission reduction needed by 2040 [6]. The main sector where CCS has been implemented so far is natural gas processing plants where raw natural gas is cleaned to produce dry pipeline natural gas. Different storage options are used including enhanced oil recovery (EOR) and storage in saline aquifers. In the electricity and heat production sector however, there are only few large scale facilities and the first one began operation on a coal power plant in Canada in 2014 [7]. Electricity and heat production is the sector with the largest share of global greenhouse gas emissions, see Figure 1.1, which means that it has a large potential to decrease its impact on global warming using CCS [8].

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[1] The IEA's Sustainable Development Scenario (SDS) outlines a major transformation of the global energy system, showing how the world can change course to deliver on the three main energy-related Sustainable development goals simultaneously.

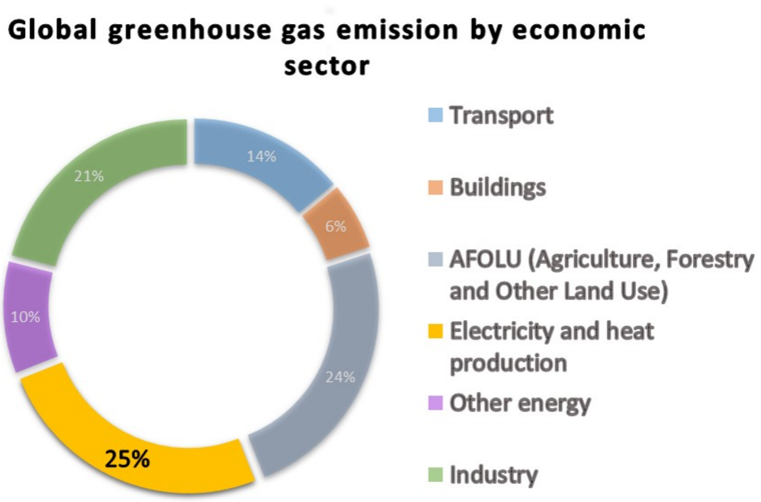


Figure 1.1: Global emission share by the main sectors based on global emissions from 2010. Source: IPCC [8].

There are three main CO_2 capture technologies; pre-combustion, oxyfuel combustion and post-combustion carbon capture (PCCC) [9]. The latter technology is relevant for this thesis, where CO_2 is removed from exhaust gases after combustion.

Even though it has a good potential, it is not possible to solve the whole climate crisis by only using CCS. Other emission mitigation measures also need to be considered like feedstock switching and efficient use of land, material and energy. One way to increase energy efficiency is to perform process integration. This is a method that aims to reduce energy consumption and emissions, and has since the 1980s been widely used in various applications to enhance the utilisation of energy [10]. The idea is to integrate complex or stand-alone systems with each other to reduce heating and cooling requirements and can be used both on individual processes and on large industrial plants.

1.2 ROLINCAP

In 2017, the EU Horizon2020 project ROLINCAP[2] started, whose purpose is to search, test and identify novel phase-change solvents which can be utilised in specifically designed rotating packed beds (RPB) processes for post-combustion CO_2 capture [11]. The aim with this is to decrease the heat consumption and equipment size for post-combustion CO_2 capture, and also to decrease the temperature required to drive the CO_2 extraction process.

Several collaboration partners are involved in the project from both industry and various universities and research institutes, each partner responsible for different parts of the project. These parts include research about thermodynamic properties, molecular design, sustainability assessment, process modelling as well as column design and pilot plant testing. Two different plants, both located in Greece, were chosen as reference plants to investigate the CCS integration potential. One of the plants is a natural gas fired combined cycle power plant with a capacity of 400 MW owned by the energy company Elpedison and the other plant is a quicklime plant owned by the chemical company CaO Hellas.

---

[2] Project title: Systematic Design and Testing of Advanced ROtating Packed Bed Processes and Phase-Change SoLvents for INtensified Post-Combustion CO_2 CAPture

The ROLINCAP project is set to be finished in 2019 and one of the last stages in the project is to create models of the reference plants, perform process integration of the emitting plants and the CO_2 capturing plants to optimise heat recovery, and finally to perform a techno-economic assessment of the integrated plants. Chalmers University of Technology is the leader of this stage and will perform a process integration on the power plant. For the quicklime plant, the Center for Research and Technology Hellas (CERTH) will perform this task and the engineering consulting company COWI will then perform techno-economic assessments for both plants in collaboration with Chalmers, Elpedison and CERTH. The assessments will have the detail of a feasibility study, and can be utilised to guide investment decisions in future CO_2 capture systems.

The most used solvent for post-combustion capture is monoethanolamine (MEA), diluted in water, where concentrations of 30wt% MEA is the benchmark solvent for PCCC [12]. MEA belongs to the group of amino alcohols, which have properties from both alcohols and amines. Two high performing phase-change solvents in the ROLINCAP project are also substances from this group of chemicals:
N-methylcyclohexylamine (MCA) and a confidential solvent hereafter denoted as the ROLINCAP solvent. From an earlier stage in the ROLINCAP project, the ROLINCAP solvent was established a promising solvent and will therefore be subject of investigation. However, MCA will also be investigated as a reference phase-change solvent since there is more information available about it.

1.3 Aim and objective

The aim of this thesis was to first integrate a PCCC process, using the solvents MEA and MCA, to a power plant and then to perform a techno-economic assessment of the capture plant. The thesis was made in cooperation with the ROLINCAP project stage described in Section 1.2, and is based on information and methods from the project. To perform these tasks, the work was divided into two main objectives:

a) **Process integration**
The first objective of the thesis was to perform a thorough process integration between the power plant and the capture plant, investigating the possibilities of using waste heat to decrease the energy demand of the integrated process.

b) **Techno-economic assessment**
The second objective was to determine and size equipment for the proposed

process integration and the capture plant, and also to make a detailed cost estimation for the added carbon capture plant.

Some specific questions have been answered in this report:

- How much energy is cost-effective to save using process integration for the different kinds of solvents compared to operating the plants separately?
- How efficient will the proposed integrated power plant be compared to the original power plant for each solvent?
- Will the phase-change solvent MCA decrease the capital expenditures (CAPEX) and operational expenditures (OPEX) of the plant compared to the conventional reference solvent MEA?
- What is the cost of capturing one tonne of CO_2 using MEA and MCA?

1.4 Limitations

As this thesis was carried through, the work to analyse and evaluate the best phase-change solvent was still ongoing in the ROLINCAP project. With the last delivered information, the ROLINCAP solvent was mentioned as a main candidate under investigation, but there were no results or data ready to be used for this thesis. Because of this, the best phase-change solvent from ROLINCAP was not evaluated in this thesis. However, a complete integration and techno-economic assessment of the reference phase-change solvent MCA was performed.

One main limitation with the thesis was to perform process integration and techno-economic assessment only on the natural gas plant owned by Elpedison. The quick-lime plant considered in the ROLINCAP project was therefore not included.

The investigation was limited to steady state operation of the power plant, with the operation properties of a base load power generation of 365 MW electricity before adding the capturing plant. Ambient conditions of 19 °C and 101325 Pa was used.

Transport and storage was not included for the captured CO_2 in this thesis since the ROLINCAP project only investigates the absorption process. However, the CO_2 should be ready for transportation, so compression of the CO_2 was included, leaving a stream of at least 98% molar pure CO_2 of 150 barg and around 40 °C as the final product.

The thesis was limited of using packed bed (PB) columns, since information about the RPB columns in the ROLINCAP project was not available at the time of the thesis.

Another limitation is that the capture plant models provided were already optimised and could not be changed in terms of process condition. The capture plant and the integration were therefore optimised separately.

2

Theory

In this chapter, the underlying theory which this thesis is based on is presented. The purpose of this is to give the reader a sufficient knowledge to understand the terminology and concepts discussed further.

2.1 Post-combustion capture

In post-combustion capture CO_2 is removed from exhaust gases after combustion. Because of this, no large changes on the existing combustion process of a plant are required, making it easier to retrofit the capture plant in existing plants. Figure 2.1 illustrates a basic sketch of an absorption based capture process. Conventional PCCC uses a chemical, typically called a solvent, to separate the carbon dioxide from the exhaust gases. The solvent, dissolved in water, extracts the CO_2 from incoming exhaust gas in an absorber column. The CO_2 is then desorbed from the solvent by letting the mixture enter a stripper column while heating it. This is called regeneration of the solvent. This way of capturing CO_2 is a proven technology, but has the downside of consuming large amounts of energy in the process. MEA is undoubtedly the currently most used solvent in post-combustion carbon capture applications due to its many benefits including low cost, biodegradability and its high reactivity towards CO_2 [13] [14].

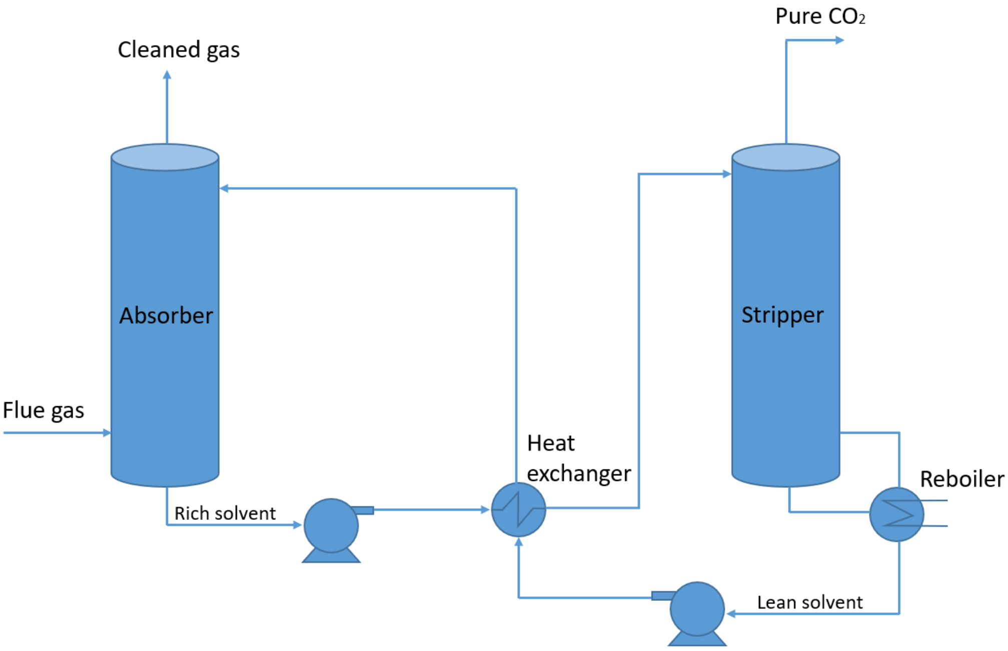


Figure 2.1: Simplified design of a post-combustion capture process

2.1.1 Fundamental equipment

To separate the carbon dioxide from the exhaust gas, two different columns are required: An absorber and a stripper. In the case of this thesis, exhaust gas from the power plant is entering the absorber column at the bottom. A recirculated counter-current flow consisting mainly of water and solvent is entering at the top. The main idea with the columns is to induce mass transfer between the two fluids, in this case CO_2 transferred between gas and solvent. A common way to induce mass transfer is to use a packed bed material inside the columns, so called packed bed columns (PBC). Packed bed material can either be structural, as in Figure 2.2, or random. Thanks to the bed, the liquid is able to spread out over the whole column area allowing the gas to react with a large share of the fluid.

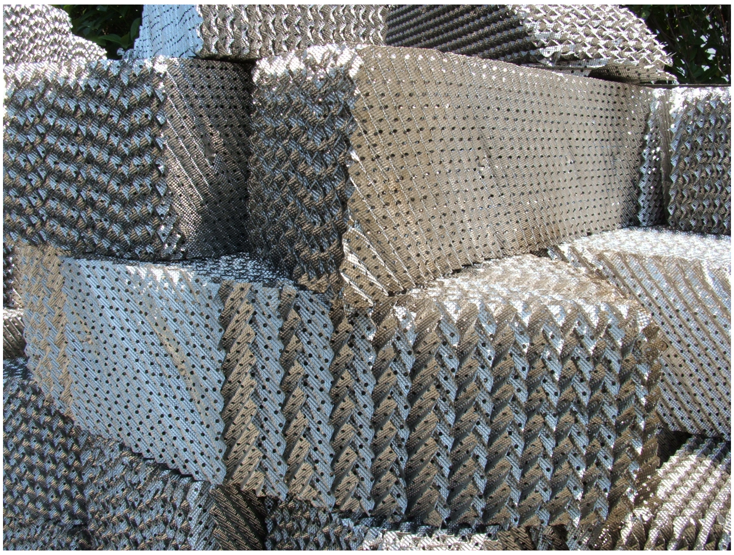


Figure 2.2: Structured packed bed material. Photograph by Luigi Chiesa. Distributed under a CC BY-SA 3.0 license.

The stripper column in ROLINCAP is similar to the absorber. It is a PB column where mass transfer is induced, but opposite to the absorber the CO_2 is stripped off the CO_2 rich solvent. Since no exhaust gas requiring much space is entering the stripper, it is smaller than the absorber. CO_2 rich solvent is entering at the top of the stripper column. In the bottom of the column, solvent is heated using a reboiler to release the CO_2 from the solvent, while simultaneously some water and solvent evaporation take place. The steam and gas is rising up through the column as the packed bed induces mass transfer of CO_2 from the liquid to the steam and gas phase. The resulting stream exiting the top of the stripper contains CO_2 and steam, which after removing the steam in a condenser can be compressed and sent to a storage location. The CO_2 lean phase is exiting the stripper at the bottom, and is recirculated back to the top at the absorber through an internal heat exchanger (HX) to heat the rich solvent flow.

2.1.2 Phase-change solvents

The new type of chemicals investigated in the ROLINCAP project is referred to as phase-change solvents. Dissolved in water, the liquid reaches a multiple phase equilibrium in the presence of CO_2 at a specific temperature, resulting in several phases with different composition in terms of solvent and water. In the case of MCA, a CO_2 rich phase and a CO_2 lean phase is formed which is illustrated in Figure 2.3

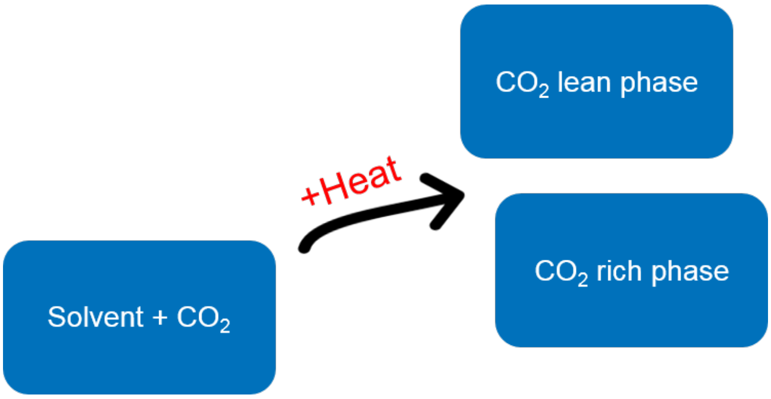
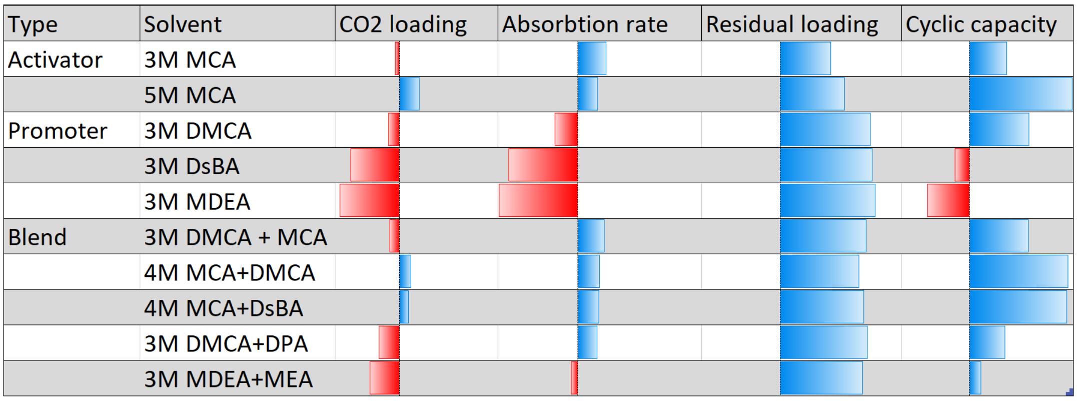


Figure 2.3: Illustration of a phase-change equilibrium in the addition of heat.

This property is useful in PCCC processes since only the CO_2 rich phase needs to enter the stripper to be processed in the reboiler. The phases can simply be separated after the absorber outlet using a decanter (a mechanical separator) and let the CO_2 lean phase be circulated back to the absorber. In the MCA flowsheets used in this thesis the solvent is heated to 101 °C for the regeneration to take place. Another benefit with MCA is that it is more reactive than MEA. Therefore, a smaller amine-water flow can be used and still reach the same capture rate as with MEA.

Solvents for this application have earlier been categorised into activators and promoters [15]. Activators are characterised by rapid reaction kinetics which is beneficial in the absorption stage. Promoters are more beneficial in the desorbtion column as they possess excellent regenerability. In [15], several solvents were tested in terms of different aspects including CO_2 loading, absorption rate, residual loading and cyclic capacity compared to the conventional solvent MEA. In every one of these aspects a high value is beneficial except for residual loading where a low value is desirable. The results are illustrated in Figure 2.4. The molar concentration of each tested solvent is stated with M (number of moles per litre). 5M MCA is beneficial in every aspect and so is two of the mixtures who also have slightly better values. The results of this study indicates that a mixture of an activator and a promoter can generate beneficial abilities from both types of solvents for the capture process.



Type	Solvent	CO2 loading	Absorbtion rate	Residual loading	Cyclic capacity
Activator	3M MCA	🟥	🟦	🟦	🟦
	5M MCA	🟦	🟦	🟦	🟦
Promoter	3M DMCA	🟥	🟥	🟦	🟦
	3M DsBA	🟥	🟥	🟦	🟥
	3M MDEA	🟥	🟥	🟦	🟥
Blend	3M DMCA + MCA	🟥	🟦	🟦	🟦
	4M MCA+DMCA	🟦	🟦	🟦	🟦
	4M MCA+DsBA	🟦	🟦	🟦	🟦
	3M DMCA+DPA	🟥	🟦	🟦	🟦
	3M MDEA+MEA	🟥	🟥	🟦	🟦

Figure 2.4: Experimental data from [15] illustrated in comparison with data for conventional solvent MEA. Blue bars indicate better performance than the respective value for MEA and red bars indicate a worse performance. All blends are mixed in proportion of 3:1.

2.1.3 Solvent degradation

Amine-based absorption-desorption, both single phase and phase-change, can suffer from solvent degradation in different ways [16]. Oxygen molecules present in the exhaust gas can react with the solvent which creates ammonia, called oxidative degradation. Due to the volatility of ammonia, the toxic substance can escape with the exhaust gases which can be harmful to the environment. Acidification is one of the consequences of ammonia emissions and it is therefore important to keep these emissions at a low level [17]. In addition to this, amines suffer from degradation in environments like in the reboiler. At these temperatures, the amines reacts with CO_2 to produce a variety of different decomposition products including ethylene amine derivatives. These derivatives are undesired since they lead to corrosion of the capture plant equipment. MCA is quite unstable at high temperatures compared to MEA while DMCA is more stable than both of these activator solvents. However, MCA is not exposed to as high temperature as MEA and thermal degradation is therefore minor for MCA, while it is more of a problem for MEA. MEA is usually stable at 120 degrees, but in the presence of CO_2, decomposition can occur at lower temperatures [18]. Another type of amine degradation products are nitrosamines which are known to be carcinogenic [19]. They are not emitted in the same extent as ammonia, but still pose an occupational hazard as they accumulate in the capture system.

2.1.4 Cleaning equipment

In order to do avoid unwanted emissions, it is necessary to add some extra cleaning equipment to the system. In the article [16], some types of cleaning equipment are proposed to countermeasure solvent losses as well as thermal and oxidative degradation products from MEA. In this article the exhaust gases pass through a water wash downstream of the absorber. In this component cold water is sprayed over the exhaust gases to remove eventual solvent following the exhaust gas exiting the absorber. The cold water and the solvent is then recirculated over the packed bed of the absorber. To avoid emissions related to solvent degradation of any kind an acid wash can be added after the water wash. This component basically work in same way as the water wash, but with an aqueous solution of acid (Sulfuric acid in this case) instead of water. Countercurrent contact between the exhaust gas and acid leads to removal of ammonia present in the exhaust gases. Directly after each of these two components there is a demister preventing droplets to follow through. To remove degradation products from the capture system, reclaimer systems can be used to purge the circulating flow [19].

2.2 Elpedison natural gas power plant

The plant under investigation in this thesis is a natural gas fired combined cycle (NGCC) power plant owned by the electricity company Elpedison, located in Thessaloniki, Greece. The power plant runs with a 265 MW gas turbine and a 135 MW triple pressure steam turbine, together yielding a net capacity of 400 MW. Heat is transferred from the gas turbine exhaust gases to the steam cycle in a heat recovery steam generator (HRSG). This type of plant is one of the most efficient forms of thermal power production, and the plant reaches an efficiency of 56% at full load. In addition to the high efficiency, the power plant also has highly flexible operation, making it suitable for both base load operation and demand variation management. Figure 2.5 shows a simplified process diagram of the power plant, where the main components and streams are displayed.

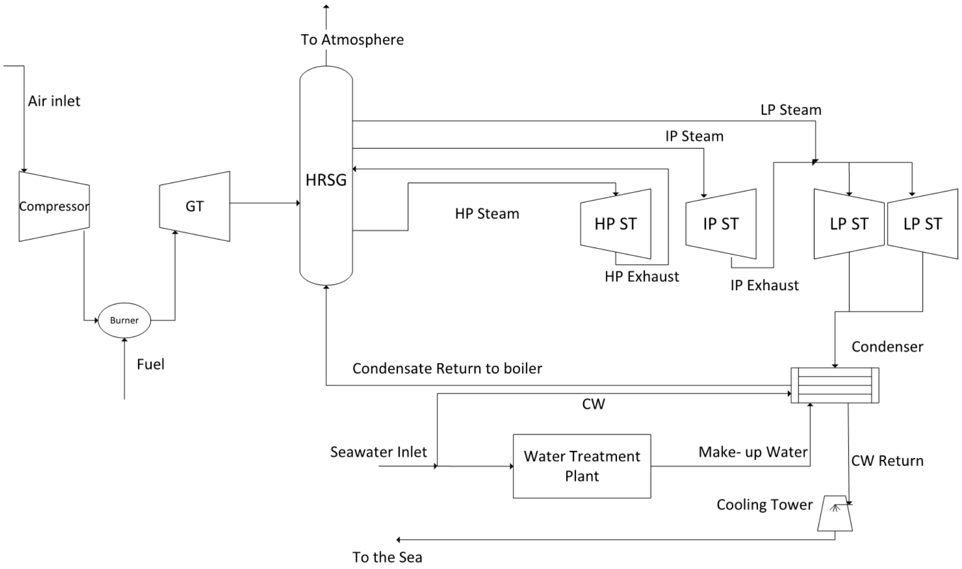

To Atmosphere
Air inlet
Compressor
GT
HRSG
LP Steam
IP Steam
HP Steam
HP ST
IP ST
LP ST
LP ST
HP Exhaust
IP Exhaust
Burner
Fuel
Condenser
Condensate Return to boiler
CW
Seawater Inlet
Water Treatment Plant
Make- up Water
CW Return
Cooling Tower
To the Sea

Figure 2.5: Schematic process diagram of main components and streams in the power plant. Abbreviations: GT - Gas Turbine, ST - Steam Turbine, CW - Cooling Water.

The gas turbine is a General Electric 9FA 2.0+ which is connected to a generator. At ISO 3977-2 conditions (15 °C, 65% humidity, 1.01 bar), the turbine has a nominal net power output of 255 MW. Depending on the ambient conditions, the gas turbine can reach a maximum net power output of 275 MW. The fuel is natural gas that is supplied from a purpose-built pipeline, and is preheated by hot intermediate pressure feed water from the steam cycle before it is combusted with compressed air of 16 bar and expanded through the turbine. The exhaust gases have a temperature of about 600 °C which is utilised in the HRSG to generate steam for the steam cycle.

The steam cycle has three pressure levels, low pressure (LP) of about 3.7 barg, intermediate pressure (IP) of about 23 barg and high pressure (HP) of about 112 barg. These pressurised steam streams are expanded in a modified Franco Tosi steam turbine, with a maximum power output of 144 MW at ISO 3977-2 conditions. After expansion in the HP turbine, the HP steam is reheated and added to the IP steam stream. The IP steam is thereafter expanded and added to the LP steam stream before final expansion in the LP turbine. The expanded steam is finally condensed at a pressure of ca 57 mbarg before it is pumped back to the HRSG with a pressure of 3.8 barg to be heated again. Figure 2.6 show a schematic diagram of the HRSG and it is connections.

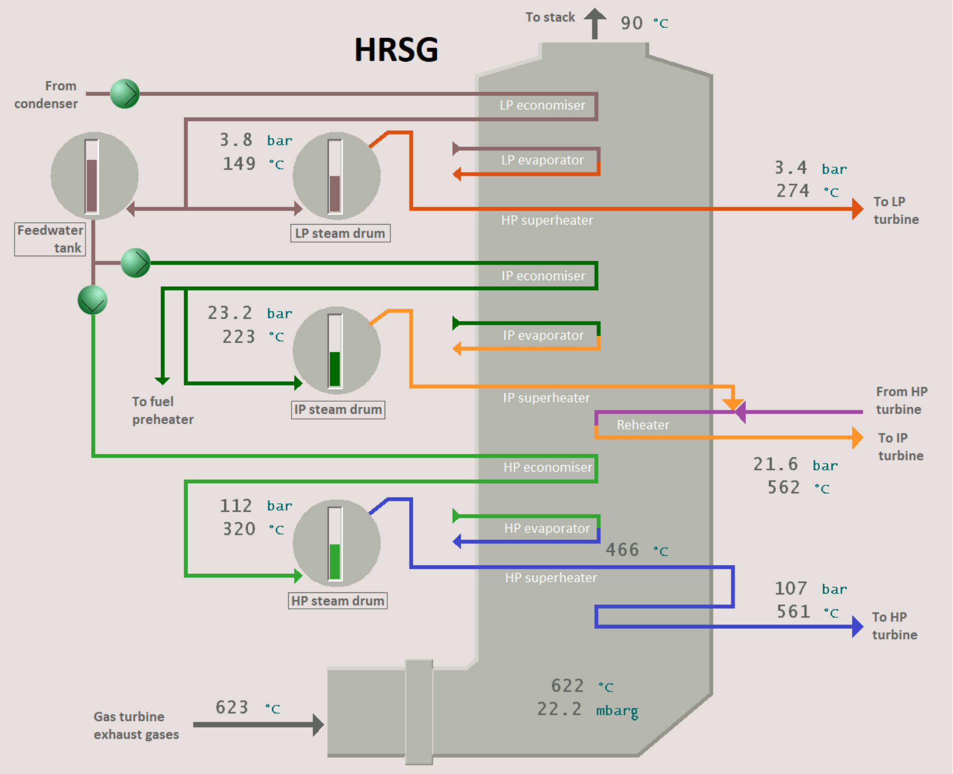

To stack
90 °C
HRSG
From condenser
LP economiser
3.8 bar
149 °C
LP evaporator
3.4 bar
274 °C
To LP turbine
HP superheater
Feedwater tank
LP steam drum
IP economiser
23.2 bar
223 °C
IP evaporator
To fuel preheater
IP steam drum
IP superheater
From HP turbine
Reheater
To IP turbine
21.6 bar
562 °C
HP economiser
112 bar
320 °C
HP evaporator
466 °C
HP steam drum
HP superheater
107 bar
561 °C
To HP turbine
622 °C
22.2 mbarg
Gas turbine exhaust gases
623 °C

Figure 2.6: Schematic diagram of the HRSG with associated streams and heating stages. Stream colour explanation: Dark grey - exhaust gas, brown - LP feedwater, dark green - IP feedwater, light green - HP feedwater, dark orange - LP steam, light orange - IP steam, blue - HP steam, purple - expanded HP steam.

The steam cycle condenser is using sea water pumped from the sea through a pipeline to condense the expanded steam. The sea water is thereafter cooled in cooling towers using mechanical induced draft, before it is returned to the sea. A part of the sea water is treated in two desalination plants to be used as make up water added in the condenser, to cover losses occurring around the steam cycle.

There are also three subsystems with smaller flows in the plant. There is a closed cooling water system that is refrigerating the machinery. This is also using sea water, but only a small fraction of the cooling water needed for the condenser cooling. There is a gland condenser that collects condensate from a HP steam bleed that is used to seal the system from atmospheric air. There is also a boiler blowdown tank, which collects hot water from the HRSG tubes to remove solid deposits from them.

2.3 Pinch analysis

Pinch analysis is a method of process integration that aims to minimise energy requirements of a specific process and consequently the costs by reducing fuel and electricity demands. Heating and cooling demands of a chemical process are identified to ultimately suggest the most efficient heat exhanger network (HEN) possible. It can be used for both designing new processes (grassroot designs) and to retrofit existing systems [20]. Firstly, the whole process is analysed thoroughly to identify existing streams with heating or cooling requirements. By following some stated ground rules for pinch analysis some specific targets can be found, e.g the pinch temperature, the minimum heating requirement and the minimum number of HX units needed. Composite curves can also be constructed which are curves that clearly show the potential of heat recovery of the streams selected. Another curve called grand composite curve (GCC) can be created which shows the net heat flow against shifted temperatures. The work then consists of designing a HEN by matching available streams using internal HXs and adding external heaters and coolers on streams where heat recovery is not possible. If the suggested design matches the calculated targets it is the best design possible in terms of maximum recovered heat.

2.3.1 Pro_PI

Pro_PI is an Excel add-in developed by Chalmers Industriteknik AB. It is a software focusing on process integration and facilitates the work in reaching the most beneficial HEN. In this add-in, stream data can be entered together with available heating and cooling utilities to find useful target values and to construct composite curves and GCCs. Graphical stream representation can also be obtained obtained in order to design complete HENs by adding internal HXs, external heaters and coolers to the streams.

2.3.2 ΔT_{min} and pinch point

When designing a HEN it must firstly be stated which minimum allowable temperature difference between hot and cold streams, ΔT_{min}, to use. In general, larger ΔT_{min} leads to smaller heat transfer area and higher utility duty while a small ΔT_{min} leads to the opposite. However, the most beneficial ΔT_{min} to use is different depending on the case. Figure 2.7 shows a general curve for total annualised cost (TAC), divided into energy and area cost, plotted against different choices of ΔT_{min}. Here it is clearly illustrated the behaviour of energy and area cost as ΔT_{min} increases. In this particular graph the lowest TAC occurs at a relatively flat part of the curve which is often the case. The choice of ΔT_{min} is therefore not very crucial for the design as long as no extreme values are chosen. Usually a value between 5 and 25 is preferred.

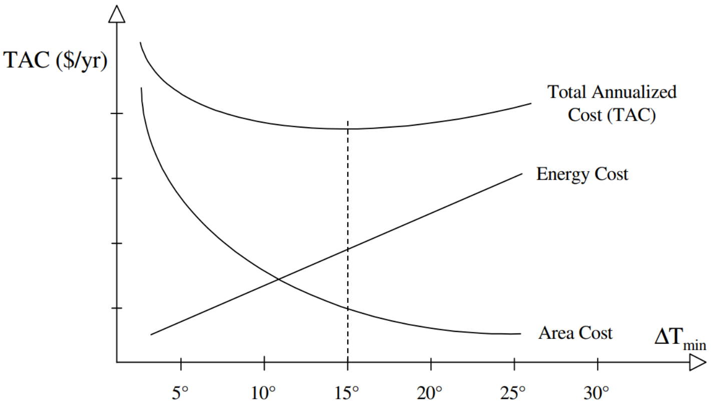


Figure 2.7: Total annualised cost of a HEN. Divided into area costs and energy costs and plotted against ΔT_{min}. Retrieved from [20].

In each system, independently on how complex it is, ΔT_{min} will in general occur at one point (temperature). This point is referred to as the pinch point of the system, hence the term pinch analysis. The pinch point separates the system of streams into two subsystems: a temperature region above the pinch temperature with a heat deficit and a region below where there is a heat surplus. The HEN is based on this pinch point and the two subsystems should be designed individually to reach a maximum energy recovery (MER) network.

2.3.3 Targets

One practical thing about Pro_PI is that it calculates some targets that are useful for the HEN design. By creating composite curves the heat recovery potential is illustrated and the minimum energy requirements are found. Figure 2.8 shows two general and simple composite curves. The blue line indicates all the selected cold streams that need to be heated and the red line indicates the hot streams that need to be cooled. Relevant targets are pointed out which can all be calculated in Pro_PI. $Q_{H,min}$ and $Q_{C,min}$ is the minimum hot and cold utility demand respectively, and Q_{HX} is the maximum internal heat exchange. ΔT_{min} is also pointed out and occurs in one point. If ΔT_{min} would be increased, the blue line has to be shifted to the right to satisfy this value.

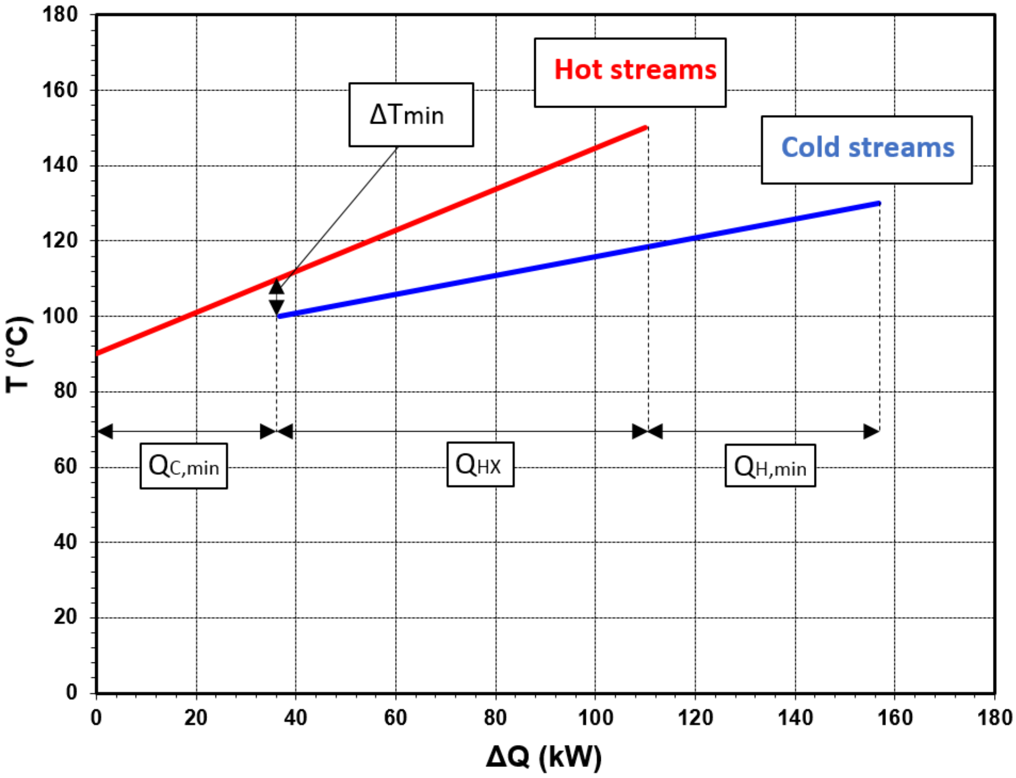


Figure 2.8: General composite curves. The curve is unrelated to this project.

2.3.4 Heat exchanger network design

The main idea when designing HENs is to first connect streams with internal HXs where it is suitable. External heaters and coolers (using available steam from the plant and cold water from the cooling system) should then be used as a complement to satisfy all remaining energy demands. In addition to this, there are three rules to consider when designing HENs.

- Do no transfer heat across the pinch
- Do not heat with heating utilities below the pinch
- Do not cool with cooling utilities above the pinch

Violating one of these rules means that there is an unnecessary use of utilities somewhere in the system where heat recovery could have been used instead. If none of these rules are violated in a design, the MER network is reached. However, that does not necessarily mean the most cost effective solution. It might not be cost effective to add too small HXs just to satisfy small energy needs. Therefore, the HENs created should be evaluated based on some valid parameters e.g cost or flexibility.

2.3.5 HEN selection

When choosing a HEN for further investigation it is a question of selecting the most cost-effective design rather than selecting the design with the largest heat recovery. It is desired to have a well integrated design in terms of heat recovery but in the same time keeping CAPEX and OPEX on acceptable levels. It is important to look at several parameters and not just one. For example, a design with many small HXs can be beneficial in terms of heat recovery, but might lead to a complex and expensive system which is not flexible to load changes. If that is the case it might be more beneficial to extend heating and cooling utilities in the system, and by that violating one of the pinch rules, to satisfy that energy need. In this thesis however, flexibility will not be considered when selecting HENs because of the given time-frame.

2.4 Costing methods

Two costing methods were used in this thesis. They have different approaches and are described in the following section.

2.4.1 Smith costing method

The costing method denoted as the Smith method is based on [21]. It is a factorial costing method and it calculates equipment costs based on a base case according to Equation 2.1.

$$C_E = C_B(\frac{Q_E}{Q_B})^M f_{\mathrm{M}} f_{\mathrm{T}} f_{\mathrm{P}} \tag{2.1}$$

where:

- C_B is the equipment cost in US$ for a base case with capacity Q_B.
- C_E is the equipment cost in US$ with capacity Q_E.
- M is a constant depending on the type of equipment.
- The factors f_{M}, f_{T} and f_{P} are related to material, temperature and pressure.

If the piece of equipment operates in a design temperature of around 0-100 °C and in a design pressure of 0,5-7 bar, the factors f_{T} and f_{P} are simply 1.0, meaning that no extra costs need to be added. The base case cost is based on a construction material, often carbon steel. If a piece of equipment needs to be constructed in another material, e.g. because of corrosion risks, f_{M} needs to be adjusted.

This way of calculating costs for specific pieces of equipment is valid in size ranges. If Q_E for a piece of equipment is larger than this range it needs to be split up into two or more equally big pieces. A number of additional factors need to be added to C_E to reach the fixed capital cost, C_F, related to that piece of equipment and account for the whole plant according to Equation 2.2. Explanations and typical values are stated in Table 2.1. Cost estimations of this type and level of detail can at best hold an accuracy of ±30%. [21]

$$\begin{aligned} C_F = C_{\mathrm{E}}(1 &+ f_{\mathrm{PIP}} + f_{\mathrm{ER}} + f_{\mathrm{INST}} + f_{\mathrm{ELEC}} + f_{\mathrm{UTIL}} + \\ &f_{\mathrm{OS}} + f_{\mathrm{BUILD}} + f_{\mathrm{SP}} + f_{\mathrm{DEC}} + f_{\mathrm{CONT}} + f_{\mathrm{WS}}) \end{aligned} \tag{2.2}$$

Table 2.1: Typical values and explanation to factors for capital cost based on pure equipment cost.

Factors for capital cost	Explanation	Typical value
f_{PIP}	Costs related to piping	0.7
f_{ER}	Erection of equipment	0.4
f_{INST}	Instrumentation and controls	0.2
f_{ELEC}	Electrical	0.1
f_{UTIL}	Utilities	0.5
f_{OS}	Off-sites	0.2
f_{BUILD}	Buildings (Including services)	0.2
f_{SP}	Site preparation	0.1
f_{DEC}	Design, engineering and construction	1.0
f_{CONT}	Contingency	0.4
f_{WS}	Working capital	0.7

These cost calculations are based on the cost level of January 2000. Therefore, the cost level need to be updated to the current cost levels. This is accounted for by using the Chemical Engineering Plant Cost Index (CEPCI) according to Equation 2.3. The CEPCI for January 2000 is 435.8 and the value for 2018 is 603.1 [22].

$$\frac{C_{2018}}{C_{2000}} = \frac{CEPCI_{2018}}{CEPCI_{2000}} \tag{2.3}$$

2.4.2 COWI costing method

At COWI a software called Cleopatra is often used to estimate costs of equipment. It is based on equipment cost from numerous of previous projects, and by entering some basic parameters (e.g heat exchanger area, pump duty etc.) a rough cost estimation can be provided for most process industry equipment. Cleopatra is able to provide pure equipment costs and an additional cost for installation. If the size of a piece of equipment is too large for the software to provide cost estimations for, it need to be divided into a number of smaller units whose size are within the valid intervals. The accuracy of this costing software and level of detail is around $\pm 20 - 30\%$ according to COWI.

2.4.3 Net present value

In the Net present value (NPV) method, all future payments are translated into today's monetary value based on an interest rate *i*. Using Equation 2.4, the annual share of the capital cost can be calculated over the given time period *n*. All future payments can then be translated into the monetary value of today using Equation 2.5. Based on which year the payment occurs it is worth a different amount of today's money, where a payment occurring 25 years from now is worth less than a payment occurring in 5 years.

$$\text{Annualised capital cost} = \text{Capital cost} \times \frac{i(1+i)^n}{(1+i)^n - 1} \tag{2.4}$$

$$\text{PV} = \frac{F}{(1+i)^t} \tag{2.5}$$

where:

- PV is present value of a cash flow F.
- t is the time until the payment occurs in years.

3

Methodology

Methodically, the thesis was divided into three parts with sub tasks, presented in the list below. The two latter parts were performed for both 30 wt% MEA solvent and 35 wt% MCA solvent.

- **Power plant modelling**
  A model of the power plant was established to get a visual overview of the plant and to get approximate results for different capture plant integrations.
- **Process integration**
  - **Pinch analysis**
    Pinch analysis including streams from both the power plant and the capture plant was used to find targets for maximum energy recovery and to find suitable streams for heat exchanging.
  - **HEN selection**
    A large number of HENs were systematically created and cost estimated to find the HEN with lowest cost. The cost estimations include both operational and capital expenditures.
- **Techno-economic assessment**
  - **Equipment design**
    The suggested pieces of equipment were dimensioned to handle the magnitude of the capture process.
  - **Process cost estimation**
    The capital and operational costs of adding the capture plant to the power plant were calculated. This was performed using two different costing methodologies. Performance metrics such as CO_2 avoidance cost were also calculated to compare the two solvents.

3.1 Project data

As this project was part of the ROLINCAP project, it was based on work from earlier project stages and also on new information from the project partners. For the power plant, data was available in tables for four base load operation points, two medium load operation points and two minimum load operation points. This data was measured and compiled in tables by the power plant company Elpedison, which had sampled the power plant in different parts of the year to get data with different seasonal conditions. By using an average of these four base load operation points, the seasonal variations could be flattened out, obtaining an average base load. The data also contained information about the operation conditions, fuel composition, process flow temperatures and pressures, turbine power outputs and exhaust gas composition. However, the information was not complete, lacking especially for the flows in the HRSG. Except from the tabulated data, there were also screenshots of the power plant control program showing flow schematics and measured operation values, shot at arbitrary times.

The design and modelling of the capture plant was performed by CERTH, who presented the components and molar flows in Process Flow Diagrams (PFD) accompanied with the associated temperatures, pressures and composition of the flows. The PFDs provided by CERTH was optimised for the average base load of Elpedison. These PFDs are based on a capture rate of 90% which means that 10% of the CO_2 in the exhaust gas is released into the atmosphere. CERTH also provided an equipment sheet that includes information about the columns, the compression train and the heat exchangers in the PFDs. However, equipment information about the decanter or the CO_2 intercoolers was not included. The relevant information from the equipment sheet is presented in Table 3.1. The compression train is the same for both solvents.

Table 3.1: Capture plant equipment data from CERTH. Abbreviation explanation: D=diameter, H=height, TS=number of theoretical stages, Q=heat transfer, A=HX area, Q_{el}=power demand, P_{out}=unit outlet pressure.

Description	**MEA capture plant**	**MCA capture plant**
Columns		
Absorber column	D=9.5 m, H=32.1 m	D=9.8 m, H=33.4 m
Stripper column	D=4.0 m, H=18,6 m	D=3.5 m, H=17.3 m
Absorber packing	TS=52 stages, H=20.8 m	TS=50 stages, H=22.4 m
Stripper packing	TS=32 stages, H=12.8 m	TS=30 stages, H=12.0 m
Heat exchangers		
Stripper condenser	Q=67.8 MW, A=2896 m^2	Q=64.4 MW, A=2789 m^2
Stripper reboiler	Q=138.6 MW	Q=68.1 MW
R/L HX	Q=75.7 MW, A=4165 m^2	Q=79.8 MW, A=4651 m^2
CO_2 compression train		
CO_2 compressor 1	Q_{el}=2.16 MW, P_{out}=0.24 MPa	
CO_2 compressor 2	Q_{el}=2.14 MW, P_{out}=0.56 MPa	
CO_2 compressor 3	Q_{el}=2.12 MW, P_{out}=1.32 MPa	
CO_2 compressor 4	Q_{el}=2.08 MW, P_{out}=3.12 MPa	
CO_2 compressor 5	Q_{el}=2.00 MW, P_{out}=7.38 MPa	
CO_2 pump	Q_{el}=2.86 MW, P_{out}=15 MPa	

The PFD of the MEA process is displayed in Figure 3.1, with the capture cycle on the left and the CO_2 compression train on the right side. In the top left corner there is a gas cleaning part based on what is described in Section 2.1.4. The corresponding PFD for the MCA capture plant is presented in Figure 3.2. A decanter and a stream splitter distinguish the phase-change process. In Table 3.2, important information from the PFD is displayed, such as column temperatures, reboiler heat load, solvent CO_2 loading and CO_2 purity of the resulting CO_2 stream.

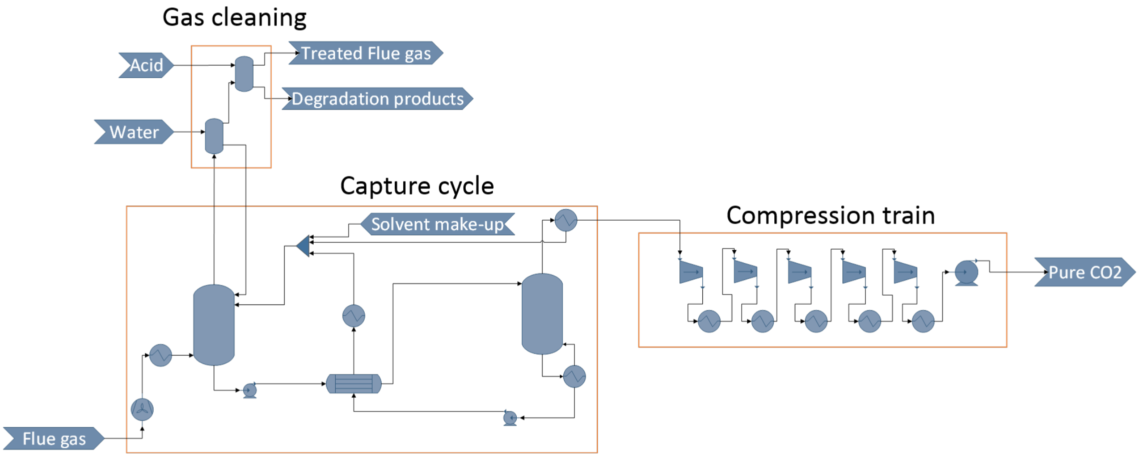

Gas cleaning
Acid
Treated Flue gas
Degradation products
Water
Capture cycle
Solvent make-up
Compression train
Pure CO2
Flue gas

Figure 3.1: The MEA based capture plant PFD presenting main components and flows.

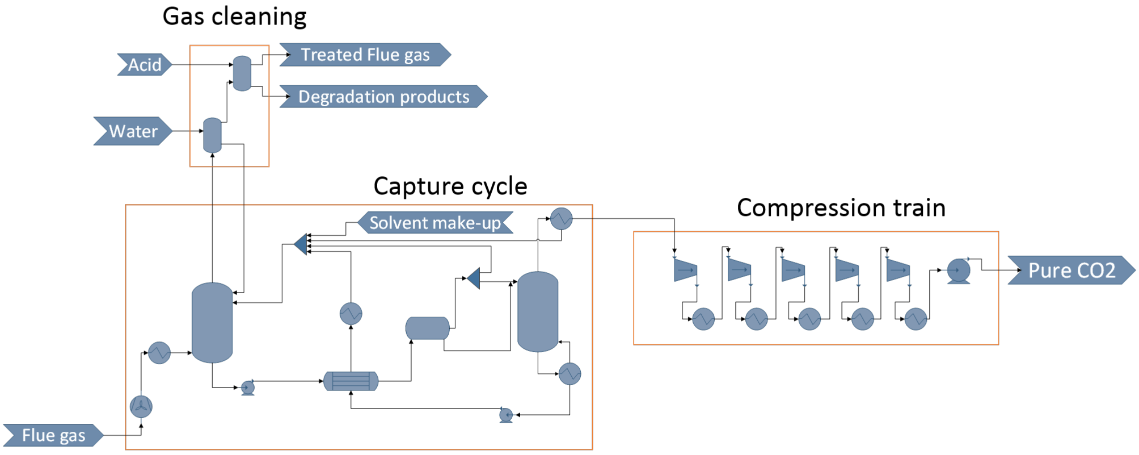

Gas cleaning
Acid
Treated Flue gas
Degradation products
Water
Capture cycle
Solvent make-up
Compression train
Pure CO2
Flue gas

Figure 3.2: The MCA based capture plant PFD presenting main components and flows.

Table 3.2: Specifications of temperatures, reboiler heat load, solvent CO_2 loading and CO_2 end stream purity.

	MEA	MCA
Absorber bottom temperature, °C	57	71
Reboiler temperature, °C	120	101
Reboiler heat load, MW	138.6	68.1
Rich solvent CO_2 loading, $mol_{CO_2}/mol_{solvent}$	0.44	0.7
Lean solvent CO_2 loading, $mol_{CO_2}/mol_{solvent}$	0.32	0.31
CO_2 purity, mol%	100	100

3.1.1 Data problematics

There were two problems with the data of the exhaust gas which this work is based on. Because of a misunderstanding in an earlier stage of the ROLINCAP project, the unit of some of the lowest concentrated components in the exhaust gas were wrongly assumed to be vol% when it really was mg/Nm^3. Because of this, an exhaust gas composition with 5 vol% CO and 0.8 vol% SO_2 instead of 5 mg/Nm^3 CO and 0.8 mg/Nm^3 SO_2 was sent to the actor responsible for the modelling of the capture plant process. The complete composition of the faulty and corrected exhaust gas is displayed in Table 3.3.

Table 3.3: The faulty and corrected exhaust gas composition as volumetric concentration. Numbers marked with * have the unit mg/Nm^3.

Exhaust gas composition [vol%]								
	N_2	Ar	H_2O	CO_2	O_2	CO	SO_x	NO_x
Faulty composition	66.9		10.0	3.3	14.0	5.0	0.8	20.1 (ppm)
Corrected composition	74.2	0.9	8.4	3.6	12.9	5.0*	0.8*	20.1*

The second problem occurred in the modelling of the capture plant process. Typically low concentrated components such as CO, SO_2 and NO_x are excluded from the model since their contribution of mass is negligible. However, because of the unit error of the exhaust gas data, these components accounted for approximately 5.8 vol% of the gas flow. This had the implication that a large portion of the actual gas mass flow was excluded from the model. Because of difficulties modelling O_2, this component had also been excluded from the model. As approximately 14 vol% of the exhaust gases consisted of O_2, a total of almost 19 vol% of the exhaust gas flow was excluded in the model.

Except the consequence that less mass was accounted for in the capture process model, this also increased the share of the remaining exhaust gas components N_2, H_2O and CO_2. Because of this, the capture process was optimised for a higher concentration of CO_2, than intended. The flow and composition of the exhaust gas as modelled is presented in Table 3.4.

Table 3.4: Molar flows and volumetric concentration of the exhaust gas components as modelled.

	N_2+Ar	H_2O	CO_2
Molar flow [mol/s]	14741.4	2200.9	732.9
Concentration [vol%]	83.40	12.45	4.15

Actions to correct this exhaust gas composition was made, but the actor responsible for the capture process modelling did not have time to get results with the corrected composition in time for this thesis. Because of this, the work proceeded using data from the model based on the composition in Table 3.4. The flow and composition of the exhaust gases may not be representative for the Elpedison power plant, but it is nevertheless a realistic flow and composition for a power plant.

3.2 Power plant modelling

To be able to predict and analyse the behaviour of the power plant when integrating the capture plant, a plant model was desired. There was a model available from an earlier stage in the ROLINCAP project, made in the modelling software Ebsilon®Professional. Ebsilon is a good tool for modelling all kinds of thermodynamic processes at steady state, which is suitable for this implementation. Furthermore, it is also good for visualisation purposes.

The Ebsilon model is displayed in Figure 3.3. In the figure, the gas turbine including a generator is located in the lower left corner, while the steam cycle is the rest of the picture. The HRSG, which is represented with the most components, is located in the upper middle of the figure, and the steam turbine with generator is located on the right side, accompanied with the steam condenser above.

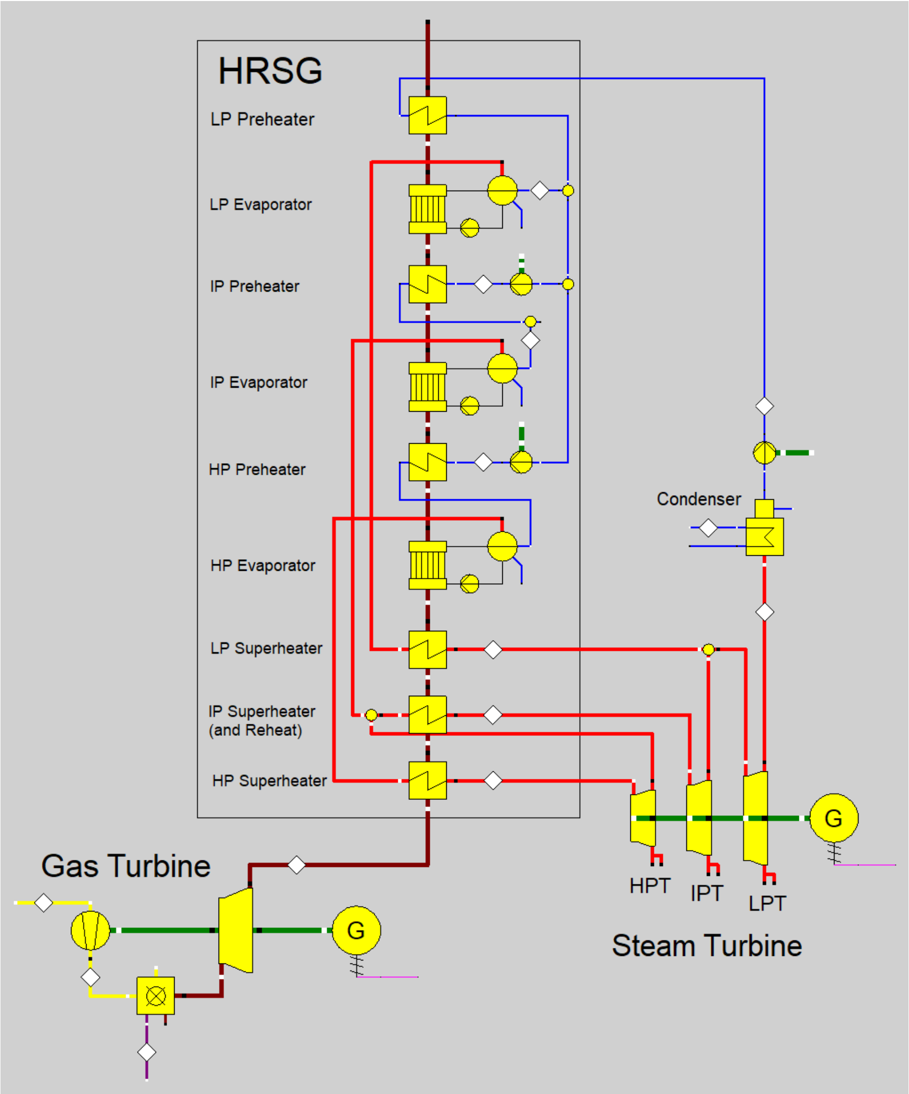

HRSG
LP Preheater
LP Evaporator
IP Preheater
IP Evaporator
HP Preheater
HP Evaporator
LP Superheater
IP Superheater
(and Reheat)
HP Superheater
Condenser
Gas Turbine
G
G
HPT
IPT
LPT
Steam Turbine

Figure 3.3: The Ebsilon model, where the gas turbine is located in the lower left corner, the steam turbine is located on the right side and the HRSG in the middle.

3.2.1 Model refinement

Early in the project, the provided Ebsilon model was refined to better suit data attained from the power plant, and a fuel preheating system that earlier was excluded from the model was added. The input data in the model was changed to average values of the four base load operation points in 2018, where two had been recorded in January and two in August to even out seasonal changes. However, some of the input data needed for the model was not available as average values but only as single operation point values from arbitrary times, so approximations was made to suit the model outputs as close to the average values as possible. The suited data concerned mainly the mass flows, temperatures and pressures in the steam cycle. In table 3.5, the most important data inputs are listed. A discrepancy in the model inputs is the *IP steam & reheat temperature*. This should be 560 °C but this value could not be achieved because it lead to negative temperature differences in heat exchangers, which is why 530 °C was used.

Table 3.5: Important model input data.

Gas Turbine	
Natural gas flow	13.16 kg/s
Combustion air flow (wet)	611 kg/s
Combustion air inlet temperature	17.1 °C
Combustion air inlet pressure	1 bar
Compressor pressure ratio	15.53
Turbine outlet pressure	1.034 bar
Turbine output	233.3 MW

Steam cycle	
HP steam pressure	3.7 bar
IP steam pressure	23 bar
LP steam pressure	112 bar
Condensing pressure	57 mbar
HP feedwater mass flow	265 t/h
IP feedwater mass flow	39.5 t/h
LP feedwater mass flow	28 t/h
Fuel preheater hot water mass flow	24 t/h
HP steam temperature	561 °C
IP steam & reheat temperature	533 °C
LP steam temperature	274 °C
Cooling water inlet temperature	20 °C

3.2.2 Model capture plant heat integration

The capture plant requires heat to regenerate the CO_2-rich solvent in the stripping column, added in the reboiler. As the capture plant was integrated with the power plant, this heat was supplied by the power plant. In Ebsilon, this was modelled by splitting and diverting a selected hot process mass flow to a HX, where it is evaporating a water stream representing the amine mixture. The evaporation temperature is controlled by specifying the corresponding pressure. A controller is used to control the water stream mass flow to make sure the stream is heated with the appropriate amount.

3.3 Process integration

The process integration was based on a method called pinch analysis. Using the pinch analysis, a large number of HENs were systematically created and evaluated economically to make a well-reasoned HEN selection for each solvent. The process integration work was made in the Excel Add-in Pro_PI.

3.3.1 Stream selection for pinch analysis

The stream selection needed to perform pinch analysis initiated the process integration. Some suggestions for process integration were stated in an earlier deliverable in the ROLINCAP project where only the fuel preheater seemed to be the most promising alternative. Since hot IP feedwater is extracted to heat the fuel to 185 °C today there is potential for heat recovery to satisfy this duty instead. Another stream included in the process integration was the exhaust gas leaving the emitting plant. For the absorber to work properly, the incoming exhaust gas needs to be about 40-50 °C. This might be of use to heat a cold stream from the capture plant. Other alternatives from the power plant, like the cooling system for machinery, were discarded because of the low mass flow. For the capture plant there were more options. In the CO_2 compression train there are 5 CO_2 compressors and after each of them there is an intercooler that cools the CO_2 stream from 126 to 40 °C. There is also a need for cooling of lean solvent down to 40 °C before recycling it back to the absorber. The largest need for heat in the capture plant is undoubtedly the stripper reboiler where a large amount of water is evaporated. In the MEA case this happens around 120 °C and for MCA around 101 °C. All of these streams were added to the process integration and they are all stated together with the potential streams from the power plant in Table 3.6. A cold stream indicates a stream with a heating demand and a hot stream indicates a stream with a cooling demand. These streams are also highlighted with blue (cold stream) and red (hot stream) in the power plant and capture plant PFDs in Figures 3.4-3.6.

Table 3.6: Potential streams for process integration

Streams included in pinch analysis	**Type of stream**
Fuel preheating	cold
Exhaust gas leaving the emitting plant	hot
CO_2 cooling in compression train × 5	hot
Lean solvent cooling	hot
Stripper reboiler	cold

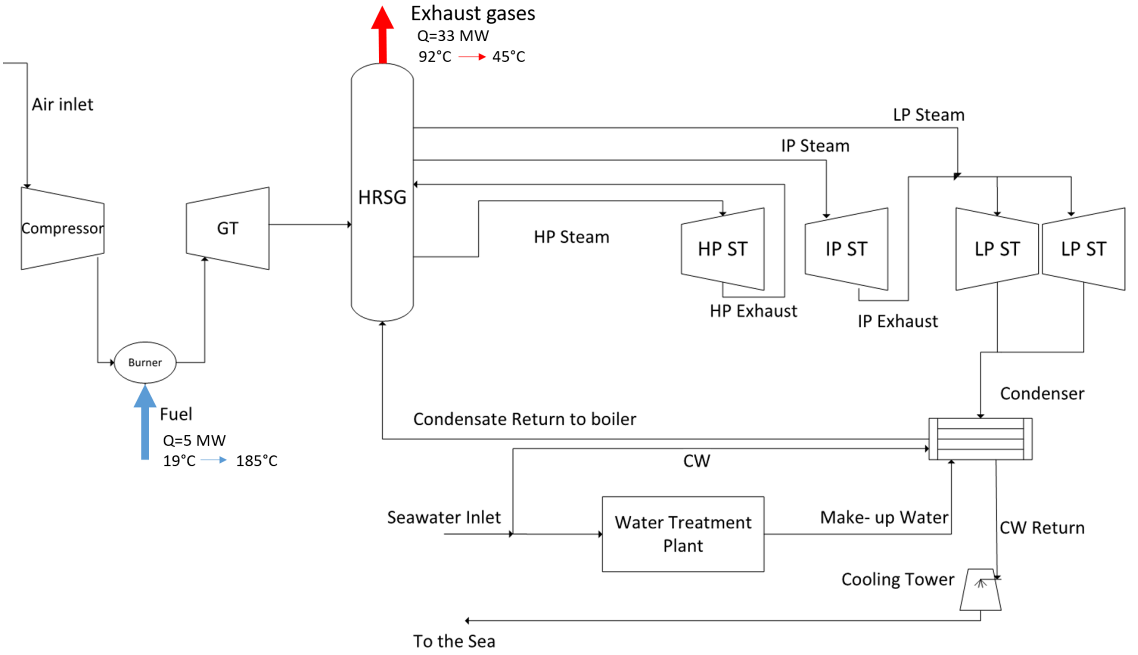

Exhaust gases
Q=33 MW
92°C → 45°C
Air inlet
LP Steam
IP Steam
HRSG
Compressor
GT
HP Steam
HP ST
IP ST
LP ST
LP ST
HP Exhaust
IP Exhaust
Burner
Fuel
Q=5 MW
19°C → 185°C
Condenser
Condensate Return to boiler
CW
Seawater Inlet
Water Treatment Plant
Make- up Water
CW Return
Cooling Tower
To the Sea

Figure 3.4: Schematics of the power plant, where streams selected for pinch analysis are highlighted with colour.

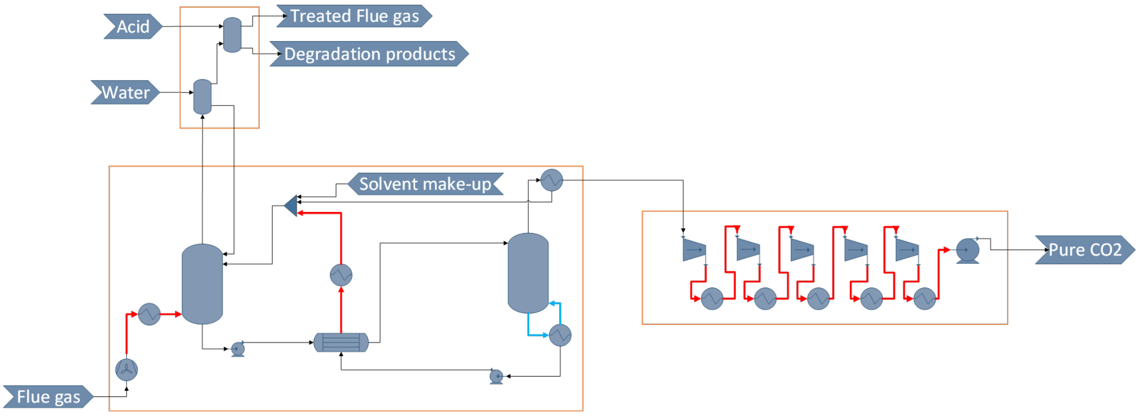

Acid
Treated Flue gas
Degradation products
Water
Solvent make-up
Pure CO2
Flue gas

Figure 3.5: MEA PFD where the streams included in the pinch analysis are highlighted.

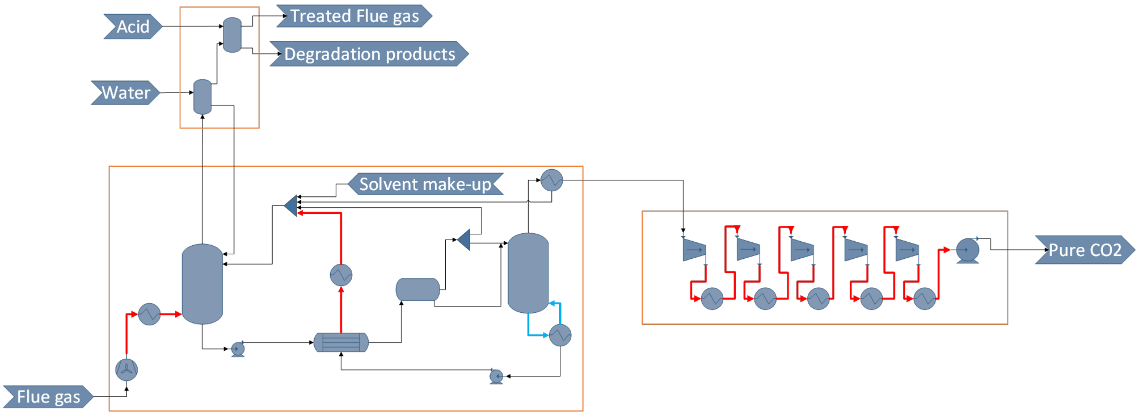


Figure 3.6: MCA PFD where the streams included in the pinch analysis are highlighted.

3.3.2 Heat exchanger network design

When designing HENs, it is hard to know which design will turn out most cost effective. Therefore a number of different designs were created for each solvent following a specific procedure. First, the MER network was constructed for each solvent for three common values of ΔT_{min}: 5, 10 and 15. Then, based on each MER network, a new design was created by removing the smallest (or one of the smallest) internal HXs. This procedure was repeated until a design with only external heaters and coolers was obtained resulting in a number of designs for each ΔT_{min} and solvent.

3.3.3 HEN selection

Only one heat exchanger network was chosen for each solvent to be analysed in the techno-economic assessment. The reason for this is that the techno-economic assessment was at this point assumed to be a rather time consuming process.

The HEN selection was performed by combining the Smith costing method, see Section 2.4.1, and the COWI costing method, see Section 2.4.2. Calculations of pure equipment costs, C_E, were first made on a case with only external heaters and coolers to get an average deviation of HX costs of the two methods. Since COWI's costing method is more time consuming than the Smith method, and also because it required support from a COWI engineer, only the Smith method was used to calculate C_E for all HENs. Using the average deviation between the two methods a cost range for each HX could be calculated. Since additional factors used to reach C_F were available only in the Smith method, these were used on both ends of the range to calculate C_F.

Integrating the capture plant with the power plant will result in a decreased electricity production leading to a decreased revenue. The extra heating requirements of the capture plant were used to add this to the cost estimations. Together with a steam cycle efficiency calculated from the Ebsilon model and a Greek electricity price (without taxes) the cost of decreased electricity production was calculated using Equation 3.1.

$$C_{REP} = Q_{\mathrm{Hreq}}\eta_{ST}tP_{\mathrm{El}} \tag{3.1}$$

where:

- C_{REP} is the cost of reduced electricity production [€/year].
- Q_{Hreq} is the heating requirements for the capture plant after integration [kW].
- η_{ST}=36.7% is the efficiency of the steam turbine [-].
- t is the number of operating hours in a year [h/year].
- P_{El} is the Greek electricity price excluding taxes [€/kWh].

Cooling cost was also added to this cost estimation model. This was calculated using Equation 3.2. The cost and operation parameters were provided by Elpedison and is stated in Table 3.7

$$C_{COOL} = V_{\mathrm{C}}tP_{\mathrm{El}} \tag{3.2}$$

where:

- C_{COOL} is cooling cost [€/year].
- V_{C} is the required flow of cooling water for the capture plant after integration [m^3/h].
- t is the number of operating hours in a year [h/year].
- P_{COOL} is the cooling water price [€/m^3].

Table 3.7: Parameters provided by Elpedison

Price for cooling water	0.08 €/m^3
Average electricity selling price	65 €/MWh
CO_2 emission allowance cost	20 €/ton
Average operating hours	4688 h/year

This cost estimation model also accounts for the time value of money using the NPV approach, see Section 2.4.3. The interest rate, $i = 0.05$, and capital cost payment period, $n = 10$, used in this cost model were based on a calculation example in [21]. One important thing to remember is that these cost estimations are only used for comparing different HEN designs. The rest of the capture plant, columns and compressors etc., was not accounted for since it did not affect which HEN design that got chosen. The design in each of the solvent cases with the lowest total NPV was chosen for the total techno-economic assessment for the whole capture plant.

3.4 Techno-economic assessment

The techno-economic assessment of the suggested designs was based on both the Smith method and the COWI method. In cases where neither of these methods could make valid cost estimations of a specific piece of equipment, vendors or manufacturers of the equipment in question were contacted for cost estimations. The parameters required for cost estimating each type of equipment with the Smith and COWI cost methods are listed in Table 3.8.

Table 3.8: Parameters required for each capture plant component to get valid results from the cost methods.

Component	**Parameters**
Heat exchangers	• Type of HX • Heat transfer area • Pressure levels • Material
Absorber and stripper columns including packing	• Height • Diameter • Material • Wall thickness • Operating Pressure • Operating temperature • Packing height
Decanter	• Vessel dimensions • Material
Pumps	• Type of pump • Pressure head • Volume flow • Material
Compressors/fans	• Volume flow • Type of compressor • Material • Motor size
Cleaning equipment (washers)	• Height • Diameter • Material

3.4.1 Equipment design

Required heat exchanger areas were calculated in the previous stage of the project using Pro_PI. Other parameters for HX, e.g material and type of HX, was determined in discussion with a COWI engineer. In order to make valid cost estimations for the rest of capture plant, and the surrounding modifications it implies, parameters were needed for the rest of the equipment. Some parameters were provided by CERTH, see Section 3.1, while others had to be calculated or approximated. Equipment in contact with aqueous amines naturally suffers from corrosion, [23]. These pieces of equipment in the PFD was therefore designed in high grade stainless steel.

The equipment required to increase the pressure of the exhaust gas before the absorber couldn't be the same for the MEA and MCA case. In the MCA case the absorber works at around atmospheric pressure, which means that only an exhaust gas fan was required to compensate for the pressure drop in the absorber. The absorber in the MEA case, however, operates close to 2 bara. Increasing the pressure of atmospheric exhaust gases up to 2 bara required a compressor since the available fan differential pressure was not enough. Both exhaust gas fan and exhaust gas compressor duties were also calculated in Ebsilon.

3.4.1.1 Column dimensioning

In the Smith method, the cost function for columns is based on column weight. As only data of column heights and diameters were provided in the CERTH equipment sheet, the wall thickness of the columns had to be calculated to obtain the column weights. The material of the columns were stainless steel grade 316 (abbreviated SS316), which has a density of 8000 kg/m^3. To dimension the wall thickness, methodology from literature [24] was used. Since the columns are not highly pressurised, the minimum practical wall thickness was used as dimensioning factor, which also includes a corrosion allowance of 2 mm. However, the thickness was only given for diameter intervals within 1-3.5 meters. Since all of the columns in this thesis are larger than that, the thickness values were plotted in Matlab and a linear fitted function was created from the tabulated values. This plot is presented in Figure 3.7. By extending the fitted function, a minimum wall thickness e could be obtained for all columns.

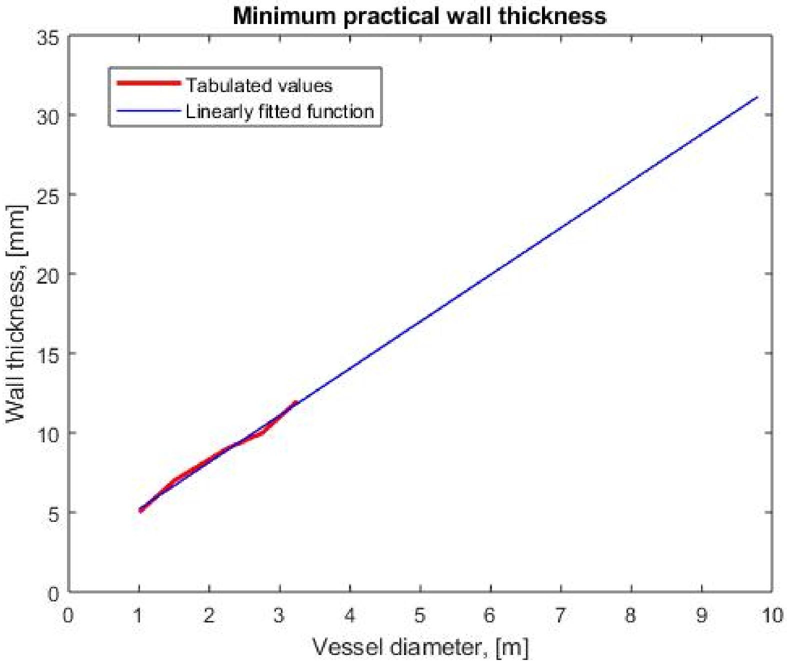


Figure 3.7: Plot of the minimum practical wall thickness values. The blue line is a linear fit of the red curve, which is the tabulated minimum practical wall thickness.

Knowing the wall thickness, the weight of the columns could be determined, using Equation 3.3. For simplification, the columns were approximated as hollow cylinders having a flat top and bottom with the same thickness as the cylinder wall.

$$m_{col} = \rho V = \rho(\frac{\pi h({D_o}^2 - {D_i}^2)}{4} + 2\frac{\pi e {D_o}^2}{4}) \tag{3.3}$$

where:

- m_{col} is the column weight [kg].
- ρ is the density [kg/m^3].
- V is the wall volume [m^3].
- h is the column height [m].
- D_i is the column inner diameter [m]
- e is the column wall thickness [m]
- $D_o = D_i + e$ is the column outer diameter [m].

The packing inside the columns also needed to be accounted for. For the models provided by CERTH a random packing was used, one inch metal pall rings with a specific surface area of 212 m^2/m^3. The cost for the specific packing was obtained by contacting a vendor which provided a price/volume (1790 £/m^3) [25]. The packing cost for each column was then calculated using the packing height and diameter of the columns.

3.4.1.2 Decanter dimensioning

Because of the large flows in the capture process and the lack of information about decanters, the decanter in the MCA capture plant was assumed to be a large tank only using gravitation to separate the phases. The size of the tank was determined using Equation 3.4.

$$V_{tank} = \frac{\dot{m}}{\rho} t_{res} \tag{3.4}$$

where:

- V_{tank} is the tank volume [m^3].
- $\dot{m}$ is the solvent mass flow [kg/s].
- ρ is the density of the solvent [kg/m^3].
- t_{res} is the residence time [s] in the tank.

The residence time required to ensure separation of the solvent phases is based on another investigation of phase change solvents [26], where a few minutes is mentioned to be enough for industrial scale processes. Because of this, a residence time of three minutes was used. From lab scale tests within the ROLINCAP project, results show that the gravitational phase separation is very quick, so three minutes is considered to be a conservative assumption. The tank was assumed to be a horizontally oriented cylinder with a diameter of three meters. Using this and the required volume V_{tank}, the length of the tank could be calculated.

3.4.1.3 Pump dimensioning

Pumps were required in the capture process, although they were not specified in the PFDs. In the MEA case two pumps were needed in the capture process, one pump after each column for the rich and lean solvent flows. In the MCA case some other pumps were required because of the decanter. It is undesired to pump the flow too much before the decanter since it will lead to turbulence in the flow and disturb the gravitational separation process. Extra pumps were therefore added in the outlets of the separation process which pump the flows all the way up to the top of the columns. Another small pump was needed in both solvent cases in the integration with the power plant. The pump was added to push the condensed LP stream from the reboiler back to the steam cycle. This pump is denoted *Reboiler pump* in this thesis.

Pump duties were calculated using Equation 3.5 [24]. The pumps have to account for both head pressure and pressure drops of the equipment. A pressure drop $P_{resistance} = 0.1$ bar was assumed for every kind of equipment in the capture plant. All the equipment was assumed to be on the same ground level, but the streams entering the columns were considered to enter at the top of the columns. The head pressure of this height is accounted for in $Q_{el,head}$. A total pumping efficiency η_{pump} =0.9 was assumed, including both motor and pump efficiencies.

$$Q_{el,pump} = \frac{Q_{el,head} + Q_{el,resistance}}{\eta_{pump}} = \frac{\dot{m}_{solvent} g h + \frac{\dot{m}_{solvent}}{\rho_{solvent}} P_{resistance}}{\eta_{pump}} \tag{3.5}$$

where:

- $Q_{el,pump}$ is the pump duty [W].
- $Q_{el,head}$ is the pump duty [W] from pressure head.
- $Q_{el,resistance}$ is the pump duty [W] from equipment pressure drop.
- η_{pump} is the total pumping efficiency [-].
- $\dot{m}_{solvent}$ is the solvent mass flow [kg/s].
- g is the gravitational acceleration [m/s^2].
- h is the head [m].
- $\rho_{solvent}$ is the density of the solvents [kg/m^3].
- $P_{resistance}$ is the pressure drop [Pa] in the equipment.

3.4.1.4 Exhaust gas fan dimensioning

The fan was initially modelled as a separate component in Ebsilon. It had to pressurise the exhaust gas enough to overcome the absorber pressure drop, which was assumed to be 0.1 bar. However, the power requirement of the fan was far beyond the limits in the intervals of the thesis costing methods, so a number of fan vendors where asked for cost estimations. In their cost estimation, they had dimensioned the fan using the required flow and pressure increase, and therefore the total power output of their recommended fan was used.

3.4.1.5 Exhaust gas compressor dimensioning

The exhaust gas compressor used in the MEA case was dimensioned by modelling it in Ebsilon. The compressor had to pressurise the exhaust gas from 1.033 bara to the operation pressure of the absorber (1.988 bara) plus the pressure drop of the absorber (0.1 bar). It was modelled with an isentropic efficiency of 0.85 and the compressor motor with an electrical efficiency of 0.93. If the compressor size was to big to fit within the Smith costing interval, it was split into a number of smaller units. The maximum limit of the interval was increased with 50% to decrease the number of units. The increase was assumed to be valid.

3.4.1.6 Cleaning equipment

Because of the low share of dust, CO, SO_2 and NO_x in the exhaust gases leaving the emitting plant, no exhaust gas cleaning was needed before the capture plant. However, to limit solvent escape and degradation product from following the gas stream out of the absorber, the equipment described in section 2.1.4 was applied to the capture process: a water wash and an acid wash.

The design parameters of the cleaning equipment were based on the same article, [16]. A model of the cleaning system located after the capture plant was provided during the project. The model was constructed in a software called Aspen and from this model the flows in and out of the columns, the height and the required number of stages could be read among other things. The suggested diameters from Aspen however led to a rather high gas velocity, which is why they were modified to achieve a reasonable value. According to [16] the equipment should use a structured packing. A price for this type of packing however could not be obtained and the same packing as in the columns was therefore also used in the cleaning equipment.

3.4.2 Cost estimation procedure

For cost estimating the whole added capture plant, both COWI and Smith costing methods was used. The equipment cost C_E was calculated using the Smith method (Equation 2.1), for every piece of equipment within the valid range of the cost equation. The equipment cost was also calculated using the COWI method. In cases where no cost estimations were available from either of the methods, vendors were contacted to provide cost estimates.

To get the total capital cost for the plant (C_F), the Smith method was used for both the Smith and COWI equipment cost estimates. All the factors presented in Table 2.1 except f_{OS} was used in the calculations of C_F since the plant is located within an industrial area. The operational costs and the NPV of the whole plant was calculated in the same way as for the HEN selection, see Section 3.3.3. To be able to compare the two solvents some specific metrics were calculated:

- Total CAPEX
- Total annual OPEX

- Total NPV
- CO_2 avoidance cost

CAPEX includes capital costs for every piece of equipment required for the capture plant. OPEX includes cooling cost, cost of decreased electricity production, electricity cost of pumps, compressors and fans, but also the cost relief in CO2 emission allowances. Total NPV includes all future costs, CAPEX and OPEX, translated into today's monetary value to consider the time value of money. The CO_2 avoidance cost, $C_{\text{avoidance}}$ [€/tonne CO_2], was calculated according to Equation 3.6, where the total NPV was divided by the total amount of captured CO_2 in the plant lifetime, $CO_{2,LT}$. The lifetime of the plant was assumed to be 25 years.

$$C_{\text{avoidance}} = \frac{\text{Total NPV}}{CO_{2,LT}} \tag{3.6}$$

In addition to this, the required reboiler energy to capture one tonne of CO_2 was calculated using the reboiler duty, Q_R [W], and the time it takes to capture 1 tonne of CO_2, t_{tonne} [s/tonne CO_2], according to Equation 3.7.

$$\text{Reboiler energy/tonne } \mathrm{CO_2} = Q_R \cdot t_{tonne} \tag{3.7}$$

3.4.3 Power plant efficiency

The efficiency η_{tot} of the power plant was calculated using Equation 3.8, where $P_{net,el}$ $[W]$ is the net power output of the power plant and Q_{in} $[W]$ is the heat input to the plant. To begin with, Q_{in} was calculated for the original power plant using the given efficiency and net power output. Since Q_{in} is constant, only $P_{net,el}$ needed to be calculated for each capture scenario. This was calculated by subtracting the decrease of electricity generation from the steam outtake and increase of electricity usage in the capture plant equipment from the original plant net output power.

$$\eta_{tot} = \frac{P_{net,el}}{Q_{in}} \tag{3.8}$$

4

Results

This chapter presents the integrated power plant model from Ebsilon and the results from the HEN selection where the most cost-effective HENs are displayed. Finally, cost figures from the techno-economic assessment are presented for each solvent.

4.1 Power plant modelling results

4.1.1 Model refinement

The model output should be as close to the plant output as possible. There were many more resulting data points in the model than recorded plant data points. The available and important data points are displayed and compared in Table 4.1. For the temperatures, the change is calculated using kelvin. The modelled values lie within an interval of ±8% from the plant values.

Table 4.1: Important model output data, comparing the output values of the model (Model output) with the given average output values (Aimed output).

	Model output	Aimed output	Deviation
Gas turbine expander inlet temperature	1229 °C	1329 °C	0.938 [-]
Gas turbine expander outlet temperature	616 °C	615.8 °C	1.000 [-]
HRSG exhaust gas exit temperature	99 °C	91.5 °C	1.020 [-]
Steam turbine output	132.17 MW	132.15 MW	1.000 [-]
Plant efficiency	56.0 %	54.8 %	1.022 [-]

4.1.2 Capture plant heat integration

An image showing the schematics of the Ebsilon model with the integrated capture plant is displayed in Figure 4.1, including the fuel preheater and the stripper reboiler. Because of the magnitude of the capture plant heat load, the power plant could only completely satisfy the need by using steam. Since the temperature in the reboiler was much lower than the steam temperatures, steam with as low temperature and heating value as possible was used. Therefore, only LP steam was modelled to heat the reboiler. Another proposed integration was between the fuel preheater and capture plant, to make use of potential waste heat from the capture plant.

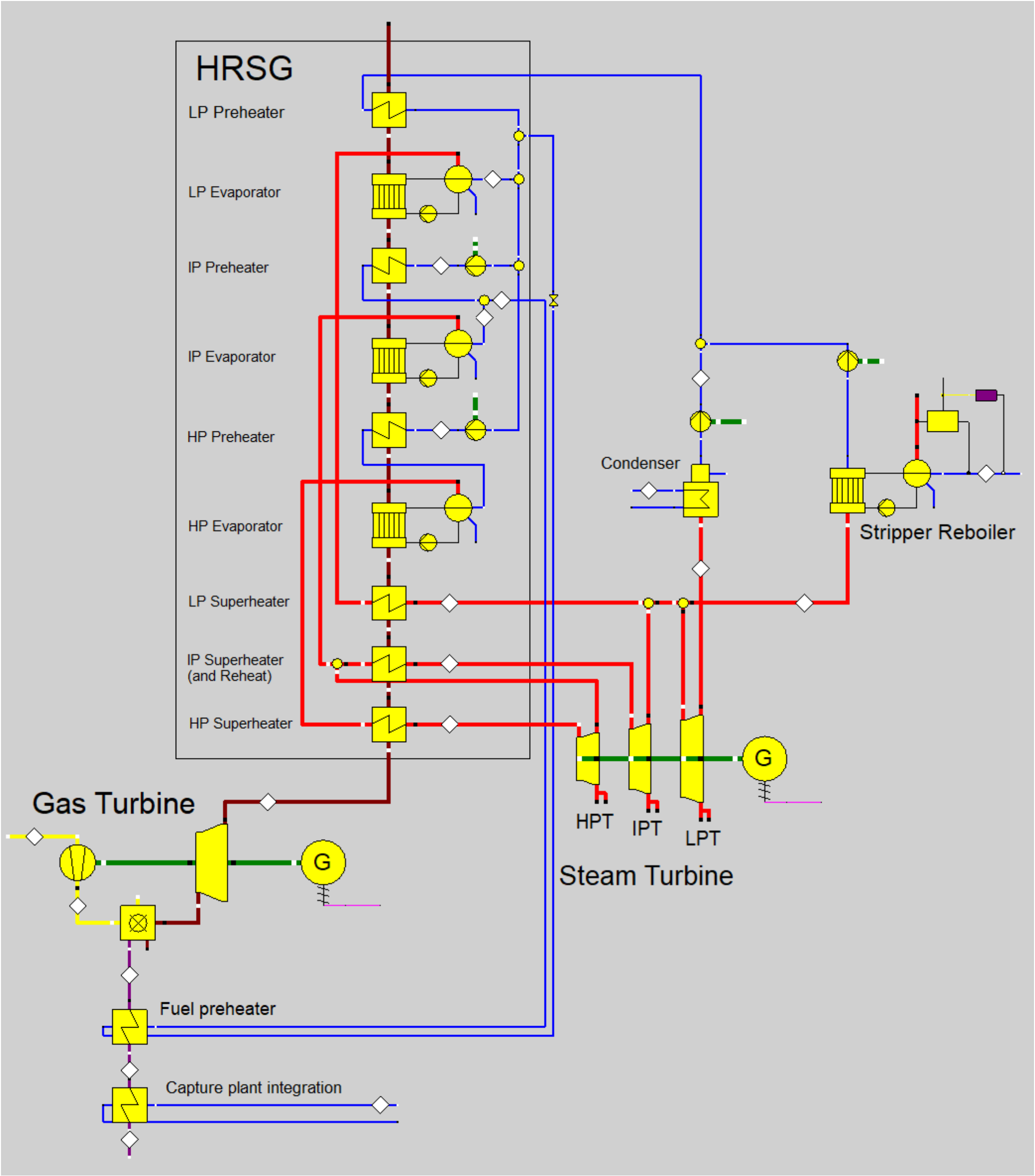

HRSG
LP Preheater
LP Evaporator
IP Preheater
IP Evaporator
HP Preheater
HP Evaporator
LP Superheater
IP Superheater (and Reheat)
HP Superheater
Condenser
Stripper Reboiler
G
HPT
IPT
LPT
Steam Turbine
Gas Turbine
G
Fuel preheater
Capture plant integration

Figure 4.1: The Ebsilon plant model, with the added fuel preheater and two capture plant integrations: LP steam to the reboiler and fuel preheating from the capture plant.

4.2 HEN design and selection results

Using the HEN design procedure described in section 3.3.2, a total number of 38 HENs were created. 8 designs each were created for MCA ΔT_{min}=5, 10 and 15 and MEA ΔT_{min}=5 which means that 7 internal HXs had to be added to reach a MER network. For MEA ΔT_{min}=10 and 15 only 2 internal HXs was needed which resulted in 3 designs each. The total NPV for each HEN design is presented in Figure 4.2 and 4.3. To make the graphs readable, the cost range is not visable. Only a mean value of the cost range is presented for each case. The cost range for the designs however are in the range of ± 2-4 %. In these figures NI represents the *Not Integrated* cases with only external heaters and coolers. NI+1 is based on the NI case, but with one internal HX added to the system. The naming of the rest of the designs follow the same principle until a MER network is reached. The MER networks are therefore denominated NI+7 or NI+2 depending on the case. The designs MEA ΔT_{min}=15 NI+1 and MCA ΔT_{min}=15 NI+2 turned out to be the most cost effective designs for respective solvent and are marked with red rings.

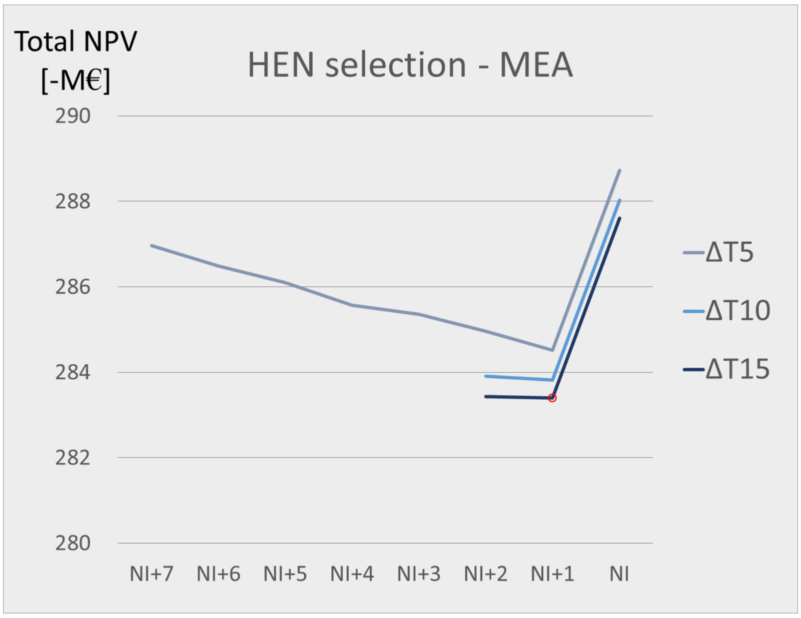


Figure 4.2: Total NPV for each HEN design in the MEA case for three different ΔT_{min}.

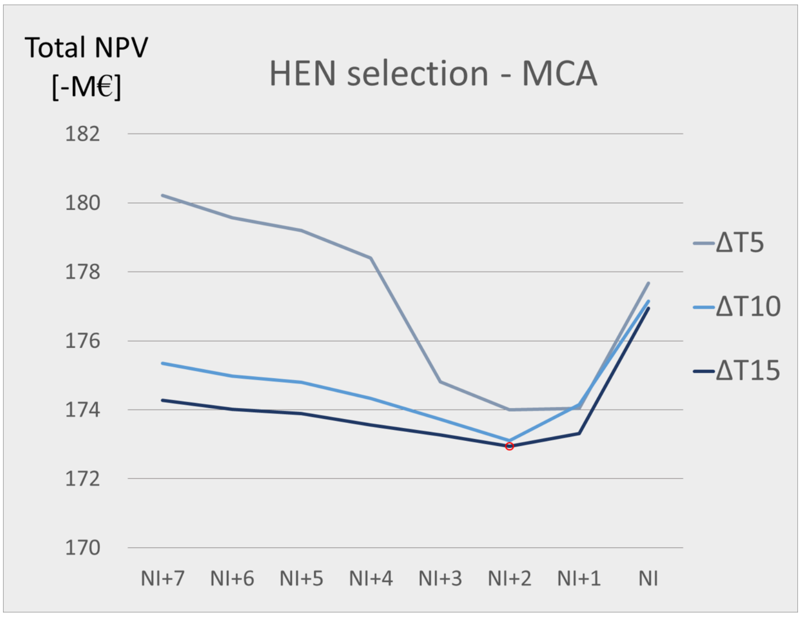


Figure 4.3: Total NPV for each HEN design in the MCA case for three different ΔT_{min}.

4.2.1 Final MEA HEN design

In the MEA case the ΔT_{min}=15 NI+1 design turned out to be the most cost effective design which means that it includes one internal HX. The heat exchanger network is illustrated in Figure 4.4 where yellow, blue and red circles indicate internal HX, external coolers and heaters. The internal HX connects the fuel preheater stream and one of the CO_2 compresson intercooling streams. The fuel stream is heated first by the internal HX to about 112 °C and the rest of the duty is satisfied with the existing fuel preheater using hot feedwater from the steam cycle. The intercooling stream is satisfied completely by the internal HX. The total heat recovery in this design is the duty of the internal HX (Int1), 2.9 MW. The composite curves for MEA ΔT_{min}=15 is shown in Figure 4.5. The maximum energy recovery is rather small compared to the total duty that needs to be satisfied using external heaters and coolers. It is however about the same size as the heat recovered in the chosen design.

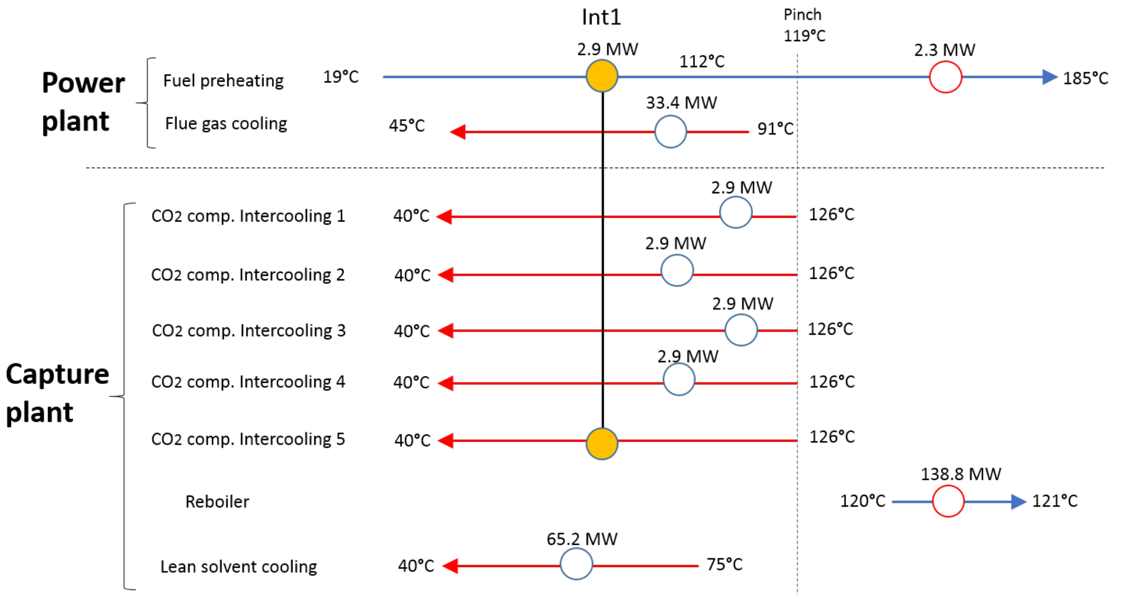

Int1
Pinch
119°C
2.9 MW
112°C
2.3 MW
Power plant
Fuel preheating
19°C
185°C
Flue gas cooling
45°C
33.4 MW
91°C
2.9 MW
CO2 comp. Intercooling 1
40°C
126°C
2.9 MW
CO2 comp. Intercooling 2
40°C
126°C
2.9 MW
CO2 comp. Intercooling 3
40°C
126°C
2.9 MW
Capture plant
CO2 comp. Intercooling 4
40°C
126°C
CO2 comp. Intercooling 5
40°C
126°C
138.8 MW
Reboiler
120°C
121°C
65.2 MW
Lean solvent cooling
40°C
75°C

Figure 4.4: Suggested HEN design for the MEA case.

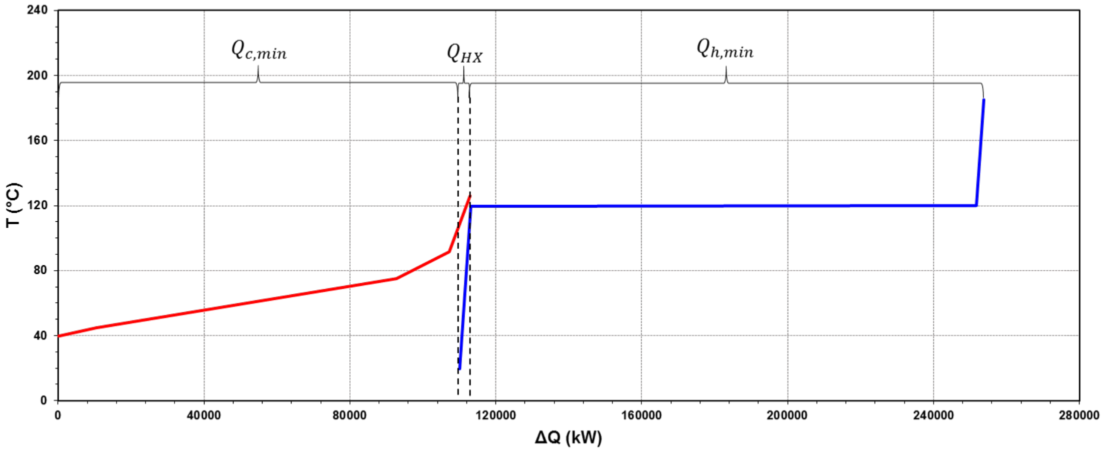


Figure 4.5: Composite curves for MEA ΔT_{min}=15

4.2.2 Final MCA HEN design

In the MCA case the ΔT_{min}=15 NI+2 turned out to be the best design in terms of costs, which means that the design inlcudes two internal HXs. The suggested HEN design is illustrated in Figure 4.6. The total heat recovery is the duties of both the internal HXs (In1 & Int2), 2.7 MW. The MCA ΔT_{min}=15 composite curve is shown in Figure 4.7. Here the maximum available heat to recover is a little larger than in the MEA case, 3.8 MW.

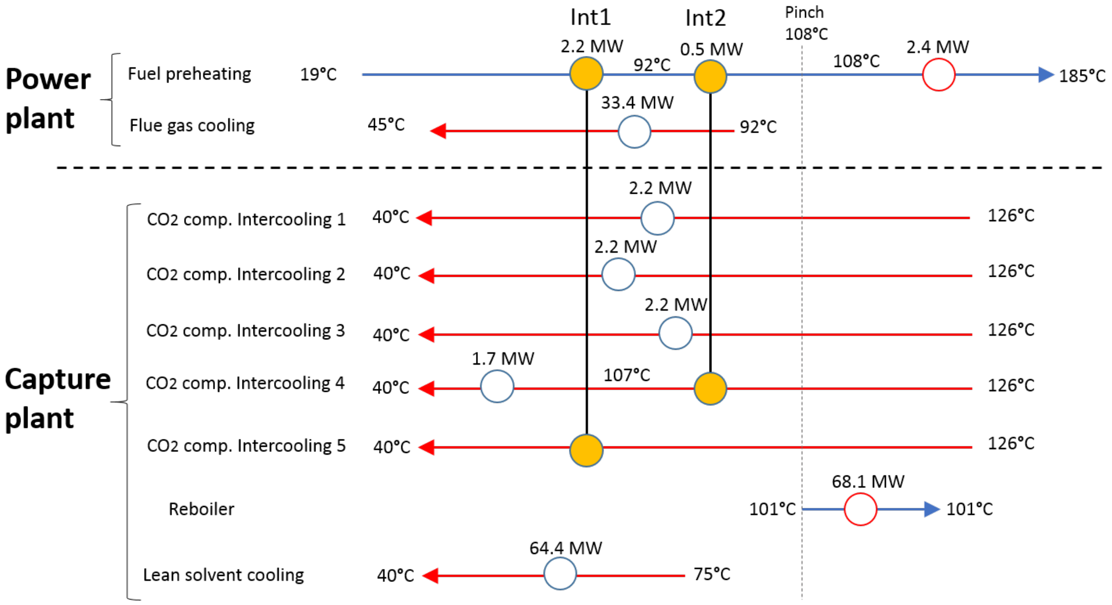

Int1
Int2
Pinch
108°C
Power plant
Fuel preheating
19°C
2.2 MW
92°C
0.5 MW
108°C
2.4 MW
185°C
Flue gas cooling
45°C
33.4 MW
92°C
Capture plant
CO2 comp. Intercooling 1
40°C
2.2 MW
126°C
CO2 comp. Intercooling 2
40°C
2.2 MW
126°C
CO2 comp. Intercooling 3
40°C
2.2 MW
126°C
CO2 comp. Intercooling 4
40°C
1.7 MW
107°C
126°C
CO2 comp. Intercooling 5
40°C
126°C
Reboiler
101°C
68.1 MW
101°C
Lean solvent cooling
40°C
64.4 MW
75°C

Figure 4.6: Suggested HEN design for the MCA case.

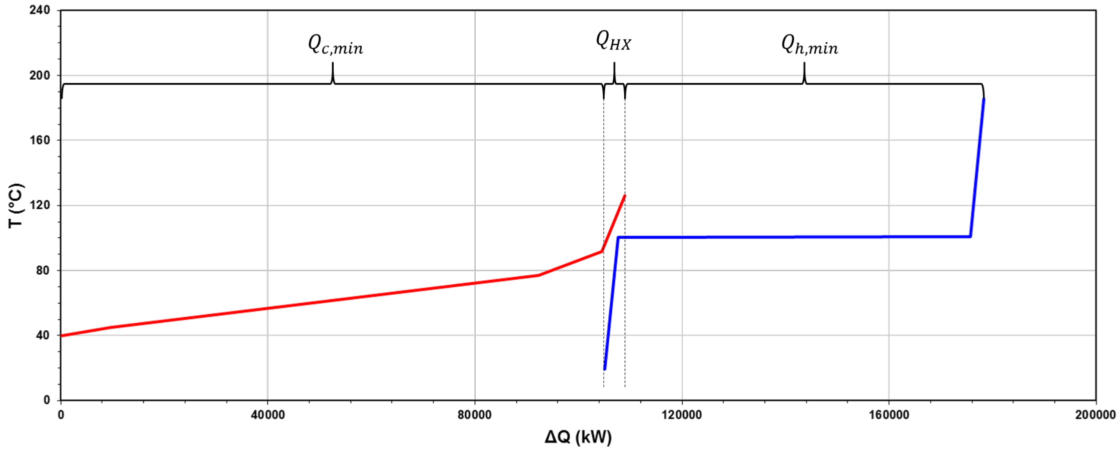


Figure 4.7: Composite curves for MCA ΔT_{min}=15

4.3 Techno-economic assessment results

The results from the techno-economic assessment of the capture plant is presented as capital cost, operating cost and total NPV. The cost figures for the MCA plant is typically lower than for the MEA plant.

Figure 4.8 illustrates the capital costs for the MEA and MCA case calculated with the COWI costing method, the Smith method and a mean value between the two methods. To differentiate the cost drivers of the plant, the costs are divided into capture plant sections: the capture cycle, CO_2 compression, gas cleaning and integration. *Capture cycle* includes equipment related to the solvent such as columns, HXs and pumps, and also the flue gas preparation before entering the absorber (cooling and pressurizing). The *CO_2 compression* includes the whole compression train including intercoolers, compressors and the CO_2 pump. The gas treatment after the absorber is included in *gas cleaning*. The integration between the power plant and capture plant is accounted for in *integration* including internal HXs and additional pumps.

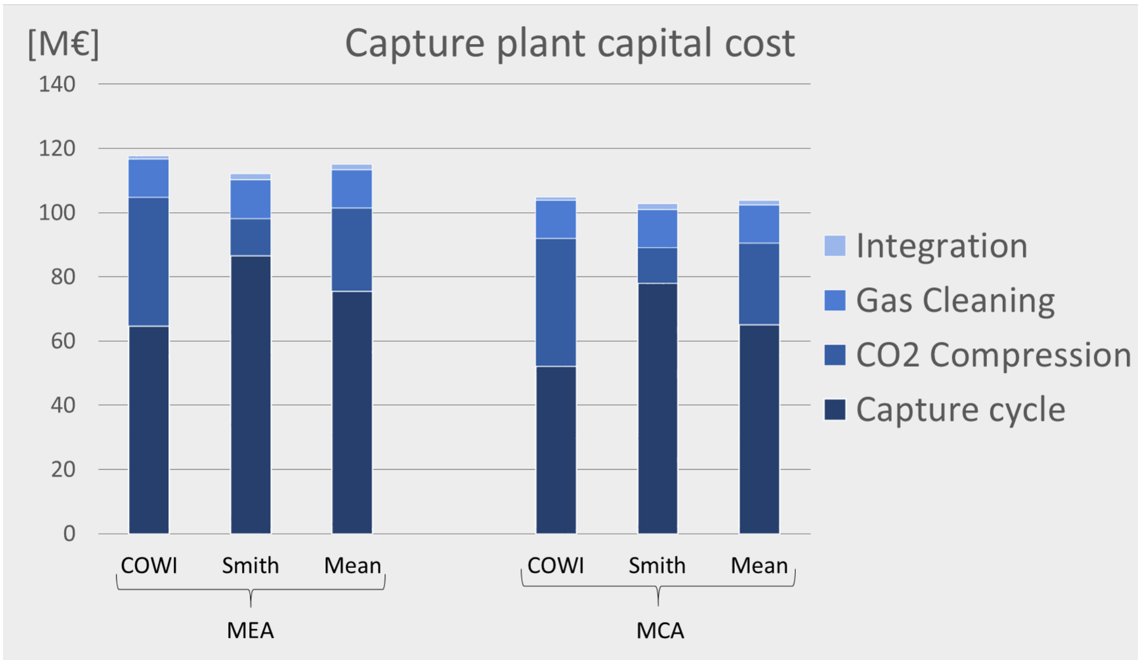


Figure 4.8: Capture plant capital cost.

The total capital cost seem to be quite similar between the solvents. The capture cycle part is bigger in the MEA case mainly because of the required exhaust gas compressor instead of an exhaust gas fan. Even though the total cost is similar between the methods, there are large differences in cost for certain types of equipment. The capital cost contribution from each type of equipment is presented in Table A.1-A.2 in appendix A.

The operating costs of MEA and MCA case are illustrated in Figure 4.9. The total operating cost is divided into cost sections: decreased electricity production, cooling cost and electricity cost. Since 90% of the CO_2 is captured, the annual cost for CO_2 emission allowances decrease with 90% from 10.9 M€ to 1.1 M€ for the power plant. For the capture plant, this decrease in cost is seen as an income of 9.8 M€ yearly, and is labelled as CO_2 emission allowance relief in Figure 4.9. Because of the CO_2 emission allowance relief, MCA have an annual OPEX of 7 M€ instead of 16.8 M€. For MEA, annual OPEX is decreased to 32 M€ , a figure more than four times larger than the annual MCA OPEX. A large contributor to the high MEA OPEX is the cost of electricity for the exhaust gas compressor.

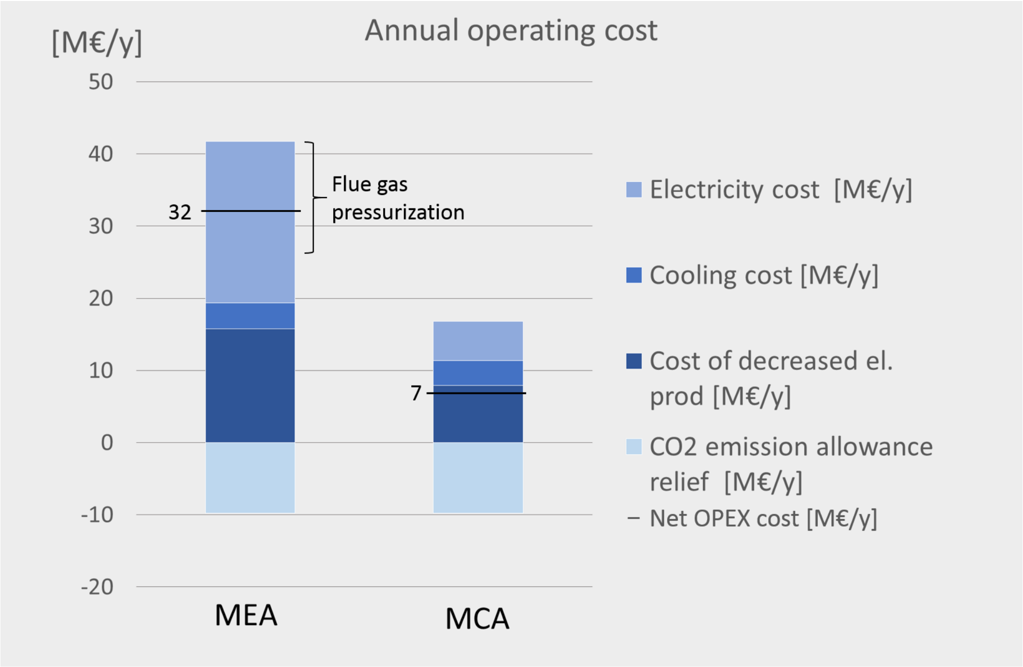


Figure 4.9: OPEX of the MEA and MCA case.

Figure 4.10 illustrates the total NPV of the MEA and MCA case separately divided into CAPEX and OPEX. Since The COWI and Smith method provided such similar results of total capital costs, only a mean value is presented in this graph.

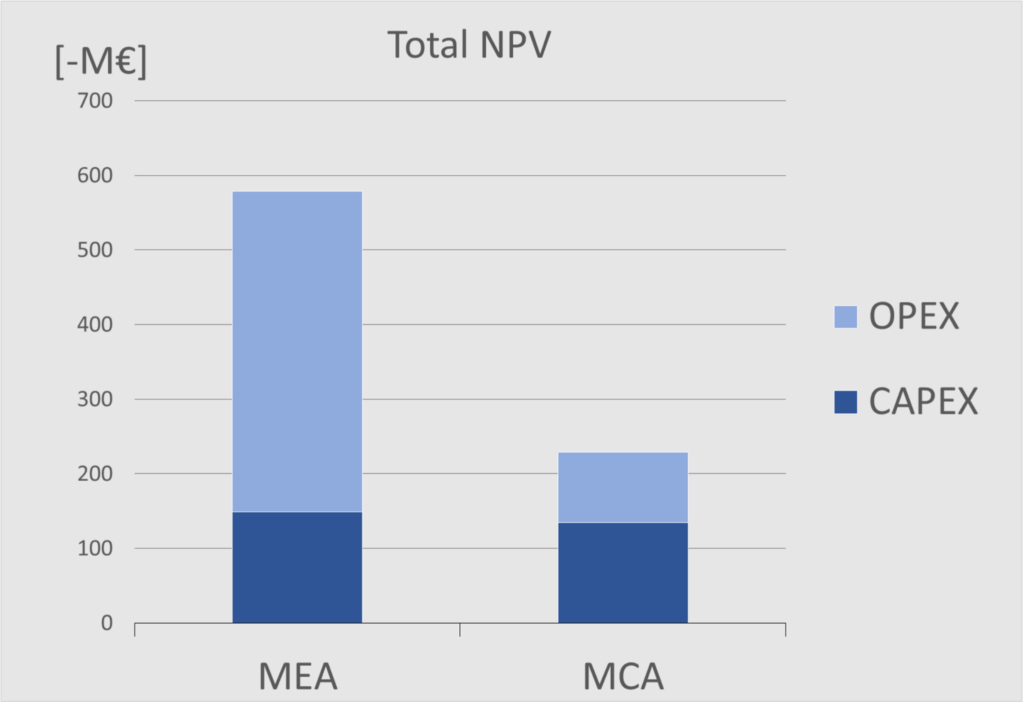


Figure 4.10: Total net present value of CAPEX and OPEX, both MEA and MCA case.

Worth observing is that the NPV in this case is a negative value because the expenditures of the process are substantially higher than the incomes. The results shows that the MEA case has an almost three times larger NPV than MCA, mostly because of the higher operating costs.

Some cost related metrics are presented in Table 4.2 including CO_2 avoidance cost and reboiler energy demand per captured tonne of CO_2. The MCA process is only about a third as costly as MEA per tonne of CO_2. Looking at the reboiler energy demand, the MEA process demands about twice as much heat for solvent regeneration than MCA. Table 4.3 compare the power plant efficiency without CO_2 capture with the two cases where CO_2 capture is applied. By adding the MCA CO_2 capture plant, the power plant will experience an efficiency penalty of 6.8 percentage points, while adding the MEA CO_2 capture plant would increase that penalty to 19.4 percentage points.

Table 4.2: Cost related metrics

	MEA	**MCA**
CO_2 avoidance cost [€/tonne CO_2 captured]	47.3	18.7
Reboiler energy demand [GJ/tonne CO_2 captured]	4.8	2.3

Table 4.3: Power plant efficiency without CO_2 capture, with MCA based CO_2 capture and with MEA based CO_2 capture.

Power plant efficiency	
Without capture	54.8 %
MCA capture	48.0 %
MEA capture	35.4 %

5

Discussion

5.1 Result discussion

The results from the techno-economic assessment show that the total capital cost of the capture plant does not differ much between the two costing methods. However, by looking into certain kinds of equipment it can be concluded that the equipment costs differ quite a lot in some cases depending on which costing method is used. This is visually clear in Figure 4.8. One reason for this is that the Smith method, in some cases, does not specify the type of equipment. For example, when cost estimating compressors only a general cost function for compressors is available in Smith, while in the COWI case the specific type of compressor was selected based on advise from COWI (centrifugal and reciprocating). The same thing applies to HXs and pumps. Columns was the only type of equipment where the equipment costs were more or less similar in both costing methods.

One promising ability of phase-change solvents mentioned in this thesis is that they have a lower regeneration temperature. This allows the use of lower grade waste heat to a greater extent compared to conventional solvents. This was not as noticeable in the results as expected. The MER network in the MCA case could recover 3.8 MW compared to 3 MW in the MEA case. This is an increase of 27%, but this turned out to be not cost effective to make use of. This phase-change ability was therefore not utilised.

All streams with a heating or cooling demand from the capture plant were not included in the pinch analysis. The already included internal HX, labelled R/L (Rich/Lean) exchanger, and the stripper condenser were left untouched. The reason for this is that the R/L exchanger was considered already included in an optimal way and that the stripper condenser operated in quite low temperatures. Also, in the HEN design it turned out to be a deficit of cold streams, so adding another low temperature hot stream would not make any large difference in the resulting design. However, including these streams would enable more possible designs of heat integration, from which a more economical design could have been found.

The HEN selection was in this thesis only based on cost. However, more variables than cost can be important to consider when selecting the best HEN alternative. For example, the flexibility of a design is an important factor. Increasing the complexity of the network will naturally decrease the flexibility and make it less stable to load changes. Including a flexibility analysis might in general change the HEN selection. On the other hand, the winning designs in this thesis consist of only 1 and 2 HXs which is almost the least complex design apart from the unintegrated case.

Three different values of ΔT_{min} was used in the HEN selection: 5, 10 and 15°C. By observing Figures 4.2-4.3 it is clear that ΔT_{min}=15°C was the most cost effective global minimum temperature difference. Since ΔT_{min}=15°C is the higher limit of the temperature interval, it is possible that the cost optimal ΔT_{min} is a value outside of this interval of 5-15°C, for example 20°C.

In the provided PFDs, the MCA process works at approximately atmospheric pressure, while the MEA process works at around 2 bara. Because of this, an exhaust gas compressor was needed to pressurise the gases to 2 bara which requires a lot of electricity and is a big part of the OPEX in the MEA process. Figure 5.1 shows the difference in OPEX between a MEA process with an absorber operating in 1 and 2 bara. The OPEX would decrease about 43%, from 31.9 M€/year to 18.3 M€/year, which would make a significant difference in the overall NPV. In Table 5.1, the power plant efficiency is presented for the different capture processes, where using a MEA capture plant pressure of 1 bara gives an efficiency that is 6.9 percentage points higher than using 2 bara. These numbers are calculated with the assumption that the absorption process would be the same as the pressure is decreased to 1 bara. However, the pressure will probably affect the absorption rate which could also change the flow rates and temperatures in the capture plant. Another solution would be to only expand the gases to 2 bara in the gas turbine, this would however decrease the produced electricity so the gain in efficiency is unsure.

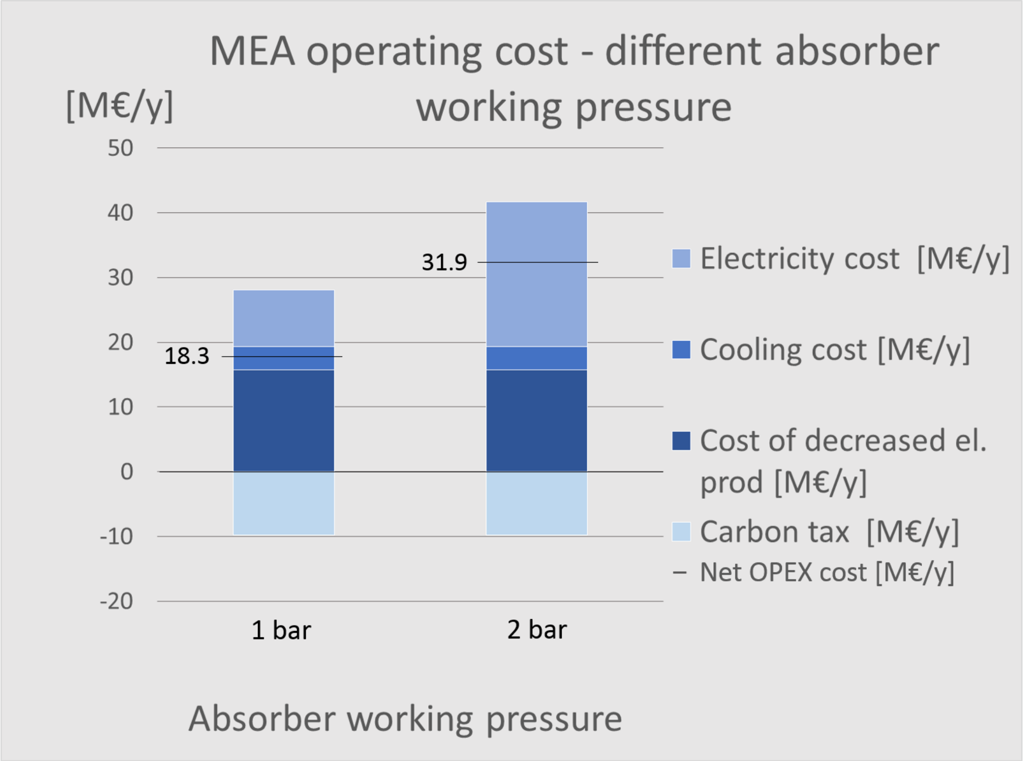


Figure 5.1: MEA OPEX using different working pressure in the absorber and in the rest of the amine cycle.

Table 5.1: Power plant efficiency without CO_2 capture, with MCA based CO_2 capture and with MEA based CO_2 capture for both 1 and 2 bara.

Power plant efficiency	
Without capture	54.8 %
MCA capture	48.0 %
1 bara MEA capture	42.3 %
2 bara MEA capture	35.4 %

The refined Ebsilon model corresponded rather well with the recorded data of the power plant. However, the model was very sensitive to changes, where a change would require several consecutive changes to make the model converge. This sensitivity arose mostly because of constraints of the HRSG streams. Because of this sensitivity, a lot of work was required to create different functional scenarios with an integrated capture plant. Because of this, the model was not used much in this aspect, but more as a reference of the original power plant, where approximate temperatures and flows of the streams were visible. The model could probably become more functional if more constraints were determined for the equipment and less constraints for the streams, but this was not investigated.

5.2 Error sources

This work involves a number of approximations which are potential sources of errors. One main error source is that the flow and composition of the exhaust gas in the CO_2 capture process model differentiate from the real exhaust gases of Elpedison. Except that the composition is wrongly calculated because of unit errors, some components were also excluded from the model. This has increased the CO_2 concentration, which can have an impact on the optimum solvent concentration and flow. Since approximately 19 vol% of the gas flow is excluded from the model, a smaller flow of solvent is used. Because of this, every equipment in the capture plant is undersized to what it should really be. However, the CO_2 concentration and exhaust gas flow used in the model are still realistic for a power plant, which makes the capture plant structure and dimensions valid as a unit. Apart from the different exhaust gas composition, the lower flow could be motivated by using a capture plant that is not designed for high power plant loads, but rather some kind of mid load. If the power plant frequently operates at mid-load, dimensioning the capture for this capacity could potentially be cost-effective.

Using random packing material instead of structured packing in the cleaning equipment is another approximation which may change the results. Structured packing is typically more expensive than random packing so the cost of cleaning equipment should therefore be larger than what is presented in this thesis.

A number of other things were not included because of lack of information. For example initial cost of solvent loading, solvent makeup cost and cost for extra cooling capacity that need to be built. The extra cooling capacity that need to be installed for both the MEA and MCA case is however about the same size, which means that similar costs need to be added to both cases. Cost of solvent would also increase CAPEX, while the solvent makeup cost would increase OPEX.

The heat exchanger networks in this thesis were made following a specific procedure. A MER network was reached for each case, but there are several ways to reach this. Because of this there is a possibility that there are other HEN networks than the ones created in this thesis that might have have been more cost effective. However, if another HEN would be better, the difference in cost is expected to be small relative to the rest of the plant.

The intercoolers in the compression train play an important role in the process integration in this thesis. They are the only link between the capture plant and power plant in terms of heat recovery. The duty of these exchangers were calculated using the mass flow, temperature difference and specific heat of the CO_2 flow. The duty of every intercooler will differ in reality since the specific heat of the flow increases with temperature and pressure indirectly. These duties were instead all approximated as the duty of the first intercooler for simplicity and since they all used the same ΔT_{min} for cooling. The carbon dioxide flow through the compression train is the same for MEA and MCA, but in the MEA case some water was present in the model. This led to differences in duties for the intercoolers between the two cases.

5.3 Outlooks

In the ROLINCAP project, the ROLINCAP solvent is mentioned as the most promising phase-change solvent in the study. It is therefore worth to investigate the potential of this solvent, which is likely to decrease the solvent regeneration energy demand and by that the cost of CO_2 capture. Also promising is the use of RPB columns instead of PB columns. It will not only decrease the column sizes and by that capital cost, the enhanced mass transfer enables use of more viscous solvents, such as a more concentrated amine solvent. Increasing the amine concentration could potentially increase the CO_2 loading per kg of solvent, which would decrease the required solvent mass flow. This would decrease the size of the rest of the equipment also, not only the columns, potentially decreasing capital costs significantly. However, since the RPB require energy to rotate, it is important that it does not use so much electricity that the operational costs rise and replace the lost capital costs.

For further work, it is recommended to perform sensitivity analysis of the plant, to determine the sensitivity to price differences in electricity, natural gas or CO_2 emission allowance cost.

6

Conclusions

Process integration has been performed on an emitting plant and a capture plant for post combustion carbon capture. Conventional solvent MEA and phase-change solvent MCA are the solvents under investigation and one cost-optimal heat exchanger network for each type of solvent has been further investigated in a techno-economic assessment. Several conclusions can be made based on the results.

Not much energy was recovered in the process integration between the power plant and capture plant. For MEA, approximately 3 MW of heat was possible to recover through process integration between the power plant and the capture plant, and almost the same amount was cost-effective to recover. However, for MCA, approximately 4 MW was possible to recover but only approximately 3 MW of this was cost-effective to make use of. These 3 MW is only a small percentage of the total reboiler heat load, so the heat recovery is considered low. Since the power plant is already optimised, there are few sources of waste heat to use in the capture plant.

The power plant suffer from large efficiency penalties from adding CO_2 capture. The power plant efficiency is decreased from 54.8% to 48.0% by adding the MCA CO_2 capture plant and to 35.4% by adding the MEA CO_2 capture plant.

Both CAPEX and OPEX are lower for the MCA capture plant than the MEA capture plant. The MCA capture process requires approximately half as much energy for solvent regeneration than the MEA CO_2 capture process and it requires less electricity for pump and fan work. The operational costs for MCA is therefore drastically lower than the MEA operational cost. Because of this, the cost of capturing one tonne of CO_2 is much lower for MCA than MEA, 18.7 €/tonne CO_2 compared to 47.3 €/tonne CO_2.

Another conclusion is that the costing methods used provide very different results for specific components, but at the same time provide very similar results when it comes to total plant capital cost. It is therefore important to investigate which costs seem more realistic based on costs from earlier projects.

A big part of the MEA operating cost arises because of the high absorber operating pressure of 2 bara, which require energy intensive and expensive compressor work of the exhaust gases. It is therefore crucial to investigate whether 1 bara is suitable for absorption or if 2 bara is required to acquire a good absorption process.

In total, phase-change solvents do seem like a promising technology compared to the conventional MEA technology. It has the potential to decrease the cost significantly, which in time could lead to actual use of post combustion CO_2 capture in power plants and not only implementations of research purpose.

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A

Appendix 1

In Tables A.1 and A.2, the contribution to the total plant CAPEX C_F from each piece of equipment is presented for each solvent. In the second column the total equipment size is presented for each equipment, which was split up into two or more units if required. For some equipment a reasonable cost was only available in one of the methods or from vendors. That cost was in these cases used for both methods.

Table A.1: Installed equipment cost for entire capture plant, MEA case.

Description	**Total sizes**	**Investment cost [k€]**			
		COWI	#	**Smith**	#
Columns					
Absorber	D=9.5m, H=32m	27,442 (23.3%)	1	25,612 (22.8%)	1
Stripper	D=4m, H=18.5m	3,244 (2.8%)	1	2,295 (2.9%)	1
Heat Exchangers					
CL1	Q=2.86MW, A=423m^2	366 (0.3%)	1	662 (0.6%)	1
CL2	Q=2.86MW, A=325m^2	318 (0.3%)	1	553 (0.5%)	1
CL3	Q=2.86MW, A=266m^2	297 (0.3%)	1	724 (0.6%)	1
CL4	Q=2.86MW, A=226m^2	297 (0.3%)	1	649 (0.6%)	1
Int1	Q=2.86MW, A=1148m^2	875 (0.7%)	1	1959 (1.7%)	1
LNCOOLER	Q=65.23MW, A=2117m^2	461 (0.8%)	2	6,731 (6.0%)	1
FGCOOLER	Q=33.36MW, A=8633m^2	4,214 (3.6%)	1	2,440 (6.52%)	3
REBOILER	Q=138.6MW, A=1357m^2	2,083 (1.8%)	1	4,974 (4.4%)	1
R/L HX	Q=75.68MW, A=4165m^2	583 (1.5%)	3	6,657 (11.9%)	2
Stripper Cond.	Q=67.82MW, A=2896m^2	800 (1.4%)	2	8,331 (7.4%)	1
Compressors					
FG Comp.	Q=51MW, ΔP=1 bar	(Smith) (13.2%)	4	3,872 (13.8%)	4
CO2 Comp. 1	Q=2.15MW, ΔP=1.4 bar	6,450 (5.5%)	1	1,721 (1.5%)	1
CO2 Comp. 2	Q=2.11MW, ΔP=3.2 bar	4,505 (3.8%)	1	1,708 (1.5%)	1
CO2 Comp. 3	Q=2MW, ΔP=7.6 bar	3,795 (3.2%)	1	1.667 (1.5%)	1
CO2 Comp. 4	Q=1.98MW, ΔP=18 bar	9,816 (8.3%)	1	1.657 (1.5%)	1
CO2 Comp. 5	Q=1.97MW, ΔP=42.6 bar	9,773 (8.3%)	1	1.653 (1.5%)	1
Pumps					
Rich solvent pump	Q=236kW, Head=17m	1,314 (3.4%)	3	603 (0.5%)	1
Lean solvent pump	Q=440kW, Head=34m	1,314 (3.4%)	3	849 (0.8%)	1
CO$_2$ pump	Q=468kW, ΔP=76.2 bar	4,653 (4.0%)	1	879 (0.8%)	1
Reboiler pump	Q=0.73kW, ΔP=0.1 bar	77 (0.1%)	1	25 (0%)	1
Other equipment					
Water wash	D=9m, H=18m	5,659 (4.8%)	1	(COWI) (5.0%)	1
Acid wash	D=9m, H=18m	6,242 (5.3%)	1	(COWI) (5.6%)	1
Total equipment cost		117 750 (100%)		112 277 (100%)	

Table A.2: Installed equipment cost for entire capture plant, MCA case.

Description	Total Size	Investment cost [k€]			
		COWI	#	Smith	#
Columns					
Absorber	D=9.8m, H=33.4m	30,267 (28.8%)	1	28,273 (27.4%)	1
Stripper	D=3.5m, H=17.3m	2,630 (2.5%)	1	2,403 (2.3%)	1
Heat Exchangers					
CL1	Q=2.23MW, A=330m^2	281 (0.3%)	1	560 (0.5%)	1
CL2	Q=2.23MW, A=254m^2	281 (0.3%)	1	468 (0.5%)	1
CL3	Q=2.23MW, A=208m^2	265 (0.3%)	1	612 (0.6%)	1
CL4	Q=2.23MW, A=258m^2	265 (0.3%)	1	508 (0.5%)	1
Int1	Q=2.23MW, A=574m^2	557 (0.5%)	1	1,223 (1.2%)	1
Int2	Q=0.49MW, A=220m^2	334 (0.3%)	1	637 (0.6%)	1
LNCOOLER	Q=64.43MW, A=2043m^2	429 (0.8%)	2	6,572 (6.4%)	1
FGCOOLER	Q=33.36MW, A=8633m^2	4,214 (4.0%)	1	2,440 (7.1%)	3
REBOILER	Q=68.08MW, A=628m^2	1,251 (1.2%)	1	2,946 (2.9%)	1
R/L HX	Q=79.82MW, A=4651m^2	647 (1.9%)	3	7,176 (13.9%)	2
Stripper Cond.	Q=64.43MW, A=2789m^2	779 (1.5%)	2	8,120 (7.9%)	1
Blowers and Compressors					
FG Blower	Q=6,7MW, ΔP=0.1 bar	1,762 (5.0%) (Vendor)	3	1,762 (5.1%) (Vendor)	3
CO2 Comp. 1	Q=2.15MW, ΔP=1.4 bar	6,450 (6.1%)	1	1,721 (1.7%)	1
CO2 Comp. 2	Q=2.11MW, ΔP=3.2 bar	4,505 (4.3%)	1	1,708 (1.7%)	1
CO2 Comp. 3	Q=2MW, ΔP=7.6 bar	3,795 (3.6%)	1	1,667 (1.6%)	1
CO2 Comp. 4	Q=1.98MW, ΔP=18 bar	9,816 (9.4%)	1	1,657 (1.6%)	1
CO2 Comp. 5	Q=1.97MW, ΔP=42.6 bar	9,773 (9.3%)	1	1,653 (1.6%)	1
Pumps					
P1	Q=20kW, Head=2m	403 (0.8%)	2	155 (0.2%)	1
P2	Q=137kW, Head=34m	784 (0.8%)	1	447 (0.4%)	1
P3	Q=56kW, Head=17m	482 (0.5%)	1	273 (0.3%)	1
P4	Q=16kW, Head=17m	228 (0.2%)	1	137 (0.1%)	1
P5	Q=80kW, Head=34m	504 (0.5%)	1	333 (0.3%)	1
CO_2 pump	Q=468kW, ΔP=76.2 bar	4,653 (4.4%)	1	879 (0.9%)	1
Reboiler pump	Q=0.36kW, ΔP=0,1 bar	44 (0%)	1	17 (0%)	1
Other equipment					
Water wash	D=9m, H=18	5,606 (5.3%)	1	(COWI) (5.4%)	1
Acid wash	D=9, H=18	6,242 (6.0%)	1	(COWI) (6.1%)	1
Decanter	D=3, H=15.5	(Smith) (1.4%)	1	1,309 (1.4%)	1
Total equipment cost		104 984 (100%)		103 083 (100%)	