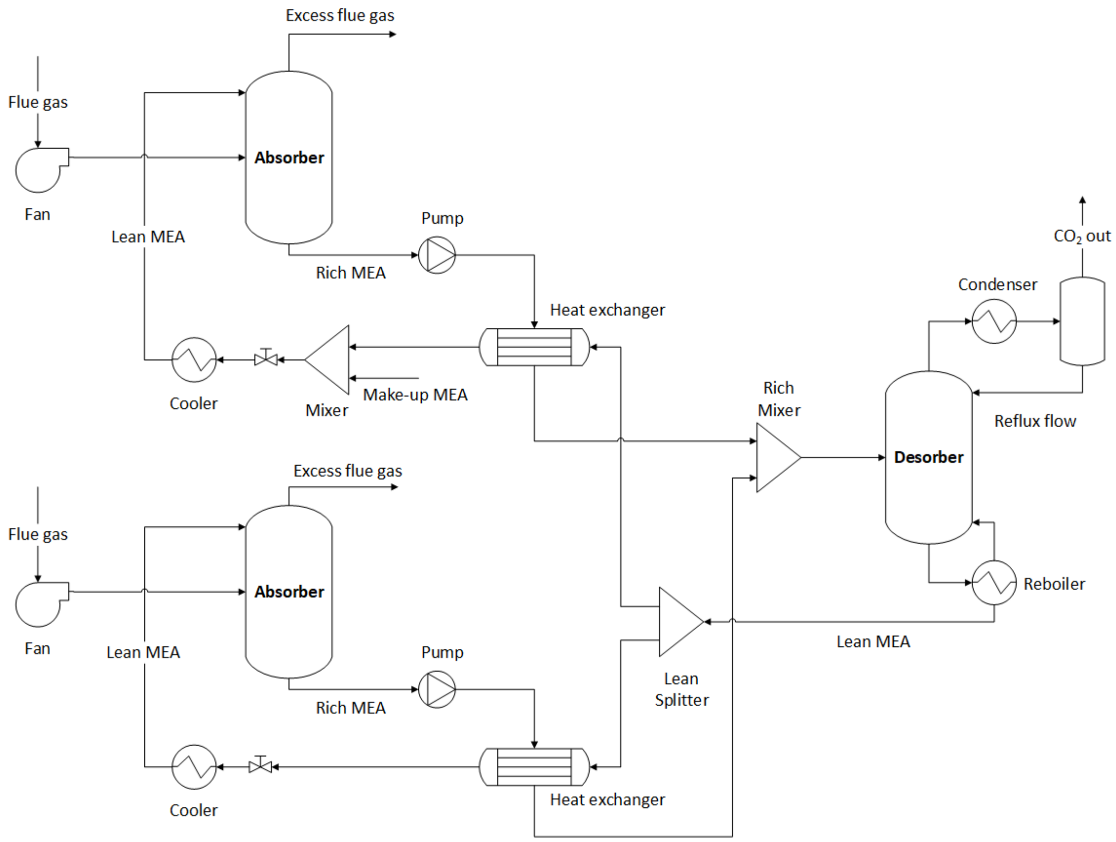

Local Infrastructures for CCS Clusters

A Case Study of Two CHP Plants in Gothenburg

Master's Thesis in Innovative and Sustainable Chemical Engineering

VICTORIA IGNELL AND EMELIE JOHANSSON

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CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2021
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Cover: Flowsheet over a post-combustion process, with monoethanolamine as solvent, for two companies with separate absorbers and a shared desorber.

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Abstract

Climate change is a global problem and measures to reduce CO_2 emissions are required. One acknowledged measure, both in Sweden and globally, is carbon capture and storage (CCS). This thesis investigates economic benefits for infrastructure cooperation in CCS implementation including solvent regeneration, liquefaction and transport to an intermediate storage.

Two combined heat and power (CHP) plants in Sävenäs owned by Renova and Göteborg Energi were investigated as a case study. The capture plants were modelled in Aspen Plus as post-combustion processes with monoethanolamine as a solvent. Heat demand and other utilities were quantified to decide the impact on the heat and electricity production at each plant. Based on the utility demand and equipment need, an economic analysis was performed. The cost for CO_2 liquefaction and transport were estimated. Four scenarios for the operation of the CHP plants including maintained fuel usage, maintained district heat delivery, collaboration to maintain district heat delivery and reduced capture rate, were compared.

The work concludes that, depending on scenario, the capture cost for separate plants at Renova and Göteborg Energi is 39-46 and 80-82 €/ton CO_2, respectively. For a capture plant with a shared infrastructure, the capture cost is 45-52 €/ton CO_2. The result shows a possible district heat recovery of 70% from the capture plant, which is important to decrease the cost of the heat demand. The heat integration with the steam cycle is important for the capture cost. Use of primary steam leads to higher variable OPEX, due to higher reductions in electricity generation, but lower impact on district heating generation, than the use of an extracted stream from the turbine. The plant utilisation is more important for the specific capture cost than the size of the CHP plant. Depending on the allocation of the cost between the companies, the possible savings are between 0.2-1.0 M€/yr corresponding to 2-5% of the total cost. Therefore, there are small economic benefits with a shared process where shared liquefaction has the most potential.

Keywords: CCS, CHP, CO_2 Transport, Liquefaction, CO_2 Storage, Cluster, CCS Infrastructure, MEA, Gothenburg.

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Victoria Ignell and Emelie Johansson, Gothenburg, June 2021

Contents

Abbreviations

CAPEX Capital Expenditures

CCS Carbon Capture and Storage

CHP Combined Heat and Power

CO_2 Carbon Dioxide

DCC Direct Contact Cooler

EIC Equipment Installed Cost

GHG Greenhouse Gas Emissions

MEA Monoethanolamine

NaOH Sodium Hydroxide

NO_2 Nitrogen Dioxide

NOK Norwegian Kroner

OPEX Operating Expenditures

SNCR Selective Non Catalytic Reduction

SO_2 Sulphur Dioxide

SO_x Sulphur Oxide

ss Stainless Steel

TDC Total Delivered Cost

Nomenclature

$\alpha - value$ The ratio between produced electricity and heat

Δh_{cond} Condensation enthalpy of water

ΔP Pressure difference

ΔT_{lm} Logarithmic temperature difference

$\dot{m}$ Mass flow

$\dot{n}$ Molar flow

$\dot{V}_{CO_2}$ Volumetric flow of CO_2 per day

η Efficiency

λ Thermal conductivity

μ Dynamic viscosity

ρ Density

A Area

c_p Heat capacity

C_{truck} Investment cost per truck

d Diameter

$f_{administration}$ Administration cost

$f_{contingency}$ Contingency factor

f_{direct} Direct cost

$f_{engineering}$ Engineering cost

$f_{equipment}$ Equipment factor

f_{inner} Friction factor inside tubes

$f_{location}$ Location factor

$f_{material}$ Material factor

f_{piping} Piping factor

$F_{Total,CS}$ Total material factor for carbon steel

$F_{total,material}$ Total material factor

G Mass velocity

h Convective heat transfer coefficient

L Length

N Number

n Operating lifetime

n_{truck} Lifetime per truck

P Pressure

p Interest rate

R Gas constant

r Annuity factor

Re Reynolds number

T Temperature

$T_{capacity}$ Transport capacity of a truck

$t_{operating}$ Operating hours per week

t_{trip} Time for one trip

t_{work} Work hours per day per driver

U Overall heat transfer coefficient

W Work

x Mole fraction

Q Heat transferred

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1 Introduction

The present carbon dioxide (CO_2) concentration in the atmosphere is the highest reached value in the last 800 000 years. Since the pre-industrial era the CO_2 concentration has increased drastically due to anthropogenic emissions. The greenhouse gas (GHG) emissions have proven to cause changes to the climate and these can be observed over the whole world. Humans, animals, plants and entire ecosystems are affected by these changes through for example an increased risk of fires and heavy rainfalls [1]. Climate change is a global problem and with the Paris Agreement, the parties should work together to ensure that the average temperature, compared to the levels before the industrialisation, should not exceed an global increase of 2 °C. However, they should also work for not exceeding 1.5 °C as it decreases the risk for great impacts from climate change [2].

To reach the targets set out by the Paris Agreement the Swedish government has established a long term goal, stating that Sweden should have reached net zero emissions of GHG until the Year 2045. After this year, negative emissions should be achieved. To accomplish this several methods and technologies have been suggested, such as shifting to biobased fuels, increase energy and material efficiency and applications of carbon capture and storage (CCS) [3]. Both the European Commission and the Intergovernmental Panel on Climate Change identifies CCS as an essential technology to decrease the emissions of GHG [4, 5].

CCS is a technology which involves capture and compression of CO_2 from industrial flue gases before transport to a permanent storage, e.g. in geological formations. The most studied and used method of CCS is using a solvent [6], commonly monoethanolamine (MEA) [7], to absorb the CO_2 [6]. The solvent needs to be regenerated and this is performed by heating in a desorber. The CO_2 is released as a gas after which the lean solvent can be reused in the absorber. When introducing CCS to an already existing industry, the heat demand for the solvent regeneration [7] and the electricity demand for the CO_2 compression [8] have to be satisfied [6]. However, CCS can lead to drastically decreased emissions of CO_2 and if CCS is implemented in biobased industries it will lead to negative emissions [6].

One industry sector where CCS can be useful is for combined heat and power (CHP) plants, which are both electricity and district heat producers [9]. Overall, considering the GHG emissions in Sweden the CHP plants had a contribution of 7% in 2019 [10]. The fuel input in CHP plants have gone more towards biofuels from fossil fuels [10] and a higher use of waste for incineration [11]. The waste can be both a biofuel and a fossil fuel depending on the different components [11]. In Sweden, the fossil based carbon share of the waste is around 30% [12].

When considering carbon capture as a measure to reduce emissions in industries, the investment and operating costs are important due to the economic risk but also for the technology value. The latter is influenced by the governmental policies through the CO_2 emissions and the associated cost, since CCS will be valuable if the cost for capturing CO_2 is less than for emitting it [13]. To decrease the cost, especially concerning the infrastructure for transport of CO_2, there exist different opportunities. One way is through having a cluster of industries which cooperate with the transport, storage and the knowledge required. Another benefit is reduced economic risk for the companies investing in CCS due to several actors sharing the risk [14]. A possibility for collaboration could also include having a shared facility for solvent regeneration. The effect of having a shared regeneration facility and transport to intermediate harbour storage is investigated in this thesis to research if the cost associated with CCS implementation can be decreased. This could present possibilities for further installation of CCS in industries, which is important in order to decrease emissions.

1.1 Aim and Scope

The work investigates possibilities for local collaboration on CCS infrastructures with the aim to identify economic benefits with shared infrastructures. Solutions for the infrastructure are evaluated based on the impact on the heat and electricity generation, heating and cooling demands in the capture plant, as well as the effects on capital and operating expenditures for the capture plant, liquefaction and CO_2 transport.

The focus of the work is on a case study of two CHP plants in Sävenäs, Gothenburg, owned by Renova and Göteborg Energi and the CCS technology considered is post-combustion capture with MEA as the solvent. The regeneration of the MEA solvent, CO_2 liquefaction and the transport to an intermediate storage in the Port of Gothenburg are considered as opportunities for collaboration.

2
Theory

The theory for CCS and CO_2 transport modes are described in this chapter. CHP plants are also described both in general and in more detail for the two companies in the case study. The policies which are of importance for CHP plants and for carbon capture plants are explained in the end of this chapter.

2.1 Carbon Capture and Storage

CCS is a category of technologies for collecting CO_2 from industrial flue gases. The captured CO_2 is transported and stored permanently in a suitable geological formation. The aim with CCS is to reduce CO_2 emissions drastically, however, this can only be achieved by addition of energy. There is ongoing research regarding the development of CCS and how to reduce the cost to make more investors prepared to take the economic risk [6].

2.1.1 Carbon Capture Technologies

There are three different categories of CCS technology, pre-combustion capture, oxyfuel capture and post-combustion capture. Pre-combustion and oxyfuel capture are both implemented to the production plants before the combustion. These technologies are more difficult to use when CCS is retrofitted into an already existing plant, whereas the opposite applies for post-combustion [6]. Therefore, the first two techniques are only described in overview.

The first method with pre-combustion CCS uses oxygen separated from air to perform gasification followed by the water gas shift reaction, which transforms carbon monoxide and water to CO_2 and hydrogen gas. The produced hydrogen can be separated from the CO_2 and used to create heat and power without any CO_2 emissions. The second method, oxyfuel combustion adds a pure oxygen flow and a recycled flue gas flow to the boiler, to shift the flue gas composition towards CO_2 and water. The third method, with post-combustion capture, is the most common as it is the easiest method to add to an already existing process. The reason being, as the name indicates, that the carbon capture is performed after the combustion, before the flue gases leave the stack [6]. Therefore, post-combustion is the method used in this thesis and is described in more detail in the following section.

2.1.1.1 Post-Combustion Technology

Post-combustion CCS can be an important technology for emission reduction, as it can be applied for combustion of all types of fuel [15]. A post-combustion carbon capture process with the use of MEA as solvent is shown in Figure 2.1.

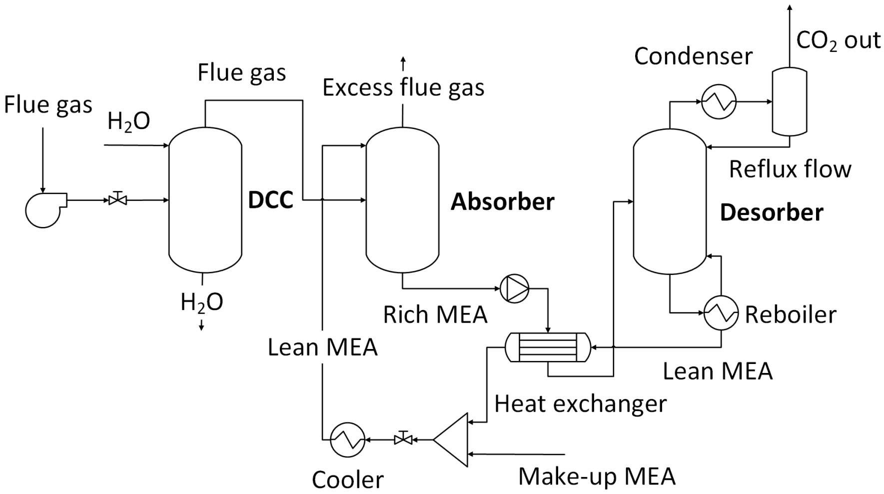

Figure 2.1: An example of a process scheme for a post-combustion process using MEA as solvent.

In the absorber the acid gas, in this case CO_2, is a reactant in a reversible reaction with the aqueous amine solution [15]. The solvent enters the column from above and the flue gas enters from below, which creates a counter flow pattern and contact between the liquid and gas phase, see Figure 2.1 [16]. A suitable temperature in the absorber can vary between 40 to 60 °C depending on the solvent used [15]. The remaining flue gases go through a water wash treatment to remove any solvent before the gases enter the atmosphere. The CO_2 is chemically bound to the solvent and leaves the absorber as rich solvent [15].

Chemical absorption has proven to be more efficient for low CO_2 concentrations [15] and aqueous amine solutions are the most common solvents with MEA as the most studied [7]. A concentration of 30 wt% MEA is often used in studies [17, 18, 19]. Furthermore, MEA has some disadvantages compared to other solvents such as high regeneration energy and requiring hydrogen to be manufactured which is currently produced from fossil fuels. Production of both MEA and the required base chemicals include high energy consumption and CO_2 emissions [7]. However, it has a relatively high reactivity and as it is the most investigated solvent it is still widely studied and used [7, 17].

The flue gases usually leave the combustion system at atmospheric pressure, resulting in a large gas flow with relatively low CO_2 concentration, between 3 to 15% depending on the fuel [15]. The low CO_2 concentration is one of the biggest disadvantages with post-combustion capture compared to the other previously mentioned methods [7]. With a low amount of CO_2, the absorption and thereby separation become less efficient [20] due to a smaller concentration gradient [16]. The flue gases from the combustion usually have a temperature above 100 °C and will therefore require pretreatment in the form of cooling before the absorption. This can be performed by a Direct Contact Cooler (DCC) which uses water to cool the flue gases, see Figure 2.1 [21]. Furthermore, the water will also wash the flue gases from particles and other contaminants [15, 22] and reduce the water content in the flue gases [22]. The pressure drop in the DCC and absorption column has to be counteracted with a blower [15].

An important aspect is the impurities present in the flue gases which can degrade the MEA, one example is sulphur oxide (SO_x). The presence of SO_x can require pretreatment before the absorption [23]. The recommendation is a concentration level below 10 ppm of sulphur dioxide (SO_2), which is also applied for nitrogen dioxide (NO_2), another impurity that can degrade MEA [24]. To reduce the SO_x concentration before the absorption, sodium hydroxide (NaOH) can be used and this can be added directly in a wash column [25].

The rich solvent is regenerated in the desorber column, usually operating between 100 to 140 °C, shown in Figure 2.1. The regeneration reaction is endothermic and a reboiler is used at the bottom of the column to provide the required heat. A condenser is used to separate the remaining water and the CO_2 flow at the top of the column. The CO_2 is compressed before transport to a storage [6]. The lean solvent exiting the desorber is used to preheat the rich solvent prior to regeneration [15] in order to recover some of the heat [16]. After the heat exchanger the lean solvent is pumped back to the absorber [15], as shown in Figure 2.1.

2.1.2 Liquefaction

Before the CO_2 is transported and stored it must be compressed [6] and for some transport modes, such as ship, it has to be liquefied [26]. In advance of the liquefaction, preparation of the CO_2 flow is required in the form of purification steps, depending on the impurities, and drying steps by condensation and adsorption [27].

There are two commercial processes of liquefaction, one using low pressure compression followed by an external refrigeration cycle and a second using high pressure compression with free liquid expansion [26]. The first method utilises two to four stages of compression with intermediate cooling to increase the pressure and to lower the temperature of the CO_2 as well as to condense the water [26, 28]. An example of the described process steps are shown in Figure 2.2 for three compressor stages [28]. The condensed water is removed from the CO_2 stream in a knockout drum between every compression stage and an adsorption dryer is used to remove the rest of the water after the last compressor stage [26, 28]. In the next step the CO_2 is cooled to liquid phase using a refrigerant, commonly ammonia [26]. A recirculation is used in combination with an extra compressor to return the gaseous CO_2 after the main compressor stages [28]. The CO_2 is compressed to a pressure around 17 bar before it is stored in an intermediate storage [26].

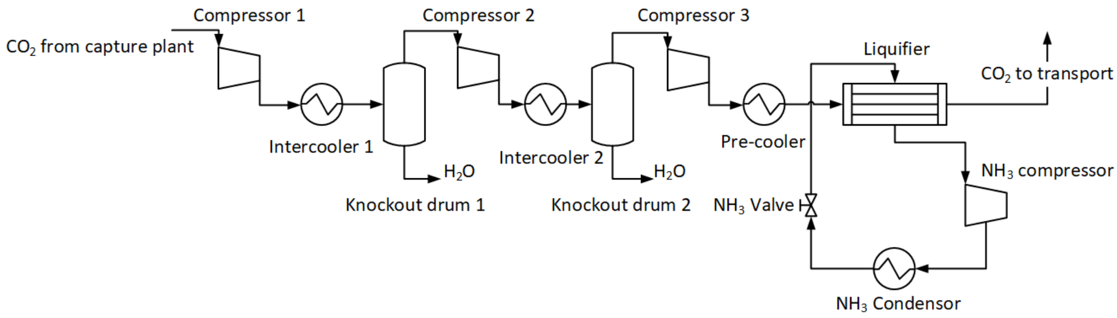

Figure 2.2: Liquefaction process with NH_3 refrigeration. Own figure based on the process in Deng et al. [28].

The second method uses compression in three stages with intercooling similar to the first method. A difference is that the gas is compressed to 73.8 bar, which is the critical pressure, and then cooled to 31.1 °C which is below its critical temperature. The gas is expanded with the use of a valve to reach a pressure around 17 bar and a temperature of -22 °C. The process is self-refrigerating during adiabatic conditions and half of the CO_2 is liquefied and the rest is recycled to the third compression step [26].

The external refrigerant process, shown in Figure 2.2, is often preferred as it has lower cost for the required energy. However, the free liquid expansion is not using a refrigerant. Additionally, for conditions around 6.5 bar and -52 °C, which is usually recommended for large-scale ship transport, the refrigerant will need lower temperature and for this case free liquid expansion is more beneficial [26]. The liquefaction method was one decision required when CCS was considered for a CHP plant in Klemetsrud, Norway, owned by Fortum Oslo Varme. The method using a refrigerant, consisting of ammonia, was chosen for the liquefaction due to it being a more used and developed method [29].

2.1.3 Transport

Transport modes available for larger volumes of CO_2 often include ship and/or pipelines on- and offshore [15]. CO_2 tanks on trains or trucks could also be an option for transport on land. However, the amount of CO_2 transported is more restricted for these modes meaning that they can be more beneficial at a smaller scale both considering the CO_2 flow and the distance [30]. These two parameters also have a large effect on the cost, and are the reason why trucks or trains can become expensive for larger volumes or longer distances [15, 30].

As mentioned previously, the CO_2 flow from the capture plant is liquefied before the transport. This step is necessary to decrease the cost of the transport, because it results in a higher density [15, 30]. A project called Northern Lights is investigating the area outside the coast of Norway for possible geological storage sites [31] and is using a pressure of 15 bar and a temperature of -25 °C for ship transport [32]. For transport in tanks, the liquid state of the CO_2 is generally preferred [15] often at a temperature below -15 °C and a pressure above 12 bar, however with a pressure above 45 bar the ambient temperature can be sufficient for the transport [30]. Insulation is required for the tanks with a possible capacity of 60 ton CO_2 but 30 ton CO_2 is more common [30]. The previously mentioned project in Klemetsrud in Oslo, Norway, considered the most appropriate transport mode to be trucks to the harbour in Oslo [33] for a distance of 12.1 km, where around 30 tons was possible to load in each truck [29].

Opportunities for companies to cooperate can exist if they are located near each other in clusters. Infrastructure for transport and compression units can be shared which can decrease the costs and imply less economic risk [14]. For example, economies of scale can be a result from a shared pipeline transport and storage injection. With fewer but larger pipelines required, the cost will be lower. In Europe there have been several identified clusters of industries with potential for cooperation [34]. A large example in Europe is in Rotterdam, the Netherlands, where there are many industries in close proximity with possible storage sites near the coast [35]. Another example of an area with several clusters of industries, with the possibility of a shared transport and storage solution, is the region of Skagerrak and Kattegat [36].

The captured CO_2 can also be collected at a hub which can be a large connection for several pipelines or an intermediate storage for several industries [14]. The next step with transport from a cluster hub to a permanent storage, can be achieved with offshore pipelines or ships, in the case of storage at sea [37]. In Sweden an option for an intermediate storage is in the Port of Gothenburg, specifically the Energy terminal, before overseas transport to a permanent storage [27]. This has been studied in the project CinfraCap which investigated the possibilities for a CCS infrastructure from the capture plant to the Port of Gothenburg [31]. The option of having a shared liquefaction for three facilities close to the intermediate storage site was investigated and around 300 million SEK (corresponding to approximately 30 million €) were possible in investment savings compared to the alternative of separate facilities. However, neither of the alternatives was proposed as the more appropriate solution because possibilities such as use of current labour and heat integration should be further included [27].

For the economy of the capture plant, the cost can be stated either with or without the compression stages in the liquefaction and therefore a cost of around 50-130 €/ton CO_2 can be seen in the literature [38]. The cost for only the capture plant can be between 50-70 €/ton CO_2 when applied to power plants with coal as fuel [16]. For other industries such as steel and pulp mills, a range of 80 to 90 €/ton CO_2 is possible, considering the whole chain from capture to storage. In these cases, the steam used for the reboiler had a large impact on the cost because the heat was assumed to be produced externally [39].

2.1.4 Storage

For intermediate storage at harbours or hubs, in the wait for ship transport to a permanent offshore storage, tanks are used with a suitable pressure to keep the liquid phase of the CO_2 as well as a low temperature [36]. Permanent storage of CO_2, both offshore and onshore, is dependent on having appropriate geological formations. For example, fields which previously contained oil or gas as well as different types of rocks depending on its porosity can be used [15]. Another example is geological formations filled with water due to the rock characteristic, these are called aquifers [40]. The characteristic of the rock above the storage space is important because this should act as a ceiling which the CO_2 can not go through in order for the storage to be permanent [36]. In order to place the CO_2 in the storage, injection techniques are required and for example the equipment used in the oil industry have been a starting point for the injection methods [15].

The potential for storage exists in several countries, which was shown in a summary by the Global CCS Institute including 29 countries and the EU [41]. When considering the Nordic countries, Norway has the most potential storage capacity after a high oil and gas extraction in the North Sea [42]. Around the west coast of Sweden, the Skagerrak and Kattegat region could also have possible storage sites [36]. These consider offshore storage but there could also be potential onshore through for example coal beds, however for larger commercial storage it has to be offshore according to the government in Sweden [40].

In Norway there is an ongoing project called Longship which considers the full chain of CCS from capture to storage. The transport is considered by the previously mentioned project Northern Lights. The storage is also included in this part of the project [43]. Ships will be used to gather the emissions to a common terminal where the transport to the permanent storage, in the North Sea, will be through pipelines [44].

2.2 Combined Heat and Power Plant

For heat and electricity production, a CHP plant can be used [9]. One method is the use of a steam cycle, see Figure 2.3, where water at a high pressure is heated to steam through heat exchanging in a boiler after which a generator is connected to a turbine for electricity production. The outgoing stream can be used in a condenser for heat production, shown as district heating, $Q_{district\ heat}$. After the condenser, a pump is required to recreate the high pressure before steam production in the boiler. The fuel which can be used in the boiler is flexible, it can for example be wood or waste [45].

The flue gases out from the boiler still have useful heat, because of a high temperature. Therefore a flue gas condenser can be used to obtain some of this heat through condensing the water present in the flue gases [46]. The additional heat acquired can be used for district heating [47].

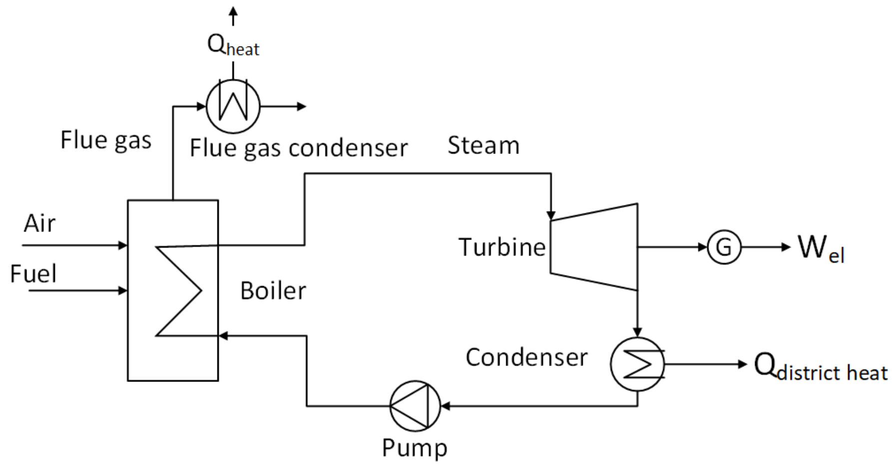

Figure 2.3: A steam cycle for a combined heat and power plant.

Two CHP plants connected to the district heating system in Gothenburg are owned by Göteborg Energi and Renova. These are used for the case study in this thesis. Both are located in Sävenäs, Gothenburg, and are described below.

2.2.1 Renova's Waste-to-Energy Plant in Sävenäs

Renova is a waste management company for the region of Gothenburg whose production started in 1972 [48, 49]. The company collects waste and the parts which can not be recycled is burned in their waste-to-energy plant in Sävenäs to produce heat and power. The production at the CHP plant is never totally shut down [48]. A reason is the income from companies and households paying for their waste to be handled. In addition, there are prohibitions for landfilling regarding waste of a certain kind, which therefore have to be incinerated [50]. The produced heat is delivered to the district heating network of Gothenburg and the electricity to the electrical grid. The incineration of waste results in carbon dioxide emissions, where two thirds are biogenic [48]. When the maximum capacity of the district heating system is reached, the excess heat is cooled down in cooling towers which use water from the river Säveån [51].

The plant has four furnaces and in addition to the combustion in the furnaces, flue gas condensation is also used to collect heat [49]. Each furnace has separate flue gas purifications [48]. Impurities in the flue gases are treated with different filters and for NO_x treatment a Selective Non Catalytic Reduction (SNCR) or a catalytic purification is used depending on the boiler. SO_2 is also treated in a specific step for each of the flue gas flows. The emission of SO_2 was 7.5 ton during 2019 [51] which corresponds to a SO_2 concentration of 0.9 ppm as a yearly average.

Renova's CHP plant in Sävenäs, further on denoted as Renova, has the permission to burn 550 000 tons of waste every year and usually around half of the incoming waste is from households [48]. In 2019 approximately 526 000 tons were incinerated and the energy regeneration was 3.4 MWh/ton. In the same year, the CHP plant in Sävenäs produced 1.5 TWh of heat and 280 GWh of electricity [51]. The production covers around 30% of the district heat and 5% of the electricity demand in Gothenburg [48].

2.2.2 Göteborg Energi's CHP Plant in Sävenäs

In 1990, Göteborg Energi was founded from another company called Energiverken and the city of Gothenburg now has the full ownership of the company [52]. Göteborg Energi is the owner of the district heating system in Gothenburg, but they also have part in other energy matters [53] such as local ownership of the electricity system [54] and sales of biogas [53]. In the Sävenäs area, Göteborg Energi owns a CHP plant which is connected to the district heating system [47], referenced with the name Göteborg Energi throughout the report. Winter is the main time for the operation of the plant and the fuel is mostly bio-based through wood chips but some of the boilers use natural gas or oil to a smaller extent. The boilers are used for producing heat and are four in total, where one of them, named HP3, is connected to a steam turbine. For HP3 and one other boiler, there are also facilities for flue gas condensation, which produce additional district heat. Near the plant there is a river called Säveån and this is used for releasing the condensed water after treatment [47].

A flue gas treatment is used after HP3, although not for SO_x. The targeted impurities are ashes and NO_x with the use of filters and SNCR respectively. The sulphur level in the fuels for the other boilers, oil or natural gas, is known. Therefore, it is only measured for HP3 due to the use of wood chips [47]. The reported SO_2 emissions during 2019 was 3 ton [47] and this corresponded to an average level of 1.6 ppm SO_2 in the flue gases.

During the year of 2019 the CHP plant produced 447 GWh heat and 41 GWh electricity with a total emission of CO_2 at 148 kton. The amount of wood chips incinerated during the year 2019 was 157 kton [47]. There are plans for increasing the amount of renewable fuels through replacing some of the old boilers [55]. The current total capacity of the CHP plant is 278 MW for the installed equipment, however 352 MW is allowed according to the original permit for the construction of the plant [47].

2.3 Emission Allowances and Carbon Capture and Storage Funding

For the use and implementation of CCS, the associated policies and funding are important. One aspect within the EU is the trade of emission allowances, which consists of an established trading system where the amount of emission allowances in the scheme is controlled and decreased over time with the aim to reduce emissions. The emissions from a company need to be covered by emission allowances and between companies a trade of the allowances occur. In the first stage the allowances can be distributed partly without cost and partly through having to buy the allowances in an auction. The former depends on several parameters such as how efficient the production is considering carbon use [56]. Consequently, the cost for the allowances varies during the year and from 2019 to 2020 the price has increased from around 20 €/ton CO_2 to 30 €/ton CO_2 [57]. The need for allowances can be reduced through CCS, because the amount of CO_2 permanently stored is not included in the total emissions. The trading scheme does not consider the use of biofuels and its emissions [43], yet waste incineration plants are included in the trading scheme in Sweden [58].

Sweden has acknowledged the need to include negative emissions, with CCS for biofuels, to be able to fulfil the climate targets [59]. One suggestion for promotion of negative emissions is through an auctioning system designed in reverse [60] where the company's capture cost should be the lowest in order to receive a compensation through the auction [61]. There is also a programme called the Industry Leap in Sweden which is a way for both industries and research projects to receive funding, such as CCS projects especially considering bioenergy [3]. Within the EU there are also possibilities for funding through an Innovation Fund between the years 2020-2030, and this is especially aimed at industries to reduce the economic risk associated with investing in for example CCS [62].

3
Method

In the following chapter the method is presented including how the Aspen Plus simulations were structured, the necessary data used for the transport calculations and the scenarios and cases studied. Further on, the method and assumptions for the economic and sensitivity analyses are described.

The system boundaries considered for the capture plant can be seen in Figure 3.1 for the two studied cases. The first case considered a separate capture plant for each company and the second case included shared desorption, liquefaction and transport infrastructure. The system boundaries did not include the intermediate and permanent storage since these steps were equal for the two cases.

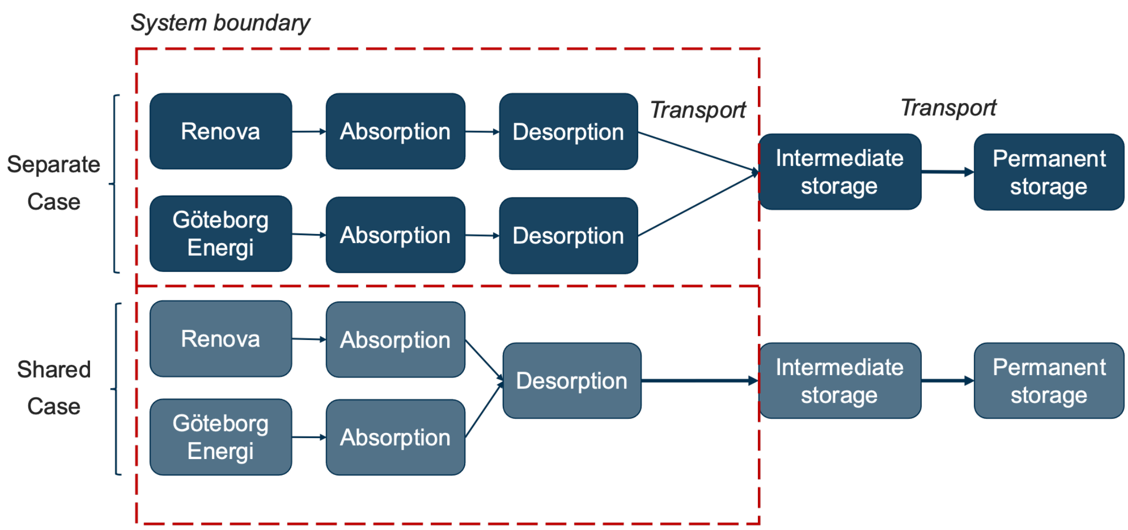

Figure 3.1: The compared cases with the system boundaries for the method.

3.1 Simulations and Equipment Sizing of a Carbon Capture Plant

To compare the cases with separate or shared regeneration and transport, the sizes of the equipment and the heating/cooling demands were required. Therefore, a model of the capture plant in Aspen Plus was used for simulations of the processes. The dimensions of the plant were determined for an on-design point, estimated based on the higher production levels and the same dimensions were further used for off-design points for lower production levels.

3.1.1 Separate Case

The separate capture plants were simulated in Aspen Plus by using a model setup by Garðarsdóttir et al. [17]. The model has since been used in other studies [13, 63] and further developed. The flowsheet of the modelled capture plant can be seen in Figure 3.2. Rate-based and equilibrium modelling were used for the absorber and desorber, respectively [17]. The lean MEA to the absorber was adjusted to obtain a CO_2 removal of 90% in the absorption column. In the absorber, a washer was included for the outgoing flue gases with an additional height of 3 m and equal diameter to the absorber. The Direct Contact Cooler (DCC) described previously was not seen to be required due to the low SO_x levels, NO_x removal and flue gas condensation at the CHP plants.

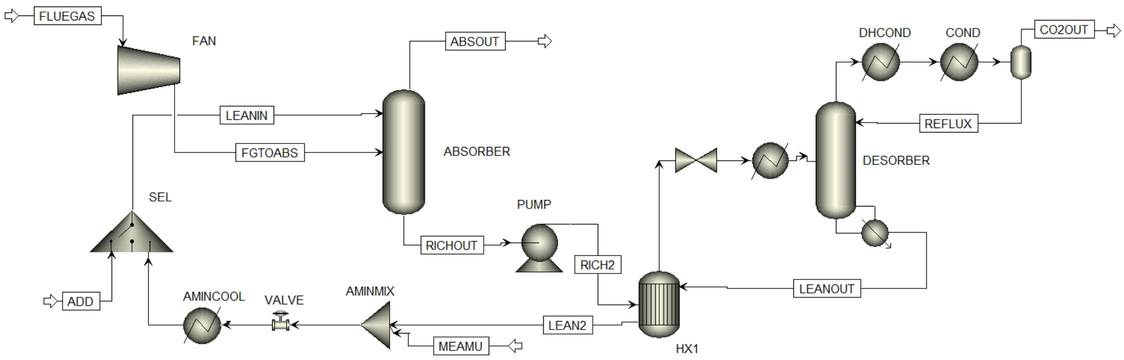

Figure 3.2: Aspen Plus model for the separate capture plant.

The assumptions used for the model are summarised in Table 3.1. The estimated inlet temperature and pressure to the capture plant were based on the capture plant being located after the flue gas condensation. The lean loading, which is the mole of CO_2 in the lean flow compared to the mole of MEA, was optimised for the on-design point in the first scenario. This was achieved through changing the lean loading in order to have as low heat demand as possible in the reboiler compared to kg CO_2 out [39]. The resulting lean loading was 0.23 and was used for the succeeding cases and scenarios.

Table 3.1: Assumed parameters for the Aspen Plus model.

Parameter	Value	Source
Flue gas inlet [bar]	1.01	
Flue gas inlet [°C]	40	
MEA [wt%]	30	[17, 18, 19]
Capture rate [%]	90	
Absorber inlet temperature [°C]	40	[17]
Reboiler temperature [°C]	120	[13]
Absorber pressure [bar]	1.06	
Desorber pressure [bar]	1.9	
Absorber height [m]	25.7	
Desorber height [m]	12.8	
Lean loading [mol CO_2/mol MEA]	0.23	

3.1.2 Shared Case

For the case with a shared desorber, the Aspen Plus model was modified as in Figure 3.3. Two of the differences were the additions of a second absorber and a mixer for the rich streams after a lean/rich heat exchanger (HX1 and HX2) since the following desorber was shared. One of these heat exchangers (HX2) represents both a double pipeline and a heat exchanger, see Figure 3.4, due to the required transport between the absorber and the desorber placed on different plants. The double pipeline was constructed with one flow in the outer pipe and one counter flow in the inner pipe. The warmer flow was placed in the outer pipe. Heat was transferred between the two flows through the wall of the inner pipe and the double pipe was assumed to be insulated and placed underground, resulting in no heat losses. The simulated heat exchanger was used to calculate the total heat transfer required, as the MEA pipeline between the companies also functioned as a heat exchanger. Out from the desorber a splitter divided the lean MEA to the two absorption columns. The split ratio was calculated based on the relation between the lean MEA into the two absorber columns, the flows LEANIN and LEANIN2 in Figure 3.3.

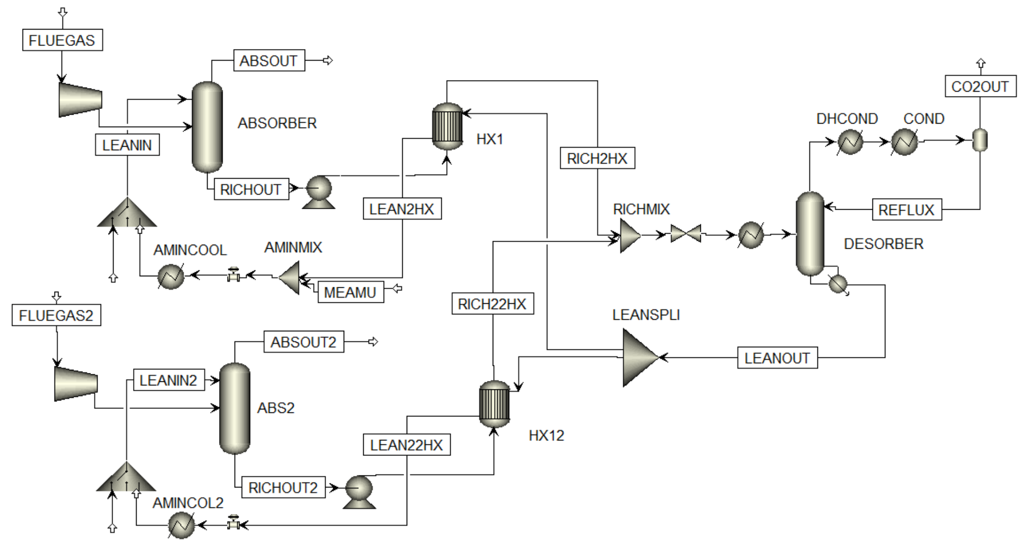

Figure 3.3: Aspen Plus model for the shared desorber.

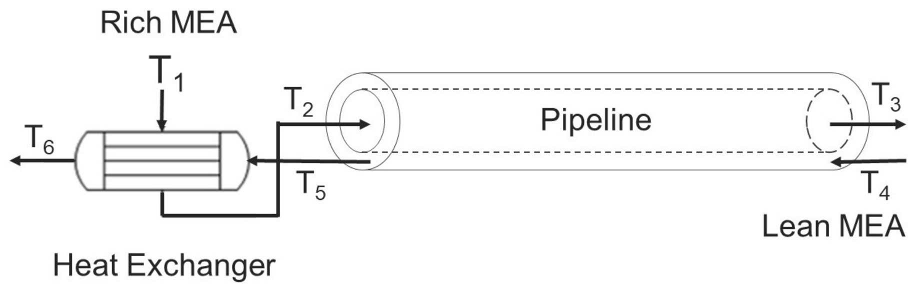

Figure 3.4: Double pipeline combined with an additional heat exchanger.

To be able to transfer the rich and lean MEA between the sites extra pump work was required. To determine the extra work, the pressure drop in the double pipeline between the sites was calculated. The diameters of the inner and outer pipes were calculated for the first design through the velocity in the pipe based on the mass flow and the density of the liquid. The inner and outer velocities were set to be smaller but close to 0.9 m/s due to design recommendations [64] and the velocities in a test facility in Norway in the city Mongstad [65]. The diameters were calculated for the first on-design point and kept the same for the following designs.

The double pipeline was assumed to behave like a double pipe heat exchanger and the pressure drop was determined with Darcy-Weisbach equation, which is described in Equation (3.1) [66]. The length, L, of the pipeline was estimated as the distance between the companies.

$$\Delta P_{inner} = \frac{f_{inner} \cdot L \cdot G_{inner}^2}{2 \cdot \rho_{inner} \cdot d_{inner}} \tag{3.1}$$

f_{inner} is the friction factor for the inner pipe and was calculated with Equation (3.2) for a turbulent Reynolds number below 10^5 and Equation (3.3) for a larger Reynolds number [67], because the pipe was assumed to be smooth [66, 67].

$$f_{inner} = \frac{0.316}{Re_{inner}^{0.25}} \qquad 2100 < Re_{inner} < 10^5 \tag{3.2}$$

$$f_{inner} = 0.0032 + 0.221 \cdot (Re_{inner})^{-0.237} \qquad 10^5 < Re_{inner} \tag{3.3}$$

The Reynolds number for the inner tube is described in Equation (3.4) and G_{inner} is the mass velocity and was calculated for the inner tube by Equation (3.5) [66].

$$Re_{inner} = \frac{d_{inner} \cdot G_{inner}}{\mu_{inner}} \tag{3.4}$$

$$G_{inner} = \frac{\dot{m}}{A_{inner}} \tag{3.5}$$

The same procedure was used to determine the pressure drop in the outer tube. However, the diameter was changed to the difference between the outer and inner diameter, which is described in Equation (3.6) and the flow area was changed to the cross-sectional area between the tubes, see Equation (3.7). All the physical properties such as density, viscosity and mass flows were changed to the outer pipe conditions [66].

$$D = d_{outer} - d_{inner} \tag{3.6}$$

$$A_{outer} = \frac{\pi(d_{outer}^2 - d_{inner}^2)}{4} \tag{3.7}$$

The pump work was calculated from the pressure drop, the density and the pump efficiency according to Equation (3.8) [66].

$$W = \frac{\Delta P}{\rho} \cdot \frac{\dot{m}}{\eta} \tag{3.8}$$

From the simulations in Aspen Plus the total heat transferred between the rich and lean stream as well as the logarithmic mean temperature difference were determined. An overall heat transfer coefficient was calculated for the pipelines by Equation (3.9) [68]. The required area of the pipeline to fulfil the total heat transfer was determined with Equation (3.10). Two different materials were compared, either a plastic in the form of polyethylene or manganese steel. The thermal conductivity of plastic was assumed to be 0.4 W/(m·K) [69] and manganese steel as 35 W/(m·K) due to it being between 22-50 W/(m·K) for different manganese levels [70]. The wall thickness of the pipe was assumed to be 10 mm due to it being a standard for an outer diameter of around 0.5-0.8 m [71].

$$\frac{1}{U} = \frac{d_{outer}}{d_{inner} \cdot h_{inner}} + \frac{d_{outer} \cdot ln(\frac{d_{outer}}{d_{inner}})}{2 \cdot \lambda} + \frac{1}{h_{outer}} \tag{3.9}$$

$$A_{required} = \frac{Q_{total}}{U \cdot \Delta T_{lm}} \tag{3.10}$$

The heat transfer coefficient, h, was calculated for the inner and outer pipe with Equation (3.11). For the inner pipe, D_e was assumed to be equal to d_{inner} and for the outer pipe it was described by Equation (3.12). The division between the viscosity of the fluid in the centre of the pipe and by the wall was assumed to be equal to 1 [66].

$$h = \frac{0.027}{D_e} \cdot \lambda \cdot Re^{0.8} \cdot \left(\frac{C_p \cdot \mu}{\lambda}\right)^{\frac{1}{3}} \left(\frac{\mu}{\mu_w}\right)^{0.14} \tag{3.11}$$

$$D_e = \frac{d_{outer}^2 - d_{inner}^2}{d_{inner}} \tag{3.12}$$

The amount of total heat transferred in the pipeline was calculated according to Equation (3.13) by the use of the ratio between the actual pipeline area and the larger required area. The additional heat transfer needed in an external heat exchanger was calculated with Equation (3.14) and the temperatures between the heat exchanger and the pipeline, T_2 and T_5 in Figure 3.4, were determined by Equation (3.15), Equation (3.16) and Equation (3.17). The heat exchanger area was obtained as in Equation (3.10).

$$Q_{pipeline} = \frac{A_{pipeline}}{A_{required}} \cdot Q_{total} \tag{3.13}$$

$$Q_{HX} = Q_{total} - Q_{pipeline} \tag{3.14}$$

$$Q_{rich} = \dot{m}_{rich} C_{p,rich}(T_3 - T_2) \tag{3.15}$$

$$Q_{lean} = \dot{m}_{lean} C_{p,lean}(T_5 - T_4) \tag{3.16}$$

$$Q_{pipeline} = Q_{rich} = -Q_{lean} \tag{3.17}$$

3.2 Liquefaction

The CO_2 flow from the capture plant was connected with a liquefaction plant to be compressed and liquefied to a pressure of 15 bar and a temperature around -25 °C. The pressure and temperature were set to match the properties used in the Northern Lights project [32]. A liquefaction process with an ammonia refrigerant cycle was used in this thesis, see Figure 2.2, likewise to the project at Klemetsrud.

Parts of the liquefaction process were modelled in Aspen Plus, see Figure 3.5. During the simulations, CO_2 and water were considered as the only components in the inlet flow and no reactions were assumed to occur in the process. The compression was assumed to consist of a sequence in 3 stages, the first compression stage increased the pressure from 1.9 bar to 3 bar. The two following stages increased the pressure to 9 and 15 bar, respectively [28].

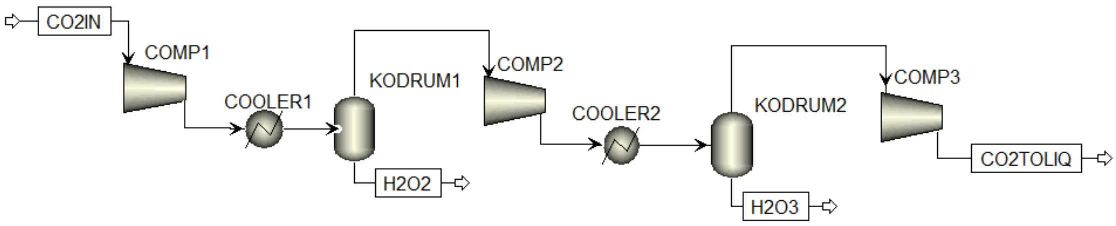

Figure 3.5: Aspen Plus model for the compression process in the liquefaction.

Between every compression stage the flow was cooled by an intermediate cooler and condensed water was removed by a knockout drum. To accomplish the last low-temperature cooling an ammonia cycle was used, however only the compression with intermediate cooling was simulated for calculation of recovered district heat [63]. The simulated model can be seen in Figure 3.5. After the ammonia cycle, a part of the flow was assumed to be recirculated, which resulted in the need of an additional compressor unit [28]. During the simulations Peng-Robinson was chosen as the equation of state [72]. For the compressors, 0.85 was assumed as the isentropic efficiency [28] and the mechanical efficiency was set as 0.95.

The work required in the additional compressors in the refrigeration cycle and the recirculation, were calculated based on the results of Deng et al. [28]. The power of the recirculation compressor was divided by the power of the CO_2 compressors, $W_{recirculation}$ based on Deng et al. [28], see Table 3.2. The work needed for the compressor in the ammonia cycle was estimated in the same manner through $W_{ammonia\ compressor}$, see Table 3.2. The relations were used in combination with the required work from the simulated CO_2 compression to obtain the power demand in the current process, see Equation (3.18).

$$W_{tot} = W_{simulation} \cdot (1 + W_{recirculation} + W_{ammonia\ compressor}) \tag{3.18}$$

In a similar way the additional cooling demands from a pre-cooler and the ammonia cycle were calculated, based on the intermediate cooling demand instead. The results from Deng et al. [28] were used for the relations, see Table 3.2, and the total cooling demand was calculated by Equation (3.19).

$$Q_{tot} = Q_{simulation} \cdot (1 + Q_{pre-cooler} + Q_{ammonia\ cycle}) \tag{3.19}$$

Table 3.2: Literature values from Deng et al. for cooling and power demands in the liquefaction [28].

Parameter	Value
$W_{recirculation}$ [MW/MW CO_2 compressors]	0.0205
$W_{ammonia\ compressor}$ [MW/MW CO_2 compressors]	0.234
$Q_{pre-cooler}$ [MW/MW intercooling]	0.5604
$Q_{ammonia\ cycle}$ [MW/MW intercooling]	2.0258

3.3 Transport and On-Site Storage

The chosen transport mode between the CHP plants and the intermediate storage was truck. For Renova, the truck alternative was considered as most feasible in the CinfraCap project. In the same project, the Port of Gothenburg with the Energy Terminal was considered for the intermediate storage [27] and was therefore chosen as the location in this thesis. The distances between the CHP plants as well as to the Energy terminal were estimated using Google Maps [73] and can be seen in Table 3.3. The distance between Renova and Göteborg Energi was measured from the centres of the facilities. The time was calculated based on an assumed velocity of 50 km/h. Liquefied CO_2 is classified as dangerous goods and has specific legislation for transport [74]. Therefore, the route between the CHP plants and the port of Gothenburg was chosen based on the recommended roads for dangerous goods from Trafikverket [75] and is shown in Figure 3.6 from Google Maps [73].

Table 3.3: Travel distances and times of one way journey for the companies in Sävenäs, Gothenburg.

Start	End	Distance [km]	Time [min]
Renova	Energy Terminal	27	32
Göteborg Energi	Energy Terminal	27	32
Renova	Göteborg Energi	1	-

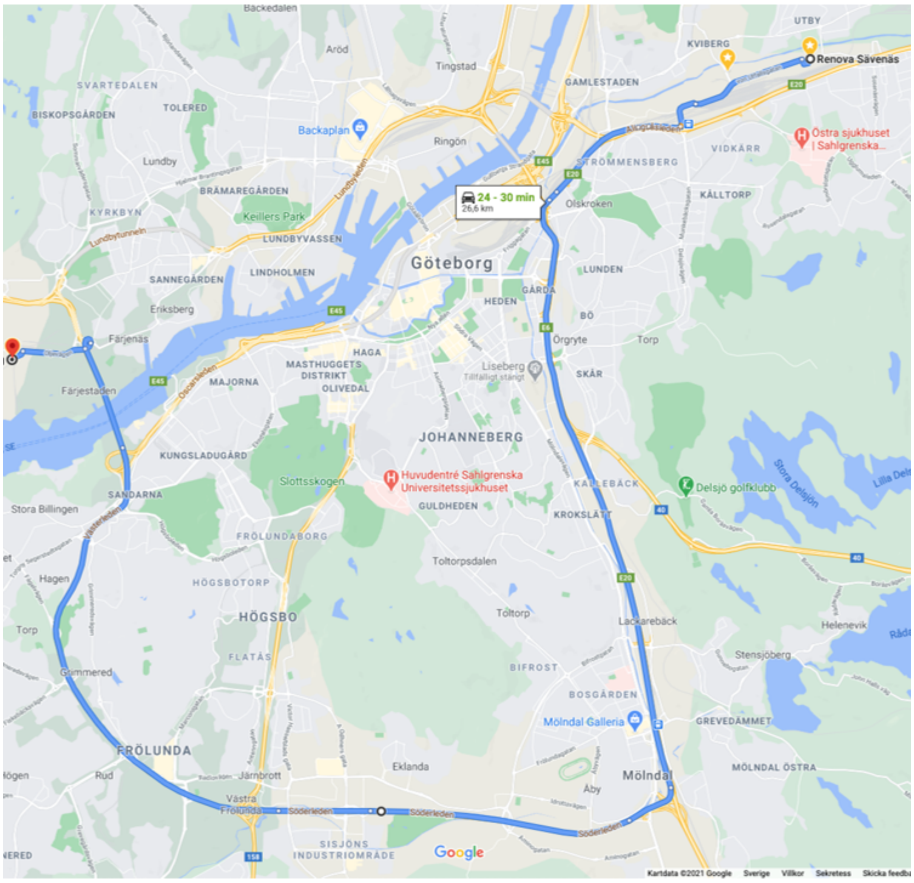

Figure 3.6: Transport route between the CHP plants and the Energy terminal in the Port of Gothenburg from Google Maps [73].

The assumed specifications for the truck transport including the loading capacity and the time required for the loading and unloading can be seen in Table 3.4. The fuel was assumed to be diesel.

Table 3.4: Truck and transport specifications.

Parameter	Value
Capacity, $T_{capacity}$ [m^3/truck]	33
Loading time [h]	0.5
Unloading time [h]	0.5
Fuel usage [l/km]	0.5

On the capture plant, an on-site storage for the CO_2 was included, with a storage time of 24 hours, based on the study in Klemetsrud [29]. The CO_2 density at 15 bar [28] and the maximum CO_2 flow was used for calculation of the storage volume.

3.4 Scenarios and Cases

To investigate if a common regeneration plant, liquefaction and transport were beneficial and how the district heat and electricity production of the CHP plants were affected by a carbon capture plant, four different scenarios were established, see Figure 3.7. The first being maintained fuel usage which was used to see the effect on the operation of the CHP plant after utilising a part of the produced steam for the reboiler in the capture plant. The second scenario considered a maintained district heat delivery from the studied system. Maintained district heat was studied to see how much more fuel would be necessary in order to fulfil the reboiler duty and the district heating demand. In this scenario the boiler capacity was assumed to be possible to increase, for example through upcoming renovations, and the calculations were used to determine how much capacity would be required. However, the cost for an increase in the capacity was not taken into account nor the permits required. The third scenario relates to the second through the aim of maintained district heat delivery, however with a consideration to the restricted capacity. Therefore when possible, the heat production was increased and if the on-design point was reached a loss in district heat would occur instead. The fourth scenario included a reduced capture rate in order to see the effects on the economy with capture plants for two CHP plants more similar in size.

For each scenario two cases were studied as previously explained in Figure 3.1, firstly having separate carbon capture infrastructures and secondly having a shared regeneration plant, liquefaction and transport. For some of the early cases in the first scenario there were a few design choices, for example the location of the shared regeneration facility, depending on which company could provide the reboiler heat, and which material should be used for the MEA pipeline between the companies.

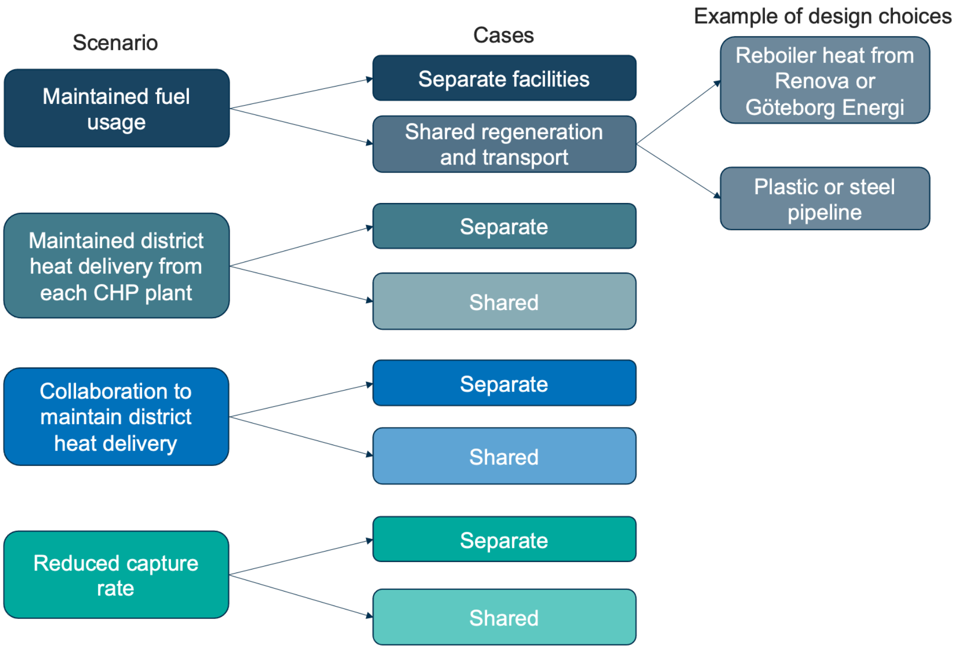

Figure 3.7: The studied scenarios and cases.

3.4.1 Scenario 1: Maintained Fuel Usage

The first scenario considered simulations based on the fuel usage, which was assumed to be maintained and therefore some of the heat produced in the CHP plant had to be used in the reboiler in the capture plant instead of being delivered as district heat. To achieve this, data for the heat production during 2019 on an hourly basis was received from Renova and Göteborg Energi.

The total CO_2 emissions and produced heat including the flue gas condensation during 2019 is shown in Table 3.5 for both Renova [51] and Göteborg Energi [47]. For Göteborg Energi, the stated heat production and emissions originated from the boiler HP3, where wood chips were used and additional fossil emissions (6 kton [47]) from the other boilers were not considered.

Table 3.5: The total CO_2 emissions and produced heat in 2019.

	Renova	**Göteborg Energi (HP3)**
Total CO_2 emissions, $m_{CO_2\ 2019}$ [kton]	546.6	142
Total produced heat, $Q_{heat\ 2019}$ [GWh]	1 500.8	421

The heat production for Renova's CHP plant is shown in Figure 3.8, including the flue gas condensation and heat production for internal use. In the duration chart Figure 3.9, the hourly produced heat and the estimated design points can be seen. The on-design was chosen for the more frequent production at a high level and is seen as the highest horizontal line. The variations in operation for the CHP plant were considered through off-design modelling to calculate the different power, heating and cooling demands. The off-design point was chosen based on the more frequent low level of heat production.

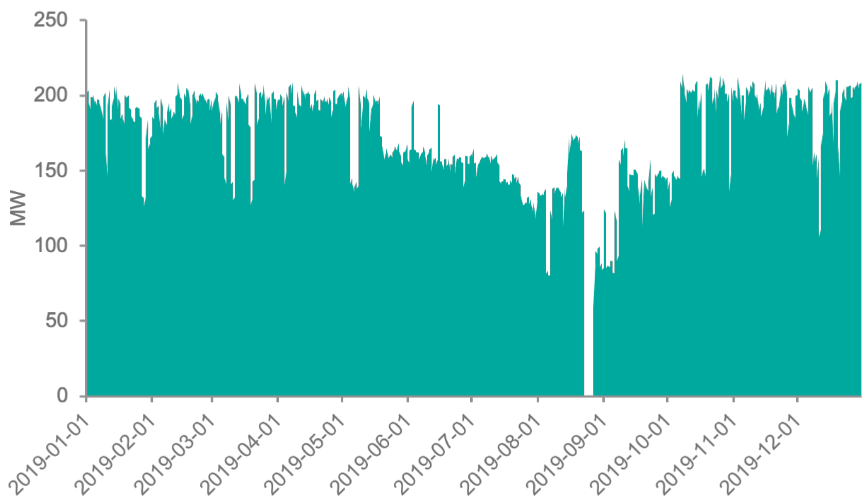

Figure 3.8: Measured produced and delivered heat for Renova in 2019.

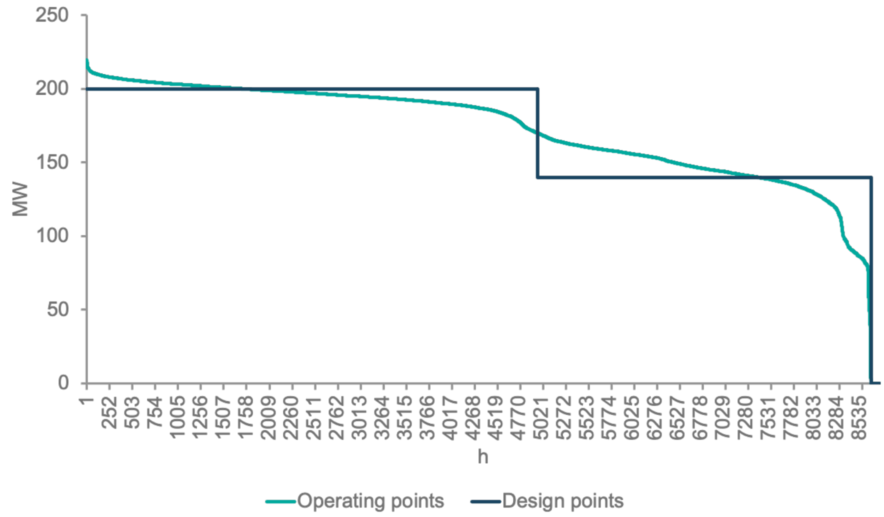

Figure 3.9: Operating and design points for Scenario 1, maintained fuel usage, during 2019 for Renova.

The produced heat was converted to a flue gas flow through the use of the total CO_2 emissions during a year. The total emissions was divided with the total produced heat in order to know the emission per kW as in Equation (3.20).

$$ton\ CO_2/kW = \frac{\dot{m}_{CO_2,\ 2019}}{Q_{heat,\ 2019}} \tag{3.20}$$

Since the hourly produced heat in 2019 was known, the CO_2 emissions for every hour were calculated, Equation (3.21).

$$\dot{m}_{CO_2} = ton\ CO_2/kW \cdot Q_{heat\ per\ hour} \tag{3.21}$$

The amount of CO_2 was converted into molar flow and the flue gas flow was estimated in Equation (3.22) with the ideal gas law and the CO_2 concentration.

$$\dot{V}_{flue\ gas} = \frac{\dot{n}_{CO_2} RT}{P \cdot x_{CO_2}} \tag{3.22}$$

The design points with the corresponding flue gas flow rates and operating hours can be seen in Table 3.6.

Table 3.6: Design points, operating hours and flue gas flows for Renova Scenario 1, maintained fuel usage.

Design point [MW]	Operating hours [h]	Flue gas flow [kNm^3/h]
On-design, 200	4722	371
Off-design, 140	1777	260
Off-design, 0	92	0

The flue gas composition can be seen in Table 3.7 and was estimated based on the environmental report from Renova in 2019 [51] and a master's thesis by Andersson [19]. The volume fraction of water was based on the vapour fraction at 1.01 bar [76] due to the flue gases being seen as saturated after the flue gas condensation before the inlet to the capture plant.

Table 3.7: Flue gas composition for Renova.

Components	Volume fraction
CO_2	0.10
H_2O	0.074
O_2	0.07
N_2	0.756

For Göteborg Energi, the same method was applied for the choices of on-design and off-design simulations. The variation in the heat production including the flue gas condensation during 2019 is seen in Figure 3.10 for the HP3 boiler. The levels which were estimated for the design points can be seen in a duration chart, Figure 3.11, including the produced heat as operating points.

Figure 3.10: Measured produced and delivered heat for Göteborg Energi in 2019.

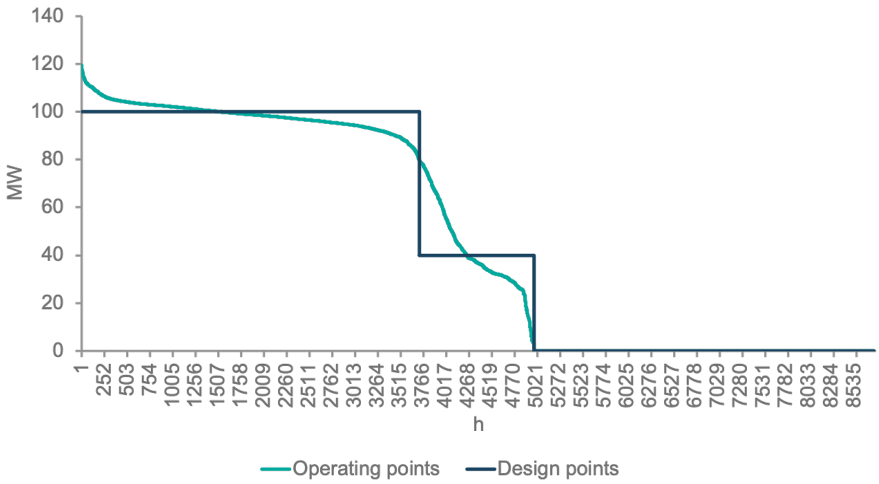

Figure 3.11: Operating and design points for Scenario 1, maintained fuel usage, during 2019 for Göteborg Energi.

Table 3.8 shows the flue gas flow rates for the simulations and the operating hours.

Table 3.8: Design points for Göteborg Energi Scenario 1, maintained fuel usage.

Design point [MW]	Operating hours [h]	Flue gas flow [kNm^3/h]
On-design, 100	1513	157
Off-design, 40	2221	63
Off-design, 0	3760	0

The assumed composition of the flue gas can be seen in Table 3.9 and the flue gas was again seen to be saturated due to the flue gas condensation before the inlet to the capture plant. For each scenario the inlet flue gas composition was seen to be constant for Renova and Göteborg Energi.

Table 3.9: Flue gas composition for Göteborg Energi.

Component	Volume fraction
CO_2	0.11
H_2O	0.074
O_2	0.07
N_2	0.746

For the case with a common regeneration unit the combinations from the two separate cases were used as the design points, with 100 MW production for Göteborg Energi and 200 MW production for Renova as the on-design point. The design points and the operating hours are shown in Table 3.10. The desorber was placed at Renova due to the use of a cheaper fuel, more operating hours and a higher capacity.

Table 3.10: Design points with corresponding operating hours for the shared case in Scenario 1, maintained fuel usage.

Renova [MW] \ Göteborg Energi [MW]	**0**	**40**	**100**
0	92	-	-
140	2895	228	562
200	773	1038	3172

3.4.1.1 Steam Cycle and District Heat Integration

The effect on the steam cycle in the CHP plant was considered through calculations of the lost production in electricity and district heat, based on the reboiler demand. The power-to-heat ratio, the α-value, was assumed constant and calculated through the total electricity and heat produced during 2019. The calculation of the α-value can be seen in Equation (3.23).

$$\alpha = \frac{W_{electricity}}{Q_{heat}} \tag{3.23}$$

The electricity production at each design point was calculated through the α-value and the produced heat. A reference case of the boiler production was set according to Equation (3.24).

$$Q_{boiler} = W_{electricity,ref} + Q_{heat,ref} \tag{3.24}$$

The losses in electricity production for the CHP plant, after implementing carbon capture, were estimated as in Table 3.11. For Renova there were possibilities to use 7 bar steam due to it being an existing steam extraction [77], however when the steam reached the reboiler the pressure was assumed to be 3.5 bar. The value of 20% reduction in electricity production was chosen based on the work by Pröll and Zerobin [78]. Göteborg Energi was assumed to use primary steam and the electricity loss was estimated based on the work by Beiron et al. [79]. The percentage was also assumed to be equal for both the on- and off-design points in the separate case. In the same work, a model of a more general CHP plant in Ebsilon Professional was used [79] and based on this model the electricity loss for each combination in the shared case was estimated as in Table 3.11 depending on the reboiler duty.

Table 3.11: Losses in electricity production for the studied CHP plants after implementation of carbon capture. The combination stands for Göteborg Energi-Renova.

	Heat production combination [MW]	**Electricity loss [%]**
7 bar steam	-	20
Primary steam, separate facility	-	56
Primary steam, shared facility	100-200	57
	100-140	44
	40-200	45
	40-140	33
	0-200	39
	0-140	26

An overall energy balance including the reboiler demand can be seen in Equation (3.25), based on a maintained boiler production, from Equation (3.24). $W_{electricity}$ is the produced electricity after losses.

$$Q_{boiler} = W_{electricity} + Q_{reboiler} + Q_{heat} \tag{3.25}$$

Through the known electricity loss and the maintained steam production, the lost heat was calculated with Equation (3.26).

$$Q_{lost\ heat} = Q_{heat,ref} - (Q_{boiler} - W_{electricity} - Q_{reboiler}) \tag{3.26}$$

For the desorber in the capture plant, opportunities for heat integration with the district heating system were considered as seen in Figure 3.12. Part of the heat from the condenser, down to 60 °C for the CO_2 flow, was used for district heat and was calculated in the Aspen Plus simulations. The saturated water out from the reboiler was cooled further by the use of district heat, prior to being returned to a suitable location in the steam cycle. Before the cooler, a valve decreased the pressure to the same level as the condenser in the steam cycle. This pressure was assumed to be 1.2 bar to have 105 °C as the condensation temperature [76], in order to have a temperature of 100 °C for district heating water [48].

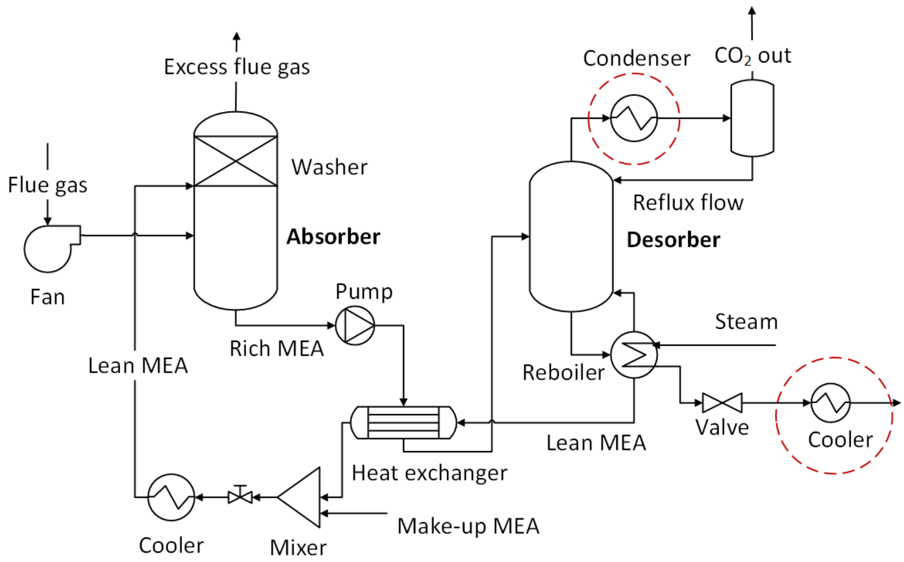

Figure 3.12: Heat integration in the capture plant.

The valve after the reboiler operated isenthalpic and the stream was cooled to saturated water at the new pressure. The possible amount of district heat could be calculated through the enthalpies and the steam flow. The latter was determined through Equation (3.27) with the condensation enthalpy of steam at 3.5 bar.

$$\dot{m}_{steam} = \frac{Q_{reboiler}}{\Delta h_{cond}} \tag{3.27}$$

The additional district heat from the condenser and the cooler after the reboiler, shown in Figure 3.12, was subtracted from the lost heat in the CHP system originating from the reboiler demand. From the liquefaction, there was also possible heat integration from the intercoolers between the compressor stages. As a result, the total lost district heat over the capture plant, liquefaction and the CHP plant was calculated in Equation (3.28). For the case of utilising primary steam from Göteborg Energi a de-superheater before the reboiler was also used for district heat.

$$Q_{lost\ DH} = Q_{lost\ heat} - Q_{condensor,\ DH} - Q_{after\ reboiler,\ DH} - Q_{intercooler,\ DH} \tag{3.28}$$

3.4.2 Scenario 2: Maintained District Heat Delivery

To maintain the district heat delivery, additional boiler duty was required through higher fuel usage, and was calculated based on the district heating and electrical efficiencies from the first scenario. For the district heating efficiency, the total district heat produced, including the heat recovered from the capture plant, was divided by the calculated boiler duty in Scenario 1 as in Equation (3.29). The electrical efficiency was based on the resulting electricity production after the losses based on the percentages in Table 3.11 and is shown in Equation (3.30). In this scenario, primary steam was assumed to be used and in the case of Renova the district heating and electrical efficiency had to be calculated with a recalculated district heat and electricity production from Scenario 1 based on primary steam instead of 7 bar steam from Table 3.11 for the separate facility.

$$\eta_{DH} = \frac{Q_{DH,\ Scenario\ 1}}{Q_{boiler,\ Scenario\ 1}} \tag{3.29}$$

$$\eta_{el} = \frac{W_{el,\ Scenario\ 1}}{Q_{boiler,\ Scenario\ 1}} \tag{3.30}$$

The district heating efficiency was used in combination with the heat production, which should be equal to the district heat production without a capture plant, in order to calculate the boiler duty required in Scenario 2, Equation (3.31).

$$Q_{boiler,\ Scenario\ 2} = \frac{Q_{DH,\ Scenario\ 2}}{\eta_{DH}} \tag{3.31}$$

The resulting boiler demand was used in combination with the electrical efficiency to calculate the production of electricity, Equation (3.32), and further on the electricity loss compared to the original production.

$$W_{el,\ Scenario\ 2} = \eta_{el} \cdot Q_{boiler,\ Scenario\ 2} \tag{3.32}$$

The flue gas flows were recalculated with the additional heat production with Equation (3.20). Based on the new flue gas flows, Aspen Plus simulations were performed. If the lost district heat was relatively large, the new loss was added to the boiler duty and new flue gas flows were recalculated and simulated. This iteration continued until the loss in district heat was below 0.01 MW. The final flue gas flows and the corresponding heat production can be seen in Table 3.12 for Renova.

Table 3.12: Design points and flue gas flows for Scenario 2, maintained district heat delivery, Renova.

Design point [MW]	**Flue gas flow [kNm^3/h]**
On-design, 219	407
Off-design, 153	284
Off-design, 0	0

Table 3.13 shows the final flue gas flows used in Aspen Plus and heat production required for Göteborg Energi.

Table 3.13: Design points and flue gas flow for Scenario 2, maintained district heat delivery, Göteborg Energi.

Design point [MW]	**Flue gas flow [kNm^3/h]**
On-design, 107	168
Off-design, 43	67
Off-design, 0	0

The fuel required for Renova was calculated with 11.3 MJ/kg waste as the heat value [51]. For Göteborg Energi, a heat value of 10.6 MJ/kg wood chips was estimated based on the fuel usage in 2019 and the produced energy [47].

For the shared regeneration facility, only the company with the regeneration plant had to increase their steam production. Renova was considered as the location for the plant and therefore had larger flue gas flows. Göteborg Energi had the same flue gas flows as in Table 3.8 for Scenario 1. Table 3.14 shows the new heat production and the corresponding flue gas flows from Renova.

Table 3.14: Design points and flue gas flows for Scenario 2, maintained district heat delivery, considering a shared desorber. The new combinations from the shared case are stated as Göteborg Energi-Renova.

Combination Scenario 2 [MW]	**Flue gas flow Renova [kNm^3/h]**
100-228	424
100-163	302
40-222	413
40-157	292
0-219	407
0-154	285
0-0	0

3.4.3 Scenario 3: Collaboration to Maintain District Heat Delivery

For the third scenario the maximum capacity was assumed to be the same as the on-design points in Scenario 1, 200 MW and 100 MW for Renova and Göteborg Energi respectively, see Section 3.4.1. However, when the CHP plants were not operating at maximum capacity the fuel usage was increased to cover the lost district heat as in Scenario 2. The additional boiler duty and the resulting flue gas flow were calculated according to Section 3.4.2. Since Renova was not operating over the existing capacity, extracted steam was assumed to be used in the separate case. Therefore, the district heating and electrical efficiencies for Renova were based on the results from Scenario 1 with extracted steam. For Göteborg Energi and the shared case, primary steam was used. The loss in electricity production was calculated using the method in Scenario 2 for the off-design points. For the on-design points, when district heat was still lost, the electricity loss was calculated according to Table 3.11.

The operating points at Renova and the design points for Scenario 3 compared to Scenario 1 are shown in Figure 3.13. The figure indicates that Renova had the same on-design point at 200 MW for both Scenario 1 and 3, but the off-design was increased in Scenario 3. The resulting flue gas flows can be seen in Table 3.15.

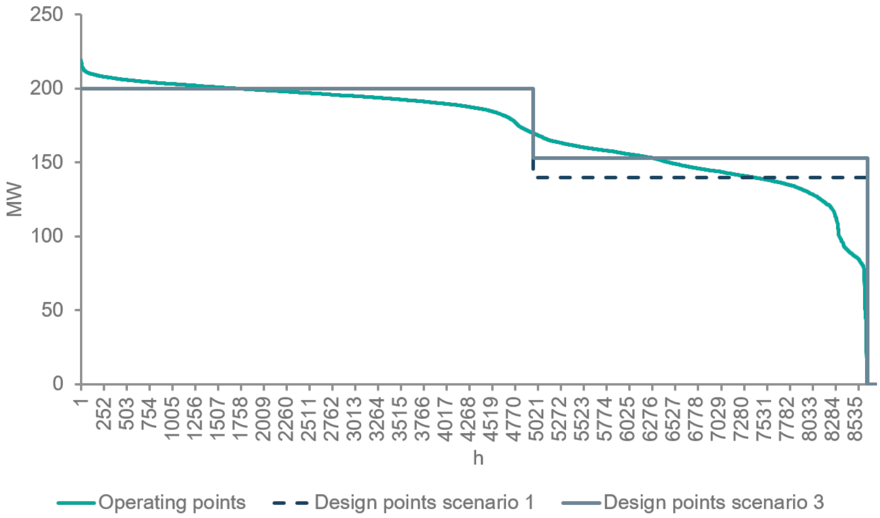

Figure 3.13: Operating and design points for Renova in Scenario 3, collaboration to maintain district heat delivery, compared to Scenario 1, maintained fuel usage.

Table 3.15: Design points and flue gas flows for Scenario 3, collaboration to maintain district heat delivery, Renova.

Design point [MW]	**Flue gas flow** [kNm^3/h]
On-design, 200	371
Off-design, 153	284
Off-design, 0	0

The design points for Göteborg Energi are presented in Figure 3.14 and it shows that the on-design point was the same for both Scenario 1 and 3, in contrast to the off-design point which had a higher value in Scenario 3. The flue gas flow for the specific design points can be seen in Table 3.16.

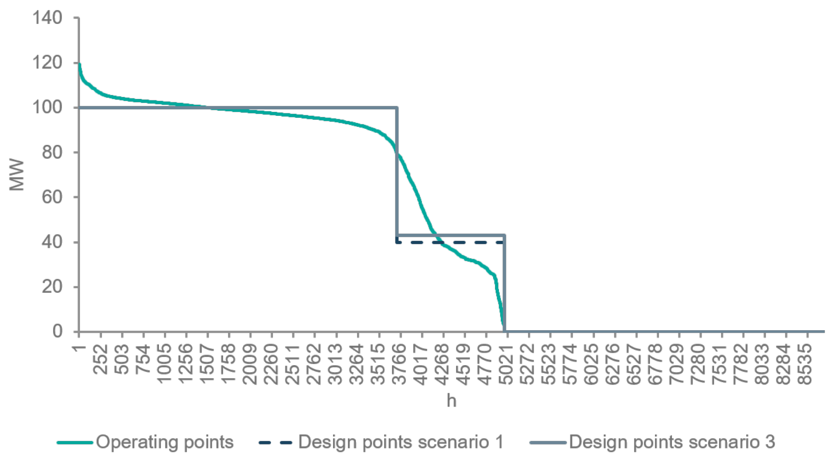

Figure 3.14: Design points for Göteborg Energi in Scenario 3, collaboration to maintain district heat delivery, compared to Scenario 1, maintained fuel usage.

Table 3.16: Design points and flue gas flows for Scenario 3, collaboration to maintain district heat delivery, Göteborg Energi.

Design point [MW]	Flue gas flow [kNm^3/h]
On-design, 100	157
Off-design, 43	67
Off-design, 0	0

For the shared facility the two CHP plants were assumed to collaborate to maintain the district heat production as much as possible. Since the reboiler was located at Renova and waste was a cheaper fuel, the additional heat was produced at Renova when possible. During maximum production at Renova, Göteborg Energi was instead assumed to increase the fuel use to cover the extra heat needed. If both plants were operating at their on-design points no extra heat could be produced and this led to a loss in district heating. The estimated design points are shown in Figure 3.15, with the respective flue gas flow in Table 3.17.

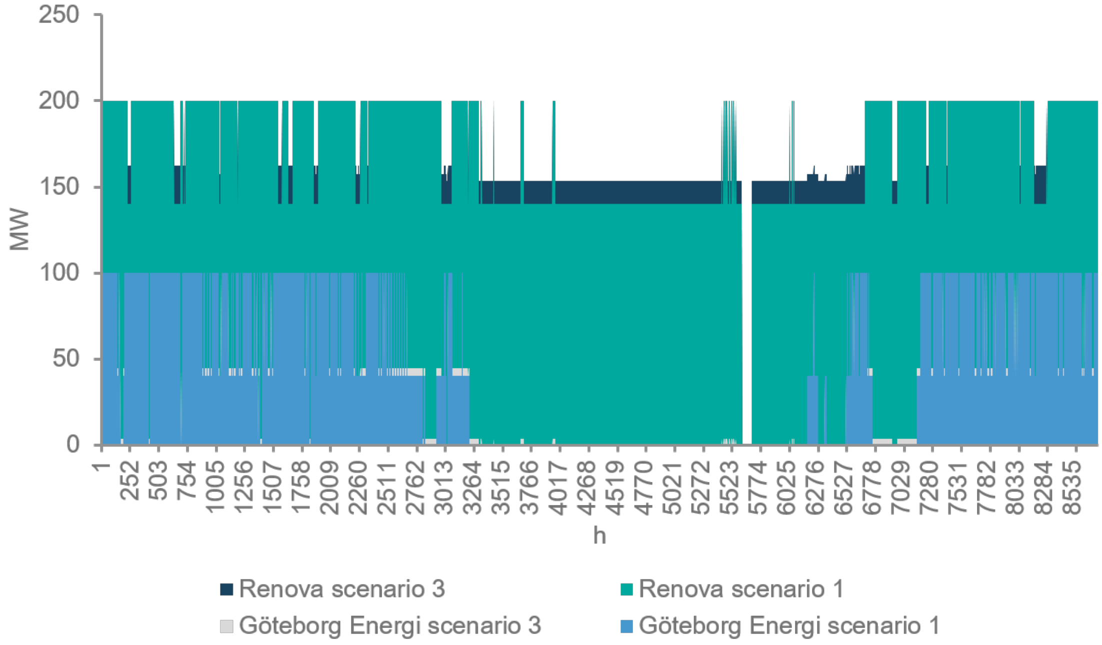

Figure 3.15: Design points for a shared facility in Scenario 1, maintained fuel usage, and Scenario 3, collaboration to maintain district heat delivery.

Table 3.17: Design points and flue gas flows for Scenario 3, collaboration to maintain district heat delivery, considering a shared desorber. In the combinations, the first number represents the design point for Göteborg Energi and the second for Renova.

Combination Scenario 3 [MW]	**Flue gas flow Göteborg Energi [kNm^3/h]**	**Flue gas flow Renova [kNm^3/h]**
100-200	157	372
100-162	157	302
44-200	69	372
40-157	63	292
4-200	4.5	372
0-154	0	285
0-0	0	0

3.4.4 Scenario 4: Reduced Capture Rate

The fourth scenario with reduced capture rate was decided based on a lower production in Renova. This was performed to see the effects from a capture plant when it was used on two processes with similar size. Each of the design points were chosen as 60% of the original production for Renova, which would represent two of their largest boilers [51]. Göteborg Energi was kept at the same points as previously. The new design points for Renova can be seen in Figure 3.16. This could also be seen as a partial capture compared to the total flue gases.

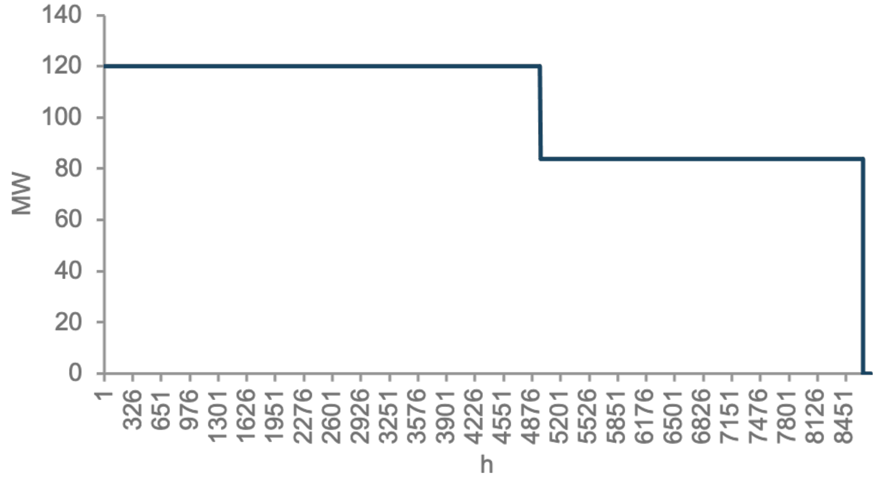

Figure 3.16: Design points for Renova with 60% of the original heat production for Scenario 4, reduced capture rate.

The corresponding flue gas flows for each design point can be seen in Table 3.18. Göteborg Energi had the same design points as in Scenario 1, Table 3.8. For the shared desorber, the combinations of the design points were used and the operating hours were the same as in Table 3.10 for Scenario 1, but with the new design points for Renova. The electricity loss was calculated in the same manner as in Scenario 1, see Table 3.11.

Table 3.18: Design points and flue gas flows for Renova in Scenario 4, reduced capture rate.

Design point [MW]	Flue gas flow [kNm^3/h]
On-design, 120	223
Off-design, 84	156
Off-design, 0	0

3.5 Economic Analysis

The economic analysis included evaluations of the capital expenditures (CAPEX) and operational expenditures (OPEX) for the carbon capture plant, the liquefaction, on-site storage and transport. The method and the assumptions for the analysis are presented in this section.

3.5.1 Capital Expenditures of the Capture Plant

The CAPEX calculation was based on the work by Eliasson and Fahrman [63], where literature values for different equipment costs were used to construct cost curves based on the power-law. Each of the curves was related to the size of the equipment, such as the volume [63]. To account for other parameters, for example material, pipes and labour cost, factors from Ali et al. [38] were used in the Enhanced Detailed Factor method. The cost from the power-law functions had to be converted to Norwegian Kroner (NOK), with the use of the exchange rate, due to the factors from Ali et al. being based on NOK. The chosen materials for each unit can be seen in Table 3.19, where stainless steel (ss) have two different treatments, either welded or machined [63]. The table also includes the material factors used for converting the material to carbon steel [38]. A flowsheet is shown in Appendix A for the included units in the CAPEX calculations.

Table 3.19: Material specifications.

Unit	Material [63]	Material factor [38]
Columns and vessels	welded ss316	1.75
Heat exchangers, coolers, reboiler	welded ss316	1.75
Fan	carbon steel	1
Pumps	machined ss316	1.30

The factors taken into consideration in the Enhanced Detailed Factor method from Ali et al. can be seen in Equation (3.33) for carbon steel [38]. The sum of these factors need to be adjusted based on the material according to Equation (3.34) which also considered the piping cost and a value of 1 for the equipment cost [38].

$$F_{Total,\ CS} = f_{direct} + f_{engineering} + f_{administration} + f_{contingency} \tag{3.33}$$

$$F_{Total,\ material} = F_{Total,\ CS} + (f_{material} - 1)(f_{equipment} + f_{piping}) \tag{3.34}$$

The equipment cost was used in combination with the final material factor [63] in order to calculate the equipment installed cost (EIC) as in Equation (3.35) [38].

$$EIC = Equipment\ Cost \cdot F_{Total,material} \tag{3.35}$$

The sum of the EIC for all of the equipment gave the Total Installed Cost (TIC) together with the exchange rate for NOK to EUR [38]. The required volume of MEA, based on 40 minutes as assumed recirculation time for the MEA flow, was used to calculate the start-up MEA investment cost. The time was based on an article by Montañés, Flø and Nord [80] and the assumption in the work by Eliasson and Fahrman [63]. For the shared case, the flow of Göteborg Energi was transported to the desorber at Renova and back, therefore 40 minutes was added to the MEA recirculation time for Göteborg Energi. These considerations for the TIC can be seen in Equation (3.36) [38, 63].

$$TIC = (\sum EIC + MEA\ investment) \cdot Exchange\ rate \cdot f_{location} \tag{3.36}$$

The electrical work and CAPEX for additional pumps were calculated based on an estimated pressure drop and the mass flow. The pumps were seven in total, see Appendix A, where one was simulated in Aspen Plus. The pump work required was calculated as in Equation (3.8). The extra pumps needed with a shared regeneration facility were previously described in Section 3.1.2. Based on the temperature differences, flows and assumed overall heat transfer coefficients the heat transfer areas for the coolers, reboiler and the heat exchanger were calculated.

An annuity factor was used to divide the TIC over the plant's lifetime in order to calculate the annual CAPEX. The interest rate was converted to an annuity factor, r, according to Equation (3.37) [38] which depended on the number of operating years, n.

$$r = \sum_{n=1}^{23} \frac{1}{(1+p)^n} \tag{3.37}$$

The technology was assumed to be nth of a kind with an assumed lifetime of 25 years, including two years for the construction of the plant [63]. Table 3.20 shows the lifetime, the assumed interest rate as well as the Chemical Engineering Plant Cost Index (CEPCI), which was used to convert all costs to the year 2016 [63].

Table 3.20: Assumed values for the CAPEX calculation.

Parameter	Value	Source
Lifetime [years]	25	[63]
Interest rate, p [%]	7.5	[38]
Cost year	2016	
Exchange rate 2016 [NOK/€]	9.7	[63]
CEPCI 2016	98	[81]
$f_{location}$, Rotterdam	1	[38]

The assumed cost of the pipeline material and installation is presented in Table 3.21 for 2016. The material was chosen based on the cost of the pipeline together with the additional lean/rich heat exchanger required.

Table 3.21: Assumed cost for the pipeline depending on material.

Material	Cost [€/m]
Manganese steel	885
Polyethylene (plastic)	619

3.5.2 Operational Expenditures of the Capture Plant

In the OPEX, electricity cost for the pumps and the fan, steam demand for the reboiler, make-up MEA and cooling demands were included. The OPEX was calculated for every hour depending on the estimated design points and summed up over the year. To take the variations of the electricity market into account, the cost of the electricity was calculated through the use of the hourly electricity prices from Nordpool in 2019 [82].

The cost for the cooling water was assumed to be constant over the year and the used value can be seen in Table 3.22. The assumed cost for MEA and the extra fuel, which was required for the second and third scenario, are also included in the table. The amount of make-up MEA needed, due to losses and degradation [83], was calculated based on the assumption of 1.2 kg MEA/ton CO_2 [83, 84]. The cost for the reboiler heat demand in Scenario 1 and 4 was calculated based on the lost district heat and the lost electricity. The monthly revenue for district heat, varying between 1-18 €/MWh over the year, was used as the steam price in combination with lost revenue from electricity production.

In Scenario 2 and for the off-design points in Scenario 3, the cost for the additional fuel was used for the estimation of the steam price. However, for the on-design points in Scenario 3, with collaboration to maintain district heat delivery, the steam cost was estimated as the lost revenue in district heat and electricity. The fuel for Renova is waste, which the company take care of in exchange for payments from the waste producers. However, because of additional costs with a higher boiler load, for example considering flue gas purification, it was assumed to not be a cost nor an income and therefore set as zero. At Göteborg Energi, the cost for the extra fuel was calculated with a wood chip price of 250 SEK/MWh for 2019, obtained from Energimyndigheten [85] and converted to €/MWh with the exchange rate in Table 3.22. The boiler efficiency was also considered when calculating the fuel cost, it was assumed to be 0.9 based on the fuel input and the produced heat in 2019 [47].

Table 3.22: Assumed cost parameters for the OPEX calculation.

Parameter	Value	Source
Cooling water cost, 15-20 °C [€/m^3]	0.02	[39]
MEA cost [€/m^3]	2000	[63]
Fixed OPEX [% of CAPEX]	6	[63]
Fuel cost for wood chips 2019 [€/MWh]	22	[85]
Exchange rate 2019 [SEK/€]	11.3	[86]

The electricity, heating, and cooling demands were obtained from the simulations in Aspen Plus for the different scenarios. The cost for each demand was calculated annually and was summed as the variable OPEX. To include maintenance, labour, replacement of material and insurances, a fixed OPEX was calculated as 6% of the CAPEX [28].

3.5.3 Capital and Operational Expenditures of Liquefaction

The method for the economy of the liquefaction plant was based on the work by Eliasson and Fahrman [63] and Deng et al. [28]. The capital cost for the process was determined based on the size and cost of the liquefaction plant in the study of Deng et al. [28]. Based on the process of Deng et al., a reference CO_2 flow, $\dot{m}_{ref}$, and a Total Delivered Cost (TDC_{ref}), see Table 3.23, was used to calculate a new TDC in Equation (3.38) with the simulated CO_2 flow [63].

An owner cost and an indirect cost was taken into consideration together with the TDC as well as a factor for the project contingency in order to obtain the TIC, see values in Table 3.23 [28]. A process contingency of 20% was assumed, which was used in the other economic analysis in this thesis.

$$TDC = (\frac{\dot{m}}{\dot{m}_{ref}})^{0.6} \cdot TDC_{ref} \tag{3.38}$$

Table 3.23: Assumed cost parameters for liquefaction.

Parameter	**Value**	**Source**
Owner cost [%]	7	[28]
Indirect cost [%]	14	[28]
Project contingency [%]	20	[28]
Process contingency [%]	20	[39]
Maintenance cost of CAPEX [%]	6	[63]
$\dot{m}_{ref}$ [kg CO_2/s]	37.31	[28]
TDC_{ref} [M€]	23.9	[28]

The OPEX for the liquefaction process was divided into fixed and variable. In the fixed OPEX the maintenance was included, see Table 3.23. The variable OPEX included the costs of the power and cooling demands calculated in the same manner as for the capture plant. From the simulations, the power needed in the three compression stages and the cooling demand in between were obtained. Power demands of the compression in the recirculation and the ammonia cycle as well as the cooling demand in the pre-cooler and the NH_3 condenser were calculated based on results from Deng et al. described in Section 3.2.

3.5.4 Capital and Operational Expenditures of Transport and On-Site Carbon Dioxide Storage

The cost for transport was calculated based on the number of trucks required and the distances. In Table 3.24, the assumed costs for 2016 such as the cost of a truck, diesel and the driver's wages can be seen.

Table 3.24: Assumed cost parameters for truck transport.

Parameter	Value
Cost per truck, C_{truck} [k€]	280
Diesel cost [€/litre]	1.2
Driver wages [k€/year]	80
Maintenance cost [% of CAPEX]	5

The CAPEX for transport was calculated as the total cost for the trucks. Therefore, the number of trips for one driver per day was calculated in Equation (3.39) based on the driver's workday, t_{work} as 8 h/day, and the time required for one trip, back and forth, see Table 3.4 in Section 3.3. The maximum CO_2 flow for a day was taken from the on-design due to it resulting in the maximum number of vehicles required.

$$N_{trips} = \frac{t_{work}}{t_{trip}} \tag{3.39}$$

Each day was assumed to consist of three shifts and the total number of trips for a truck was therefore three times the trips for one driver. The number of trucks, N_{trucks}, was calculated through Equation (3.40). Truck specifications are seen in Table 3.4.

$$N_{trucks} = \frac{\dot{V}_{CO_2}}{N_{trips/truck} \cdot T_{capacity}} \tag{3.40}$$

The annual CAPEX for the trucks was calculated with Equation (3.41). The truck lifetime, n_{truck}, was estimated to be 10 years.

$$Annual\ CAPEX = \frac{C_{truck} \cdot N_{trucks}}{n_{truck}} \tag{3.41}$$

The OPEX was divided into fixed and variable, where the former included the maintenance cost and driver cost and the latter the diesel cost. The maintenance cost was calculated based on the CAPEX of the trucks, see Table 3.24. The number of drivers was calculated with consideration of working 40 h/week and an operating time of 24 h/day, $t_{operating}$. Equation (3.42) shows how the number of drivers, $N_{drivers}$, was calculated and the cost of the drivers was estimated through the wages.

This was seen as a part of the fixed OPEX because the number of drivers was assumed to be constant during the year.

$$N_{drivers} = \frac{7 \cdot t_{operating}}{t_{work}} \cdot (N_{trucks}) \tag{3.42}$$

The variable OPEX, the diesel cost, was calculated based on the number of trips per day which was estimated by the use of the CO_2 flow/day and the loading capacity of the truck. Consideration to the different design points were taken through the different CO_2 flows.

For the on-site storage, the cost was divided into the CAPEX for the tank and for the loading pump as well as the OPEX for the loading pump. The cost estimations and the power demand for the pump can be seen in Table 3.25. The previously used electricity cost was applied and for each operating hour the same power demand was assumed based on the pump capacity. The fixed OPEX for the loading pump was also considered in the form of the maintenance cost. The annual CAPEX was calculated based on the previously used annuity factor.

Table 3.25: Cost estimations for on-site storage.

Parameter	Value
CAPEX tank [kSEK/m^3]	47.53
CAPEX loading pump [kSEK]	571
Power demand loading pump [kW]	2.28
Maintenance cost [% of CAPEX loading pump]	4

3.6 Sensitivity Analysis

The sensitivity analysis was performed for the third scenario because it was seen as the most realistic in practice. Six parameters were investigated with an increase and decrease of 50%. The CAPEX was chosen as one of the parameters due to the size of the cost and the dependence on several factors such as the material cost and the installation [38]. The electricity and the steam cost were also two parameters included in the sensitivity analysis due to the yearly fluctuations and possible future changes [39]. The fourth parameter was the cooling utilities and the associated cost due to it being a limited resource for the two CHP plants. The fifth parameter considered the transport through the diesel cost. During a year the diesel cost varies and the truck fuel might also be replaced during the plant lifetime, which would impact the price. The last parameter was the fuel price for the CHP plants. When extra incineration to maintain district heat delivery was considered, the fuel price could have more influence on the total cost through market variations and was therefore included in the sensitivity analysis.

3.7 Business Cases

In order to divide the cost for the shared plant between the companies, three different business cases were investigated. The first cost allocation was calculated based on the separate costs and how large share each of the companies had compared to the sum of the separate cost. The second allocation method used their respective emissions as base. The third method included dividing the cost savings equal between the companies, 50/50. The business cases were compared for Scenario 1 and 3.

4

Results and Discussion

The steam and cooling demand as well as the loss in district heat and electricity production are shown in this chapter. CAPEX and OPEX, both variable and fixed, are presented for each scenario and discussed. For Scenario 3, collaboration to maintain district heat delivery, the results for the sensitivity analysis can be seen. A comparison between the costs for each scenario and the results of allocating the shared cost between the companies are presented and discussed. A discussion of future work is also included in this chapter.

4.1 Scenario 1: Maintained Fuel Usage

For Scenario 1 with maintained fuel usage, results for the utility demands concerning cooling water and heat as well as the amount of lost district heat and electricity are shown. The results for the transport both concerning the solvent pipeline and the truck transport is shown for Scenario 1. The economic results are also presented in total and for the separate parts such as CAPEX and variable OPEX.

4.1.1 Impact on CHP Production and Utility Demands

In Table 4.1 the amount of recovered district heat is shown. The presented cooling demand is only considering the use of cooling water for lower temperatures. Both the district heat and cooling demand include the contribution from the liquefaction. The heat demand in form of steam and lost electricity in the steam cycle are also shown for the different design points. Around 70% of the heat demand in the reboiler is recovered as district heat. The specific heat demand in the reboiler is around 3.3 MJ/kg CO_2 for all on-design cases, however for different off-design points it varies between 3.2 to 3.3 MJ/kg CO_2.

Table 4.1: Steam and cooling demand for the design points in Scenario 1 with maintained fuel usage and recovered district heat and lost electricity production. The combinations are Göteborg Energi-Renova.

	Design point [MW]	Steam demand at 3.5 bar [MW]	Recovered district heat [MW]	Cooling demand 15-20°C [MW]	Lost electricity [MW]
Separate, Renova	200	61	43	34	7.4
	140	41	29	21	5.2
Separate, Göteborg Energi	100	28	21	16	2.1
	40	11	7.7	5.4	0.86
Shared	100-200	89	62	52	21
	100-140	68	48	37	11
	40-200	71	50	40	17
	40-140	51	35	27	8.6
	0-200	61	43	34	15
	0-140	41	28	21	6.8

The yearly heat demand is similar for the sum of the separate capture plants for Renova and Göteborg Energi compared to the shared heat demand, see Figure 4.1. The same applies for the recovered district heat from the capture plant and the liquefaction. The loss in heat and electricity production, the difference between the heat demand and recovered district heat, is divided differently depending on the use of extracted steam or primary steam. For Renova, with the use of extracted steam, the lost district heat is higher compared to the other plants but the lost electricity is lower. For Göteborg Energi and the shared case, primary steam was used causing a higher loss in the electricity production. The other utilities such as power and cooling for the capture plant and the liquefaction do not have a large difference between the shared case and the sum of the separate plants, see Figure 4.2.

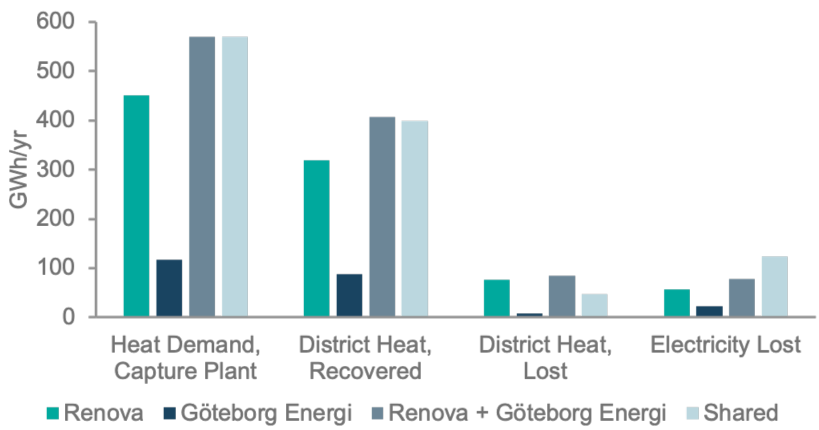

Figure 4.1: Heat demand, lost district heat and lost electricity for Scenario 1 with maintained fuel usage. The steam used had a pressure of 3.5 bar. The district heat recovered also includes the heat recovery from the liquefaction.

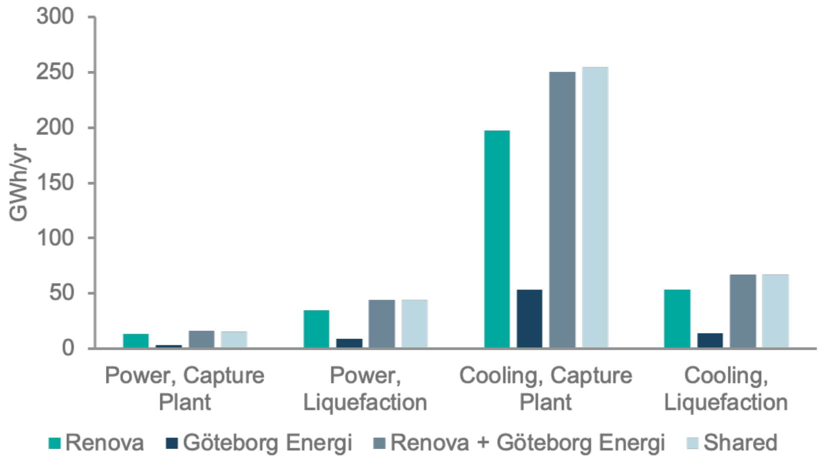

Figure 4.2: Power and cooling demand for Scenario 1 with maintained fuel usage. The cold water used for cooling had a temperature between 15-20 °C.

4.1.2 Economic Evaluation

The total cost for the considered system, capture plant to intermediate storage, is visualised in Figure 4.3. The bars and the left axis show the annual cost and the dots relate to the right axis for the specific cost. The summarised cost for the two separate plants is shown in the figure and was calculated based on the sum of the annual costs for the two plants divided by the total yearly captured CO_2. For the separate cases, Göteborg Energi has a much higher specific cost compared to Renova due to less operating hours. The shared case has a slightly smaller total cost although a larger variable OPEX compared to the summarised separate cases. For the shared case, both the CAPEX and the fixed OPEX are lower costs.

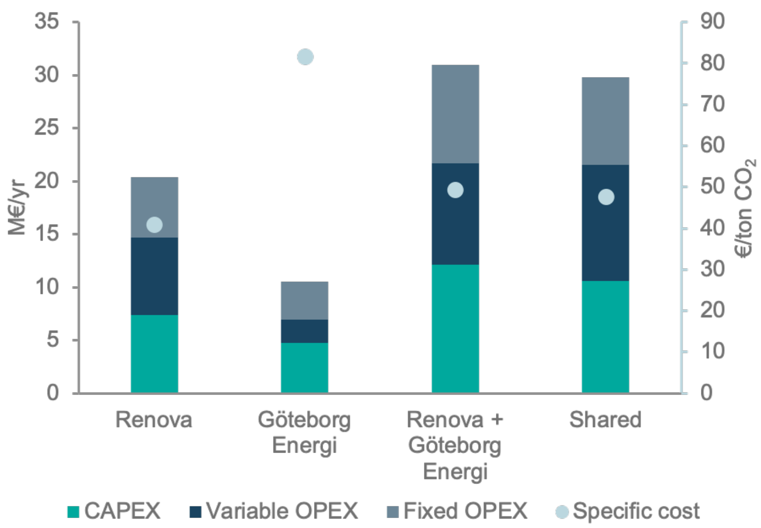

Figure 4.3: Annual and specific total cost for the studied system in Scenario 1, maintained fuel usage. The annual cost, the bars, corresponds to the left axis and the specific cost, the dots, is presented at the right axis.

The specific costs for Scenario 1 are based on the captured amount of CO_2, corresponding to a capture rate of 90%, shown in Table 4.2 for each case.

Table 4.2: Captured CO_2 in Scenario 1 with maintained fuel usage for each facility corresponding to a 90% capture rate.

	Captured CO_2 [kton/year]
Separate, Renova	500
Separate, Göteborg Energi	130
Shared	630

The increase in variable OPEX in the shared case stems from the use of primary steam. For Renova in the separate case, as mentioned previously, extracted steam was used resulting in a higher district heat loss but less reduction of produced electricity, which leads to a lower variable OPEX. If primary steam would be required for Renova's separate capture plant, the cost would increase with around 9% to 45 €/ton CO_2. The cost for the sum of Renova and Göteborg Energi would increase with 6% to 52 €/ton CO_2, leading to a larger difference between the cases and larger possible savings with the shared case.

The presented results include the possibility for district heat integration, which lowers the cost. If the integration would not be considered the annual costs would increase with 15-20% due to the larger loss in income from district heat. Heat integration becomes an important aspect in order to lower the costs.

The loss in electricity production is also seen in the results for the specific variable OPEX, Figure 4.4. The largest contribution comes from the required and lost power, where the latter has the greater share. It can also be seen that the cost for lost district heat is lower for the shared case which again corresponds to the lower lost annual production, as previously seen in Figure 4.1. The largest differences relates to the use of extracted or primary steam. The cooling demands only include the cooling water, at 15-20 °C, and not the use of district heat. As mentioned previously, Göteborg Energi has large values due to the lower operating hours and emissions.

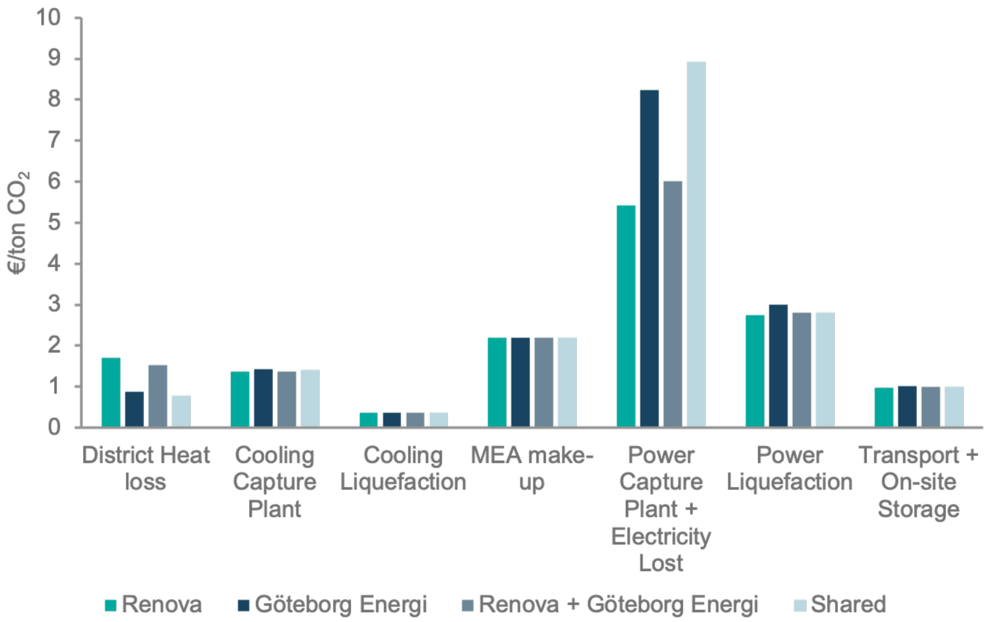

Figure 4.4: Specific variable OPEX for Scenario 1 with maintained fuel usage.

In Table 4.3 the diameter of the absorber and desorber can be seen for each case. The absorber diameters for the separate capture plants are equal to the shared case since the absorbers were separate. However, for the shared desorber a greater diameter is required due to a larger flow rate. The cross-sectional area of the shared desorber is 2% larger, however the surrounding area required for two columns compared to one has not been included.

Table 4.3: Diameter of the columns for separate capture plants and a shared desorber for Scenario 1, maintained fuel usage.

		Absorber diameter [m]	**Desorber diameter [m]**
Separate	Renova	7.6	4.2
Separate	Göteborg Energi	5.0	2.8
Shared	Renova	7.6	5.1
	Göteborg Energi	5.0	

The annual CAPEX for the capture plant, liquefaction, transport and on-site storage, are presented in Figure 4.5. The cost of the shared desorber is slightly lower compared to the two separate desorbers. Similarly, the liquefaction and pump costs are lower despite a larger capacity. In the figure, the cost for the heater and the coolers is also seen to be a bit lower for the shared case and the largest difference originates from a lower reboiler cost due to the use of only one reboiler. The transport and on-site storage do not result in a difference. The absolute possible savings in CAPEX for the capture plant is around 6.4 million € and for the liquefaction plant around 9.9 million €.

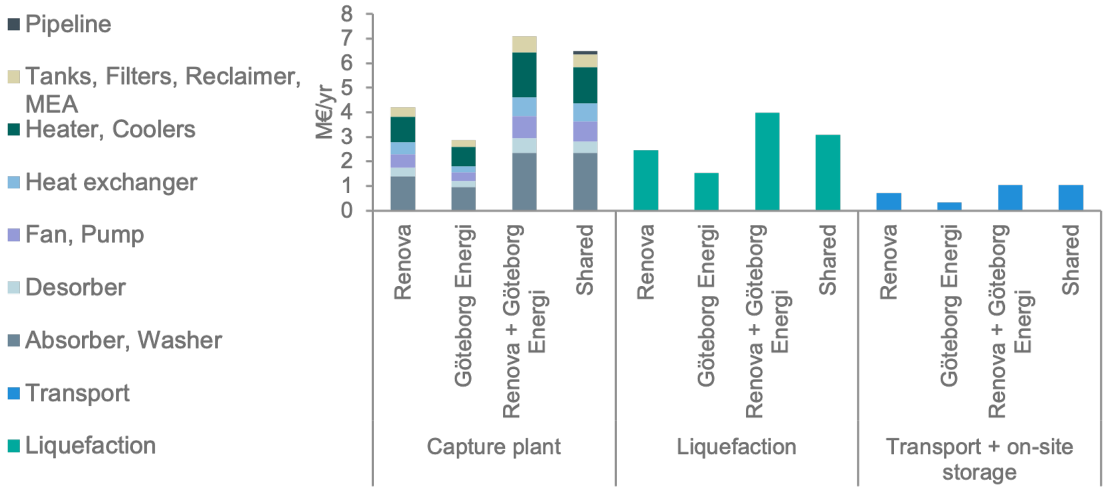

Figure 4.5: Annual CAPEX for Scenario 1 with maintained fuel usage including the MEA start-up cost.

Furthermore, the steel pipeline cost is not a large contribution in the annual CAPEX, as seen in Figure 4.5. The detailed cost of the pipeline depending on the material choice can be seen in Table 4.4. The cost for the polyethylene pipeline together with the additional heat exchanger is higher compared to steel due to the higher amount of transferred heat in the steel pipeline. Therefore, a smaller additional heat exchanger is required and the total CAPEX is lower. Consequently, the manganese steel pipeline is used in all cases. The costs were calculated with the heat exchanger after the absorber and the pipeline before the desorber. If the pipeline is placed first, the CAPEX would increase with 5% due to less efficient heat exchanging depending on the temperature differences.

Table 4.4: Heat transfer area, pipeline cost and the additional heat exchanger depending on pipeline material for Scenario 1 with maintained fuel usage.

Design		**Transferred heat [MW]**	**Inner/outer pipe diameter [m]**	**Heat transfer area [m^2]**	**CAPEX [k€]**
1	Manganese steel pipeline	9.6	0.5/0.71	1570	890
	Additional heat exchanger	27.7		2110	2180
2	Polyethylene pipeline	0.53	0.5/0.71	1570	620
	Additional heat exchanger	36.8		2730	2760

The transport required after the capture plant does not differ between the shared case and the summarised separate cases, see Table 4.5, because it was calculated based on the captured CO_2 flow which in total became equal. In practice, there could be further differences, considering for example the administration and logistics, where it could be beneficial to have one storage site where the trucks load. A challenge could instead be the restricted loading capacity, if several trucks would load at the same time from one storage site.

Table 4.5: Results for transport and on-site storage considering number of trucks, drivers and storage size for the on-design point in Scenario 1, maintained fuel usage.

	On-site storage [m^3]	**N$_{trips}$ [No./day]**	**N$_{drivers}$ [No.]**	**N$_{trucks}$ [No.]**
Separate, Renova	1515	46	17	4
Separate, Göteborg Energi	700	22	9	2
Shared	2215	68	26	6

The transport and on-site storage are not large contributions to the specific cost, which is seen in Figure 4.6. The overall cost is shown separately for the capture plant, the liquefaction and transport and on-site storage. Due to the larger variable OPEX the shared capture plant has a higher specific cost. The process part with possible savings is the liquefaction through a lower CAPEX and fixed OPEX.

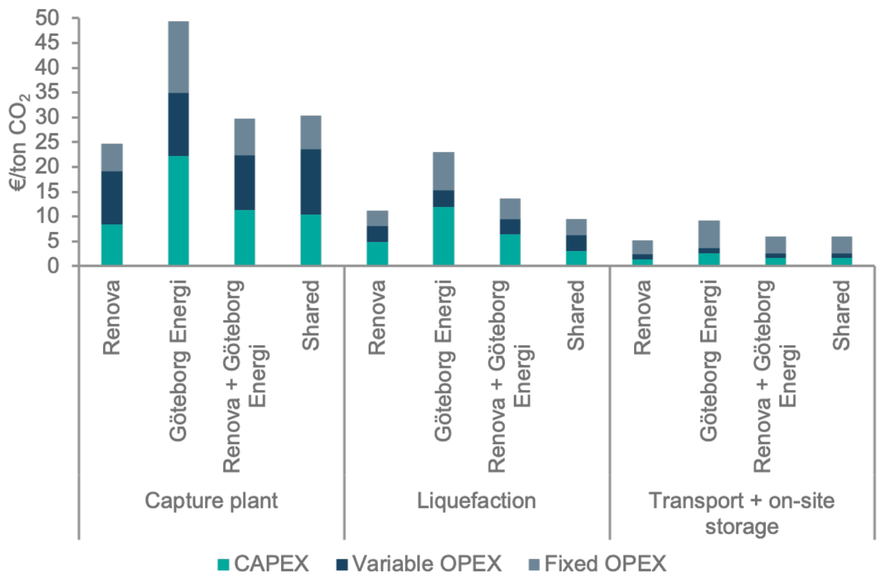

Figure 4.6: Specific cost for the capture plant, liquefaction, transport and on-site storage considering CAPEX, variable OPEX and fixed OPEX for Scenario 1 with maintained fuel usage.

Renova was used as the location for the desorber, however the cost for the alternative with Göteborg Energi was calculated and is 3% more expensive in specific cost but around 16% higher in annual cost. A part of the reason for a higher cost is the requirement for Göteborg Energi to use more fuel to cover the reboiler duty. In addition, the recovered district heat from a higher fuel usage can not be utilised during all times of the year and therefore requires a larger cooling demand.

4.2 Scenario 2: Maintained District Heat Delivery

For Scenario 2, the results are presented in the form of the required capacity and fuel usage to maintain the district heat delivery. The economic results are also presented as specific costs depending on the captured amount of CO_2.

4.2.1 Capacity and Fuel Usage

To be able to produce the required heat, the capacity in the boilers had to be investigated. In Table 4.6, the capacity available today and the required capacity in Scenario 2 are stated in MW. The existing capacity for each facility was estimated as the maximum production in 2019. The separate plants do not exceed the available capacity, therefore the capture plant can be implemented without expanding the capacity of the boilers. For the shared plant, a higher capacity than available today is required, which means that the CHP plant has to be rebuilt. For the separate capture plant at Renova the required capacity is almost reaching the maximum capacity. This could lead to problems in the operation of the plant and the boiler may need to be rebuilt to have larger margin.

Renova requires permissions for incineration of waste and their current allowed fuel usage and the fuel usage for Scenario 2 is stated in Table 4.6. The results show that the fuel usage in Scenario 2 does not exceed the permissions and no new permissions are needed. Göteborg Energi has no restrictions of how much fuel can be incinerated due to the use of wood chips. However, the company's fuel usage for 2019 was 157 kton, which is in the same order as the fuel usage for Scenario 2.

Table 4.6: Present capacity of the CHP plants and the required capacity in Scenario 2, maintained district heat delivery. The desorber was placed at Renova for the shared case.

Plant	Existing capacity [MW]	Required capacity [MW]	Allowed fuel usage [kton/yr]	Fuel usage Scenario 2 [kton/yr]
Separate, Renova	220	219	550	535
Separate, Göteborg Energi	120	107	-	160
Shared, Renova	220	228	550	547

4.2.2 Economic Evaluation

The total cost for the second scenario with maintained district heat delivery is presented in Figure 4.7 and shows a lower annual and specific cost for the shared facility compared with the separate capture plants at Renova and Göteborg Energi together.

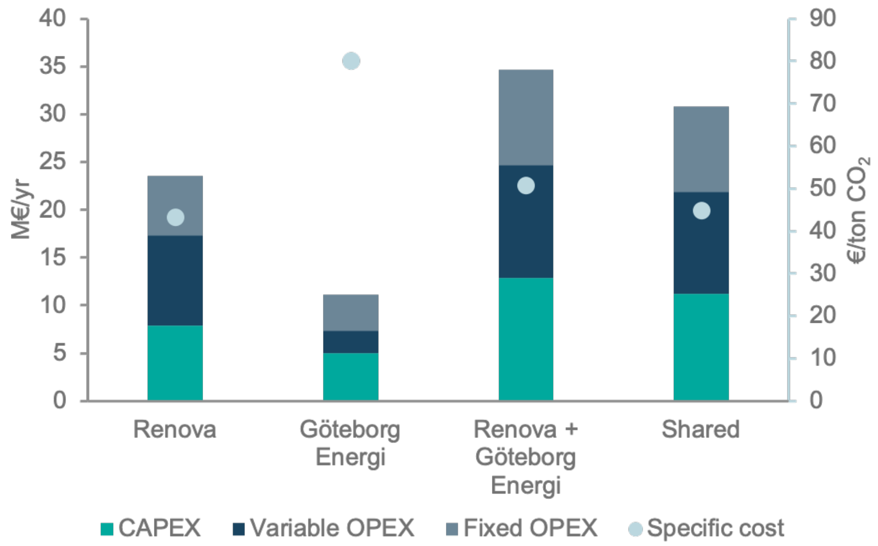

Figure 4.7: Total cost for Scenario 2 with maintained district heat delivery.

The difference in cost is larger compared to Scenario 1, see Figure 4.3. The reason being that in Scenario 2, both Renova and the shared plant used primary steam and the reduction of electricity production was more similar between the plants. This can also be recognised in Figure 4.8 as the specific cost for the power demand and power lost is lower for the shared plant compared to Renova and Göteborg Energi together. Furthermore, the figure shows the specific cost for the extra fuel since additional incineration was required to maintain the district heat delivery. Only Göteborg Energi has an extra fuel cost as the fuel at Renova was assumed to have no cost.

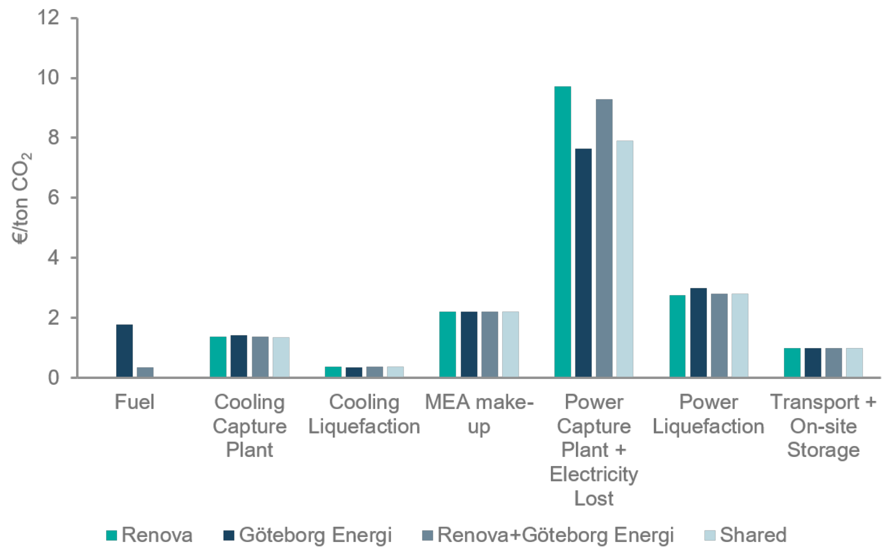

Figure 4.8: Specific variable OPEX for Scenario 2 with maintained district heat delivery.

Figure 4.9 shows the overall cost including the CAPEX and OPEX with a similar cost saving to Scenario 1 for the shared liquefaction. However, in Scenario 2, a cost saving with the shared capture plant can also be seen as a consequence of the utilisation of primary steam for all cases. The specific transport cost does not differ between the cases and compared to Scenario 1 slightly more drivers, in total 30, and trucks, in total 7, are required.

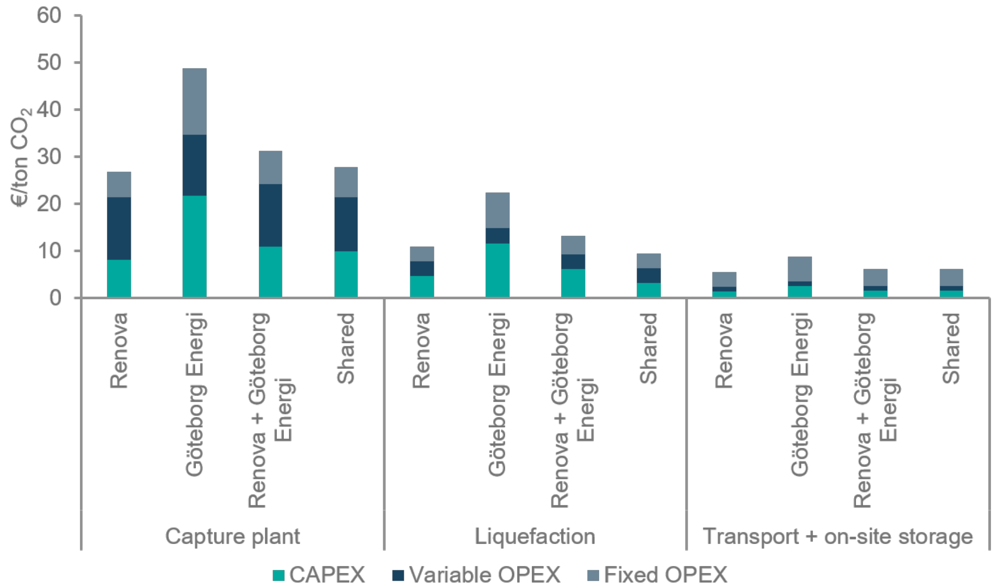

Figure 4.9: Specific cost for the capture plant, liquefaction, transport and on-site storage considering CAPEX, variable OPEX and fixed OPEX for Scenario 2 with maintained district heat delivery.

4.3 Scenario 3: Collaboration to Maintain District Heat Delivery

The results for Scenario 3 are presented in the following section, with the required fuel usage and the cost. The results from the sensitivity analysis are also shown and discussed. Scenario 3 is used to investigate how the district heat delivery will be affected if the maximum capacity can not be increased. The scenario can be more realistic in practice as no expansion of the plants is needed. The fuel usage is only increased when capacity is available and in other cases district heat is lost.

4.3.1 Fuel Usage

The fuel usage of the plants are presented in Table 4.7 regarding both the allowed amount and the required fuel in Scenario 3. For Göteborg Energi, there is no restriction for the usage of wood chips in contrast to the waste incineration at Renova. It can be seen that Renova still uses less fuel than the restricted maximum for both the separate and the shared plant. For Göteborg Energi, as mentioned, the fuel usage in 2019 was 157 kton which is around the needed amount in Scenario 3. Based on the fuel usage being lower in Scenario 3 than the allowed amount, it can be possible to increase the fuel use without a reconstruction of the CHP plants. Additionally, new permissions for a rebuild or higher fuel incineration will not be required, which can otherwise take time and be difficult to obtain.

Table 4.7: Allowed fuel usage and required amount in Scenario 3, collaboration to maintain district heat delivery.

Case	Allowed fuel usage [kton/yr]	Fuel usage Scenario 3 [kton/yr]
Separate, Renova	550	537
Separate, Göteborg Energi	-	158
Shared (Göteborg Energi, Renova)	-, 550	160, 537

In this work the start-up time and cost for the CHP plant owned by Göteborg Energi are not considered. In Scenario 3, Göteborg Energi is started several times for short periods and at low capacity, see Figure 3.15 in Section 3.4.3. This might not be applied in practice and if not, the district heat loss will be greater. However, there can be potential to save heat in the system for shorter times through accumulator tanks and thereby maintain the district heat delivery. To be certain more studies on the CHP plant should be performed.

4.3.2 Economic Evaluation

In Figure 4.10 the total cost is presented as both annual and specific cost for each case. The shared capture plant has less total cost although the difference is small compared to the separate cases. As for Scenario 1, the variable OPEX is larger for the shared case due to Renova using extracted steam in the separate case.

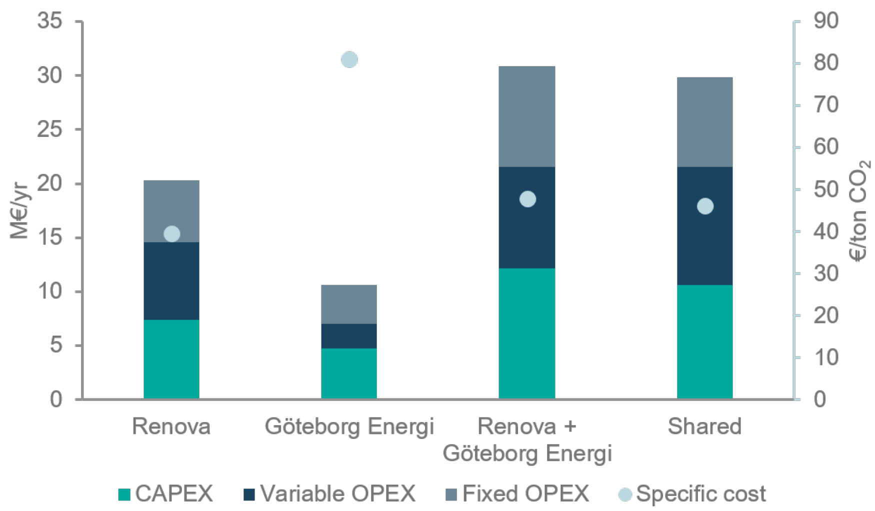

Figure 4.10: Total cost for the capture plant, liquefaction, transport and on-site storage for Scenario 3, collaboration to maintain district heat delivery.

In Figure 4.11, the total specific cost is presented for the capture plant, liquefaction plant, transport and on-site storage. The CAPEX, variable and fixed OPEX can also be seen for each case. Similar to Scenario 1, the cost saving comes from the shared liquefaction plant.

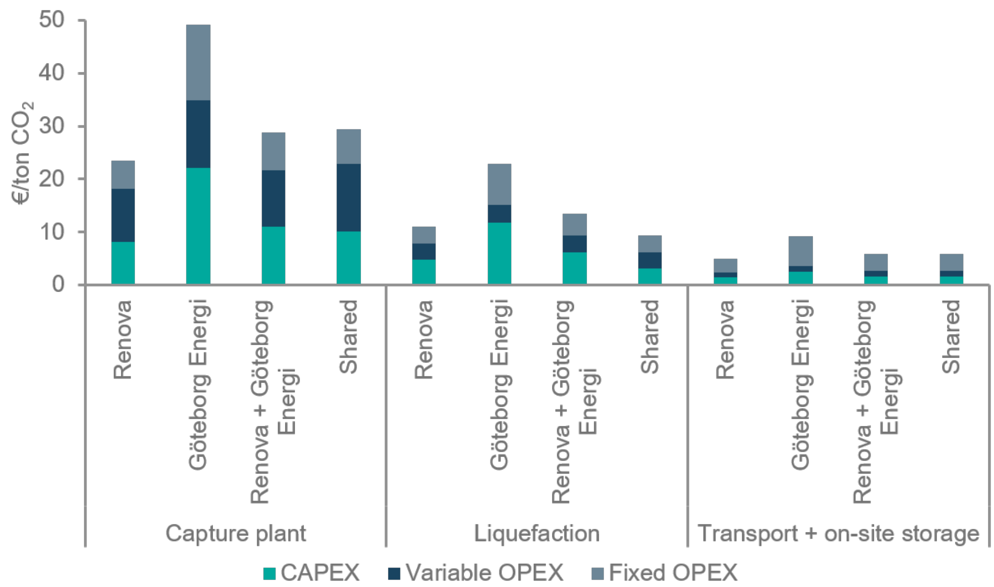

Figure 4.11: Specific cost for the capture plant, liquefaction, transport and on-site storage considering CAPEX, variable and fixed OPEX for Scenario 3 with collaboration to maintain district heat delivered.

4.3.3 Sensitivity Analysis

A sensitivity analysis was performed on Scenario 3 with an increase and a decrease of 50% for six parameters including diesel price, cooling water price, fuel price, steam price, electricity price and total CAPEX. The result for Renova is shown in Figure 4.12 and in this case CAPEX is the parameter with most impact on the specific cost due to CAPEX being a large share of the total cost. Electricity price is also an important parameter as the power consumption and lost power are a substantial part of the variable OPEX. Since, the cost is sensitive to the electricity price, it is important to consider the varying electricity market in the evaluations. Furthermore, diesel price, cooling water price, and steam price, have less effect on the total cost and the fuel price has no effect on the cost for Renova, as waste is assumed to have zero cost. Steam price refers to the district heat price, which in turn relates to lost revenue with a lower district heat production.

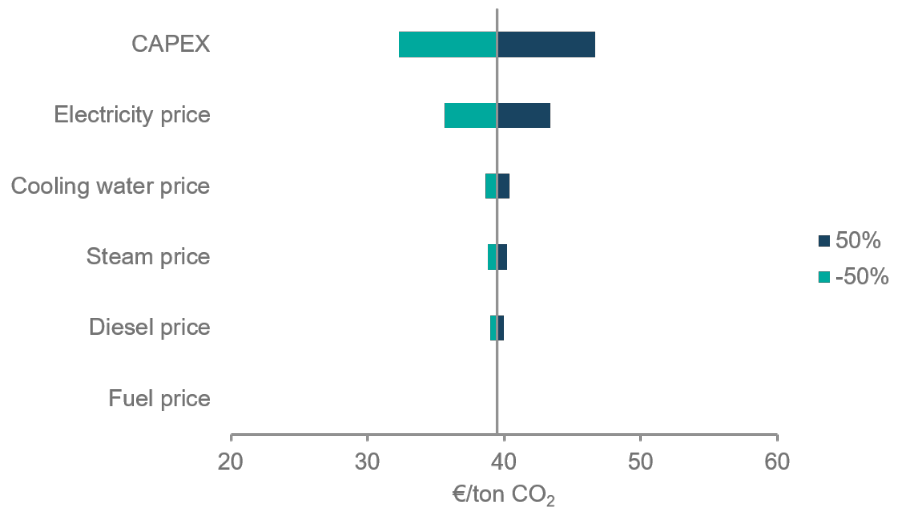

Figure 4.12: Sensitivity analysis for Renova considering Scenario 3, collaboration to maintain district heat delivery.

In Figure 4.13 the sensitivity analysis for the separate case with Göteborg Energi is presented. Since the plant is smaller compared to the others, CAPEX has a greater influence on the total cost. Another factor is the operating hours as the CAPEX is distributed over less production hours for Göteborg Energi and thereby also a lower amount of CO_2 captured. This results in CAPEX having a larger influence on the total specific cost. The electricity price has a higher effect on the total cost compared to Renova because primary steam was used. In this case, the fuel price has an effect on the cost compared to Renova. Although, the impact is very small due to the low additional fuel use.

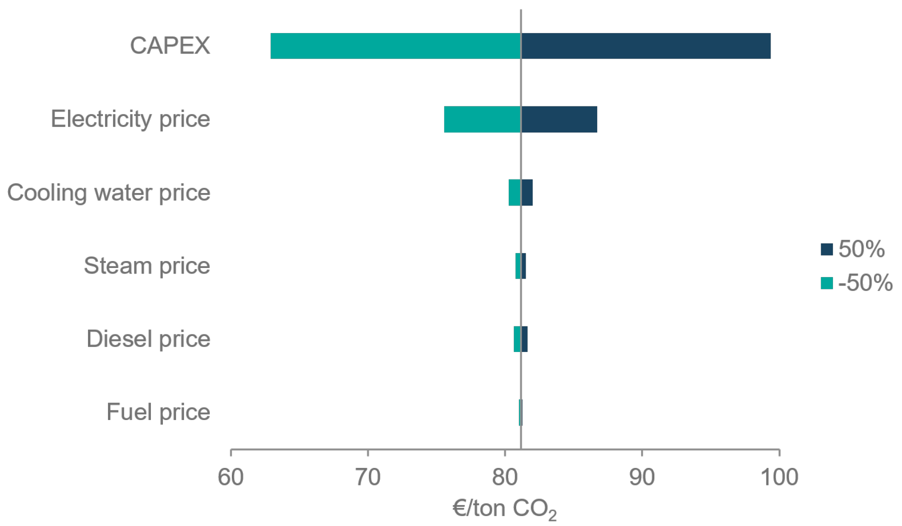

Figure 4.13: Sensitivity analysis for Göteborg Energi considering Scenario 3, collaboration to maintain district heat delivery.

The sensitivity analysis for the shared case, see Figure 4.14, has a similar result as the other plants. One difference is the slightly lower effect from the steam price since less district heat is lost. Further, the cost has a larger sensitivity to the electricity price due to the use of primary steam and thereby a greater reduction in electricity production compared to Renova, which used extracted steam.

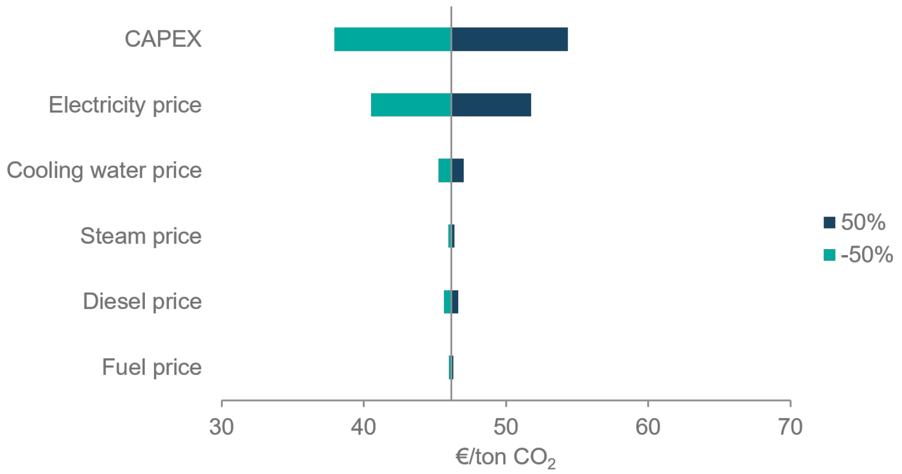

Figure 4.14: Sensitivity analysis for the shared case considering Scenario 3, collaboration to maintain district heat delivery.

Overall, the CAPEX and the electricity price are the most influential parameters to consider. CAPEX affects the total cost with about ± 20% when it is increased and decreased by 50%. When the electricity price is varied by ± 50% it affects the total cost around ± 10%. This indicates a sensitivity in the result and that uncertainties in the assumptions for estimation of CAPEX and electricity cost can lead to variations in the result.

4.4 Scenario 4: Reduced Capture Rate

The economic results for Scenario 4, reduced capture rate, are presented and discussed in this section. The results are presented as a total cost as well as separate costs for the capture plant, liquefaction, transport and on-site storage.

4.4.1 Economic Evaluation

Figure 4.15 presents the economic results for the last scenario with reduced capture rate and shows similar results as in the first scenario with maintained fuel usage. The larger variable OPEX for the shared case, once again originated from the use of primary steam. For Renova, utilisation of extracted steam results in a lower electricity loss and a lower variable OPEX. The total cost for the shared case is still slightly lower than the separate case, which again is mainly due to a lower liquefaction cost.

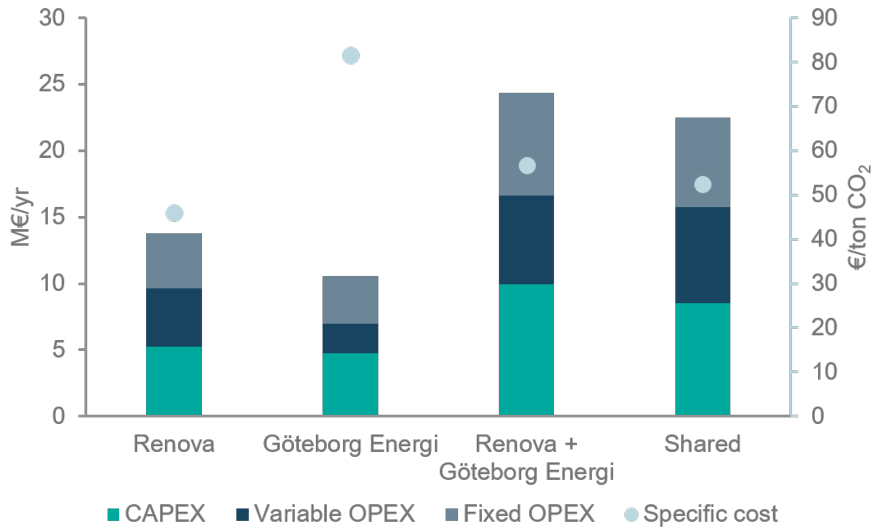

Figure 4.15: Total cost for Scenario 4 with reduced capture rate.

Figure 4.16 shows the total cost for the capture plant, liquefaction plant, transport and on-site storage divided into CAPEX, variable and fixed OPEX. For the cost savings, the result is similar to Scenario 2, where the capture plant has a possible cost saving although it is smaller compared to the saving from the liquefaction.

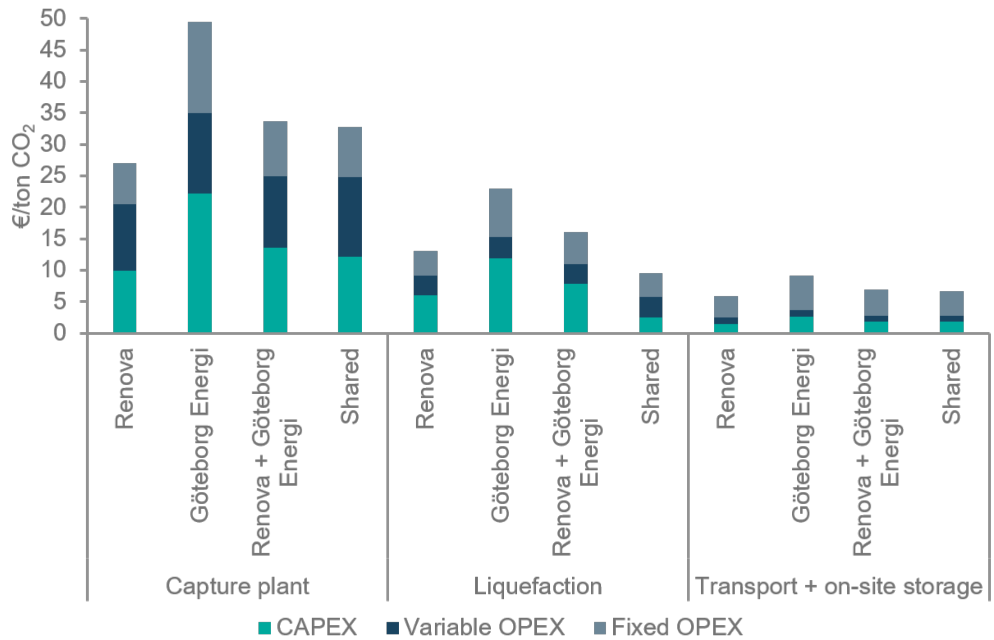

Figure 4.16: Specific cost for the capture plant, liquefaction, transport and on-site storage considering CAPEX, variable and fixed OPEX for Scenario 4 with reduced capture rate.

4.5 Comparison between Scenarios

Figure 4.17 shows a comparison between the results for the different scenarios. The shared case has the lowest total cost for all scenarios compared to the sum of the separate plants. The largest difference between the shared case and the sum is for Scenario 2, maintained district heat delivery, since Renova had to use primary steam. In the other scenarios, extracted steam was used for Renova and therefore the differences in the total cost are lower. This indicates that it is more beneficial to have a shared process if the same type of steam is used leading to a similar electricity reduction. Therefore, the available utilities are of importance.

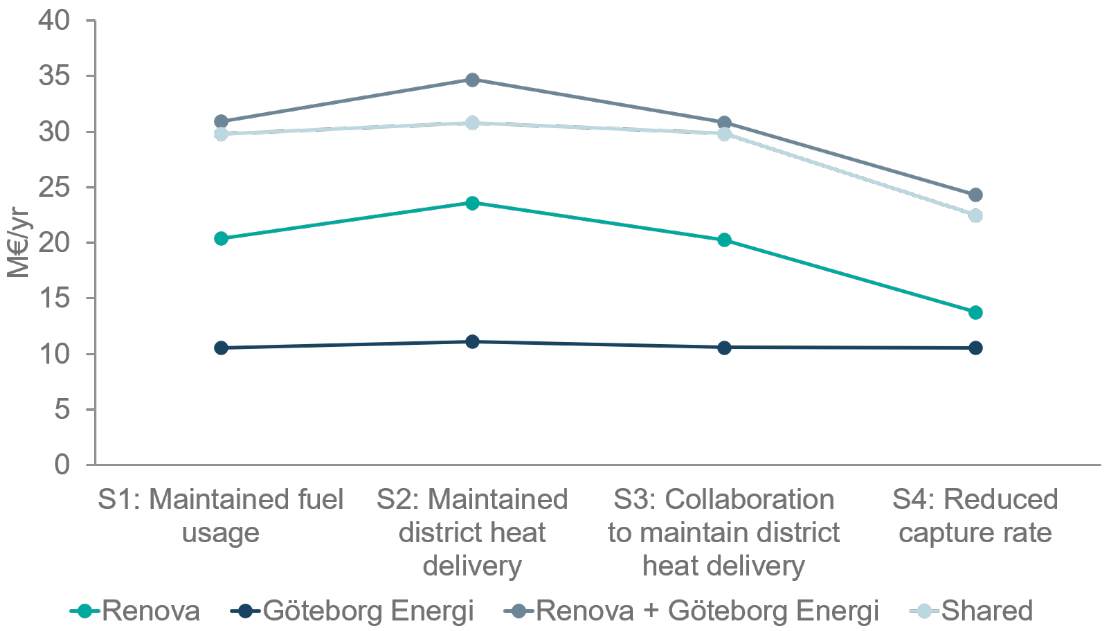

Figure 4.17: Total annual cost for the different cases and scenarios. S: Scenario.

Figure 4.18 shows that the specific cost for the shared plant is closer to Renova's cost for the maintained district heat delivery. The cost varies around 40-80 €/ton CO_2, which can be compared to the large cost range in the literature between 50-130 €/ton CO_2. The resulting cost for this case study can be considered as relatively low, especially for Renova, and one reason can be the low steam price due to the CHP plants being heat producers. Another reason can be the exclusion of a DCC and the large share of recovered district heat.

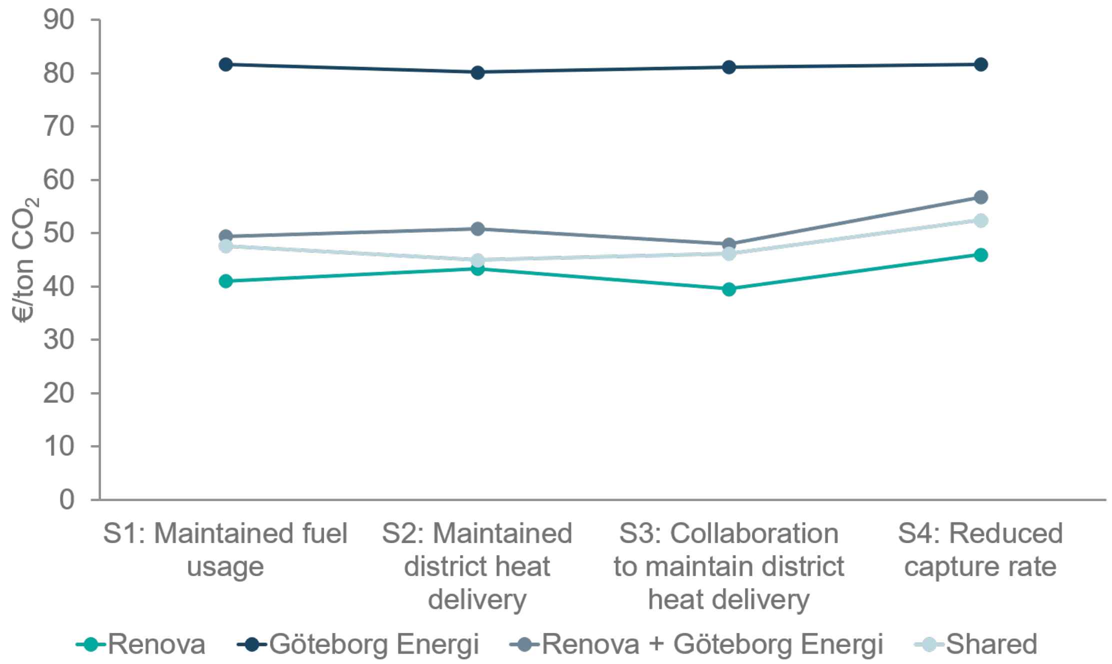

Figure 4.18: Total specific cost for the different cases and scenarios. S: Scenario.

For Scenario 4, the specific cost is slightly higher due to the lower CO_2 flow. The cost difference between the separate cases and the shared is slightly larger compared to Scenario 1. However, the specific costs are quite similar despite different heat productions. The cost is seen to be more dependent on the operating hours compared to the amount of heat produced. Renova has more operating hours in both scenarios, which leads to a low specific cost. The opposite applies for Göteborg Energi, which has a lower operating time and a higher specific cost. If the operating hours for Göteborg Energi would be equal to Renova, the annual cost would increase with 2% and the specific cost would decrease with 30%. Therefore, the operating hours have a large influence on the cost and are also contributing to the lower specific cost for Renova even when the heat production is lowered. With higher operating hours, the usage of the capture plant increases which leads to a higher OPEX and captured amount of CO_2 resulting in a more distributed cost. The capture plant at Göteborg Energi is dimensioned for the maximum flue gas flow from the CHP plant, resulting in a large CAPEX. However, the plant is rarely using its full capacity, consequently the capture cost is distributed over a lower amount of captured CO_2 compared to the dimensioned level. This results in a higher specific cost.

As the specific cost for the shared case in Scenario 4 is smaller compared to the separate case, it can be beneficial to collaborate even for two smaller sites or when partial capture is used. This can create a flexibility as the whole plant does not have to be integrated in the capture system at once.

Moreover, when the heat production is increased to cover for the lost district heat, the specific cost once again does not show large differences. Scenario 2 with maintained district heat delivery has the largest fuel increase and therefore a slightly larger cost. Since the fuel is not considered to have a cost for Renova, this contributes to the low cost despite the higher fuel usage. There are also possibilities for an additional income for Renova due to a larger waste incineration which has not been considered.

4.5.1 Business Cases

In Figure 4.19, the resulting cost reductions from the allocated shared costs divided by the separate costs are shown for different allocation methods. Both Scenario 1, maintained fuel usage, and Scenario 3, collaboration to maintain district heat delivery, are shown and larger savings are possible for the first scenario. Overall, the savings are in an interval of 0.2-1.0 M€/yr and between 2-5%. This indicates that the shared case can be cost beneficial even though the relative saving would not be large.

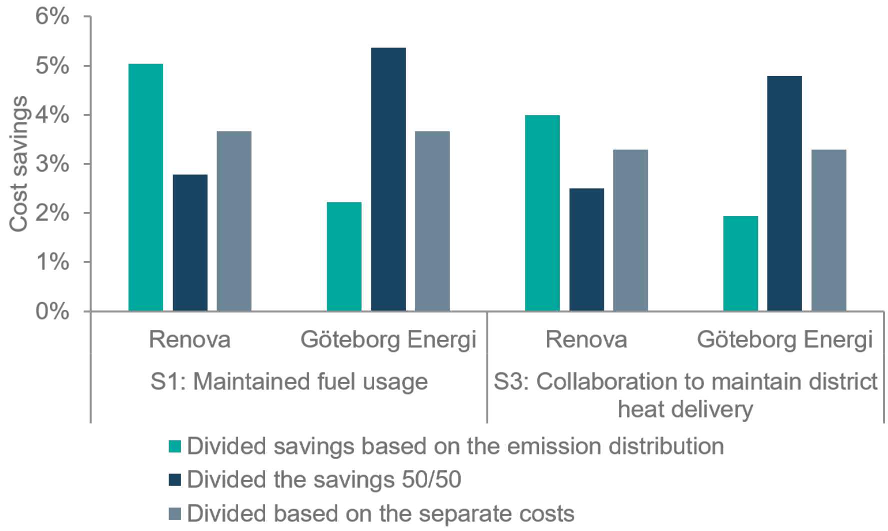

Figure 4.19: Reduction in total cost for Renova and Göteborg Energi after dividing the shared cost using different allocation methods for Scenario 1, maintained fuel usage, and Scenario 3, collaboration to maintain district heat delivery. S stands for Scenario.

For the shared plant, large cooperation and decisions between the companies would be required in order to divide the costs. The allocation method affects the possible savings for each company, as seen in Figure 4.19. For Scenario 3, the importance of mutual decisions and cost allocation increases due to the additional collaboration to maintain district heat. This requires more work and can be a challenge. On the other hand, a collaboration can decrease the economic risk since two companies share the investment and the use of the plant.

4.6 Discussion of collaboration alternatives, operating possibilities and importance of policies

There are potential cost savings with a shared infrastructure. For instance the CAPEX for the shared case in Scenario 1, with maintained fuel usage, has around 16 million € in absolute savings compared to separate capture plants. However, when taking OPEX into account the largest saving comes from a shared liquefaction. This can be compared to the result from the CinfraCap report [27], which included a study of different liquefaction infrastructures, where a shared liquefaction plant had a lower cost compared to separate liquefaction plants. For the OPEX of the capture plant, the steam utility plays a significant role in the possible savings when comparing the separate and shared case. Another alternative for cooperation between the companies are separate capture plants with shared liquefaction and transport.

Since both Renova and Göteborg Energi use the small river Säveån as a cooling water resource, their cooling capacities are limited and the cooling demand in the capture plants can be hard to fulfil. It is therefore important to minimise the cooling demand and one option is to cool against the district heating network. This does not only reduce the cooling water demand but also the lost district heat from the CHP plant. In this study it has been shown that the total costs can be reduced by 15-20% if the capture plant is integrated with the district heating system. District heat integration also leads to a smaller additional fuel usage in Scenarios 2 and 3 to fulfil the lost district heat. Further integration has not been taken into account in this study, one example is the amine cooler before the absorber, see Figure 2.1 in Section 2.1.1.1. The cooler decreases the lean solvent temperature to 40 °C and in this study a cold water flow with temperatures between 15-20 °C has been used. In Appendix B, the temperatures and pressures for the stages in the capture plant can be seen. There can be possibilities to cool the lean solvent against the district heating network as a preheater before the district heat water goes to further heating. To be sure further investigations must be performed.

The cooling tower at Renova and the option to change the power-to-heat ratio is not considered in this work. In the cooling tower, heat not used for district heating is cooled, mainly during the summer when the district heating demand is low. The cooling tower might not be equally necessary when having a capture plant, due to the lower district heat production if the fuel usage is not increased.

The power-to-heat ratio is assumed constant in this work. If either electricity or district heat is seen to be more valuable, the ratio of the production between them can be changed, which affects the revenue. If no loss of district heat is wanted, an alternative can be to reduce the electricity production. The operation of the CHP plants in combination with the capture plant has to be further investigated to find an optimal operation strategy.

The current emissions from Renova's waste-to-energy plant have to be covered by emission allowances and with the current price, the cost of the capture plant is higher than the potential reduction in allowances. However, the price of the emission allowances had a large increase between 2019 and 2020 and a further increase is a condition for the capture plant at Renova to be beneficial. The policies concerning the capture of biogenic emissions are necessary for the capture plant to be an economic alternative for Göteborg Energi. If reversed auctions would be implemented this can present an income instead of a cost for the capture plant. These types of incentives are necessary for the realisation of a capture plant at Göteborg Energi. Since there are current policies for capture of fossil emissions it can be beneficial for Renova to build a capture plant earlier than Göteborg Energi. The shared case can be constructed in a way where Renova invests in a larger desorber in order for Göteborg Energi to join when policies are implemented.

4.7 Future Work

Future work can include further investigations with the possible benefits of having separate capture plants but a shared liquefaction. In this thesis, the required transport from separate desorbers to a liquefaction plant is not considered. In order to conclude if only a shared liquefaction would be more beneficial compared to the current shared case, the additional transport would have to be regarded.

Another aspect for further investigation is the land requirement for the equipment, which can be lower for a shared plant and can present further savings. As seen in the results, the diameters for the two separate absorbers result in a lower area compared to the shared absorber, however the equipment can require less total area for the shared case if the space between the equipment is taken into account. The land requirement can be especially important due to restrictions in the available space since the plants would be located in the outer parts of Gothenburg. Therefore, it can be important to consider both the availability and the cost of the land.

Future work can also include development of a more detailed CHP model for each company in order to reduce the uncertainty in the electricity reduction. The values used in this thesis are for a more general CHP model and are estimates in order to account for the electricity loss.

Other cost aspects which can be investigated further are related to the washer since it in this study only is seen as an additional CAPEX. However, the water stream, cooler and pumps for the washer have not been considered. The additional equipment and streams required for the washer are not seen to affect the comparison of the result since the washer is still separate in both cases. Other extra equipment, which has not been taken into account, include the extra cleaning or drying of the CO_2 gas prior to the liquefaction. The choice of cleaning stage is dependent on the impurities of the CO_2 gas stream and to determine what kind of equipment is necessary, further studies of the impurities must be performed. The cleaning and drying equipment will result in a higher total cost. This can lead to further cost savings with a shared liquefaction plant, because of the need for one cleaning facility instead of two.

Safety and health aspects with the solvent, ammonia refrigerant and transport of CO_2 should be considered for the implementation of the capture plant including the required permissions for the usage and handling of these chemicals. Further investigations can be performed regarding if more suitable solvent and refrigerant alternatives exist.

Lastly, an important aspect for further studies is that the results from this thesis is for a case study. Therefore, additional case studies with other operating times and emissions will be necessary to see if other cases can have more or less economic benefits. For example, studying if two plants with similar operating hours can have more economic benefits from a shared plant.

5 Conclusion

The thesis is a techno-economic investigation of the local infrastructure needed for implementation of carbon capture and storage and the importance of sharing such infrastructure.

The result shows economic benefits with a shared infrastructure, although small. The liquefaction presented the most potential. The total cost saving of sharing infrastructure may be 2-5% of the total annual cost or 0.2-1.0 M€/yr depending on the cost allocation. Separate capture plants with a shared liquefaction could be an economic alternative and is recommended for further investigation.

Furthermore, efficient heat integration between the steam cycle, the district heating, and the capture plant is shown important to reduce the operating cost and minimise the need for external cooling. Around 70% of the reboiler heat demand could be recovered as district heat resulting in 15-20% lower total cost. The available steam utility has a large influence on the cost, where extracted steam is more beneficial compared to primary steam because of the lower loss in electricity production, district heat recovery and the relative cost of steam compared to electricity.

The size of the CHP plant does not show a large influence on the cost due to the low cost difference for Renova between scenario 1, maintained fuel usage and scenario 4, reduced capture rate. Plant utilisation is also an important parameter for the total cost through the higher specific cost for Göteborg Energi of 80-82 €/ton CO_2 due to lower operating hours compared to Renova with 39-46 €/ton CO_2. The shared case had a specific cost between 45-52 €/ton CO_2. Future work can include additional case studies of CHP plants with different sizes and plant utilisation in order to evaluate the economic benefits with a shared infrastructure further.

Bibliography

1. IPCC-The Intergovernmental Panel on Climate Change. Chapter Climate Change 2014 Synthesis Report Summary for Policymakers Summary for Policymakers. Tech. rep. 2014. Available from: https://www.ipcc.ch/site/assets/uploads/2018/02/AR5_SYR_FINAL_SPM.pdf
2. UNFCCC. ADOPTION OF THE PARIS AGREEMENT - Paris Agreement. Tech. rep. 2015. Available from: https://unfccc.int/sites/default/files/english_paris_agreement.pdf
3. Ministry of the environment and Government offices of Sweden. Sweden's long-term strategy for reducing greenhouse gas emissions. Tech. rep. 2020. Available from: https://unfccc.int/sites/default/files/resource/LTS1_Sweden.pdf
4. European Commission and European Union. Carbon Capture and Geological Storage | Climate Action. Available from: https://ec.europa.eu/clima/policies/innovation-fund/ccs_en
5. Meyer L and Rubin E. Summary for Policymakers A Special Report of Working Group III of the Intergovernmental Panel on Climate Change IPCC Special Report Carbon Dioxide Capture and Storage Summary for Policymakers. Tech. rep. 2005
6. Gibbins J and Chalmers H. Carbon capture and storage. Energy Policy 2008; 36:4317–22. DOI: 10.1016/j.enpol.2008.09.058. Available from: https://www.sciencedirect.com/science/article/pii/S0301421508004436
7. Luis P. Use of monoethanolamine (MEA) for CO2 capture in a global scenario: Consequences and alternatives. Desalination 2016 Feb; 380:93–9. DOI: 10.1016/j.desal.2015.08.004
8. Jackson S and Brodal E. Optimization of the Energy Consumption of a Carbon Capture and Sequestration Related Carbon Dioxide Compression Processes. Energies 2019 Apr; 12:1603. DOI: 10.3390/en12091603. Available from: https://www.mdpi.com/1996-1073/12/9/1603
9. Hester RE and Harrison RM. 1.2.4.2 Combined Heat and Power. *Energy Storage Options and Their Environmental Impact*. Royal Society of Chemistry, 2019. Available from: https://app.knovel.com/hotlink/khtml/id:kt011TWU51/energy-storage-options/combined-heat-power
10. Naturvårdsverket. Utsläpp av växthusgaser från el och fjärrvärme. 2020 Dec. Available from: http://www.naturvardsverket.se/Sa-mar-miljon/Statistik-A-O/Vaxthusgaser-utslapp-fran-el-och-fjarrvarme/

11. Naturvårdsverket. Bränsleanvändning för el- och fjärrvärmeproduktion. 2019 Sep. Available from: https://www.naturvardsverket.se/Sa-mar-miljon/Statistik-A-O/Bransleanvandning-for-el--och-fjarrvarmeproduktion/
12. Blomqvist EW and Jones F. Bestämning av andel fossilt kol i avfall som förbränns i Sverige. Tech. rep. Malmö: Avfall Sverige, 2012 Jan
13. Biermann M, Wolf J, and Mathisen A. Reducing the Cost of Carbon Capture in Process Industry Final report II. Tech. rep. Gothenburg: Chalmers University of Technology, 2019
14. Global CCS institute. The Global Status of CCS. Special Report: Understanding Industrial CCS Hubs and Clusters. Tech. rep. Melbourne: Global CCS institute, 2016 Jun
15. Intergovernmental Panel on Climate Change. Carbon Dioxide Capture and Storage. Tech. rep. New York: Cambridge University Press, 2005. Available from: https://www.ipcc.ch/site/assets/uploads/2018/03/srccs_wholereport.pdf
16. Wang Y, Zhao L, Otto A, Robinius M, and Stolten D. A Review of Post-combustion CO2 Capture Technologies from Coal-fired Power Plants. *Energy Procedia*. Vol. 114. Elsevier Ltd, 2017 :650–65. DOI: 10.1016/j.egypro.2017.03.1209
17. Garðarsdóttir SÓ, Normann F, Andersson K, and Johnsson F. Postcombustion CO2 capture using monoethanolamine and ammonia solvents: The influence of CO2 concentration on technical performance. Industrial and Engineering Chemistry Research 2015 Jan; 54:681–90. DOI: 10.1021/ie503852m
18. Martinez Castilla G, Biermann M, Montañés RM, Normann F, and Johnsson F. Integrating carbon capture into an industrial combined-heat-and-power plant: performance with hourly and seasonal load changes. International Journal of Greenhouse Gas Control 2019 Mar; 82:192–203. DOI: 10.1016/j.ijggc.2019.01.015
19. Andersson J. An investigation of carbon capture technologies for Sävenäs waste-to-energy plant. PhD thesis. Luleå: Luleå University of Technology, 2020
20. Leung DY, Caramanna G, and Maroto-Valer MM. An overview of current status of carbon dioxide capture and storage technologies. 2014. DOI: 10.1016/j.rser.2014.07.093
21. Ustadi I, Mezher T, and Abu-Zahra MR. Potential for Hybrid-Cooling System for the CO2 Post-Combustion Capture Technology. *Energy Procedia*. Vol. 114. Elsevier Ltd, 2017 Jul :6348–57. DOI: 10.1016/j.egypro.2017.03.1771
22. Scherffius JR, Reddy S, Klumpyan JP, and Armpriester A. Large-Scale CO 2 Econamine FG Plus SM Electric Generating Station Selection and/or peer-review under responsibility of GHGT. Energy Procedia 2013; 37:6553–61. DOI: 10.1016/j.egypro.2013.06.587. Available from: www.sciencedirect.com
23. Adams DMB. Flue gas treatment for CO2 capture. Tech. rep. IEA Clean Coal Centre, 2010 Jun :61
24. Rao AB and Rubin ES. A technical, economic, and environmental assessment of amine-based CO2 capture technology for power plant greenhouse gas control. Environmental Science and Technology 2002 Oct; 36:4467–75. DOI: 10.1021/es0158861

25. Moser P, Schmidt S, Sieder G, Garcia H, Ciattaglia I, and Klein H. Enabling post combustion capture optimization-The pilot plant project at Niederaussem. *Energy Procedia*. Vol. 1. 1. 2009 Feb :807–14. DOI: 10.1016/j.egypro.2009.01.107
26. Zahid U, An J, Lee U, Choi SP, and Han C. Techno-economic assessment of CO2 liquefaction for ship transportation. Greenhouse Gases: Science and Technology 2014 Dec; 4:734–49. DOI: 10.1002/ghg.1439
27. COWI. CinfraCap-Gemensam infrastruktur för transport av koldioxid. Tech. rep. Göteborg: COWI, 2021 Mar
28. Deng H, Roussanaly S, and Skaugen G. Techno-economic analyses of CO2 liquefaction: Impact of product pressure and impurities. International Journal of Refrigeration 2019 Jul; 103:301–15. DOI: 10.1016/j.ijrefrig.2019.04.011
29. Klemetsrudanlegget and Project CCS Carbon Capture Oslo. Concept Study Report. Tech. rep. Fortum Oslo Varme, 2017
30. Johnsson F and Kjärstad J. Avskiljning, transport och lagring av koldioxid i Sverige. Behov av forskning och demonstration. Tech. rep. Gothenburg: Department of Space, Earth and Environment, Chalmers University of Technology, 2019. Available from: https://data.geus.dk/nordiccs/map.xhtml
31. Göteborgs hamn. CinfraCap - flytande koldioxids väg till hamnen. Available from: https://www.goteborgshamn.se/hamnens-projekt/cinfracap/
32. Northern Lights project team. Northern-Lights-FEED-report-public-version. 2020
33. Gassnova and Gassco. Feasibility study for full-scale CCS in Norway Contents. Tech. rep. Ministry of Petroleum and Energy, 2016
34. ZEP. Identifying and Developing European CCS Hubs. Tech. rep. Zero emissions platform, 2016 Apr. Available from: http://www.sccs.org.uk/images/expertise/reports/catalysing/downloads/SCCSConference2014Report.pdf
35. i24c and Element Energy. Deployment of an industrial Carbon Capture and Storage cluster in Europe: A funding pathway. Tech. rep. Industrial Innovation for Competitiveness and Element Energy Limited, 2017 Aug. Available from: www.i24c.eu
36. Bjørnsen D, Kjærstad J, Langlet D, Mathisen A, Aagaard P, and Anundskås A. Carbon Capture and Storage in the Skagerrak/Kattegat region Final report. Tech. rep. 2012 Feb. Available from: http://www.ccs-skagerrakkattegat.eu/.
37. European Technology Platform for Zero Emission Fossil Fuel Power Plants. The Costs of CO2 Transport Post-demonstration CCS in the EU. Tech. rep. 2011
38. Ali H, Eldrup NH, Normann F, Skagestad R, and Øi LE. Cost Estimation of CO2 Absorption Plants for CO2 Mitigation – Method and Assumptions. International Journal of Greenhouse Gas Control 2019 Sep; 88:10–23. DOI: 10.1016/j.ijggc.2019.05.028
39. Garðarsdóttir SÓ, Normann F, Skagestad R, and Johnsson F. Investment costs and CO2 reduction potential of carbon capture from industrial plants – A

Swedish case study. International Journal of Greenhouse Gas Control 2018 Sep; 76:111–24. DOI: 10.1016/j.ijggc.2018.06.022
40. Mortensen GM, Erlström M, Nordström S, and Nyberg J. Geologisk lagring av koldioxid i Sverige. Lägesbeskrivning avseende förutsättningar, lagstiftning och forskning samt olje-och gasverksamhet i Östersjöregionen. Tech. rep. Sveriges geologiska undersökning, 2017
41. Global CCS Institute. Global Storage Portfolio. Tech. rep. Global Carbon Capture and Storage Institute, 2016 Mar
42. Anthonsen KL, Aagaard P, Bergmo PES, Erlström M, Fareide JI, Gislason SR, Mortensen GM, and Snæbjörnsdottir SÓ. CO2 Storage Potential in the Nordic Region. Energy Procedia 2013; 37:5080–92. DOI: https://doi.org/10.1016/j.egypro.2013.06.421. Available from: http://www.sciencedirect.com/science/article/pii/S1876610213006644
43. Norwegian Ministry of Petroleum and Energy. Longship-Carbon capture and storage. Tech. rep. Norwegian Ministry of Petroleum and Energy, 2020
44. Northern Lights. Northern Lights Project description. Available from: https://northernlightsccs.com/en/about
45. Darrow K, Tidball R, Wang J, and Hampson A. Catalog of CHP Technologies. Tech. rep. U.S. Environmental Protection Agency. Combined Heat and Power Partnership., 2017 Sep
46. Selivanovs J, Vigants E, Priedniece V, Veidenbergs I, and Blumberga D. Flue gas treatment multi-criteria analysis. *Energy Procedia*. Vol. 128. Elsevier Ltd, 2017 :379–85. DOI: 10.1016/j.egypro.2017.09.056
47. Göteborg Energi. Miljörapport 2019 Sävenäsverket. Tech. rep. Göteborg: Göteborg Energi, 2020
48. Renova AB. Från avfall till ren energi -Avfallskraftvärmeverket i Sävenäs. Tech. rep. Göteborg
49. Renova AB. SAMRÅDSUNDERLAG Sävenäs avfallsanläggning. Tech. rep. Göteborg, 2017. Available from: https://www.renova.se/globalassets/om-renova/samradsunderlag-170307.pdf
50. Förbränningsskatteutredningen. Brännheta skatter! Bör avfallsförbränning och utsläpp av kväveoxider från energiproduktion beskattas? Tech. rep. Stockholm: Statens Offentliga Utredningar, 2017 :545
51. Renova. Miljörapport 2019. Tech. rep. 2020
52. Göteborg Energi. Vår historia - 150 år av energi. Available from: https://www.goteborgenergi.se/om-oss/vilka-vi-ar/historia
53. Göteborg Energi. Vårt energisystem. Available from: https://www.goteborgenergi.se/om-oss/vad-vi-gor/vart-energisystem
54. Göteborg Energi. Så funkar svenska elnätet. Available from: https://www.goteborgenergi.se/om-oss/vad-vi-gor/vart-energisystem/sa-funkar-elnatet
55. Göteborg Energi. Mer Biobränsle i Sävenäsverket. Available from: https://www.goteborgenergi.se/hallbaraihop/12/Mer-biobransle-i-Savenasverket

56. Zetterberg L, Burtraw D, Stensson DE, Paulie C, and Roth S. Europe's choice-Facts and function of the EU emissions trading system. Tech. rep. Gothenburg: Mistra Indigo, 2014. Available from: http://indigo.ivl.se
57. Nilsson M. EUs utsläppshandel 2020. Tech. rep. Stockholm: Landsorganisationen i Sverige, 2020 Dec
58. Naturvårdsverket. Verksamheter som omfattas. 2020 Nov. Available from: http://www.naturvardsverket.se/Stod-i-miljoarbetet/Vagledningar/Utslappshandel---vagledningar/For-anlaggningar-krav-pa-tillstand-och-rapportering/Verksamheter-som-omfattas/
59. Miljödepartementet and Infrastrukturdepartementet. Sverige godkänner regler som möjliggör export av koldioxid avsedd för lagring under havsbotten. 2020 Jun. Available from: https://www.regeringen.se/pressmeddelanden/2020/06/sverige-godkanner-regler-som-mojliggor-export-av-koldioxid-avsedd-for-lagring-under-havsbotten/
60. Klimatpolitiska vägvalsutredningen. Vägen till en klimatpositiv framtid. Betänkande från Klimatpolitiska vägvalsutredningen. Tech. rep. Stockholm: Statens Offentliga Utredningar, 2020
61. Pousette A. Energimyndigheten föreslår omvända auktioner som stöd för bio-CCS. 2021 Apr. Available from: http://www.energimyndigheten.se/nyhetsarkiv/2021/energimyndigheten-foreslar-omvanda-auktioner-som-stod-for-bio-ccs/
62. European Commission. Innovation Fund. Available from: https://ec.europa.eu/clima/policies/innovation-fund_en
63. Eliasson Å and Fahrman E. Utilization of Industrial Excess Heat for CO 2 Capture Effects on Capture Process Design and District Heating Supply. PhD thesis. Gothenburg, Sweden: CHALMERS UNIVERSITY OF TECHNOLOGY, 2020
64. Kohl AL and Nielsen RB. Mechanical Design and Operation of Alkanolamine Plants. *Gas Purification*. 5th ed. Houston: Gulf Publishing Company, 1997. Chap. 3:187–277
65. Hjelmaas S, Storheim E, Flø NE, Thorjussen ES, Morken AK, Faramarzi L, De Cazenove T, and Hamborg ES. Results from MEA Amine Plant Corrosion Processes at the CO2 Technology Centre Mongstad. *Energy Procedia*. Vol. 114. Elsevier Ltd, 2017 :1166–78. DOI: 10.1016/j.egypro.2017.03.1280
66. Swamee PK, Aggarwal N, and Aggarwal V. Optimum design of double pipe heat exchanger. International Journal of Heat and Mass Transfer 2008 May; 51:2260–6. DOI: 10.1016/j.ijheatmasstransfer.2007.10.028
67. Nakayama Y. Flow in Pipes. *Introduction to Fluid Mechanics*. 2nd ed. Elsevier, 2018 Jan. Chap. 7:135–61. DOI: 10.1016/b978-0-08-102437-9.00007-3. Available from: https://app.knovel.com/web/view/khtml/show.v/rcid:kpIFME0012/cid:kt011PH9R2/viewerType:khtml//root_slug:7-flow-in-pipes/url_slug:flow-in-pipes?b-q=Introduction%20to%20Fluid%20Mechanics&sort_on=default&b-group-by=true&b-sort-on=default&b-content-type=all_references&include_synonyms=no&b-toc-cid=kpIFME0012&b-toc-root-slug=&b-toc-url-slug=flow-in-

pipes&b-toc-title=Introduction%20to%20Fluid%20Mechanics%20(2nd%20Edition)&page=6&view=collapsed&zoom=1

68. Welty JR, Rorrer GL, and Foster DG. Fundamentals of Mumentum, Heat and Mass Transfer. 6th ed. New York: John Wiley &Sons Singapore Pte. Ltd, 2015 :209–undefined
69. Biron M. Thermoplastics. *Material Selection for Thermoplastic Parts - Practical and Advanced Information for Plastics Engineers.* Waltham: Elsevier, 2016. Chap. 2.3.1.1. Available from: https://app.knovel.com/hotlink/pdf/id:kt01102GKF/material-selection-thermoplastic/thermoplastics
70. Bejan A. Metallic Solids. *Convection Heat Transfer.* 4th ed. Hoboken: John Wiley & Sons, 2013. Chap. B.2. Available from: https://app.knovel.com/hotlink/pdf/id:kt011B72A1/convection-heat-transfer/properties-metallic-solids
71. Storm K. Standard and Line Pipe - Wall Thickness. *Industrial Process Plant Construction Estimating and Man-Hour Analysis.* Elsevier, 2019. Chap. 2.11. Available from: https://app.knovel.com/hotlink/pdf/id:kt0122EAUA/industrial-process-plant/standard-line-pipe-wall
72. Øi LE, Eldrup N, Adhikari U, Bentsen MH, Badalge JL, and Yang S. Simulation and cost comparison of CO2 liquefaction. *Energy Procedia.* Vol. 86. Elsevier Ltd, 2016 Jan :500–10. DOI: 10.1016/j.egypro.2016.01.051
73. Google. Google maps directions for driving from Renova, Sävenäs, to Energihamnen, Gothenburg. Available from: https://www.google.com/maps/dir/Renova+S%C3%A4ven%C3%A4s,+G%C3%B6teborg/Energihamnen+G%C3%B6teborg,+Bentylgatan,+G%C3%B6teborg/@57.7086279,11.8957898,12z/data=!4m19!4m18!1m10!1m1!1s0x464ff43db4e12f99:0x9ae7c87175fbb363!2m2!1d12.0515318!2d57.7322649!3m4!1m2!1d11.9380601!2d57.6459471!3s0x464ff293716d9e4b:0xfc3e3a9178992060!1m5!1m1!1s0x464f8cea91fb2a39:0x9885e72855b63bdc!2m2!1d11.8801262!2d57.6978066!3e0
74. Myndigheten för samhällsskydd och beredskaps föreskrifter om transport av farligt gods på väg och i terräng. *Anna Asp, Myndigheten för samhällsskydd och beredskaps författningssamling.* 2021. Available from: www.nj.se/offentligapublikationer
75. Trafikverket. NVDB på webb. Available from: https://nvdb2012.trafikverket.se/SeTransportnatverket
76. Mörtstedt SE and Hellsten G. Data och Diagram. Energi- och kemitekniska tabeller. 7th ed. Stockholm: Liber AB, 1999
77. Eriksson M. Dynamisk simulering av avfallskraftvärmeverk. PhD thesis. Göteborg: Institutionen för reglerteknik, 2008 Sep
78. Pröll T and Zerobin F. Biomass-based negative emission technology options with combined heat and power generation. Mitigation and Adaptation Strategies for Global Change 2019 Oct; 24:1307–24. DOI: 10.1007/s11027-019-9841-4
79. Beiron J, Normann F, and Johnsson F. A Case Study of the Potential for CCS in Swedish Combined Heat and Power Plants. Proceedings of the 15th Greenhouse Gas Control Technologies Conference 15-18. Tech. rep. 2021

80. Montañés RM, Flø NE, and Nord LO. Experimental results of transient testing at the amine plant at Technology Centre Mongstad: Open-loop responses and performance of decentralized control structures for load changes. International Journal of Greenhouse Gas Control 2018 Jun; 73:42–59. DOI: 10.1016/j.ijggc.2018.04.001
81. Spek M van der, Roussanaly S, and Rubin ES. Best practices and recent advances in CCS cost engineering and economic analysis. International Journal of Greenhouse Gas Control 2019 Apr; 83:91–104. DOI: 10.1016/j.ijggc.2019.02.006
82. Nordpool. Historical Market Data- elspot-prices_2019_hourly_eur. 2020. Available from: https://www.nordpoolgroup.com/historical-market-data/
83. Knudsen JN, Jensen JN, Vilhelmsen PJ, and Biede O. Experience with CO2 capture from coal flue gas in pilot-scale: Testing of different amine solvents. *Energy Procedia*. Vol. 1. 1. Elsevier, 2009 Feb :783–90. DOI: 10.1016/j.egypro.2009.01.104. Available from: https://www.sciencedirect.com/science/article/pii/S1876610209001052
84. Voldsund M, Gardarsdottir SO, De Lena E, Pérez-Calvo JF, Jamali A, Berstad D, Fu C, Romano M, Roussanaly S, Anantharaman R, Hoppe H, Sutter D, Mazzotti M, Gazzani M, Cinti G, and Jordal K. Comparison of technologies for CO 2 capture from cement production—Part 1: Technical evaluation. Energies 2019 Feb; 12. DOI: 10.3390/en12030559
85. Energimyndigheten. Läget på energimarknaderna - Biodrivmedel och fasta biobränslen. Tech. rep. Energimyndigheten, 2020 Jan
86. Forex Bank. EURO-EUR. Available from: https://www.forex.se/valuta/eur
87. Biermann M, Ali H, Sundqvist M, Larsson M, Normann F, and Johnsson F. Excess heat-driven carbon capture at an integrated steel mill – Considerations for capture cost optimization. International Journal of Greenhouse Gas Control 2019 Dec; 91. DOI: 10.1016/j.ijggc.2019.102833

A

Flowsheet for Economic Evaluation

The unit operations for the carbon capture plant which are included in the economic analysis, except filters as well as a pump for the cold utility, are seen in Figure A.1 for the case with separate plants. The flowsheet is created for the current process but based on the work by Biermann et al. [87].

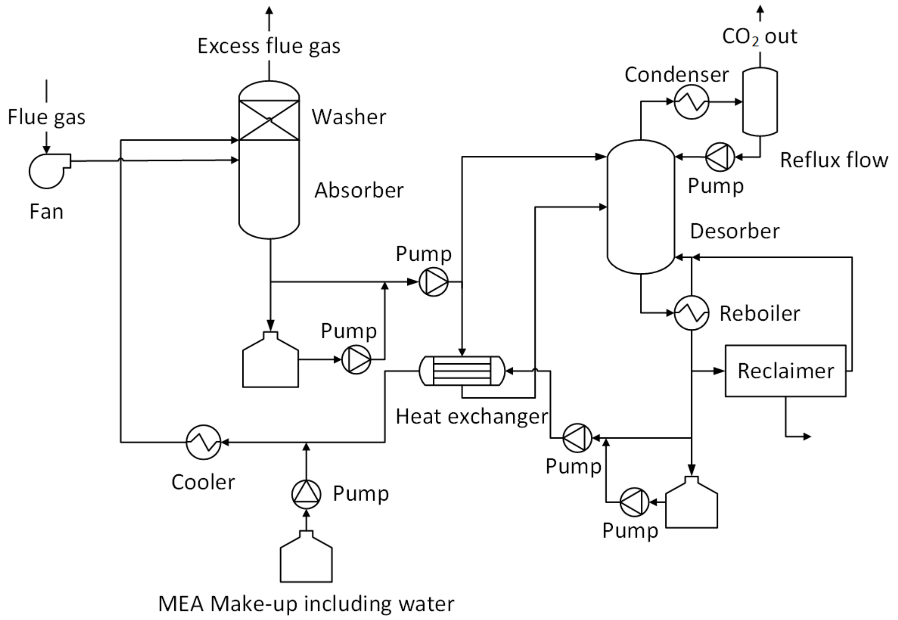

Figure A.1: The included units in the economic analysis shown in a flowsheet for a separate capture facility based on Biermann et al. [87].

In Figure A.2, the shared case with two absorbers for each company and one desorber is shown. The required pipeline and additional pumps are also visible, however only one make-up stream for MEA was assumed. A pump for the cold utility was also added for this case as well as filters similar to the separate case.

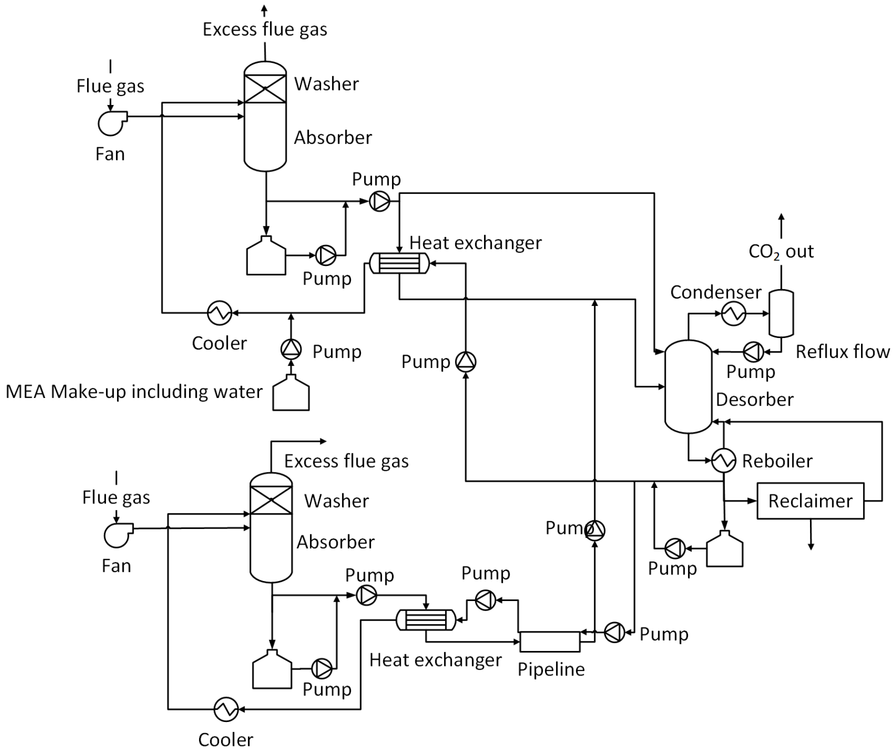

Figure A.2: The included units in the economic analysis shown in a flowsheet for a shared regeneration facility based on Biermann et al. [87].

B

Temperature and Pressures for the Capture Plant

In Figure B.1, the temperatures and pressures of the carbon capture plant can be seen. The possible heat integration is represented with an additional condenser to show the two temperature stages and a cooler for the condensed steam after the reboiler.

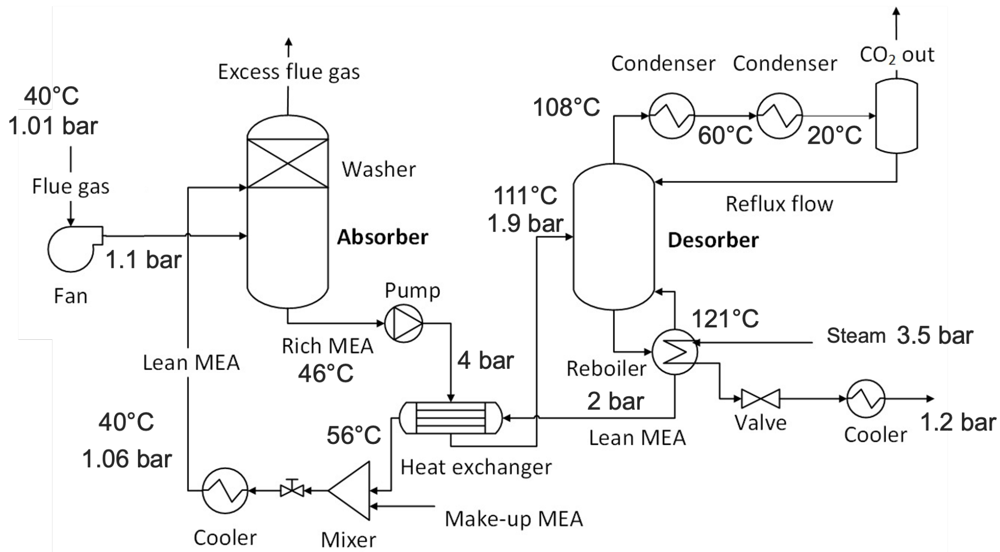

Figure B.1: Temperatures and pressures in the carbon capture process including the possible heat integration.

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