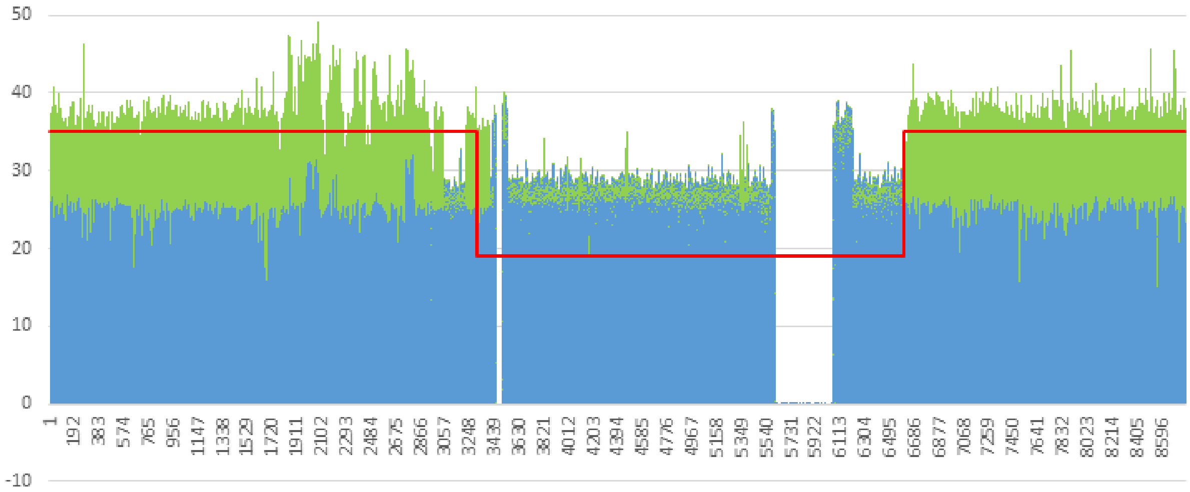

Design of Partial CO_2 Capture from Waste Fired CHP Plants

Master's thesis within the Sustainable Energy Systems programme

SIMON ÖBERG

Department of Energy and Environment
Division of Energy Technology
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2017

Master's thesis 2017

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Master's Thesis 2017
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Cover: The heat generation from steam and flue gas condensers at Lillesjöverket and the average summer and winter heat load in Uddevalla during 2016.

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Abstract

In the Paris Agreement in 2015, it was stated that the temperature increase should be kept well below 2 °C above pre-industrial levels. As a response to the Paris Agreement, the Swedish government in February 2017 presented a climate reform to make Sweden CO_2 neutral by 2045, and even CO_2 negative beyond 2045. This requires extensive actions. One of the most required technical solutions is carbon capture and storage (CCS). One way of implementing CCS in a cost-efficient manner is in form of partial CO_2 capture, where only a fraction of the emissions are captured.

An important aspect of the Swedish energy system is the extensive use of waste incineration in combined heat and power (CHP) plants. This work evaluates the possibilities of making the Swedish waste fired CHP plants CO_2 neutral by applying partial capture in order to capture the fossil share of the CO_2 emissions.

This work investigates the implementation of carbon capture through a post-combustion absorption process with monoethanolamine (MEA) as absorbent. Two design alternatives are evaluated for partial CO_2 capture, based on either a high absorption rate for a fraction of the flue gas flow, or a low absorption rate for 100 % of the flue gas flow. The two designs are evaluated for a generic CHP plant, based on the average size of a Swedish waste fired CHP plant. Also, a specific case of Lillesjöverket is analyzed. Since seasonal variations in the heat load occur, both a constant annual operation and a seasonally optimized operation are considered for the design of the partial capture unit.

The results show that the generic plant, which have fossil CO_2 emissions of 64 200 ton per year, have a specific cost of 94-98 EUR/ton CO_2 for a constant annual operation of the capture unit, but that the specific cost may be decreased to 86 EUR/ton CO_2 if a seasonal operation is applied. Due to the small size of the capture unit, the capital cost is dominating and stands for more than 80 % of the specific cost. The capture unit should, thus, be designed to minimize the capture cost.

The effect of implementing carbon capture on the other product streams of the CHP plant is significant, especially for Lillesjöverket that besides heat and electricity also produces pellets. In case waste incineration should be made CO2-neutral, the other products have to be prioritized based on price and season in order to maximize the revenue, and it can be concluded that the addition of a capture unit requires a system perspective analysis already in the planning and design phase.

Keywords: partial capture, post-combustion, waste incineration, seasonal operation.

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Simon Öberg, Gothenburg, June 2017

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1 Introduction

This thesis evaluates the possibilities and correlated costs for applying carbon dioxide absorption to a waste fired combined heat and power plant. A brief introduction to the current situation, both globally and locally in Sweden will be given in this chapter, along with the aim and scope of the thesis and restrictions of what is not included.

1.1 Background

Scientists have agreed for many years that climate change is real and that it is due to anthropogenic activities, yet the transition towards renewable and more sustainable energy sources are still falling behind. However, the signing by the majority of the worlds leaders of the Paris Agreement at COP21 (Conference Of the Parties) in December 2015 was an important step. The ambitious goals of the agreement states that the global temperature should be kept well below a 2 °C increase, compared to pre-industrial levels, and that efforts to limit the temperature rise to 1.5 °C should be pursued [1]. These ambitious goals requires large-scale actions and real-world implementations of mitigation technologies, which are available, to the existing system, but also investments in research and development of new technologies for future systems with other requirements.

In order to monitor and estimate the current situation, a carbon budget have been established for the 2 °C scenario (2DS). This carbon budget have been estimated to some 980 Gt of CO_2 for the energy sector, starting from 2014 [2], and with annual emissions expected to increase in this sector from the current 30 Gt to 35 Gt in 2030 [1], immediate measures are crucial. The two main measures to reduce emissions in the 2DS is energy efficiency measures and substitution of fossil energy sources to renewable alternatives [1]. The third most important measure is Carbon Capture and Storage (CCS). Several reports by the International Energy Agency (IEA) have stated the importance of CCS, and in the Technology Roadmap from 2013, CCS is estimated to contribute to one sixth of the reduction of CO_2 emissions in 2050. Between 2015 and 2050 CCS is also assumed to be responsible for 14 % of the cumulative emission reductions compared to a business-as-usual approach, which would yield an average global temperature rise of 6 °C [3].

With CCS established as a key-technology for reaching the ambitious goals of the Paris Agreement, and the fact that fossil fuels will continue to be a main energy source for decades to come, the decision for large scale implementation may seem obvious. This is however not the case. Implementation of CCS comes with large costs, partly for the investment, but also for the energy penalty that reduces the ability to produce and sell electricity, or other products that are an output of the process. In order to overcome this problem, the concept of partial CO_2 capture have been introduced. Partial capture implies that only a fraction of the CO_2 is captured, with the benefit of both lower investment costs and lower energy penalty. A lower investment cost and a more flexible way of implementing the capture process is two of the main reasons why partial capture have become a more and more interesting topic. Ideally, all CO_2 should be captured, but until either local or global legislation requires that, the industry will minimize their effort to current legislation in order to minimize the additional cost. Partial capture can thus facilitate the implementation of CCS to meet initial demand on reducing emissions, and simultaneously enable the technology to mature and develop to become a more competitive technology to mitigate climate change [4].

The concept of applying partial capture CCS on Swedish waste fired CHP plants became in February 2017 an even more interesting topic. The Swedish government then presented a suggestion to a new climate reform which stated that Sweden should be CO_2 neutral by 2045, and have net-negative emission levels beyond 2045. The fossil related emissions from waste incineration stands for 4 % of the total emissions in Sweden [12], and if both fossil and biogenic emissions are taken into account, the emissions account for 9.5 %. If the goal of the climate reform is to be met, these emissions have to be addressed.

The Swedish electricity production reached in 2012 an all time high of 162 TWh of produced electricity. The system is based on over 90 % CO_2 neutral technologies such as hydro-, wind- and nuclear power [9], and the emissions from electricity production is low. The remaining 10 % primarily comes from thermal combined heat and power (CHP) plants and industrial CHP facilities operated with various types of fuel [9]. Even though the thermal CHP plants only cover a minor part of the electricity production, they have an important role in producing heat to the district heating system. The district heating system delivered close to 60 TWh of heat in 2013, losses included, out of which thermal CHP plants produced 40 % [9]. As mentioned, the district heating system has a close connection to thermal CHP plants, and since these plants are responsible for approximately 22 % of all emissions that originate from a larger point-wise emission sources in Sweden [10], they are important to lower the national CO2 emissions and of interest for applying CCS.

In many municipalities that have invested in the CHP technology, waste is the main fuel type. This is due to regulations against putting combustible and organic waste in landfills, and the fact that utilities that uses waste as fuel are paid to do so since it provides a service to the municipalities and their waste handling [9]. The fuel is on average of about 60 % biomass and 40 % fossil origin [11]. Partial capture of CO_2 is, thus, interesting for this type of process. By applying partial capture corresponding to the fossil share, these plants can become CO_2 neutral, and if the capture process is expanded further, they will become CO_2 negative, so called Biomass Energy Carbon Capture and Storage (BECCS). By utilizing the concept of BECCS, emissions in other sectors that are difficult to capture, such as heavy transports, may be compensated for.

1.2 Aim and Scope

The aim of this thesis is to design and evaluate a partial CO_2 absorption process for a Swedish waste fired CHP plant. The work will minimize the monetary cost per mass-unit of CO_2 captured. The preliminary objective is to design a process that achieves CO_2 neutrality for the CHP plant, and then evaluate the potential for further improvements. The design is limited to the post-combustion capture process and is divided into two design paths:

- Separation rate path; 100 % of the flue gas flow is processed at a low capture rate.
- Slip stream path; a fraction of the flue gas flow is processed at 90 % capture rate.

The two design paths are optimized and evaluated for a generic case and a specific case - Lillesjöverket CHP plant. The generic case is based on an average sized Swedish waste fired CHP plant with no restrictions on energy supply for the capture process. Both cases have the objective to capture at least the fossil share of the CO_2 in the flue gas stream, but the specific case is, in contrary to the generic case, limited to the specific conditions related to Lillesjöverket. A more detailed description of the design optimization can be found in Chapter 4.

The thesis is limited to only evaluate the cost for capturing and compressing the CO_2, and thus excludes the cost for transportation and the issue with a suitable storage location close to the site. Practical issues at the site such as space availability and building permits are also excluded in this work, and emissions due to transportation of the waste and the residuary ashes are also excluded since they are unconnected to the combustion process.

2 Carbon Capture and Storage

CCS can be divided into three consecutive steps; capture and compression, transportation and long term storage. Only the capture and compression part is considered in this report. This chapter explains the fundamental aspects of CCS in general, and go in detail for some aspects that is of particular importance for this work.

2.1 CO_2 Capture Technologies

There are several technologies available for the purpose of capturing CO_2 from various types of streams, and among these are post-, pre- and oxy-fuel combustion the most common [5]. Each of these technologies have their specific benefits and drawbacks, and since this work considers partial capture of CO_2 from already existing plants, the post-combustion technique is selected. Post-combustion separates the CO_2 from the flue gas at the end of the tail pipe, and can thus comparably easy be retrofitted to any existing plant [5]. Below follows a more detailed description of post-combustion, partial capture, and description of the difference between CCS and BECCS.

2.1.1 Post-combustion

The post-combustion technology can utilize several different separation techniques, which is well described in the literature [5]. The separation technique used in this project is a chemical absorption process between a liquid solvent and the gaseous CO_2. The solvent and the CO_2 form a weakly bonded temporary compound before the process is reversed by the addition of heat. By letting the reversed processes occur in different containers, the CO_2 can be extracted from the other flue gases, which is shown in the simple schematic in Figure 2.1.

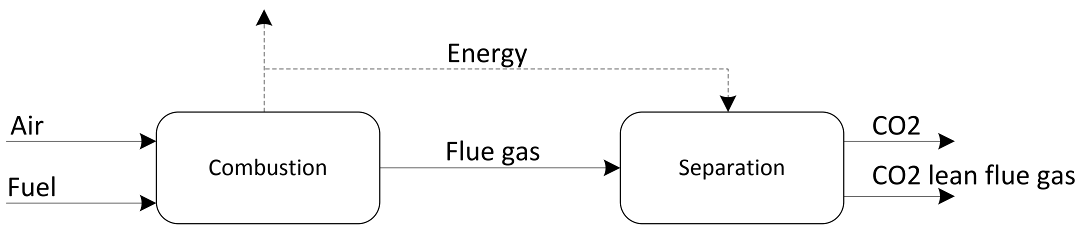

Figure 2.1: Schematic of the post-combustion absorption process.

There are many solvents to choose among, but the most commonly used is monoethanolamine (MEA), which is also considered a benchmark and is often used as reference solvent when new solvents are evaluated. Amine solvents like MEA are sensitive to impurities such as excess oxygen, nitrous oxides and sulphur dioxide, which forms heat stable salts and other compounds that results in a degradation of the solvent. MEA is also degraded when temperatures reach above 120 °C, which thus becomes an upper temperature limit in the stripper where the CO_2 is released from the solvent by the addition of heat [5].

A more detailed description of the absorption process is shown in Figure 2.2, in which it can be seen that the lean solvent meets the flue gas in a counter current flow in the absorber. The CO_2 is absorbed by the lean solvent, which is re-defined as a rich solvent when it exits the absorber since it holds more CO_2 than prior to the absorber. The absorption rate is based on chemical reactions and is thus affected by a number of parameters, such as lean-loading, concentration of CO_2 in the flue gas and the residence time, which is affected by the absorber height. The lean loading is a measurement of the amount of CO_2 per amount of MEA, and should be in the range of 0.2-0.3 mol CO_2/mol MEA. The rich-loaded solvent after the absorber should contain about 0.5 mol CO_2/mol MEA. The temperature in the absorber is also important in order to maintain the chemical reactions. The reactions are exothermic, and thus the temperature increases, but if the temperature increases too much the reactions slow down, and in the worst case the reactions could be reversed. Because of this, the lean solvent is cooled to 40 °C and the flue gas to 45-50 °C before they enter the absorber [5].

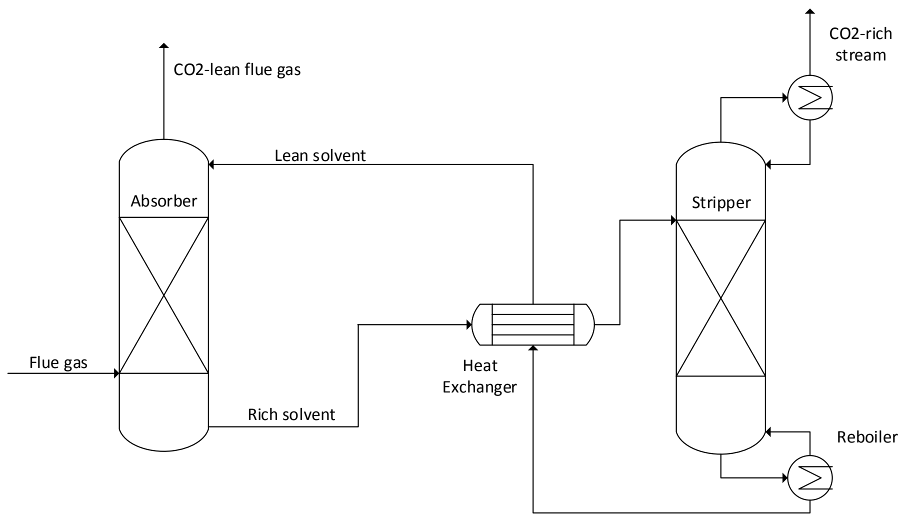

Figure 2.2: A basic configuration of the chemical absorption process.

The lean and rich solvent streams are heat exchanged in order to improve the energy efficiency of the process. The rich solvent is then pumped to the top of the stripper where it flows downwards and is gradually heated up to 100-120 °C in order to break the bonds between MEA and CO_2 [5]. The released CO_2 exit at the top of the stripper after water and other compounds have been condensed. The heat that facilitates the process is added in the reboiler at the bottom of the stripper and usually comes from saturated steam at 3-4 bar taken from the steam cycle. This heat duty is the main energy penalty for the whole process and has a significant effect on the overall efficiency of the plant.

2.1.2 Partial Capture

Despite the globally accepted view on climate change and the need for significant reductions of CO_2 emissions, the implementations of the CCS technology is moving slowly. There are more than one reason for this development, but some of the more important aspects are the high investment and operational costs and the lack of regulations regarding CO_2 emissions. Because of this, the concept of partial capture of CO_2 has been introduced. Partial capture implies that a fraction of the total CO_2 is captured either at one site or in a region as a way to minimize investment and operational cost in order to meet initial restrictions on CO_2 emissions.

Since most larger point sources of CO_2 have multiple CO_2 emission sources, such as multiple power units with individual exhaust pipes, two main approaches for partial capture have been identified. The first approach is to capture a relatively low fraction of CO_2 from each unit, whereas the second approach suggests capturing a high fraction of CO_2 in one or a few of the emission sources. The benefit of the second approach is that when new stricter emission regulations are introduced, new capture units could be added to the unabated emission sources without influencing the other capture units. An increased capture for the first approach would imply larger modification that would affect all processes at the site with a longer overall activity shutdown. [4]

One of the most important benefits of utilizing partial capture is that the technology will be used and thus develop and mature to become a more competitive and economical CO_2 abatement technology. Other benefits with partial capture is that it enables the combination of carbon capture and generation fluctuations. This can be particularly interesting for plants with daily or seasonally variations, since the capture unit could be sized to process only a part of the flue gas flow at a constant level instead of being sized in order to follow the fluctuations of the whole plant. Partial capture could also be beneficial when only a fraction of the CO_2 originate from fossil sources, such as the case with waste incineration. [4]

2.2 Transportation and Storage

When the CO_2 has been captured, it has to be transported to a suitable storage site. This transport can be carried out either by pipelines or ships, depending primarily on the amount of CO_2 that is captured, but also on the specific location of both the plant site and the storage site. Studies on the Nordic countries have shown that for volumes below 5.25 Mton CO_2 per year, ship transports have a lower specific cost, whereas for volumes larger than that, pipelines are less costly [6]. If pipelines are to be used for transportation, the CO_2 has to be compressed to a supercritical state, which implies that it is compressed to around 100 bar [7]. If instead ships are used, the CO_2 only has to be compressed to 7 bar [8], which of course requires much less energy, but then the energy for the shipping has to be added on top of compression work.

The site for CO_2 storage can be either on- or off-shore, and there are mainly three types of geological formations which are considered, and they are oil and gas reservoirs, unminable coal beds and deep saline aquifer formations. If either oil and gas reservoirs or unminable coal beds are used, additional oil or gas can be extracted, which could be seen as an economical benefit for these storage alternatives, but rather contradicting if the main goal is to release less CO_2 to the atmosphere. Deep saline aquifers on the other hand, does not hold this benefit, but the global storage capacity in these formations is estimated to be far more extensive than the the other two alternatives. [8]

2.3 CCS and BECCS

As mentioned in the introduction, there is a potential of reaching negative emissions due to the specific composition of the fuel in the case of waste incineration. The concept of BECCS is an extension of the CCS concept that uses biogenic fuel instead and thus generates negative emissions. According to several reports by the IEA and the Intergovernmental Panel of Climate Change (IPCC), CCS technologies are important tools to reduce CO_2 emissions in order to mitigate global warming. If there however is an overshoot of the atmospheric concentration of CO_2 due to a too slow reduction of the emissions, BECCS technologies together with afforestation is the only measures that could reverse that overshoot and enables us to meet the aim of a global temperature rise below 2 °C by the year 2100 [13]. In this project, the initial goal is to capture the fossil CO_2 and thus generate a CO_2 neutral CHP plant. The potential for extending the capture to become CO_2 negative will be evaluated based on the economical performance due to different tax policy scenarios, as is depicted in Figure 2.3.

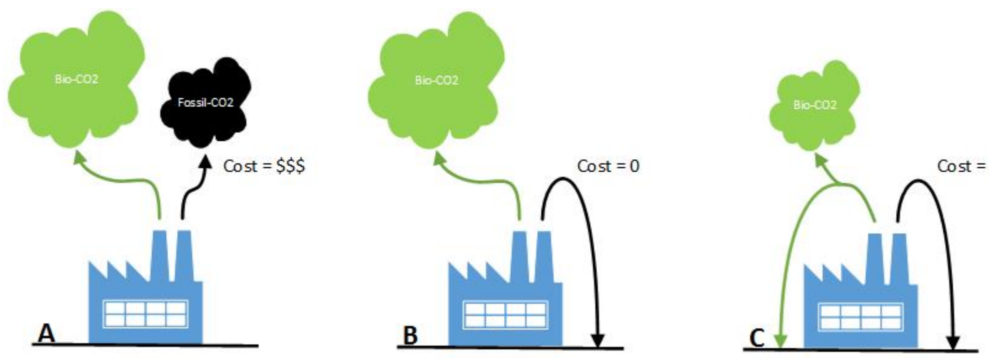

Figure 2.3: Development paths for CO2 emissions from waste incineration depending on regulation scenarios. A: no capture due to low fossil CO2 tax. B: capture of fossil CO2 due to tax incentives. C: capture of biogenic CO2 due to negative costs for negative emissions.

3 Waste Fired CHP Plants

The main purpose for the Swedish waste fired CHP is to generate heat to the district heating system, which supplies 58 % of the total energy consumption in the residences and housing sector [9]. Another purpose is to produce electricity, even if this sometimes is considered as a by-product of the heat production. Yet another important purpose is the incineration of waste, which serves as an energy recycling process of a flow of materials that otherwise is difficult to recycle. This chapter aims to describe the waste incineration process at a detailed level for a general CHP plant in a Swedish context, and with additional specifications for the reference plant, Lillesjöverket.

3.1 The General Plant

Even though modern waste incineration plants are extremely complex as a whole, the combustion process is rather straight forward. The fuel, which usually is a mix of household and industrial waste, is normally combusted in a grate incinerator, which is the most common technique and is used in 90 % of all waste incineration plants in Europe [17]. Different types of grate incinerators exist, but the most common is the reciprocating grate where the fuel is pushed forward by the relative movements of the grate's different sections, and preheated air is supplied from below the grate. The movement of the grate and the air supplied from below controls the combustion and there are primarily two main concerns in this part of the process, to reach a complete combustion and to keep a stable temperature profile in the flue gas train which should be below the maximum acceptable limit. This limit is due to compounds in the fuel that forms corrosive salts at high temperatures which sticks to the heat transfer surfaces in the super-heater, the first heat exchanger in the flue gas train, and reduces the heat transfer capacity. Due to the variations in calorific heat value of the fuel, additives may have to be added in the combustion zone in order to keep a stable temperature profile [17].

The above described grate fired furnace and flue gas train can be seen in the upper-left part of Figure 3.1, together with all the other processes that occur in a state-of-the-art waste fired CHP plant. The steam that is generated in the heat exchangers after the furnace is limited to a temperature of about 400 °C due to the risk of fouling on the heat exchanger surfaces [17], and is used in a steam turbine to produce electricity before it is condensed to supply the district heating system with heat. The district heating system is supplied with pressurized water a around 90 °C and 4 bar, which returns to the plant at a temperature of 40 °C. Apart from the heat generated by condensing steam, heat can also be produced through flue gas condensation. This is seen in the lower-right side of Figure 3.1 where the return water from the district heating system is preheated before it reaches the steam condenser.

The waste is normally kept in a bunker close to the furnace where it is delivered by trucks on a daily basis. The waste is then continuously mixed in order to even out materials from the same delivery with either extremely high or low energy content in order to facilitate a more stable combustion process. The ashes that remains after the combustion corresponds to 15-20 weight-% of the incoming waste and is recycled and used as filling materials at construction sites when possible [17].

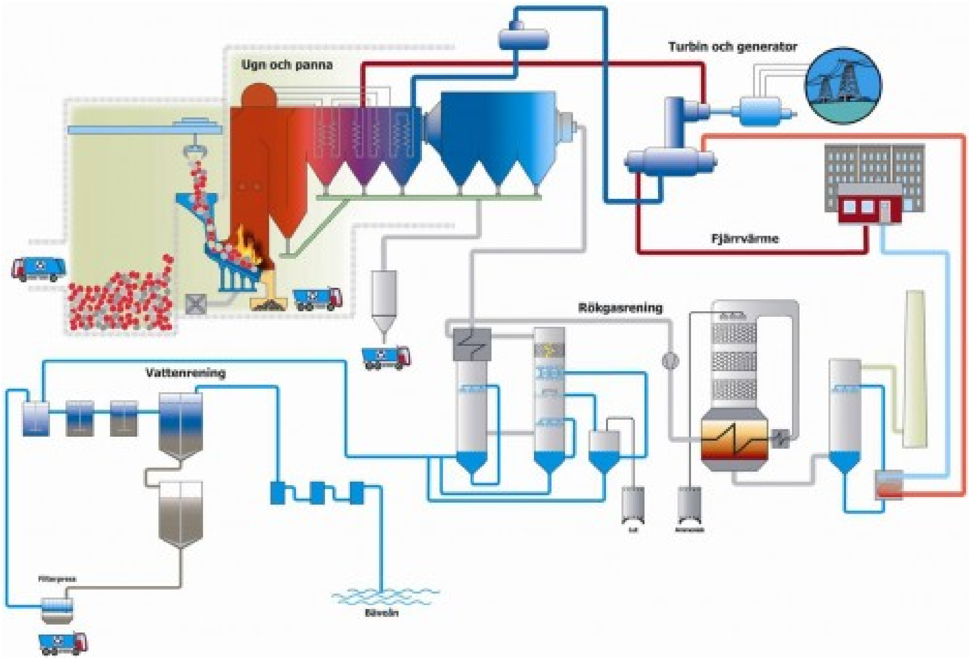

Figure 3.1: The configuration of a state-of-the-art waste fired CHP plant [16].

3.1.1 Flue gas treatment

The flue gas treatment system can be divided into three main parts, reduction of particles, acid gas emissions and nitrogen oxides. There are mainly three techniques to capture the particles, which usually is the first step in the cleaning process, and those are electrostatic precipitator (ESP), bag filters and cyclones. Cyclones have the lowest energy consumption of the three alternatives, but cyclones also have the lowest capture rate for particles smaller than 5 µm. Both the ESP and bag filters are generally very efficient particle removers, with concentrations as low as 15-25 mg/m^3, compared to 200-300 mg/m^3 for the cyclones. Both the ESP and bag filters however consumes more energy than the cyclone, and the most energy is consumed by bag filters due to the large pressure drop that needs to be compensated with a fan. [17]

The reduction of acid gas emissions, primarily HCl and SO_2, is done in either a dry or wet process. In the dry process, lime or sodium bicarbonate is fed into a reactor as dry powder to react with the acid compounds. The resulting products are solid and is normally captured in a bag filter downstream. The wet process takes place in a so-called scrubber where the flue gases are mixed with water to form a very acidic solution with a pH between 0-1. In this first stage the HCl is removed, but due to the low pH, the SO_2 remains. The removal of SO_2 instead occurs in a second stage where the pH is controlled close to neutral. Caustic soda solution or lime milk is added to absorb the SO_2.[17]

Nitrogen oxides (NO_x) emissions can be reduced in several different ways. The initial way to reduce NO_x is to control the combustion such that the combustion is complete and that it does not reach too high temperatures, since that is when NO_x is formed. The combustion can be controlled by fuel or air staging. It is however hard to reach the emission limits just by controlling the combustion, and thus additional measures are required. The most common technology is selective catalytic reduction (SCR), which is located at the end of the flue gas train, just before the stack. Ammonia (NH_3) is added to the flue gas in a zone that also contains a catalyst that facilitates the reaction. The temperature should be in the range of 230-320 °C, and the flue gases must thus be reheated since the temperature after all the other cleaning processes is seldom above 100 °C. This technology is also sensitive to SO_2, and the cleaner the flue gases are before the SCR, the lower the temperature can be and thus less heat is required to reheat the flue gas flow. [17]

3.1.2 Waste as a fuel

As been described in the previous section, the combustion of waste requires a extensive flue gas treatment system in order to reach acceptable emission levels of various toxic compounds. This, combined with the varying energy content and the limited steam temperature are significant drawbacks with waste incineration. There must however be some benefits with this technology since the installed capacity doubled in Sweden between 2000 and 2008 and the import of foreign waste increased by 200 % between 2008 and 2013 [18].

The initial driving force was due to regulations aiming to reduce the use of landfill, but later the economic and environmental incentives have been the main driving forces for this development [18]. Despite the additional investments cost due to the advanced flue gas treatment system, the technology has proven to be economically compatible and feasible, mainly because the plants are paid to incinerate the waste, in contrary to the normal case where the fuel is an expense.

The environmental benefit from waste incineration has been proven to be quite substantial. Even though there are direct emissions at the combustion site of about 750 kg CO_2-equivalents per ton of waste, the net emissions for the whole system is reduced by some 500 kg CO_2-equivalents per ton of waste if it is imported from an average European Union country [18]. This net reduction of emissions is due to the reduction of primarily methane emissions from landfills, the substitution of fossil fuels in the district heating system and the fact that the electricity produced at these sites substitutes electricity from other fossil fueled power plants on the continent [18]. The report also clearly shows that the additional emissions due to transports from other countries are small compared to the other aspects taken into consideration.

3.2 Lillesjöverket - a reference plant

Lillesjöverket is a waste fired CHP plant in the south-western part of Sweden, located in the municipality of Uddevalla. The plant started its production in 2009 and was at the time allowed to incinerate 98 000 tons of waste per year. That number has now been increased to 130 000 tons per year, but for the past years the plant has processed between 118 000-125 000 tons per year [19]. This makes Lillesjöverket smaller than the average sized plant in Sweden, which incinerates close to 148 000 tons per year [20]. The plant produces both heat and electricity, and in 2016 the delivered heat and electricity was 265 000 MWh and 67 000 MWh, respectively.

The furnace has a thermal capacity of 46.2 MW and generates steam at 400 °C and 40 bar. This steam is used in a turbine with two extractions from which the steam is extracted to provide the district heating system with heat and to preheat the feedwater. The turbine has a maximum effect of 10 MW, the district heating system has an effect of up to 40 MW, and the plant has an overall efficiency above 90 %. Lillesjöverket is known for their high availability, which is defined as actual operating hours over planned operating hours, and in 2016 the plant achieved 100 % availability [19]. The plant is however shut down two times per year, one shorter period in the beginning of the summer for a quick overview, and one longer period at the end of the summer for the annual revision. This can be seen in Figure 3.2 where a simplified load curve is shown.

The graph in Figure 3.2 contains three lines. The red line is the load curve for the district heating system to which the plant is connected, based on the load for 2016 [19]. As can be seen, the load curve is simplified to only contain two load levels, a summer and a winter load. In reality, the load changes every hour and is affected by multiple aspects, but despite this the simplification is considered reasonable in this project since its main purpose is to identify the most distinguished trends in the load and to identify periods with excess heat. The other two lines, the heat from condensation of steam and flue gases, are however quite close to reality with some minor exceptions. The main thing that should be highlighted is the excess of heat during the summer months, and this despite that the flue gas condenser is bypassed. Due to this excess of heat, Lillesjöverket has invested in a pellets production process that utilizes this heat in the drying process of transforming sawdust to pellets.

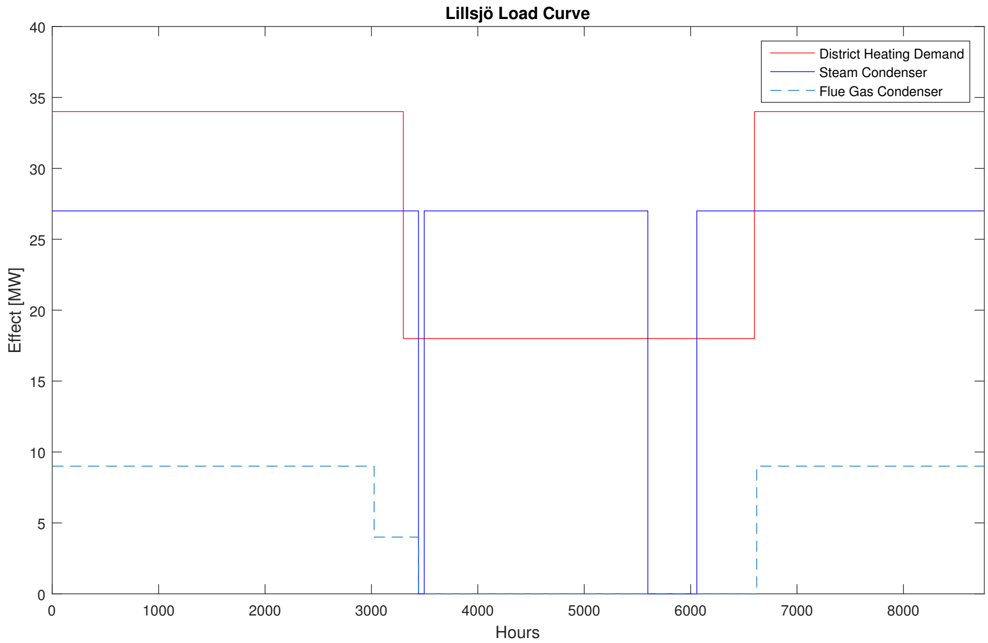

Figure 3.2: A simplified load curve for Lillesjöverket.

Except for the two periods where the plant is shut down, the plant is operating at a constant load at all hours, and this is facilitated by the pellets production. Without this new feature, the load would have to decrease during the summer, resulting in less income from waste incineration and higher thermal stresses on the furnace due to temperature variations, which would reduce the technical lifetime of the furnace. This constant load requires between 14-15 tons of waste per hour, with an average energy content of around 11 MJ/kg. The resulting annual CO_2 emissions are approximately 110 000 tons, out of which 40 %, 44 000 tons, originate from fossil compounds. The initial goal with partial capture is thus to capture those 44 000 tons in order to make the plant CO_2 neural.

A model of Lillesjöverket is shown in Figure 3.3, where red flows represents steam, blue flows water and brown flows flue gas. Steam is delivered to the turbine, from which two steam extractions are made to preheat the feedwater and provide heat to the district heating system. The district heating system is in the bottom right of the figure an is provided with heat from three heat exchangers, two steam condensers and one flue gas condenser. The furnace is modeled with two combined components, a steam generator and a combustion chamber, which are seen in the upper left part of the picture. The flue gas treatment system consists of an electrostatic precipitator (ESP), a combi-scrubber, a selective catalytic reduction (SCR), and a number of heat exchangers.

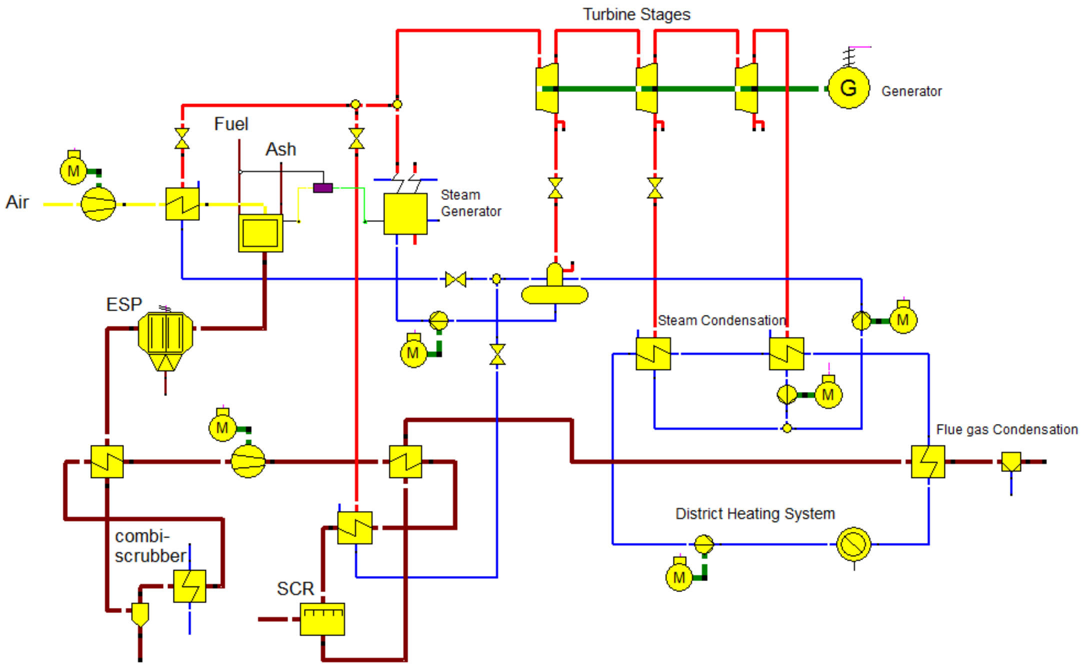

Figure 3.3: The process of Lillesjöverket created in Ebsilon Professional.

4 Methods

Two main design paths have been identified for partial capture [4]. These two design paths are described in this chapter, along with the software that have been used to dimension the CO_2 capture process equipment and to evaluate the process performance. A description of two case studies and the cost estimation methodology is also given in this chapter.

4.1 Design Paths for Partial CO2 Capture

The two design paths for partial capture both originate from the same MEA-based (monoethanolamine) post-combustion process where 100 % of the flue gas flow is treated with a capture rate of 90 %. This design is used as a reference for comparison of the cost of partial capture. The two design paths are referred to as *slip stream path* (SSP) and *separation rate path* (SRP) and are visualized in Figure 4.1. In the SSP design, the flue gas stream is split in two streams out of which only one is led into the absorber where the capture rate is kept at 90 %, and the other stream is released to the atmosphere as usual. For the separation rate path, the whole flue gas flow is led into the absorber, but the capture rate is altered from the reference case conditions by changing a number of parameters to fulfill a desired capture rate, such as the absorber diameter and the solvent flow rate.

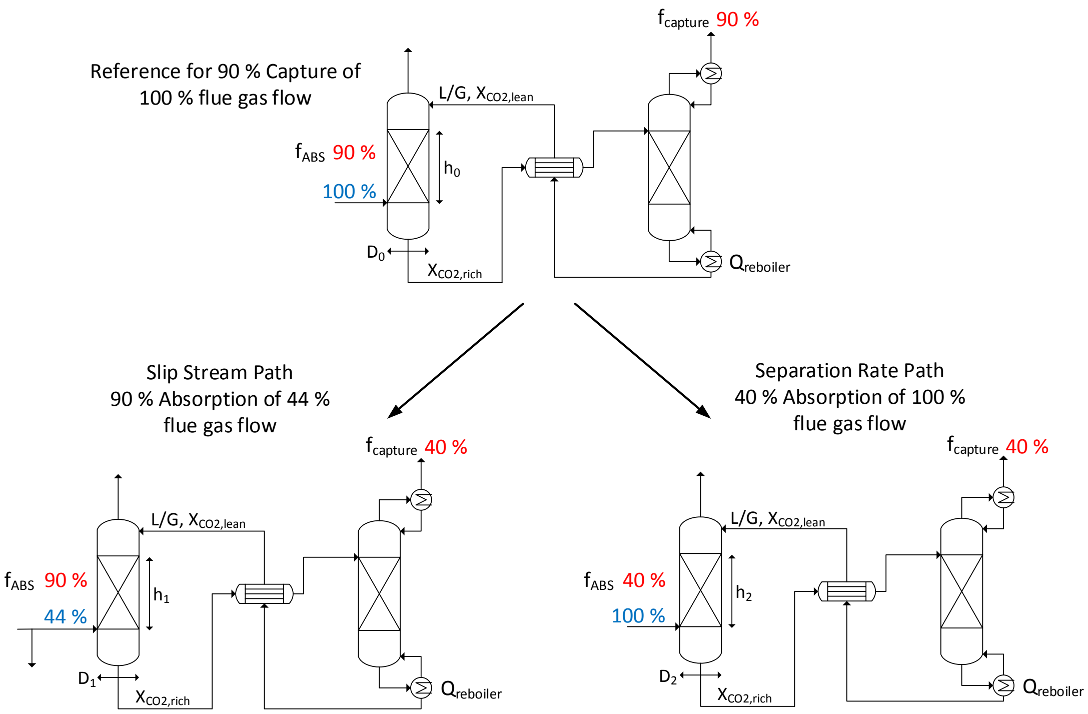

Figure 4.1: Design paths for partial capture of CO_2 with post-combustion. The liquid to gas ratio (L/G), lean and rich loading, and the fraction of CO_2 absorbed and captured are shown in the figure.

4.1.1 Design Path Optimization

The reference design and the two design paths were dimensioned for a specified flue gas flow, with the aim to reach an overall capture rate of 40 % of the total CO_2 emitted from the power plant, which corresponds to the fossil share. With a specific capture rate of 90 % for the slip stream path design, the flue gas flow led to the absorber then only has to be 44 % of the total flow, as can be seen in Figure 4.1. Once the design paths were dimensioned for the given flue gas flow and desired capture rate, the diameters for the absorber and stripper were fixed. A sensitivity analysis was then conducted in order to evaluate the effect of changing several key parameters, such as the absorber height and solvent flow, for the performance of the capture unit process.

4.2 Generic Case

The generic case of waste incineration is based on a theoretical average sized Swedish waste fired CHP plant, which was estimated by dividing the total amount of waste incinerated in one year by the number of active plants [20]. The incineration was assumed to occur at a constant load over the whole year, and the flue gas flow was extrapolated from Lillesjöverket, as a function of the annual amount of waste incinerated.

The generic case evaluates the economic performance of the previously mentioned design paths and the reference design by estimating the investment costs CAPEX and the fixed operating cost, through the external partner TelTek. These costs, together with defined variable operating costs for steam and electricity consumed by the capture unit are used to calculate the specific cost for CO_2 capture. This approach does however not consider other adverse effects of implementing a capture unit to a real CHP plant. Except for the fact that the implementation would contribute with additional costs from the use of steam and electricity, it would also limit the maximum power output from the plant and thus its ability to meet the demand at all hours. This would create a situation where other units in the system would have to start, and this is not included in the calculations. Also, it is assumed that all heat and electricity required for the capture unit is available.

4.2.1 Off-design Performance

An evaluation of the off-design performance was also conducted for the different design paths. Here, the definition of off-design is the operation of the capture unit at a different capture rate than what it was designed for. It could be argued that the SRP-design is an off-design operation of the reference design, but since the SRP-design has been dimensioned specifically for a capture rate of 40 %, it is considered as on-design when the capture rate of 40 % is maintained.

The evaluation of the off-design performance was conducted by a sensitivity analysis where the following key parameters were varied; The solvent flow rate and the reboiler heat input were the main parameters analyzed, and the critical and limiting aspects considered in the process were flooding in the absorber and stripper, violation of temperature limits in the heat exchanger and the temperature of the lean solvent leaving the stripper.

4.2.2 Design Point Definition

The SRP design was initially designed to capture 40 % of the CO_2, since that is the overall target. This design thus led to some specific dimensions and characteristics regarding flexibility and off-design performance. It was then realized that higher capture rates could be of interest, and thus the design had to be changed, but still be within the SRP design concept. This is where the expression *design point* is introduced. For the initial design, the design point is 40 %, since that is for which capture rate the dimensions, such as diameters and heights, have been fixated. Additional models with different design points were then generated which had higher capture rate set points that led to other dimensions and characteristics.

4.3 Lillesjöverket - a case study

In order to evaluate the cost and other effects of capturing CO_2 from CHP plants, a capture unit is integrated to an existing waste fired CHP plant model. In this case study, Lillesjöverket, which is about 20 % smaller than the generic plant, is used as a reference plant. The effect of implementing a capture unit will be evaluated for two different cases, which are described in the following subsections.

For the first case, it is ignored that Lillesjöverket recently invested in a pellets production process where excess heat during the summer months is used to dry sawdust. The reason for not including the pellets production in the first cases is that it is not considered representative for the majority of Swedish waste plants. The second case does however evaluate the possibilities and effects of implementing a capture unit to Lillesjöverket, including the pellets production.

The excess heat currently used in the pellets production arises from the fact that the heat demand in the district heating system is lower during the summer, and this seasonal variation is a key aspect when evaluating partial CO_2 capture. The excess heat present during the summer could be used to capture a significant part of the fossil CO_2 during that time and thus reduce the impact on the heat and electricity production for the remaining part of the year. A simplified load curve, see Figure 3.2, has therefore been established from the data given from Lillesjöverket, which is used to evaluate the following cases.

4.3.1 CASE 1: Lillesjöverket excluding pellets production

Due to the specific characteristics of Lillesjöverket and its load curve, it was assumed that a more efficient design point could be found than the one described in section 4.1.1. The objective is to find a design that can maximize the use of excess heat during the summer, and by that minimize the costs, and at the same time be used in a reasonable low-load operation during the rest of the year. The low-load operation during winter is necessary since the target amount of 40 % fossil CO_2 cannot be captured during the summer.

4.3.2 CASE 2: Lillesjöverket including pellets production

In this case, the pellets production is taken into account in the model, and the objective is to find the best suitable capture design and implementation in order to fulfill the goal of capturing the fossil share of the emissions. This case may include additional equipment such as heat pumps, electric or biomass boilers in order to facilitate the capture process due to the lack of excess heat in the system due to the pellets production.

4.4 Process Modeling Tools

Two different software tools were used to model the investigated processes, Aspen Plus and Ebsilon Professional. The former was used to model and optimize the CO_2 absorption process and to dimension the process equipment, while the latter was used to extract flue gas data from a detailed model of the reference CHP plant and to evaluate the effect on the plant once the absorption process was implemented.

4.4.1 Aspen Plus

Aspen Plus V8.8 is a commercial process simulation software used for conceptual design and optimization of various chemical processes. The software handles mass and energy balances, and includes a large amount of chemical and physical property data that enables the program to handle chemical reactions, mass transport and equilibrium calculations.

For this thesis, the process configuration shown in Figure 4.2 is chosen. The configuration is referred to as rich stream splitting (RSS), and as can be seen in the figure, the rich solvent stream is split into two streams on its way to the stripper, with only one of these streams being preheated with a heat exchanger. The two streams are injected to the stripper at different heights, with the unheated stream injected at the top and the heated stream further down. This configuration creates a temperature distribution with the lowest temperature at the top of the stripper and the highest temperature at the bottom where the heat from the reboiler is injected. The solvent is thus gradually heated to the temperature where the CO_2 is released. The benefit with this configuration is an improved energy efficiency, compared to a configuration without this stream splitting, which is achieved with a minimal and very cost effective transformation of the process. This is a well known configuration that has been used and evaluated in several academic articles [14].

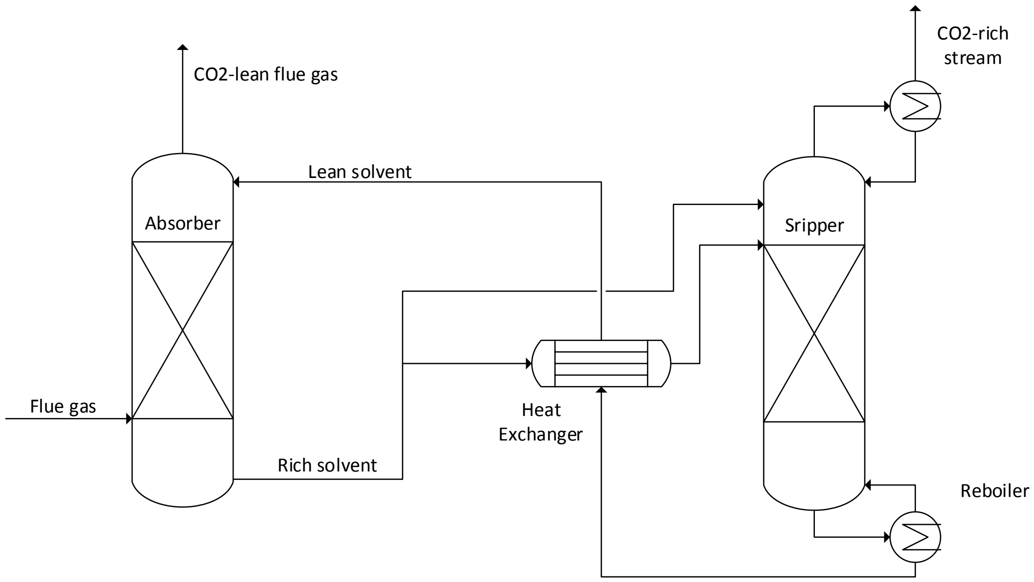

Figure 4.2: Absorption process configuration with the RSS configuration.

4.4.2 Ebsilon

Ebsilon Professional is a simulation tool that simulates thermodynamic processes and is used for designing and optimizing thermal power plants. Due to the high flexibility and broad range of components, the software can be used to model virtually any thermodynamic cycle. This software was primarily used to evaluate the change in electricity and heat production when a capture unit was implemented, but it was also used to identify limitations on steam extraction and possibilities to use excess heat to preheat the district heating water in order to improve the overall performance. More details about the modeling can be found in Chapter 5.

4.5 Cost Estimations

Cost calculations are vital to the project and are performed by the external partner TelTek. TelTek evaluates the investment and operation cost for the equipment required, and thus a number of design parameters have to be identified i order to carry out the cost estimations. These design parameters are typically the absorber and stripper diameter and height, pump capacity and the heat exchanger area and temperature range. A complete list of the components included in the cost estimation can be found in Appendix 1.

Further cost calculations were however required and below follows a description of how those calculations were performed. Since the cost estimations by TelTek only were performed for the three designs in the generic case, the reference design, the separation rate path (SRP) and the slip stream path (SSP), a scaling method was used in order to calculate the investment cost (CAPEX) and fix operational cost (fix OPEX) for SRP designs in between the ones already calculated by TelTek, which is the reference design and the SRP designed with a capture rate (CR) of 40 %. The scaling of CAPEX and fix OPEX is performed according to equation 4.1 and 4.2, where the scaling factor, α, is determined using the existing values for CAPEX provided by TelTek for the reference design and the SRP_{40} design. It should be emphasized that SRP_{90} becomes the same design as the reference design. The scaling factor is calculated in Appendix 2. The maintenance factor in equation 4.2 ranges between 2 % and 6 %, with a base value of 4 %.

$$CAPEX_{SRP,i} = CAPEX_{SRP,40} \left(\frac{CR_{SRP,i}}{CR_{SRP,40}} \right)^{\alpha} \quad [EUR] \tag{4.1}$$

$$OPEX_{fix,SRP,i} = X_{maintenance} * \frac{CAPEX_{SRP,i}}{annuityfactor} \quad [EUR] \tag{4.2}$$

The price of steam was also calculated. This was done by accumulating the loss of income from electricity and heat to the amount of steam extracted to the capture unit. A steam extraction was added in the Ebsilon model, as can be seen in Figure 5.4, and for various amounts of steam extracted, the effects on the electricity and heat generation were monitored. The cost is calculated in euros per ton of steam used in the capture unit according to equation 4.3 below. The cost for steam is based on the losses from electricity and heat income due to the steam extraction, and the difference in generated power output is thus multiplied with the price of electricity and heat, respectively. The loss of income is then divided with the steam flow in tonnes per hour since the desired unit for the steam cost is euros per hour. The cost changes considerably between the winter and summer period since the price for heat changes significantly and the fact that the heat production is more affected than the electricity production when steam is extracted to the capture unit. The price for electricity and heat is displayed in Table 4.1 along with the amount of hours allocated to summer and winter, which is also shown in Figure 3.2.

$$C_{steam} = \frac{(P_{el,max} - P_{el,ext.})C_{el} + (Q_{heat,max} - Q_{heat,ext.})C_{heat}}{\dot{m}_{steam,ext.}} \quad [EUR/ton] \tag{4.3}$$

Table 4.1: The price for electricity and heat during summer and winter.

	Electricity [EUR/MWh]	Heat [EUR/MWh]	Time [Hours]
Summer	27	5	2750
Winter	27	21	5470

During the summer period, prior the implementation of a capture unit, it was assumed that the plant decreases its furnace load to 70 % of its nominal value due to the decreased demand for heat and the limited cooling capacity. When the capture unit was implemented, it was then assumed that the plant increased the furnace load in order supply the heat required in the capture unit as well as maintaining the district heating output. The furnace load can however not increase further than 100 %, and if more heat was required, the district heating output was decreased to the benefit of the capture unit, leading to less income from district heating. This also entails two other changes regarding the economics of the plant. The change in electricity production between 70 % furnace load and the increased load with steam extraction to the capture unit generates additional income, and the additional combustion of waste also generates an additional income. The calculations are shown below in equation 4.4 and 4.5.

$$I_{electricity} = (P_{el,extraction} - P_{el,70\%})C_{el} \quad [EUR/h] \tag{4.4}$$

$$I_{waste} = \frac{m_{waste,ton/yr}}{t_{hrs/yr}} * I_{EUR/ton} \left(\frac{\dot{m}_{fuel,extraction} - \dot{m}_{fuel,70\%}}{\dot{m}_{fuel,100\%}} \right) \quad [EUR/h] \tag{4.5}$$

5 Modelling

This chapter describes how the models in Aspen Plus and Ebsilon Professional have been set up and how they have been run. The focus is to explain the most important aspects in detail and to some extent omit less important components in the models that can be seen as basic and thus easy to find information about in literature referred to in the text.

5.1 Aspen Plus

The MEA-based absorption process is modeled in the Aspen Plus V8.8 software, and apart from the simple schematic of the model in Figure 4.2, a more detailed schematic is shown in Figure 5.1. The solvent is a 30 weight-% MEA solution with a pressure of 1 [bar] on the absorption side and 2 [bar] on the stripper side. The packing material used in the columns is Sulzer Mellapak 250Y which is a commonly used packing material in this type of processes. Aspen Plus uses the built in electrolyte nonrandom two-liquid (ELECNRTL) method and the Redlich-Kwong equation of state for computing the the physical properties of the liquid and vapor phase, respectively. A more detailed description of the process model can be found in the work by Gardarsdóttir [21].

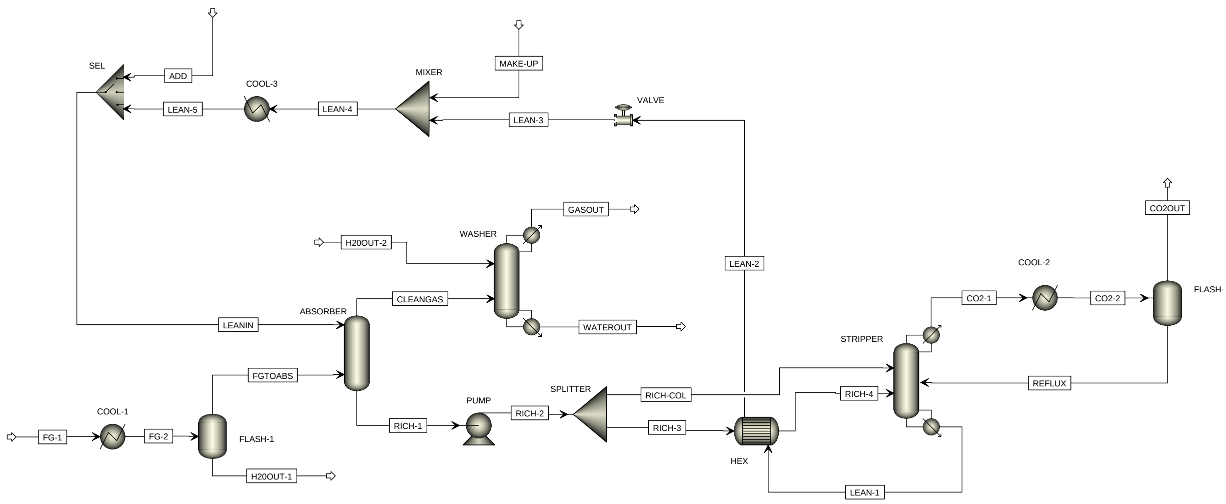

Figure 5.1: A process flow-sheet over the absorption process in Aspen Plus.

Two process units not shown in the Aspen flow sheet in Figure 5.1 have to be mentioned. The first unit is the flue gas fan that is implemented prior to the absorber column. The fan increases the flue gas pressure before it enters the absorber, but the pressure increase is only moderate and is due to the pressure losses in the absorber and washer. The pressure increase over the fan is thus set to the cumulative pressure drop in the absorber and washer, and the model of this unit is illustrated in Figure 5.2. The fan has an isentropic efficiency of 0.865 and a mechanical efficiency of 0.98.

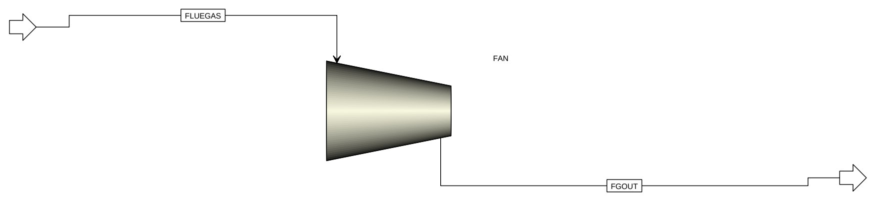

Figure 5.2: Aspen model of the flue gas fan.

The second and more energy consuming process is the CO_2 compression unit placed downstream of the capture unit. The clean CO_2 is compressed in a four-staged axial compressor with a pressure ratio of 2.99 for each stage, reaching a pressure of 80 [bar] at the outlet. The CO_2 is inter-cooled between each of the compression stages to 25 °C in order to improve the efficiency, which is 0.72 for all stages, and to further remove moisture. Once the CO_2 stream has been compressed to a liquid state, it is further pressurized to 110 [bar] with a pump, as is shown in Figure C.2. The pump has an efficiency of 0.796, and since the transportation and injection to final storage site is excluded in this work, this is the last step for the CO_2 in this process assessment. The cost for compressing the CO_2 is included in the cost calculations.

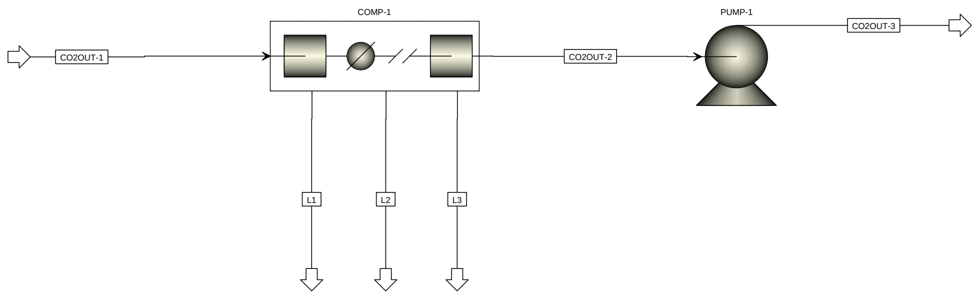

Figure 5.3: The compression unit of the pure CO_2 stream.

5.1.1 Generic Case

The models for the generic cases are based on the flue gas (FG) properties displayed in Table 5.1 and 5.2. The flue gas composition is based on weight percentage in Table 5.1, and the capture rate (CR) is specified for both the specific capture rate in the absorber and the total for the whole flue gas flow in Table 5.2. This is because the SSP-design only process a fraction of the flue gas flow. The CO_2 emissions displayed in Table 5.2 is the amount of CO_2 not captured in the absorber and thus released to the atmosphere. This figure is one of the most important design parameters and will be described in detail below.

Table 5.1: Flue gas properties and composition for the generic CHP plant.

$\dot{m}_{FG}$ [kg/s]	T [°C]	P [bar]	x_{N_2} [wt.%]	x_{CO_2} [wt.%]	x_{O_2} [wt.%]	x_{H_2O} [wt.%]
31	50	1.04	0.724	0.175	0.006	0.095

Table 5.2: Chosen specifications for each design path for the generic case.

	$\dot{m}_{FG}$ [kg/s]	CR [%]	CR_{tot} [%]	CO_2 emissions [kg/s]	Captured CO_2 [kg/s]
Reference	31	90	90	0.54	4.88
SRP	31	40	40	3.26	2.17
SSP	13.78	90	40	0.24	2.17

The dimensioning is governed by a number of design specifications in Aspen, which specifies constraints for some key aspects such as emission levels and stream compositions. One of these design specifications is the one controlling the CO_2 concentration in the solvent stream leaving the stripper, also denoted lean loading. The lean loading is set as a design specification and is defined in equation 5.1. The lean loading is controlled by adjusting the reboiler heat duty, and for the design point the lean loading was set to 0.32 mol CO_2 per mol MEA [22].

$$x_{CO_2} = \frac{n_{CO_2} + n_{HCO_3^-} + n_{CO_3^{-2}} + n_{MEACOO^-}}{n_{MEA} + n_{MEAH^+} + n_{MEACOO^-}} \tag{5.1}$$

When the lean loading is specified, the heat duty in the reboiler depends mainly on the solvent mass flow and the rich loading, which denotes the concentration of CO_2 in the stream leaving the absorber. The pressure in the stripper also affects the lean loading, but since the pressure was set to a constant value of 2 [bar], this parameter was no longer considered. The mass flow controlling design specification varies the solvent mass flow in order to reach a certain capture rate, or as it is defined in Table 5.2, the amount of CO_2 remaining in the exhaust gas. By doing this, both the reboiler heat duty and the absorber and stripper diameter is influenced. The absorber diameter is affected by a constraint on the fractional capacity, also referred to as flooding. For a specific diameter, there is only a certain amount of gas and liquid that can meet in the column before it becomes flooded, and a constraint to adjust the diameter to only reach 78 % of this flooding limit was set for operation at the design point.

As can be seen in Figure 5.1, the absorber and washer are modelled as separate components even though they sit on top of each other in a single column in reality. This simplifies the simulation, and the diameter of the washer was linked to adjust to the absorber diameter through a calculator block in Aspen Plus. The height of the washer was however dimensioned with a constraint to keep the concentration of MEA below 1 ppm on a molar basis in the exhaust gas.

In addition to what has already been specified about the model, Table 5.3 presents all the other settings that does not change for any of the designs. The absorber diameter, solvent flow rate and heat exchanging area in the heat exchanger are parameters that changes with the different design paths and are thus a result of the dimensioning and are therefor presented in Chapter 6.

Table 5.3: Fixed input data for the CO_2 absorption model that is valid for all cases in this study.

Absorber		
- Stages	30	
- Height	25	[m]
- P	1	[bar]
- $T_{solvent,in}$	40	[°C]
- $T_{gas,in}$	40	[°C]
Washer		
- Stages	10	
- P	1	[bar]
Stripper		
- Stages	20	
- Height	12	[m]
- P	2	[bar]
Reboiler		
- Type	Circulation w/o baffle	
Heat Exchanger		
- $\Delta T_{hotout,coldin}$	10	[K]
- ΔT_{min}	1	[K]
- ΔP	0	[bar]
- U	1500	[W/m^{2}K]
Splitter		
- Split fraction	0.8	

When the dimensions for a chosen design path had been made, the dimensions of the model were fixated in order to perform an off-design evaluation. This means that the design specifications described previously were inactivated, and the absorber diameter and heat exchanger area were locked, the reboiler heat duty varied within a reasonable range and the solvent flow rate optimized for each of the different heat duties. This off-design evaluation meant that the capture rate changed, as well as the lean loading. The limitations for off-design operation were identified in mainly three components, flooding in the absorber, temperature violations in the heat exchanger and too high temperatures in the CO_2 lean stream leaving the stripper. An upper limit for flooding was set to 0.83, violations in the heat exchanger could to some extent be avoided by changing the splitter fraction and the solvent stream was kept below 122 °C in order to keep the degradation of the MEA-based solvent within reasonable limits [24].

5.1.2 Lillesjöverket

The difference between Lillesjöverket and the generic plant is both the size of the flue gas flow and the flue gas composition, as can be seen in Table 6.4. This difference has little effect on the absorption process and thus only the best performing partial capture design from the generic analysis was applied to Lillesjöverket. The cost of capturing CO_2 from Lillesjöverket was then evaluated both with and without the pellets production, referred to as case 1 and 2 in Chapter 4.

Table 5.4: Flue gas properties and composition for Lillesjöverket and the generic plant.

	$\dot{m}_{FG}$ [kg/s]	T [°C]	P [bar]	x_{N_2} [wt.%]	x_{CO_2} [wt.%]	x_{O_2} [wt.%]	x_{H_2O} [wt.%]
Lillesjöverket	25	50	1.04	0.737	0.146	0.072	0.045
Generic plant	31	50	1.04	0.724	0.175	0.006	0.095

5.2 Ebsilon Professional

The model used to represent the CHP plants was originally created in collaboration between scientists at Chalmers University of Technology and engineers at Uddevalla Energi in order to simulate Lillesjöverket. The model was constructed before the plant was equipped with a pellets production unit. The model is described in detail in the course Heat and Power Systems Engineering given at Chalmers [23], and is shown in Figure 3.3. In Table 5.5-5.7, data is given for several of the most important parameters. The fuel composition is given in Table 5.5, and temperatures and pressures for steam and district heating water is given in Table 5.6 together with steam extraction pressures. The last table specifies the mass flows of fuel, flue gas and steam in the steam cycle, as well as the electricity and heat output, for both the generic plant and Lillesjöverket for both 100 % and 70 % furnace load. The model used for the generic plant is an up-scaled version of the model of Lillesjöverket.

Table 5.5: Fuel composition used in the model.

NCV [MJ/kg]	C [wt.%]	H [wt.%]	O [wt.%]	N [wt.%]	S [wt.%]	Cl [wt.%]	H2O [wt.%]	ASH [wt.%]
11.1	0.298	0.042	0.154	0.006	0.002	0.006	0.36	0.132

Table 5.6: Primary steam conditions and extraction pressures as well as pressure and temperatures for the district heating system.

$P_{primary}$ [bar]	$T_{primary}$ [°C]	$P_{ext,1}$ [bar]	$P_{ext,2}$ [bar]	P_{end} [bar]	T_{DH} [°C]	P_{DH} [bar]
40	400	6	1.286	0.4	90/40	4

Due to a lower heating demand during the summer, the furnace load is assumed to be reduced to 70 % during this period, and data for this part load is therefore displayed together with the full load data for both plants in Table 5.7. When the capture unit is added, it was assumed that the furnace will be operated at full load during the summer since more heat is required in the absorption process, and the additional electricity produced and not used in the capture unit would generate an income, as well as the increased amount of waste combusted, which would lead to a lower capture cost.

Table 5.7: Key data for the generic plant and Lillesjöverket at 100 % and 70 % furnace load.

	$\dot{m}_{\mathrm{fuel}}$ [kg/s]	$\dot{m}_{FG}$ [kg/s]	$\dot{m}_{steam}$ [kg/s]	$\dot{m}_{DH}$ [kg/s]	P_{el} [MW]	Q_{heat} [MW]
Generic plant						
- 100 %	4.86	31	19.8	188	12.34	39.4
- 70 %	3.39	21.7	13.82	131.3	8.61	27.5
Lillesjöverket						
- 100 %	3.95	25	16.08	152.8	10.0	32
- 70 %	2.76	17.7	11.26	107	7.0	22.4

5.2.1 Steam Extraction

The steam required in the reboiler is extracted from the first turbine stage, also referred to as the first extraction point, at a pressure of 6 bar, which is shown in Figure 5.4. The mass flow of the extracted steam is controlled by *Controller 1* such that the specified heat duty in the reboiler for the capture unit is met. Due to this extraction, less steam is used to produce electricity and heat, and in order to maintain the mass flow in the steam cycle, *Controller 2* adjusts the mass flow in the district heating system. If the district heating mass flow is not controlled, the model would increase the mass flow in the steam cycle and the model would no longer represent the same power plant. The condensate leaving the reboiler is saturated liquid at 3 bar and 133 °C, which is then used to preheat the district heating stream before it is led back to the feed water through the deaerator.

The condensate could also have been feed directly to the feed water stream, but then the district heating system would have been even more penalized by the extraction than it already is. Since all heat production is located downstream of the extraction point, in contrary to the electricity production where more than 50% of the shaft power is generated by the first turbine stage, the heat production becomes more affected by the extraction than the electricity production.

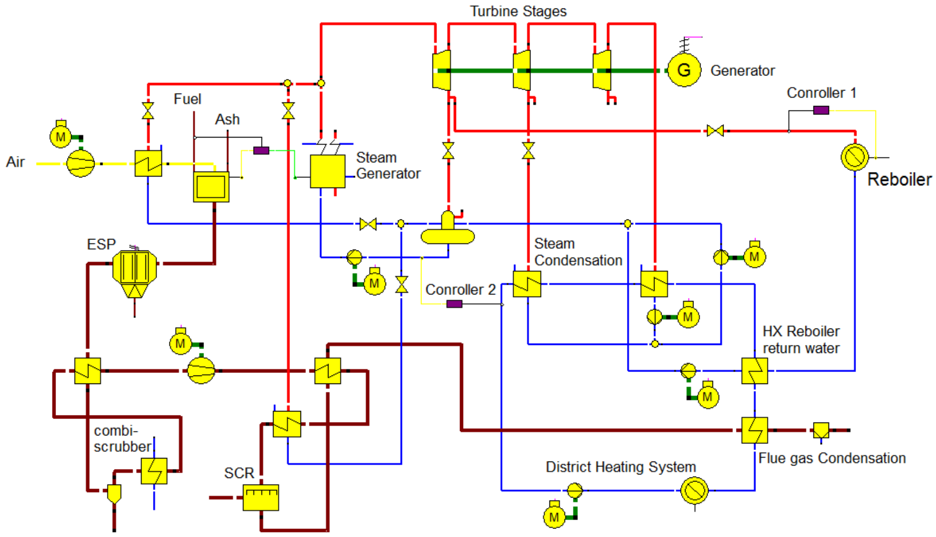

Figure 5.4: A process figure of the waste-fired CHP plant with the additional steam extraction to the reboiler. Note the two controllers that are used to control the extracted mass flow and maintain the steam cycle mass flow by changing the district heating system mass flow.

5.2.2 Pellets Production

An extraction of heat to the pellets production process is required for Case 2. This is done by using district heating water at 90 °C in a heat exchanger with a 30% ethylene-glycol mixture with distillate water on the cold side. It is three plate heat exchanger with a maximum effect of 5350 kW each that is used for this purpose, and the diverted stream is controlled by *Controller 3* that calculates the correct flow in order to reach the desired thermal effect. The pellets production process is shown in the bottom of Figure 5.5, below the district heating system.

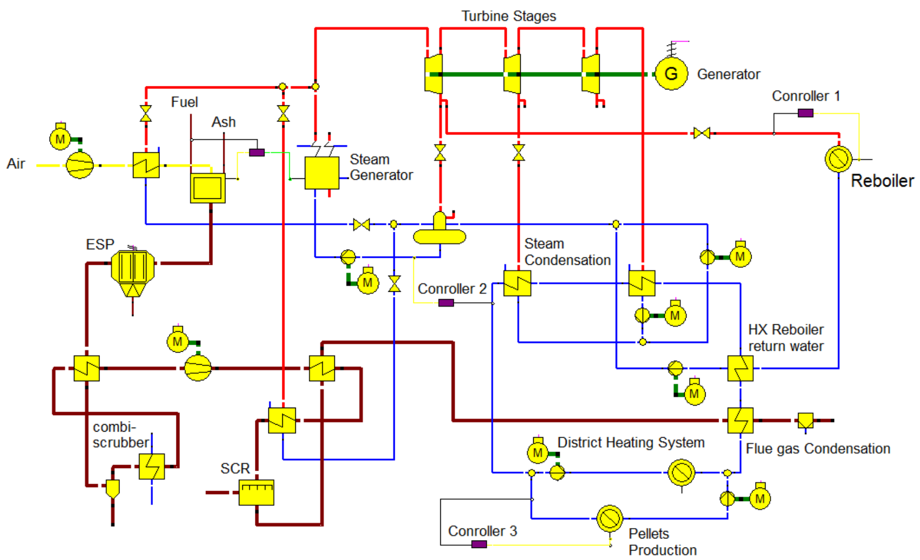

Figure 5.5: A process figure over the plant including both the steam extraction for partial capture and the pellets production process.

5.2.3 High Pressure Steam Extraction

Lillesjöverket has the ability to by-pass the turbine with high pressure steam in order to benefit the heat production. This is done by extracting steam at 40 [bar] and 400 °C before it enters the turbine and instead lead to the second of the two steam condensers in the district heating system. This was done in a separate model since there is no possibility to switch streams on and off in Ebsilon. The high pressure steam extraction is seen in Figure 5.6.

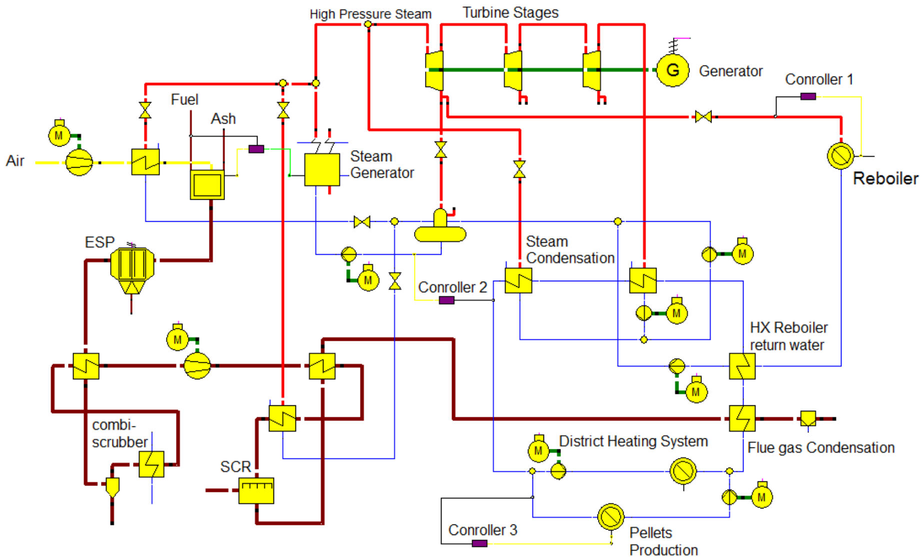

Figure 5.6: A process figure over the plant including both the high pressure steam extraction and the pellets production process.

6 Results and Discussion

This chapter presents and discusses the results obtained from the simulations and the cost estimations of the design paths described. The results are divided in generic and specific case results, and a large part of the chapter covers the analysis of the SRP design and the effect of seasonal operation optimization.

6.1 Results for the Generic Case

In Table 6.1, five key economical aspects are presented for the three designs that have been evaluated. The specific cost of capturing one ton of CO_2 is significantly lower for the generic reference design, compared to the other designs. The generic reference design is in fact the same design as the SSP_{40} model, but with a considerably larger flue gas flow, and as can be calculated from the table, the specific cost is 29 % lower for the reference design. One should however be careful to judge the results only on this parameter. The investment cost for the reference design is 42 % higher than for the SSP_{40} design, which implies that the risk for the investors is much greater in the full capture case, even though the specific cost is lower.

Table 6.1: The economical result for the three designs. Note that the OPEX and specific cost is calculated for a constant capture rate and reboiler heat duty throughout the year, additional income for waste incineration and electricity production is not included and that the amount of captured CO_2 differentiate significantly.

	Investment [kEUR]	CAPEX [kEUR/a]	fix OPEX [kEUR/a]	OPEX [kEUR/a]	Specific Cost [EUR/ton CO_2]	CO_2 Captured [ton/yr]
Reference	41 389	4 117	1 656	3 872	67	144 500
SRP_{40}	31 342	3 118	1 254	1 929	98	64 200
SSP_{40}	29 071	2 892	1 163	2 004	94	64 200

6.2 Evaluation of the SRP Design

The SRP design path was evaluated for a number of different design points, which are defined in section 4.2.2, and the results of this evaluation will be presented in this section. The evaluation has been divided in to three areas according to the subsections below.

6.2.1 Off-design Operation Range

In total five models of the SRP design were generated, each with different design point. The off-design operation range for these designs is shown in Figure 6.1, and as can be seen in the figure, the capture rate can increase with 10 %-units from the design point, and decrease with about 15-20 %-units. A specification of the limitations that defines the operation range is found in section 5.1.1. If a capture rate in the interval of 38-79 % is desired, it can be seen that several design points can fulfill the requirement within their off-design range. In Figure 6.1, it can however not be determined which design point that should be chosen before any other with regards to energy efficiency, and thus more information is required.

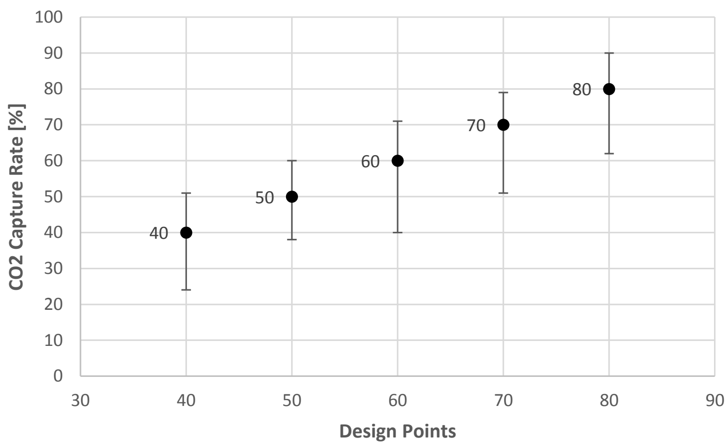

Figure 6.1: The operation range with regards to the capture rate for the SRP design at five different design points.

The additional information added to the graph in Figure 6.2 is the reboiler heat duty. It can now be seen that for a specific capture rate, e.g. 50 %, three design points are clearly eligible, but the required reboiler heat duty varies considerably. Thus, from an energy point of view, it is now possible to select the most efficient design point for a desired capture rate. It should be noticed that since the curves are non-linear, the specific heat required, Q_{spec} [MJ/kg$_{CO_2}$], is increasing for increased capture rates, independently of design point. All design points have a specific heat duty within the range of 2.8-4.0 MJ/kg$_{CO_2}$, with the lower value at the lower end of the capture rate bound.

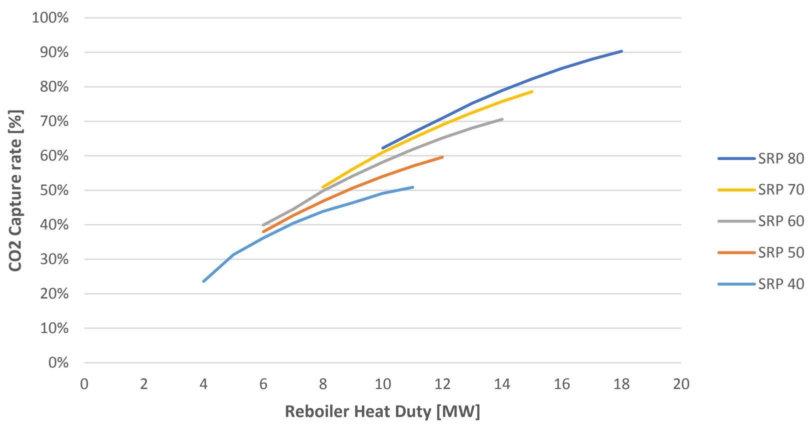

Figure 6.2: The capture rate as a function of reboiler heat duty for the five different design points of the SRP design.

6.2.2 Summer and Winter Operation

The point of evaluating the off-design performance of the SRP design is to utilize the reduced need for heat and the corresponding lower cost for steam during the summer. The hypothesis is that if more CO_2 is captured during summer, the cost for capturing the target 40 % on an annual basis would decrease. Figure 6.3 shows the operation costs during the summer period, which is defined as 2750 hours. As can be seen, the absolute cost for steam, cooling water and electricity increases with higher design points since more CO_2 is captured. The cost for engineers and operators is however constant since all models, independent of design point, are assumed to operate during the whole summer period. The capture units are operated at its maximum capture rates according to Figure 6.1, and the main aspect that should be noticed during the summer period is the relation between the cost for steam and the other parameters. Even though steam is the most expensive parameter, it is not considerably more expensive than cooling water and electricity and does not in any way stand out as a critical parameter. The specific heat demand, in MJ/kg CO_2, does increase when the capture unit is operated at its maximum capture rate, but this is compensated for by the low steam price during the summer.

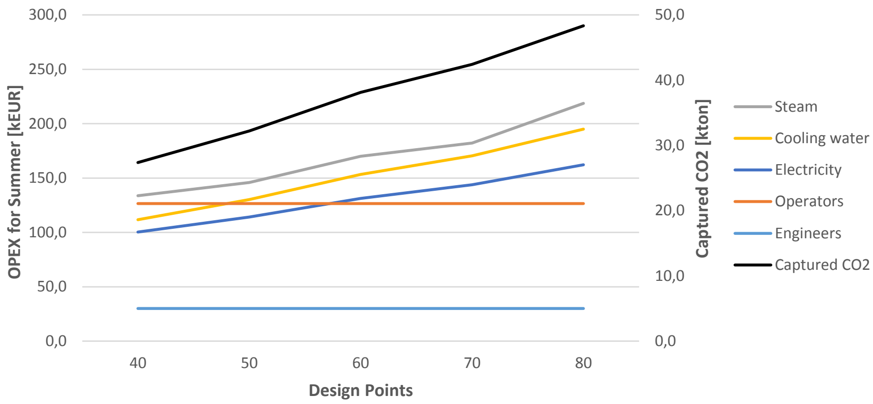

Figure 6.3: Variations of variable cost parameters for different design points during the summer period. Note that the cost on the y-axis is for the whole summer period.

Two similar graphs for the winter period, which is defined as 5470 hours, are shown in Figure 6.4 and 6.5, and they have a completely different appearance than Figure 6.3. The reason to why there are two graphs for the winter period is that the capture rate (CR) can either be maximized or minimized according to the operation range shown in Figure 6.1. The capture rate could of course be maximized or minimized for the summer as well, but due to the lower steam price, it is not considered reasonable to minimize the capture rate during the summer. The costs in Figure 6.4 is calculated for a maximized capture rate, which means less operational hours but a higher steam demand, and the opposite applies to Figure 6.5, where the minimum capture rate is used. There are mainly two reasons to why the steam cost has changed so dramatically for the winter period, and the first and most obvious reason is that the steam price is considerably higher due to a higher price on heat. The other reason, which also is the reason to why the slopes are negative, is that if more CO_2 is captured during the summer, less have to be captured during the winter, and thus less steam is required. Here it should be highlighted that the cost for summer and winter has to be compared for the same design point, since the design point will stay constant once the unit is built. The cost for engineers and operators are, in contrary to the summer, not constant during the winter. This cost is dependant on the operation hours of the capture unit, and since models with a lower capture rate requires more operation hours to reach the targeted amount of captured CO_2, they have higher costs for engineers and operators.

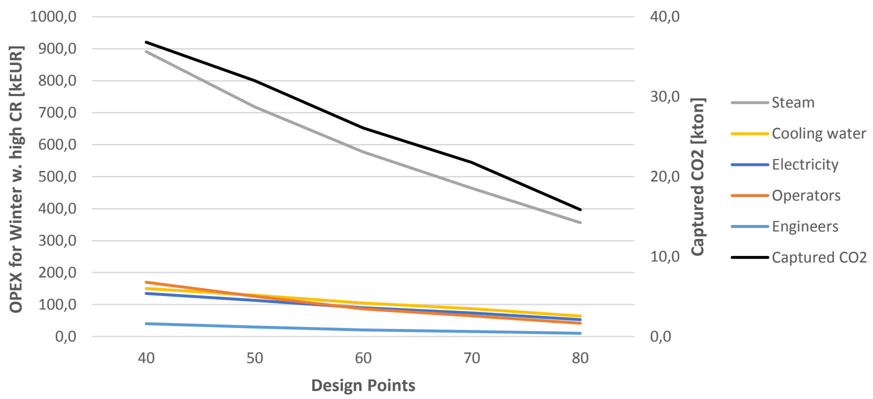

Figure 6.4: Variations of variable cost parameters for different design points during the winter period with a high capture rate. Please notice the large difference on the y-axis scale compared to Figure 6.3 and that the cost is for the whole winter period.

The results show that even if the costs for engineers and operators increases, the cost for steam decreases even more and a lower total operational cost is achieved for all design points for the case with a lower capture rate during the winter. It should however be mentioned that the difference is small between the two approaches, and that other aspects may play a more important role in the decision making of how to run the capture unit. One example could be that all heat generated is required in the district heating system during the coldest period of the winter, when the price for heat also reaches its maximum, and that the emissions for that period could be captured during another period when there is more heat available for the capture unit.

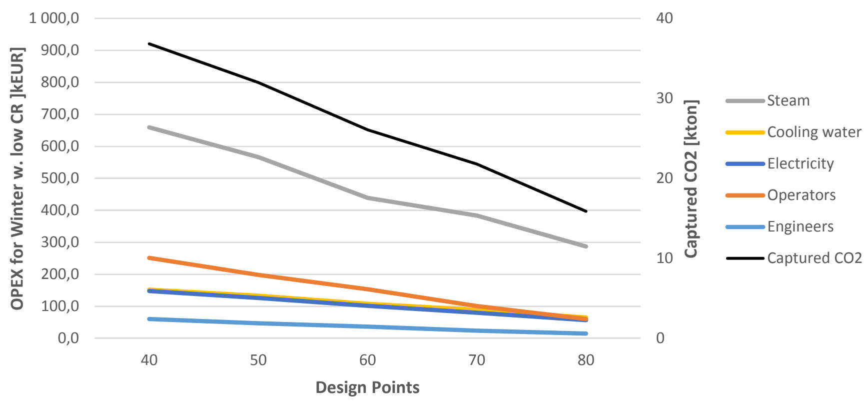

Figure 6.5: Variations of variable cost parameters for different design points during the winter period with a low capture rate. Please notice the large difference on the y-axis scale compared to Figure 6.3 and that the cost is for the whole winter period.

6.2.3 Cost for different design points

The total annual cost of capturing CO_2 consists of both fixed and variable costs, and in Figure 6.6 these costs are presented for the five different SRP design points, along with the generic reference design with a design point of 90 %. It should be highlighted that the operational and specific costs are calculated based on an annual capture rate of 40 %, meaning the 64.2 kton CO_2 originating from fossil sources. The CAPEX and fix OPEX figures were calculated through equation 4.1 with a scaling factor alpha equal to 0.343. As can be seen, the distribution of OPEX between summer and winter is changing with increasing design point, and as was presumed, the total OPEX cost is decreasing with higher design point. Unfortunately, despite the decreasing OPEX, the total annual cost is increasing with increased design point due to the increase of CAPEX and fix OPEX costs. This implies that a partial capture unit should be built as small as possible in order to minimize the capital cost, and thus minimizing the specific cost of capturing CO_2. The specific cost is shown as a black line above the bars in Figure 6.6, and the lowest specific cost is 86 EUR/ton CO_2 calculated for the lowest design point with a capture rate of 40 %. The 86 EUR/ton CO_2 is 10 % lower than the specific cost presented in Table 6.1 for the SRP_{40} model, and thus also lower than the specific cost for the SSP_{40} model with a specific cost of 94 EUR/ton CO_2. This is a result of operating the capture unit at maximum capture rate during the summer and then lowering the capture rate during the winter instead of letting the capture unit operate at constant load throughout the year. It is thus evident that even if the operation of the capture unit cannot compensate for the increasing capital costs, it can still lower the specific cost quite significantly.

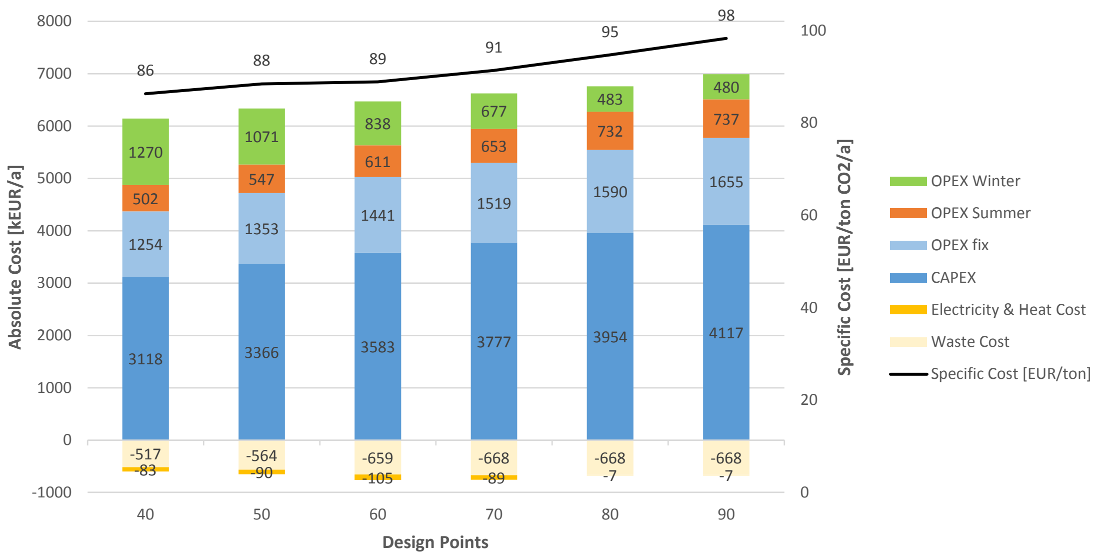

Figure 6.6: The distribution between fix and variable costs for the different SRP design points and the generic reference model on right side. The black line above the bars shows the specific cost in EUR/ton CO_2 for the different design points. Note that additional income for waste incineration and electricity production during the summer is included as a negative cost.

6.3 Sensitivity Analysis

Due to uncertainties in the cost estimations and the assumptions made, a sensitivity analysis was conducted in order to evaluate the robustness of the results. The first analysis is shown in Figure 6.7 where the capital and operational costs per captured ton of CO_2 is plotted separately with a variation of +/– 30 % for each design point of the SRP design. From the figure it is clear that even if the real cost would deviate significantly from the results in this work, the capital cost will always be considerably larger than the operational costs.

Another interesting aspect in the sensitivity analysis is the rate of change of the specific cost for CAPEX and OPEX, and which parameters that have the largest effect on them. The interval between the design points 40 % and 50 % was studied in more detail, and it was found that whilst the OPEX cost decreased with 0.33 EUR/ton/%-design point, the CAPEX cost increased with 0.54 EUR/ton/%-design point. The capital cost is thus not only significantly higher than the operational cost, it also increases more than the operational cost decreases. In order to evaluate the sensitivity in the rate of change, the interest rent and pay-back period were varied with +/– 25 %, see Table 6.2, and not even with an interest rent 25 % lower does the capital cost increase less than the operational cost decreases. The conclusion of this sensitivity analysis is thus that the result is robust even for large changes on important parameters affecting the specific cost.

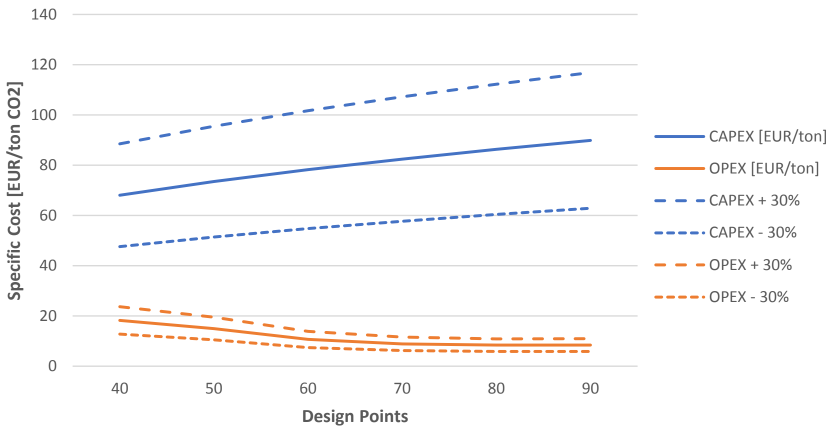

Figure 6.7: The specific OPEX and CAPEX costs for the different design points of the SRP design. Note that CAPEX also includes fix OPEX costs.

Table 6.2: The change of CAPEX due to changes in the interest rent and pay-back period. The numbers should be compared with the base value for CAPEX cost of 0.54 EUR/ton/%-design point and the value of – 0.33 EUR/ton/%-design point.

	Interest rate +25 %	Interest rate –25 %	Pay-back time +25 %	Pay-back time –25 %
Rate of change [EUR/ton/%-d.p.]	0.60	0.49	0.51	0.60

Yet another sensitivity analysis was performed by comparing the effect on the specific cost by varying parameters affecting either the capital cost or the operational cost. The result is shown in Table 6.3 where the four parameters have been varied with +/– 25 %, and the values below each of the parameters express how the total specific cost is affected. As is shown, the parameters related to the capital cost have a much larger effect on the resulting specific capture cost than the parameters related to the operational costs. The presented changes on the specific cost in Table 6.3 is an average value for the effects on the different design points of the SRP design.

Table 6.3: The resulting effects on the specific cost by varying four key parameters related to either the capital or operational costs with +/– 25 %.

	Interest rate	Pay-back time	Waste price	Heat Price
+25 %	+ 8.7 %	– 5.5 %	– 2.7 %	+ 1.5 %
– 25 %	– 8.4 %	+ 10 %	+ 2.7 %	– 1.5 %

6.4 Lillesjöverket

The results for the evaluation of partial capture from Lillesjöverket have been divided in three subsections; general results and Lillesjöverket with and without pellets production. The characteristics of the capture unit will first be described, followed by an evaluation of its impact on Lillesjöverket.

6.4.1 Lillesjöverket - General Results

Table 6.4 presents the economical performance of the capture unit designed for Lillesjöverket. The capture unit is a SRP design with a design point of 40 %, a decision that was made after the economic evaluation of the generic results that was presented in Figure 6.6. The values presented in Table 6.4 are based on a constant annual load, and even if the investment cost is 16 % lower than for the generic SRP_{40} design, the specific cost of capturing CO_2 is 20 % higher than the corresponding value for the generic SRP_{40} design.

Table 6.4: The economical result for Lillesjöverket. Note that the OPEX and specific cost is calculated for a constant capture rate and reboiler heat duty throughout the year and that additional income for waste incineration and electricity production is not included.

	Investment [kEUR]	CAPEX [kEUR/a]	fix OPEX [kEUR/a]	OPEX [kEUR/a]	Specific Cost [EUR/ton CO_2]	CO_2 Captured [ton/yr]
Lillesjöverket	26 304	2 617	1 052	1 428	118	43 200

The specific operational cost for Lillesjöverket is 33 EUR/ton CO_2 and is calculated by dividing the OPEX cost with the amount of CO_2 captured. This is 10 % more than the specific operational cost for the generic SRP_{40} design, but an even more interesting comparison is to compare the specific operational cost for CO_2 capture with the operational cost for the conventional and already existing flue gas treatment. The operational cost for the conventional flue gas treatment have been estimated in collaboration with engineers at Lillesjöverket to 11.7 EUR/ton CO_2. The cost of capturing CO_2 is thus three times higher, all-though it should be mentioned that these figures does not include capital costs and that the operational cost for capturing CO_2 is calculated for a constant annual load without including revenues from additional waste incineration and electricity production.

The operation range and off-design performance of the capture unit designed for Lillesjöverket is shown in Figure 6.8. The maximum capture rate is similar to that of the generic cases with an upper bound approximately 10 %-units higher than the design point. The lower capture rate limit is however even lower than for the generic cases, which may be positive if the steam extraction would have to be reduced.

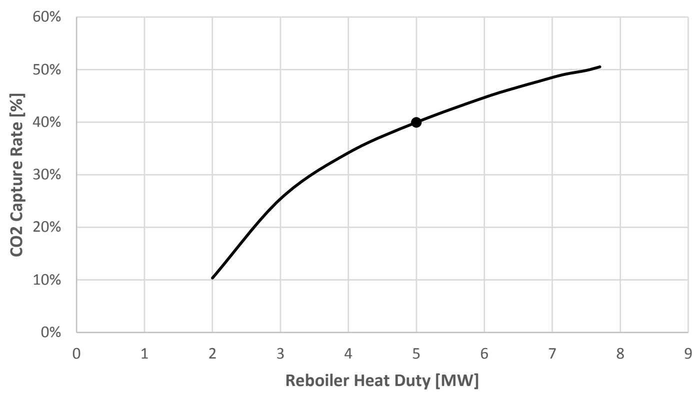

Figure 6.8: The capture rate as a function of reboiler heat duty for the capture unit designed for Lillesjöverket. The filled circle marks the design point.

6.4.2 Lillesjöverket excluding Pellets Production

In the previous section, the characteristics of the capture unit was described in Figure 6.8, and this section intend to describe how the capture unit affects Lillesjöverket. Two alternatives for steam extraction were identified; normal steam extraction that is visualized in Figure 5.4, and high pressure steam (HPS) extraction that is visualized in Figure 5.6. The extraction of steam to the capture unit has a negative effect on the production of electricity and heat, and Figure 6.9 shows how the electricity and heat production is affected for different reboiler duties. For the normal extraction, which is the solid line in Figure 6.9, it is easy to see that the heat production is penalized much harder than the electricity production, which is due to the fact that the majority of the electricity is produced in the high pressure turbine which is located up-stream of the extraction point. The HPS extraction was implemented since the heat production was considered as more important in the local energy system, since electricity could be provided from other sources, and as can be seen in Figure 6.9, the implementation has a significant impact on the electricity and heat production. The reference point is full generation for normal steam extraction, which is why the heat generation is more than 100 % for the HPS extraction with a reboiler heat duty of 2 MW.

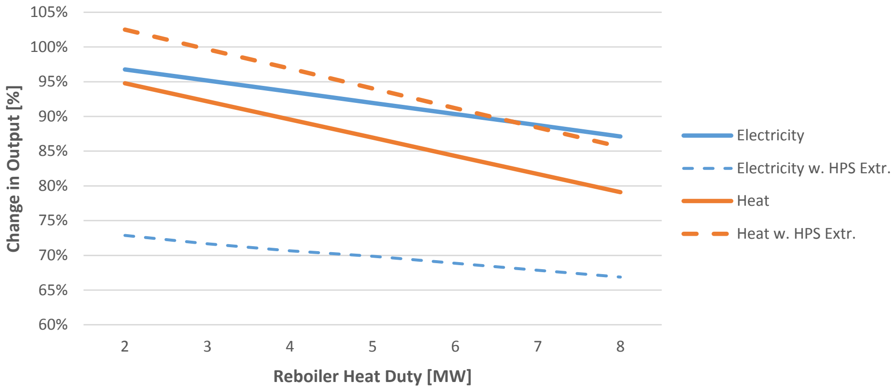

Figure 6.9: The change in electricity and heat output from Lillesjöverket at different reboiler heat duties for normal and high pressure steam (HPS) extraction.

The specific cost in Table 6.4 is calculated for a constant load of 5.02 MW in the reboiler and a capture rate of 40 %. If the operation of the capture unit is instead seasonally optimized, the reboiler would require 7.7 MW during the summer and 4.1 MW during the winter. The resulting maximum capacities for electricity and heat generation are displayed in Table 6.5 for both normal and HPS extraction. These values should be compared to the maximum output of electricity and heat of 10 MW and 32 MW, respectively, if no capture unit is present. The different alternatives for steam extraction give rise to the question of how the products should be prioritized. Variation in market price could change the order of priority, and thus change the production. This optimization problem becomes even more evident when more products streams are added, such as pellets which is discussed in section 6.4.3.

If the capture unit is operated according to the seasonal optimized operation described in the previous paragraph, the specific cost is decreased to 106 EUR/ton of CO_2, as is presented in Table 6.6. This is a 10 % decrease of the specific cost presented in Table 6.4. This is a result of an optimized use of cheap steam during the summer and the inclusion of income for the additional waste incineration and electricity production during the summer when the plant otherwise was assumed to decrease its furnace load to 70 % of its nominal value.

Table 6.5: Lillesjöverkets expected output of electricity and heat during summer and winter for normal and HPS extraction. Figures are presented in MW.

	Summer	Winter
Reboiler Duty	7.7	4.1
Normal extraction		
- Electricity	8.78	9.36
- Heat	25.56	28.57
HPS extraction		
- Electricity	6.73	7.07
- Heat	27.64	30.90

Table 6.6: Capital and operational costs related to the capture unit designed for Lillesjöverket.

CAPEX	2 617	kEUR
fix OPEX	1 052	kEUR
OPEX summer	361	kEUR
OPEX winter	990	kEUR
Waste	-365	kEUR
Electricity	-58	kEUR
Total	4 596	kEUR
Specific Cost	106	EUR/ton

6.4.3 Lillesjöverket including Pellets Production

The pellets production process has a maximum thermal capacity of 16 MW heat which is taken from the district heating system, see Figure 5.5. The left graph in Figure 6.10 shows the effects on electricity and heat production for different thermal loads in the pellets process during the summer, when there is reboiler heat duty in the capture unit of 7.7 MW. The right graph shows the corresponding results for the winter period, with a reboiler heat duty of 4.1 MW. These graphs can thus be used to determine either the pellets or heat production as a function of the other, since the electricity production is constant for all thermal loads in the pellets process.

The maximum heat output during the summer when the pellets process is operating at full capacity reaches only 11.83 MW, which is considerably lower than the average summer heat load of 18 MW shown in Figure 3.2. During the summer, the heat demand in the district heating is occasionally as low as 8-10 MW [19], but for the major part of the summer the power plant will not be able to supply the district heat demand and the pellets production, whereas two alternatives arises. Either could the pellets production be down prioritized and shifted to the winter in order to make more heat available for the district heating system, or another unit in the district heating system could be started to cover the remaining demand. This optimization problem has however not been within the scope of this work.

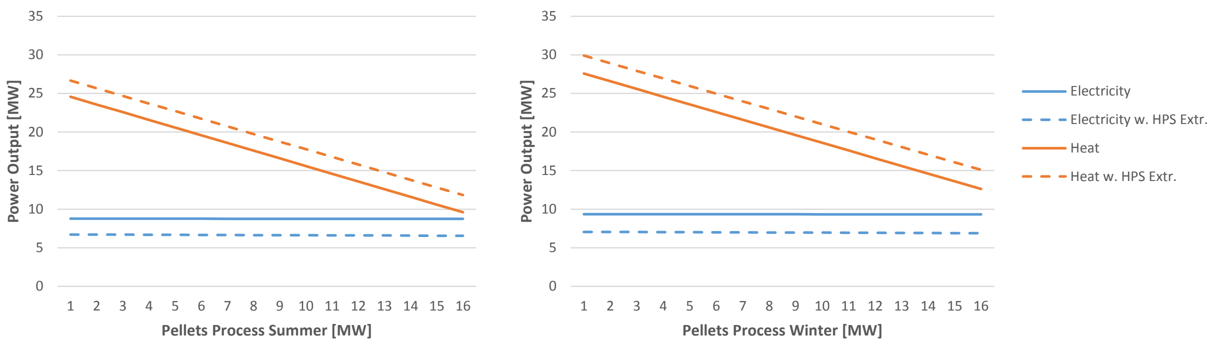

Figure 6.10: The change in electricity and heat output from Lillesjöverket during summer (left) and winter (right) at different loads in the pellets process for both normal and HPS extraction.

The efficiency of the pellets process is dependent on the ambient temperature, and during the summer the maximum production rate is 16 ton pellets per hour, whereas the maximum production rate only reaches 11 ton pellets per hour during the winter. The annual target amount is 44 000 tons of pellets, which is met if the process run at full load during the whole summer period:

$$16\ \text{MW} \times 1\ \text{ton/MW/h} \times 2750\ \text{h} = 44\ 000\ \text{ton}$$

If this cannot be full filled, pellets have to be produced during the winter, when both the process is less efficient and the price for heat is higher, leading to a higher production cost and lower profits.

6.5 Summary

This section summarizes and analyzes the results for two fundamentally different approaches, constant annual operation and seasonal optimized operation. The cost for constant load operation has been calculated for the three generic cases and for Lillesjöverket, which is a SRP_{40} design, whereas the cost for a seasonal optimized operation has only been calculated for the generic SRP design and Lillesjöverket.

6.5.1 Constant Annual Operation

The economical performance for the four cases used for a constant annual load is shown in Table 6.7. An interesting finding is the share of the total annual cost that is a fixed cost. The fixed annual cost includes CAPEX and fix OPEX, and is 69.4 % of the total annual cost for the generic SRP_{40} design, and 72.0 % for Lillesjöverket, which are the same design but for different flue gas flows. The flue gas flows are 31 kg/s and 25 kg/s for the generic SRP_{40} and Lillesjöverket, respectively, and from this it can be calculated that the fixed share of the annual cost decreases with 0.43 % for each kg/s that the flue gas flow is increased. The same pattern is found when comparing the reference design with the SSP_{40} design, where the fixed share is 59.9 % for the reference design, and 66.9 % for the SSP_{40} design. For this comparison the flue gas flows are 31 kg/s and 13,8 kg/s for the reference and SSP_{40} model, respectively, and here the share of fixed annual cost decreases with 0.41 % per kg/s of increased flue gas flow. The share of fixed cost is thus equally sensitive to changes in the flue gas flow for the two comparisons, and even if these figures cannot be linearized for all designs with different sizes, it consolidates the relationship between variable and fixed costs for designs in this size range.

Table 6.7: The economical result for the four designs. Note that the OPEX and specific cost is calculated for a constant capture rate and reboiler heat duty throughout the year and that the amount of captured CO_2 varies significantly between the designs.

	Investment [kEUR]	CAPEX [kEUR/a]	fix OPEX [kEUR/a]	OPEX [kEUR/a]	Specific Cost [EUR/ton CO_2]	CO_2 Captured [ton/yr]
Reference	41 389	4 117	1 656	3 872	67	144 500
SRP_{40}	31 342	3 118	1 254	1 929	98	64 200
SSP_{40}	29 071	2 892	1 163	2 004	94	64 200
Lillesjöverket	26 304	2 617	1 052	1 428	118	43 200

6.5.2 Seasonal Optimized Operation

Seasonal optimized operation implies that the operation of the capture unit is maximized during periods with low operation cost, such as periods with low steam price, and thus to minimize the operation at other more expensive periods. A prerequisite for this concept is that the capture unit have the ability to operate at off-design, which the SRP design have. The economic results of operating two SRP designs of different size is presented in Table 6.8, and the specific cost has decreased with 12 % and 10 % for the generic SRP design and Lillesjöverket, respectively, compared to the corresponding values for a constant annual load in Table 6.7. Below follows a deeper analysis of the specific cost divided in OPEX and CAPEX where CAPEX includes fix OPEX costs.

Table 6.8: A comparison between the generic SRP design and the SRP design for Lillesjöverket. Note that these results are based on a seasonal optimized operation.

	Investment [kEUR]	CAPEX [kEUR/a]	fix OPEX [kEUR/a]	OPEX [kEUR/a]	Specific Cost [EUR/ton CO_2]	CO_2 Captured [ton/yr]
Generic SRP_{40}	31 342	3 118	1 254	1 172	86	64 200
Lillesjöverket	26 304	2 617	1 052	928	106	43 200

Specific OPEX Comparison

Table 6.9 presents a comparison between the operational costs for the two SRP designs previously discussed in Table 6.8. All parameters are separated and the relative change in specific cost is presented in the last column. The largest relative change appears for the operators and engineers that both have an increase of 49 %. This is because the cost for operators and engineers are dependent on the operation time and not the amount of CO_2 captured. This means however that the specific cost related to operators and engineers will increase for smaller capture units if they are operated in the same way as a lager capture unit. The total change in specific OPEX costs is 18 %, or 3.2 EUR/ton CO_2, which is less than the increase in specific cost for operators and engineers. The increase in electricity cost is mainly caused by the compressor which is a reciprocating compressor, instead of an axial compressor, for Lillesjöverket due to the smaller amount of CO_2. The increase in specific cost for steam is due to the fact that Lillesjöverket has a lower concentration of CO_2 in the flue gas compared with the generic flue gas, and thus the absorption process becomes less efficient [25]. The decrease in cooling water is also due to the composition of the flue gas, which for Lillesjöverket contains less water and thus also require less cooling capacity.

Table 6.9: A comparison between the specific OPEX cost for two capture units of the same design but with different dimensions due to different flue gas flows. The unit for OPEX is kEUR/a and EUR/ton CO_2 for specific cost.

	OPEX Genric SRP	OPEX Lillesjöverket	Specific cost Generic SRP	Specific cost Lillesjöverket	Relative Change
Electricity	247	173	3.8	4.0	4 %
Cooling water	264	125	4.1	2.9	– 30 %
Steam	793	586	12.3	13.6	10 %
Operators	378	378	5.9	8.8	49 %
Engineers	90	90	1.4	2.1	49 %
Waste and elec.	– 600	– 423	– 9.3	– 9.8	5 %
Total	1 172	929	18.3	21.5	18 %

Specific CAPEX Comparison

A similar comparison was conducted for the capital cost and how the cost for different components changes between the two capture units. In Table 6.10 and 6.11 the equipment cost and installation factor for all components is given for the generic SRP design and Lillesjöverket, and the resulting specific cost is found in the last column. It can now be calculated that the OPEX share of the total specific cost is just above 20 % for both capture units, and they are thus similar in that aspect.

Table 6.10: Specified capital costs and installation factors for all components of the generic SRP design. All costs are in kEUR and specific cost in EUR/ton CO_2.

	Equip. cost	Installation fact.	Total cost	CAPEX	Specific cost
Pump & fan	168	11.56	1 943	193	3
Compressor	1 934	3.51	6 793	676	10.5
Absorber	1 760	3.53	6 217	618	9.6
Washer	366	6.18	2 263	225	3.5
Stripper	313	6.20	1 938	193	3.0
Flash	206	8.17	1 681	167	2.6
Tank	175	11.14	1 952	194	3.0
Heat exchanger	581	9.84	5 715	569	8.9
Reboiler	248	7.81	1 937	193	3.0
Filters	49	18.45	904	90	1.4
Total CAPEX			31 342	3 118	48.6
fix OPEX				1 254	19.5
Total					68.1

Table 6.11: Specified capital costs and installation factors for all components of the capture unit designed for Lillesjöverket. All costs are in kEUR and specific cost in EUR/ton CO_2.

	Equip. cost	Installation fact.	Total cost	CAPEX	Specific cost
Pump & fan	138	12.36	1 703	169	3.9
Compressor	1 459	3.70	5 398	537	12.4
Absorber	1 313	3.97	5 206	518	12.0
Washer	324	6.45	2 088	208	4.8
Stripper	269	6.66	1 792	178	4.1
Flash	178	8.46	1 506	150	3.5
Tank	175	11.14	1 952	194	4.5
Heat exchanger	367	11.11	4 072	405	9.4
Reboiler	203	8.19	1 665	166	3.8
Filters	49	18.45	904	90	2.1
Total CAPEX			26 287	2 615	60.5
fix OPEX				1 051	24.3
Total					84.9

The resulting relative change from the figures presented in Table 6.10-6.11 is presented in Table 6.12. As can be seen, the equipment cost have decreased for most components, except for tanks and filters which are the same for both capture units, which is expected since Lillesjöverket is a smaller plant. The decrease in equipment cost is however evened out with increased installation factors, which thus is one reason to why the specific cost is higher for the smaller capture unit. Furthermore, except for the large relative increase for tanks and filters, which should be revised if it really is the case, the only component that stands out is the heat exchangers which only have a relative increase of 6 % for the specific capital cost.

The installation factor includes the cost for all additional components such as pipes, measurement equipment and other supporting structures in the surrounding of the specified equipment. In the case of a pump for the generic SRP design, the installation factor is 11.56, which is multiplied with the equipment cost for the pump in order to calculate the total cost for the pump and the surrounding necessities, see Table 6.10. The cost for the surrounding necessities for a smaller pump, e.g. Lillesjöverket, is however more or less the same as for the generic design, but since the pump has a lower equipment cost, the installation factor is increased to 12.36 in order to generate the same cost for the surrounding necessities. The installation factor is also affected by type of equipment and its material. A component in a cheaper material comes with a lower cost, but since that does not affect the cost for surrounding requirements, the installation factor is increased. The total relative change in specific capital cost is 25 %, but if the installation factor would have been kept constant, the specific capital cost would only have increased with 16 %. The installation factor thus have an significant effect on the change in specific capital cost between the two cases.

Table 6.12: The comparison and change in equipment cost, installation factor and specific cost between the generic SRP design and the capture unit designed for Lillesjöverket. Note that the change is defined as how the cost for Lillesjöverket has changed compared to the generic case.

	Change in equipment cost	Change in installation factor	Relative change in specific cost
Pump & fan	– 18 %	7 %	30 %
Compressor	– 25 %	5 %	18 %
Absorber	– 25 %	12 %	24 %
Washer	– 12 %	4 %	37 %
Stripper	– 14 %	7 %	37 %
Flash	– 13 %	4 %	33 %
Tank	0 %	0 %	49 %
Heat exchanger	– 37 %	13 %	6 %
Reboiler	– 18 %	5 %	28 %
Filters	0 %	0 %	49 %
Total CAPEX			25 %
fix OPEX			25 %
Total			25 %

7 Conclusion

This thesis evaluates the implementation of partial CO_2 capture to Swedish waste fired CHP plants, with the aim to capture the fossil share of the CO_2 emissions in order to make the plants CO_2 neutral. The results of this work shows that implementation of partial CO_2 capture on Swedish waste fired CHP plants is technically possible. The extracted steam will lower the production of heat and electricity from the plant, but since the plants does not normally operate at full load during the summer, the effect on these product streams during this period is small or non-existing. The revenues during summer may in-fact increase with the implementation of a capture unit due to an increased rate of waste incineration, and, thus, an increased electricity production. The cost related to the CO_2 capture is, however, significant. There are, however, no other options for waste incineration to become CO2 neutral and the cost indicates what must be accepted by the plant owners for continued operation.

It was found that a seasonal optimized operation has a significant effect on the specific price for capturing CO_2, compared to a constant annual load operation. The main reason to this is the excess capacity during summer and the large difference in the price for heat between summer and winter. This difference in heat prices causes a large difference in the cost of extracting steam, and the operation of the capture unit should thus be minimized during periods with high heat prices.

From this work, it can also be concluded that Swedish CHP plants are relatively small CO_2 emission sources, and that this has an important impact on the cost of the capture units according to the economy-of-scale concept. The CAPEX share of the total specific cost is over 80 %, and it has been shown that the capital cost increases faster than the operational cost decreases when larger capture units are used to capture a fixed amount of CO_2. It can thus be established that for CO_2 sources of this size, the capture unit should be optimized for low investment costs rather than low operation costs. Due to the size of these plants, the cost for transportation of the CO_2 may be considerable, and it would be beneficial if other CO_2 sources existed in the vicinity of the plant in order to reduce the transportation cost.

The large dependence of capital costs is also visible in the sensitivity analysis that was conducted. Changes to parameters related to the capital costs have a much larger effect on the specific capture cost than changes done to parameters related to the operational cost. Furthermore, the type of cost estimation applied in this work is normally considered to have an accuracy of $+/-$ 30 %, and even if the result is varied within this range, the same conclusions can be made. The result is thus considered as robust and not on the edge of pointing towards a different conclusion.

The case study of Lillesjöverket showed that the plant may capture its fossil share of CO_2 on an annual basis and still deliver between 89-97 % of its nominal heat production during winter. However, it showed difficult to combine CCS with other uses of the heat like pellets production. The capture unit was assumed to be required for operation, and the remaining products that have to be prioritized is then heat, pellets and electricity. Heat is today often prioritized by CHP plants because it is limited by its local distribution and cannot be replaced by other remote heat sources. The price characteristics for these products would however be of great significance for the results of solving such an optimization problem, to which also other units in the district heating system could be added. The conclusion of the case study of Lillesjöverket is that the implementation of a capture unit creates optimization problems in the local district heating system, and that a systems perspective is required early in the planing phase.

7.1 Future Work

An idea for future work is to merge the two design approaches and create a hybrid between the slip stream path and the separation rate path in order to see if there is any combination that could perform better than the results in this work. Other aspects that influence the capital and operational costs such as stream splitting configurations and over-sizing of reboiler heat input or other parameters could also be of interest to investigate further. It would also be interesting to see how a power plant should be designed if the capture unit was to be included all ready in the design phase instead of retro-fit, and which capture unit design then would be the best choice.

The system perspective mentioned previously could also be an interesting topic for future work. Different CO_2 tax scenarios could be included in order to evaluate the effect of an CO_2 tax, or potentially determine at which level such a tax should be set. Moreover, it could also be interesting to include the transportation and storage, which was omitted in this work, in order to get an overview of the practical potential for applying carbon capture and storage to Swedish waste fired CHP plants.

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A Appendix 1 - Capture Unit Dimensions

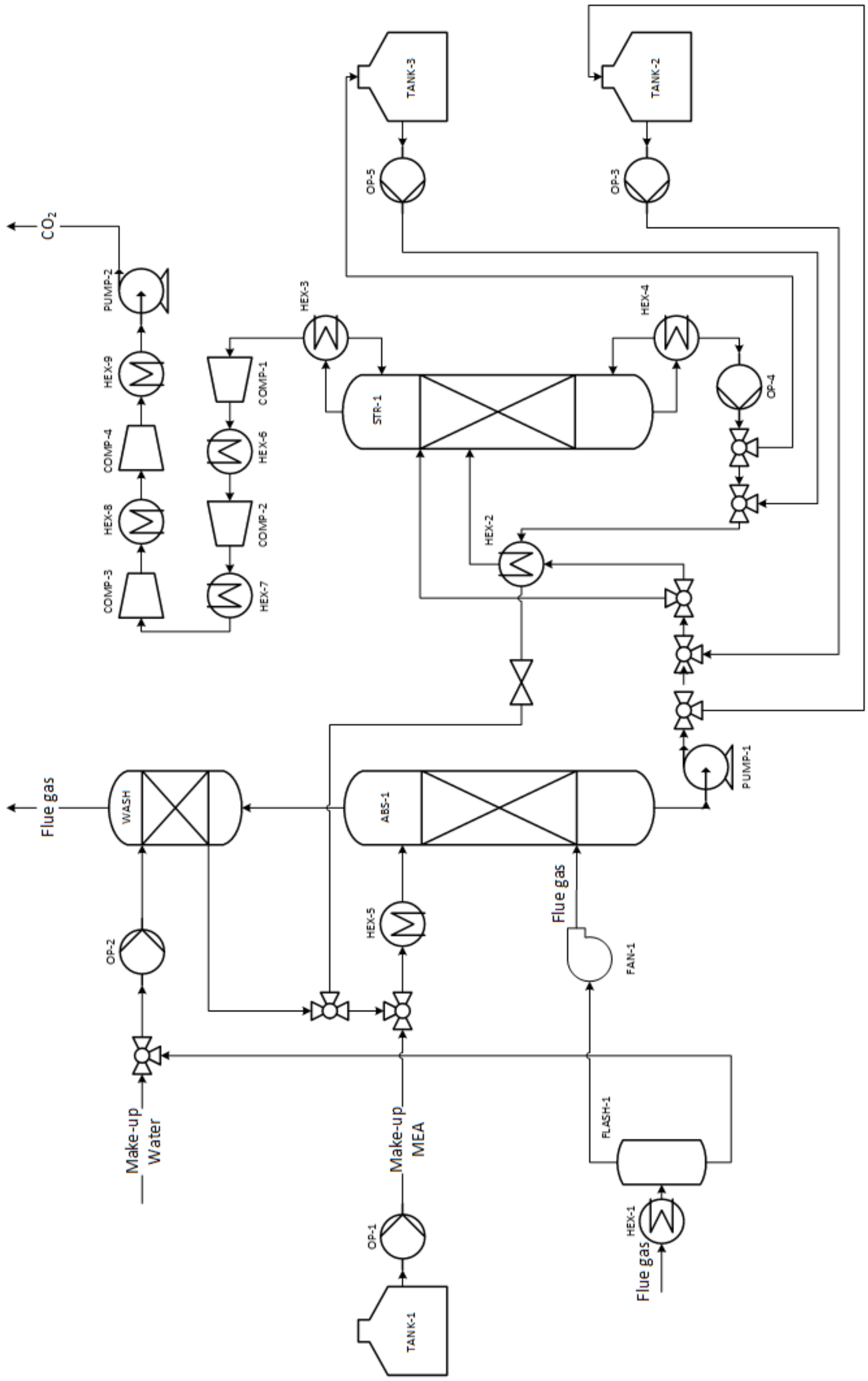

Figure A.1: Schematic overview of the capture unit.

Reference Design

Table A.1: Equipment list presenting all units and characteristic dimensions for the reference model.

Column	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
ABS-1	Sulzer Mellapak 250Y	1.0	40 - 47.4	25 (35.7)	4.24
STR-1	Sulzer Mellapak 250Y	2.0	20 - 121	12 (17.1)	2.36
WASH	Sulzer Mellapak 250Y	1.0	40 - 59	3.1 (4.47)	4.24

Heat Exch.	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Area [m^2]
HEX-1	Shell and Tube	1.0 - 1.0	10 - 50	2 839	515
HEX-2	Shell and Tube	2.0 - 2.0	47 - 121	22 355	1 696
HEX-3	Shell and Tube	1.0 - 1.0	10 - 98	4 570	184
HEX-4	Reboiler	2.0 - 2.0	120 - 133	17 405	2 046
HEX-5	Shell and Tube	1.0 - 1.0	10 - 57	5 940	246
HEX-6	Shell and Tube	1.0 - 5.0	10 - 131	483	58
HEX-7	Shell and Tube	1.0 - 8.9	10 - 138	556	64
HEX-8	Shell and Tube	1.0 - 26.8	10 - 139	572	65
HEX-9	Shell and Tube	1.0 - 80	10 - 140	1 483	168

Flash	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
FLASH-1	Vertical	1.0	40 - 40	6.57	4.06

Compressor	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
COMP-1	Axial	1.0 - 3.0	20 - 131	489	2.72
COMP-2	Axial	3.0 - 8.9	25 - 138	492	0.92
COMP-3	Axial	8.9 - 26.8	25 - 139	476	0.29
COMP-4	Axial	26.8 - 80	25 - 140	434	0.09

Pump	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Pump-1	Centrifugal	1.0 - 2.0	62 - 62	17	0.097
Pump-2	Centrifugal	80 - 110	25 - 32	36.6	0.007
OP-1	Centrifugal	1.0 - 5.6	20 - 20	58	-
OP-2	Centrifugal	1.0 - 5.6	20 - 20	58	-
OP-3	Centrifugal	2.0 - 6.6	62 - 62	58	-
OP-4	Centrifugal	2.0 - 6.6	120 - 120	58	-
OP-5	Centrifugal	2.0 - 6.6	120 - 120	58	-

Flue gas fan	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Fan-1		1.0 - 1.04	50 - 54	129.4	27.8

Tank	Type	Pressure [bar]	Temperature [°C]	Volume [m^3]	
TANK-1	Make-up MEA	1.0	40 - 40	10	
TANK-2	Buffer tank	2.0	59 - 59	10	
TANK-3	Buffer tank	2.0	120 - 120	10	

Slip Stream Path

Table A.2: Equipment list presenting all units and characteristic dimensions for the slip stream path design.

Column	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
ABS-1	Sulzer Mellapak 250Y	1.0	40 - 50.3	25 (35.7)	2.91
STR-1	Sulzer Mellapak 250Y	2.0	20 - 120	12 (17.1)	1.61
WASH	Sulzer Mellapak 250Y	1.0	40 - 59	2.22 (3.17)	2.91

Heat Exch.	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Area [m^2]
HEX-1	Shell and Tube	1.0 - 1.0	10 - 50	1 261	229
HEX-2	Shell and Tube	2.0 - 2.0	50 - 120	11 016	872
HEX-3	Shell and Tube	1.0 - 1.0	10 - 94	1 681	70
HEX-4	Reboiler	2.0 - 2.0	120 - 133	7 709	852
HEX-5	Shell and Tube	1.0 - 1.0	10 - 60	3 643	145
HEX-6	Shell and Tube	1.0 - 5.0	10 - 131	215	26
HEX-7	Shell and Tube	1.0 - 8.9	10 - 138	247	28
HEX-8	Shell and Tube	1.0 - 26.8	10 - 139	254	29
HEX-9	Shell and Tube	1.0 - 80	10 - 140	660	75

Flash	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
FLASH-1	Vertical	1.0	40 - 40	4.54	2.71

Compressor	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
COMP-1	Axial	1.0 - 3.0	20 - 131	218	1.21
COMP-2	Axial	3.0 - 8.9	25 - 138	219	0.41
COMP-3	Axial	8.9 - 26.8	25 - 139	212	0.13
COMP-4	Axial	26.8 - 80	25 - 140	193	0.039

Pump	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Pump-1	Centrifugal	1.0 - 2.0	50 - 50	6.7	0.051
Pump-2	Centrifugal	80 - 110	25 - 32	13.9	0.003
OP-1	Centrifugal	1.0 - 5.5	20 - 20	30	-
OP-2	Centrifugal	1.0 - 5.5	20 - 20	30	-
OP-3	Centrifugal	2.0 - 6.5	50 - 50	30	-
OP-4	Centrifugal	2.0 - 6.5	120 - 120	30	-
OP-5	Centrifugal	2.0 - 6.5	120 - 120	30	-

Flue gas fan	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Fan-1		1.0 - 1.04	50 - 54	57.5	12.2

Tank	Type	Pressure [bar]	Temperature [°C]	Volume [m^3]
TANK-1	Make-up MEA	1.0	40 - 40	10
TANK-2	Buffer tank	2.0	59 - 59	10
TANK-3	Buffer tank	2.0	120 - 120	10

Separation Rate Path

Table A.3: Equipment list presenting all units and characteristic dimensions for the separation rate path design.

Column	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
ABS-1	Sulzer Mellapak 250Y	1.0	40 - 42	25 (35.7)	3.61
STR-1	Sulzer Mellapak 250Y	2.0	20 - 121	12 (17.1)	1.57
WASH	Sulzer Mellapak 250Y	1.0	40 - 59	2.22 (3.17)	3.61

Heat Exch.	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Area [m^2]
HEX-1	Shell and Tube	1.0 - 1.0	10 - 50	2 839	515
HEX-2	Shell and Tube	2.0 - 2.0	42 - 121	11 364	880
HEX-3	Shell and Tube	1.0 - 1.0	10 - 92	1 433	61
HEX-4	Reboiler	2.0 - 2.0	119 - 133	7 160	797
HEX-5	Shell and Tube	1.0 - 1.0	10 - 51	1 878	84
HEX-6	Shell and Tube	1.0 - 5.0	10 - 131	215	26
HEX-7	Shell and Tube	1.0 - 8.9	10 - 138	247	28
HEX-8	Shell and Tube	1.0 - 26.8	10 - 139	254	29
HEX-9	Shell and Tube	1.0 - 80	10 - 140	660	75

Flash	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
FLASH-1	Vertical	1.0	40 - 40	6.57	4.06

Compressor	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
COMP-1	Axial	1.0 - 3.0	20 - 131	218	1.21
COMP-2	Axial	3.0 - 8.9	25 - 138	219	0.41
COMP-3	Axial	8.9 - 26.8	25 - 139	212	0.13
COMP-4	Axial	26.8 - 80	25 - 140	193	0.04

Pump	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Pump-1	Centrifugal	1.0 - 2.0	42 - 42	6	0.046
Pump-2	Centrifugal	80 - 110	25 - 32	14	0.003
OP-1	Centrifugal	1.0 - 5.4	20 - 20	27	-
OP-2	Centrifugal	1.0 - 5.4	20 - 20	27	-
OP-3	Centrifugal	2.0 - 6.4	62 - 62	27	-
OP-4	Centrifugal	2.0 - 6.4	120 - 120	27	-
OP-5	Centrifugal	2.0 - 6.4	120 - 120	27	-

Flue gas fan	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Fan-1		1.0 - 1.04	50 - 54	129.4	27.8

Tank	Type	Pressure [bar]	Temperature [°C]	Volume [m^3]
TANK-1	Make-up MEA	1.0	40 - 40	10
TANK-2	Buffer tank	2.0	59 - 59	10
TANK-3	Buffer tank	2.0	120 - 120	10

Lillesjöverket

Table A.4: Equipment list presenting all units and characteristic dimensions for the capture unit designed for Lillesjöverket.

Column	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
ABS-1	Sulzer Mellapak 250Y	1.0	40 - 41	25 (35.7)	3.28
STR-1	Sulzer Mellapak 250Y	2.0	20 - 121	12 (17.1)	1.31
WASH	Sulzer Mellapak 250Y	1.0	40 - 59	2.16 (3.09)	3.28

Heat Exch.	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Area [m^2]
HEX-1	Shell and Tube	1.0 - 1.0	10 - 50	261	47
HEX-2	Shell and Tube	2.0 - 2.0	41 - 120	8 119	623
HEX-3	Shell and Tube	1.0 - 1.0	10 - 93	1 068	45
HEX-4	Reboiler	2.0 - 2.0	119 - 133	5018	558
HEX-5	Shell and Tube	1.0 - 1.0	10 - 50	1 266	57
HEX-6	Shell and Tube	1.0 - 5.0	10 - 131	144	17
HEX-7	Shell and Tube	1.0 - 8.9	10 - 138	166	19
HEX-8	Shell and Tube	1.0 - 26.8	10 - 139	171	19
HEX-9	Shell and Tube	1.0 - 80	10 - 140	444	50

Flash	Type	Pressure [bar]	Temperature [°C]	Height [m]	Diameter [m]
FLASH-1	Vertical	1.0	40 - 40	6.04	3.76

Compressor	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
COMP-1	Axial	1.0 - 3.0	20 - 131	147	0.82
COMP-2	Axial	3.0 - 8.9	25 - 138	147	0.27
COMP-3	Axial	8.9 - 26.8	25 - 139	143	0.09
COMP-4	Axial	26.8 - 80	25 - 140	130	0.03

Pump	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Pump-1	Centrifugal	1.0 - 2.0	41 - 41	4.5	0.032
Pump-2	Centrifugal	80 - 110	25 - 32	9.3	0.002
OP-1	Centrifugal	1.0 - 5.4	20 - 20	19	-
OP-2	Centrifugal	1.0 - 5.4	20 - 20	19	-
OP-3	Centrifugal	2.0 - 6.4	50 - 50	19	-
OP-4	Centrifugal	2.0 - 6.4	120 - 120	19	-
OP-5	Centrifugal	2.0 - 6.4	120 - 120	19	-

Flue gas fan	Type	Pressure [bar]	Temperature [°C]	Capacity [kW]	Flow [m^3/s
Fan-1		1.0 - 1.04	50 - 54	103.4	21.9

Tank	Type	Pressure [bar]	Temperature [°C]	Volume [m^3]
TANK-1	Make-up MEA	1.0	40 - 40	10
TANK-2	Buffer tank	2.0	59 - 59	10
TANK-3	Buffer tank	2.0	120 - 120	10

B Appendix 2 - Scaling Factor Calculation

Equation 4.1 was solved for the parameter alpha, α, by using the CAPEX cost for the full capture reference design with a capture rate of 90 % and the SRP_{40}, which were both calculated by TelTek.

$$CAPEX_{SRP,i} = CAPEX_{SRP,40}\left(\frac{CR_{SRP,i}}{CR_{SRP,40}}\right)^{\alpha} \quad [EUR]$$

$$\alpha = \frac{LOG\left(\frac{CAPEX_{SRP,90}}{CAPEX_{SRP,40}}\right)}{LOG\left(\frac{CR_{90}}{CR_{40}}\right)} = \frac{LOG\left(\frac{4117}{3118}\right)}{LOG\left(\frac{90}{40}\right)} = 0.343$$

C Appendix 3 - Data from Lillesjöverket

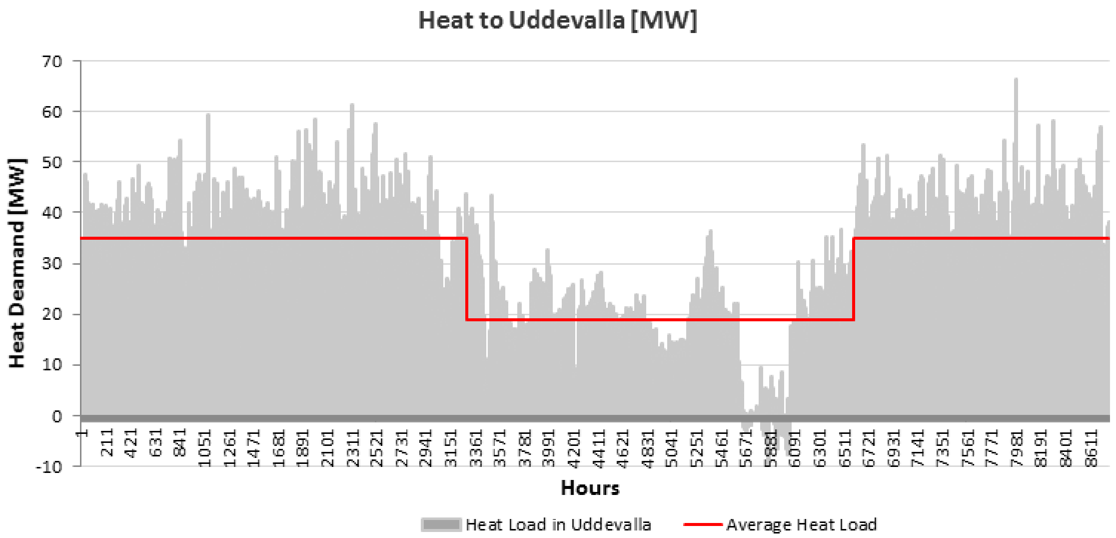

Figure C.1: The measured heat load in the district heating system in Uddevalla in 2016. The red line presents the average load during summer and winter.

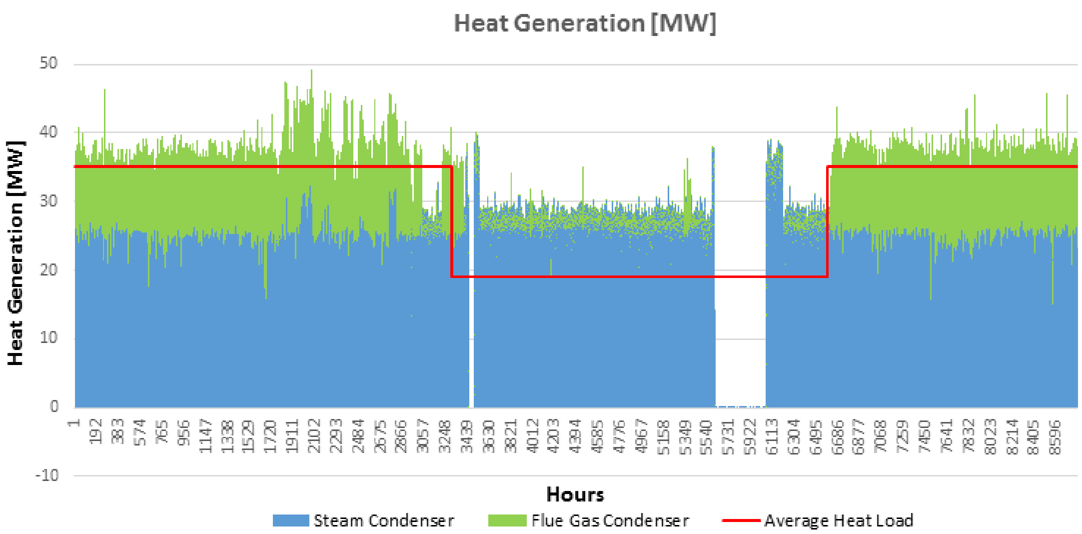

Figure C.2: The heat generated from the steam and flue gas condensers at Lillesjöverket during 2016. The red line represents the average heat load during summer and winter.