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Conclusion

A pulverized coal (PC) plant was found to be the best fit for a new power plant on Svalbard. The technology is commercially available, and no research and development is required. A maximum boiler temperature of 800°C was assumed, together with subcritical pressure in the steam cycle. District heating from a backpressure steam turbine was found to be a better option than a central heat pump, both practically and economically.

Abstract

Different types of coal-fired power plants were considered as options for a new power plant at Svalbard. Conventional technology was found to be the best fit, and a pulverized coal plant was modeled in detail. As the current plant does not have any flue gas treatment, the new plant was designed to handle CO_2, sulfur, NO_x, dust particle and mercury emissions. In a literature search, a seawater scrubber and amine solution carbon capture were found to be suitable for this task.

The plant was modeled in Aspen HYSYS according to design basis and data given by Longyearbyen Bydrift. Four cases were considered and studied in detail. The base case generates electric power from three steam levels, and utilizes the existing district heating network in Longyearbyen to use remaining heat from the steam cycle. In the heat pump case, electric power is generated from two steam levels, fully condensing the steam. It was assumed that the power could be used in a central heat pump or in consumer bought heat pumps, consuming the power more efficiently. The last two cases consider how increasing the steam pressure or temperature affects the plants thermal efficiency.

Economic analysis was performed on all major equipment, using order-of-magnitude scaling and the factorial method. Variable costs, revenues and working capital were estimated together with capital costs to perform investment analysis on the investment.

By analyzing case study data from Aspen HYSYS it was found that the base case is preferable over the heat pump case, both in efficiency and in economic perspective. The case studies on steam temperature and pressure confirmed that higher values will give a rise in thermal efficiency.

Further research is recommended on optimizing the steam cycle, as number of steam levels, steam pressure and temperature highly affect the thermal efficiency. Research and development is recommended on amine solution carbon capture, as the expense of carbon capture and storage is the economic bottleneck of the project.

Preface

This project was written as a part of the M.Sc. degree in Chemical Engineering at the Norwegian University of Science and Technology in the course TKP4170 Process Design Project.

We would like to thank our supervisor, Professor Sigurd Skogestad for his invaluable help, insight and motivation throughout the project period.
We are deeply grateful, and his support has helped us throughout the whole project, providing knowledge we will carry on in our degree.

Declaration of Compliance

We hereby declare that this is an independent report according to the exam regulations of the Norwegian University of Science and Technology.

Trondheim, November 21, 2013

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1 Design Basis

The design for the new plant was based upon the current plant data. It was assumed that the electrical energy demand would double, and that the need for district heating would increase with 50 %. A summary of the data from the current plant and the design basis for the new plant are listed in Table 1.1.

Table 1.1: Plant data for the current plant and design basis for the new plant.

	Current plant	New plant
Electrical energy	4.8 MW	9.6 MW
Thermal energy (district heating)	8.0 MW	12 MW
Coal	25000 ton/year	60000 ton/year (calculated)
Diesel	390 000 liters/year	307 000 liters/year (average of the last four years)

The simplifications and assumptions that are made are listed below.

- The coal is assumed to be pure carbon with the same net calorific value as the coal on Svalbard (~7300 kcal/kg), shown in Figure D.1 in Appendix C .
- The boiler is modeled as a combustion reactor with a maximum temperature of 800°C, followed by a heat exchanger.
- The boiler is assumed to combust the coal completely.
- The boiler is assumed to have a minimum temperature approach of 10°C, this is obtained by adjusting the flow of water in the steam cycle.
- The flue gas desulfurization is not included in the model.
- The carbon capture facility is modeled as a pure component splitter with a given heat requirement for re-boiling the rich amine solution in the stripper column.
- The HP steam temperature and pressure are set to 600°C and 165 bar respectively.
- The outlet pressure of the HP steam turbine is set to 49 bar, and is reheated to 600°C.
- The vacuum pressure created by the condenser is set to 0.01 bar.
- The seawater is assumed to be 4°C.
- The district heating network is modeled as heat exchangers with an inlet temperature of 120°C, an outlet temperature of 80°C and a pressure drop of 5 bar. This is obtained by adjusting the flow of water in the district heating network.
- The compressor train for the compression of carbon dioxide for storage is modeled as two compressors and a pump with intercooling using heat exchanger with cold seawater.
- The inter-stage pressures for the CO_2 compressor train are found by trial and error to yield the lowest amount of work needed.
- The adiabatic efficiencies of all the compressors and pumps are assumed to be 85%.
- The split ratio between the LP steam turbine and the IP steam turbine are chosen to yield a total district heating of 12 MW.
- The amount of air blown into the boiler is chosen to yield a maximum combustion temperature of 800°C.

- The amount of heat needed to the reboiler of the stripper is based on a 90% CO_2 capture, and an amount of heat needed per kilo CO_2 removed $\left(3242\ \frac{\text{kJ}}{\text{kg}\ CO_2}\right)$.
- The amount of electricity needed for the heat pump is found by an estimate for the heat pumps coefficient of performance $\left(3\ \frac{\text{MW of heat}}{\text{MW of electricity}}\right)$ [1], and a basis of 12 MW for district heating.
- The price of electricity was assumed to be 1 NOK/kWh.
- The price of district heating was assumed to be 0.5 NOK/kWh.
- Constant yearly costs and revenues.
- Constant depreciation rate of 10%.
- Constant amount of depreciation of 20%.
- 0% tax on Svalbard.

2 Introduction to Coal-Fired Power Plants

For more than 100 years, coal-fired power plants have generated the major portion of the worldwide electric power [2] with a current (2011) market supply share of 41.2% [3]. Coal is the largest growing source of primary energy worldwide, despite the decline in demand among the OECD countries, due to China's high increase in demand [4]. The Chinese coal consumption and production account for more than 45% of both global totals, and it has been estimated that their share will pass 50% by 2014 because of their high demand for cheap energy [4]. This will drastically increase the world total CO_2-production which will contribute greatly to the global warming and other environmental effects such as ocean acidification [5]. It will therefore be of great importance to develop clean and efficient coal plants which can produce electricity that can compete with the prices of the cheap, polluting coal plants that currently exists. Some instances of such plants have been proposed as alternatives to the conventional coal-fired power plant and they will be given an introduction in this report.

2.1 Conventional Coal-Fired Power Plants

Conventional coal-fired power plants use pulverized coal (PC) or crushed coal and air as a fuel to the furnace. The coal is pulverized by crushing and fed to the reactor at ambient pressures and temperatures and burned in excess of air. The excess of air is introduced to lower the furnace temperature which makes the equipment cheaper as it does not have to withstand extreme temperatures, and it also reduces the formation of NO_x. NO_x is formed at high temperatures and is a pollutant that has a negative effect on the health of humans besides contributing to acidic precipitation [6]. The hot flue gas from the furnace is used to heat up the boiler which produces high pressure (HP) steam. This steam is in turn expanded in a turbine arrangement that generate electrical power. The low pressure (LP) steam is then condensed and re-fed to the boiler. The hot flue gas contains pollutants and aerosols which have to be removed before the gas is vented through the stack to the atmosphere. Pollutants that have to be removed include mercury, NO_x and SO_2. The nitrous oxides are usually removed using selective catalytic reduction (SCR) where ammonia is used as a reducing agent [7]. The sulfur, mercury and other solid matter is normally removed as solid matter by reducing the sulfurous oxide using lime and water, and then passing the flue gas through an electrostatic precipitator or a fabric filter. The slurry is then collected for safe deposition. Conventional coal plants operating using subcritical (sC) conditions, which will result in low overall plant efficiency [8]. A conceptual process flow diagram of this power plant is shown in Figure 2.1.

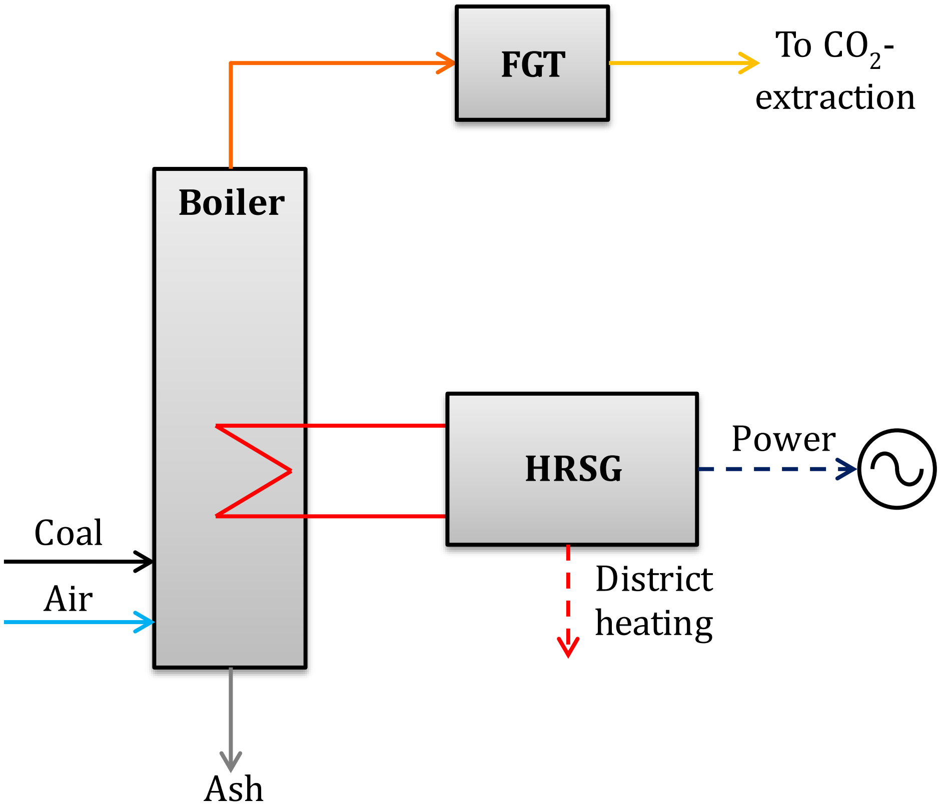

Figure 2.1: Simplified process flow diagram for the conventional coal plant with district heating.
HRSG = heat recovery steam generator.
FGT = flue gas treatment (desulfurization, mercury removal, dust removal etc.)

2.1.1 Supercritical Coal Fired Power Plants

The efficiency of the plant can be increase by using supercritical (SC) steam conditions with higher pressure. The plant efficiency is increasing both for increasing pressure drop and increasing temperature. There is therefore a constant development of better equipment that can withstand higher steam pressures and temperatures [8]. Some examples of conditions are listed in

Table 2.1. The ultra supercritical configuration is currently under development and is expected to be available in 2015 [8]. A typical heat recovery steam generator design is shown in Figure 2.2.

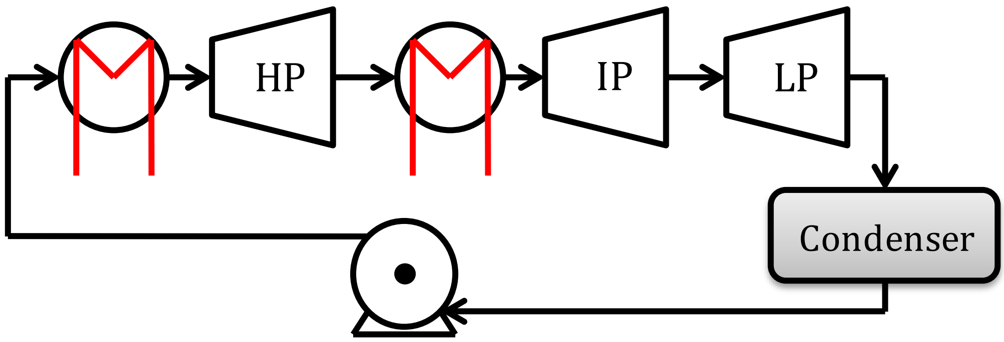

Figure 2.2: Heat recovery steam generator cycle with three pressure levels, HP, IP and LP.
HP = high pressure.
IP = intermediate pressure.
LP = low pressure.

Table 2.1: Some typical HP steam conditions [8]

	Temperature [°C]	**Pressure** [bar]
Depleted	< 500	< 115
Subcritical (sC)	500-600	115-170
Supercritical (SC)	500-600	230-265
Ultra supercritical	~730	~345

2.2 Integrated Gasification Combined Cycle (IGCC) Coal-Fired Power Plants

IGCC power plants feed compressed oxygen and a slurry of coal and water to a gasifier. The gasifier converts the fuel to synthesis gas (syngas) which is then treated to remove sulfur, mercury and aerosols. The syngas is then brought to a combustor with compressed air diluted with nitrogen in a turbine. The flue gas is then used to create steam by passing it through an HRSG. This steam is passed through a series of turbines, as with the conventional plant. The efficiency gain this method has compared to the conventional plant is that the combustor turbine operates at a very high temperature (~1500°C), but it also has to have an air separation unit (ASU) to achieve reasonable conversion rates for the gasification process [9]. The IGCC power plants require large investments because of all the advanced utilities such as a fluidized bed reactor for gasification and an air separation unit. The IGCC power plants can achieve up to 3% higher efficiencies which can be worth the investment in the long run, especially for huge power plants [8].

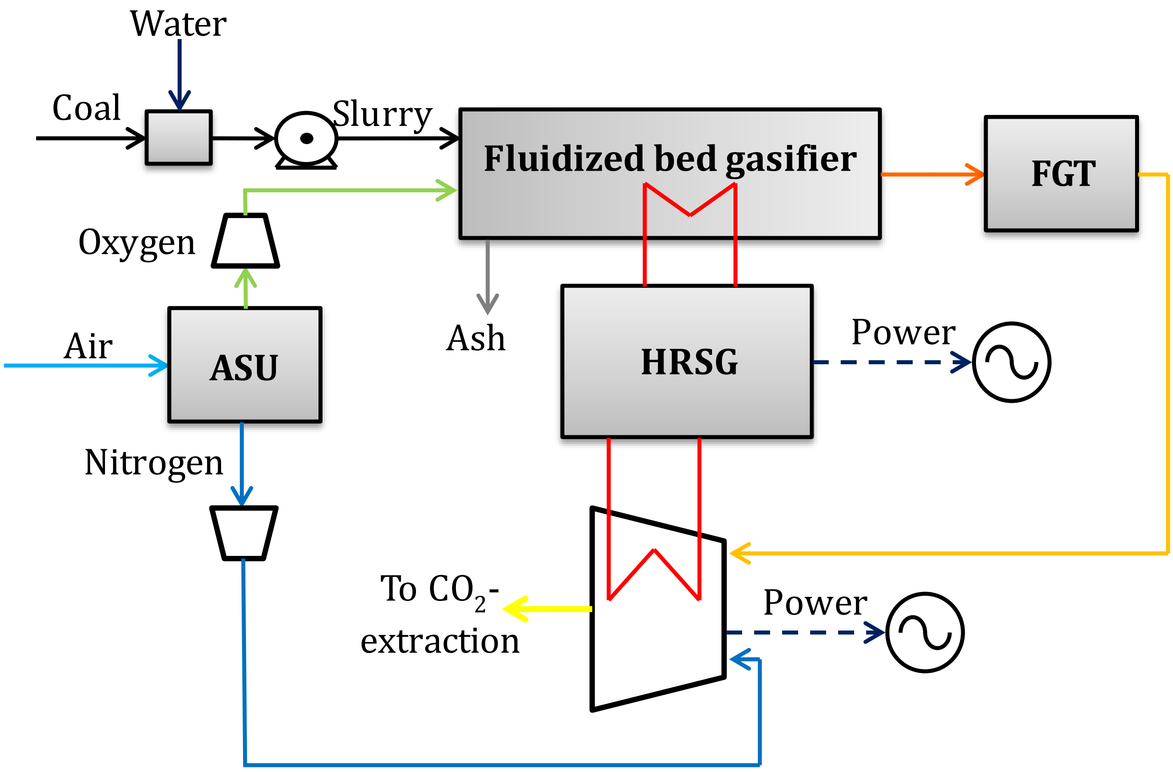

Figure 2.3: Simplified process flow diagram for the IGCC power plant.

2.3 Oxygen-Fired Coal Combustion Power Plants (Chemical Looping Combustion)

Oxygen-fired coal combustion power plants, also known as chemical looping combustion, burn PC with pure oxygen which creates a flue gas that has a very high carbon dioxide concentration. This has the advantage that the flue gas can be injected directly into storage after desulfurization, and cleaning. This technology is currently under development and several pilot plants have been built [10]. Unlike the other power plant designs, this design does not suffer a significant loss in efficiency when carbon capture and storage (CCS) is implemented. For a conventional power plant the loss in efficiency can be up to 14%, while the oxygen-fired power plant only suffers losses of around 3% [8] [11]. Another advantage is that there will not be any formation of nitrous oxides due to the lack of nitrogen in the feed, however the concentration of sulfur oxide will increase due to the flue gas recycle. This is on the other hand not seen as a major problem as sulfur oxide can be treated by introducing lime in the reactor. However this technology is currently not available commercially.

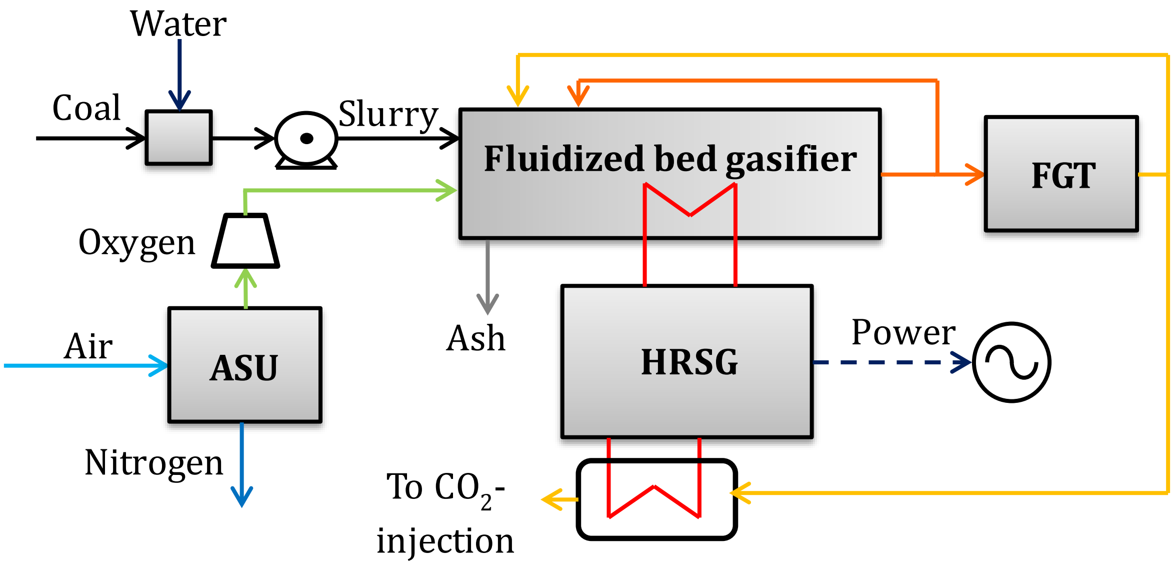

Figure 2.4: Simplified process flow diagram for the oxygen-fired coal combustion power plant.

3 Introduction to Flue Gas Treatment

3.1 CO_2 Capture

Energy supply from fossil fuels is associated with large emissions of CO_2 and account for 75% of the total CO_2 emissions. CO_2 emissions will have to be cut by 50% to 85% to achieve the goal of restricting average global temperature increase to the range of 2°C to 2.4°C [5]. Industry and power generation have the potential to reduce the emission of greenhouse gases by 19% by 2050, by applying carbon capture and storage [12]. There are three basic systems for CO_2 capture.

- Post-combustion capture
- Pre-combustion capture
- Oxygen fuelled combustion capture

3.1.1 Post-Combustion Capture

CO_2 captured from flue gases produced by combustion of fossil fuel or biomass and air is commonly referred to as post-combustion. The flue gases are passed through a separator where CO_2 is separated from the flue gases. There are several technologies available for post-combustion carbon capture from the flue gases, usually by using a solvent or membrane. The process that looks most promising with current technologies is the absorption process based on amine solvents. It has a relatively high capture efficiency, a high selectivity of CO_2 and the lowest energy use and cost in comparison with other technologies. In absorption processes, CO_2 is captured using the reversible nature of chemical reactions of an aqueous alkali solution. Amine solutions are most common for carbon capture. After cooling the flue gas it is brought into contact with solvent in an absorber at temperatures of 40°C to 60°C. The regeneration of solvent is carried out by heating in a stripper at elevated temperatures of 100°C to 140°C. This requires a lot of heat from the process, and is the main reason why CO_2 capture is expensive [13].

Membrane processes are used for CO_2 capture at high pressure and higher concentration of carbon dioxide. Therefore, membrane processes require compression of the flue gases; as a consequence this is not a feasible solution with available technology as of 2013. However, if the combustion is carried out under high pressure, as with the IGCC process, membranes can become a viable option once they achieve high separation of CO_2 [14].

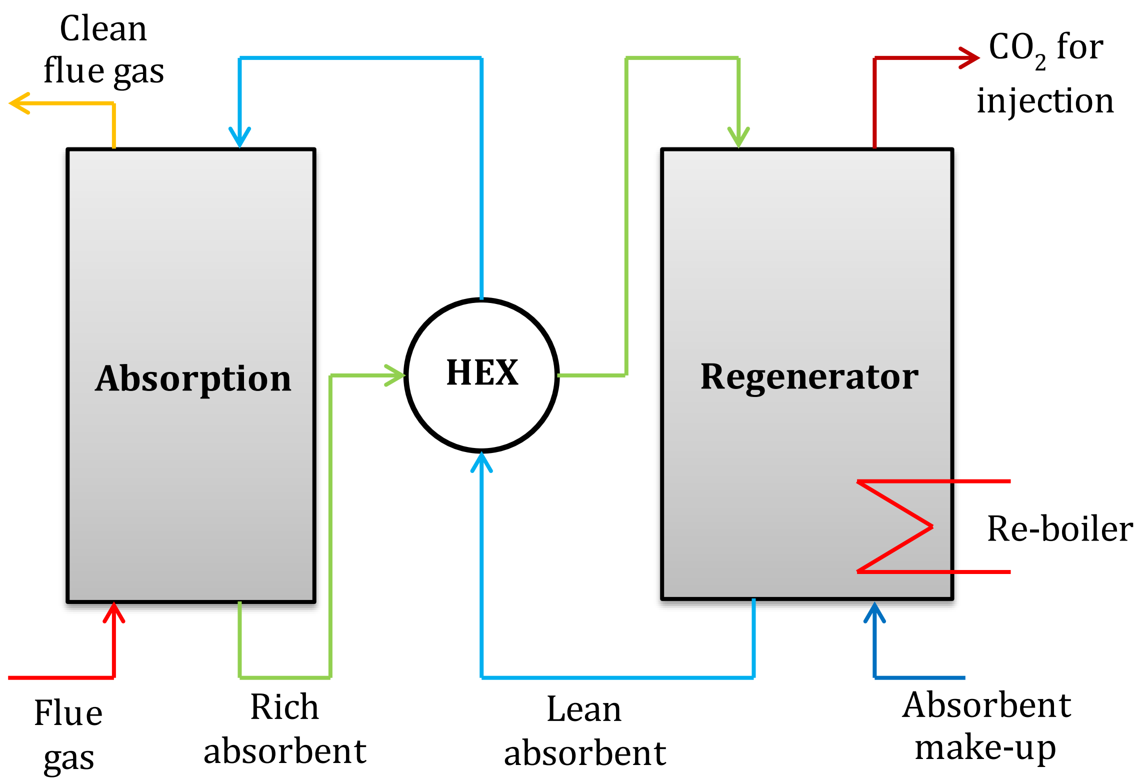

Figure 3.1: Conceptual process flow diagram of the absorption process.
HEX = heat-exchanger used to minimize the total heat needed for separation of carbon dioxide.

3.1.2 Pre-Combustion Capture

Pre-combustion capture involves reacting fuel with oxygen or air and adding water, and converting the carbonaceous material into synthesis gas containing carbon monoxide and hydrogen. During the conversion of fuel into synthesis gas, CO_2 is produced via water-shift reaction. CO_2 is then separated from the synthesis gas using a chemical or physical absorption process resulting in H_2 rich fuel which can be further combusted with air. Pressure swing adsorption is commonly used for the purification of syngas to high purity of H_2, however, it does not selectively separate CO_2 from the waste gas, which requires further purification of CO_2 for storage. The chemical absorption process is also used to capture CO_2 from syngas at partial pressure below 1.5 MPa. The solvent removes CO_2 from the shifted syngas by mean of chemical reaction which can be reversed by high pressure and heating. The physical absorption process is applicable in gas streams which have higher CO_2 partial pressure or total pressure and also with higher sulfur contents. This process is used for the capturing of both H_2S and CO_2, and one commercial solvent is Selexol [8].

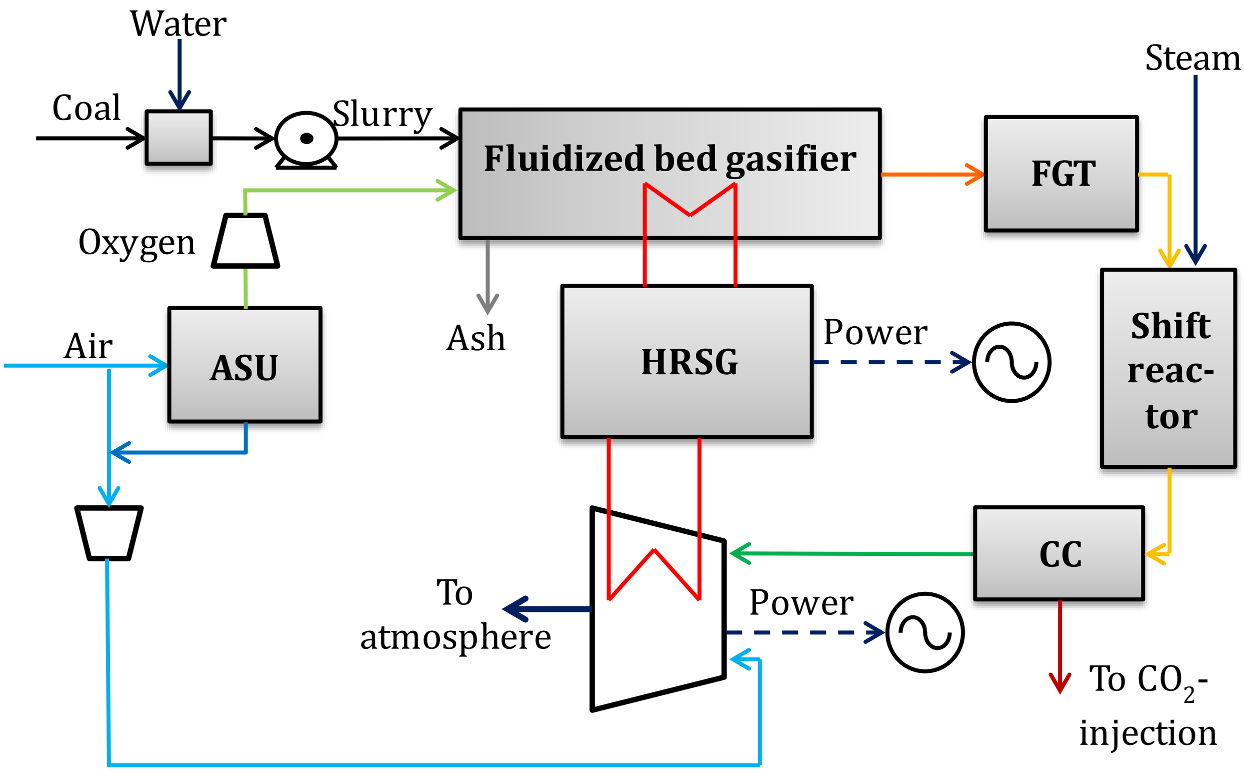

Figure 3.2: Schematic drawing of an integrated gasification combined cycle plant with pre-combustion carbon caption. The carbon monoxide from the gasifier is converted to carbon dioxide and hydrogen by reacting it with steam in the shift reactor. The carbon dioxide is captured in the carbon capture unit (CC, see Figure 3.1), and pure hydrogen is burned with nitrogen diluted air.

3.1.3 Oxygen-Fired Combustion

In oxygen-fired coal combustion, the flue gas is almost free of nitrogen-gases and after removal of sulfur the flue gas is 90 % CO_2, rest H_2O. There is no need for further CO_2-capture, and the CO_2 can be compressed and stored [11].

3.1.4 Carbon Storage

After the carbon dioxide has been compressed it can be injected into storage. CO_2 is usually stored in geological formation at depths of 1000 m or more [15]. Hence high pressures are required before injection, which has the advantage that CO_2 can be injected as a supercritical fluid. This will reduce the pipeline diameter and, consequently capital cost [16]. The oil industry is experienced with geological difficulties, and may provide expertise on the geological formation and how they will react to carbon dioxide injection. The storage site and reservoir has to be closely monitored to ensure that the carbon dioxide does not escape into the atmosphere or nearby drinking water supplies. With careful design of injection and appropriate monitoring of well pressure and local CO_2-concentrations, it can be ensured that the injected carbon dioxide remains underground for thousands of years [15].

3.1.5 Economics of CO_2 Capture

CO_2 capture is an expensive process both in capital costs and variable costs. Post combustion absorption by amine solution can add as high as a 29% electrical power penalty even with state of the art CO_2 capture technology [8]. The capital costs depend highly on the flow rate of flue gas, as this increases the regenerator and compressor size. The variable costs are also increased with higher flue gas flow, as more sorbent is required, and the cost of CO_2-transport and storage will increase [17]. Improving process configurations and solvent capacity can majorly reduce power demand for the regenerator. Such improvements include: absorber intercooling, stripper interheating, flashing systems and multi-pressure stripping, though all of these will come at the expense of complexity and higher capital costs. Table 3.1 shows the development in MEA absorption systems from year 2001 to 2006 [18].

Table 3.1: Economics of CO_2 capture by MEA scrubbing [18].

Year of design	2001	2006
MEA [weight percent]	20	30
Power used [MWh/ton]	0.51	0.37
@ $ 80/MWh [$/ton CO_2 removed]	41	29
Capital cost [$/ton CO_2 removed per year]	186	106
@ 16%/year [$/ton CO_2 removed]	30	17
Operating and maintenance cost [$/ton CO_2 removed]	6	6
Total cost [$/ton CO_2 removed]	77	52
Net CO_2 removal with power replaced by gas [%]	72	74

The environmental impact is commonly expressed as cost per pollutant removed or cost per pollutant avoided. Cost of CO_2 per tom removed is different from cost of CO2 avoided and cost of CO_2 avoided is given as [19] :

$$\text{cost of CO}_2\text{ avoided (\$/tonne)} = \frac{(\$/\text{kWh})_{\text{capture}} - (\$/\text{kWh})_{\text{reference}}}{(\text{tonne CO}_2/\text{kWh})_{\text{reference}} - (\text{tonne CO}_2/\text{kWh})_{\text{capture}}} \quad (1)$$

Table 3.2 shows the cost variability and representative cost values for power generation and CO_2 capture, for the three fuel systems respectively. The cost of electricity is lowest for the NGCC, regardless of CO_2 capture. Pulverized Coal plant has lower capital cost without capture while IGCC plant has lower cost when current CO_2 capture is added in the system.

Table 3.2: Summary of reported CO_2 emissions and costs for a new electric power plant with and without CO_2 capture base don current technology (excluding CO_2 transport and storage costs) [20].
MW_{ref} = reference plant net output
Rep. value = representative value
NGCC = natural gas combined cycle plant.

Cost and Performance Measures	PC Plant		IGCC Plant		NGCC Plant	
	Range low-high	*Rep. value*	Range low-high	*Rep. value*	Range low-high	*Rep. value*
Emission rate w/o capture [kg CO_2/MWh]	722-941	795	682-846	757	344-364	358
Emission rate with capture [kg CO_2/MWh]	59-148	116	70-152	113	40-63	50
Percent CO_2 reduction per kWh [%]	80-93	85	81-91	85	83-88	87
Capital cost w/o capture [$/kW]	1100-1490	1260	1170-1590	1380	447-690	560
Capital cost with capture [$/kW]	1940-2580	2210	1410-2380	1880	820-2020	1190
Percent increase in capital cost [%]	67-87	77	19-66	36	37-190	110
Cost of CO_2 avoided [$/t CO_2]	42-55	47	13-37	26	35-74	47
Cost of CO_2 captured [$/t CO_2]	29-44	34	11-32	22	28-57	41
Power penalty for capture [% MW_{ref}]	22-29	27	12-20	16	14-16	15
Thermal efficiency w/o capture [8]	36.8%-39.3%	38.1%	39.0%-42.1%	40.2%	50.2%	50.2%
Thermal efficiency w/ capture [8]	26.2%-28.4%	27.3%	31.0%-32.6%	31.6%	42.8%	42.8%

3.2 Flue Gas Desulphurization

SO_2 has a harmful effect both on humans and the environment. Exposure to higher concentrations of SO_2 is the cause of many harmful diseases. SO_2 affects the environment by reacting into acids and is a major source of acid rain [21]. Combustion of sulfur-containing compounds such as coal is a major source of SO_2 generation. Removal of sulfur from solid fuels is not practical, so the sulfur is removed from the flue gas after combustion of coal. The removal of sulfur oxide from the flue gasses is achieved by physical or chemical absorption process [22]. There are two commonly used industrial processes for the desulphurization of flue gasses [23], wet and dry scrubbing.

3.2.1 Wet scrubbing

In wet scrubbing, a solvent is used for the absorption of SO_2. Typically water is considered to be the cheapest solvent. It's washing capacity however is very limited and huge quantity of water has to be used. Approximately 75 tons of water is used per ton of flue gas, and even then 5% of SO_2 remains [22].

In advanced processes, flue gas is treated with an alkaline slurry in an absorber, where SO_2 is captured, shown in Figure 3.3. The most commonly used slurry is composed of limestone, which reacts with the sulfur. The sulfur removal efficiency is 98 % for wet slurry scrubbing processes. Carbon dioxide removal units include a polishing scrubber which lowers flue gas SO_2 content from 44 ppmv to 10 ppmv [8].

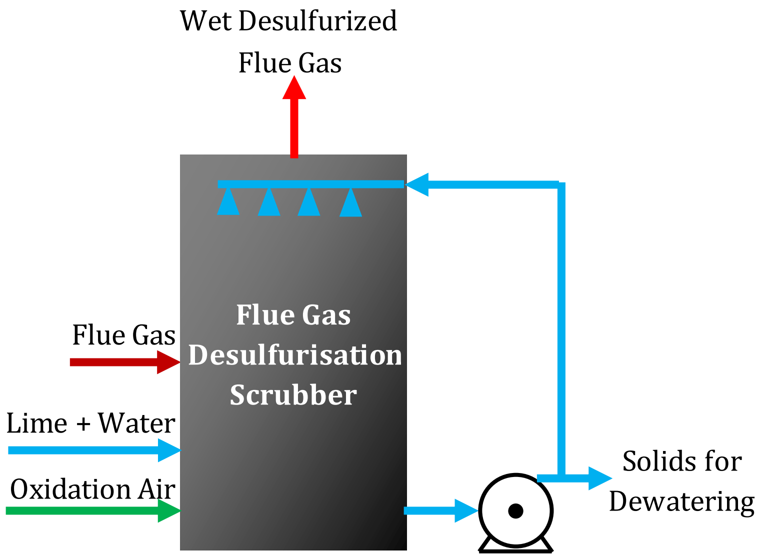

Figure 3.3: Flue gas desulfurization via wet scrubbing using limestone slurry [24].

Wet scrubbing process is mostly used with higher sulfur contents with economic efficiency of 95 to 98%. The main disadvantage of wet scrubbing is acidic environment which can be corrosive. Therefore, corrosion resistant material is required for the construction which increases capital cost of the plant. The other disadvantage includes consumption of large quantity of water for the process.

3.2.2 Dry Scrubbing

Dry scrubbing is useful for coal with lower sulfur content. It also has the major advantage of maintaining a higher temperature of the emission gasses, while wet scrubbing decreases the temperature to that of water used [21]. Lime is normally used as sorbent-agent during the dry scrubbing process. The dry sorbent reacts with the flue gas at 1000 °C to remove SO_2. In dry scrubbing adiabatic saturation approach is normally applied. Adiabatic saturation is required to achieve high SO_2 removal, thereby carefully controlling the amount of water [21].

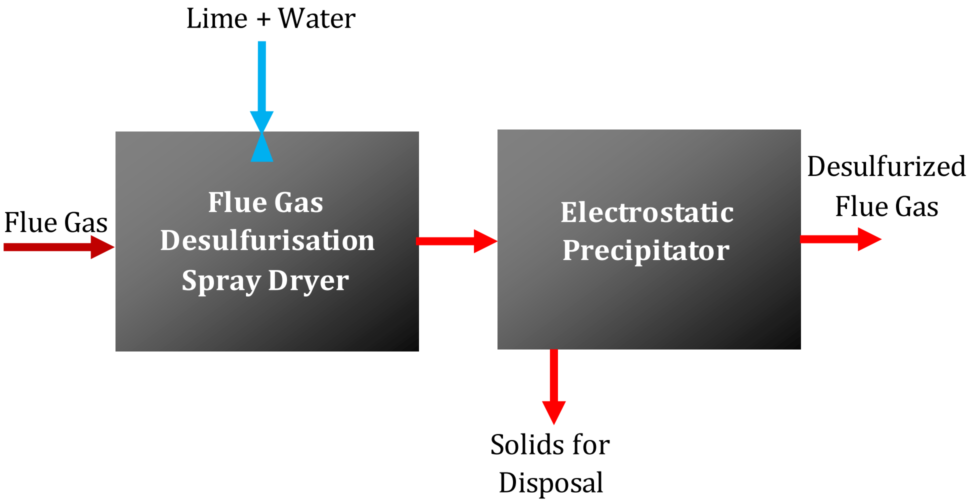

Figure 3.4: Flue gas desulfurization via dry scrubbing using limestone [24].

Relatively poor conversion is the main disadvantage of dry sorbent process, which increases the operational costs. Dry scrubbing uses a minimal amount of water, and leaves the flue gas dry, reducing risk of corrosion. Relatively dry calcium sulfite is obtained as by-product with fly ash [21].

3.3 NO_x Removal

Selective catalytic reduction (SCR) approach is applied for removal of NO_x, where the flue gas is reduced with ammonia over a catalyst to nitrogen and water. The SCR has a efficiency of 80 to 90 % for NO_x removal. The optimal temperature range is 650 – 750 °C, with vanadium and titanium as catalyst, and the following reaction is carried out during the process:

$$4NH_3 + 4NO + O_2 \rightarrow 4N_2 + 6H_2O \tag{2}$$

Ammonia is either injected in pure form under pressure or in an aqueous solution. Instead of ammonia, urea can also be used for the process. The main challenge of the process is full conversion as emission of the NH_3 is highly undesirable. The unreacted NH_3 can also oxidize SO_2 to SO_3 that further reacts with NH_3 to form ammonium bisulfate. Ammonium bisulfate is a sticky solid material which can plug the equipment. The catalyst activity is very critical with SCR because catalyst can cost up to 15 – 20% of the capital cost [7].

4 Process Descriptions

4.1 Current Plant at Svalbard

A process flow diagram for the existing plant at Svalbard is shown in Figure 4.1.

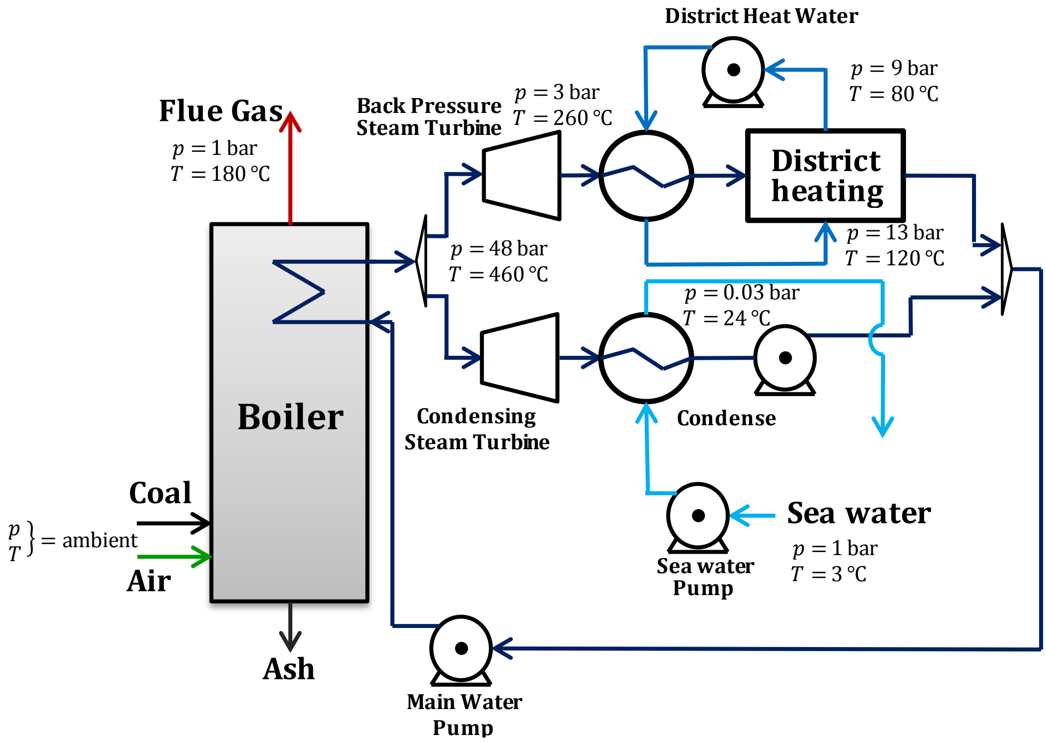

Figure 4.1: Process flow diagram of current plant at Svalbard

4.1.1 Boiler

Coal is crushed and fed to the boiler along with air, where it combusts at high temperatures forming flue gas and ash. The air is blown into the boiler by fans. Water is circulated through the evaporator-drums in the boiler to yield pressurized steam. The flue gas is sent directly to the stack and out to the atmosphere.

4.1.2 Steam Cycle

The hot steam from the boiler is split into two streams, one of which is sent to a condensing steam turbine, the other to a backpressure steam turbine. The steam from the condensing steam turbine is quenched in a condenser that is cooled with seawater; this allows the outlet pressure from the turbine to be relatively low, depending on how much it is cooled. The backpressure steam turbine has a relatively high backpressure, which yields moderately high temperatures in the outlet stream. This heat is exploited by heating up cold water from the district heating network. The two different streams are then combined, sent through a pump and back to the boiler.

4.1.3 District Heating

The district heating water circulates Longyearbyen, and is used to heat houses and tap water, before it is sent back to the power plant for reheating. The water is pumped to a heat exchanger where it is reheated by the condensing steam from the backpressure steam turbine. After which it is sent back to the district heating network.

4.1.4 Gas Treatment

The flue gas is currently not treated before it is sent trough the stack to the atmosphere.

4.2 Proposed Pulverized Coal plant with Carbon Capture and Storage (Base Case)

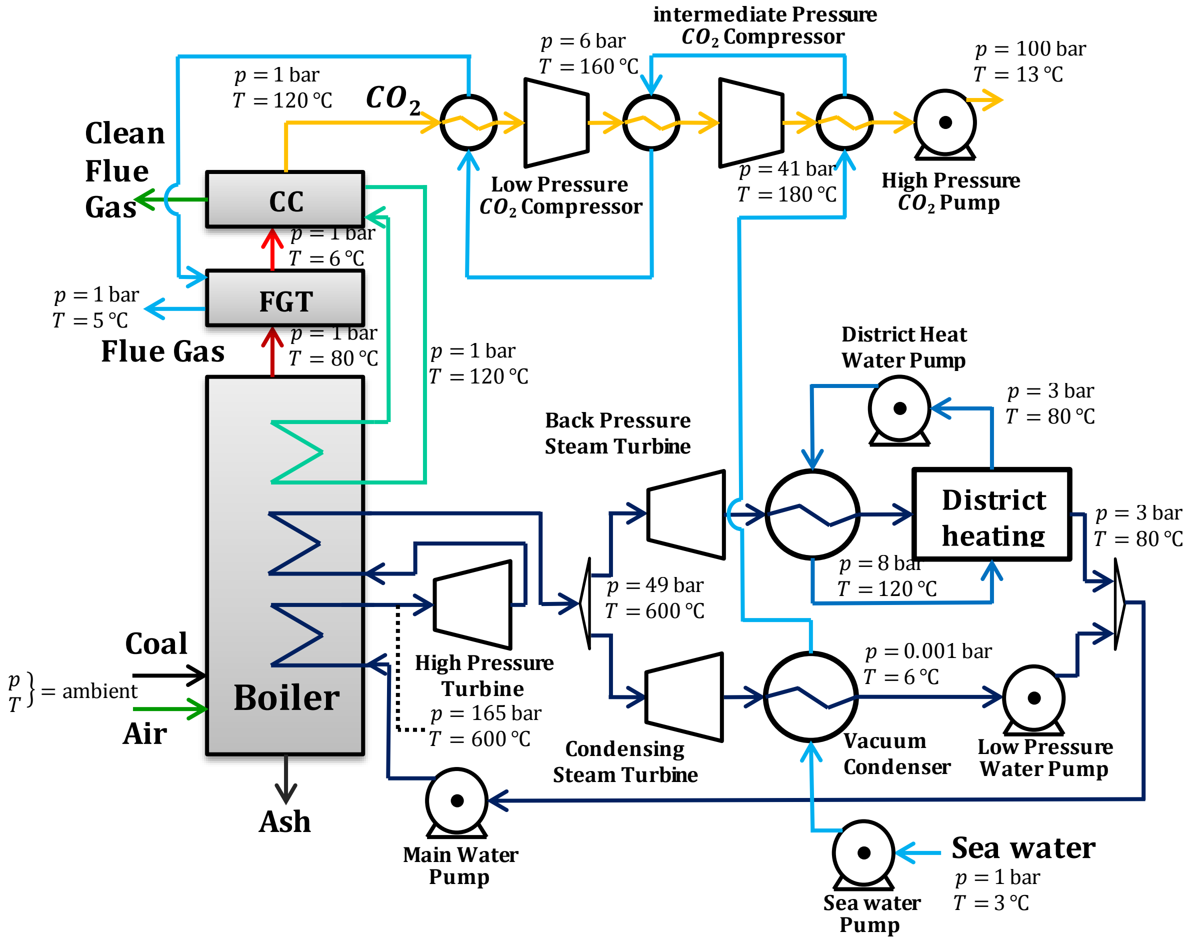

Figure 4.2: Process flow diagram of the proposed design.
FGT = flue gas treatment unit, and
CC = carbon capture unit.

4.2.1 Pulverized Coal Boiler

Coal is pulverized and fed to the boiler along with air, where it combusts at high temperatures forming flue gas and ash. Water is circulated through the evaporator-drums in the boiler to yield high pressure (HP) steam, through the re-heater to re-heat the intermediate pressure (IP) steam, while re-boiler amine solution is fed through the economizer, recovering heat from the flue gas.

4.2.2 Steam Cycle

HP steam is run through a HP turbine, and the re-heated IP steam is run through a split to the IP turbine or the low pressure (LP) turbine. After the LP turbine, the steam condenses using seawater from the seawater pump. Because of the low temperature of the seawater, low pressures are obtained. After the IP turbine, the water still has a high temperature and pressure, which is used for district heating. Residual condensed steam from both the IP and LP turbine is pumped back up to a high pressure and once again fed into the boiler evaporator drums.

4.2.3 Flue Gas Treatment

The flue gas out of the boiler is first desulfurized in the flue gas desulfurization unit (FGD), using a seawater scrubber. The seawater is used to cool steam and CO_2 earlier in the process, and then fed to the seawater scrubber. As the concentration of sulfur in the water out is low, it can safely be pumped back to the ocean. As the flue gas is quenched through seawater, dust particles and mercury is also removed.

After FGD, the flue gas goes through the carbon capture (CC) facility, using an amine solution. Flue gas with low CO_2-content is sent out to the environment, while the captured CO_2 is sent to compression and storage. Here the CO_2 is cooled with seawater and compressed two times, before being pumped as a supercritical fluid down into geological formations.

4.2.4 District Heating

The proposed pulverized coal power plant will use the same district heating network as is already built on Svalbard, giving a steam split-factor between IP and LP turbines. The district heating water circulates Longyearbyen, and is used to heat houses and tap water. After being used, the water is sent back to the power plant. The cycle introduces a pressure drop, which is equalized by the district heating water pump.

4.3 Case Studies

In the report two different cases will be studied, one using the existing district heating loop on Svalbard, another introducing a central heat pump.

4.3.1 Base Case

The base case uses the current district heating network on Svalbard, and produces 12 MW of district heating, with double the current electric power, giving 9.6 MW of electric energy. The split of steam to the backpressure IP turbine for district heating and LP turbine for pure electric generation is adjustable, and was adjusted to fit the design basis in Table 1.1, using as little coal as possible.

4.3.2 Heat Pump Case

The heat pump case is an attempt at maximizing electric energy produced, to use the electric energy in heat pumps instead of district heating. This model has the possibility of selling excess electric power to the neighboring town of Barentsburg, but is highly dependent on good coefficient of performance for the heat pump. Investment analysis is performed on both cases, using the same amount of coal, to compare them to each other.

5 Flowsheet Calculations

5.1 Base Case

The plant is modeled using Aspen HYSYS, based on the results from the report “Cost and Performance Baseline for Fossil Energy Plants” by the National Energy Technology Laboratory [8].

5.1.1 Flow diagram

The process flow diagram from the Aspen HYSYS model is shown in Figure 5.1, a larger figure is shown in Appendix F .

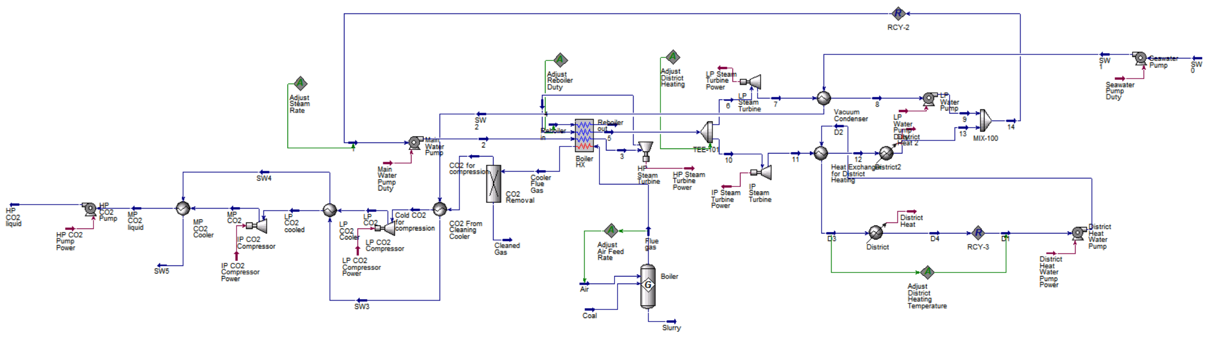

Figure 5.1: The process flow diagram from the Aspen HYSYS model for the base case.

5.1.2 Stream Data

The important stream data for the different streams in the Aspen HYSYS model are shown in Table 5.1.

Table 5.1: Stream data for the base case, calculated in Aspen HYSYS.

Stream name	Vapor fraction	Temperature [°C]	Pressure [bar]	Mass flow [kg/s]
Air	1.00	10.00	1.0	70.4
Coal	0.00	10.00	1.0	1.90
Flue gas	1.00	799.99	1.0	72.3
Slurry	0.00	799.99	1.0	0.00
1	0.00	40.36	3.0	9.08
2	0.00	41.30	165.0	9.08
3	1.00	600.00	165.0	9.08
4	1.00	407.07	49.0	9.08
5	1.00	600.00	49.0	9.08
6	1.00	600.00	49.0	4.87
7	0.90	5.88	0.0	4.87
8	0.00	5.88	0.0	4.87
9	0.00	5.99	10.0	4.87
10	1.00	600.00	49.0	4.21
13	0.00	80.00	3.0	4.21
14	0.00	40.44	3.0	9.08
D1	0.00	80.00	3.0	63.8
D4	0.00	80.00	3.0	63.8
D2	0.00	80.04	8.0	63.8
D3	0.00	120.00	8.0	63.8
11	1.00	349.87	8.0	4.21
12	0.00	120.00	8.0	4.21
Reboiler in	0.00	120.00	2.0	9.93
Reboiler out	1.00	120.00	2.0	9.93
Cleaned Gas	1.00	79.42	1.0	65.3
CO2 for compression	1.00	79.40	1.0	6.97
MP CO2	1.00	177.95	41.0	6.97
MP CO2 liquid	0.00	6.00	41.0	6.97
HP CO2 liquid	0.00	13.31	100.0	6.97
SW 1	0.00	3.00	1.2	2560
SW 2	0.00	4.00	1.2	2560
SW 0	0.00	3.00	1.0	2560
SW5	0.00	4.39	1.2	2560
LP CO2	1.00	163.23	6.1	6.97
LP CO2 cooled	1.00	10.00	6.1	6.97
SW4	0.00	4.13	1.2	2560
Cold CO2 for compression	1.00	10.00	1.0	6.97
SW3	0.00	4.04	1.2	2560
Cooler Flue Gas	1.00	79.40	1.0	72.3

5.1.3 Compositions

The composition of each stream from the Aspen HYSYS model is shown in Table 5.2.

Table 5.2: Composition data for the base case, calculated in Aspen HYSYS, where x_i is the mole fraction of component i.

Stream name	x_{H_2O}	x_{CO_2}	x_{O_2}	x_{N_2}	x_C
Air	0.000	0.000	0.210	0.790	0.000
Coal	0.000	0.000	0.000	0.000	1.000
Flue gas	0.000	0.065	0.145	0.790	0.000
Slurry	0.000	0.065	0.145	0.790	0.000
1	1.000	0.000	0.000	0.000	0.000
2	1.000	0.000	0.000	0.000	0.000
3	1.000	0.000	0.000	0.000	0.000
4	1.000	0.000	0.000	0.000	0.000
5	1.000	0.000	0.000	0.000	0.000
6	1.000	0.000	0.000	0.000	0.000
7	1.000	0.000	0.000	0.000	0.000
8	1.000	0.000	0.000	0.000	0.000
9	1.000	0.000	0.000	0.000	0.000
10	1.000	0.000	0.000	0.000	0.000
13	1.000	0.000	0.000	0.000	0.000
14	1.000	0.000	0.000	0.000	0.000
D1	1.000	0.000	0.000	0.000	0.000
D4	1.000	0.000	0.000	0.000	0.000
D2	1.000	0.000	0.000	0.000	0.000
D3	1.000	0.000	0.000	0.000	0.000
11	1.000	0.000	0.000	0.000	0.000
12	1.000	0.000	0.000	0.000	0.000
Reboiler in	1.000	0.000	0.000	0.000	0.000
Reboiler out	1.000	0.000	0.000	0.000	0.000
Cleaned Gas	0.000	0.000	0.155	0.845	0.000
CO2 for compression	0.000	1.000	0.000	0.000	0.000
MP CO2	0.000	1.000	0.000	0.000	0.000
MP CO2 liquid	0.000	1.000	0.000	0.000	0.000
HP CO2 liquid	0.000	1.000	0.000	0.000	0.000
SW 1	1.000	0.000	0.000	0.000	0.000
SW 2	1.000	0.000	0.000	0.000	0.000
SW 0	1.000	0.000	0.000	0.000	0.000
SW5	1.000	0.000	0.000	0.000	0.000
LP CO2	0.000	1.000	0.000	0.000	0.000
LP CO2 cooled	0.000	1.000	0.000	0.000	0.000
SW4	1.000	0.000	0.000	0.000	0.000
Cold CO2 for compression	0.000	1.000	0.000	0.000	0.000
SW3	1.000	0.000	0.000	0.000	0.000
Cooler Flue Gas	0.000	0.065	0.145	0.790	0.000

5.1.4 Summary of Key Results

A summary of the key results from the flowsheet calculation is shown in Table 5.3. In addition, the composite curves for all the heat exchangers are shown in Appendix E

Table 5.3: A summary of the results from the simulation of the base case.

Object	Value
District heating power output	12.0 MW
Net electrical power output	9.6 MW
Amount of coal needed	60000 ton/year (1.90 kg/s)
Thermal efficiency	34.7%
Heat needed for CO_2-removal	22.6 MW

5.1.5 Steam Temperature

A case study was performed on the model in Aspen HYSYS, which studies the effect of the temperature of the steam. The thermal efficiency is calculated as:

$$\eta_{\text{thermal}} = \frac{\text{Useful energy out}}{\text{Useful energy in}} = \frac{W_{\text{electric}} + Q_{\text{district heat}}}{m_{\text{C}}\ \text{NCV}} \tag{3}$$

Where η_{thermal} is the thermal efficiency, W_{electric} is the net electric power produced, $Q_{\text{district heat}}$ is the power provided to district heating, m_{C} is the mass flow of coal and NCV is the coal's net calorific value. In the rest of the report, the thermal efficiency is used exclusively. The results are shown in Figure 5.2. The steam temperature's effect on efficiency is highly dependent on the boiler design, which can be seen from the composite curves in Figure 5.3.

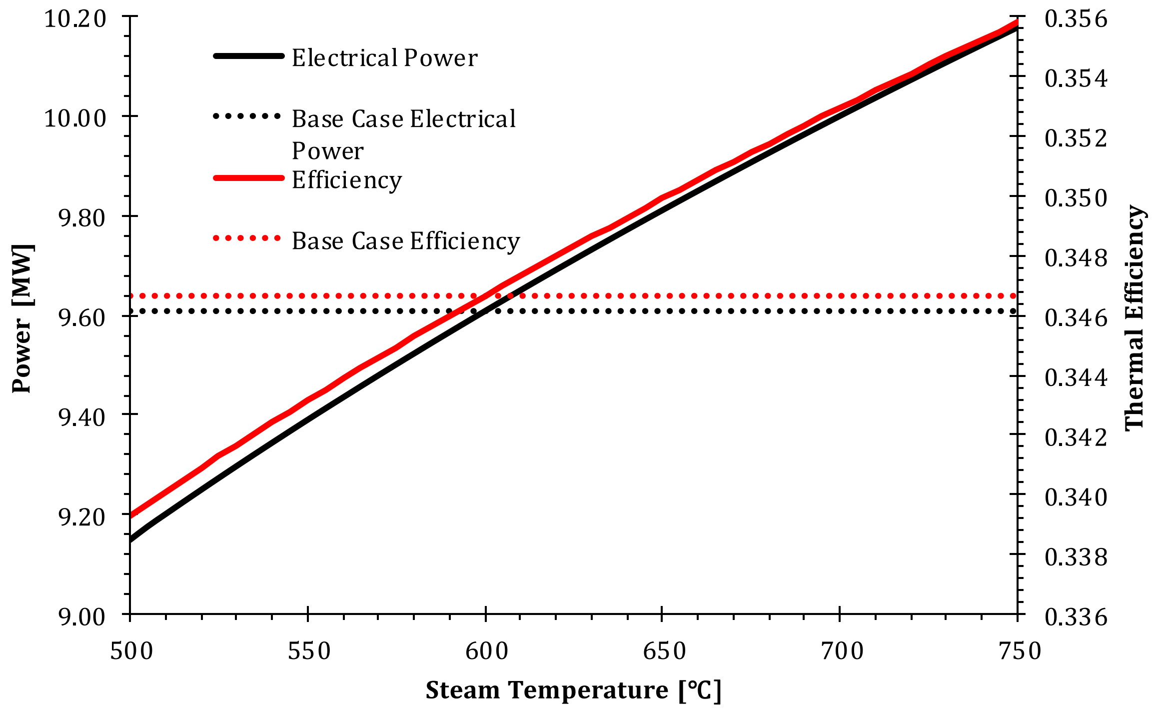

Figure 5.2: The effect on net power output and thermal efficiency as a function of temperature in the combustion chamber of the boiler.

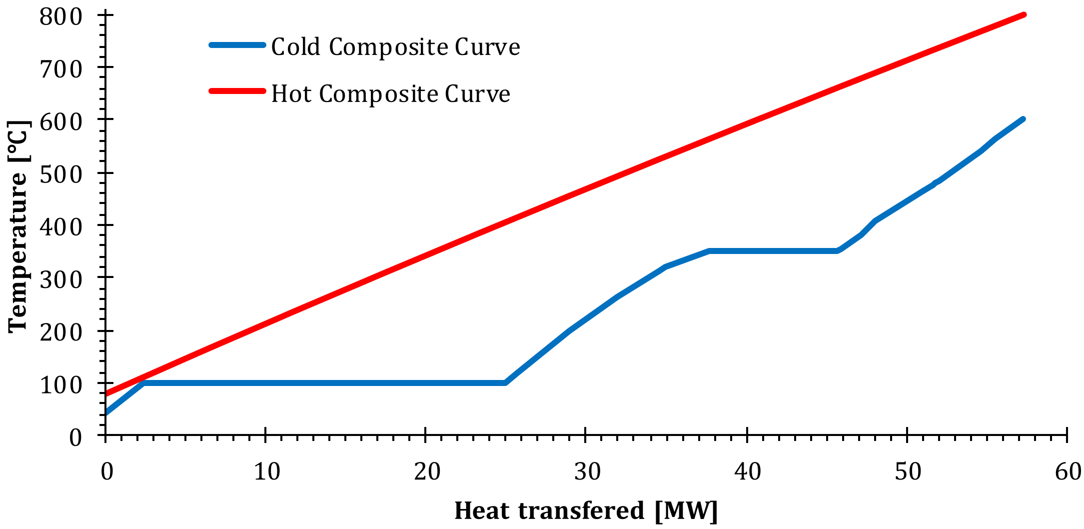

Figure 5.3: Plot of hot and cold composite curve for the pulverized coal boiler. The pinch temperature is set to 10°C.

5.1.6 Steam Cycle Pressure

A case study was performed on the model in Aspen HYSYS, which studies the effect of the highest pressure in the steam cycle. The results are shown in Figure 5.4.

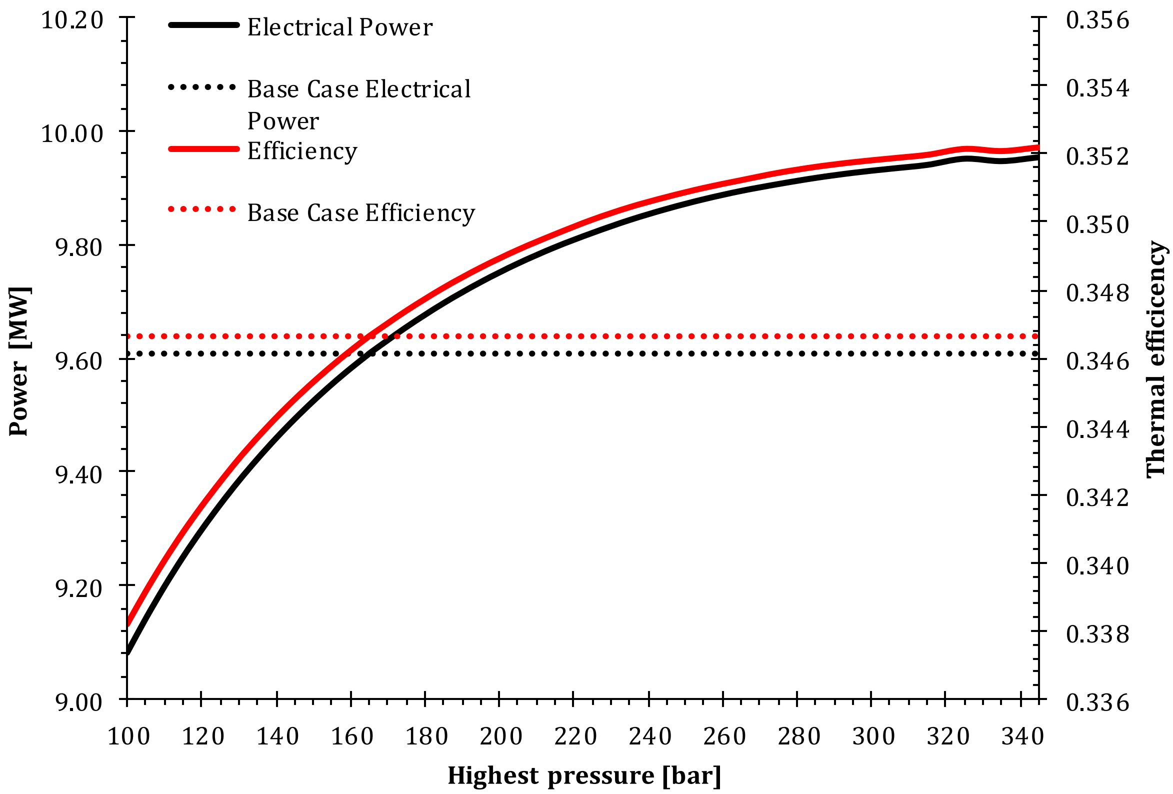

Figure 5.4: The effect on net power output and thermal efficiency as a function of maximum pressure in the steam cycle.

5.2 Heat Pump Case

The heat pump case is modeled in a similar manner as the base case, but without the district heating network and the split between the LP steam turbine and the IP steam turbine.

5.2.1 Flow Diagram

The process flow diagram from the Aspen HYSYS model is shown in Figure 5.5.

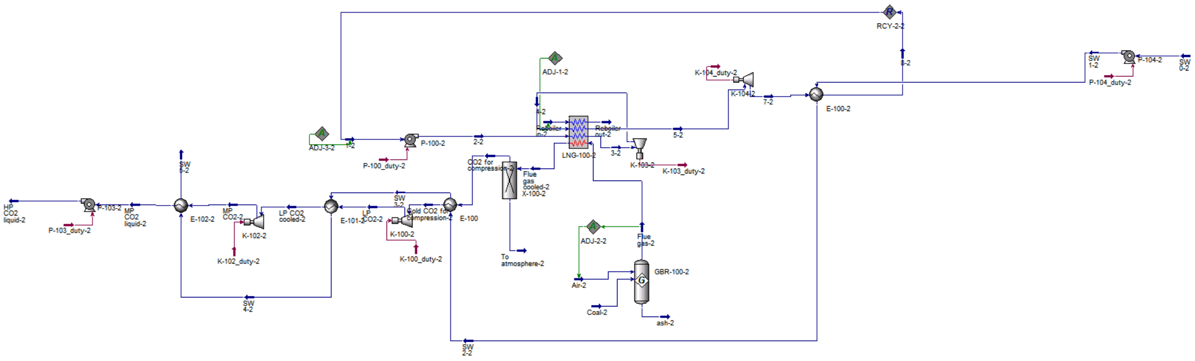

Figure 5.5: The process flow diagram from the Aspen HYSYS model for the heat pump case.

5.2.2 Stream Data

The important stream data for the different streams in the Aspen HYSYS model are shown in Table 5.4.

Table 5.4: Stream data for the heat pump case, calculated in Aspen HYSYS.

Stream name	Vapor fraction	Temperature [°C]	Pressure [bar]	Mass flow [kg/s]
Air	1.00	10.00	1.01	70.4
Coal	0.00	10.00	1.00	1.90
Flue gas	1.00	799.99	1.00	72.3
ash	0.00	799.99	1.00	0.00
1	0.00	5.88	0.01	9.09
2	0.00	6.65	165.00	9.09
3	1.00	600.00	165.00	9.09
4	1.00	407.07	49.00	9.09
5	1.00	600.00	49.00	9.09
7	0.90	5.88	0.01	9.09
8	0.00	5.88	0.01	9.09
Flue gas cooled	1.00	60.50	1.00	72.3
Reboiler in	0.00	120.0	1.01	9.93
Reboiler out	1.00	120.0	1.01	9.93
To atmosphere	1.00	60.53	1.00	65.3
CO2 for compression	1.00	60.50	1.00	6.97
LP CO2	1.00	163.23	6.10	6.97
LP CO2 cooled	1.00	10.00	6.10	6.97
MP CO2	1.00	177.95	41.00	6.97
MP CO2 liquid	0.00	6.00	41.00	6.97
HP CO2 liquid	0.00	13.31	100.00	6.97
SW 1	0.00	3.00	1.20	3183
SW 2	0.00	4.50	1.20	3183
SW 0	0.00	3.00	1.00	3183
Cold CO2 for compression	1.00	10.00	1.00	6.97
SW 3	0.00	4.52	1.20	3183
SW 4	0.00	4.60	1.20	3183
SW 5	0.00	4.80	1.20	3183

5.2.3 Compositions

The composition of each stream from the Aspen HYSYS model is shown in Table 5.5.

Table 5.5: Stream data for the heat pump case, calculated in Aspen HYSYS, where x_i is the mole fraction of component *i*.

Stream name	x_{H_2O}	x_{CO_2}	x_{O_2}	x_{N_2}	x_C
Air	0.000	0.000	0.210	0.790	0.000
Coal	0.000	0.000	0.000	0.000	1.000
Flue gas	0.000	0.065	0.145	0.790	0.000
ash	0.000	0.065	0.145	0.790	0.000
1	1.000	0.000	0.000	0.000	0.000
2	1.000	0.000	0.000	0.000	0.000
3	1.000	0.000	0.000	0.000	0.000
4	1.000	0.000	0.000	0.000	0.000
5	1.000	0.000	0.000	0.000	0.000
7	1.000	0.000	0.000	0.000	0.000
8	1.000	0.000	0.000	0.000	0.000
Flue gas cooled	0.000	0.065	0.145	0.790	0.000
Reboiler in	1.000	0.000	0.000	0.000	0.000
Reboiler out	1.000	0.000	0.000	0.000	0.000
To atmosphere	0.000	0.000	0.155	0.845	0.000
CO2 for compression	0.000	1.000	0.000	0.000	0.000
LP CO2	0.000	1.000	0.000	0.000	0.000
LP CO2 cooled	0.000	1.000	0.000	0.000	0.000
MP CO2	0.000	1.000	0.000	0.000	0.000
MP CO2 liquid	0.000	1.000	0.000	0.000	0.000
HP CO2 liquid	0.000	1.000	0.000	0.000	0.000
SW 1	1.000	0.000	0.000	0.000	0.000
SW 2	1.000	0.000	0.000	0.000	0.000
SW 0	1.000	0.000	0.000	0.000	0.000
Cold CO2 for compression	0.000	1.000	0.000	0.000	0.000
SW 3	1.000	0.000	0.000	0.000	0.000
SW 4	1.000	0.000	0.000	0.000	0.000
SW 5	1.000	0.000	0.000	0.000	0.000

5.2.4 Summary of Key Results

A summary of the key results from the flowsheet calculation is shown in Table 5.6. In addition, the composite curves for all the heat exchangers are shown in Appendix E

Table 5.6: A summary of the results from the simulation of the heat pump case. A coefficient of performance of 3 was used to obtain these results.

Object	Value
District heating power output	12.0 MW
Net electrical power output	9.4 MW
Amount of coal needed	60000 ton/year (1.90 kg/s)
Thermal efficiency	34.3%
Heat needed for CO_2-removal	22.6 MW

5.2.5 Coefficient of performance

The overall thermal efficiency was calculated for different values of the heat pump's coefficient of performance. A plot of the result is shown in Figure 5.6.

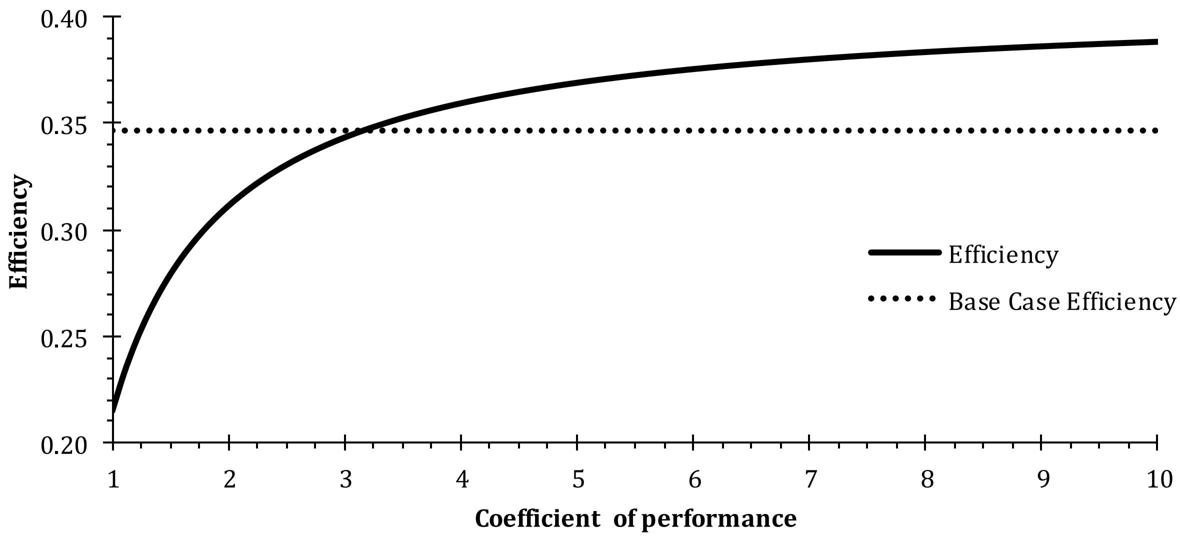

Figure 5.6: A plot of overall thermal efficiency as a function of the heat pump's coefficient of performance.

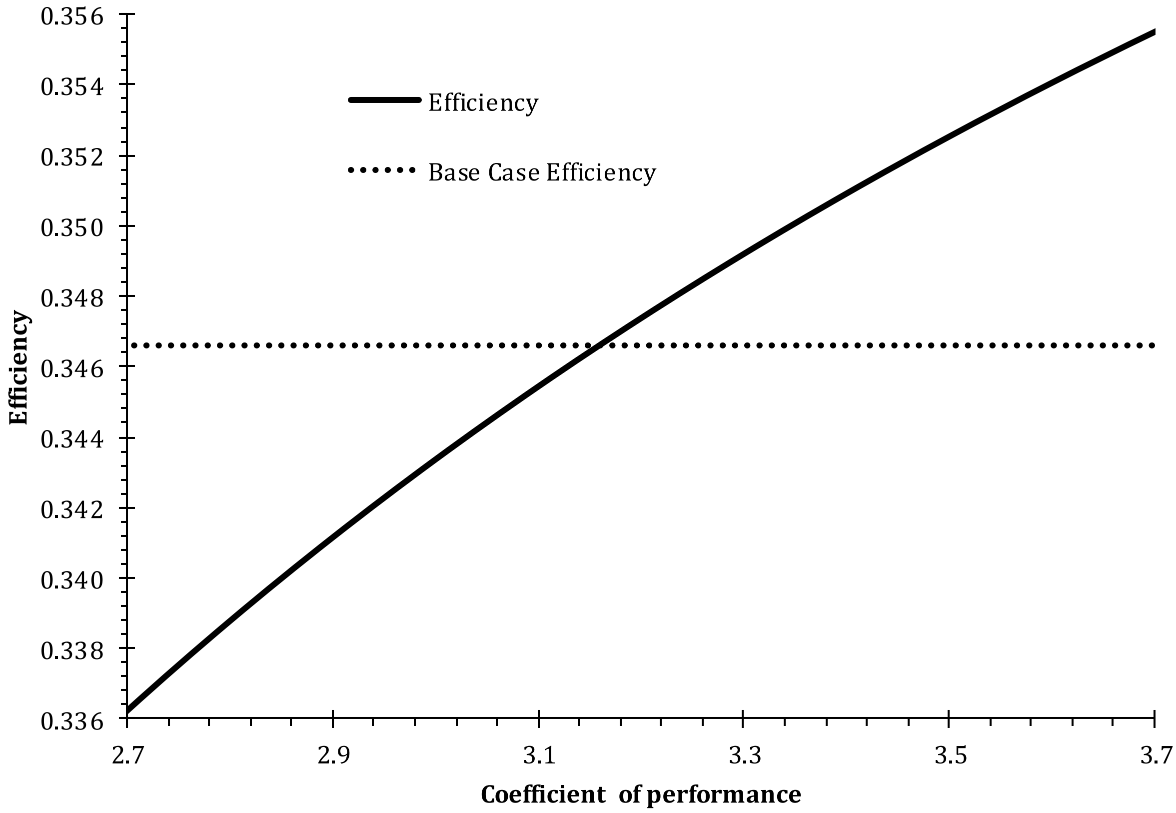

Figure 5.7: An excerpt of Figure 5.6 for comparison with Figure 5.2 and Figure 5.4.

As can be seen, a high coefficient of performance is required to compete with the base case. This is further addressed in Section 8.2.4.

6 Cost Estimation

6.1 Capital Costs of Major Equipment

Cost estimations were performed on major equipment for both the district heating case and the heat pump case. In this section, only costs for the base case are shown, while detailed cost estimations for both cases are shown in Appendix A and Appendix B . Cost estimations were also performed for the base case but without district heating, this is shown in Appendix C .

6.1.1 Pulverized Coal boiler

The pulverized coal (PC) boiler cost was estimated by an order-of-magnitude scaling, using the following equation [25]:

$$C_2 = C_1\left(\frac{S_2}{S_1}\right)^n = C_1\left(\frac{S_2}{S_1}\right)^{0.67} \tag{4}$$

where C_2 is the cost of a plant with capacity S_2, C_1 is the cost of a plant with capacity S_1 and n is the exponent, which can be assumed to be 0.67 for PC plant boilers [26]. The capacity used for calculations was electrical power produced in MW, assuming no district heating. By using data for a PC plant with CO_2-capture [8], a cost of $ 33 807 600 (on Jan. 2013 basis by using CEPCI [27]) was obtained. This includes costs for piping, instrumentation and equipment erection. This cost has some uncertainty, with the original boiler being 50 times larger than the boiler estimated.

6.1.2 Heat exchangers

The heat exchangers were assumed to be U-tube shell and tube exchangers, with the following formula and tabulated values from Sinnott [25]:

$$C = a + bS^n = 24\,000 + 46 \times S^{1.2} \tag{5}$$

where a and b are cost constants, the capacity parameter S is the heat transfer area of the exchanger in m^2 and n is the exponent. To calculate the required area, the overall heat transfer coefficient was estimated from tables in Sinott [25].For given values of exchanger area, constants and exponent, the cost of the exchangers were estimated, shown in Table 6.1. Here the Vacuum Condenser was divided into several exchangers with capacity inside the 1000 m^2 limit of formula (5). The values are on Jan. 2013 basis by using CEPCI [27]. This does not include any material factors, piping, instrumentation or other cost factors.

Table 6.1: Estimations of heat exchanger costs, this does not include any material factors, piping, instrumentation or other cost factors.

Heat Exchanger	U [W/m²K]	Capacity [m²]	Cost [$]
Heat Exchanger for District Heating	1200	105	43 500
Four Vacuum Condensers	1200	4 × 984	977 200
CO_2 Cooler	50	313	83 300
Low Pressure CO_2 Cooler	100	213	63 100
Intermediate Pressure CO_2 Condenser	700	108	43 900

6.1.3 Turbines

The costs of the steam turbines were estimated using the same formula as for the boiler, using an exponent of 0.67 for steam turbines [26]. The cost of the High Pressure, Intermediate Pressure and Low Pressure turbines were estimated, and are shown in Table 6.2 on Jan. 2013 basis by using CEPCI [27]. This does not include any material factors, piping, instrumentation or other cost factors.

Table 6.2: Estimation of turbine cost, this does not include any material factors, piping, instrumentation or other cost factors.

Turbines	Capacity [kW]	Cost [$]
High Pressure Turbine	2991	951 300
Intermediate Pressure Turbine	2097	749 900
Low Pressure Turbine	6785	1 646 900

6.1.4 Compressors

The compressors for CO_2-injection were assumed to be centrifugal and estimated from tabulated values in Sinnott using the following formula, inserted for constants and exponent:

$$C = a + bS^n = 490\,000 + 16\,800 \times S^{0.6} \tag{6}$$

where S is the size parameter, here compressor power in kW. For given values of power, constants and exponent the cost of Low Pressure CO_2 Compressor and Intermediate Pressure

CO_2 Compressor were estimated to be \$ 2 652 400 and \$ 2 681 700, respectively. The values are on a Jan. 2013 basis by using CEPCI [27], and do not include any material factors, piping or other cost factors.

6.1.5 Pumps

In similar fashion to compressor estimations, pumps were assumed to be single-stage centrifugal pumps, with the following formula and tabulated values from Sinnott [25]:

$$C = a + bS^n = 6900 + 206 \times S^{0.9} \tag{7}$$

where the capacity parameter S is liters feed per second. For given values of feed, constants and exponent, the cost of the pumps were estimated, shown in Table 6.3. The values are on Jan. 2013 basis by using CEPCI [27]. The cost of the Seawater pump has a large uncertainty, due to it being outside of the interval of Formula (7). This does not include any material factors, piping, instrumentation or other cost factors.

Table 6.3: Estimations of pump costs, this does not include any material factors, piping, instrumentation or other cost factors.

Pump	Capacity [L s^{-1}]	Cost [\$]
High Pressure CO_2 Pump	7.8	14 300
Main Water Pump	9.1	14 600
Low Pressure Water Pump	4.8	13 500
District Heat Water Pump	66.2	27 600
Seawater Pump	2501	421 800

6.1.6 Flue Gas Desulfurization

For flue gas desulfurization a wet scrubber was chosen, using wastewater from seawater heat exchangers. Scaling was performed in a similar manner to that of the boiler for a flue gas wet desulfurization unit [28]. Using plant produced electricity in MW as a scaling variable, a cost of \$ 5 896 400 (on Jan. 2013 basis by using CEPCI [27]) was obtained. This includes piping, instrumentation and other cost factors.

6.1.7 Carbon Capture Facility

Cost of the carbon capture (CC) facility was based on a plant utilizing amine solution technology. The capital costs of the CC units were estimated as a given percentage of the capital cost of the PC plant without carbon capture. A conservative estimate may be found in Table 3.2 as 87%. Hence the capital cost of the CC facility was estimated to be \$ 97 880 000 (on Jan. 2013 basis by using CEPCI [27]). This includes piping, instrumentation and other cost factors.

6.1.8 Heat Pump Costs

For the heat pump case, heat pump costs were estimated using a cost estimate based on the thermal output of the heat pump [29]:

$$C = 14000 \frac{\mathrm{NOK}}{\mathrm{kW}} \cdot Q_{\mathrm{district\,heating}} \tag{8}$$

The estimated capital cost of the heat pump is NOK 168 000 000, which includes installation and other cost factors. This cost is not used in the base case.

6.1.9 Total Equipment Costs

In the total equipment costs, all equipment was modified using the following formula and factors from Sinnott [25]:

$$C = \sum_{i=1}^{i=M} C_{\mathrm{e,i,CS}} [(1 + f_{\mathrm{p}}) f_{\mathrm{m}} + (f_{\mathrm{er}} + f_{\mathrm{el}} + f_{\mathrm{i}} + f_{\mathrm{c}} + f_{\mathrm{s}} + f_{\mathrm{l}})] \tag{9}$$

where C is the total cost of the plant, including engineering costs, $C_{\mathrm{e,i,CS}}$ is purchased cost of equipment i in carbon steel, M is the total number of pieces of equipment, f_{p} is the installation factor for piping, f_{m} is the materical factor for exotic alloys, f_{er} is the installation factor for equipment erection, f_{el} is the installation factor for electrical work, f_{i} is the installation factor for instrumentation and process control, f_{c} is the installation factor for civil engineering work, f_{s} is the installation factor for structures and buildings and f_{l} is the installation factor for lagging, insulation, and paint. A typical value for the factors combined for a piece of equipment in carbon steel is 3.2, and 3.7 for 304 stainless steel.

For equipment handling CO_2 and seawater, 304 stainless steel was found to be resistant enough [25], while other pieces of equipment were estimated as carbon steel. To calculate the total fixed capital cost, C_{FC}, the following formula and factors from Sinnott were used [25]:

$$C_{\mathrm{FC}} = C(1 + \mathrm{OS})(1 + \mathrm{DE} + \mathrm{X}) \tag{10}$$

where OS is the offsite cost, DE is design and engineering cost and X is contingenacy costs. The total fixed capital costs are summarized in Table 6.4 on U.S. Gulf Coast Jan. 2013 basis.

Table 6.4: Total estimated equipment costs, with material factors, piping and other cost factors included.

Equipment	Cost [$]
Pulverized Coal Boiler	40 569 200
Heat Exchanger for District Heating	139 200
Vacuum Condenser	3 126 900
CO_2 Cooler	306 400
Low Pressure CO_2 Cooler	232 300
Intermediate Pressure CO_2 Condenser	161 600
Total Exchanger Costs	**3 966 500**
High Pressure Turbine	3 044 200
Intermediate Pressure Turbine	2 399 700
Low Pressure Turbine	5 270 100
Total Turbine Costs	**10 714 000**
Low Pressure CO_2 Compressor	9 760 700
Intermediate Pressure CO_2 Compressor	9 868 800
Total Compressor Costs	**19 629 500**
High Pressure CO_2 Pump	52 500
Main Water Pump	46 800
Low Pressure Water Pump	43 100
District Heat Water Pump	88 300
Seawater Pump	1 552 100
Total Pump Costs	**1 782 800**
Wet Flue Gas Desulfurization	5 896 400
Carbon Capture Facility	97 900 000
Total Fixed Capital Costs	341 028 400
Total Fixed Capital Costs [NOK]	**1 977 964 700**

The total equipment cost was calculated from $ U.S. Gulf Coast Jan. 2013 basis into Norwegian Krone 2013 basis (NOK) [30], yielding NOK 1 977 964 700 as total equipment costs. The corresponding estimate without a carbon capture facility was estimated as NOK 799 069 800.

For the heat pump case, the total equipment cost was estimated to be NOK 2 798 851 400.

6.2 Variable Costs

6.2.1 Labor

Number of required operators per shift is given by [31]:

$$N_{\text{operators}} = (6.29 + 0.23 \times N_{\text{Units}})^{0.5} \tag{11}$$

where N_{Units} is number of process units at the plant. With 12 process units, it yields 4 operators per shift.

Table 6.5: Estimation of the labor costs.

Labor	
Number of units	20
Number of operators	4
Shifts per day	6
Employed operators	24
Salary per year	NOK 433 000
Labor costs per year	**NOK 10 392 000**

6.2.2 Diesel Costs

Diesel is burned to generate power and district heating when the coal plant is down for maintenance and for peak loads. Yearly diesel costs are shown in Table 6.6.

Table 6.6: Estimated yearly diesel costs [32].

Diesel consumption	
Diesel consumption in 2012 [l]	390417
Diesel consumption in 2011 [l]	530287
Diesel consumption in 2010 [l]	72907
Diesel consumption in 2009 [l]	236728
Mean diesel consumption [l]	307585
Price of diesel [NOK/l]	NOK 12.00
Diesel costs per year	**NOK 3 691 000**

6.2.3 Cost of Coal

Yearly amount of coal burned was used together with price of coal. The coal price was found by matching Svalbard quality of coal with similar coal reserves [33]. This is used to calculate yearly cost of coal, shown in Table 6.7.

Table 6.7: Estimated yearly cost of coal.

Coal consumption	
Coal price [$/short ton]	65.5
Coal price [$/ton]	72.2
Coal consumption [ton/year]	60000
Coal costs per year	$ 4 332 000
Coal costs per year	**NOK 25 125 700**

6.2.4 Operation and Maintenance

Yearly operation and maintenance costs were calculated using statistical analysis of financial performance of earlier power plants [34], giving the estimated costs in Table 6.8.

Table 6.8: Estimated yearly costs of operation and maintenance.

Operations and Maintenance costs	
O&M [$/kW]	71
$W_{electric}$ [kW]	13421.5
Total O&M costs [$/year]	$ 952 900
Total O&M costs [NOK/year]	**NOK 5 527 000**

6.2.5 Chemicals

The estimation of yearly costs for chemicals was obtained by scaling of a PC plant with a carbon capture facility [8]. The results are summarized in Table 6.9.

Table 6.9: Estimated yearly costs of chemicals. This includes carbon, MEA solvent, lye, corrosion inhibitor, ammonia and other chemicals.

Cost of Chemicals	
Chemicals [$/kWh]	$ 0.00359
W_{electric} [kW]	13421.5
W [kWh]	117572753
Total O&M costs [$/year]	$ 422 100
Total O&M costs [NOK/year]	**NOK 2 448 100**

6.2.6 Total Variable Costs

By adding all the variable costs, the total yearly variable cost is estimated to be NOK 47 183 800.

6.3 Revenues

It is assumed that the price of electricity on Svalbard is 1.00 NOK/kWh, and that the price of district heating is 0.50 NOK/kWh. The revenue is estimated in Table 6.10.

Table 6.10: Estimated yearly revenues.

Revenue	
Price of electricity [NOK/kWh] [35]	NOK 1.00
Price of district heating [NOK/kWh] [35]	NOK 0.50
Total amount of electricity per year [kWh]	84 034 680
Total amount of district heating per year [kWh]	105 120 000
Revenue from electricity	NOK 84 034 700
Revenue from district heating	NOK 52 560 000
Total yearly revenue	**NOK 136 594 700**

6.4 Working Capital

The working capital was estimated as the cost of 60 days of raw material (coal and chemicals) for production and 2% of the total fixed investment cost (for spare parts), as recommended by NETL [8]. This estimation results in a working capital of NOK 40 673 800.

7 Investment Analysis

The investment analysis was done by estimating net present value (NPV) and internal rate of return (IRR) for each of the two cases (heat pump versus district heating). NPV was estimated using the following formula [25]:

$$\mathrm{NPV} = -C_{\mathrm{FC}} + \sum_{n=1}^{t} \frac{CF_n}{(1+i)^n} \tag{12}$$

where CF_n is the cash flow in year n, t is the project life in years and i is the interest rate in percent/100. The IRR was calculated by setting equation (12) equal to zero, and solving for i.

7.1 Base Case

Figure 7.1 shows the estimated NPV as a function of years after investment, using the estimated capital cost, variable costs, revenues and working capital: It has been assumed constant yearly costs, a constant depreciation rate of 10%, a constant 20% amount depreciation and 0% tax on Svalbard.

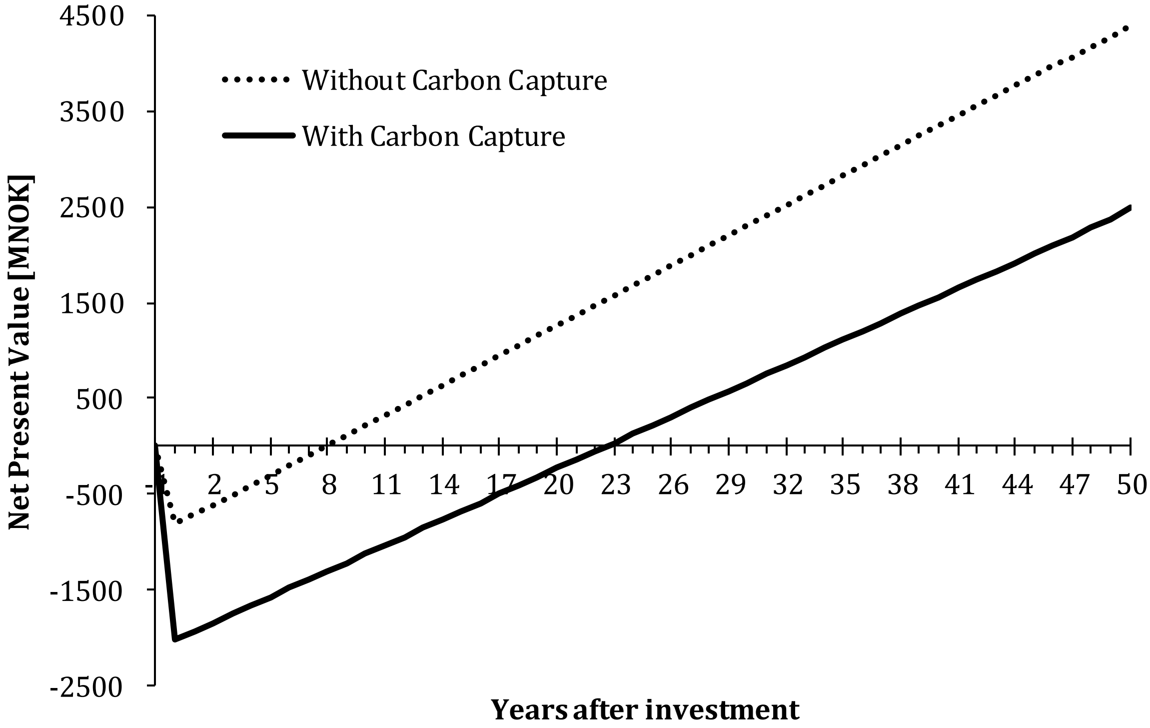

Figure 7.1: A plot showing the base case's net present value as a function of years after investment with and without carbon capture and storage.

The IRR of the base case project was calculated to be 3.8% with carbon capture, and 11.1% without carbon capture.

7.2 Heat Pump Case

Figure 7.2 shows the estimated NPV as a function of years after investment using the estimated capital cost, variable costs, revenues and working capital: It has been assumed constant yearly costs, a constant depreciation rate of 10%, a constant 20% amount depreciation and 0% tax on Svalbard.

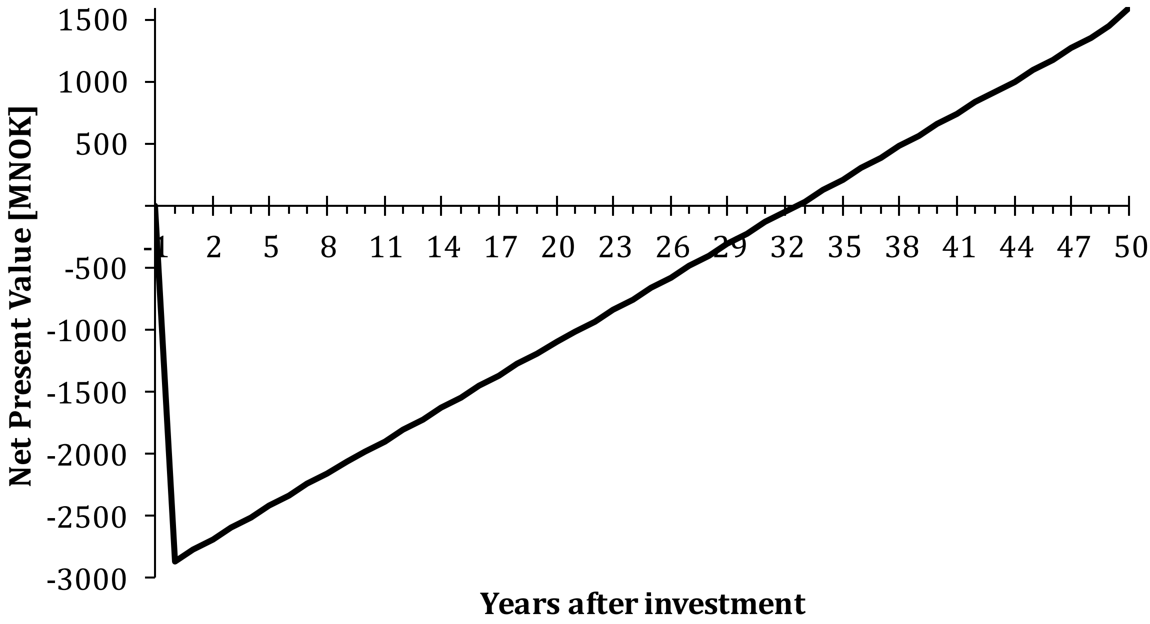

Figure 7.2: A plot showing the heat pump case's net present value as a function of years after investment.

The IRR of the project was calculated to be 1.9%.

The IRR was also estimated for a case where no district heating is produced, and yielded an IRR of 2.2% (which assumes that all the consumers will have to obtain their own heat pumps, or some other form of heating).

8 Discussion

8.1 Plant Choices

Many considerations have to be done regarding choice of plant. Weighing the high efficiency of Integrated Gasification Combined Cycle (IGCC) plants against the in-development Chemical Looping (CL) plants with simple carbon capture and storage (CCS) opportunities, or the already conventional pulverized coal (PC) plant. Choice of steam cycle is also important, as number of turbines and pressure levels highly affect the efficiency. Lastly, flue gas treatment has to be discussed, where the different options are available at different prices.

8.1.1 Plant Type

While IGCC and CL plants were considered in the beginning, they were found unsuitable for the planned project at Svalbard. These plants take advantage of size, as the air separation unit (ASU) and catalyzed reactors come at a large capital and variable cost. If, however, the plant is large enough, the ASU and reactor can repay themselves many-fold with the increased efficiency of the plant [8]. Both options also put strains on equipment, especially so with pre-combustion capture, where high purity hydrogen is combusted inside the turbine, yielding high temperatures. Using a gas turbine for combined cycle puts a lot of strain on gas treatment, as the turbines are sensitive to sulfur and carbon dioxide contents.

With its remote location, a power plant at Svalbard can easily find itself for long periods of time without spare parts, making well-developed technology the preferred choice. Instead of in-development technology, state of the art conventional technology was found to be the best fit. PC plants fit this choice well, as the idea of burning coal to yield steam is over 200 years old [36]. In a PC plant, the same principle applies, but the coal is pulverized into coal dust that burns more efficiently. In Appendix E composite curves for the boiler are shown, proving that the design is within the constraints of utilizable energy.

Steam turbine technology is well developed for PC plants, no gas turbine or reactor is required, and air is blown into the boiler, without need for compression or cryogenic distillation. All of these factors help keep the capital and variable costs low, giving a better investment perspective.

8.1.2 Steam Cycle

The current plant has a steam cycle design with two turbines, one of which is a condensing turbine, and the other is a backpressure turbine used to produce district heating. The proposed design has an extra turbine because they have a much higher steam pressure, as the current technology has a maximum allowable pressure drop. It also permits for reheating in between pressure levels which increases the efficiency [8]. A higher steam pressure also improves the overall efficiency, which is evident in Figure 5.4.

The efficiency may be further improved by allowing a higher maximum pressure, but due to support limitations on Svalbard, a more robust design was chosen. A higher steam temperature may also have a positive effect, as can be seen in Figure 5.2.

8.1.3 Flue Gas Treatment

The plant has to be within Norwegian emission regulations, putting requirements on the dust particles, sulfur, mercury and carbon dioxide output of the plant. As Svalbard already has plans for a Flue Gas Desulfurization unit (FGD) using seawater, a seawater scrubber was used in the design. A seawater scrubber is cheaper; both in capital cost, as the scrubber itself is cheap, and in variable costs as seawater is considered to be free [21]. A seawater scrubber would also deal with the mercury [23] and dust particles [8], which is harder to achieve using a dry scrubber [21]. Alternatively an electrostatic filter could be used to remove dust particles and mercury derivatives.

Svalbard also takes part in a project [37], aiming at a CO_2-free Svalbard by 2025, which requires the plant to have carbon dioxide capture and storage. As geological formations for storage exist in close vicinity of Longyearbyen, the plan is feasible. At atmospheric pressure, amine solution is preferred for capture, but puts further strain on the capital costs, as well as using heat from the boiler as re-boiler duty. During economic evaluation, CSS capital costs were estimated as a worst case scenario from Table 3.2, to be as much as 87% of the PC plant itself, with an electric power penalty of up to 29%.

8.2 Case Studies

8.2.1 Base Case

In the base case, it is assumed that the current district heating network on Svalbard can be used, which will reduce the capital cost significantly. District heating has the advantage of yielding high overall plant efficiency, because most of this heat is not feasible for production of electrical power. The current power plant was calculated, by Equation (4), to have an efficiency of 48.3%

(including diesel generators). The proposed design has an efficiency of 34.8%, which is quite high considering that it includes carbon capture.

8.2.2 Steam Temperature

A case study was performed on the model in Aspen HYSYS, yielding results which point towards a correlation that higher steam temperatures yield a higher power generation, and consequently a higher thermal efficiency. These results apply only for the power plant modeled, and may vary with varying steam cycles, boiler choices and plant size. Temperature considerations will have to be done, as the higher temperature will lead to higher heat exchanger area in the boiler and increased corrosion of the steam turbines [8]. Safety of the employees is also of concern, and risk analysis is important for choosing a desired design.

8.2.3 Steam Cycle Pressure

A case study was performed on the model in Aspen HYSYS, yielding results which point towards a correlation that higher pressure in the steam cycle gives a higher power generation, and consequently a higher thermal efficiency. These results apply only for the power plant modeled, and may vary with varying steam cycles, boiler choices and plant size. For the plant to use supercritical and ultra-supercritical pressure, equipment and safety considerations will have to be done. The increased capital cost from materials and complexity will have to be assessed. This is especially true as ultra-supercritical steam generation is still under development, and not available commercially [8]. It is still possible to build a pilot plant, but this will require a lot of expertise and support, which might not be suitable at a remote location such as Svalbard.

8.2.4 Heat Pump Case

In the heat pump case, a central heat pump at the plant was considered. The heat pump would provide hot water for the district heat network, and obtaining heat from the ocean. However, since Svalbard has a cold climate, a seawater heat pump would not be viable. Nevertheless, this option could become sustainable if geothermal heat is used instead of seawater. The efficiency of the design with seawater as the heat source is calculated to be 34.3% which is lower than the base case. In Figure 5.6, the efficiency is plotted against the coefficient of performance, and it is apparent that even a high performance heat pump would yield rather low efficiency. Although, higher than the base case with a sufficiently effective heat pump.

8.3 Investments

8.3.1 Cost estimations

In cost estimation of the PC boiler, it was assumed that capacity scaling was sufficient, as no price for a boiler in the right capacity range was found. The boiler used for scaling was almost 50 times larger, which results in a high uncertainty. The steam turbine costs were estimated in an equivalent manner, but with the scaling capacity much closer to the estimated steam turbine capacities. The FGD and heat pump costs were also estimated using the aforementioned method, and were assumed to moderately accurate, being inside a given capacity range.

Heat exchanger, compressor and pump costs were estimated using a slightly more accurate method, as described earlier. However, the seawater pump were larger than the stated interval, hence its cost estimation will be somewhat inaccurate.

The cost of the carbon capture facility was estimated as 87% of the total capital cost of the whole plant (without carbon capture), the worst case scenario from Table 3.2. This estimate might be too large, as there is a lot of ongoing research into improving the cost of carbon capture by amine absorption.

The total fixed capital cost of the plant was calculated the factorial method presented earlier. The factors were obtained from Sinnott [25], and might have some degree of uncertainty as the expenses will be higher on a remote location such as Svalbard.

The cost of labor was estimated from the number of units on the plant. Only operators were considered, and all other cost for other personnel (administrative, maintenance etc.) are not included. This is fairly inaccurate, and the total labor cost will probably be higher.

The usage of diesel was assumed to be constant, which might be wrong as the new plant may experience some difficulties during startup which will increase the need for backup power, and consequently diesel.

It is assumed that the coal have to be bought at a relatively high price because the net calorific value of the coal on Svalbard is high.

The cost of chemicals was estimated by scaling chemical requirements from a PC plant with CO_2 capture, and might have some uncertainty associated with it, due to the fact that the plant used for scaling is 50 times larger.

For the estimation of revenues it is assumed that the plant is operating at maximum capacity and that all the electricity and heating produced will be bought by the consumers. This assumption is inaccurate, as the electricity needed at Svalbard will not automatically double as soon as the new plant is installed.

The working capital is estimated as 60 days of coal supply, and 2% of the total plant cost for spare part etc. The National Energy Technology Laboratory recommends that only 0.5% of the total plant cost is needed for spare parts, however, Svalbard is a remote location and it was consequently assumed that it will need four times the amount of spare parts.

8.3.2 Investment analyses

The net present value (NPV) was estimated by assuming that the yearly expenses and revenues are constant throughout the whole lifetime of the project. This is incorrect, but yields a good indication of the project's value for comparing to other projects. In Figure 7.1 and Figure 7.2 it is evident that the base case has a larger NPV than the heat pump case, and will therefore be a better investment. In Figure 7.1, the NPV for the base case without district heating is shown, and it has a much higher NPV than the base case. This indicates that CCS is the economic bottleneck, and it might be worthwhile to wait for more efficient technology to become readily available.

The base case has a higher internal rate of return than the heat pump case, which is also an indication that it is the better investment. When it is assumed that the consumers obtain their heat by some other form than district heating, a higher IRR is obtained, but this is still lower than the base case. However, both cases have a huge NPV in the 50 year horizon-perspective, and if a sufficiently efficient heat pump is developed, it may be worth the extra investment.

9 Conclusion and Recommendations

A pulverized coal (PC) plant was found to be the best fit for a new power plant on Svalbard. The technology is commercially available, and no research and development is required. Oxygen-fired combustion is a lucrative option, as the carbon capture is more efficient, but it will have to be developed further before being implemented at a secluded location as Svalbard.

A maximum boiler temperature of 800°C was assumed, together with subcritical pressure in the steam cycle. It is recommended to do further studies, as it was shown that increased temperature and pressure gives higher efficiency. Supercritical pressures are already conventional, but a plant at Svalbard needs to consider its isolated location. The boiler has been greatly simplified; hence further studies are needed for obtaining actual combustion temperature and heat exchanging possibilities.

District heating from a backpressure steam turbine was found to be a better option than a central heat pump, both practically and economically. Even if consumers obtain their own heat pumps, the base case is preferable.

Carbon capture and storage (CCS) is also a field largely in development. Price of equipment is expected to fall, and efficiency is expected to rise. CCS is the bottleneck in both economical and efficiency-wise for the power plant, and was studied further in another project.

References

[1] Heat Pump & Thermal Storage Technology Center of Japan, "Data Book on Heat Pump & Thermal Storage Systems 2011-2012," HPTCJ, 2011.

[2] H. Termuehlen and W. Emsperger, Clean and Efficient Coal-Fired Power Plants, New York: ASME Press, 2003, pp. 1-2.

[3] International Energy Agency, "Key World Energy Statistics," 2013. [Online]. Available: http://www.iea.org/publications/freepublications/publication/KeyWorld2013_FINAL_WEB.pdf. [Accessed 18 October 2013].

[4] International Energy Agency, "Medium-Term Coal Market Report 2012," IEA Publications, Paris, 2012.

[5] Intergovernmental Panel on Climate Change, "Climate Change 2013: The Physical Science Basis: Headline Statements from the Summary for Policymakers," 27 September 2013. [Online]. Available: http://www.ipcc.ch/news_and_events/docs/ar5/ar5_wg1_headlines.pdf. [Accessed 19 October 2013].

[6] Clean Air Technology Center, "Nitrogen Oxides (NOx), Why and How They Are Controlled," United States Environmental Protection Agency, Durham, 1999.

[7] J. U. Watts, A. N. Mann and D. L. Russel Sr., "An Overview of NOx Control Technologies Demonstrated under the Department of Energy's Clean Coal Technology Program," National Energy Technology Laboratory, Pittsburg, 2000.

[8] National Energy Technology Laboratory, "Cost and Performance Baseline for Fossil Energy Plants," Pittsburgh, 2013.

[9] O. Maurstad, "An Overview of Coal Based Integrated Gasification Combined Cycle (IGCC) Technology," Cambridge, 2005.

[10] B. J. P. Buhre, L. K. Elliott, C. D. Sheng, R. P. Gupta and T. F. Wall, "Oxy-fuel combustion technology for coal-fired power generation," *Progress in Energy and Combustion Science,* vol. 31, no. 4, pp. 283-307, 2005.

[11] Y. Oki, J. Inumaru, S. Hara, M. Kobayashi, H. Watanabe, S. Umemoto and H. Makino, "Development of oxy-fuel IGCC system with CO2 recirculation for CO2 capture," *Energy Procedia,* vol. 4, pp. 1066-1073, 2011.

[12] International Energy Agency, "CO2 Capture and Storage," OECD, 2008.

[13] D. Berstad, R. Anantharaman and P. Nekså, "Low-temperature CO2 capture technologies - Applications and potential," *International Journal of Refrigeration,* vol. 36, no. 5, pp. 1403-1416, 2013.

[14] H. Zhai and E. S. Rubin, "Membrane-based CO2 Capture Systems for Coal-fired Power Plants," National Energy Technology Laboratory, Pittsburgh, 2012.

[15] M. Blunt, "Carbon Dioxide Storage," Imperial College of London, London, 2010.

[16] E. S. Rubin, C. Chen and A. B. Rao, "Cost and Performance of Fossil Fuel Power Plants with CO2 Capture and Storage," *Energy Policy,* vol. 35, no. 9, pp. 4444-4454, 2007.

[17] E. S. Rubin and A. B. Rao, "A Technical, Economic and Environmental Assessment of Amine-Based CO2 Capture Technology for Power Plant Greenhouse Gas Control," *Environmental Science & Technology,* vol. 36, no. 20, pp. 4467-4475, 2002.

[18] G. T. Rochelle, "Amine Scrubbing for CO2 Capture," *Science,* vol. 325, no. 5948, pp. 1652-1654, 25 September 2009.

[19] R. K. Varagani, F. Châtel-Pélage, P. Pranda, M. Rostam-Abadi, Y. Lu and A. C. Bose, "Performance Simulation and Cost Assessment of Oxy-Combustion Process for CO2 Capture from Coal-Fired Power Plants," in *Fourth Annual Conference on Carbon Sequestration*, Alexandria, 2005.

[20] E. S. Rubin, C. Chen and A. B. Rao, "Comparative Assessment of Fossil Fuel Power Plants with CO2 Capture and Storage," in *Proceedings of 7th Internation Conference on Greenhouse Gas Control Technologies*, Vancouver, 2005.

[21] R. K. Srivastava, J. Wojciech and C. Singer, "SO2 Scrubbing Technologies: A Review," *Environmental Progress,* vol. 20, no. 4, pp. 219-228, December 2001.

[22] H. Kettner, "The Removal of Sulfur Dioxide from Flue Gases," *Bulletin of The World Health Organization,* vol. 32, no. 3, p. 421, 1965.

[23] C. E. Miller, T. J. Feeley, III, W. W. Aljoe, B. W. Lani, K. T. Schroeder, C. Kairies, A. T. McNemar, A. P. Jones and J. T. Murphy, "Mercury Capture and Fate Using Wet FGD at Coal-Fired Power Plants," U.S. Department of Energy, Pittsburgh, 2006.

[24] Power Technology, "Power Technology - A future for coal power," [Online]. Available: http://www.power-technology.com/. [Accessed 17 November 2013].

[25] R. Sinnott and G. Towler, Chemical Engineering Design, 5th ed., Burlington, Massachusetts: Elsevier, 2009, pp. 301, 313, 819, 1101.

[26] M. Sciazko and T. Chmielniak, "Chapter 10: Cost Estimates of Coal Gasification for Chemicals and Motor Fuels," in *Gasification for Practical Applications*, 2012, p. 263.

[27] Chemical Engineering, "Economic Indicators," *Chemical Engineering,* pp. 71-72, April 2013.

[28] U.S. Enviromental Protection Agency, "Air Pollution Control Technology Fact Sheet," U.S. Enviromental Protection Agency, 2002.

[29] The International Renewable Energy Agency, "Heat Pumps: Technology Brief," IRENA, 2013.

[30] XE Currency, "USD/NOK-kart," November 2013. [Online]. Available: http://www.xe.com/currencycharts/?from=USD&to=NOK&view=1Y. [Accessed November 2013].

[31] R. Turton, R. C. Bailie, W. B. Whiting and J. A. Shaewitz, Analysis, Synthesis and Design for Chemical Processes, 2nd ed., Hamilton in Castleton, New York: Prentice Hall, 2002, p. 193.

[32] Norsk Petroleumsinstitutt, "Statistikk," 24 June 2013. [Online]. Available: http://www.np.no/priser/. [Accessed 10 November 2013].

[33] U.S. Energy Information Administration, "Coal News and Markets Report," 4 November 2013. [Online]. Available: http://www.eia.gov/coal/news_markets/. [Accessed 10 November 2013].

[34] L. Giorgio and M. Mauro, "Small–medium sized nuclear coal and gas power plant: A probabilistic analysis of their financial performances and influence of CO2 cost," *Energy Policy,* p. 6360–6374, 2010.

[35] R. Blomstrand, Interviewee, *Longyearbyen Board Member.* [Interview]. 28 10 2013.

[36] H. Termuehlen and W. Emsperger, Clean and Efficient Coal-Fired Power Plants: Development Toward Advanced Technologies, 1st ed., New York, New York: The American Society of Mechanical Engineers, 2003, pp. 5-25.

[37] Nordic Network for Sustainable Energy Systems in Isolated Locations, "NordSESIL.net," 2006. [Online]. Available: http://nordsesil.wikispaces.com/CO2+free+Svalbard+by+2025. [Accessed 16 November 2013].

[38] F. Holland, F. Watson and J. Wilkinson, Perry's Handbook, 6th ed., New York: McGraw Hill,, 1984.

[39] Statistisk Sentralbyrå, "Earnings in electricity supply," 14 March 2013. [Online]. Available: http://www.ssb.no/en/arbeid-og-lonn/statistikker/lonnkraft/aar/2013-03-14. [Accessed 9 November 2013].

[40] U.S. Department Of Labor, "Consumer Price Index," October 30 2013. [Online]. Available: ftp://ftp.bls.gov/pub/special.requests/cpi/cpiai.txt. [Accessed 9 November 2013].

List of symbols and abbreviations

Symbol	Unit	Description
ASU		Air Separation Unit
a	\$	Constant for Cost Estimation
b		Constant for Cost Estimation
C	\$	Cost
$C_{\mathrm{e,i,CS}}$	\$	Purchased Cost of Equipment in Carbon Steel
C_{FC}	\$	Total Fixed Capital Cost
C_1	\$	Cost of Plant with Capacity S_1
C_2	\$	Cost of Plant with Capacity S_2
CC		Carbon Capture
CCS		Carbon Capture and Storage
CEPCI		Chemical Engineering Plant Cost Index
CL		Chemical Looping
COP		Cost of Electricity
DE	\$	Design and Engineering Cost
FGD		Flue Gas Desulfurization
FGT		Flue Gas Treatment
f_{c}		Installation Factor for Civil Engineering Work
f_{el}		Installation Factor for Electrical Work
f_{er}		Installation Factor for Equipment Erection
f_{i}		Installation Factor for Instrumentation and Process Control
f_{l}		Installation Factor for Lagging, Insulation and Paint
f_{m}		Installation Factor for Exotic Alloys
f_{p}		Installation Factor for Piping
f_{s}		Installation Factor for Structures and Buildings
HEX		Heat Exchanger
HRSG		Heat Recovery System Generator
HP		High Pressure
IGCC		Integrated Gasification Combined Cycle
IP		Intermediate Pressure
IRR		Internal Rate of Return
LP		Low Pressure
M		Total Number of Pieces of Equipment
MEA		Monoethanolamine
m_{C}	$\mathrm{kg\,s^{-1}}$	Mass flow of Coal
$N_{\mathrm{operators}}$		Number of Operators
N_{Units}		Number of Units
NCV	$\mathrm{cal\,g^{-1}}$	Net Colorific Value
NGCC		Natural Gas Combined Cycle
NOK		Norwegian Krone
NPV	MNOK	Net Present Value
n		Exponent Factor for Order of Magnitude
OECD		Organization for Economic Co-operation and Development
OS	\$	Offsite Costs
O&M	$\$\,\mathrm{kW^{-1}}$	Operation and Maintenance Cost
PC		Pulverized Coal
$Q_{\mathrm{district\,heat}}$	MW	Power provided to District Heating
S_1		Capacity of Plant with Cost C_1
S_2		Capacity of Plant with Cost C_2
sC		Subcritical
SC		Supercritical

Symbol	Unit	Description
SCR		Selective Catalytic Reduction
U	$W\,m^{-2}K^{-1}$	Overall Heat Transfer Coefficient
$W_{electric}$	MW	Net Electric Power Produced
X	$	Contingency Cost
x_i		Mole Fraction of Component i
$\eta_{thermal}$		Thermal Efficiency

Appendix A - Cost estimation for the base case

A.1 Cost of major equipment

The cost estimation of the major equipment is shown in Table A.1.

Table A.1: Cost estimation of the major equipment.

Boiler	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	13.42154714
C1	$ 339 189 000.00
S1 [MW]	550
I2	630.2
I1	525.4
Capital Cost of Boiler	**$ 33 807 637.06**

HP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	2.990730163
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of HP Turbine	**$ 951 313.24**

IP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	2.096839296
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of IP Turbine	**$ 749 896.70**

LP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	6.784556249
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of LP Turbine	**$ 1 646 918.71**

District Heat Exchanger	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	126061.1955
U [W/m2 K]	1200
A [m2]	105.0509963
I2	630.2
I1	525.4
Capital Cost of District Heat Exchanger	**$ 43 490.90**

CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	15645.45174
U [W/m2 K]	50
A [m2]	312.9090347
I2	630.2
I1	525.4
Capital Cost of CO2 Cooler	**$ 83 268.56**

Vacuum Condensers	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	4724201.521
U [W/m2 K]	1200
A [m2]	3936.834601
Number of condensers	4
A per condenser [m2]	984.2086503
I2	630.2
I1	525.4
Capital Cost per Condenser	$ 244 288.95
Capital Cost of Vacuum Condensers	**$ 977 155.79**

LP CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	21301.19111
U [W/m2 K]	100
A [m2]	213.0119111
I2	630.2
I1	525.4
Capital Cost of LP CO2 Cooler	**$ 63 129.62**

IP CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	75330.38298
U [W/m2 K]	700
A [m2]	107.6148328
I2	630.2
I1	525.4
Capital Cost of IP CO2 Cooler	**$ 43 922.57**

LP CO2 Compressor	**(490000+16800*P^0.6)*(I2/I1)**
P [kW]	960.7527304
I2	913.9
I1	525.4
Capital Cost of LP CO2 Compressor	**$ 2 652 375.18**

IP CO2 Compressor	**(490000+16800*P^0.6)*(I2/I1)**
P [kW]	987.014704
I2	913.9
I1	525.4
Capital Cost of IP CO2 Compressor	**$ 2 681 738.22**

HP CO2 Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	7.801010982
I2	913.9
I1	525.4
Capital Cost of HP CO2 Pump	**$ 14 278.32**

Main Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	9.116187496
I2	913.9
I1	525.4
Capital Cost of Main Water Pump	**$ 14 620.95**

LP Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	4.768238621
I2	913.9
I1	525.4
Capital Cost of LP Water Pump	**$ 13 463.61**

District Heat Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	66.18057501
I2	913.9
I1	525.4
Capital Cost of District Heat Water Pump	**$ 27 595.24**

Seawater Pump	**(6900+206*V^0.9)*(I2/I1)**
qtot [l/s]	2500.640907
I2	913.9
I1	525.4
Capital Cost of Seawater Pump	**$ 421 755.00**

Flue Gas Desulfurization	**C*P*(I2/I1)**
P [kW]	13421.54714
C [$/kW]	250
I2	692.9
I1	394.3
Capital Cost of Flue Gas Desulfurization	**$ 5 896 392.35**

CO2 Removal	**C = x*C1*(I2/I1)**
Capital Cost Without Carbon Capture [$]	$ 156 035 250.47
Percent Increase with Carbon Capture	87 %
I2	499.6
I1	692.9
Capital Cost of CO2 Removal	**$ 97 879 973.57**

Installation factors	
Equipment erection factor, f_{er}	0.5
Piping factor, f_p	0.6
Instrumentation and Control factor, f_i	0.3
Electrical factor, f_{el}	0.2
Civil factor, f_c	0.3
Structures and Building factor, f_s	0.2
Lagging and Paint factor, $_{fl}$	0.1
Material, f_m: Carbon steel	1
Material, f_m: Stainless steal	1.3
Boiler	1.2
HP Turbine	3.2
IP Turbine	3.2
LP Turbine	3.2
District Heat Exchanger	3.2
Vacuum Condenser	3.2
CO2 Cooler	3.68
LP CO2 Cooler	3.68
IP CO2 Cooler	3.68
LP CO2 Compressor	3.68
IP CO2 Compressor	3.68
HP CO2 Pump	3.68
Main Water Pump	3.2
LP Water Pump	3.2
District Heat Water Pump	3.2
Seawater Pump	3.68
Flue Gas Desulfurization	1
CO2 Removal	1

Installed Capital Costs	
Boiler	$ 40 569 164.47
HP Turbine	$ 3 044 202.35
IP Turbine	$ 2 399 669.44
LP Turbine	$ 5 270 139.88
District Heat Exchanger	$ 139 170.88
Vacuum Condenser	$ 3 126 898.52
CO2 Cooler	$ 306 428.31
LP CO2 Cooler	$ 232 317.02
IP CO2 Cooler	$ 161 635.05
LP CO2 Compressor	$ 9 760 740.66
IP CO2 Compressor	$ 9 868 796.65
HP CO2 Pump	$ 52 544.21
Main Water Pump	$ 46 787.03
LP Water Pump	$ 43 083.55
District Heat Water Pump	$ 88 304.78
Seawater Pump	$ 1 552 058.42
Flue Gas Desulfurization	$ 5 896 392.35
Offsites	0.4
Design and Engineering	0.25
Contingency	0.1
Total fixed capital cost (without CO2 removal)	$ 156 035 250.47
	NOK 905 004 452.10
Total fixed capital cost (with CO2 removal)	$ 341 028 400.52
	NOK 1 977 964 723.00

A.2 Variable costs

The estimation of the variable costs is shown in Table A.2.

Table A.2: Estimation of the variable costs.

Labor	
Number of units	20
Number of operators	4
Shifts per day	6
Employed operators	24
Salary per year	NOK 433 000.00
Labor costs per year	**NOK 10 392 000.00**

Diesel consumption	
Diesel consumption in 2012 [l]	390417
Diesel consumption in 2011 [l]	530287
Diesel consumption in 2010 [l]	72907
Diesel consumption in 2009 [l]	236728
Mean diesel consumption [l]	307584.75
Price of diesel [NOK/l]	NOK 12.00
Diesel costs per year	**NOK 3 691 017.00**

Coal consumption	
Coal price [$/short ton]	65.5
Coal price [$/ton]	72.20017637
Coal consumption [ton/year]	60000
Coal costs per year	$ 4 332 010.58
Coal costs per year	**NOK 25 125 661.38**

Operations and Maintenance costs	
O&M [$/kW]	71
P [kW]	13421.54714
Total O&M costs [$/year]	$ 952 929.85
Total O&M costs [NOK/year]	**NOK 5 526 993.11**

Cost of Chemicals	
Chemicals [$/kWh]	$ 0.00359
P [kW]	13421.5
W [kWh]	117572753
Total O&M costs [$/year]	$ 422 086.18
Total O&M costs [NOK/year]	**NOK 2 448 099.86**

Total	
Total variable costs [$/year]	**NOK 47 183 771.35**

A.3 Revenues

The estimation of the yearly revenue is shown in Table A.3.

Table A.3: Estimation of the yearly revenue.

Revenue	
Price of electricity [NOK/kWh]	NOK 1.00
Price of district heating [NOK/kWh]	NOK 0.50
Total amount of electricity [kWh]	84034680
Total amount of district heating [kWh]	105120000
Revenue from electricity	NOK 84 034 680.00
Revenue from district heating	NOK 52 560 000.00
Total revenue	**NOK 136 594 680.00**

A.4 Working capital

The estimation of the working capital is shown in Table A.4.

Table A.4: The estimation of the working capital.

Working capital	
Value of raw materials in inventory (60 days)	NOK 1 114 538.70
Spare parts (2 % of total plant cost)	NOK 39 849 181.64
Total Working Capital	**NOK 40 963 720.34**

Appendix B - Cost estimation for the heat pump case

B.1 Major Equipment

The cost estimation of the major equipment is shown in Table B.1.

Table B.1: Cost estimation of the major equipment.

Boiler	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	13.42154714
C1	$ 339 189 000.00
S1 [MW]	550
I2	630.2
I1	525.4
Capital Cost of Boiler	**$ 33 807 637.06**

HP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	2.99386541
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of HP Turbine	**$ 951 981.30**

IP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	0
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of IP Turbine	**$ -**

LP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	12.65709666
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of LP Turbine	**$ 2 501 016.97**

District Heat Exchanger	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	0
U [W/m2 K]	1200
A [m2]	0
I2	630.2
I1	525.4
Capital Cost of District Heat Exchanger	$ -

Vacuum Condensers	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	12261753.54
U [W/m2 K]	1200
A [m2]	10218.12795
Number of condensers	10
A per condenser [m2]	1021.812795
I2	630.2
I1	525.4
Capital Cost per condenser	**$ 254 206.86**
Capital Cost of Vacuum Condensers	**$ 2 542 068.57**

CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	14777.81911
U [W/m2 K]	50
A [m2]	295.5563823
I2	630.2
I1	525.4
Capital Cost of CO2 Cooler	**$ 79 663.40**

LP CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	22420.49441
U [W/m2 K]	100
A [m2]	224.2049441
I2	630.2
I1	525.4
Capital Cost of LP CO2 Cooler	**$ 65 306.34**

IP CO2 Cooler	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	88258.8128
U [W/m2 K]	700
A [m2]	126.0840183
I2	630.2
I1	525.4
Capital Cost of IP CO2 Cooler	**$ 47 090.88**

LP CO2 Compressor	**(490000+16800*P^0.6)*(I2/I1)**
P [kW]	960.7527304
I2	913.9
I1	525.4
Capital Cost of LP CO2 Compressor	**$ 2 652 375.18**

IP CO2 Compressor	**(490000+16800*P^0.6)*(I2/I1)**
P [kW]	960.7527304
I2	913.9
I1	525.4
Capital Cost of IP CO2 Compressor	**$ 2 652 375.18**

HP CO2 Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	7.801010982
I2	913.9
I1	525.4
Capital Cost of HP CO2 Pump	**$ 14 278.32**

Main Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	8.895505456
I2	913.9
I1	525.4
Capital Cost of Main Water Pump	**$ 14 563.82**

LP Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	0
I2	913.9
I1	525.4
Capital Cost of LP Water Pump	**$ -**

District Heat Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	0
I2	913.9
I1	525.4
Capital Cost of District Heat Water Pump	**$ -**

Seawater Pump	**(6900+206*V^0.9)*(I2/I1)**
qtot [l/s]	3109.246726
I2	913.9
I1	525.4
Capital Cost of Seawater Pump	**$ 510 502.50**

Flue Gas Desulfurization	**C*P*(I2/I1)**
P [kW]	13421.54714
C [$/kW]	250
I2	692.9
I1	394.3
Capital Cost of Flue Gas Desulfurization	**$ 5 896 392.35**

CO2 Removal	**C = x*C1*(I2/I1)**
Capital Cost Without Carbon Capture [$]	$ 220 792 351.45
Percent Increase with Carbon Capture	87 %
I2	499.6
I1	692.9
Capital Cost of CO2 Removal	**$ 138 501 713.29**

Central heat pump	
Cost per kW [NOK/kW]	NOK 14 000.00
Heat provided [kW]	12000
Capital Cost of Central heat pump	**NOK 168 000 000.00**

Installation factors	
Equipment erection factor, f_{er}	0.5
Piping factor, f_p	0.6
Instrumentation and Control factor, f_i	0.3
Electrical factor, f_{el}	0.2
Civil factor, f_c	0.3
Structures and Building factor, f_s	0.2
Lagging and Paint factor, $_{fl}$	0.1
Material, f_m: Carbon steel	1
Material, f_m: Stainless steal	1.3
Boiler	1.2
HP Turbine	3.2
IP Turbine	3.2
LP Turbine	3.2
District Heat Exchanger	3.2
Vacuum Condenser	3.2
CO2 Cooler	3.68
LP CO2 Cooler	3.68
IP CO2 Cooler	3.68
LP CO2 Compressor	3.68
IP CO2 Compressor	3.68
HP CO2 Pump	3.68
Main Water Pump	3.2
LP Water Pump	3.2
District Heat Water Pump	3.2
Seawater Pump	3.68
Flue Gas Desulfurization	1
CO2 Removal	1
Heat Pump	1

Installed Capital Costs	
Boiler	$ 40 569 164.47
HP Turbine	$ 3 046 340.16
IP Turbine	$ -
LP Turbine	$ 8 003 254.32
District Heat Exchanger	$ -
Vacuum Condenser	$ 8 134 619.43
CO2 Cooler	$ 293 161.31
LP CO2 Cooler	$ 240 327.32
IP CO2 Cooler	$ 173 294.44
LP CO2 Compressor	$ 9 760 740.66
IP CO2 Compressor	$ 9 760 740.66
HP CO2 Pump	$ 52 544.21
Main Water Pump	$ 46 604.22
LP Water Pump	$ -
District Heat Water Pump	$ -
Seawater Pump	$ 1 878 649.18
Flue Gas Desulfurization	$ 5 896 392.35
Heat Pump	$ 28 965 517.24
Offsites	0.4
Design and Engineering	0.25
Contingency	0.1
Total fixed capital cost (without CO2 removal)	$ 220 792 351.45
	NOK 1 280 595 638.40
Total fixed capital cost (with CO2 removal)	$ 482 560 589.57
	NOK 2 798 851 419.53

B.2 Variable Costs

The estimation of the variable costs is shown in Table B.2.

Table B.2: Estimation of the variable costs

Labor	
Number of units	18
Number of operators	4
Shifts per day	6
Employed operators	24
Salary per year	NOK 433 000.00
Labor costs per year	**NOK 10 392 000.00**

Diesel consumption	
Diesel consumption in 2012 [l]	390417
Diesel consumption in 2011 [l]	530287
Diesel consumption in 2010 [l]	72907
Diesel consumption in 2009 [l]	236728
Mean diesel consumption [l]	307584.75
Price of diesel [NOK/l]	NOK 12.00
Diesel costs per year	**NOK 3 691 017.00**

Coal consumption	
Coal price [$/short ton]	65.5
Coal price [$/ton]	72.20017637
Coal consumption [ton/year]	60000
Coal costs per year	$ 4 332 010.58
Coal costs per year	**NOK 25 125 661.38**

Operations and Maintenance costs	
O&M [$/kW]	71
P [kW]	13421.54714
Total O&M costs [$/year]	$ 952 929.85
Total O&M costs [NOK/year]	**NOK 5 526 993.11**

Cost of Chemicals	
Chemicals [$/kWh]	$ 0.00359
P [kW]	13421.5
W [kWh]	117572753
Total O&M costs [$/year]	$ 422 086.18
Total O&M costs [NOK/year]	**NOK 2 448 099.86**

Total	
Total variable costs [NOK/year]	**NOK 47 183 771.35**

B.3 Revenues

The estimation of the yearly revenue is shown in Table B.3.

Table B.3: Estimation of the yearly revenue.

Revenue	
Price of electricity [NOK/kWh]	NOK 1.00
Price of district heating [NOK/kWh]	NOK 0.50
Total amount of electricity [kWh]	117412204
Total amount of district heating [kWh]	105120000
Heat pump COP	3
Electricity needed for heat pump	35040000
Available electricity	82372204
Revenue from electricity	NOK 82 372 204.19
Revenue from district heating	NOK 52 560 000.00
Total revenue	**NOK 134 932 204.19**

B.4 Working Capital

The estimation of the working capital is shown in Table B.4.

Table B.4: Estimation of the working capital

Working capital	
Value of raw materials in inventory (60 days)	NOK 1 114 538.70
Spare parts (2 % of total plant cost)	NOK 55 977 028.39
Total Working Capital	**NOK 57 091 567.09**

Appendix C - Cost estimation for the base case without carbon capture

C.1 Cost of major equipment

The cost estimation of the major equipment is shown in Table C.1.

Table C.1: Cost estimation of the major equipment.

Boiler	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	13.42154714
C1	$ 339 189 000.00
S1 [MW]	550
I2	630.2
I1	525.4
Capital Cost of Boiler	**$ 33 807 637.06**

HP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	5.02153665
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of HP Turbine	**$1 346 211.64**

IP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	2.096839296
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of IP Turbine	**$ 749 896.70**

LP Turbine	**C2 = C1(S2/S1)^0.67*(I2/I1)**
S2 [MW]	15.3701151285023
C1	$ 834 000.00
S1 [MW]	3
I2	657.7
I1	575.4
Capital Cost of LP Turbine	**$ 2 848 567.17**

District Heat Exchanger	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	126061.1955
U [W/m2 K]	1200
A [m2]	105.0509963
I2	630.2
I1	525.4
Capital Cost of District Heat Exchanger	**$ 43 490.90**

Vacuum Condensers	**(24000+46*A^1.2)*(I2/I1)**
UA [W/K]	10702471.7025945
U [W/m2 K]	1200
A [m2]	8918.726419
Number of condensers	9
A per condenser [m2]	990.9696021
I2	630.2
I1	525.4
Capital Cost per Condenser	$ 246 066.61
Capital Cost of Vacuum Condensers	**$ 2 214 599.52**

Main Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	15.13710685
I2	913.9
I1	525.4
Capital Cost of Main Water Pump	**$ 16 138.04**

LP Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	10.8022358
I2	913.9
I1	525.4
Capital Cost of LP Water Pump	**$ 15 053.08**

District Heat Water Pump	**(6900+206*V^0.9)*(I2/I1)**
q [l/s]	66.18057501
I2	913.9
I1	525.4
Capital Cost of District Heat Water Pump	**$ 27 595.24**

Seawater Pump	**(6900+206*V^0.9)*(I2/I1)**
qtot [l/s]	5665.09248
I2	913.9
I1	525.4
Capital Cost of Seawater Pump	**$ 867 388.10**

Flue Gas Desulfurization	**C*P*(I2/I1)**
P [kW]	13421.54714
C [$/kW]	250
I2	692.9
I1	394.3
Capital Cost of Flue Gas Desulfurization	**$ 5 896 392.35**

Installation factors	
Equipment erection factor, f_{er}	0.5
Piping factor, f_p	0.6
Instrumentation and Control factor, f_i	0.3
Electrical factor, f_{el}	0.2
Civil factor, f_c	0.3
Structures and Building factor, f_s	0.2
Lagging and Paint factor, $_{fl}$	0.1
Material, f_m: Carbon steel	1
Material, f_m: Stainless steal	1.3
Boiler	1.2
HP Turbine	3.2
IP Turbine	3.2
LP Turbine	3.2
District Heat Exchanger	3.2
Vacuum Condenser	3.2
Main Water Pump	3.2
LP Water Pump	3.2
District Heat Water Pump	3.2
Seawater Pump	3.68
Flue Gas Desulfurization	1

Installed Capital Costs	
Boiler	$ 40 569 164.47
HP Turbine	$ 4 307 887.24
IP Turbine	$ 2 399 669.44
LP Turbine	$ 9 115 414.95
District Heat Exchanger	$ 139 170.88
Vacuum Condenser	$ 7 086 718.48
Main Water Pump	$ 51 641.72
LP Water Pump	$ 48 169.87
District Heat Water Pump	$ 88 304.78
Seawater Pump	$ 3 191 988.22
Flue Gas Desulfurization	$ 5 896 392.35
Offsites	0.4
Design and Engineering	0.25
Contingency	0.1
Total fixed capital cost (without CO2 removal)	$ 137 770 641.27
	NOK 799 069 719.38

C.2 Variable Costs

The estimation of the variable costs is shown in Table C.2.

Table C.2: Estimation of the variable costs

Labor	
Number of units	10
Number of operators	3
Shifts per day	6
Employed operators	18
Salary per year	NOK 433 000.00
Labor costs per year	**NOK 7 794 000.00**
Diesel consumption	
Diesel consumption in 2012 [l]	390417
Diesel consumption in 2011 [l]	530287
Diesel consumption in 2010 [l]	72907
Diesel consumption in 2009 [l]	236728
Mean diesel consumption [l]	307584.75
Price of diesel [NOK/l]	NOK 12.00
Diesel costs per year	**NOK 3 691 017.00**
Coal consumption	
Coal price [$/short ton]	65.5
Coal price [$/ton]	72.20017637
Coal consumption [ton/year]	31000
Coal costs per year	$ 2 238 205.47
Coal costs per year	**NOK 12 981 591.71**
Operations and Maintenance costs	
O&M [$/kW]	71
P [kW]	13421.54714
Total O&M costs [$/year]	$ 952 929.85
Total O&M costs [NOK/year]	**NOK 5 526 993.11**
Cost of Chemicals	
Chemicals [$/kWh]	$ 0.00359
P [kW]	13421.5
W [kWh]	117572753
Total O&M costs [$/year]	$ 422 086.18
Total O&M costs [NOK/year]	**NOK 2 448 099.86**
Total	
Total variable costs [NOK/year]	**NOK 32 441 701.69**

C.3 Revenues

The estimation of the yearly revenue is shown in Table C.3.

Table C.3: Estimation of the yearly revenue.

Revenue	
Price of electricity [NOK/kWh]	NOK 1.00
Price of district heating [NOK/kWh]	NOK 0.50
Total amount of electricity [kWh]	83771908.33
Total amount of district heating [kWh]	105120000
Heat pump COP	3
Electricity needed for heat pump	35040000
Available electricity	82372204
Revenue from electricity	NOK 83 771 908.33
Revenue from district heating	NOK 52 560 000.00
Total revenue	**NOK 136 331 908.33**

C.4 Working Capital

The estimation of the working capital is shown in Table C.4.

Table C.4: Estimation of the working capital

Working capital	
Value of raw materials in inventory (60 days)	NOK 770 351.56
Spare parts (2 % of total plant cost)	NOK 15 981 394.39
Total Working Capital	**NOK 16 751 745.95**

Appendix D Net Calorific Value of the Coal on Svalbard

The net calorific value and contents of the coal are shown in Figure D.1.

Longyearbyen 06.09.2013

Periode: 2013 08
Produsert i perioden: 3 807 tonn

Samtlige analyseresultat er oppgitt på tørr basis.

Småkull:

Uke	Tonn	Aske	Flykt	Fix. C	Svovel	NCV	FSI
2013 31	162	8,7 %	37,2 %	53,5 %	0,59 %	7453	8,0
2013 32	243	8,2 %	38,3 %	52,8 %	0,66 %	7522	8,5
2013 33	351	10,7 %	36,9 %	51,7 %	0,72 %	7311	8,0
2013 34	486	7,8 %	38,7 %	52,9 %	0,62 %	7549	8,5
2013 35	432	6,7 %	38,8 %	53,9 %	0,58 %	7619	8,5
Sum / Gjsn	1 674	8,3 %	38,1 %	53,0 %	0,6 %	7504	8,3

Middelgods:

Uke	Tonn	Aske	Flykt	Fix. C	Svovel	NCV	FSI
2013 31	378	20,9 %	33,4 %	45,1 %	0,62 %	6376	8,5
2013 32	405	18,7 %	35,2 %	45,6 %	0,50 %	6580	7,5
2013 33	432	12,8 %	37,3 %	49,1 %	0,83 %	7087	8,5
2013 34	432	9,9 %	38,7 %	50,8 %	0,60 %	7395	7,5
2013 35	486	8,0 %	39,2 %	52,2 %	0,59 %	7509	8,5
Sum / Gjsn	2 133	13,7 %	36,9 %	48,8 %	0,63 %	7023	8,1

Underkorn <5mm: ________

Figure D.1: Analysis of the coal on Svalbard, received from Jørn Myrlund at Longyear Energiverk.

Appendix E – Composite Curves

E.1 Boiler

The hot and cold composite curves of the pulverized coal boiler in the base case are shown in Figure E.1.

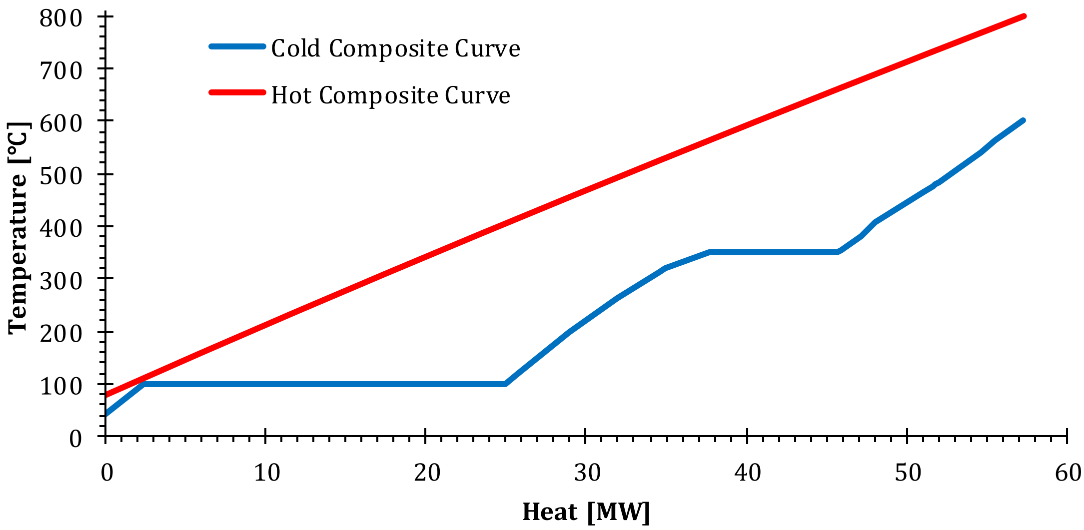

Figure E.1: Plot of hot and cold composite curve for the pulverized coal boiler. The pinch temperature is set to 10°C.

E.2 District Heat Exchanger

The hot and cold composite curve of the district heat exchanger in the base case are shown in Figure E.2.

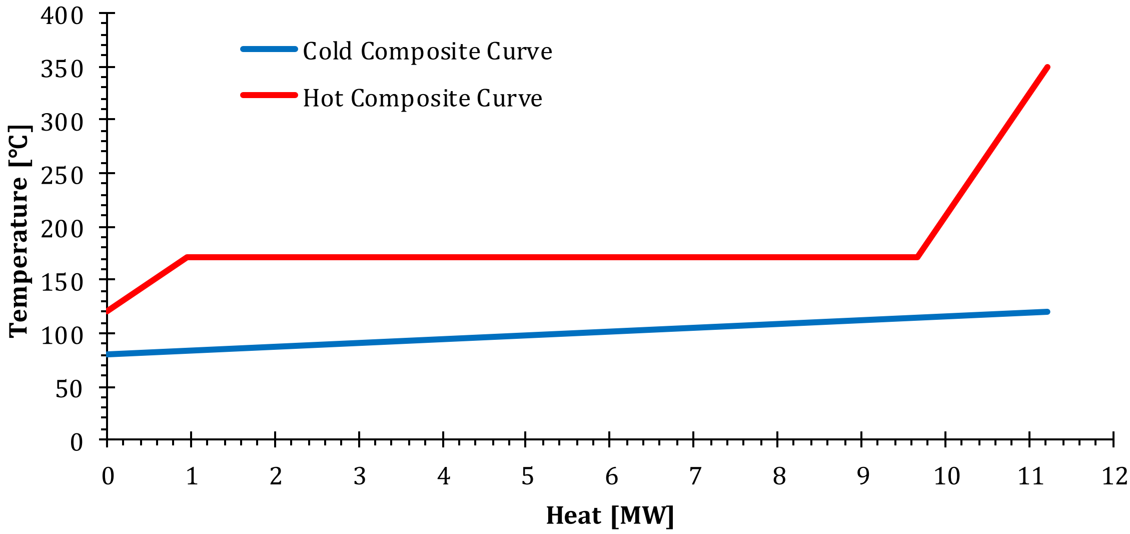

Figure E.2: Plot of hot and cold composite curve for the district heat exchanger.

E.3 Vacuum Condenser

The hot and cold composite curves of the vacuum condenser in the base case are shown in Figure E.3.

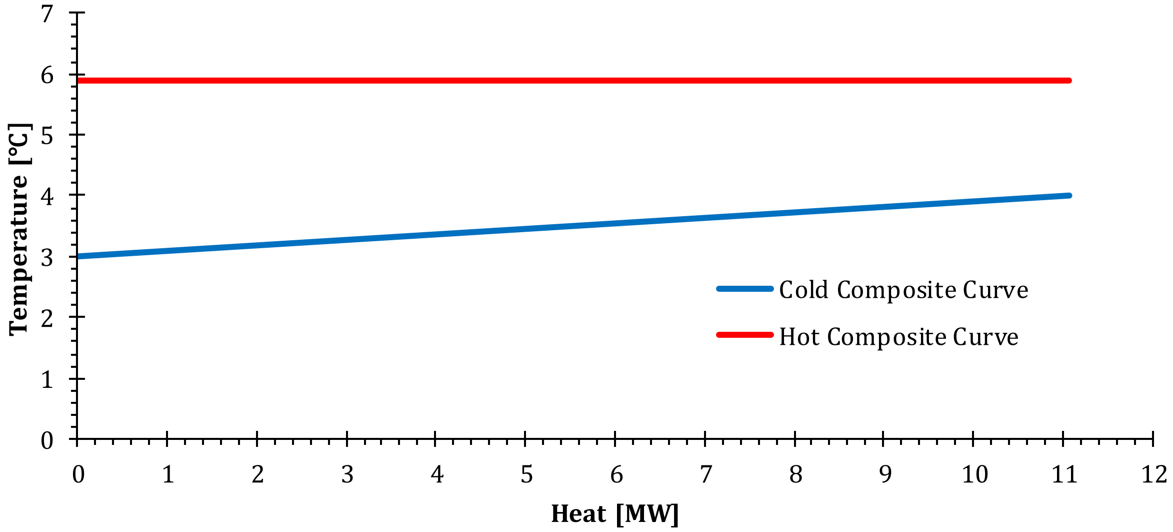

Figure E.3: Plot of hot and cold composite curve for the condenser.

E.4 CO_2 Cooler

The hot and cold composite curves of the CO_2-Cooler in the base case are shown in Figure E.4.

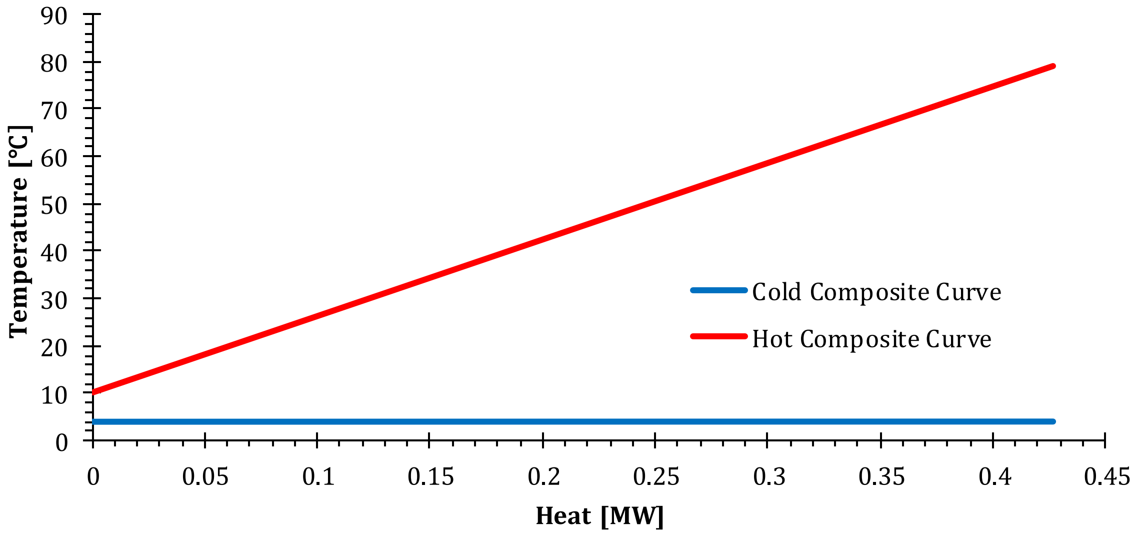

Figure E.4: Plot of hot and cold composite curve for the CO_2-Cooler.

E.5 LP CO_2-Cooler

The hot and cold composite curves of the LP CO_2-Cooler in the base case are shown in Figure E.5.

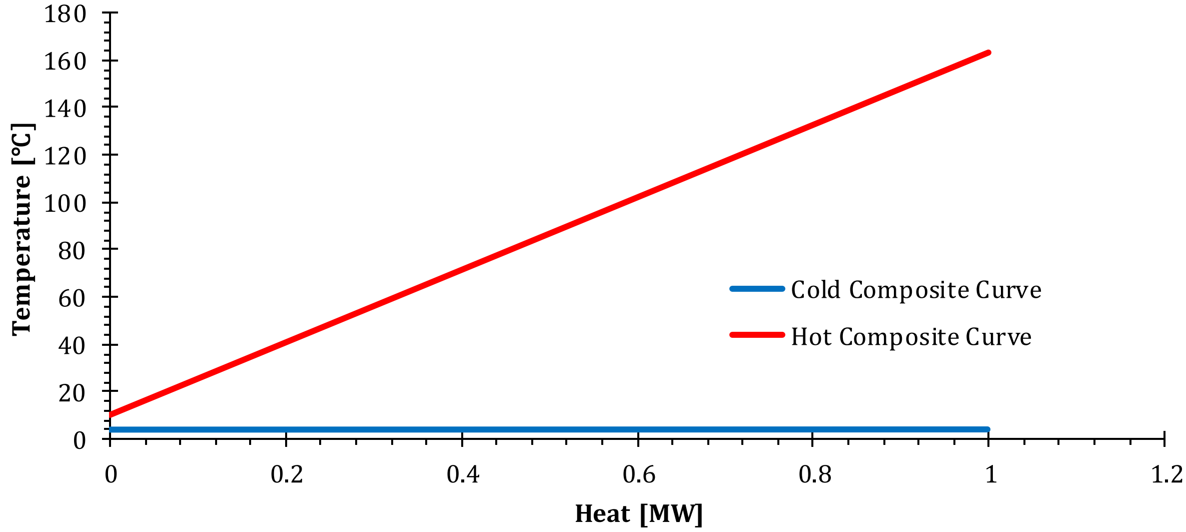

Figure E.5: Plot of hot and cold composite curve for the LP CO_2-Cooler.

E.6 IP CO_2-Cooler

The hot and cold composite curves of the LP CO_2-Cooler in the base case are shown in Figure E.6.

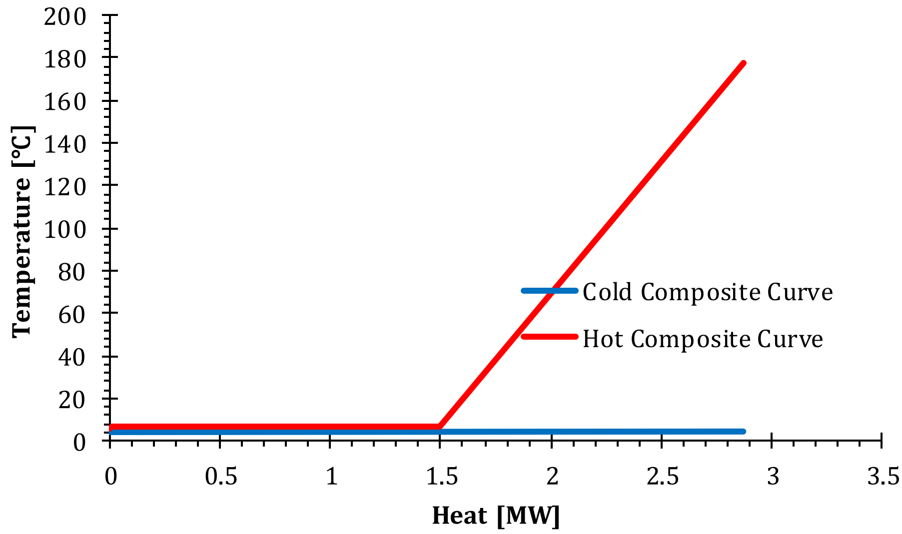

Figure E.6: Plot of hot and cold composite curve for the LP CO_2-Cooler.

Appendix F – Aspen HYSYS Flowsheets

F.1 Base Case

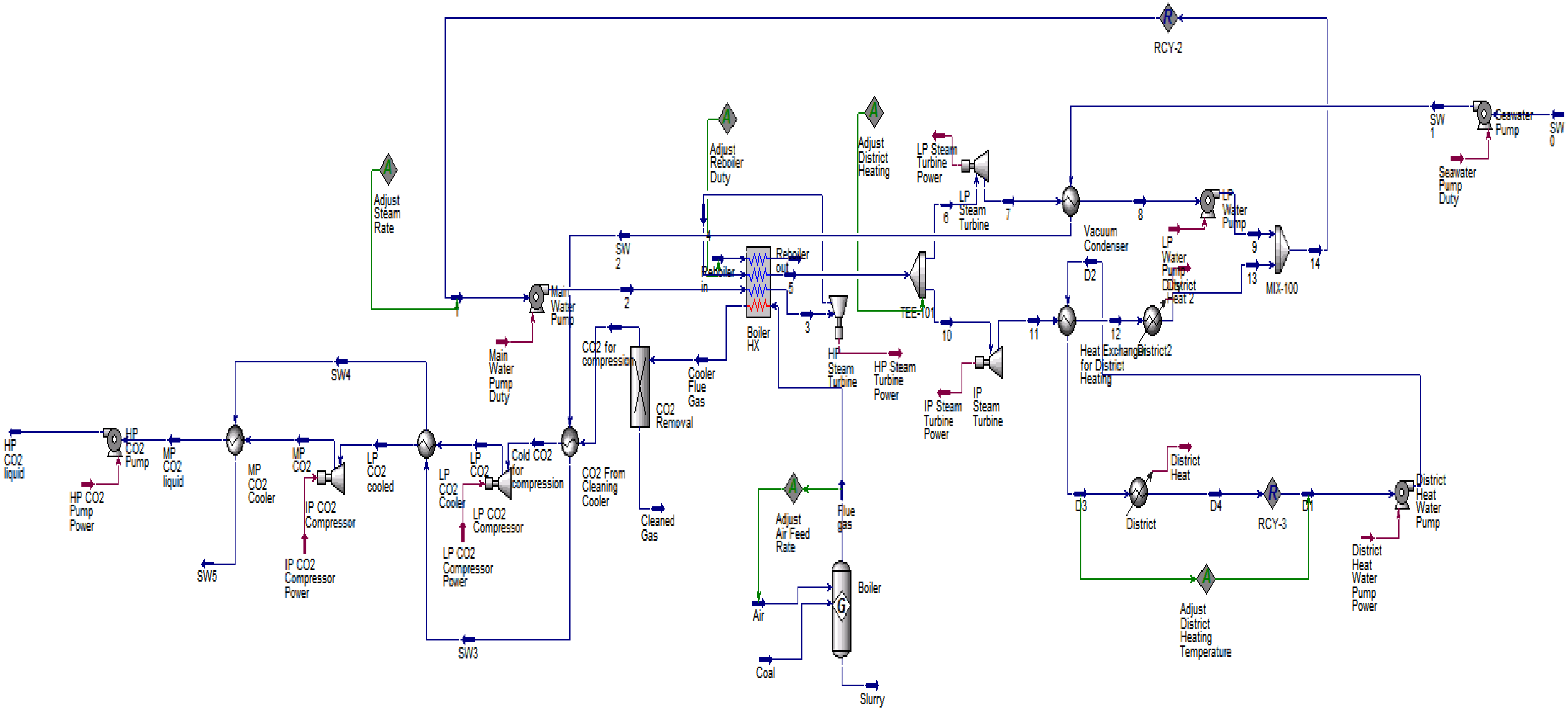

Figure F.1: A larger image of the Aspen HYSYS flowsheet.

F.2 Heat Pump Case

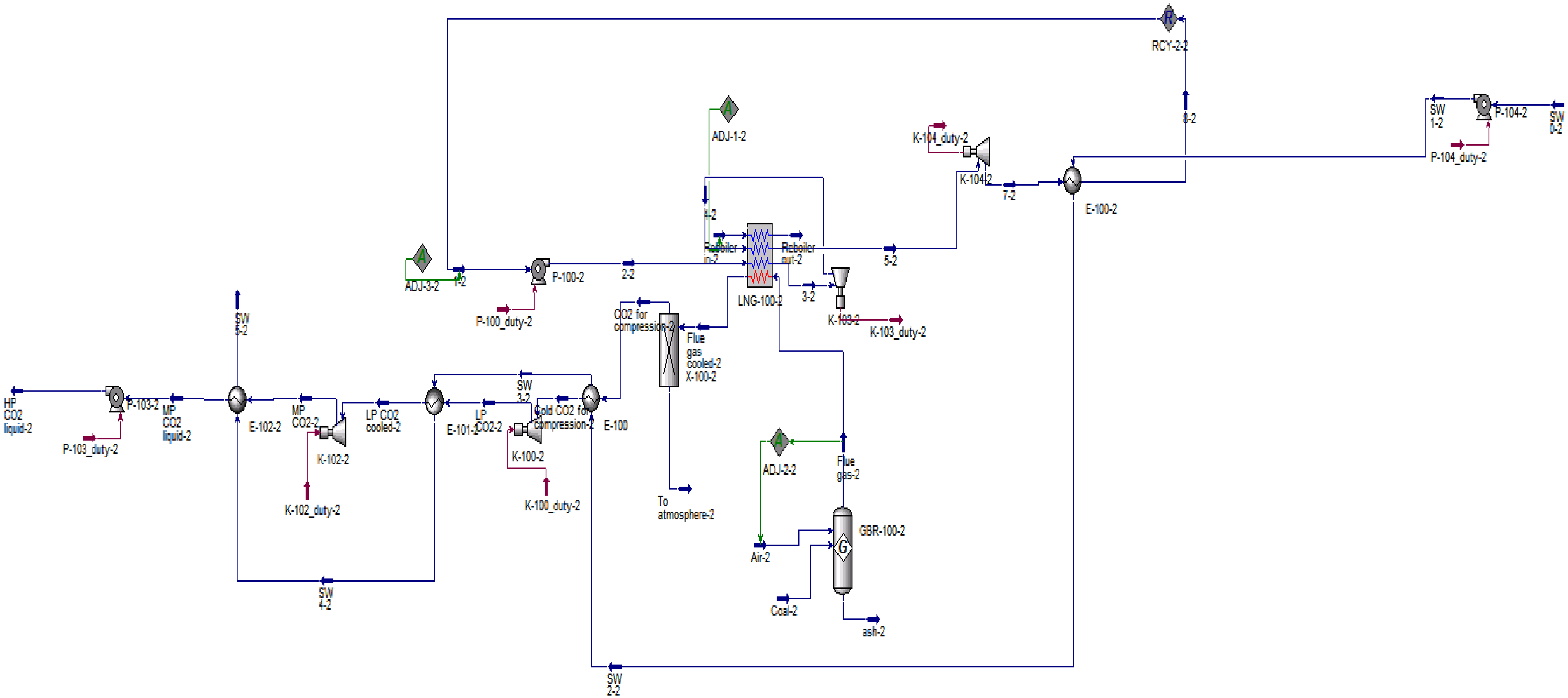

Figure F.2: A larger image of the Aspen HYSYS flowsheet.