Investigation of a CO-production of Substitute Natural Gas and Biomethanol Plant Compared to Stand Alone Biomethanol Plant

Master Thesis
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2019-06-04

Contents

Abstract

Climate change is one of humanities greatest challenges; the greenhouse gases in the atmosphere are still increasing. There is not one solution that will fix this problem several different products, raw material and technologies are required to take the steps towards a renewable society. Biomethanol and substitute natural gas, are two fossil free energy carrier, they can both be produced through gasification.

Biomethanol is produced in an energy consuming process were the limited conversion rate, creates a need to recycle and thereby compress a large recycle feed. This thesis objective is to investigate if a co-production of substitute natural gas and biomethanol can be a feasible process, using mature technology. Simulations of the processes is performed in ASPEN plus. Three different cases were set up.

In case 1 the produced syngas from the gasifier is cleaned using cyclone, OLGA, Rectisol and a guard bed. A shift is used to get the correct ratio between CO and H_2.The gas is then compressed further before entering the methanol synthesis. The flashed off gas is then lead to a methanation step and further down to a gas upgrading step, while the liquid stream is lead to distillation. This way producing both substitute natural gas and biomethanol is produced.

In case 2 the produced syngas from the gasifier is reformed and then cleaned using cyclone, OLGA, Rectisol and a guard bed. A shift is used to get the correct ratio between CO and H_2.The gas is then compressed further before entering the methanol synthesis. The flashed off gas is recycled back to the methanol reactor inlet after being compress up to 75 bar again. The liquid stream is lead to distillation, biomethanol being the only product.

In case 3 the produced syngas from the gasifier is cleaned using cyclone, OLGA, Rectisol and a guard bed. A shift is used to get the correct ratio between CO and H_2.The gas is then compressed further before entering the methanol synthesis. The flashed off gas is recycled back to the methanol reactor inlet after being compress up to 75 bar again. The liquid stream is lead to distillation. The purge stream is lead through a gas turbine, before entering a furnace to create high and medium pressure steam, which is used to produce electricity in turbines.

The simulation results shows that case 2 has the highest thermal efficiency, 0.63, followed by case 1, 0.47 and case 3, 0.21.

The feasibility study shows that case 1 and 2 are feasible with present value method. The discount rate is set to 10% and the plants economical life time, is assumed to be 30 years. Case 2 is showed to be the most feasible investment, during this conditions and assumptions. Case 3 showed a negative present value.

This thesis indicates that the theory of co-production being more feasible then a stand-alone biomethanol plant is not the case during these conditions.

Acknowledgements

First of all, I want to express my gratitude to my supervisor Christian Hulteberg, associate professor, department of Chemical engineering, for him sharing his knowledge and providing me with great guidence.

I also want to express my gratitude to professor Hans Karlsson, department of Chemical engineering, for his time and knowledge.

I want to thank professor Ola Wallberg, department of Chemical engineering for sharing his knowledge.

I want to thank my family and friends for all their support and faith in me. At last, I want to share my graditude and love to Mathias Agmell, for always standing in my corner with love and encouragment, this would not be possible without you.

Symbols and abbreviations

BFB	Bubbling Fluidized Bed
BFW	Boiler Feed Water
CFB	Circulating Fluidized Bed
ECN	Energy Research Centre of the Netherlands
ESP	Electrostatic Precipitator
GC	Gas Chromatography
GROT	Forest Residues - Branches and Tree-tops
HT	High Temperature
ICI	Imperial Chemical Industries
LP	Low Pressure
LT	Low Temperature
MFV	Minimum Fluidizing Velocity
MT	Medium Temperature
OLGA	Oil Based Gas Scrubber
PAH	Polyaromatic Hydrocarbon
SNG	Subsitute Natural Gas
SSD	Superheated Steam Dryer
WGS	Water Gas Shift

Chapter 1

Introduction

In order to reduce the emissions of greenhouse gases different technologies must be used. In Sweden, electricity and heat production have relatively low emissions of greenhouse gases. The greatest bottleneck to achieve the national climate commitments lies in the transportation sector. Most of the fuel used today are fossil fuels, in 2015 14.7 %, 13.1TWh of the 89.1 TWh energy used in the transport sector were from a renewable energy source. The total energy consumption in Sweden 2015 were 375TWh, this means that transport sector make up about 25% [1].Therefore, even if the most energy efficient way to use biomass is as a fuel in a combined heat and power plant, it might be environmentally smarter to use it to create transportation fuel or other products. In that case, Biomethanol and substitute natural gas (SNG) are two possible products. But in the production of methanol, the high pressure and the limited conversion per pass through the reactor, makes it a very energy consuming process.

1.1 Purpose and aim

The objectives of this thesis are to design a plant that produces SNG and bio-methanol by gasification of biomass, to find available and proven technology to use and to make a feasibility study to find out if co-production can be economically more profitable than a pure methanol plant. Due to avoiding recycle the process gas back to the methanol synthesis after the flash step, and instead use this gas to produce a methane rich gas in a methanation step and thereby save the compression cost for the recycled stream. This study will be achieved through;

- A literature study over available technologies.
- Technology for the different process steps will be chosen out of the literature study.
- The chosen process will then be simulated.
- Calculations of the feasibility for the two cases will thereafter be evaluated.

The biomass input will approximately be 520 MW. The biomass input will consist of branches and tops (GROT) from forest residues. Three cases, with one co –production plant with methanol and SNG, one pure methanol production plant and one methanol plant that combust the purges gas will be performed and compared. Figure 1.1 below shows a schematic picture of the process from biomass to product for the co-production plant.

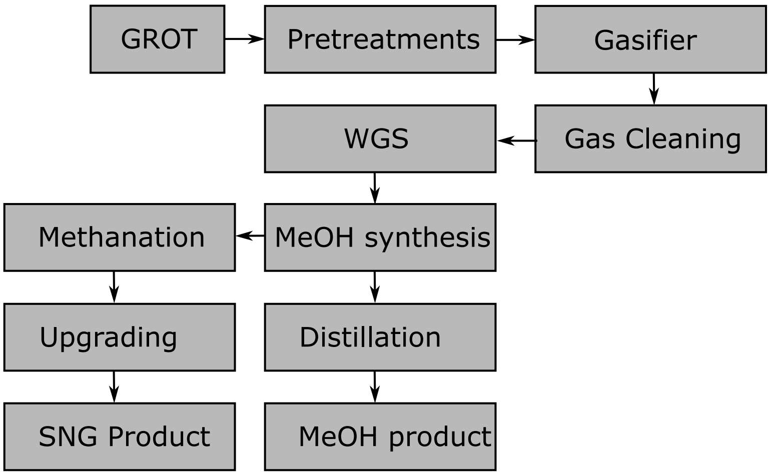


Figure 1.1: Schematic illustration of the process

Chapter 2

Background

In the study the raw material in the process will be biomass and the main products will be methanol (MeOH) and SNG. The selected biomass will be GROT from logging.

2.1 Methanol

Methanol is the simplest alcohol with the formula CH_3OH. It is a colorless and clear liquid at room temperature and has a mild odor. Methanol can be used as fuel, it has advantages as low flame temperature that gives low energy loss, and it also has low CO, NOX and hydrocarbon emissions compared to gasoline fuel. When being produced from renewable raw material it also has a low climate impact. [2]

2.1.1 Purity Standards and Procuct Price

When sold as fuel or a chemical to the industry the purity of the methanol needs to be at least 99.85%. At this purity the methanol sales price are 360 EUR/Tonne for European Posted Contract price according to Methanex [3].

2.1.2 Production of Methanol

Methanol is produced by reacting natural gas or SNG over a Cu/ZnO catalyst. The reaction pressure should be between 50-100 bars, and the temperature should be held between 220-275°C. The methanol conversion is limited to 25% per pass through the reactor by thermodynamics [4]. To achieve a higher yield produced methanol must be separated from the product stream before the stream is recycled back to the inlet of the reactor. The reactions for the methanol synthesis are given in Eq. 2.1.

$$\begin{aligned} CO + 2H_2 \leftrightarrow CH_3OH \qquad & \Delta H_R^{\circ} = -91\frac{kJ}{mol} \\ CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O \qquad & \Delta H_R^{\circ} = -53\frac{kJ}{mol} \\ CO_2 + H_2 \leftrightarrow CO + H_2O \qquad & \Delta H_R^{\circ} = -41\frac{kJ}{mol} \end{aligned} \tag{2.1}$$

2.2 Substitute Natural Gas

Biogas or SNG is a good replacement to natural-gas, instead of being a fossil energy source it is made from biomass and has therefore almost none net impact on the greenhouse effect. The main component in SNG is methane, CH_4.

Organic matter can be transformed into methane by microorganisms during the right circumstances. This can be used in order to produce biogas, this technique is called anaerobic digestion, but it is best suited for organic matter with high water content that is easy to break down. For organic matter that is drier and harder to break down, a process that is built on gasification can be used instead; the produced gas is called SNG [5, 6].

The syngas from the gasifier are reacted to form methane. The main reactions are;

$$\begin{aligned} CO + 3H_2 \leftrightarrow CH_4 + H_2O \qquad \Delta H^0_{298} = -206\frac{kJ}{mol} \\ CO_2 + 4H_2 \leftrightarrow CH_4 + 2H_2O \qquad \Delta H^0_{298} = -165\frac{kJ}{mol} \end{aligned} \tag{2.2}$$

The reactions are exothermic and therefore heat must be removed to control the temperature.[7]
The sales price for SNG on the European market is 0.535 SEK/kWh. [8]

2.3 Biomass

The biomass that is being used in this plant will mainly be GROT, therefore there will not be a more thorough presentation of other kind of fuel. The moisture content in GROT varies widely between 35-55%, according to [9] the most common moisture content is between 40-50%, in this report a moisture content in GROT of 45% will be assumed.

Before the biomass can be gasified it has to be dismembered into finer particles and dried, the minimum particle size is determined by the gasifier. The biomass should also be dried, before entering the gasifier to a water content somewhere between 10-20% [10].

Chapter 3

Available Production Technology

There are different equipment and methods that can be used to convert biomass to methanol and SNG. Some are old and proven technology, some are quite new and some are still in development progress. The most common and promising new technologies will be reviewed and compiled.

3.1 Pretreatment Technology

Before the biomass can be gasified it has to be dismembered into finer particles and dried, because the gasifier can not handle large pieces of biomass and if the biomass is wet a big part of the energy put into the gasifier will be used to dry the material, this will make the overall efficiency lower. The biomass should have a moisture content between 10-15% before it enters the gasifier. The raw material can be delivered in different ways, ready to use, grinded or without any preparation [4, 11].

If the selected process has a pressurized gasifier the biomass must be pressurized before it is feed into the gasifier. There are also different feeding systems to choose from [12].

3.1.1 Shreding

The biomass must be shredded before entering the dryer and the gasifier, the gasifier is the one equipment that determines how small the biomass has to be. If the biomass does not have the required size it can lead to problems with feeding and inhibit bed fluidization. The two most used technologies to achieve the size reduction are knife chippers and hammer mills [13].

3.1.2 Drying

Biomass has a low dry content when it arrives to the production site, drying biomass takes a lot of energy. To get a better efficiency in the gasifier the biomass can be dried before it enters the gasifier using excess heat from downstream process. The biomass has a moist content of about 45% before it is dried. Drying the biomass below 12% is inefficient because of inner transportation problems as a result of the surface no longer being moist, drying to much can also lead to the biomass self-igniting. Other reasons for drying the biomass before entering the gasifier are to make sure the gasifier gets a feed of consistent moisture content, which results in an even process that is easier to control and to avoid the biomass clogging the feeding system [13].

There are two main ways of providing the heat to dry the biomass, these are by direct drying and by indirect drying. Direct drying is when the drying medium comes in contact with the biomass and indirect when the biomass is not in contact with the heating source. The direct drying can use hot air or superheated steam as a drying medium [14].

To avoid the release of volatile compounds in the dryer, biomass should not exceed a temperature greater than 100 °C [13].

3.1.2.1 Rotery Dryers

Rotary driers are the most common dryer type used in big industrial scale. There are many variations of rotary driers but a general description of the dryer is that it consists of a circulating drum where the biomass passes through and gets in contact with the drying medium. Inside the drum, flights are mounted with the purpose to increase the contact between biomass and the drying medium. The rotary drum dryer can handle a wide range size distribution of the biomass going into the dryer. Larger particles will have a longer residence time than the smaller ones. The advantages with rotary dryers are that they are not sensitive for the particle size of the material or the size diversity. They can operate with high temperature on the drying medium, they have the largest capacity of dryers and they have a robust technology giving them a low maintenance cost. The drawbacks are that they take a lot of space and have the highest fire hazard. This because it is hard to control the moisture content of the material [14].

3.1.2.2 Flash Dryers

Flash dryers have a short resident time, about 30 seconds, were the biomass are mixed with a high-velocity hot air stream before they are separated in a cyclone. Before entering the flash dryer the biomass has to be shredded or grinded to reduce its size, so it is small enough to be suspended in the drying medium. The advantages of flash dryers are that they have lower fire risk and they are small in comparison to rotary dryers. The drawback of the flash dryer is that it needs very small particles [14].

3.1.2.3 Superheated Steam Dryers

Superheated Steam Dryers (SSDs) operates similar to flash dryers, with the difference that the drying medium in this case is steam instead of air. The steam used to dry the material is recycled into the dryer with about 90%, the 10% remaining percent is representing the removed water from the material. The advantages of SSDs are that it is easy to recover heat and a low fire hazard. Some of the drawbacks of SSDs are that the biomass needs to have a small particle size and great capital costs [14].

3.1.3 Pressurising

Pressurizing the process makes the plant smaller in size but also increases the requirements on the plant to handle the pressure. For some processes increased pressure can favor the sought chemical reaction, the methanol synthesis is favored by high pressure.

3.1.3.1 Lock Hoppers

Lock hoppers is a batch-wise system where inert gas pressurize a closed vessel, filled with biomass, until the desired pressure has been reached. The pressurized biomass is then emptied in a feed bin to allow a continuous flow of biomass into the gasifier. To transfer the biomass from the feed bin to the gasifier a screw feeder could be used. The inert gas can be N_2 or CO_2 [12].

3.2 Gasification

In gasification the biomass is transformed into syngas. The gasification is done in three steps pyrolysis, combustion and reduction and is carried out in an environment with a limited O_2 access. In the pyrolysis step the biomass will decompose and become pyrolysis gas and char. It is the volatile compounds in the biomass that creates the gas and the rest will take the form of char. The pyrolysis starts when the biomass reaches a temperature of about 400° C. This step does not need any oxygen, it only needs heat. [15]

Gasification can be carried out at with or without pressure. Increased pressure increases the partial pressures of reactive components such as H_2O and CO_2 this leads to a higher reaction rates, the higher partial pressure of O_2 also leads to a higher oxidation rates and thereby higher temperatures in the gasifier which gives a higher conversion rate. The increased pressure will also have reducing effect on the conversion rate because of the thermodynamic resistance to increase the amount of molecules [12]. Pressurized gasification makes the gasifier smaller which makes investment cost lower but if the pressure is to high this decrease in investment will change into an increase due to the need of greater dimensions and materials. High pressure increases the methane and CO_2 content of the produced gas and reduces the H_2 and CO content [4, 16].

3.2.1 Fixed Bed Gasifiers

Fixed bed gasifiers also called moving bed gasifiers, are an old type of gasifier and were used during World War 2 when gasoline was a scarce resource.

There are two main types of this kind of gasifier updraft gasifier and downdraft gasifier, an illustration can be seen in Figure 3.1. In a fixed bed gasifier the fuel will be transported slowly through the gasifier on a fixed bed. The fuel is introduced at the top of the gasifier and gravity is pulling it down towards the bottom. The difference between the two in design is that the produced gas is removed in the bottom of the downdraft gasifier and in the top of the updraft. The second difference is that the oxidizing medium is feed in the middle of the downdraft gasifier and in the bottom of updraft gasifier. Fixed bed gasifiers that gasify biomass produce a product gas that contains a lot of tars. The operation temperature varies inside of the reactor from 500-1,000 °C. A great disadvantage of the fixed bed gasifiers are besides the high tar content in the product gas the fact that they have a low capacity, only 10-15 ton of dry biomass per hour [4].

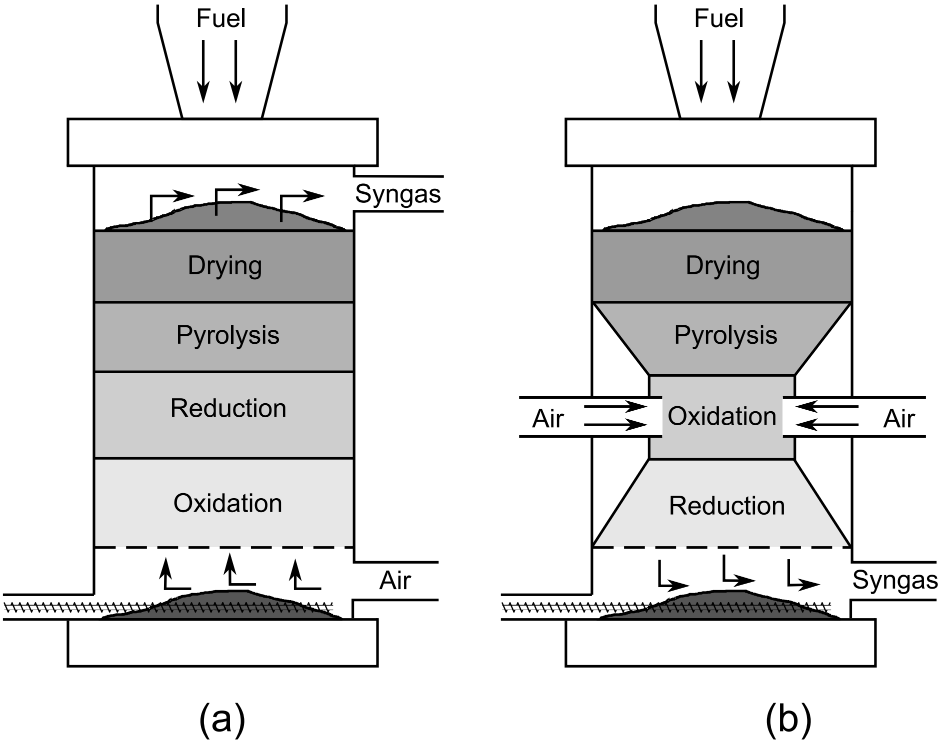


Figure 3.1: Illustraion of an updraft gasifier (a) and a downdraft gasifier (b).

3.2.2 Fluidized Bed Gasifiers

The fluidized bed gasifiers have a floating bed that is made up of particles often sand but can also be made of a material that have some catalytic characteristics. The fluidization medium is injected at the bottom of the bed and must hold the Minimum Fluidizing Velocity (MFV), the velocity that make the bed fluidize. When MFV is upheld the gasifier will operate as a Bubbling Fluidized Bed (BFB) and at higher velocity it will operate as a Circulating Fluidized Bed (CFB), were the bed material will get so high velocity so it travels out of the gasifier together with the produced gas. Fluidized bed gasifiers have a good heat transport and the temperature is even throughout the reactor [17].

The Fluidized Bed gasifiers can operate with both air, O_2 or steam as oxidizing medium and with or without pressure. A problem with fluidized bed gasifiers are that ash related problem can occur at low temperatures. The problem is that the sand agglomerates because of sticky or melted ash, this leads to defluidized bed. It is hard to get rid of the agglomerated bed and replacing it is expensive. The agglomeration temperature varies depending on the fuel and the bed material, a cleaner fuel with a low alkali metal content will have an increased agglomeration temperature. Other ways of avoiding this problem is too often exchange the bed material, adding mineral binding products and holding down the temperature in the gasifier [4].

3.2.2.1 Bubbling Fluidized Bed Gasifiers

In the BFB gasfier the gasifying medium inlet is in the bottom of the gasifier and the velocity is balanced so the bed is fluidized, at about 2-3 m/s. With the velocity the size and speed of the bubbles can be tuned. This affects the mixing and heat transfer between the particles. The biomass fuel is fed into the gasifier above or into the bed.

BFBs can handle a high throughput compared to Fixed Bed gasifiers. They have a high carbon conversion due to good mixing, great heat transfer, good particle/gas contact and the long residence time. Another good character of the BFB gasifer is that it produces a product gas with low tar content. When it comes to specifications of the fuel the BFB gasifier is forgiving in both particle size and moisture content, as well as fluctuations in them both. The BFB gasifer can operate at pressures varying from 1-35 bar and temperatures between 650-950 °C.

The disadvantages with this kind of gasifier is that it produces a syngas rich in particulates and if large bubble size occur it can result in gas go through the bed unreacted. As with all fluidized bed gasifiers there are also a risk off bed agglomeration. Another disadvantages is that the technology can be considered advanced and more so if it is operated with elevated pressure [4]. An illustration of the design of a BFB gasifier can be seen in Figure 3.2.

3.2.2.2 Circulated Fluidized Bed Gasifiers

CFB gasifiers has a higher gas velocity then the BFB, this results in some of the bed material following the product gas out of the gasifier. To separate and to recycle the bed material a cyclone is used. Typically velocity used is between 5-10 m/s.

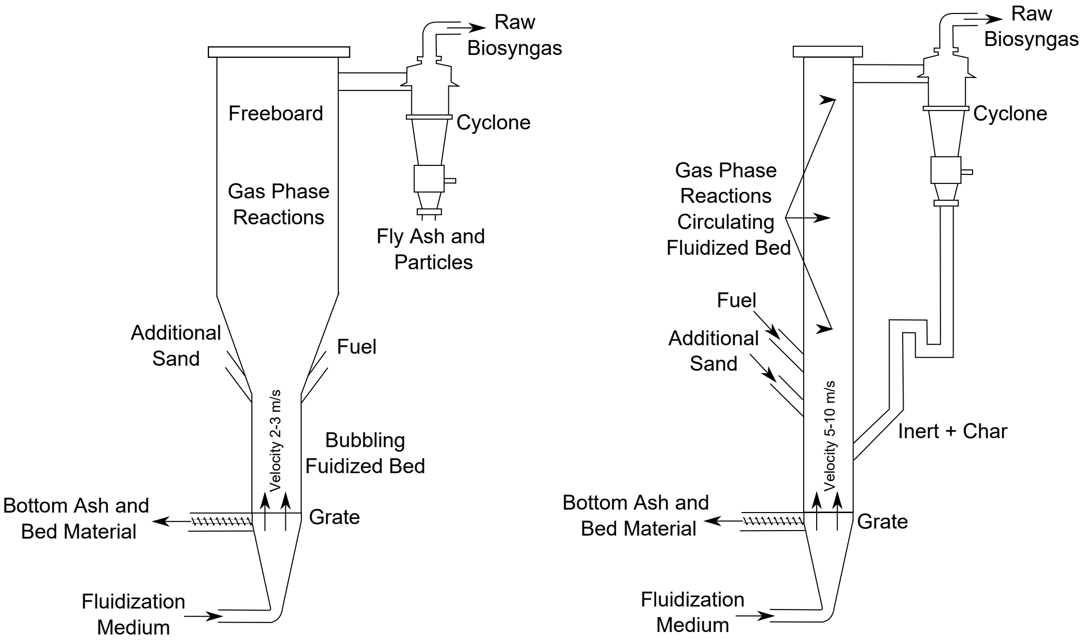
CFB gasifiers can handle a high throughput compared to Fixed Bed gasifiers and BFB gasifiers. They have a high carbon conversion due to good mixing, great heat transfer, good particle /gas contact and the long residence time. Another good character of the CFB gasifer is that it produces a product gas with low tar content. When it comes to specifications of the fuel the CFB gasifier is forgiving in both particle size and moisture content, as well as fluctuations in them both. It can also handle a fluctuation in feed quantity. The CFB gasifier can operate at pressures varying from 1-19 bar and temperatures between 800-1,000 °C.

The disadvantages with this kind of gasifier is that it produces a syngas rich in particulates and if large bubble size occur it can result in gas go through the bed unreacted. As with all fluidized bed gasifiers there are also a risk of bed agglomeration. Another disadvantages is that the technology can be considered advanced and more so if it is operated with elevated. It should be considered that a high particle size on the fuel leads to a high velocity that may result in erosion of the equipment. Compared to the BFB gasifier the CFB gasifier have a lower heat exchange efficiency, in the direction of the solid flow temperature gradients can arise [4]. An illustration of the design of a CFB gasifier can be seen in Figure 3.2.

3.2.3 Entrained Flow Gasifier

Entrained flow gasifiers are operated at elevated pressure and high temperature. The biomass must be grinded into very small particles in order to achieve a good carbon conversion. The biomass particles should have a size below 0.1 mm. Figure 3.3 show an illustration of an entrained flow gasifier.

When the biomass comes into the gasifier it reacts with the oxygen and/or steam almost instantaneously. The gasifier operates at temperatures around 1,200 °C and pressures between 20- 50 bar. The high temperature leads to an almost tar free product gas. The produced gas will have a low content of carbon dioxide and on methane. The main disadvantages is that a large part of the injected energy will be transformed into sensible heat, that needs to be taking care of in a good way to make the overall process efficient. The pretreatment of the biomass is more expensive due to the high demands on the size [4].



(a) (b)

Figure 3.2: Bubbling Fluidized Bed Gasifier (a) and Circulating Fluidized Bed Gasifier (b).

3.2.4 Indirect Gasifiers

Indirect gasification differs from the other gasification methods because no combustion takes place in the part of the gasifier were the gas is produced, instead an external heat source is used to provide the required heat. This can for example be done by using steam but it can also be combustion of the produced gas, char or any other fuel. This results in an almost nitrogen free gas with relatively high methane content. A nitrogen free product gas reduces the size of the rest of the plant, this is economically favorable. [4]. Figure 3.3 shows a schematic view of the design of an indirect gasifier.

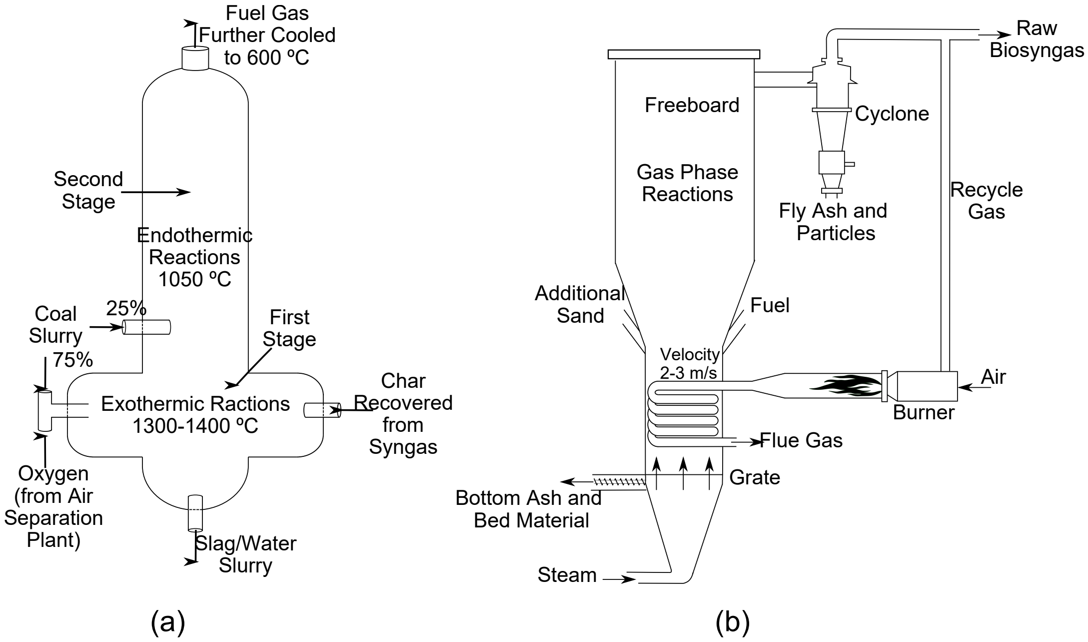


Figure 3.3: Entrained Flow Gasifier (a) and Indirect Gasifier (b).

3.3 Gas Cleaning

In the process gas leaving the gasifier there are a lot of unwanted impurities such as particles, tars, carbon dioxide, water and sulfur compounds. This needs to be reduced before the gas can enter the methanol synthesis. The produced SNG must also be cleaned in order to achieve the requirements of the gas grid.

3.3.1 Particle Removal

Particles that leaves the gasifier together with the gas is mostly ash, bed material and coke. They are the first impurities that need to be separated out of the gas. There are many ways to remove the particles, through a cyclone, filter, scrubber or an electrostatic precipitator.

3.3.1.1 Cyclone

In the cyclone the particles are separated from the gas by using the centrifugal force. This method has its advantages of being reliable and cheap. The parameter that decides the working conditions in form of temperature and pressures is the choice of material. The disadvantage of a cyclone is that it isn't efficient enough to meet the requirements for downstream processes [18]. Figure 3.4 shows a schematic view of a cyclone.

3.3.1.2 Barrier Filter

A barrier filter separates by size with porous materials made out of metal or ceramic [18]. The particles are caught in the filter until cleaning, this is preformed by of blowing pressurized gas in the opposite direction making the particles fall down in a collector [19]. The minimum size of the particles that get separated depends on the mesh size of the filter. When using barrier filters to separate out very small particles the pressure drop must be taken in to account [18]. Figure 3.4 show an illustration of a barrier filter.

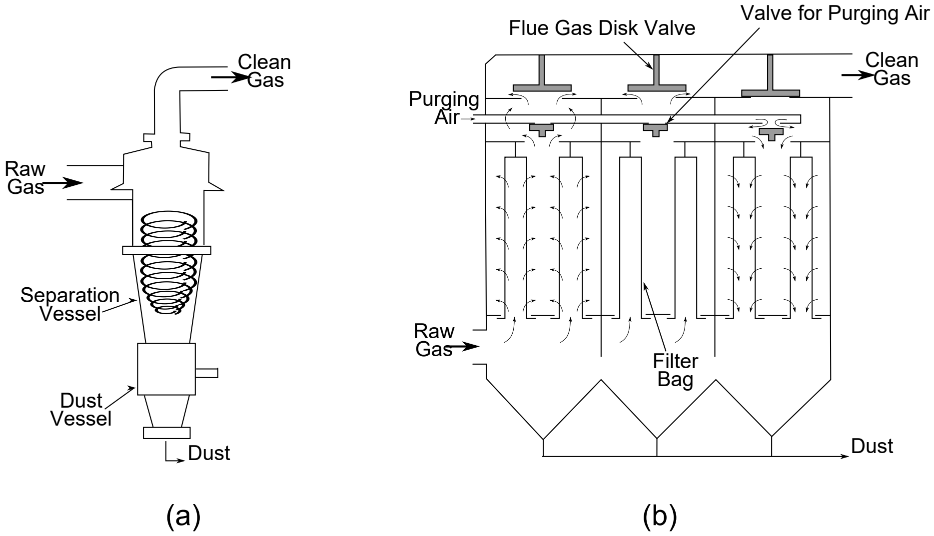


Figure 3.4: Cyclone (a) and Barrier Filter (b).

3.3.1.3 Electrostatic Precipitator

Electrostatic Precipitator (ESP) uses electrical forces to separate the gas from the particles. It does it by charging the particles with high voltage, the charged particles will then move towards the precipitator electrode. The particles will stay there until the regeneration when the power turns off and a hammer hits the precipitator, this makes the particles fall down into a collector [20]. Figure 3.5 shows a schematic view of an ESP.

3.3.1.4 Scrubber

Both water based and oil based scrubbers can be used to remove particles. Due to higher affinity the oil based scrubbers can remove smaller particles then the water scrubbers. The oil scrubber medium is more expensive then the use of water based medium. The big disadvantage of using water is that a water cleaning plant has to be installed because of the dirty water needs taking care of. The dirty oil medium on the other hand can just be send back to the gasifier were it will work as fuel [18]. Figure 3.5 shows an illustration of a scrubber.

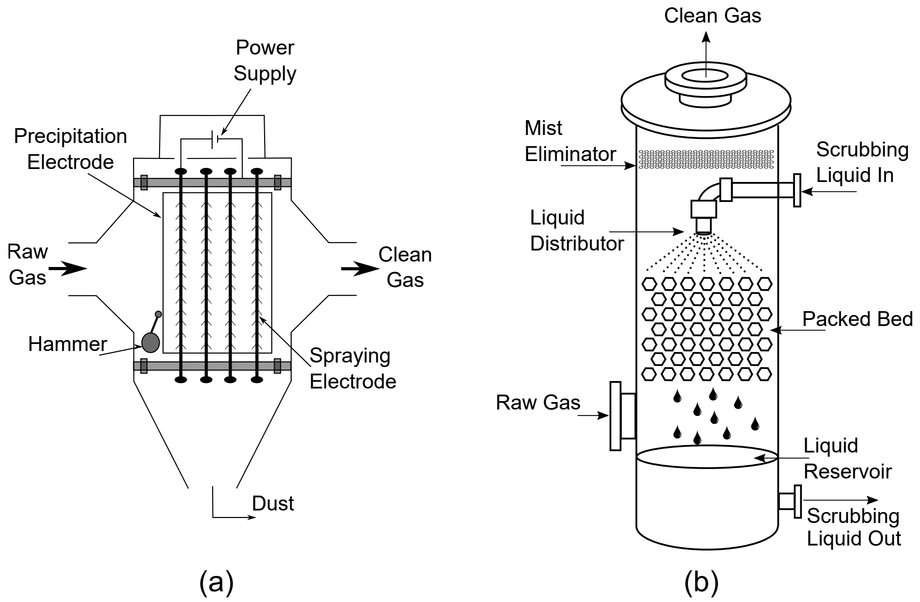


Figure 3.5: Electrostatic Precipitator (a) and Scrubber (b).

3.3.2 Tar Removal

In the gasifying reaction tars are formed, these are made up by hydrocarbons with a higher molecular weight then benzene [21]. The removal of these tars is one of the hardest problems when it comes to gas cleaning. The kind of tar produced in the gasification depends on the reaction temperature. A lower temperature then 750 °C gives lower molecular weight tars and a temperature over 750 °C gives higher molecular weight tars. Tars are a problem because they create fouling in the equipment when the temperature reaches the tar dew point. This leads too high operating cost because of the increased maintenance and decreasing operational time [22]. Parameters that affect the dew point most are concentration and composition. Tars can be divided into six different classes, Energy Centre of Netherlands (ECN) uses the classification system shown in Table 3.1.

3.3.2.1 Physical Tar Removal

Physical removal of tars is most commonly done by using a wet scrubber.

3.3.2.1.1 Oil Based Scrubbers

A relatively new technology developed by ECN, Energy Centre of Netherlands, is an oil based scrubber system called (OLGA). Instead of focusing on reducing the concentration of tar ECN set out on reducing the tar dew point, and thereby getting rid of the tar related problems. The tar loaded syngas is first cooled by the oil, this step make the heavy tars condensate and dissolved in the oil. The removed tars are then separated from the scrubbing oil and returned to the gasifier together with unrecovered scrubbing oil. The syngas goes into the absorber were the light tars are absorbed by the scrubbing liquid. The scrubbing liquid is then sent to the stripper to regeneration then the light tars and some scrubbing oil is returned to the gasifier. By returning the tars to the gasifier the total process gets more efficient. Before the syngas reaches the Olga process the syngas has to be cooled down and particles removed, fine particles are removed in the OLGA. The level of how much the gas needs to be cooled is based on the medium used in the process; the temperature interval reaches between ca 320-500°C. Using a medium allowing a higher

Table 3.1: ECN classification system for tars, adpoted from [23].

Class	Type	Examples
1	GC undetectable tars.	biomass fragments, heaviest tars (pitch)
2	Heterocyclic compunds. These are components that generally exhibit high water solubility.	phenol, cresol, quinoline, pyridine
3	Aromatic components. Light hydrocarbons, which are important from the point view of tar reaction pathways, but not in particular towards condensation and solubility.	toluene, xylenes, ethylbenzene (excluding benzene)
4	Light polyaromatic hydrocarbons (2-3 rings PAHs). These components condense at relatively high concentrations and intermidiate temperatures.	naphthalene, indene, biphenyl, antracene
5	Heavy polyaromatic hydrocarbons ($\geq$ 4-rings PAHs). These components condense at reatively high temperature at low concentrations.	fluoranthene, pyrene, crysene
6	GC detectable, not identified compounds.	unknowns

temperature has the advantage of having less fouling between the cyclone and the OLGA process. The OLGA system does not affect the methane content in the gas. To achieve the required level of particles a hot gas filter made of ceramic can be used. Ceramic hot filters are expensive and should in first hand be used for small production scale. Another disadvantage is that they are fragile and easily breaks. Instead of having a hot gas filter the OLGA system can be equipped with an ESP. A cyclone still has to be used to separate the gas from the bigger particles but with an ESP the OLGA can manage the small particle and separate them from the gas. Using an ESP also gives ca 1% higher chemical efficiency [24].

3.3.2.2 Chemical Tar Removing Processes

In chemical tar removing systems, heat and catalyst is used to crack the tars. There are two main approaches thermal cracking and catalytical cracking.

3.3.2.2.1 Thermal Cracking

The tar is cracked at high temperatures, to reach this high temperature some of the produced gas is combusted. In order to achieve this some oxygen must also be present. The reactions taking place are given in Eq. 3.1. After the cracker a particle removal step must be installed to remove the dust and soot that have been formed [25].

$$\begin{gathered} CH_4 + 2O_2 \rightarrow CO_2 + 2H_2O \\ C(tar) + O_2 \rightarrow CO_2 \end{gathered} \tag{3.1}$$

3.3.2.2.2 Catalytical Cracking

By using a catalyst in the cracking process a lower operational temperature can be used. The catalyst can be used in the gasifier as well and there by a separate cracking reactor is unnecessary.

The material of the catalyst is a metal or a metal-oxide. Nickel, alumina, dolomite and limestone is material that are commonly used. By using catalytical cracking in the gasifier a lower investment cost is achieved, if a separate cracking reactor is used the investment cost is lower for that unit, because the catalyst enables a lower temperature and that results in lower requirements on the material. Methane lowers temperature of the syngas this results in less combustion of the biomass/produced gas. But on the downside, is that it is not a fully tested technic [15].

3.4 Removal of Inorganic Impurities and CO_2

The inorganic components in the gas need to be removed because they cause catalyst poisoning in downstream equipment. CO_2 is produced in both the gasifier and the water shift gas step, the CO_2 also needs to be removed in order to not inhibit the MeOH synthesis and to keep the scales on the equipment down. Upgrading of the SNG also implicate removal of CO_2, N_2 and O_2. For the MeOH synthesis the syngas should have a CO_2 content of 4-8% to make the reaction about 100 times faster [4]. There are several ways to remove these impurities; the most common and tested will be reviewed.

3.4.1 Physical absorption

In physical absorption the solubility of the unwanted components in a special medium is used. In the absorber the gas and the solvent are brought into contact counter currently. There are three different types of absorption that can be used to remove CO_2, chemical, physical and hybrid absorption [4]. Physical absorption capacity is better than chemical absorption capacity at high partial pressures, at low partial pressure it is the result is the opposite [7].

3.4.1.1 Selexol

Selexol is a physical absorption medium that is made of dimethyl ether of polyethylene glycol. This solvent is stable and is neither reactive nor degrading. Selexol has a disadvantage of its lighter components can be distilled off during regeneration; therefore attention must be kept on the solvent properties so that they are sufficient. The Selexol process removes H_2S, CO_2, COS, NH_3, HCN, metal carbonyls and mercaptans and other sulfur compounds. [7]

There will be some loss of methane when using Selexol due to the solubility of methane in Selexol. This loss is reduced by the use of a flash step in the process were the flashed gas that is rich of methane is recycled to the inlet. [26]

3.4.1.2 Rectisol

The Rectisol process was developed by Lurgi together with Linde in the mid-1950s. The process is a physical washing system that uses methanol as the absorbing medium. The operating conditions are favored by low temperatures, and high pressures. [27] The need of refrigeration is a disadvantage of this process because of the cost aspect. Most of the cooling can be accomplished with heat exchanging within the process. An advantages is that methane are not absorbed. Components that are absorbed by the methanol are CO_2, H_2S, COS, NH_3, nickel and iron carbonyls and other sulfur compounds. [7]

3.4.2 Pressure Swing Adsorption

Pressure Swing Adsorption (PSA) can be used to remove CO_2, O_2, and N_2 from the gas. The separation is determined by the molecular size. The adsorption is performed at elevated pressure and the regeneration of the adsorption columns is performed at pressure reduction. The adsorption is usually carried out on zeolites or active carbon. Hydrogen sulphide is a poison to the adsorption material and should be removed from the syngas before it enters the PSA. The PSA plant has working cycle that contains different phases, to get a continuous gas flow the PSA plant is made up by 4 different columns that is working in different phases of the cycle. First is the adsorption stage were the compressed gas passes through the column and the CO_2, O_2, and N_2 is adsorbed on to the adsorption material. The upgraded gas, now rich on methane, is then removed from the column. The column is then starting the regeneration phase were the pressure is lowered in several stages, before the last stage the remaining gas in the column which still contains high levels of methane is recycled to the PSA inlet. The pressure is then lowered further to finish the regeneration and the gas that is desorbed is mainly the CO_2, O_2, and N_2. The gas will also contain some amount of methane that will be lost. The last phase of the cycle is the pressurizing of the column which is done by balancing it to a column that has finished the adsorption phase before the raw SNG completes the pressurizing. [26]

3.5 Methanol Synthesis

The optimal (H_2-CO_2)/ (CO+CO_2) ratio should be 2 or just slightly over 2. [7] The synthesis of methanol is around 100 times faster when CO2 is present. The presence of CO2 also slows down the catalyst deactivating. (3). For maximum activity and selectivity the feed should have a CO2 content of 4-8%. (28). The H2O content of the feed should also be kept down to avoid active site blocking. [4]

3.5.1 Catalysts

The most common catalysts of today are low temperature and low pressure catalyst as the Cu/ZnO/Al_2O_3 catalyst that have a lifetime of 3-5 years [7]. They operate at temperatures from 220-275 °C and pressures between 50-100 bar. Around 1 kg MeOH can be produced per liter of catalyst and hour [4].

The main causes of catalyst deactivating are poisoning and Cu-sintering. Cu-sintering are is favored of temperatures above 300 °C to minimize the sintering the temperature should not exceed this temperature. It is also important to keep the chloride in the syngas on a minimum because it has been correlated with a higher sintering rate.

The most important poisons to get rid of are H_2S, chlorine and alkali metals. The H_2S gets absorbed on the catalyst surface and thereby deactivates it, the ZnO in the catalyst works as guard by attracting the H_2S and forms ZnS and $ZnSO_4$ that goes out the reactor with the gas and thereby keeping the copper sites free and minimizing the deactivation. The Cu/ZnO catalysts can absorb 0.4 wt% of sulfur before the activity goes below 70%. The maximum content of H_2S in the gas should be 0.1 ppm.

Chlorine is a strong poison and should be none existing in the gas neither should it be used during maintenance or any other situation where it can come in contact with the catalyst. If introduced as HCl it activates sintering by producing copper chloride. Chlorine can also form compounds that will block active sites, increases sintering rate by other mechanism then the mentioned, it can make H_2S a worse poison, and get the ZnO to form Zn halides that creates more poisoning and sintering problems.

Metals can also poison the catalyst, the most common are the heavy metals, alkali metals and arsenic. [4]

3.5.2 Methanol Reactors

The methanol synthesis can be carried out in both adiabatic reactors and isothermal reactors. The adiabatic reactors are cooled by quenching which means that the temperature is controlled by letting the feed gas enter the reactor at different levels this leads to a saw-tooth formed temperature profile in the reactor. The isothermal reactors uses indirectly cooling were the cooling medium recover the excess heat from the reaction [7]. There are many different versions of both isothermal and adiabatic reactors the most common will be described below.

3.5.2.1 Lurgi Methanol Converter

In 1969 Lurgi came with an isothermal reactor that has become one of the most commercial isothermal methanol converters. The reactor works at an almost isothermal temperature at 230-265 °C and is cooled by water flowing on the side of the catalyst tubes. The reactor operates at pressures between 50-100 bar. Figure 3.6 shows a schematic view of a Lurgi converter. The Lurgi catalyst is made of Cu/ZnO/Cr_2O_3 and it has a lifetime of 36-48 months [4].

3.5.2.2 ICI Low Pressure Quench Converter and ARC Converter

The ICI LP quench converter is the most commonly used adiabatic reactor system for methanol production. The gas is injected to the reactor at different depths and is made up of both fresh and recycled gas. The catalyst is kept in a single bed. The reactor operates at pressures between 50-100 bar. The outlet temperature is 270 °C. The catalyst is made of Cu/ZnO/Al_2O_3 and has a life time of about 36-48 months. Figure 3.6 shows an illustration of an ICI LP quench converter.

The ARC Converter is an improved version of the ICI LP quench converter with the main difference that the catalyst is not kept in one bed, but instead is divided and placed on several distribution plates. The cooling is still made by quenching were a mix of fresh and recycled gas is injected at different places in the reactor in this case it is injected in between the catalyst beds. The injection is very well distributed over the beds making the temperature profile over the catalyst surface even. Due to good gas distribution in the reactor the temperature is lower than in ICI LP quench reactor. These properties

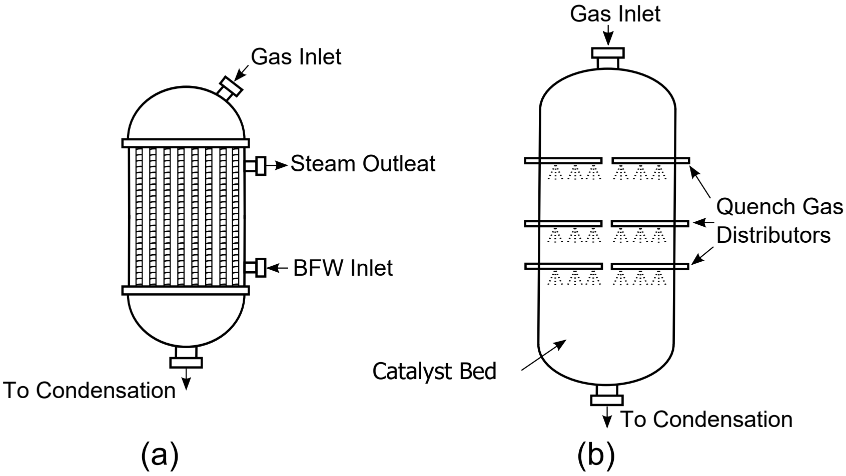


Figure 3.6: Lurgi converter (a) and ICI LP quench converter (b).

prolong the lifetime of the catalyst, increase the activity and decrease the by-product formation. The negative effect of this more complicated system is that the catalyst loading time increases [4].

3.6 Methane reformation

For a standalone methanol plant the methane produced in the gasifier is unwanted. Instead, it is favorable to reform this methane into hydrogen. This can be done using different methods the two most common is auto thermal reforming and steam methane reforming. For co-production of MeOH and SNG the methane could be kept and used as an inert gas in the methanol synthesis, that take up some of the reaction heat. If the syngas consist of very high amount of methane the up scaling cost of the equipment should be taken into consideration. The methane is unwanted in the syngas for a standalone methanol plant, a reformation of the CH_4 content can be performed to produce CO and H_2, through the reaction below.

$$CH_4 + H_2O \rightarrow CO + 3H_2 \tag{3.2}$$

The reaction is endothermic and therefor has a heat demand. The temperature in the reactor should be kept at around 1,000°C. [28]

3.7 Water gas shift

The methanol and methanation processes both needs a special H_2/CO ratio to obtain a good yield. To get the right ratio a Water Gas Shift (WGS) is used. The water gas shift reaction is almost independent of the pressure but have a strong correlation with temperature. The reaction is presented in Eq. 3.3.

$$CO + H_2O \leftrightarrow CO_2 + H_2 \tag{3.3}$$

The reaction is conducted over catalysts, there are several different kinds of catalyst that can be used with different process properties. [29] describes their different catalysts as follows:

- High temperature (HT) CO shift conversion at about 300 to 450 °C down to approx. 2.5% CO on dry basis at the reactor outlet.
- Medium temperature (MT) CO shift conversion or so-called isothermal shift conversion at about 220 to 270 °C down to approx. 0.5% CO on dry basis at the reactor outlet.
- Low temperature (LT) CO shift conversion at about 180 to 250 °C down to approx. 0.2% CO on dry basis at the reactor outlet.

3.8 Chosen Process Technology

When the process technology for the three cases were chosen so the same technology was chosen as far as possible, this was done to make the inaccuracy of the simulations and investment similar and the result more reliable.

The chosen pretreatments technologies are knife chippers and hammer mills for the size reduction of the biomass. This was the only real option found in the literature. A rotary dryer, with steam as drying medium was chosen, is chosen because it is widely used and is a robust technology. There are a large variety of gasifiers to choose between, in the literature CFB-gasifiers were the most common when it came to gasify biomass. It has reasonable demands on the preparation of the material and gives a good product. Therefor an oxygen blown CFB will be selected. Data over the gasifier performance were delivered from one of the supervisors, and was data given by E.ON. The data can be found in Appendix iv, Gasifier data. Downstream the gasifier there will be a cyclone that will clean the gas from debris that has escaped the gasifier together with the gas. In the pure methanol production case 2, a reformer will be chosen to reform the methane. To clean the gas from tars the OLGA-process will be used, it was chosen in discussions with supervisor, because it was considered an interesting solution. The system will be equipped with an ESP. Due to the data over the gasifier are lacking information of tar components; this step will not be simulated. Information from literature concerning costs will be included in the feasibility study.

Inorganic impurities and CO_2 will be removed from the gas with the Rectisol process. This is a well-used process that is robust and reliable. The Selexol is not selected because it absorbs methane which is a desired component that will become a final product in one of the cases and therefore not an option. PSA is not an option because of the decrease in pressure for the process gas. The removed H_2S will go to a Claus process; this process will not be included in this thesis. A guard bed will also be chosen to insure a long catalyst life time.

The methanol synthesis will be carried out in an isotherm reactor. Because of the low convergence the product gas will be cooled off and flashed to remove the produced methanol. In case one, the flashed gas will be led to the methanation step. In case two the pure methanol plant the flashed gas will be pressurized and then recycled back to the reactor. The methanation will be carried out in an isotherm reactor; the product gas will then be upgraded using a PSA. The PSA will not be simulated but will be included in the feasibility study.

A schematic scheme over the chosen technology for each case are shown in Figure 3.7, Figure 3.8 and Figure 3.9.

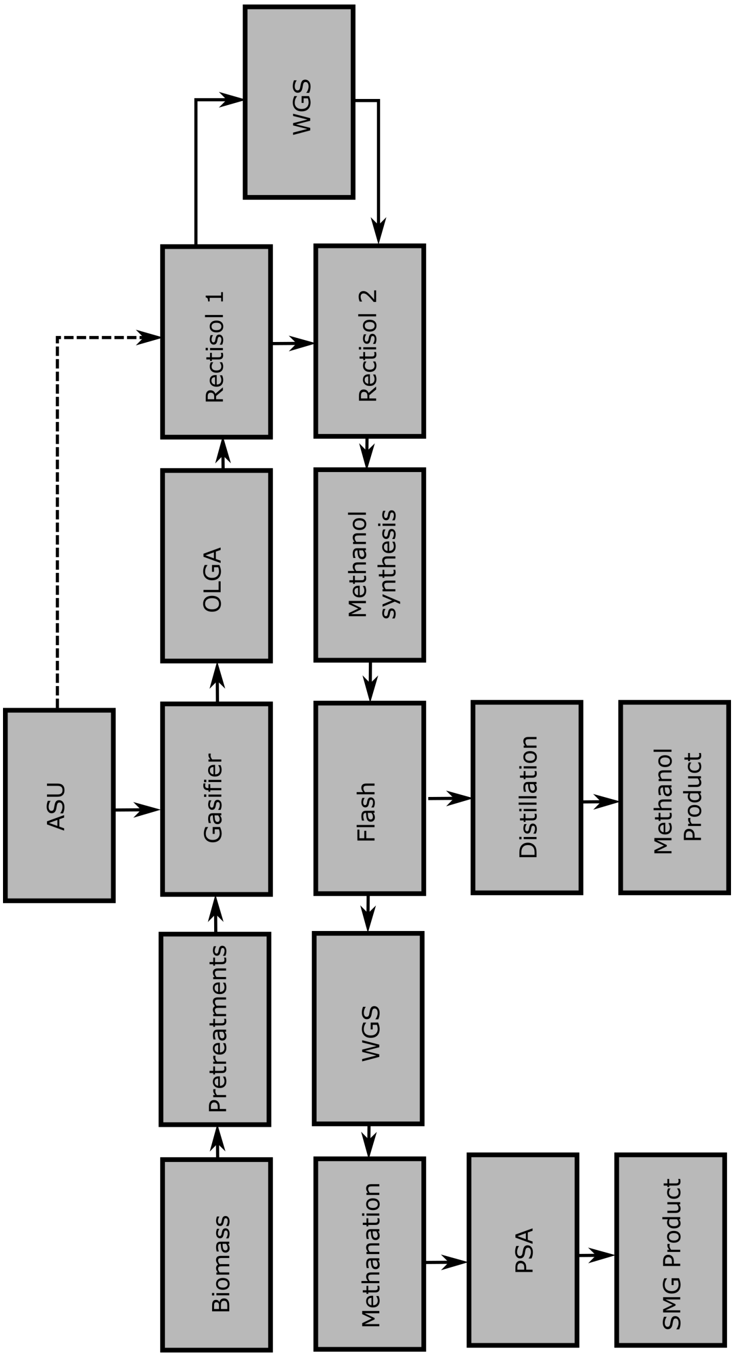


Figure 3.7: A schematic view over case 1.

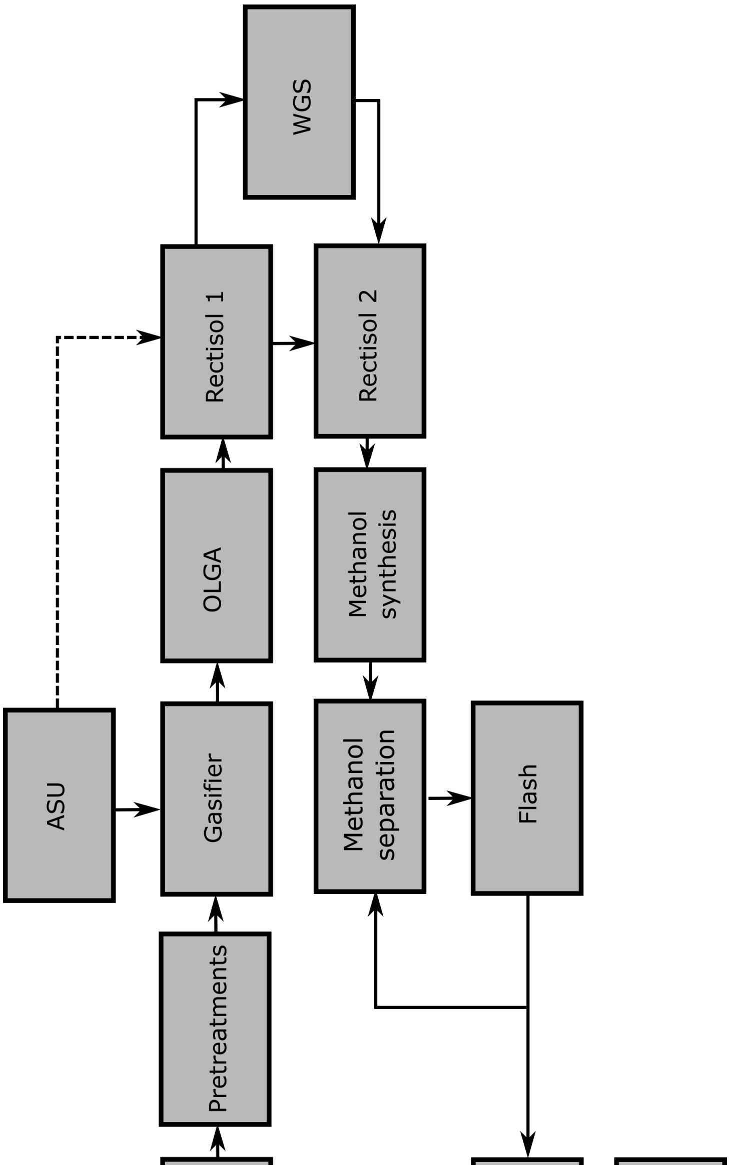


Figure 3.8: A schematic view over case 2.

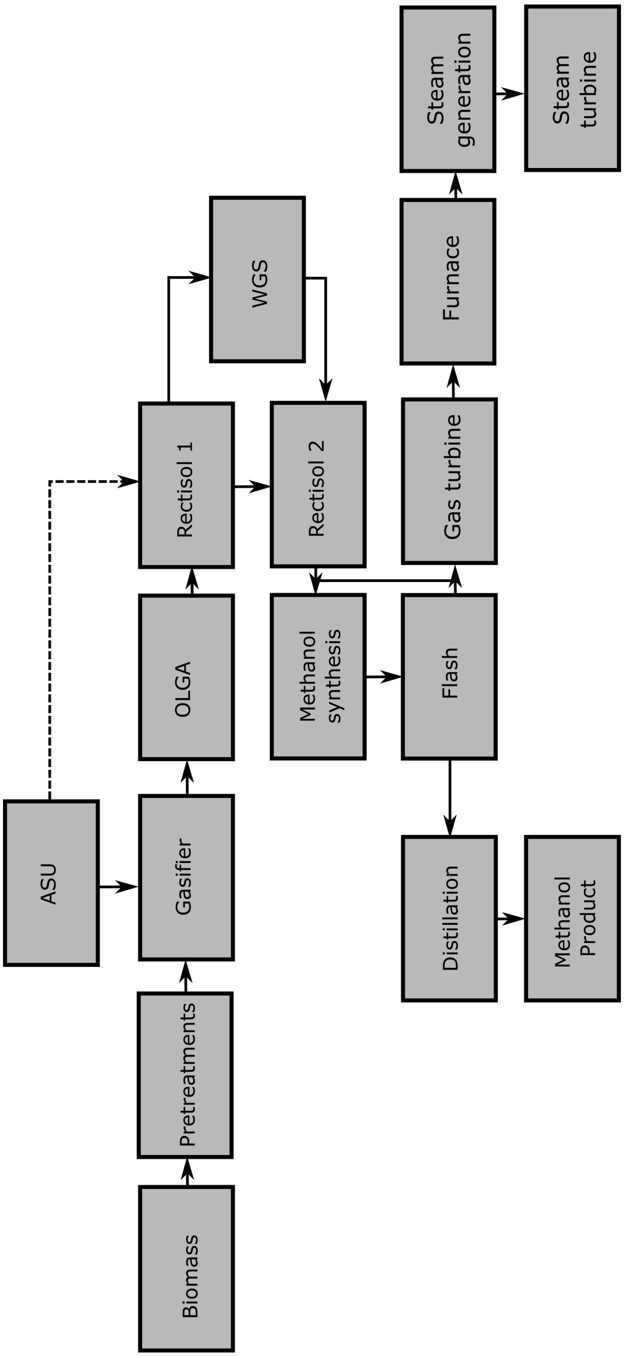


Figure 3.9: A schematic view over case 3.

Chapter 4

Simulation

The simulation was carried out using ASPEN (Advanced System for Process Engineering) plus V8.8, a process modelling tool. ASPEN contains a great physical properties library for chemical substances and have the ability to simulate complex systems and it´s heat-energy balances.

In this thesis the ability to simulate stationary systems were used. The components in the simulated system must be specified, this could limit the accuracy of the result depending the accuracy of the specified components. Due to limitations of handling solids in ASPEN no solids were considered. The input data, such as heater temperature, reactor reactions, and system pressure, will be specified in the *design specifications.* To achieve the correct input of water to the WGS and the methanation step *Designs Specs* were used. The specifications for these can be seen in Appendix ii, Design specs.

4.1 Simulation Specific Configuration

The property method chosen was Peng-Rob throughout the process, this because it was one of the methods recommended, when using the diagrams for choosing a property method provided in the Aspen plus Guideline document. Comparing the recommended property methods to each other by information given in the guideline, Peng-Rob seemed the most fitting. [30] All reactors were based on REquil, an equilibrium reactor, except the reformer which was simulated as an RGibbs reactor that is based on Gibbs free energy. The columns were all set up as Radfrac. The compressors were all run as isentropic compressors with and the default efficiency of 0.72, the pumps were also have the default efficiency, and this is variable value. It is assumed that the heat-exchangers have full efficiency, meaning the heat transfer only takes place between the cold and the hot side and not to the surrounding area.

The simulation was complex and it was hard to get the simulations to converge, this was very time consuming. It also meant that small adjustment often lead to the simulation not converging and it was hard to optimize the process in the simulation. In order to regenerate the methanol from both stripper and rectifier a second column, that didn´t recycle back the methanol to the process, had to be added after each step. There were also problems getting the ASPEN Analyzer to perform an optimization of the heat exchanger network. The program complained about segmentation temperature being too small, an adjustment was made to this temperature but then the program stopped working while running. Therefore the heat and cooling demand is based on the grand composite curve and the calculated targets.

4.2 Simulation results

The results from the simulation shows that Case 2, has the highest efficiency of $\dot{Q}_{MW}$. The yield results of all three cases are calculated in appendix vi, and the result is presented in Figure 4.1.

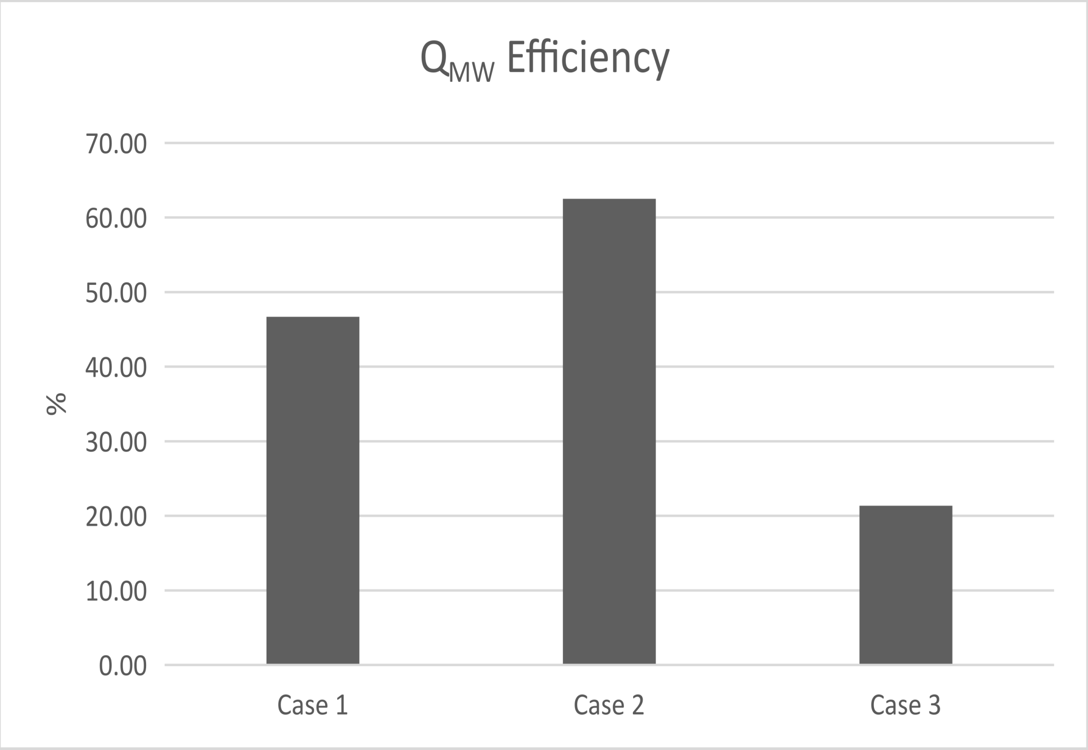


Figure 4.1: The $\dot{Q}_{MW}$ yield for the three cases.

The carbon efficiency is substantially higher in the case 2 compared to cases 1 and 3, as shown in Figure 4.2.

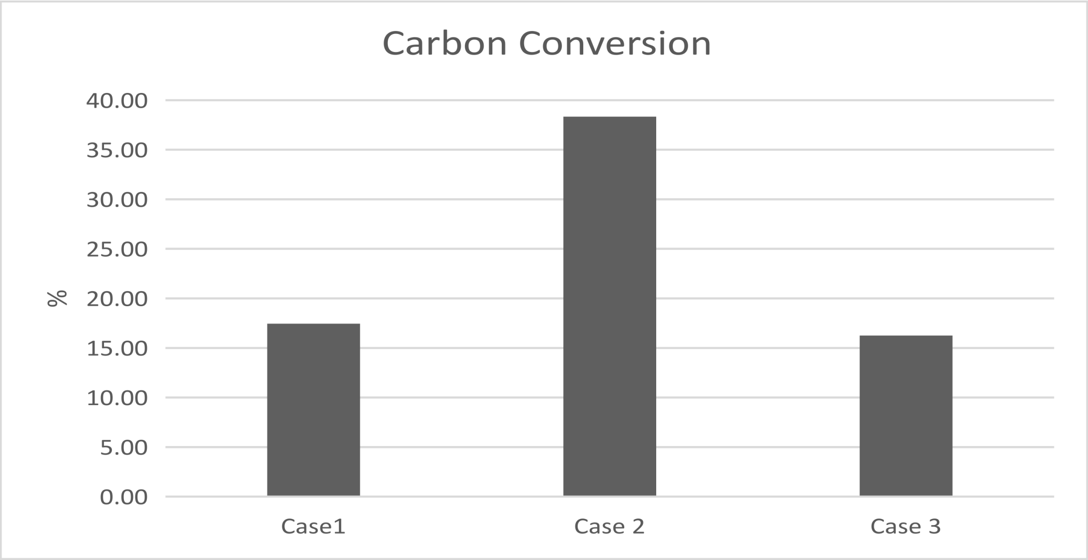


Figure 4.2: Carbon efficiencies for the three cases .

The input of raw material was similar in the three cases, the produced varied in both amount and type, the resulting product flow is presented in Figure 4.3.

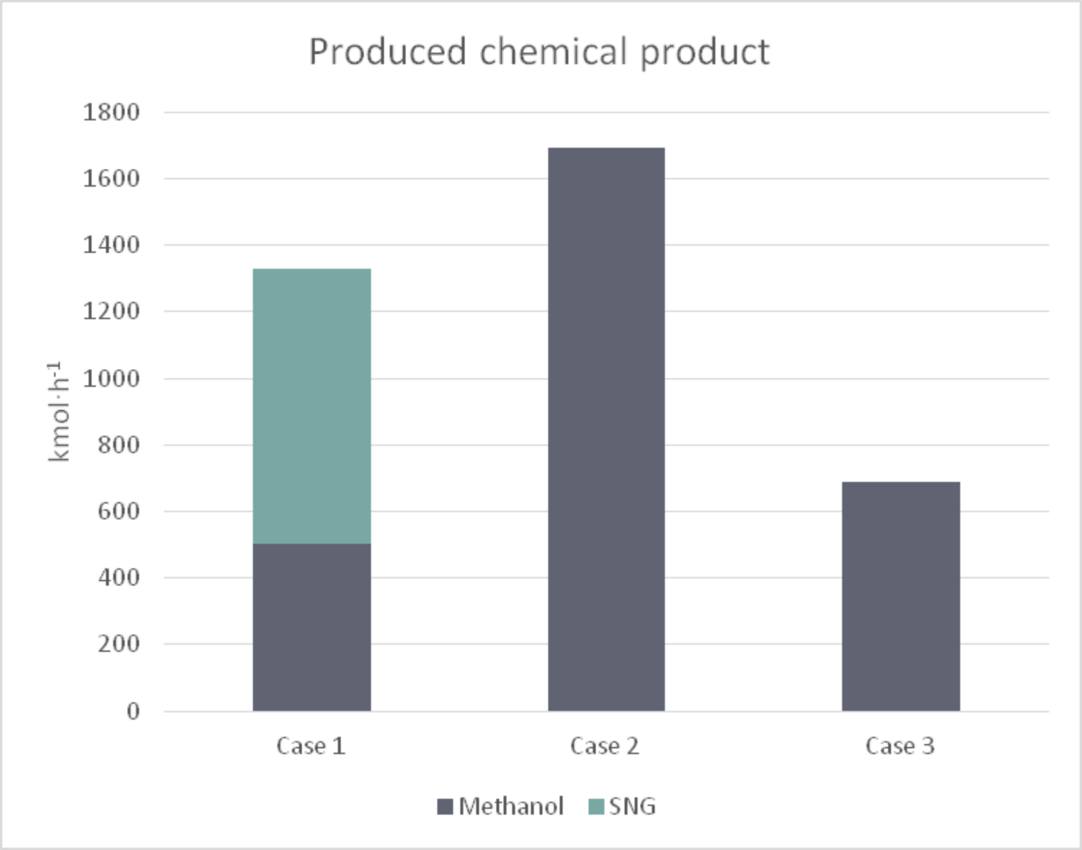


Figure 4.3: Mole flow of chemical products in the three cases.

The energy content in the three cases is presented per product for the three cases in Figure 4.4. The energy content are highest in case 2, and by far the lowest in case 3. In case 1 The produced SNG carries the most energy.

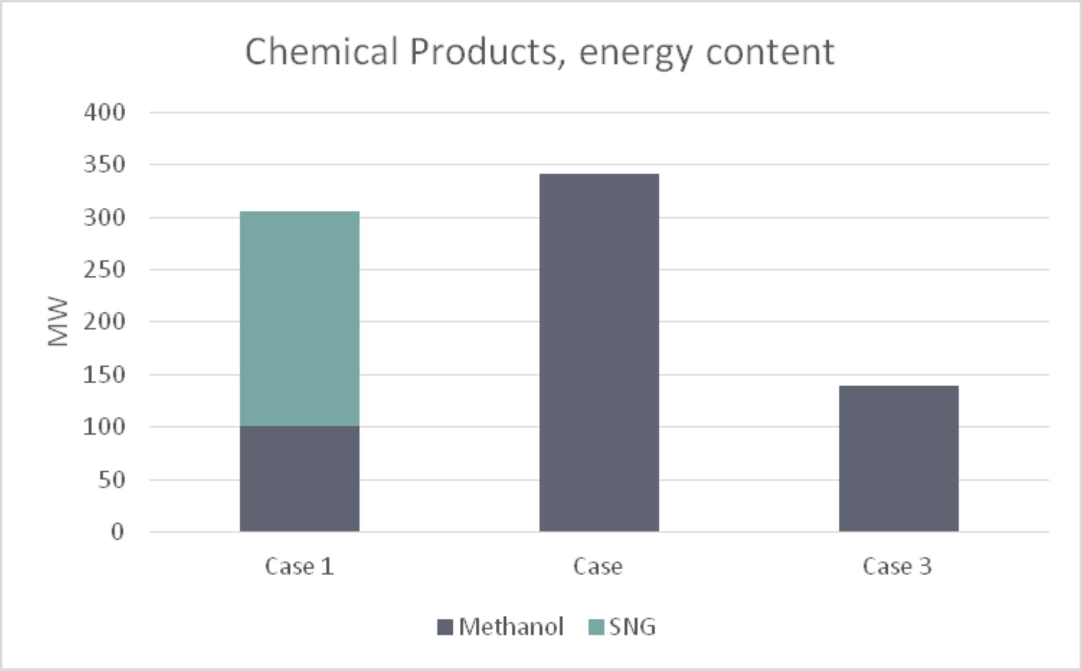


Figure 4.4: Energy content of the three cases per product.

The electricity consumption for the different cases are presented in Figure 4.5, for case three the produced electricity is included as a negative value. The charts shows that case 2 has the highest consumption follow by case 1.

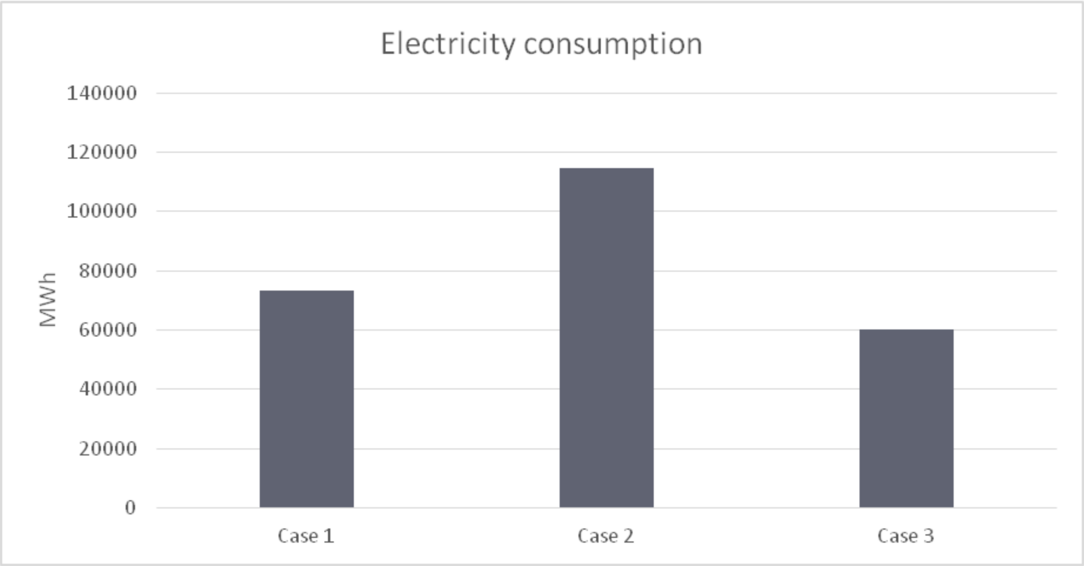


Figure 4.5: Electricity consumption.

In Figure 4.6, the electricity consumption divided on pumps and compressors per case is shown. The compressors make up the highest electricity consumption. Case 1 has considerable lower electricity consumption from its compressors compared to the other two cases.

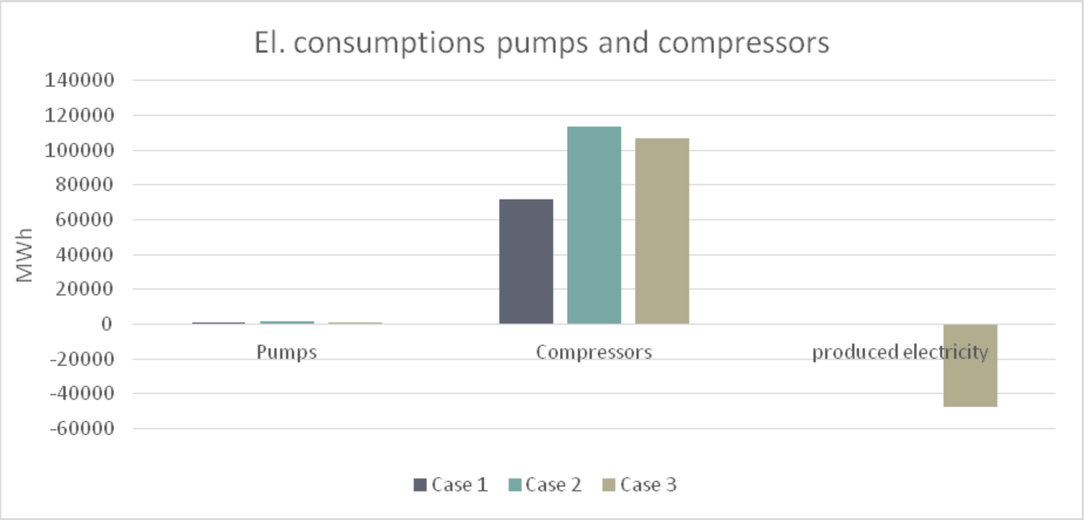


Figure 4.6: Shows electricity consumption divided on pumps and compressors.

Chapter 5

Feasibility Study

When calculating the feasibility of the investment present value method is used. It calculates the present value of the future receipts, payments and investments.

It uses:
G – Initial investment
I – Yearly receipts
P – Yearly payments
$a = I - P$ – Yearly excess receipts
n – Economical lifetime of investment
S – Residual value
i – Discount rate

G, includes everything that has with the initial investment completion. i, includes both the inflation and the real rate of interest, a nominal rate of interest. Following is assumed; n is set to 30 years, i is set to 10% including an inflation predicted to be 2%. S is set to have no value. The Present Value can then be calculated by Eq. 5.1.

$$Present\,Value = -G + a\frac{(1+i)^n - 1}{i\,(1+i)^n} \tag{5.1}$$

The investment cost is determined by the size of the plant and are considered fixed cost that cannot be influenced after the plant has been built. Yearly payments e.g. variable costs, consist of consumption of raw material, chemicals, electricity, salaries and taxes. The variable costs are mainly determined by the production volume. Some of the cost follows the production volume strictly while others are more elastic such as personal cost. This plant makes earnings on several different products; methanol, SNG, district heating and electricity certificate. Because the electricity is produced out of renewable energy source electricity certificate is obtained, these can be sold on and can thereby give earnings. It is assumed that all products can be sold to full price. In this study maximum production for 8 000 h per year is assumed. It is also assumed that the plant is located in Sweden.

5.1 Investment Costs

The objective of this thesis is to research if a combined production of methanol and SNG can be feasible and perhaps more profitable than a standalone MeOH plant Investment Cost are estimated by base costs given in literature, it is assumed that it is a NTH plant, a mature plant. In literature it is estimated that this kind of calculations have an accuracy of ± 30%. The estimated investment cost of a component are calculated according to Eq. 5.2-5.3. The base cost, C_0, is the cost for the component at capacity S_0. The Scaling factor f ranges generally between 0.6 and 0.8.

$$C = C_0 \left(\frac{S}{S_0}\right)^f \tag{5.2}$$

The base costs often have a capacity limit, also given by the literature. If the required capacity is greater than the given upper limit multiple trains can be used. The estimated cost for the additionally installed trains are considered to have a lower investment cost than the first. The trains are given the same size. The trained cost, Cm, depends on the number of trains, n, and the scaling exponent, m, with the value of 0.9, and is calculated according to Eq. 5.3:

$$C_m = Cn^m \tag{5.3}$$

To estimate the direct cost some literature use BOP and others use installation factor and each equipment's direct cost will be estimated individual using given method. Some investment costs were calculated using Ulrich method, in the online program provided by www.ulrichvasudesign.com

For contingencies an additional 18% will be added to the investment cost and then further 30% for surrounding buildings and the like, according to supervisor Christian Hulteberg.

The given base costs from literature is valid for different years and to get them all up to date the Chemical Engineering Plant cost index, CEPCI, were used. The value used for 2018 was 603.1. The investment cost are also given in different currencies and therefor they have all been converted to MSEK.

The area size for the Heat Exchanger Network (HEN) were based on simulations performed in ASPEN Plus, and the target area calculated from grand composite curve.

5.1.1 Investment cost case 1

The details and results of the investment cost calculations are presented in Table 5.1.

Table 5.1: Details and results of the investment cost calculations

Process Unit	S_0	Unit of Capacity	C_0	Base Cost Unit	f_s	f_i	S_r	MSEK 2019	Ref
Conveyers	33.5	wet tonne/h	0.41	M€2002	0.8	2	125.5	41.47	a
Grinding	33.5	wet tonne/h	0.48	M€2002	0.6	2	125.5	42.82	a
Storage	33.5	wet tonne/h	1.16	M€2002	0.65	2	125.5	106.79	a
Dryer	33.5	wet tonne/h	8.5	M€2002	0.8	2	125.5	859.76	a
Ironremoval	33.5	wet tonne/h	0.43	M€2002	0.7	2	125.5	40.85	a
Feedingsystem	33.5	wet tonne/h	0.48	M€2002	1	2	125.5	55.04	a
Air separation unit 99.5% O_2	576	tonne O_2/ day	27.9	M€2002	0.75	1.3	323.15	385.68	a
CFB	68.8	dry tonne/h	44.3	M€2002	0.7	1.69	69.02	1 230.9	a
Cyclone	34.2	m^3gas/s	3	M€2002	0.7	2	15.21	55.81	a
OLGA	1	N m^3/h gas feed	0.0002	M€2005	1	2	54 742.37	303.50	a
Guard Bed (ZnO+active C)	8	m^3gas/s	0.024	M€2002	1	3	323.15	47.71	a
Rectisol	200000	N m^3/h gas feed	20	M€2003	0.65	1.86	54 742.37	231.27	a
WGS reactor	8819	kmol CO+H_2/h	12.2	M€2002	0.65	1.81	431.98	50.99	a
Compressor CM1	13.2	MWe	12.9	M€2002	0.85	1.86	2.18	85.16	a
Compressor CM2	13.2	MWe	12.9	M€2002	0.85	1.86	2.74	103.43	a
Compressor CM3	13.2	MWe	12.9	M€2002	0.85	1.86	2.37	91.43	a
Methanol reactor. MEOHR	87.5	tMEOH/h	3.5	M€2002	0.72	2.1	16.67	36.53	a
Methanol product separation/purific.	4.66	kg MEOH/s	1.72	M$2002	0.291	1.15	4.63	29.06	c
WGS reactor	708	Kmol CO + H_2/h	12.2	M€2002	0.65	1.81	431.98	70.30	a
methanation reactor 2. CH4R2	175	$MW_{HHV.SNG}$	4.38	M€2012	0.72	2.1	215.69	118.71	d
methanation flash	8.85	m^3	0.013	M€2010	0.67	2.1	746.16	6.28	d
PSA	9600	kmol	32.6	M€2003	0.7	1.86	312.64	89.05	b
Shell-and-tube heat exchanger	80	m^2	0.032	M$2000	0.68	1.86	41280	69.38	c
Turbine HP		kW		$2004			4642	12.07	e
Turbine MP		kW		$2004			145	2.52	e
Compressor Lpsteam		kW		$2004			259	10.70	e
							TOT	4226	

a)Hamelinck, C. N., Faaij, A. P., den Uil, H., & Boerrigter, H. (2004). Production of FT transportation fuels from biomass; technical options, process analysis and optimisation, and development potential. Energy, 29(11), 1743-1771.

b)Liu, G., Larson, E. D., Williams, R. H., Kreutz, T. G., & Guo, X. (2010). Making Fischer– Tropsch fuels and electricity from coal and biomass: performance and cost analysis. Energy & Fuels, 25(1), 415-437. c)Kemp, I. C. (2005). Chemical Process Design and Integration, Robin Smith, John Wiley & Sons Ltd (2005), 687 pp, Hardback: ISBN 0 471 48680 9,£ 85.00, Paperback: ISBN 0 471 48681 7,£ 39.95. Ulrich d)Database, supplied by www.ulrichvasudesign.com/econ.html

5.1.2 Investment cost case 2

The details and results of the investment cost calculations are presented in Table 5.2.

Table 5.2: Details and results of the investment cost calculations

Process Unit	S_0	Unit of Capacity	C_0	Base Cost Unit	f_s	f_i	S_r	SEK 2019	Ref
Conveyers	33.5	wet tonne/h	0.41	M€2002	0.8	2	125.5	41.47	a
Grinding	33.5	wet tonne/h	0.48	M€2002	0.6	2	125.5	42.82	a
Storage	33.5	wet tonne/h	1.16	M€2002	0.65	2	125.5	106.79	a
Dryer	33.5	wet tonne/h	8.5	M€2002	0.8	2	125.5	859.76	a
Ironremoval	33.5	wet tonne/h	0.43	M€2002	0.7	2	125.5	40.85	a
Feedingsystem	33.5	wet tonne/h	0.48	M€2002	1	2	125.5	55.04	a
Air separation unit 99.5% O_2	576	tonne O_2/ day	27.9	M€2002	0.75	1.3	323.15	385.68	a
CFB	68.8	dry tonne/h	44.3	M€2002	0.7	1.69	69.02	1 230.9	a
Reformer	100	m3NTP/S	31.1	M€2002	0.6	2.3	15.21	379.01	a
Cyclone	34.2	m^3gas/s	3	M€2002	0.7	2	15.21	55.81	a
OLGA	1	N m^3/h gas feed	0.0002	M€2005	1	2	76591.18	424.63	a
Guard Bed (ZnO+active C)	8	m^3gas/s	0.024	M€2002	1	3	21.28	3.14	a
Rectisol	200000	N m^3/h gas feed	20	M€2003	0.65	1.86	76591.18	287.69	a
WGS reactor	8819	kmol CO+H_2/h	12.2	M€2002	0.65	1.81	1997.51	137.97	a
Compressor CM1	13.2	MWe	12.9	M€2002	0.85	2.1	4.01	161.41	a
Compressor CM2	13.2	MWe	12.9	M€2002	0.85	2.1	4.46	176.69	a
Compressor CM3	13.2	MWe	12.9	M€2002	0.85	2.1	3.74	152.13	a
Methanol reactor. MEOHR	87.5	tMEOH/h	3.5	M€2002	0.72	2.1	54.59	85.85	a
Recycle Compressor	13.2	MWe	12.9	M€2002	0.85	2.1	0.16	10.44	a
Methanol product separation/purific.	4.66	kg MEOH/s	1.72	M$2002	0.291	1.15	15.16	41.05	c
WGS reactor	708	Kmol CO + H_2/h	12.2	M€2002	0.65	1.81	431.98	70.30	a
Shell-and-tube heat exchanger	80	m^2	0.0324	M$2000	0.68	1.86	160960	156.49	c
Cooling tower	10	m^3/h	0.00443	M$2000	0.63	1.86	11242	10.14	c
Incinerator-steam production		kW		$2004			2.97	27.78	e
Compressor Lpsteam		kW		$2004			416.80	16.17	e
							TOT	4890	

5.1.3 Investment cost case 3

The details and results of the investment cost calculations are presented in Table 5.3.

Table 5.3: Details and results of the investment cost calculations

Process Unit	S_0	Unit of Capacity	C_0	Base Cost Unit	f_s	f_i	S_r	MSEK 2019	Ref
Conveyers	33.5	wet tonne/h	0.41	M€2002	0.8	2	125.5	41.5	a
Grinding	33.5	wet tonne/h	0.48	M€2002	0.6	2	125.5	42.8	a
Storage	33.5	wet tonne/h	1.16	M€2002	0.65	2	125.5	106.8	a
Dryer	33.5	wet tonne/h	8.5	M€2002	0.8	2	125.5	859.8	a
Ironremoval	33.5	wet tonne/h	0.43	M€2002	0.7	2	125.5	40.8	a
Feedingsystem	33.5	wet tonne/h	0.48	M€2002	1	2	125.5	55.0	a
Air separation unit 99.5% O_2	576	tonne O_2/ day	27.9	M€2002	0.75	1.3	323.15	385.7	a
CFB	68.8	dry tonne/h	44.3	M€2002	0.7	1.69	69.02	1 230.9	a
Cyclone	34.2	m^{3}gas/s	3	M€2002	0.7	2	15.21	55.81	a
OLGA	1	N m^3/h gas feed	0.0002	M€2005	1	2	54 742.37	303.50	a
Guard Bed (ZnO+active C)	8	m^{3}gas/s	0.024	M€2002	1	3	323.15	47.7	a
Rectisol	200000	N m^3/h gas feed	20	M€2003	0.65	1.86	54 742.37	231.3	b
WGS reactor	8819	kmol CO+H_2/h	12.2	M€2002	0.65	1.81	431.98	50.99	a
Compressor CM1	13.2	MWe	12.9	M€2002	0.85	1.86	2.18	85.2	a
Compressor CM2	13.2	MWe	12.9	M€2002	0.85	1.86	2.74	103.4	a
Compressor CM3	13.2	MWe	12.9	M€2002	0.85	1.86	2.37	91.4	a
Methanol reactor. MEOHR	87.5	tMEOH/h	3.5	M€2002	0.72	2.1	22.30	45.1	a
Recycle Compressor	13.2	MWe	12.9	M€2002	0.85	1.86	0.00	0.4	a
Methanol product separation/purific.	4.66	kg MEOH/s	1.72	M$2002	0.291	1.15	6.14	31.6	c
Shell-and-tube heat exchanger	80	m^2	0.0324	M$2000	0.68	1.86	434400	343.8	c
Cooling tower		m^3/s		M$2004			2.10	13.5	e
Furnace		kW		$2004			11761	581.2	e
Turbine HP		kW		$2004			10338	17.3	e
Turbine MP		kW		$2004			598.62	4.8	e
Gas turbine		kW		$2004			525	64	e
Compressor Lpsteam		kW		$2004			360.35	4.2	e
							TOT	4781	

5.1.4 Compilation over investment cost results

The distribution of the investment costs for the three cases, divided on the different process sections, is presented in Figures 5.1-5.3.

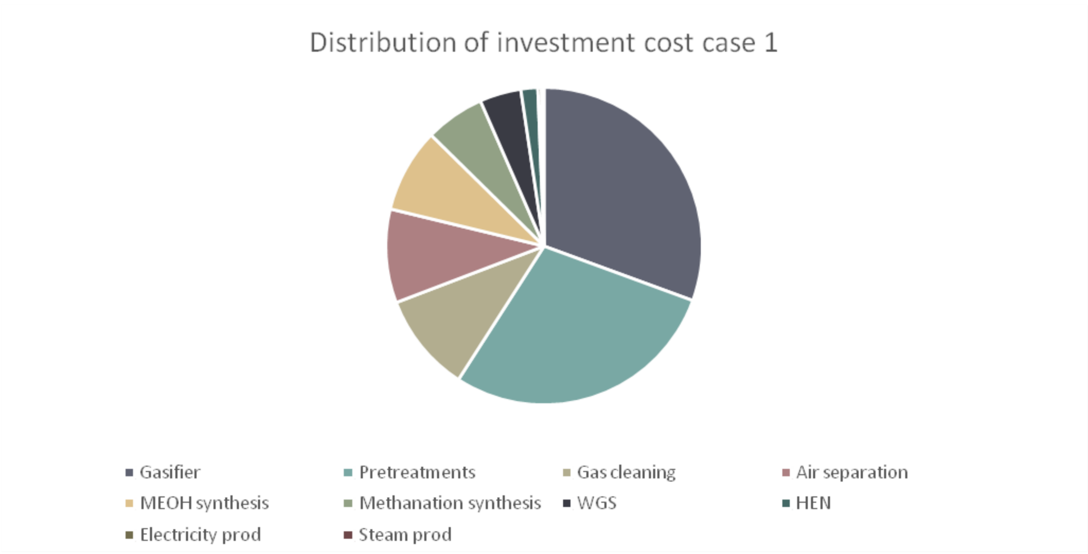


Figure 5.1: The distribution of investment costs in case 1.

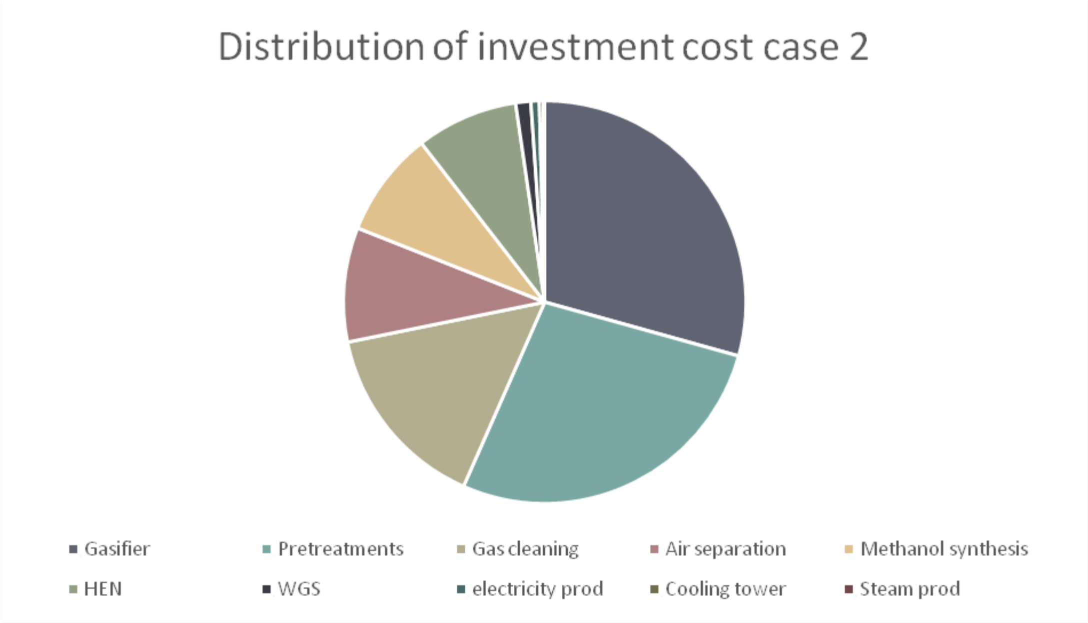


Figure 5.2: The distribution of investment costs in case 2.

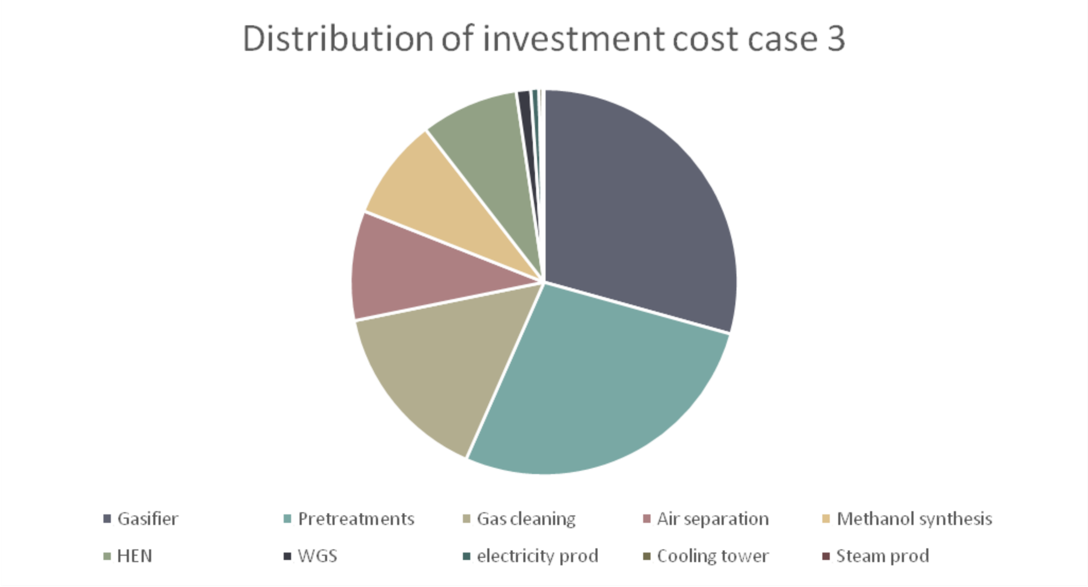


Figure 5.3: The distribution of investment costs in case 3.

The total investment cost for the three cases are compiled in Table 5.4.

Table 5.4: Compilation of the total investment costs for the three cases.

	Case 1	Case 2	Case 3
Investment Cost. MSEK	4 226	4 890	4 781
$f_{contingency}$ =1.18	4 987	5 770	5 642
$f_{building}$ =1.3	6 483	7 501	7 334
G	6 483	7 501	7 334

The results show that cases 1 has the lowest investment cost, followed by case 2 and case 3 having the highest investment cost.

5.1.5 Variable Costs

The Variable cost are made up of electricity cost, raw material, salaries and so on. It is assumed that the electricity, and raw material are the variable costs that differs between the cases. Other costs are not included in the feasibility study. Electricity produced by gas turbine in the simulation is included as a negative post. The dryer is not in the simulation, therefore the steam requirements for the dryer are calculated in Appendix xii, and added to the LP steam requirement. The steam into the gasifier are calculated in Appendix ix, Steam demand gasifier.

The variable costs used to calculate the total variable costs per case is presented in Table 5.5.

The results of the variable costs for each case are presented in Table 5.6, and is also presented in Figure 5.4. The result shows that case 1 has the lowest variable cost, followed by case 3. Case 2 has the highest cost.

Table 5.5: Variable cost per source.

Source		Unit	Ref.
Biomass	70	SEK/MWh biomass	[30]
Electricity	1	SEK/kWh	assumed
OLGA	140	SEK/tonne biomass	[24]
Water Cooling	1.39	SEK/m^3	[24]
ASU	0.22	kW/$Nm^3 \cdot s$	[31]
PSA	0.3	kWh/$Nm^3 \cdot s$	[32]

Table 5.6: Total varible cost for each case.

Variable Costs	Case 1	Case 2	Case 3
Biomass	298.65	300.7	298.55
Refrigeration	10.54	9.36	18.93
Steam	-2.1	-3.3	-2.9
Cooling Water	-	96.8	64.8
OLGA	126.1	126.1	126.1
ASU	0.02	0.02	0.02
PSA	0.00002	-	-
Pumps	1.2	1.4	10.2
Compressors	71.96	113.3	106.4
TOT	508.5	655.6	622.2

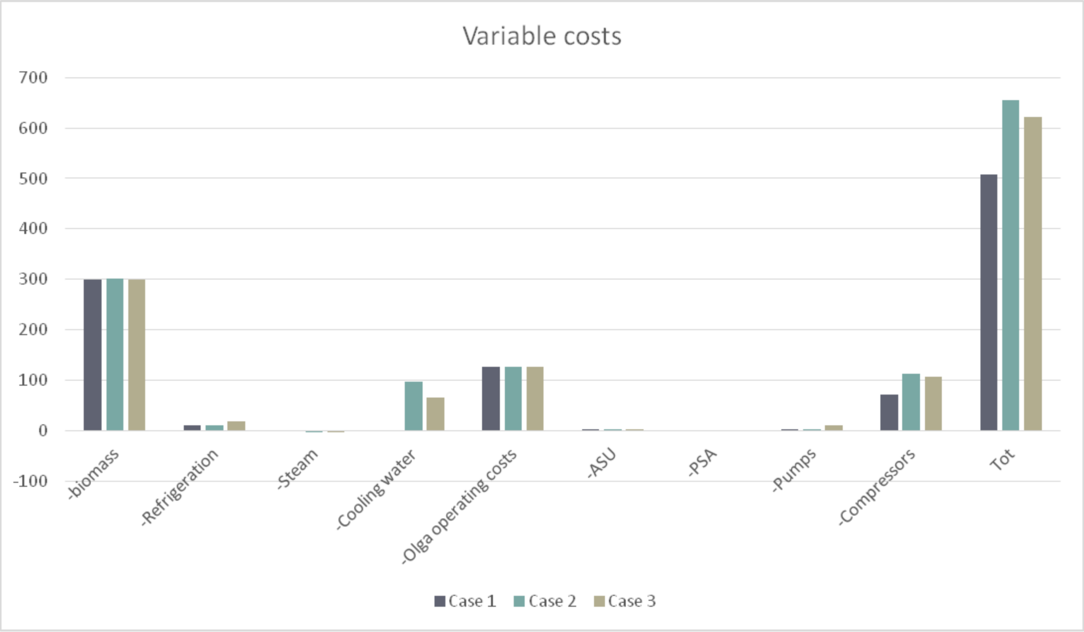


Figure 5.4: Variable costs for each case.

5.2 Revenue

The products vary per case so does the amount. For this calculations it is assumed that the returning water from district heating have a temperature of 50°C [33] and △T is 30°C in the heat exchanger it is also assumed that the district heat water is pure water. All excess LP steam is assumed to be used as district heating, with efficiency in the heat exchanger at 100%. The electrical certificate system in Sweden gives earnings for green produced electrical kWh; therefore also this is considered a product. The sales price for each product is compiled in Table 5.7.

Product	Price	Unit	Ref
Methanol	3873.6	SEK/tonne	[3]
SNG	0.535	SEK/kWh	[8]
District heat	0.72	SEK/kWh	[34]
Electricity	1	SEK/kWh	assumed
Electrial certificate	0.00668	SEK/kWh	[35]

Table 5.7: Sales price for each product.

The result of the revenue for the different cases is shown in Table 5.8. The result shows that case two has a revenue that is about 30% higher than case 1 and 2 times case 3.

Table 5.8: Revenues per case.

	Case 1	Case 2	Case 3
Products	MSEK		
Methanol	469.36	1 651.72	655.40
SNG	766.45	-	-
District heat	-	-	70.1
Electricity	38.30	-	82.7
Electrical certificate	2.56	-	7.5
Total Revenue	1 276.7	1 651.7	815.7

The way the revenues are divided on the different products in each case is presented in Figure 5.5-5.7.

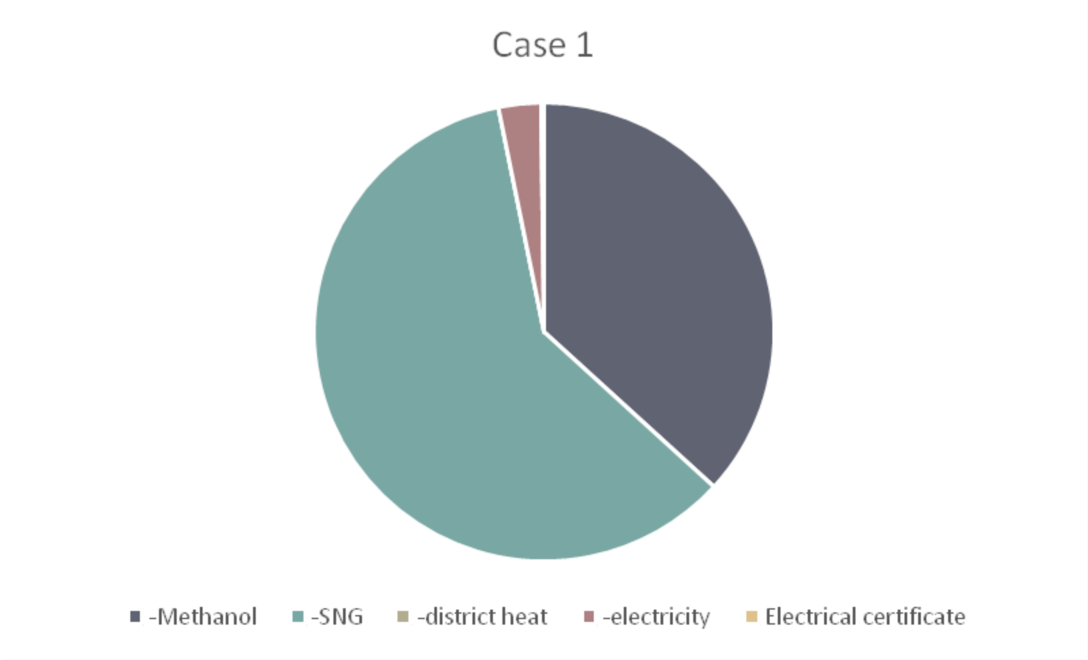


Figure 5.5: Revenues distribution in case 1.

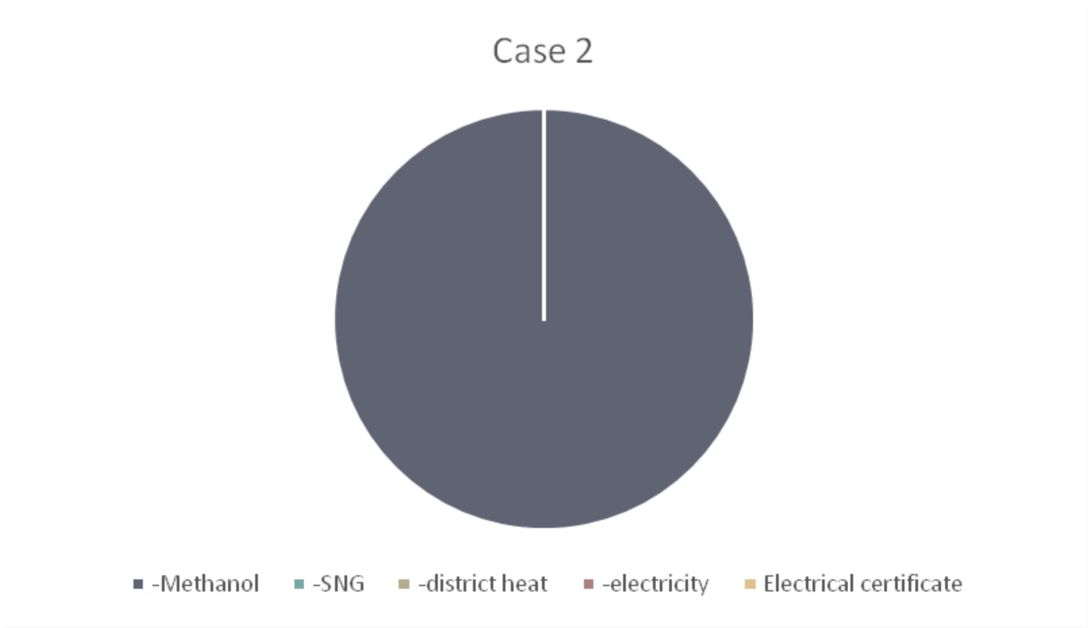


Figure 5.6: Revenues distribution in case 2.

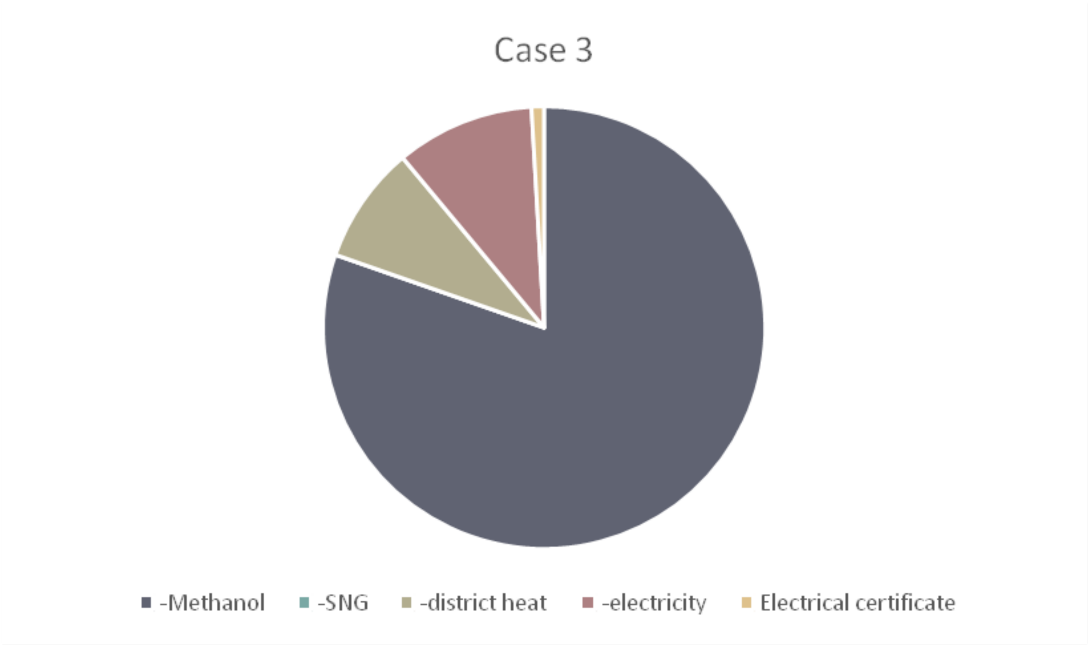


Figure 5.7: Revenues distribution in case 3.

5.3 Feasibility Results

The feasibility calculated with the present value, the results are compiled in Table 5.9.

Table 5.9: Compilation of feasability study.

	Case 1	Case 2	Case 3
	MSEK		
G	4 226	4 890	4 781
I	1 276.7	1 651.7	815.7
U	508.5	655.6	622.2
a	768.2	996.1	193.5
$Present\,Value$	3 015	4 499	-2 957

The result of the study show that case 1 and 2 both are feasible under the given assumptions. The result also show that case 2 are the most profitable investment of the two. The result also shows that case 3 is not feasible.

Chapter 6

Conclusions and future work

In times when the climate change is an everyday discussion it is important that every stone is turned to find new way to produce green energy for different use.

The results in this thesis indicate that the co-production plant is a feasible investment but not the most feasible of the three different cases. It shows that case 1 has the lowest investment cost and the lowest variable cost, but case 2 has the highest revenues. This indicates that the sales price on the products has a great impact on the result. A sensitivity study on sales price and demand could be carried out as future work to see how these factors contributes to the result. The feasibility study is not distinct due to the lack of some variable cost such as labor and assumptions that have to be investigated more proper, to get a more accurate result this should be investigated.

This study indicates that it is more profitable to produce the methanol without co-production with SNG. The result shows that case 1 uses less compressor energy then case 2 and 3.The decreased electricity use achieved when the syngas is not recycled, are not enough to make up the revenue loss. But for some applications it might be favorable depending on the marketing of products. It would be interesting in future work to see if another product with higher revenue, that can be co-produced with MeOH could result in a better outcome.

The chosen technologiess were chosen to work for all three cases and might not been optimal for every case. A gasifier that has less methane in its product gas could for example maybe be a better choice for case 2 and 3. It may therefore be a good idea to optimize the process for a pure methanol plant, to investigate if the difference in feasibility increases. That could decrease the incentive to erect a CO-production plant.

The results in this thesis contradict the theory that co-production with SNG is more feasible than pure bio-methanol production.

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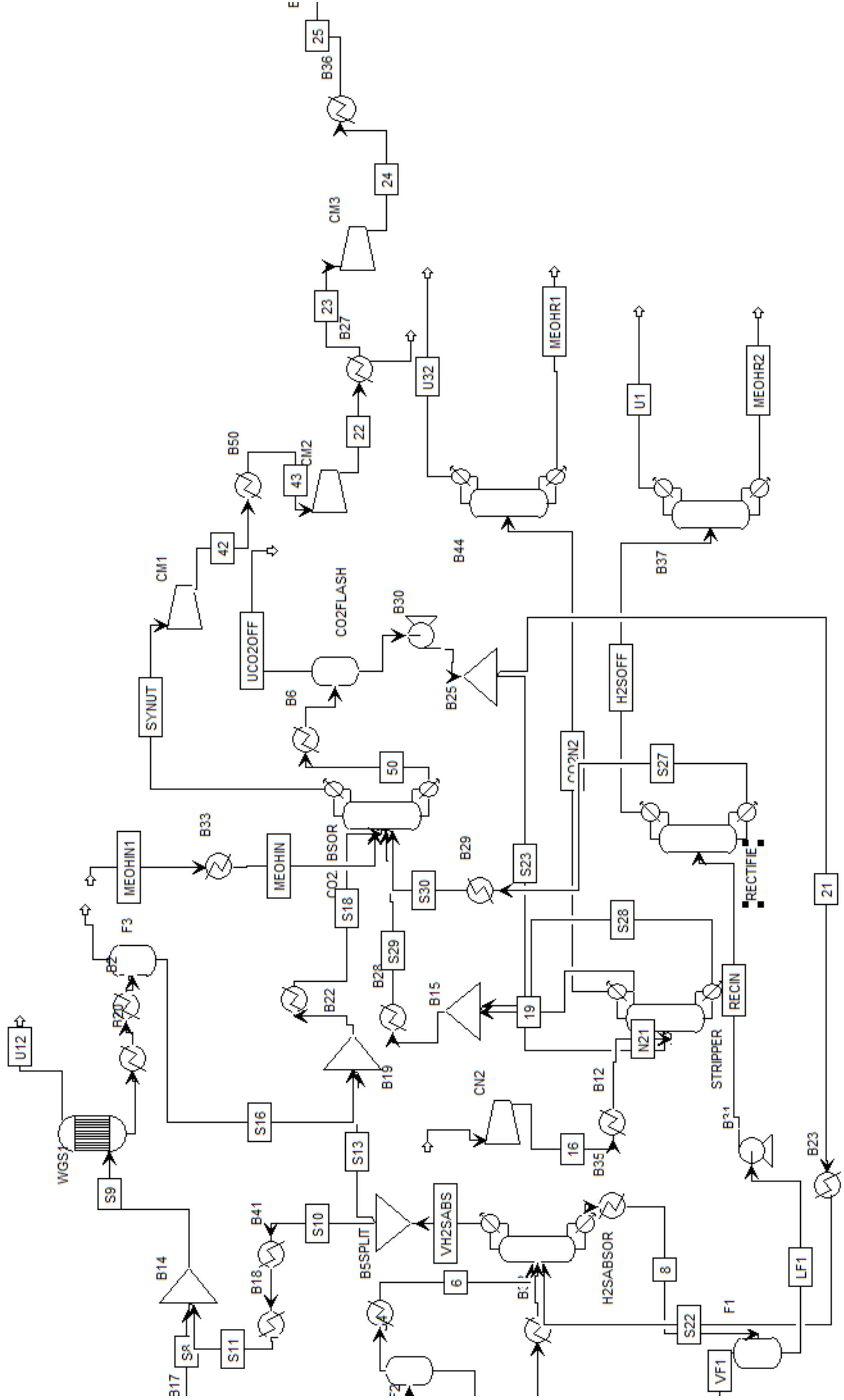
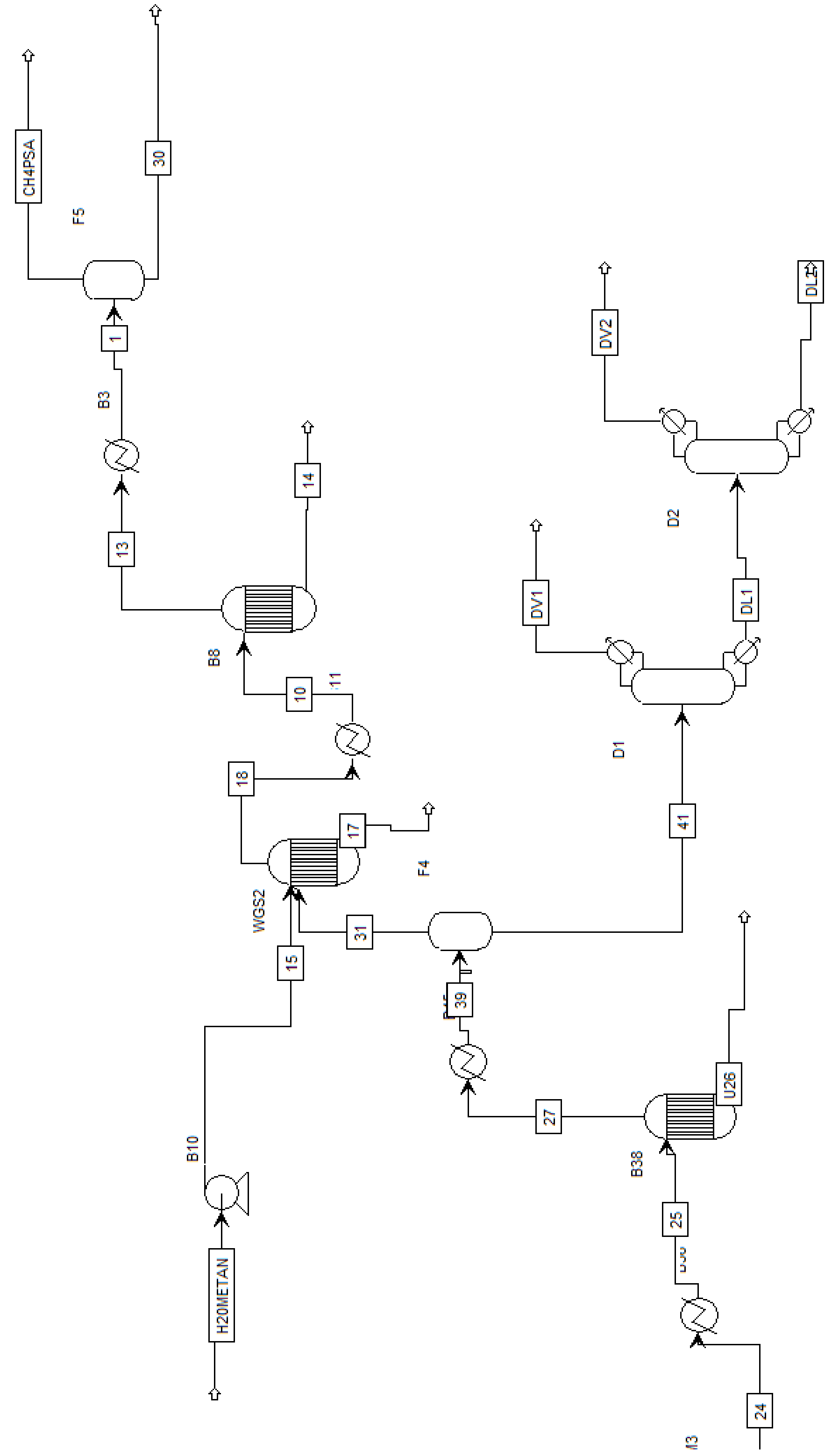
Appendix i, Simulation process schemes

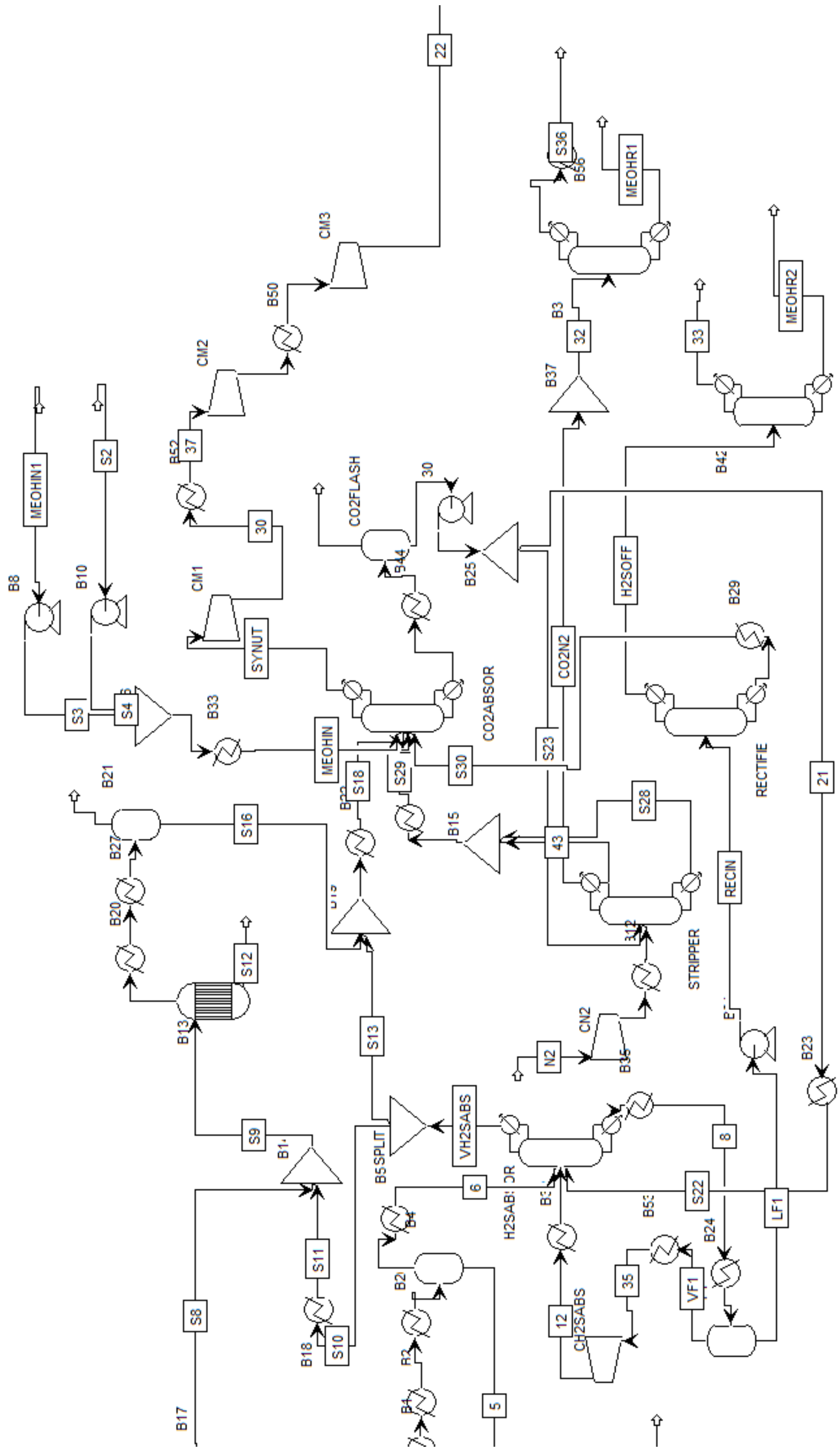
Case 1 Process scheme part 1 of simulation



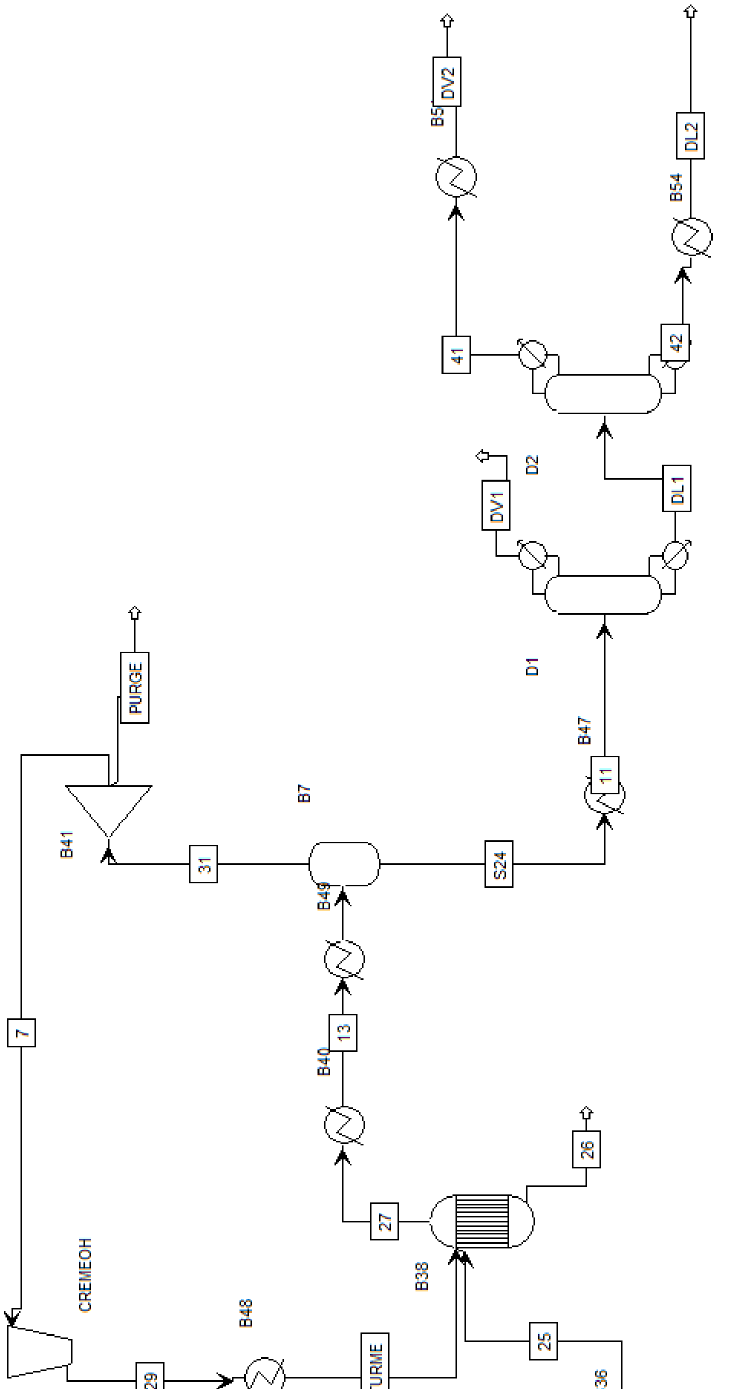
Case 1 Process scheme part 2 of simulation



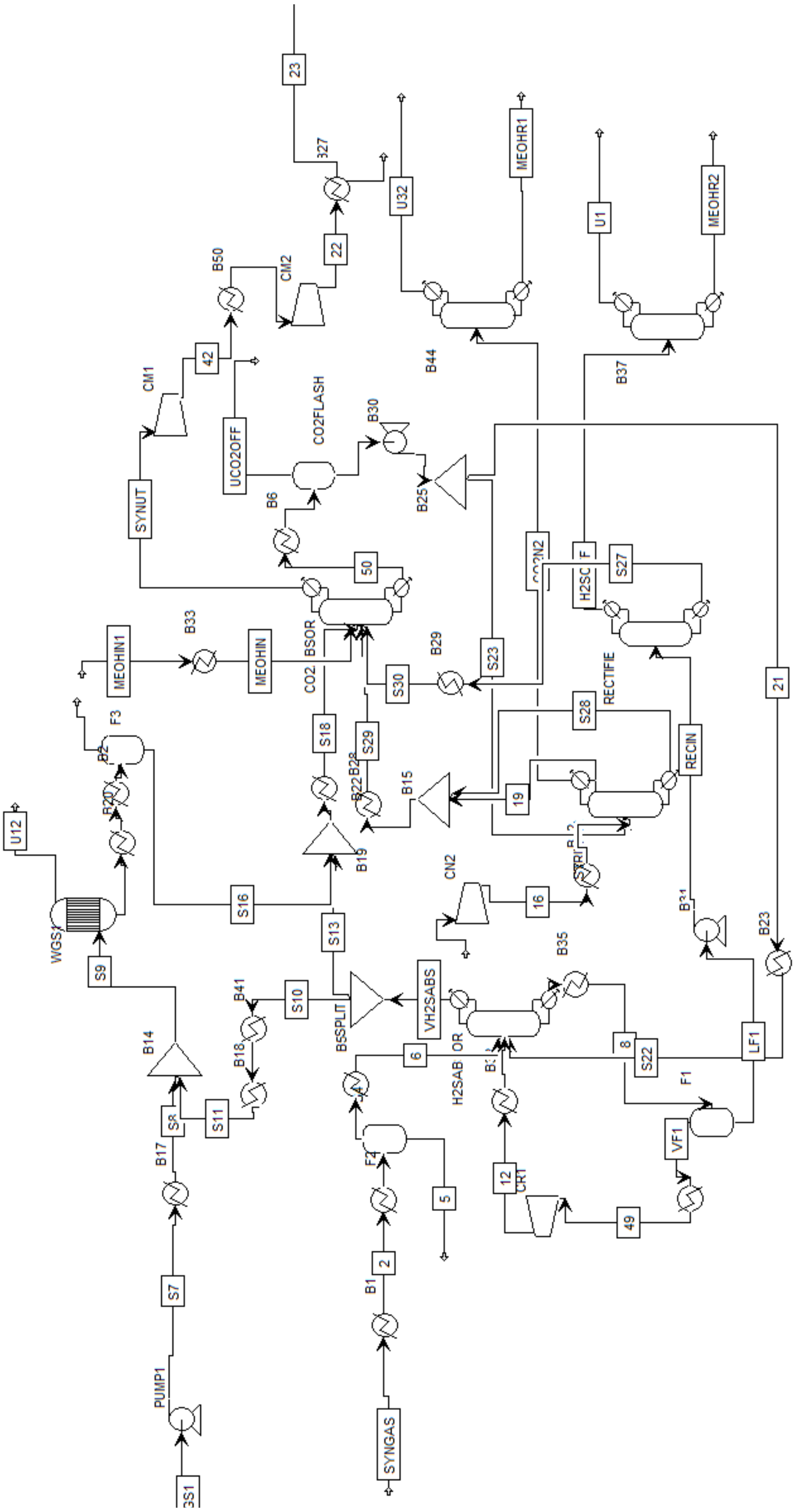
Case 2, process scheme part 1 of simulation



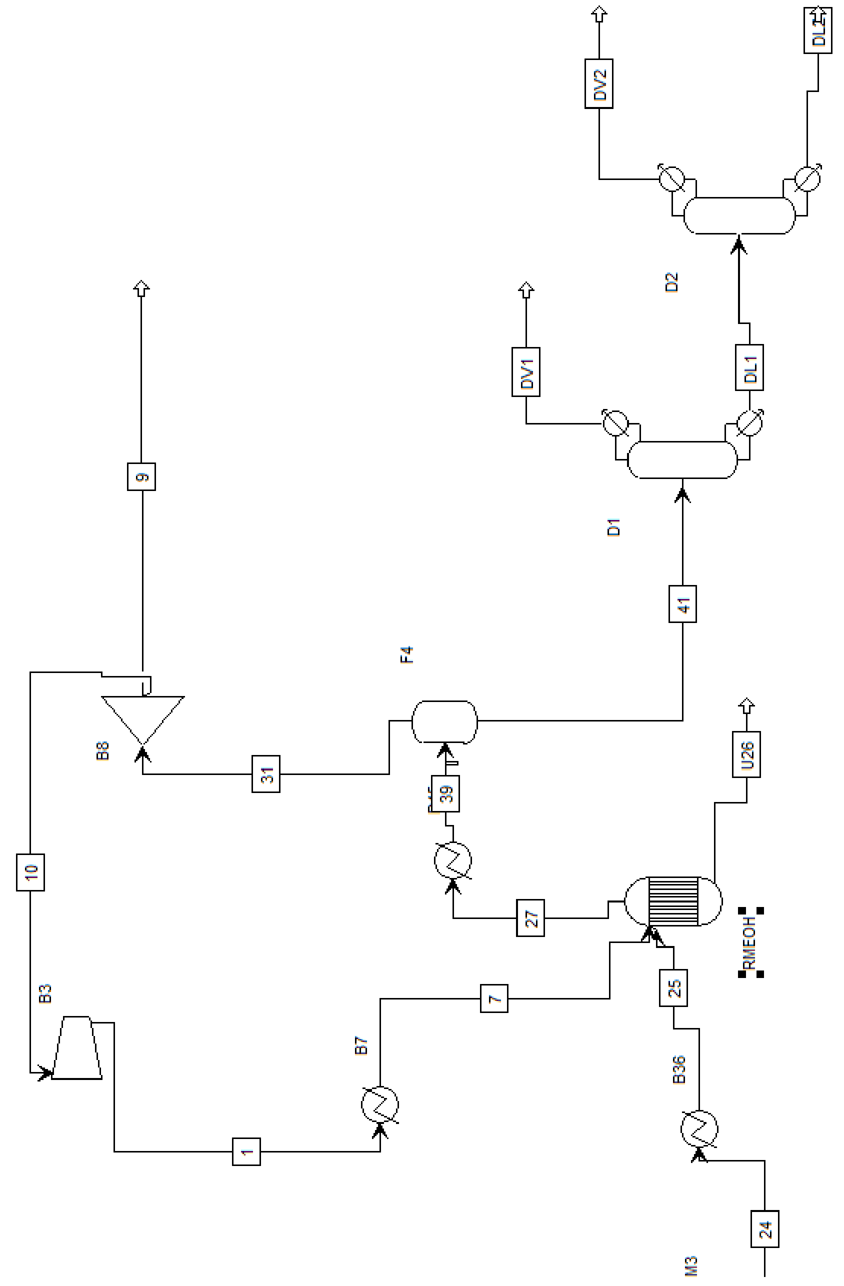
Case 2, process scheme part 2 of simulation



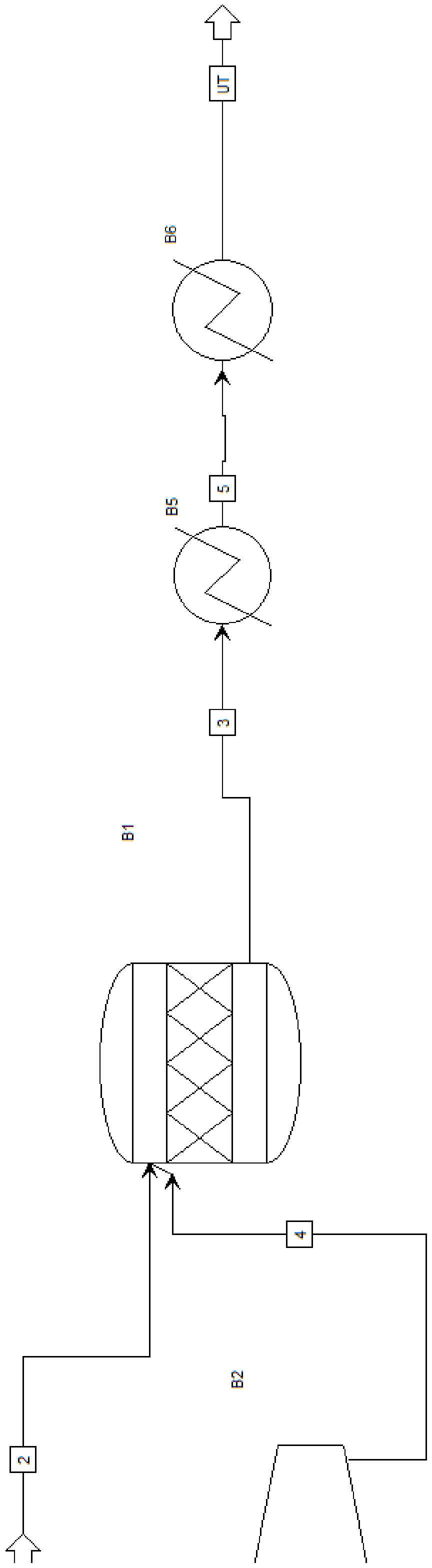
Case 3, process scheme part 1 over simulation



Case 3 process scheme part 2 over simulation



Case 3 process scheme part 3 over simulation



Appendix ii, Design specs

Design Specifications, DS-1 is the input for the WGS step and H2O is the input for the methanation step.

Design Specs		
Name	DS-1	H2O
Specification	**H2/CO**	**CO-H2O**
Specification target	**2,1**	**0**
Specification tolerance	**0,1**	**0,01**
Lower bound	**0**	**0**
Upper bound	**4500**	**3000**

Appendix iii, Target Loads retrieved from ASPEN Enalyzer

Case1

Utility	Inlet T, C	Outlet T,C	Target load(kJ/h)
Refrigerant 1	-25	-24	53644086
Refrigerant 2	-40	-39	4575690
Refrigerant 3	-66	-64	28067545
Cooling Water	25	20	1,92E+08
LP Steam Generation	*124*	*125*	*4918191*
MPGEN	*174*	*175*	*9352641*
HPGEN	*249*	*250*	*1,52E+08*

Case2

Utility	Tin	Tout	Target Load(kJ/h)
Refrigerant 1	-25	-15	43 147 407,14
Refrigerant 2	-40	-39	3872320,518
Refrigerant 3	-66	-64	28 732 144,18
Cooling Water	20	25	471 294 763,9
LP Steam	125	124	106 577 394,1
Fired Heat (1200)	1200	1000	10 704 033,01

Case 3

Utility	Tin	Tout	Target Load kj/h
Refrigerant 1	-25	-24	125 780 142,3
Refrigerant 2	-40	-39	4575689,901
Refrigerant 3	-66	-64	28 067 544,91
Cooling Water	30	35	315 773 112,8
LP Steam	125	124	65 634 263,84
MP Steam Generation	174	175	38 482 925,08
HP Steam Generation	249	250	338 321 045,4

Appendix iv, Gasifier data

		From EON	**Input process**
Fuel after dryer			
Lower heatingvalue	*MJ/kg*	16.601	
Mass flow	*kg/s*	2.9631	31.28124812
coal flow	*kmol/s*	0.11175	1.179737261
T	*Celsius*	180.1	
Nitrogen from fuel injection		0	0
Steam to Gasifier			
Mass flow	*kg/s*	1.3038	13.76412922
Temperature	*Celsius*	400	4222.773193
Oxygen to gasifier			
Mass flow	*kg/s*	0.51106	5.39522617
M	*kg/kmol*	32	337.8218554
Mole flow	*kmol/ s*	0.015970625	0.168600818
Temperature	*Celsius*	25	263.9233245
Nitrogen from O2 Prod			
Mass flow	*kg/s*		17.08488287
M	*kg/kmol*	28	28
Mole flow	*kmol/ s*		0.610174388
Char and ash from gasifier			
Mass flow	*kg/s*	0.10612	1.120301728
Temperature	*Celsius*	850	8973.393034

Char flow	*kmol/s*	0.00559	0.059013255
Product from gasifier			
Lower heating value(dry gas)	*MJ/nm3*	11402	120370.1499
Mass flow	*kg/s*	4.64744	49.06271262
Mol Flow(without ash)	*kmol/s*	0.2221	2.344694815
Coal Flow	*kmol/s*	0.11175	1.179737261
Temperature	*Celsius*	850	850
Pressure	*bar*	10	10
vol%			
CO		13.69	13.69
CO2		22.282	22.282
CH4		11.827	11.827
H2		24.916	24.916
N2		0.127	0.127
H2S		0.015	0.015
H2O		27.144	27.144
ash		0.0167	0.0167

The gasifier is assumed to have a efficiency of 0.8

Appendix v, Product streams

Case 1

Mole Flow kmol/hr	**Methanol**	**SNG. after PSA**
CO	0	
CO2	3.3103E-10	2.334
CH4	0	826.2
H2	0	1.3
N2	0	7.2
H2S	5.5789E-12	
H2O	0.5149658	
CH4O	502.735	

Case 2

Mole Flow kmol/hr	**Methanol**
CO	0
CO2	6.15938E-08
CH4	0
H2	0
N2	0
H2S	6.9682E-10
H2O	1.433406
CH4O	1693.567

Case 3

Mole Flow kmol/hr	Methanol
CO	0
CO2	1.04598E-09
CH4	0
H2	0
N2	0
H2S	4.0703E-11
H2O	3.896854
CH4O	690.1031

Appendix vi, Calculated efficiencies for the three cases

	Case 1	Case 2	Case 3
MW in	533.3	536.7	533.3
MWout	309.6	358	148.7
Efficiency	0.47	0.63	0.21
Carbon efficiency	0.17	0.40	0.16

Appendix vii, Cooling costs and electricty revenue

The electricity price is assumed to be 1 SEK/kWh
The price for 1 m^3 cooiling water is assumed to be 1.1 SEK

Case 1

Utility	**Target load(kJ/s)**	ΔH kJ/kg	$\dot{v}_{water}$. m^3/h	Cost MSEK/ year
Cooling Water	53333	*42*	*4580*	40.0
		Φ or ε	*Wel kJ/s*	
LP Steam	1366,	*0,0051*	*7*	*0,06*
MP Steam Generation	2598	*0,016*	*42*	*0,3*
HP Steam Generation	42222	*0,11*	*4644*	*37.0*

Case 2

Utility	Target load kJ/h	ε	Wel. kJ/s	MSEK/year
LP Steam	106 577 394.1	0.0051	151	1.2
Steam to dryer	187 652 652.1	0.0051	266	2.1
		ΔH kJ/kg	$\dot{v}_{water}$. m^3/h	
Cooling Water	471 294 763.9	42	11 242	96.8

The required need of fire heat is assumed to be supplied by biomass, with an efficiency off 0.8.
Target load is 10 704 033 kJ/h, biomass have a heating value of 16.60 MJ/kg, this gives an extra biomass requirement of 0,18kg/s

Case 3

Utility	Target load kJ/h	ΔH kJ/kg	$\dot{v}_{water}$. m^3/h	Cost MSEK/year
Cooling Water	315 773 112.8	42	7 532	64.8
		Φ or ε	Wel kJ/s	MSEK

LP Steam	-65 634 263.84	0,0051	-92,9	-0,7
MP Steam Generation	38 482 925.08	0,056	598,6	4,8
HP Steam Generation	338 321 045.4	0,11	10337,6	82,7
Steam to dryer	-188 732 451.6	0,0051	267,4	2,1

Appendix viii, Electricity requirements for refrigeration

Electricity requirements for refrigeration Case 1

Refrigerating utilities	**H1 kJ/kg**	**H2 kJ/kg**	**H3 kJ/kg**	**Wel kj/kg**	**Qref kJ/kg**	**Target load KJ/s**	$\dot{m}_{ref}$ **Kg/s**	**W_{el} Kj/s**	**MSEK**
Ref 1	394	405	184	11	210	**14901.14**	70.96	780.54	
Ref 2	389	397	153	8	236	**1271.02**	5.39	43.09	
Ref 3	375	390	138	15	237	**7796.54**	118428.5	493.45	
TOT								1317.07	10.54

Electricity requirements for refrigeration Case 2

Refrigerating utilities	**H1 kJ/kg**	**H2 kJ/kg**	**H3 kJ/kg**	**Wel kj/kg**	**Qref kJ/kg**	**Target load KJ/s**	$\dot{m}_{ref}$ **Kg/s**	**W_{el} kJ/s**	**MSEK**
Ref 1	394	405	184	11	210	**11985.40**	255448	627.81	
Ref 2	389	397	153	8	236	**1075.64**	19 388.52	36.46	
Ref 3	375	390	138	15	237	**7 981.15**	118 428.5	505.14	
TOT									9.36

Eletricity requirements for refrigeration Case 3

Refrigerating utilities	**H1 kj/kg**	**H2 kj/kg**	**H3 kj/kg**	**Wel kj/kg**	**Qref kJ/kg**	**Target load kJ/s**	$\dot{m}_{ref}$ **Kg/s**	**KJ/s**	**MSEK**
Ref 1	394	405	184	11	210	34938.91	166.37	1830.13	

Ref 2	389	397	153	8	236	1271.02	5.39	43.09	
Ref 3	375	390	138	15	237	7 796.54	32.90	493.45	
TOT									18.93

Appendix ix Electricity consumptions compressors and pumps

Case 1

Compressors	**kJ/s**	**MWh**	**MSEK**
CM1	2668.17	21345.4	21.34539816
CM2	2627.05	21016.43	21.01642632
CM3	1891.59	15132.72	15.1327172
CN2	1806.09	14448.73	14.4487284
CR1	2.20	17.59517	0.017595173
TOT	**8995.11**	**71960.87**	**71.96**
Pump			
B10	0.24843693	1.987495	0.001987495
B30	122.91184	983.2947	0.98329472
B31	33.6433827	269.1471	0.269147062
Pump1	2.05684035	16.45472	0.016454723
TOT	**158.86**	**1270.88**	**1.27**

Case 2

Compressors	**kJ/s**	**Mwh**	**MSEK**
CM1	4297.71	34381.68	34.38
CM2	4908.83	39270.64	39.27
CM3	2991.4	23931.2	23.93
CN2	1806.09105	14448.73	14.45
CREMEOH	158.4	1267.2	1.27
CH2SABS	0.0075	0.06	0.00006
TOT	**14162.44**	**113299.5**	**113.30**
PUMPS			
B9	0.015	0.12	0.00012
B16	6.97	55.76	0.056
B30	142.57	1140.56	1.14
B31	25.76	206.08	0.21
B46	3.77	30.16	0.03
TOT	**179.09**	**1432.68**	**1.37**

Case 3

Compressors	**kJ/s**	**MWh**	**MSEK**
B3	4140.84	33126.72	33.1
CM1	2847.60	22780.8	22.8
CM2	2627.05	21016.4	21.0
CM3	1891.59	15132.72	15.1
CN2	1806.09	14448.72	14.4
CR1	2.20	17.6	0.01
TOT	**15075.8**	**120606.3**	**106,4**
Pumps			
B30	123	984	9.9
B31	33.64	269.12	0.3
Pump1	2.06	0.01648	0.02
TOT	**158,7**	**1257,24**	**10,21**

Appendix x, Variable costs for units not included in simulation

UNIT	Electrical demand	Requirements	total	Ref
ASU	0.3 kW/Nm^3	0.73Nm^3/s	0.22kW	(34)
PSA(Case2)	0.3kwh/ Nm^3	0.081	0.024kwh	(35)

	ASU. MSEK	PSA
Case1	0.0176	0.000024
Case 2	0.0176	
Case 3	0.0176	

OLGA

Given from literature are that the total cost of utilities are 1.1 €/tonne biomass, electricity consumption making up for 26%, steam consumption 66% and cooling water consumption 8%.

The calculation was built on a electricity price of 0.07 €/kWh$_e$, heat price of 4 €/ GJ and 0.1 €/m^3 cooling water, with prices in 2005 years currency The price for the scrubbing liquid is 9€/ tonne biomass (33) This information was used to calculate the costs for the cases

CASE	Biomass input to process tonne/s	Utility Cost SEK/tonne	Cepci kvot 2019/2005	Utility cost MSEK
Case 1	0.03128	108.7	1.288	126.1
Case 2	0.03128	108.7	1.288	126.1
Case 3	0.03128	108.7	1.288	126.1

Appendix xi, Steam demand gasifier

There is a demand of steam to the gasifier. it is assumed that energy for this can be come from extra biomass input.
The steam energy input requirments calculated;

$\dot{m}_{steam}$ =5.4 kg/s
$T_{water_0} = 15°C$
$Cp_{water} = \frac{4.2kJ}{kg * C}$
$T_{steam_1} = 179.88°C$
$P_1 = 10\ bar$
$H_1 = 2\ 777.9$
$P_0 = 1\ bar = P_2$
$T_2 = 100°c$
$H_2 = 2\ 675.25$

$$Q_1 = \left(\left(T_2 - T_{water_0}\right) * Cp + H_{vap} + (H_1 - H_2)\right) * \dot{m}_{steam}$$

$$Q_1 = \left((100 - 15) * 4.2 + 2257.71 + (2777.9 - 2675.25)\right) * 5.4$$

$$= 2717.36 \frac{kJ}{kg} * 5.4 = 14673.7kJ$$

This energy demand is assumed to be met by an increase of biomass this . translates to approx. 1 kg biomass. with an efficiency of 0.8 and a heating value of biomass at 16601kJ/kg

$$\dot{m}_{Biomass} = \frac{\frac{14\ 673{,}7}{0.8}}{16\ 601} = 0.88kg$$

Appendix xii, Energy Consumption Dryer

From the information given by the gasifier output data supplied by E.ON Gas Sverige AB,
Appendix iv, Gasifier data,
$\mathrm{LHV}_{\text{biomass out from dryer}}$=16.601 MJ/kg,

From this moisture content, after dryer, is calculated:

$x_{predryer} = 45\%$

$\dot{m}_{biomassa_{afterdryer}} = 21.69\frac{kg}{s}$

$\dot{m}_{watercontent_{afterdryer}} = 2.51\frac{kg}{s}$

$x_{afterdryer} = 11.57\%$

$\dot{m}_{biomass_{nowater}} = 19.17\frac{kg}{s}$

$\dot{m}_{biomass_{45\%}} = 34.86\frac{kg}{s}$

$\dot{m}_{removed_{water}} = 13.17\frac{kg}{s}$

$T_0 = 10°C$

$T_1 = 100°C$

$\Delta T = 90°C$

$\Delta H_{vap} = 2\ 260\ kJ/kg$

$Cp_{vatten} = 4{,}2\frac{kJ}{kg} * C$

$\dot{Q}_{heat} = 44\ 562\ kJ/s$

Verkningsgraden antas vara 0,8, detta ger

$\dot{Q}_{heat_{0,8}} = 52\ 426\ kJ/s$

Appendix xiii, Electric consumption for heat generation.

	H_1 kJ/kg	H_2 kJ/kg	H_3 kJ/kg	$Q\dot{o}u_{t}$ kJ/kg	kJel/kg	ε
MPsteam, T=175-150	2 747	2 780	741,1	2 005	33	0.016
LP Steam T=125-110	2 691	2 680	525	2 166	11	0.0051
HP Steam T=250-225	2 803	2 850	1 085	1 718	47	0.027

Appendix xiv, electricity production from generated steam.

Electricity production from Steam

	H_1 kJ/kg	H_2 kJ/kg	H_3 kJ/kg	Qin(kJ/s)	W(kJ/s)	Φ
MPsteam	2 690	2 825	440	2 385	135	0.056
HP steam	2 740	2 950	1 086	1 864	210	0.11

LP steam is assumed to be used as district heating, with efficiency in the heat exchanger at 100%

Appendix xv, Heat exchanger network area

Case	Area m^3
Case1	41 280
Case2	160 960
Case 3	
Case3, part 1,2	216 300
Case 3, part 3	218 100
TOT Case 3	434 400