



# **Process Simulation in Aspen Plus of an Integrated Ethanol and CHP plant**

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## Abstract

The purpose of this master thesis project was to construct a process model and to evaluate the impact of integrating an ethanol plant with a co-generation plant for heat and electric power production. The simulation program Aspen Plus was used to construct the process model and simulations were performed to investigate different conditions for an integrated plant. An economic analysis was done as well. The input data for the ethanol production are based on simulations performed at Lund University. To construct the model for the combined heat and power (CHP) plant, data from a real CHP plant integrated with a steam dryer were used.

The ethanol plant in the process model is based on lignocellulosic raw material, using the enzymatic hydrolysis process. In the ethanol plant it is necessary to add steam in the pretreatment, distillation and evaporation process steps, which is one of the reasons to integrate the ethanol plant with a CHP plant. A solid lignin co-product is produced in the ethanol process, which can be used as fuel after drying. For this purpose a steam dryer was used in the model as well.

The process model for the integrated plant consisted of pretreatment, distillation, evaporation and steam drying representing the ethanol plant, and for the CHP plant a high-pressure turbine, low-pressure turbines, steam reformer, feed-water tanks and feed-water preheaters. A separate process model was constructed for the steam boiler, featuring a combustion chamber, an economiser, an evaporator, a superheater and an air preheater. As the integrated plant was compared with a CHP plant without extractions to an ethanol plant and steam dryer, a model was constructed for the stand-alone CHP plant as well.

Several simulations were performed varying the total heat demand for the ethanol plant between 37.2-71.6 MW, depending on the plant capacity and the efficiency of the different process steps. The secondary steam from the pretreatment was reused within the ethanol plant and the secondary steam from the dryer was reused in a steam reformer to produce more steam for a low-pressure turbine. For one simulation the secondary steam from the dryer was reused in the ethanol plant. The district heating demand was varied over the year, using heat either from the CHP plant or waste heat from the ethanol plant. In one of the simulations wood fuel and lignin fuel were mixed in the steam boiler.

The results from the simulations, along with an economic evaluation, indicate that integrating an ethanol plant and a CHP plant is profitable for both plants. The ethanol production process should be as effective as possible and both the secondary steams from pretreatment and steam drying should be used in the ethanol plant, to keep the productions costs down. The waste heat, if any, from the ethanol plant should be used for district heating, even though the produced electricity is reduced. It is preferable to sell the lignin fuel than to use it in the steam boiler of the CHP plant.

Regarding the process models, there are improvements to be made. The steam-boiler should be connected to the steam-cycle and the entire CHP plant should be integrated with an entire ethanol plant. The interconnecting streams in the process models could be optimised as well. But the process models constructed are good enough to be used for simulations of integrated plants with other conditions.

### Keywords:

ethanol production, combined heat and power production, integration of ethanol and CHP plant, process simulation, economic evaluation

## Sammanfattning

Syftet med examensarbetet var att konstruera en processmodell för att utvärdera effekten av att integrera en etanolanläggning med en kraftvärmeanläggning. Simuleringsprogrammet Aspen Plus användes för att konstruera processmodellen och simuleringar genomfördes för att utreda olika förutsättningar för en integrerad anläggning. Även en ekonomisk analys av de olika fallen genomfördes. Effektbehov, ångtryck mm för etanolanläggningen är baserade på simuleringar som genomförts vid Lunds universitet. För att kunna konstruera modellen för kraftvärmeverket användes grunddata från en verklig anläggning integrerad med en ångtork.

Etanolanläggningen i processmodellen använder lignocellulosabaserat råmaterial och den enzymatiska hydrolysisprocessen. Vid etanolframställningen är det nödvändigt att tillsätta ånga i förbehandling, destillation samt indunstning, vilken är ett av motiven till att integrera etanolanläggning med ett kraftvärmeverk. Från etanolprocessen erhålls en fast ligninåterstod som kan användas som bränsle efter torkning. För detta ändamål ingick även en ångtork i modellen.

Processmodellen för den integrerade anläggningen består av förbehandling, destillation, indunstning och ångtorkning för etanoldelen och för kraftvärmedelen en högtrycksturbin, lågtrycksturbiner, ångomvandlare, matarvattentankar samt matarvattenförvärmare. För ångpannan konstruerades en separat processmodell med en förbränningskammare, ekonomiser, förångare, överhettare och luftförvärmare. Eftersom den integrerade anläggningen har jämförts med ett kraftvärmeverk utan koppling till en etanolanläggning och ångtork, konstruerades ytterligare en modell för ett kraftvärmeverk med enbart el- och fjärrvärmeproduktion.

Ett antal simuleringar genomfördes, där effektbehovet för etanolanläggningen varierades mellan 37.2-71.6 MW, beroende på etanolanläggningens storlek samt effektiviteten för de olika processtegen. Sekundärångan från förbehandlingen i etanolprocessen återanvändes inom etanolanläggningen och sekundärångan från ångtorken användes i en ångomformare för att producera mer ånga till en lågtrycksturbin. En simulering genomfördes där sekundärångan från torken återanvändes inom etanolanläggningen. Fjärrvärmebehovet varierades över året, där antingen värme från kraftvärmeverket eller spillvärme från etanolanläggningen användes. En simulering genomfördes där träbränslet blandades med ligninbränsle.

Resultaten från simuleringarna, med en kompletterande ekonomisk analys, indikerar att en integrerad anläggning är lönsam både för etanolanläggningen och för kraftvärmeverket. Produktionsprocessen för etanol bör optimeras i så hög utsträckning som möjligt samt sekundärångan från förbehandlingen och ångtorken bör utnyttjas inom etanolanläggningen för att minimera produktionskostnaden. Det är att rekommendera att eventuell spillvärme från etanolanläggningen används för fjärrvärme, även om producerad elektricitet reduceras något. Det är att föredra att sälja ligninbränslet framför att använda det i kraftvärmeverkets ångpanna.

De konstruerade processmodellerna fungerar väl, men kan förbättras. Ångpannan bör kopplas ihop med ångturbincykeln och hela kraftvärmeanläggningen integreras med en fullständig etanolanläggning. Även de interna strömmarna mellan olika enhetsoperationer bör optimeras. Modellerna är dock tillräckligt bra för att kunna användas för integrerade anläggningar med andra förutsättningar.

### Nyckelord:

etanolproduktion, kraftvärmeproduktion, integrering av etanol- och kraftvärmeanläggning, processimulering, ekonomisk utvärdering

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# 1. Introduction

## 1.1 Background

The global climate change, which is mainly due to greenhouse gas emissions, has increasingly become a critical energy and environmental policy issue. The transport sector is a major contributor to these emissions, and by displacing gasoline and diesel with bio-fuels and making the transport system more effective, it is possible to reach a sustainable transport system. Bio-fuels have a large potential of reducing CO<sub>2</sub> emissions throughout the fuel cycle, as the vehicle combustion of bio-fuels does not contribute to net emissions of CO<sub>2</sub>. An increased use of bio-fuels reduces the oil import as well.

Ethanol is one of the alternatives to gasoline and diesel. Today ethanol is produced commercially from sugar-based materials (e.g. sugar beet and sugar cane) as well as from starch-based materials (e.g. wheat and corn). Lignocellulosic materials, such as woody biomass, herbaceous crops (e.g. switch grass) and agricultural residues (e.g. corn stover and wheat straw) are also potential raw materials, but the processes for producing ethanol from these materials are still under development. Compared to conventional gasoline, the net reduction of greenhouse gas emissions from the production and use of lignocellulosic ethanol is estimated to 70-90 % (1). By combining enzymatic hydrolysis of cellulose to produce ethanol and using biomass as process fuel makes it possible to reduce greenhouse gas emissions (CO<sub>2</sub>-equivalent) to near zero. If the ethanol production is integrated with a combined heat and power (CHP) plant and the co-generated electricity is used to displace coal-fired power, the emissions of greenhouse gases can be reduced further.

Etek Etanolteknik AB has, at the request of the Swedish Energy Agency, established a pilot plant in Örnsköldsvik in order to develop the ethanol processes based on woody biomass. The plant is unique and the first continuously operating plant in Sweden using wood as raw material. However, there will be no ethanol production for commercial purposes; the capacity lays around 500 litres ethanol per day, and the pilot will be operated continuously for periods of around 2-3 weeks. There are essentially three different hydrolysis processes for converting lignocellulosic material to sugar; concentrated and dilute-acid hydrolysis and enzymatic hydrolysis. The method used in Örnsköldsvik is the dilute-acid process, but in the near future the enzymatic hydrolysis will be studied as well.

It is possible to integrate an ethanol production plant with other operations, such as combined heat and power (CHP) production, sawmill or pulp industries, in a so called bio-energy combination plant. This thesis considers the synergy effects of combining the ethanol production and a lignin-fuel dryer with a co-generative plant for heat and power production, using biomass as fuel. The ethanol production can be used to increase the power output from the CHP plant during the warmer half of the year as well as extend the working season, by drawing off some of the steam produced in the CHP plant.

## 1.2 Aim and Methodology of the Thesis

This master thesis project has been performed in collaboration with Etek in Örnsköldsvik and under the auspices of the Master of Science Program in Energy Technology at the Department of Applied Physics and Electronics, Umeå University. The main purpose of the project was to construct a process model in Aspen Plus and to perform simulations of an ethanol production plant, based on lignocellulosic material, combined with a CHP plant. The model was supposed to give indications for the optimal management and reliable data for the electrical power output for this combination.

The input data for the ethanol production are based on simulations performed in Aspen Plus at Lund University (2). To construct the model for the CHP plant, data from a real CHP plant integrated with a steam dryer were used. Three process models were constructed, one for the steam boiler, one for the steam-cycle with extractions to an ethanol plant and one stand-alone steam-cycle.

### ***1.3 Outline of the Thesis***

The thesis begins with a presentation of the ethanol production process (Chapter 2), and of the working process for a steam dryer and steam reformer (Chapter 3). The technologies for combined heat and power production are described in Chapter 4. As mentioned earlier the simulation program Aspen Plus has been used to construct a model of a combined ethanol and CHP plant and this program is presented in Chapter 5. Chapter 6 gives an overview of the process models constructed. The results received are presented and discussed in Chapter 7 and the conclusions from this study are given in Chapter 8.

## 2. Raw Material for Ethanol Production

Ethanol can be produced using sugar-based materials, starch-based materials or lignocellulosic materials. In Sweden there is one factory (Agroetanol in Norrköping) producing ethanol from wheat, and at Domsjö Fabriker in Örnsköldsvik, lignocellulosic-based ethanol is produced as co-product. The technologies for processing lignocellulose to ethanol are however still under development. Typical lignocellulosic materials are wood chips, sawdust, energy wood, straw, reed canary grass, or municipal solid waste free from plastics. The lignocellulosic materials consist, up to 90-98 %, of the three polymeric materials cellulose, hemicellulose and lignin. Producing ethanol from biomass involves hydrolysis of cellulose and hemicellulose, followed by fermentation of simple sugar molecules to ethanol, after which the ethanol then is separated from the water by distillation. In Table 1 the typical compositions of some lignocellulosic materials are presented.

*Table 1: Composition of lignocellulosic materials of dry material (2)*

	<b>Glu</b>	<b>Xyl</b>	<b>Man</b>	<b>Gal</b>	<b>Ara</b>	<b>Lig</b>
<b>Softwoods</b>						
Radiata pine	42.8	5.9	11.3	2.5	1.6	27.2
Spruce wood	41.9	6.1	14.3	-	1.2	27.1
Spruce bark	27.3	3.2	3.2	-	4.1	37.1
Douglas fir wood	46.1	3.9	14.0	2.7	1.1	27.3
Douglas fir bark	14.7	1.8	2.5	1.6	2.8	38.8
<b>Hardwoods</b>						
Salix	41.5	15.0	3.0	2.1	1.8	25.2
<b>Other</b>						
Sugar cane bagasse	40.2	22.5	0.5	1.4	2.0	25.2
Corn stover	36.1	21.4	1.8	2.5	3.5	17.2

Glu = Glucan, Xyl = Xylan, Man = Mannan, Gal = Galactan, Ara = Arabinan, Lig = Lignin, Ext = Extractives  
-, below detection limit

### 2.1 Composition of Wood

#### 2.1.1 Cellulose

Cellulose is the main component of wood and comprises 35-50 % of dry wood by mass (3). The linear polymer is built of D-glucose linked together by  $\beta$ -1,4-glucosidic bonds. Part of the cellulose molecules are organised in crystalline form, with multiple hydrogen bonds between the chains. There are regions of lower crystallinity, i.e. amorphous regions. These regions are particularly susceptible to enzymatic hydrolysis. Due to the hydrogen bonds, cellulose is very hard to dissolve in common solvents.

#### 2.1.2 Hemicellulose

Hemicellulose, which comprises 20-30 % of dry wood by mass, is a branched polymer of different monomeric sugars, such as the hexoses D-glucose, D-mannose and D-galactose and the pentoses D-xylose and L-arabinose (3). Acetyl groups or uronic acid groups are attached to the main chain or branches. The polymer links covalently to lignin, and through hydrogen bonds

to cellulose. As it lacks the crystallinity, it does not contribute substantially to the structural properties of wood.

### **2.1.3 Lignin**

Lignin comprises 15-30 % of dry wood by mass and is a highly complex, three-dimensional aromatic polymer of different phenyl-propane units, bonded together by ether and carbon-carbon bonds. The lignin contributes to the mechanical strength of wood and there are few micro-organisms, which can decompose lignin, compared to cellulose and hemicellulose.

### **2.1.4 Extractives and Ash**

Extractives and ash are also components in wood, but are usually very low, below 5 % and around 1 % respectively. The three main groups of softwood extractives are terpenes, fats and waxes but there are also other compounds present such as n-alkanes. The inorganic compounds that remain after burning the organic matter are called ash.

### 3. General Description of Ethanol Production

Ethanol production, using wood as raw material, consists of several process steps, which are shown in Figure 1. From 100 kg dry wood it is possible to get 20-30 kg ethanol, 20-30 kg carbon dioxide and 30-45 kg lignin pellets with a dry-matter content of 85-90 %. The variations depend on the raw material and process. The carbon dioxide can be used to produce carbonic acid and the lignin pellets can be used as a fuel.

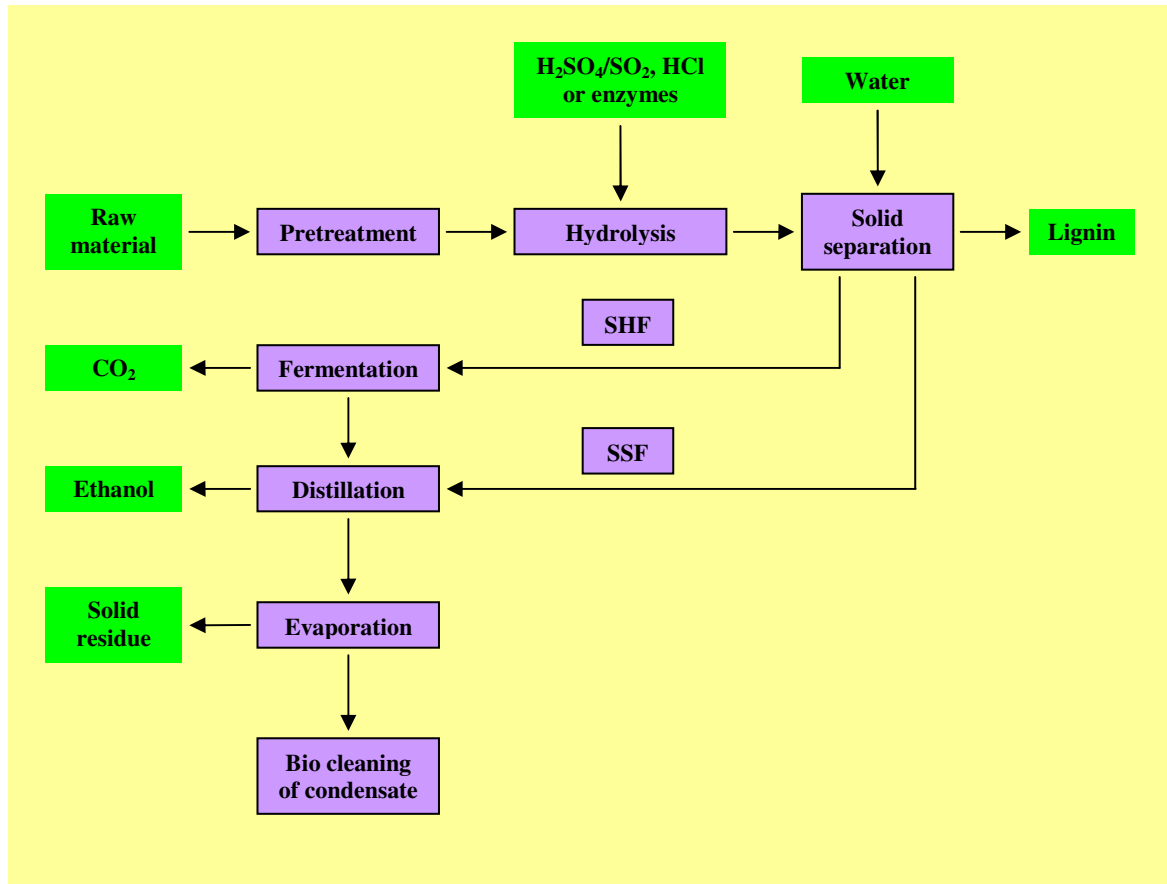


Figure 1: Schematic flow diagram for an ethanol production plant (2, 4)

#### 3.1 Pretreatment

In the enzymatic hydrolysis process it is necessary to pretreat the material to make the cellulose accessible to the enzymes. In an efficient pretreatment method, both the material loss and co-product formation are minimal, and a high reaction rate with high sugar yields is achieved. The pretreatment techniques can be divided into four different categories:

- Mechanical methods – the material is mechanically fragmented by milling or grinding.
- Mechanical-chemical methods – a combination of mechanical forces and chemical reactions such as heat treatment, with high-pressure water or steam.
- Chemical methods – acids, bases or solvents are added to swell the material or dissolve the hemicellulose and lignin. Dilute-acid pre-hydrolysis is a common example.
- Biological methods – lignin is broken down with enzymes or micro-organisms.

## **3.2 Hydrolysis**

There are three different kinds of hydrolysis processes that can be used to convert lignocellulosic materials to ethanol; concentrated-acid hydrolysis, dilute-acid hydrolysis, and enzymatic hydrolysis. Hydrolysis means splitting (lysis) of the  $\beta$ -1,4-glucosidic bonds with addition of water (hydro) into two molecular fragments. One fragment takes up a proton, while the other accepts a hydroxyl group.

### **3.2.1 The Acid Hydrolysis Process**

The acid hydrolysis process can be carried out either with concentrated or diluted acids, such as sulphurous, sulphuric, hydrochloric, hydrofluoric, phosphoric, nitric or formic acid. By using concentrated acids, such as sulphuric ( $H_2SO_4$ ) and hydrochloric (HCl), it is possible to operate the processes at low temperature and high yields are achieved, 90 % of theoretical glucose yield (5). The major drawbacks are that the acids cause equipment corrosion and energy demanding acid recovery.

Both in the concentrated-acid and dilute-acid hydrolysis processes the cellulose and hemicellulose are degraded to monomeric sugars. These sugars can be converted to ethanol by yeast and bacteria. In the dilute-acid hydrolysis the acid consumption is relatively low, but high temperatures are required to convert cellulose to glucose. High temperatures increase the rate of hemicellulose sugar decomposition, which can cause inhibition in the fermentation stage, and equipment corrosion. To decrease sugar degradation, a two-stage process has been developed, where the hemicellulose is broken down in the first step and the cellulose in the second step. To dissolve the hemicellulose and cellulose sulphuric acid and sulphur dioxide ( $SO_2$ ) are commonly used catalysts.

### **3.2.2 The Enzymatic Hydrolysis Process**

In the enzymatic hydrolysis cellulolytic enzymes (cellulases) are used to break down the cellulose to sugars. Hydrolysis and fermentation can be run separately (SHF = Separate Hydrolysis and Fermentation) or simultaneously (SSF = Simultaneous Saccharification and Fermentation) after the pretreatment. The advantage of SHF is that each step can be run under optimal conditions. The released sugar, however, inhibits the enzymes during the hydrolysis. In SSF the sugar will not inhibit the enzymes as it is immediately fermented. The produced ethanol will act as an inhibitor, but not as strongly as glucose. Another drawback of SSF is the difficulty of recycling the yeast since it will be mixed with lignin.

## **3.3 Solid Separation**

The sugar solution is separated from the solid material, which consists mostly of lignin but also of cellulose and sugar. The lignin, which has a dry content of about 30 %, can be combusted directly in an industrial boiler. It is also possible to treat the material further by drying and pelletisation.

## **3.4 Fermentation**

After hydrolysis the sugar solution is pumped to temperature controlled ( $35^\circ C$ ) reactor tanks, where the sugar is fermented to ethanol with the help of commercial bakers' yeast. The process water from the hydrolysis may have a too low pH for the yeast bacteria, and by adding NaOH and sulphuric acid the pH is adjusted to about 4.5. To be able to ferment the sugar of hardwood and agricultural residues it is necessary to use genetically modified yeast, as they consist of pentoses. Nutrients and antifoaming agents are also added in the fermentation process. The fermentation is performed in agitated tanks and produces equal amounts of ethanol

and carbon dioxide. The carbon dioxide may be cleaned for use in soft drinks, fire extinguishers etc.

### **3.5 Distillation**

The ethanol concentration in the stream from the fermentation is low, 2-5 %, which is why distillation is necessary to reach a concentration of about 95 %. The ethanol solution is heated in a distillation column, where the ethanol is steamed and rises, and the water runs downwards. At the top, the steamed ethanol passes a cooler and condenses. The cleaned ethanol is stored for further transportation.

### **3.6 Evaporation**

The stillage contains some solid materials as well as volatile organic compounds. The stream is concentrated by evaporation, where the solid part can be directly combusted in an industrial boiler. The evaporated stream, which contains the volatile compounds, is cooled and condensed. The volatile organic compounds consume oxygen when they are broken down, which is why cleaning is necessary before letting out the stream into a recipient. This is done by a so-called bio cleaning; the organic compounds may be used to produce biogas through anaerobic fermentation.

### **3.7 Steam Drying and Steam Reforming**

Drying means that liquid is removed from a material by evaporation. To be able to evaporate the liquid a high-temperature gas is passed over the material, and in general the liquid that should be evaporated is water and the gas often used is air. However there exist various kinds of drying techniques, and in this chapter the principles of backpressure drying or drying in superheated steam is described.

The wet fuel passes through one or more heat exchangers, where the water in the fuel is heated and superheated by superheated steam of 10 bar or more (6). Due to the superheating, a heat transfer to the wet fuel occurs, the water in the fuel evaporates, and the produced fuel steam is separated from the fuel.

The pressure of the fuel and the produced fuel steam vary from 2 to 5 bar (6). As the energy content of the fuel steam is quite high, it is possible to recover the heat in a steam reformer. A steam reformer is usually a heat exchanger, with clean water on one side (often the condensate of the high-pressure steam used in the steam dryer) and the contaminated steam separated from the wet fuel on the other side. The clean condensate is converted to steam again, with a lower pressure though than the incoming steam to the dryer, and can be used in for example a low-pressure turbine.

## 4. Combined Heat and Power Production

Electricity from fuels such as natural gas, oil, diesel fuel, propane, coal, wood, wood-waste and biomass can be produced in either conventional power generation plants or in CHP plants. In a conventional power plant the efficiency is low; being around 35-40 % efficient (7) and the waste heat is rejected to rivers, lakes, oceans or the atmosphere. Co-generation is the simultaneous generation of heat and power, where the waste heat is used for process heating in industries or for district heating. Through the utilisation of the heat, the efficiency of a CHP plant can reach about 70-90 % (7) or even more.

The co-generation technology has been used since the beginning of the 20<sup>th</sup> century, and systems can operate for at least 20 years. The main components of a CHP plant based on steam power are:

- A steam boiler
- A prime mover (engine)
- An electricity generator
- A heat recovery system
- A control system

### 4.1 Different Methods for Power Generation

A steam-turbine (Rankine) cycle, a gas-turbine (Brayton) cycle or a combined cycle can be used as the power cycle in a CHP plant. The Rankine cycle, see Figure 2, is the most common technology to generate electricity from fuels. The fuel boils water to steam and the steam continues to a turbine where it expands. The turbine is connected to a generator, which produces electricity. In a CHP plant the waste heat is used to heat the district heating water. The condensed water is then recirculated to the steam boiler as feed water.

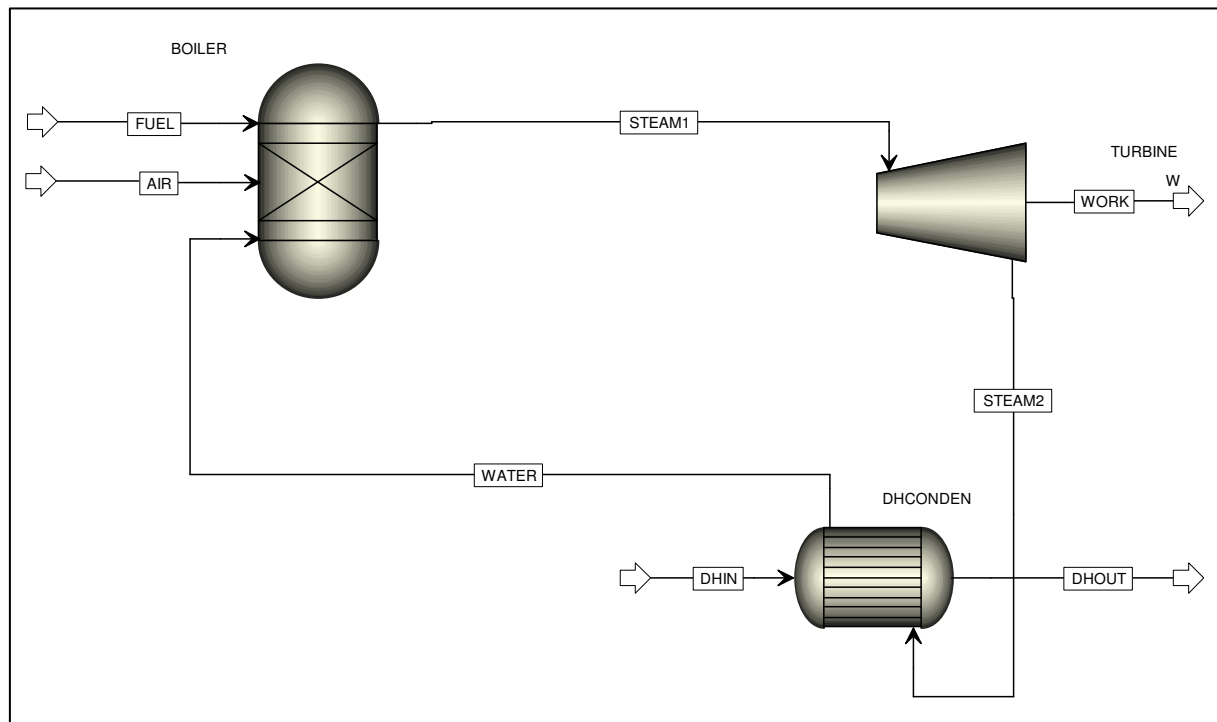


Figure 2: An ideal steam-turbine CHP plant (7, 8)

In the Brayton cycle, see Figure 3, the turbine is driven by the expansion of the gas during combustion. The pressure of the air needed for the combustion is increased in a compressor. The combustion chamber and gas turbine are flowed through by a mix of air and burning gases. As in the case of a steam-turbine cycle, the exhaust gases can be used to heat the district heating water. Gas turbines are very fast to start and stop, why they often are used for top-loads.

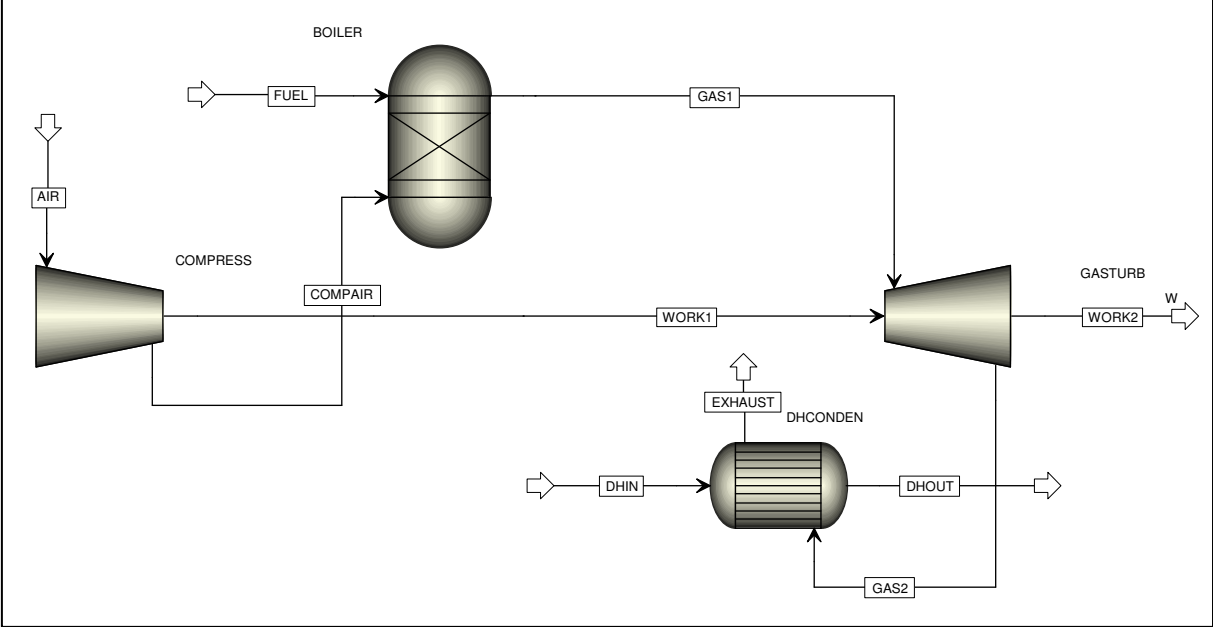
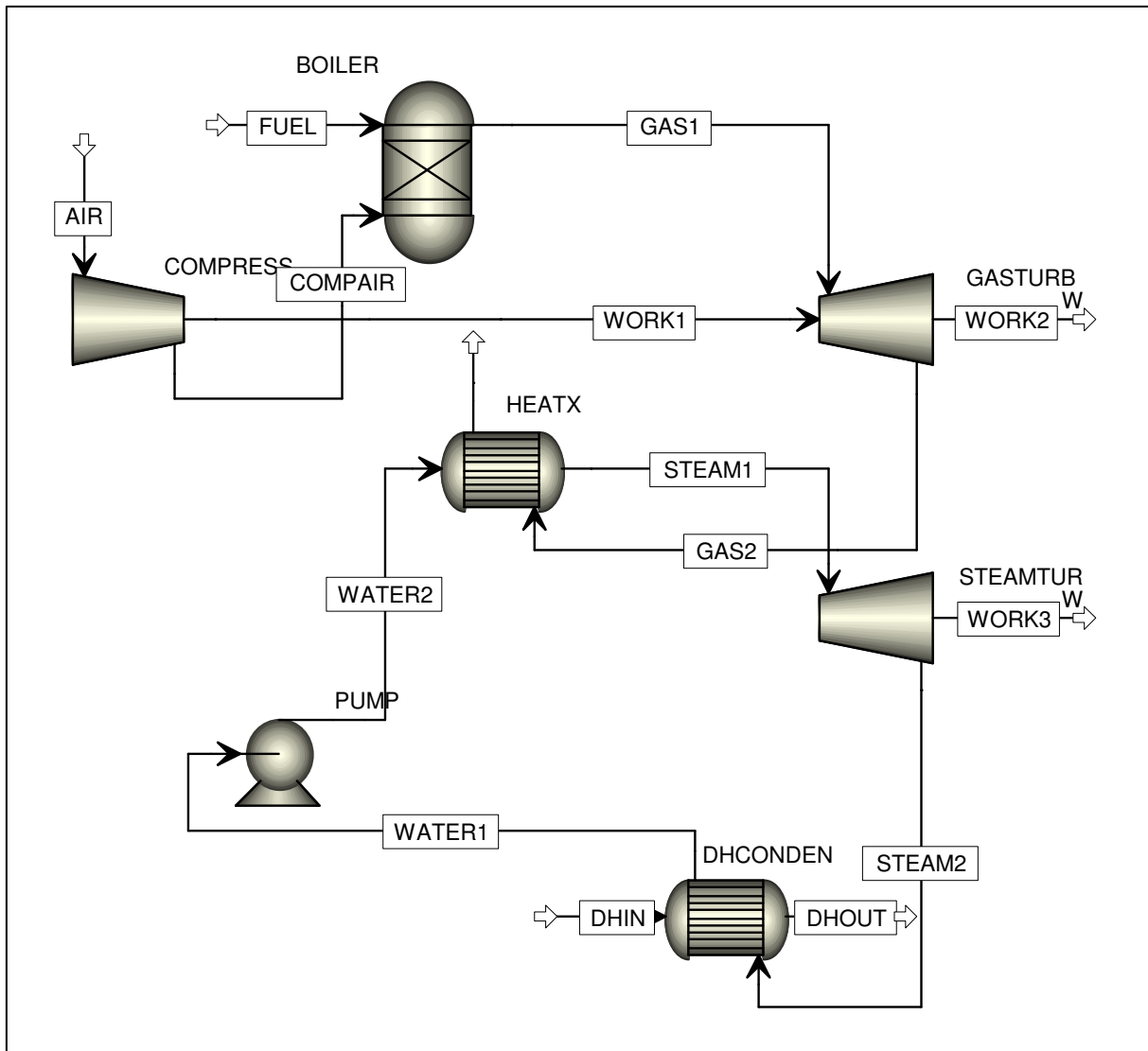


Figure 3: Gas-turbine CHP plant (7, 8)

The combined cycle is a combination of a gas-turbine cycle and a steam-turbine cycle, and has a higher thermal efficiency than either of the cycles executed individually. An example of a combined gas-steam cycle is shown in Figure 4. The combined cycle uses gas as a fuel and when the expanded gas has passed the gas-turbine, the high-temperature exhaust gases are used for steam production in a steam turbine. In a combined cycle power plant there are two generators producing electric power.



**Figure 4: Combined gas-steam CHP plant (8)**

Reciprocating engines can also be used in co-generation and they are operating on the same principles as their petrol and diesel engine automotive counterparts. The system differs very little from that of gas turbines, but reciprocating engines give a higher electrical efficiency. However it is more difficult to use the thermal energy as the temperatures are lower and it is dispersed between exhaust gases and engine cooling systems.

### **4.2 Steam-cycle CHP Plant**

The main components of a CHP plant are a steam boiler and a steam turbine connected to a generator. The steam is produced in the steam boiler at high pressures and continues to the turbine, where it expands and runs a generator, which produces electricity. After passing through the turbine the temperature and pressure of the steam drop, and continues to a warm-water condenser. In the warm-water condenser the district heating water is heated from about 30-60°C to about 70-120°C (9), depending on the backpressure of the turbine. The steam is cooled down and condenses to water, which is recirculated to the steam boiler as feed water.

To enable heat supply even when the turbine is not working, direct condensers are often installed to heat the district heating water directly. By equipping any of the turbines with a cold-water condenser it is possible to cover top loads of the electricity need. Hot-water accumulators are used to level out brief variations of the heat needed. To operate the district heating net

it is necessary to maintain the pressure of the system, which is often coordinated with the expansion tank. Steam or a special pressure maintenance pump is used for this purpose.

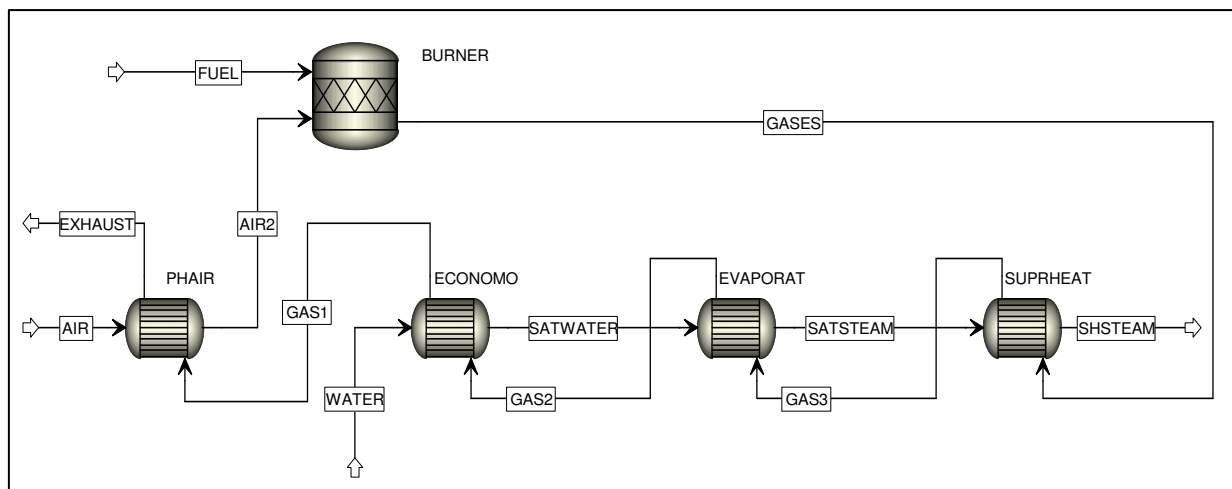
#### 4.2.1 Steam Boilers

The task of the steam boiler in a steam power process is to transform the feed water to steam of a suitable state for the process. A typical steam boiler is built up of a number of heat exchangers, where the flue gases are used to raise steam from the feed water. In co-generation, most of the steam goes directly to the turbine and is thereafter used for heating of the district heating water. The power generated by the steam is also often used to operate equipment within the plant.

A simple steam boiler can be described as a vessel partly filled with water, which is supplied with sufficient heat to form steam. The vessel is connected to a tube through which the steam can evaporate. If the outgoing steam equals the steam produced by the supplied heat, the pressure in the vessel is constant. If the amount of outgoing steam is less than the amount produced, the pressure rises, and if it is greater the pressure in the vessel drops. By changing the heat supply and/or the amount of outgoing steam, the pressure in the boiler can be regulated.

Figure 5 shows a flow-through boiler, which has been used in the process model. The main components of a steam boiler are:

- Fuel side: fuel preparation equipment and combustion equipment.
- Water-steam side: feed water pump, economiser, evaporator, and super heater.
- Air-exhaust side: air ventilator, air pre-heater, exhausts ventilator, exhaust-treatment equipment, and smokestack



**Figure 5: Connection scheme for a steam boiler of flow-through type**

In the combustion chamber the fuel is mixed with air. To get a complete combustion a surplus of air is needed and for wood the air factor ( $n$ ) lies between 1.25-1.40 (9), which means a surplus air of 25-40 % of the theoretical amount of air. The flue gases are used to convert the feed-water to superheated steam as well as to preheat the air. To get an effective heat transfer the temperature of the flue gases must remain above that of the fluid to be heated, where 30°C is a typical temperature difference (10).

The feed-water enters the steam boiler and in the economiser, if it is not constructed for steam generation, the water is heated to 25°C (9) below the saturation temperature to prevent steam generation due to operation irregularity. The saturated water continues to an evaporator, where the minimum pressure drop for the boiler's lowest load is set to 5 bar (9). Depending on the

top load the maximal pressure drop varies between different boilers. The fluid enters the superheater as saturated steam and the pressure drop lies between 5-8 % (9) of the construction pressure. The superheated steam then enters the turbine cycle.

By preheating the air needed for the combustion process, the efficiency of the boiler is increased and the exhaust gases are cooled to 120°C or less (10). Especially for wet fuel the air preheating is necessary to get a good combustion. 200-300°C are typical temperatures for wood (9), depending on the moisture content of the fuel.

### **4.2.2 Steam Turbines**

High-pressure steam is expanded within the turbine to produce mechanical energy, which can be used to drive an electricity generator. The power produced depends on how much the steam pressure can be reduced through the turbine. Typical inlet steam conditions, to get viable power generation, are 42 bar/400°C or 63 bar/480°C (10). The higher the turbine inlet pressure, the greater the power output, but higher steam pressures entail progressively greater boiler capital and running costs. As the steam cycles produce a large amount of heat compared with the electrical output, it is more cost-effective to integrate an incinerator with a steam-turbine-based co-generation unit. There are two types of steam turbines, according to the outgoing steam pressure of the turbine:

- back-pressure turbines, in which the exit pressure is greater than the atmospheric pressure;
- condensing turbines, in which the exit pressure is lower than the atmospheric pressure and a condenser is required.

The simplest arrangement is the back-pressure turbine, where the expansion of the steam is terminated at a higher pressure and the exit steam has a heating content corresponding to the need of the heat user. The main purpose of the plant is to produce heat, not electricity. The working process of a back-pressure turbine is similar to those for high-pressure and middle pressure turbines. A back-pressure turbine can be combined with a condensing turbine. The steam first expands in the back-pressure turbine and afterwards some “pass-out” steam is extracted to be used for other purposes and some continues through the condensing turbine. However, such extractions reduce the electrical output.

In a condensing turbine the outflow is directly connected to a condenser, i.e. heat exchanger, where the outflow steam is condensed. The temperature in the condenser determines the steam pressure at the outflow of the turbine. Condensing turbines are used to produce electrical energy, and to maximise the power output the steam is expanded down to vacuum with the help of the condenser. To ensure that the condenser cooling water in a district heating co-generation picks up enough heat, the turbine condenser can be operated near or even above atmospheric pressure. As with back-pressure turbines, pass-out steam can be extracted from condensing turbines as well. The drainage steam is used to pre-heat the feed water. The main advantages of the steam turbine system are its high overall efficiency, if the heat produced can be used for district heating, in industries etc, and that the heat to power ratios can be varied through flexible operation, the system is however quite slow to start up. Other advantages are that any type of fuel may be used, there are a wide range of sizes available and the system has a long working life, which can motivate the high investment costs.

### **4.2.3 Electricity Generators**

In a generator the mechanical energy of a turbine is converted to electricity. There are two types of generators, synchronous and asynchronous. Synchronous generators are more frequent-

ly used in CHP units, even though they usually are more expensive than asynchronous units with outputs below 200 kW<sub>e</sub>.

The main difference between these two generator types is that the synchronous generator can operate isolated from other generating plants and the grid. It can act as a standby generator as it can continue to supply power during grid failure. An asynchronous generator can only operate parallel with other generators and will cease to operate if it is disconnected from the mains or if the mains fail. Connection and interface to the grid is, however, simple.

## 5. Process Simulation in Aspen Plus

### 5.1 Introduction to Aspen Plus

The simulations of the integrated CHP and ethanol plant have been carried out using Aspen Plus, which is a chemical process simulation modelling software. The program was originally developed by the Massachusetts Institute of Technology for the U.S. Department of Energy to evaluate synthetic fuel technologies (11).

The program includes a library of standard unit operation blocks (e.g. pumps, heat exchangers, reactors, splitters), which represent processes taking place in an actual chemical plant. The simulation of a process plant is done by specifying configurations of unit operations and the flow of material, heat and work streams. Aspen Plus also has an extensive components database containing physical properties of a large number of pure components. Within the program there exist mathematical routines (convergence algorithms) for solving different equations of material and energy balances as well as equilibrium equations. Aspen Plus uses a sequential-modular approach to flow sheet convergence, where mass and energy balances for individual unit operation blocks are computed sequentially.

Besides the unit operation blocks Aspen Plus utilizes two more mechanisms to simulate chemical processes; FORTRAN blocks and design specifications. FORTRAN blocks are used for feed-forward control and allow incorporation of user code into a model to control variables in an Aspen Plus flow sheet. Design specification is used for feedback control, allowing the user to set particular design values for any flow sheet variable or function of flow sheet variable. The design specification varies the manipulating flow sheet variable, feed stream or block input, to achieve the design value. To compute design specification function values, FORTRAN statements can be used within the design specification block. It is possible to call any user-provided subroutine from either a design specification or a FORTRAN block.

Aspen Plus is also able to handle recycle streams, using a feature called “tear-streams”. Stream and block variables have to be manipulated iteratively, to converge upon the mass and energy balance, until it obtains a solution.

To be able to construct a process model in a flow sheet program the following three steps are necessary:

- Flow sheet definition: all inlet streams to the system have to be defined as well as different unit operations and their interconnecting streams. The flow sheet also indicates all outlet (product) streams.
- Chemical components: all chemical components in the system, from reactants to intermediates and products must be specified.
- Operating conditions: the operating conditions, such as temperature, pressure, heat duties etc., for every unit operation must be specified. All input streams have to be completely defined.

### 5.2 NREL Database

Physical property data for the different components of wood are not available in the standard Aspen Plus property databases. Inserting the available properties of these components in each simulation is quite tedious and mistakes are easily made. Therefore, the National Renewable Energy Laboratory (NREL) in the USA has constructed an in-house NREL Aspen Plus database, where the properties of the components can be called upon in the performed simula-

tions. The components from the NREL database used in the process model are cellulose, xylan and lignin.

## 6. Process Modelling of Ethanol and CHP Production

### 6.1 Case Studies and Limitations

Three process models were constructed to be able to perform the simulations in Aspen Plus; one for the steam boiler, one for the steam-turbine cycle integrated with an ethanol plant, and one turbine-cycle for a stand-alone CHP plant. Figure 6 shows a connection scheme for the turbine-cycle integration with an ethanol plant. The reason for not connecting the steam boiler with the turbine cycle was that the turbine process model was quite complicated to run and small changes in flow rates caused great changes within different unit operations. Performing the simulations separately made it easier to work with the models and gave small differences when an energy balance was done to control the results. The input data for the ethanol plant is based on simulations with Aspen Plus performed at Lund University (2). The model for the turbine-cycle is based on a real CHP plant integrated with a steam dryer.

Simulations of four different cases were done for every month of the year. The main reason for performing monthly simulations of the integrated plant is that the district heating need varies over the year. The lower district heating need during the warmer half of the year makes it possible for the CHP plant to supply the ethanol plant with more steam. The heat, needed for the evaporation and the distillation steps in the ethanol plant, depends on which kinds of processes are used, ethanol concentration after fermentation, energy optimisation, etc. The critical issue at a given process configuration is the amount of fresh water added to the system. In this thesis heat demands referring to a non-optimised ethanol process were used since the main purpose was to investigate the effect of integrating the ethanol process with a CHP plant. The case with the lowest heat demand had duties of 12.5 MW for the distillation and 10.9 MW for the evaporation, and the case with the highest heat demand had 18.1 MW and 20.7 MW respectively. It is possible that the waste heat from the ethanol plant could supply the district heating net. To be able to evaluate the effects of 10 MW waste heat, simulations for both the best and worst case were performed for all months except June, July and August, which have a district heating need less than 10 MW. Simulations were done for a stand-alone CHP plant as well, so that a comparison of the results could be made.

The maximum flow rate of the feed-water to the boiler was set to 46 kg/s with an incoming temperature of around 230°C, which gives a maximum heat effect of 76 MW and a maximum electric effect of 38 MW. Wood with 50 wt-% DM<sup>1</sup> was used as fuel. One simulation, though, was made mixing wood with the lignin fuel produced from the ethanol plant.

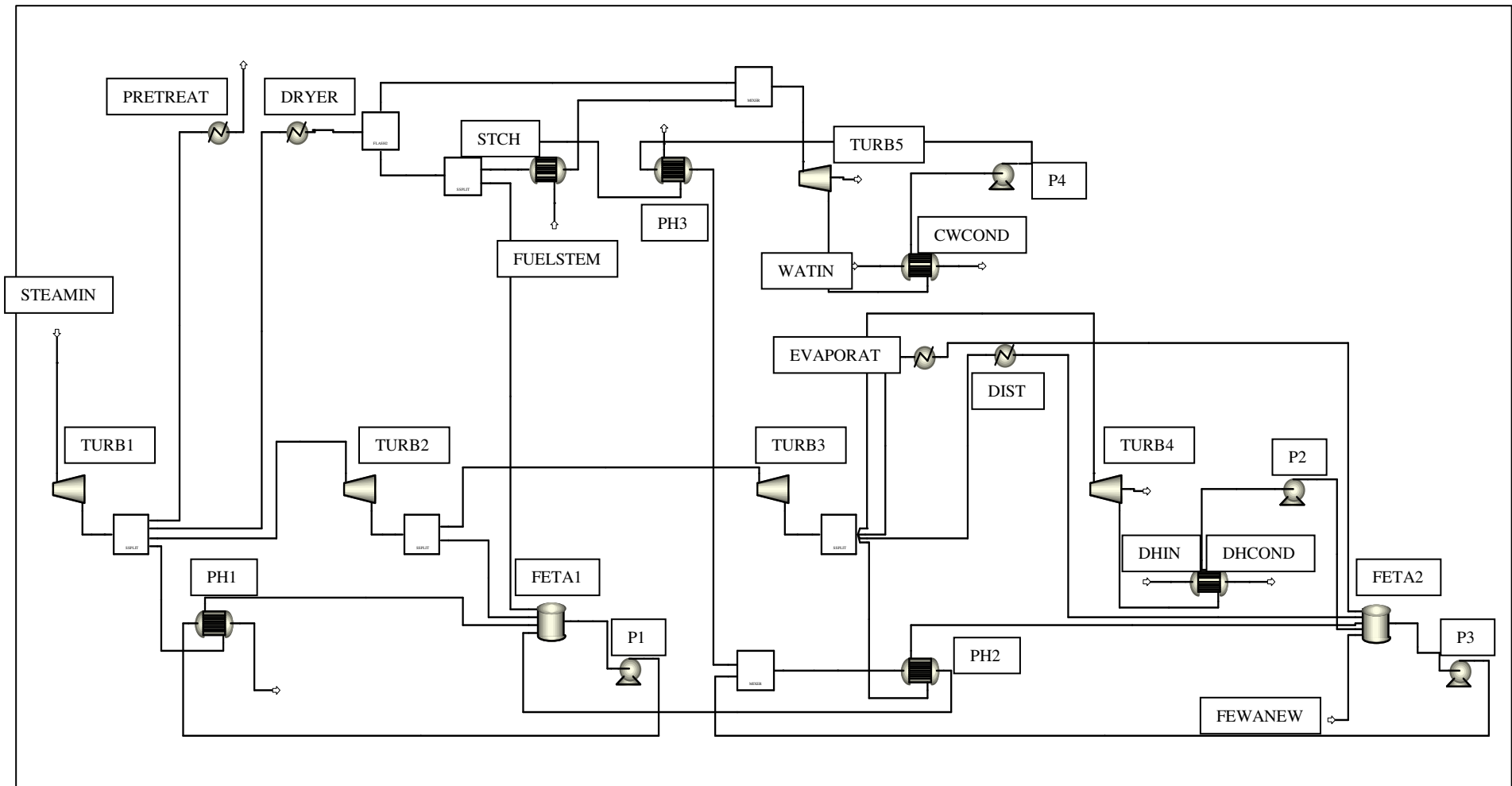
In a real steam-turbine cycle, the isentropic efficiency and the outgoing pressure depend on the incoming flow rate of the steam. However, the isentropic efficiency and outgoing turbine pressure were set to be the same in all the simulations, independently on the amount of incoming steam, which was another way of simplifying the model. This is supposed to be an approximation far enough as the steam flow rate varied between 28-46 kg/s. In the simulations for the stand-alone CHP plant the variations of the flow rate lied between 2-40 kg/s. Due to the larger span of the flow rate, both the isentropic efficiency and the outgoing pressure were varied in the model of the stand-alone CHP plant.

The ethanol production plant used in the process model, was supposed to give about 46,000 tons of ethanol per year. To produce this amount 200,000 ton raw material is needed. The secondary steam from the pretreatment was supposed to be reused in the distillation and evaporation steps, and the secondary steam from the dryer was recirculated to the steam-turbine cycle. Two additional simulations of July, as it is the month with the lowest district heating

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<sup>1</sup> DM = dry matter

need, were made to be able to assess the impact of increasing the size of the ethanol plant to 69,000 tons annually, as well as the impact of using the secondary steam (fuel steam) from the dryer within the distillation and evaporation instead of using it in the steam reformer.



**Figure 6: Connection scheme for CHP plant integrated with an ethanol plant. Ethanol plant: PRETREAT = Pretreatment ( $p = 25$  bar), DRYER = Steam dryer ( $p = 25$  bar), EVAPORAT = Evaporation ( $p = 4$  bar), DIST = Distillation ( $p = 4$  bar), CHP plant: TURB1-2 = High-pressure turbine, TURB3-4 = Low-pressure turbine, TURB5 = low-pressure turbine, CWCOND = Cold-water condenser, DHCOND = district heating condenser, STCH = Steam reformer, PH1-3: Feed-water preheaters, FETA1-2 = Feed-water tanks, P1-4 = Pumps, Incoming streams: STEAMIN = superheated steam from boiler, FUELSTEM = Evaporated steam from wet fuel, WATIN = cooling water, DHIN = Incoming district heating water, FEWANEW = Fresh feed-water**

## 6.2 Ethanol Production Plant

Figure 7 shows the material streams and heat needed for different operations within the ethanol plant. The enzymatic hydrolysis process was used for ethanol production, but the conditions for the pretreatment have been changed somewhat as compared to the model used in Lund, which is described down below. Even though the prerequisites are changed, the heat duty is set to be the same, as it gives small errors compared with the total balance. In the flow-sheet model the pretreatment, lignin fuel dryer, evaporation and distillation are represented by the unit operation “HEATER”, which only has one inlet and one outlet stream. The heaters’ working pressure is supposed to be the same as the ingoing steam pressure, 25 bar (pretreatment and dryer) and 4 bar respectively (evaporation and distillation) and the outgoing vapour fraction is set to 0 for all the heaters, the condensates are supposed to be saturated as they are reused as feed-water.

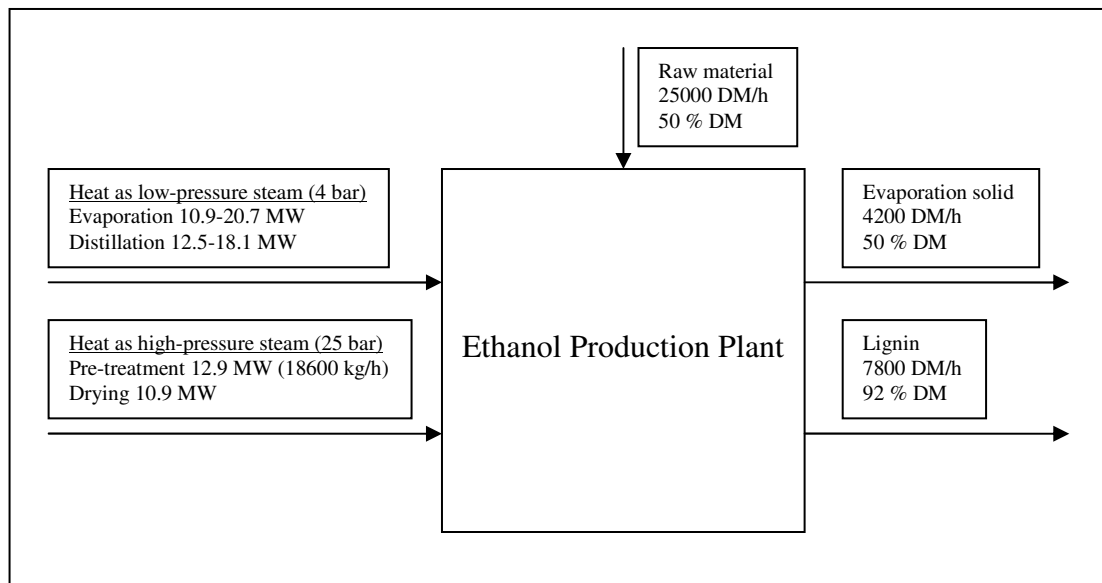


Figure 7: Energy and steam flows for an annual ethanol production of 46000 ton. DM = dry matter. (2)

### 6.2.1 Pretreatment

In the pretreatment process fresh wood chips of spruce are impregnated with sulphur dioxide and thereafter heated directly with high-pressure steam for about five minutes. To process 25,000 kg DM/h with a dry-matter content of 50 %, 18.6 ton steam/h is needed, and instead of saturated steam at 22 bar superheated steam at 25 bar was used. The pressure is decreased in two steps, first to 4 bar and thereafter to 1 bar. These two steps are not considered in this model. As the steam is mixed with the raw material, the condensate from the pretreatment is contaminated and can not be used as feed-water in the CHP plant. The secondary steam, which lies around 6.2 MW, is however reused within the ethanol process, and the lost steam is compensated with 18.6 ton/h of fresh feed-water.

### 6.2.2 Distillation and Evaporation

The distillation process step consists of two strippers and one rectifier with top stage pressures of 3, 1.25 and 0.35 bar. In the process model the distillation heater needs steam with an effect between 15.8-21 MW, depending on the energy optimisation in plant. As the secondary steam from the pretreatment is reused, the heat duty needed is 12.5 MW for the case with the lower heat demand and 18.1 MW for the case with the higher heat demand.

The stillage goes to a liquid-solid separation step, from which the solid lignin continues to a dryer. The liquid phase continues to the evaporation step, where it is concentrated from about 3-4 % DM to 50 % DM. The evaporator system consists of five effects where the pressure in the first effect is 3 bar and in the last 0.2 bar. The steam effects needed for the evaporation is 13.8-24 MW and 10.9-20.7 MW when the pretreatment steam is reused. The clean condensates from the evaporation and the distillation are recirculated to an open feed-water heater in the CHP plant.

### **6.2.3 Steam Drying**

The solids separated from step 1 and 2 of the distillation, having temperatures of 130°C and 115°C respectively, are mixed and flashed to a temperature of about 100°C and a pressure of 4 bar. Superheated steam at 25 bar is used to dry the lignin solid from 20-40 % DM to 92 % DM. An effect of 11.4 MW is needed in this step. To represent this process in the model the clean condensate of 25 bar from the dryer is flashed down to 4 bar, where the vapour fraction goes directly to a low-pressure turbine. It should be clear that there is no fuel stream present in the process model, only the superheated steam heating the fuel and the secondary fuel steam separated from the fuel.

The contaminated fuel steam, which has a temperature of 150°C, a pressure of 4.5 bar and an enthalpy of 10.5 MW, is used to convert some of the liquid fraction from the flashed condensate to steam in a steam reformer. The steam continues to the low-pressure turbine, which has an isentropic efficiency of 0.31 and an outgoing pressure of 0.02 bar.

After expanding in the turbine, the steam is cooled in a cold-water condenser. The cooling water is supposed to be taken from a nearby lake or river. To be able to bring the water back to the source, the temperature difference between the outgoing and incoming cooling water cannot be too high. The flow rate was adjusted so that the temperature difference lay around 6°C.

Before the condensate goes back to the CHP plant it is preheated, with the surplus energy of the fuel steam, in a low-pressure feed-water heater. The hot outlet temperature was specified to 19°C. The reason for specifying the hot stream outlet temperature is that the energy content of the fuel steam is the limiting factor, when the heating of the cold stream should be as much as possible.

## **6.3 Combined Heat and Power Plant**

The CHP plant consists of a steam boiler, one high-pressure and one low-pressure turbine, two feed-water tanks, two feed-water preheaters, one district heating condenser and three pumps, which increase the final feed-water pressure to 175 bar. This chapter will describe the steam-cycle in the CHP plant, and the steam boiler process model will be further described in next chapter. To construct the process model for the steam-turbine cycle data from a real plant was used. The maximum feed-water flow rate was set to 46 kg/s. The flow rate was not increased any further, because then it would have been necessary to decrease the isentropic efficiency of the high-pressure turbine.

### **6.3.1 Specifying Streams and Components**

In an Aspen Plus process model simulation it is necessary to define the inlet streams; the outlet streams and internal streams are calculated by the program. In the turbine-cycle, integrated with an ethanol plant, there are five inlet streams, whose input specifications are shown in Table 2. As for the ethanol process, the only component considered is water.

*Table 2: Specifications for inlet streams of the turbine cycle*

Model name	Description	Specifications
DHIN	Return flow from district heating net	Temperature: 47°C Pressure: 10 bar Flash Options: liquid only
FEWANEW	Fresh water to substitute the steam "lost" to the pre-treatment	Temperature: 5°C Pressure: 1 bar Flash Options: liquid only
FUELSTEM	Evaporated steam from the dried fuel	Temperature: 150°C Pressure: 4.5 bar
STEAMIN	Incoming steam to the turbine-cycle from the steam boiler	Temperature: 540°C Pressure: 140 bar
WATIN	Water from a river, lake etc, which is used to cool the steam from the low-pressure turbine connected to the steam dryer	Temperature: 5°C Pressure: 1 bar Flash Options: liquid only

For more realistic simulations it would have been better to vary the temperature for FEWANEW and WATIN over the year. To simplify the process model this was not done.

### 6.3.2 Unit Operations

The input specifications for all the unit operation blocks, except those already described in Chapter 6.2, are presented in Table 3.

*Table 3: Specifications for unit operations blocks of the turbine-cycle*

Model name	Description	Specifications
DHCOND	Warm-water condenser for district heating	Hot stream outlet vapour fraction: 0
FETA1	Feed-water tank	Outlet pressure: 6 bar
FETA2	Feed-water tank	Outlet pressure: 4 bar
P1	Feed-water pump	Pressure: 175 bar Pump efficiency: 0.7
P2	Feed-water pump	Pressure: 1 bar Pump efficiency: 0.7
P3	Feed-water pump	Pressure: 10 bar Pump efficiency: 0.7
PH1	Closed feed-water preheater	Hot stream outlet vapour fraction: 0
PH2	Closed feed-water preheater	Hot stream outlet vapour fraction: 0
TURB1	High-pressure turbine	Outlet pressure: 25 bar Isentropic efficiency: 0.9
TURB2	High-pressure turbine	Outlet pressure: 6 bar Isentropic efficiency: 0.89
TURB3	Low-pressure turbine	Outlet pressure: 4 bar Isentropic efficiency: 0.38
TURB4	Low-pressure turbine	Outlet pressure: 0.5 bar Isentropic efficiency: 0.33

As shown in Table 3 the hot stream outlet vapour fraction was set to 0 for all heat exchangers. The reason for this is that most of these streams will continue to a pump, why it is desirable to have as a high liquid fraction as possible, as the pumps only accept a vapour fraction less than 0.01 %.

A MIXER unit operation block was used to represent the feed-water tanks. The outlet streams from the feed-water tanks continue to two different pumps. As mentioned above it is necessary that the incoming stream to a pump has a low vapour fraction to make the system converge. In the versions of the process model, a FLASH block, which separates a stream in one liquid and vapour phase, was used to solve this problem. However, this unit operation block does not represent a real feed-water tank and using that block gave low electrical output from the turbines. By changing to a MIXER block, the flow rates could be varied manually and the electrical output increased.

The high-pressure turbine was represented by two turbines, TURB1 and TURB2, in the process model, as a certain amount of steam is extracted to the pre-treatment, the dryer and a feed-water preheater at 25 bar. The outgoing pressure was set to 6 bar, where some of the steam continues to the low-pressure turbine and some goes to a feed-water tank.

In the low-pressure turbine, which is represented by two turbines as well, steam is extracted at 4 bar to the evaporation, distillation and to a second feed-water preheater. The outgoing steam pressure was set to 0.5 bar and continues to a warm-water condenser in order to heat the district heating water. The condensate is then circulated to the second feed-water tank.

The isentropic efficiencies for the turbines are quite low, as shown in Table 3. The reason for setting such low isentropic efficiencies was to guarantee that the outgoing streams from the turbines had vapour fraction 1.

## **6.4 Steam boiler**

The process flow diagram for the steam boiler is shown in Figure 6, Chapter 4.2.1. A fuel stream and an air stream are fed to a reactor and a water stream is fed to an economiser. The flue gases produced in the reactor are used to preheat the incoming air as well as heating the water to superheated steam.

### **6.4.1 Specifying Streams and Components**

Wood with a dry-matter content of 50 wt-% was used as fuel in the simulations, and one simulation was done mixing wood with lignin fuel of 92 wt-% DM. In the simulations the FUEL stream consists of both the MIXED (water) and the CISOLID (solid components in wood/lignin) substream. Stream classes are used to define the structure of simulation streams. The CISOLID substream is used for homogeneous solids that have a defined molecular weight. The average composition of wood and lignin is shown in Table 4. C6 represents cellulose, glucan, mannan and galactan, the lignin represents lignin, extractives and acetyl groups and C5 represents hemicellulose, xylan and arabinan.

*Table 4: Average composition of wood and lignin fuel (2)*

<b>Component</b>	<b>Wood, wt-%</b>	<b>Lignin, wt-%</b>
C6	56,4	19
Lignin	32,7	81
C5	8,9	
Ashes	2	
<b>Total</b>	<b>100</b>	<b>100</b>

The AIR and WATER streams consist only of the MIXED substream. The components represented in the AIR stream are 0.21 O<sub>2</sub> and 0.79 N<sub>2</sub>, and in the WATER stream H<sub>2</sub>O. The input pressure for the AIR and FUEL streams was set to 1 bar, but the temperatures were varied over the year, see Table 5.

*Table 5: Temperature variations over the year for the air and the fuel streams*

<b>Month</b>	<b>Air/fuel temperature, °C</b>
January	-9.2
February	-6.8
March	-4.1
April	2.5
May	8.5
June	13.1
July	14.7
August	14.1
September	9.5
October	3.0
November	-3.3
December	-7.2

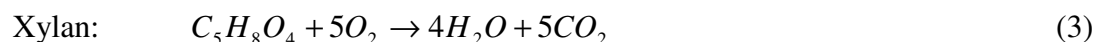
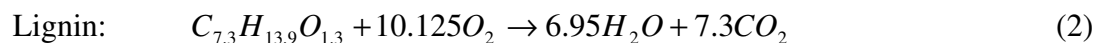
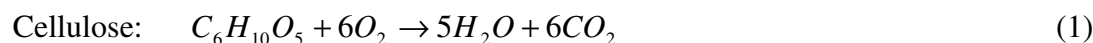
The pressure of the WATER stream was set to 175 bar, and the inlet temperature was varied depending on the flow rate of the water. Table 6 does not show the flow rates used in the simulations; as there are about 30 different simulations it only shows the approximate temperature for a certain flow rate of feed-water. The amount of fuel needed depends on the amount of water that should be converted to steam, as well as on the outdoor temperature. Therefore an average is given for the fuel needed for a certain feed-water flow.

**Table 6: Flow rate and temperature of feed-water and fuel needed**

Feed-water flow, kg/s	Temperature, °C	Fuel flow, ton/h
10	160	14.7
20	180	29.4
30	200	44.1
40	220	58.7
50	240	73.4

### 6.4.2 Unit operations

The reactor in the steam boiler process model is represented by the unit operation block “RGIBBS”, which is set to be both isobaric and adiabatic. As the pressure and heat duty are set to certain values, see Table 7, the program asks for reactions within the block. Wood fuel has about 2 wt-% ashes, and since the ashes never react in the combustion process, they are not considered in the process model. This gives 58 wt-% cellulose, 33 wt-% lignin and 9 wt-% xylan. The combustion reactions are taken from NREL’s document regarding the physical property database for bio-fuel components and the efficiency was supposed to be 95 % for all the reactions shown below (12).



The air preheater, economiser, evaporator and superheater are represented by the unit operation “HEATX”. The input specifications for these blocks are shown in Table 7. The feed-water is heated to saturated water in the economiser, to saturated steam in the evaporator and finally to superheated steam in the superheater. The flue gases from the reactor are used for this purpose as well as preheating the incoming air. The air flow was controlled by a design specification, so that the ratio of oxygen in the outgoing flue gases and the incoming air was kept at 0.2.

**Table 7: Input specifications for unit operation blocks of the steam boiler**

Model name	Description	Specifications
BURNER	Combustion chamber	Pressure: 1 bar Heat duty: 0 W
ECONOMO	Economiser	Cold stream outlet temperature: 330°C
EVAPORAT	Evaporator	Cold stream outlet vapour fraction: 1 Cold side outlet pressure: 150 bar
SUPRHEAT	Superheater	Cold stream outlet temperature: 540°C Cold side outlet pressure: 140 bar
PHAIR	Air preheater	Cold stream outlet temperature: 250°C

For one simulation the wood fuel was mixed with the lignin fuel. As the lignin fuel has a higher content of dry matter, there is less need for preheating the air. The cold stream outlet temperature in the air preheater was then set to 235°C.

## 7. Results and Discussion

The ethanol plant used in the process model needs 50,000 kg/h raw material with a dry-matter content of 50 wt-% to be able to produce 5,813 kg/h ethanol. Within the ethanol production process, based on lignocellulosic material with enzymatic hydrolysis, pretreatment, distillation and evaporation need steam, which can be supplied either by a CHP plant or by a separate steam boiler. The heat demand for the distillation and the evaporation varies between 12.5-18.1 MW and 10.9-20.7 MW respectively, depending on the amount of fresh water added to the process. Steam of 4 bar and 144°C is used in these processes. The pretreatment is supplied by 18,600 kg/h superheated steam at 25 bar, which corresponds to about 12.9 MW. The secondary steam of 6.2 MW from the pretreatment is reused within the distillation and evaporation processes.

The ethanol production gives two solid co-products, lignin<sup>2</sup> and syrup. The syrup is evaporated to a dry-matter content of 50 wt-% and the lignin is dried in a steam dryer. To dry the lignin from a dry-matter content of 30 wt-% to 92 wt-%, the heat demand is around 11.4 MW. As for the pretreatment superheated steam at 25 bar is used for this process and the secondary fuel steam of 10.5 MW is reused in a steam reformer, connected to a low-pressure turbine.

Four different simulation cases were established to consider the heat demand variations for the distillation and evaporation as well as to investigate the impact of using waste heat from the ethanol plant for district heating. All of these cases were compared with a CHP plant with no extractions to an ethanol plant and fuel dryer. The simulations were performed for every single month of the year and the four cases are:

- Case I: the total effect needed for the ethanol plant and lignin dryer is 47.7 MW.
- Case II: the total effect needed for the ethanol plant and lignin dryer is 63.1 MW.
- Case III: the ethanol plant has the same heat demand as in Case I and 10 MW waste heat is used for district heating.
- Case IV: the ethanol plant has the same heat demand as in Case II and 10 MW waste heat is used for district heating.

To be able to evaluate if both the ethanol plant and the CHP plant could profit by connecting the plants, an economic analysis has been done. For these calculations the integrated plant was supposed to be run 8,760 hours per year. The stand-alone CHP plant however was supposed to be run 7,536 hours per year, as the district heating need is not high enough during the summer months. Integrating an ethanol plant with a CHP plant gives the CHP plant a higher income for district heating and electricity, but higher costs due to an increased fuel consumption. The differences between the incomes and fuel costs have been compared to a stand-alone CHP plant for all the cases, to be able to estimate the profitability. Table 8 shows the prices for the CHP plant that have been used in the analysis. The steam price was varied to be able to estimate when an integrated plant gets a higher difference between incomes and costs than for the stand-alone CHP plant.

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<sup>2</sup> Which is actually a mixture of lignin and residual cellulose

**Table 8: Prices used for economic calculations**

District heating, SEK/MWh	Electricity, SEK/MWh	Steam, SEK/MWh	Fuel, SEK/MWh
479	400	10-250	140

The steam costs are not included in the production costs for ethanol shown in Table 9, as the steam costs vary depending on the amount of steam and the price of it from the CHP plant. The steam price from the steam boiler needed to run the ethanol plant during the whole year was set to 250 SEK/MWh, independently of the steam price from the CHP plant. All of the costs are retail prices and are used to be able to compare the different cases.

**Table 9: Breakdown of costs, SEK/L ethanol (2)**

Raw material	1.40
Chemicals	0.29
Enzymes	0.64
Co-products	-1.03
Syrup	-0.20
Other utilities	0.22
Maintenance, insurance	0.35
Capital costs	1.24/1.10 <sup>3</sup>
Labour	0.21
Total excl. steam costs	3.12/2.98

Two additional simulations were done in order to estimate the impact of using the secondary fuel steam from the steam dryer within the ethanol process (distillation and evaporation) instead of in a steam reformer, and of integrating a larger ethanol plant. These simulations were done for the month of July. Another simulation was done for January, mixing the dried lignin fuel with wood.

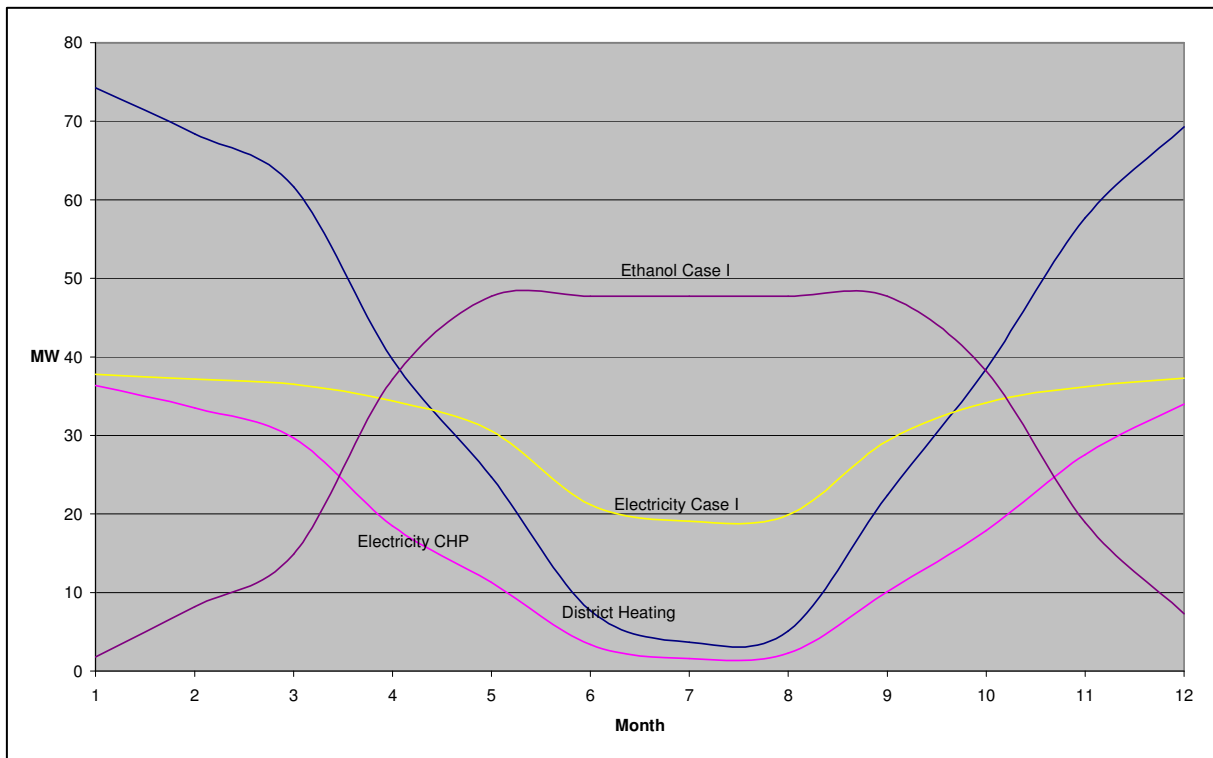
## **7.1 Results from Case I**

Figure 8 and Table 10 show the district heating and steam demand for the ethanol plant supplied by the CHP plant over the year. The electrical output for the integrated plant and the stand-alone CHP plant are also represented there. As shown in Figure 8 the CHP plant can not supply the ethanol plant with its total effect need during the whole year. To be able to run the ethanol plant on its full capacity all the year, it is necessary to install a steam boiler of 45.9 MW for this case, which is used during October-April.

<sup>3</sup> The higher capital cost is for an ethanol plant with a capacity of  $64.5 \cdot 10^6$  litres/year and the lower capital cost is for a plant with a capacity of  $96.7 \cdot 10^6$  litres/year.

**Table 10: District heating and steam to ethanol plant and electrical output for a stand-alone CHP plant and Case I**

Month	District heating from CHP plant, MW	Electrical output, stand-alone CHP plant, MW	Electrical output, Case I, MW	Steam to ethanol plant, Case I, MW
Jan	74.3	36.4	37.8	1.8
Feb	68.4	33.5	37.2	8.2
Mar	61.7	29.7	36.5	14.9
Apr	39.7	18.5	34.4	37.2
May	24.7	11.3	30.6	47.7
Jun	7.7	3.4	21.2	47.7
Jul	3.7	1.6	19.1	47.7
Aug	5.1	2.3	19.9	47.7
Sep	22.4	10.1	29.3	47.7
Oct	38.5	17.9	34.2	38.2
Nov	57.7	27.6	36.2	18.9
Dec	69.3	34.0	37.3	7.3



**Figure 8: Variations over the year of district heating, steam and electricity produced in the CHP plant of Case I, compared with the electrical output from a stand-alone CHP plant**

The CHP plant gets its profit from selling heat to the district heating net, electricity to the grid and steam to the ethanol plant. The CHP plant with no extractions to an ethanol plant only gets its income from district heating and electricity. The integrated CHP plant can sell about 76 % more energy than the stand-alone CHP plant, but instead the fuel need increases by

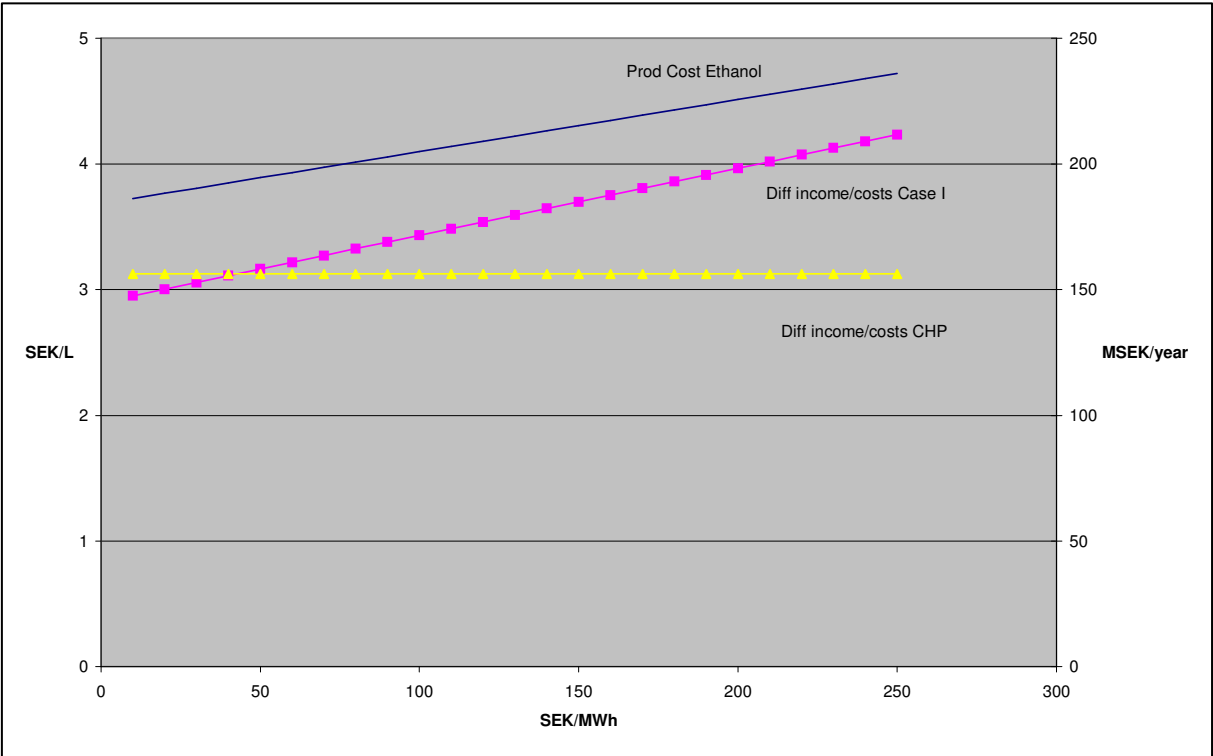
82 %. Table 11 shows the annual production of district heating, electricity and steam as well as the fuel needed for the stand-alone CHP plant and the integrated plant of Case I.

**Table 11: Annual production of district heating, electricity, steam as well as fuel demand for a stand-alone CHP plant and for Case I**

	Electricity, GWh	District heating, GWh	Steam, GWh	Fuel, GWh
<b>Case I</b>	272	344	267	921
<b>Stand-alone CHP plant</b>	162	339		505

The annual ethanol production was supposed to be about  $64.5 \cdot 10^6$  litres, which corresponds to a total need of energy for the ethanol plant of 418 GWh. When the electricity price is 400 SEK/MWh, the district heating price 479 SEK/MWh and the fuel price 140 SEK/MWh, the difference between the incomes and the costs becomes larger for the integrated CHP plant of Case I than the stand-alone CHP plant at a steam price of 43 SEK/MWh or more.

The production costs for ethanol at a steam price of 43 SEK/MWh from the CHP plant and 250 SEK/MWh from its own steam boiler is 3.86 SEK/litre, where 0.76 SEK/litre is the effective steam cost. Figure 9 shows the difference between incomes and costs in million SEK for the integrated plant of Case I and a stand-alone CHP plant as well as the production costs per litre for ethanol, when the steam price from the integrated CHP plant is varied between 10-250 SEK/MWh.



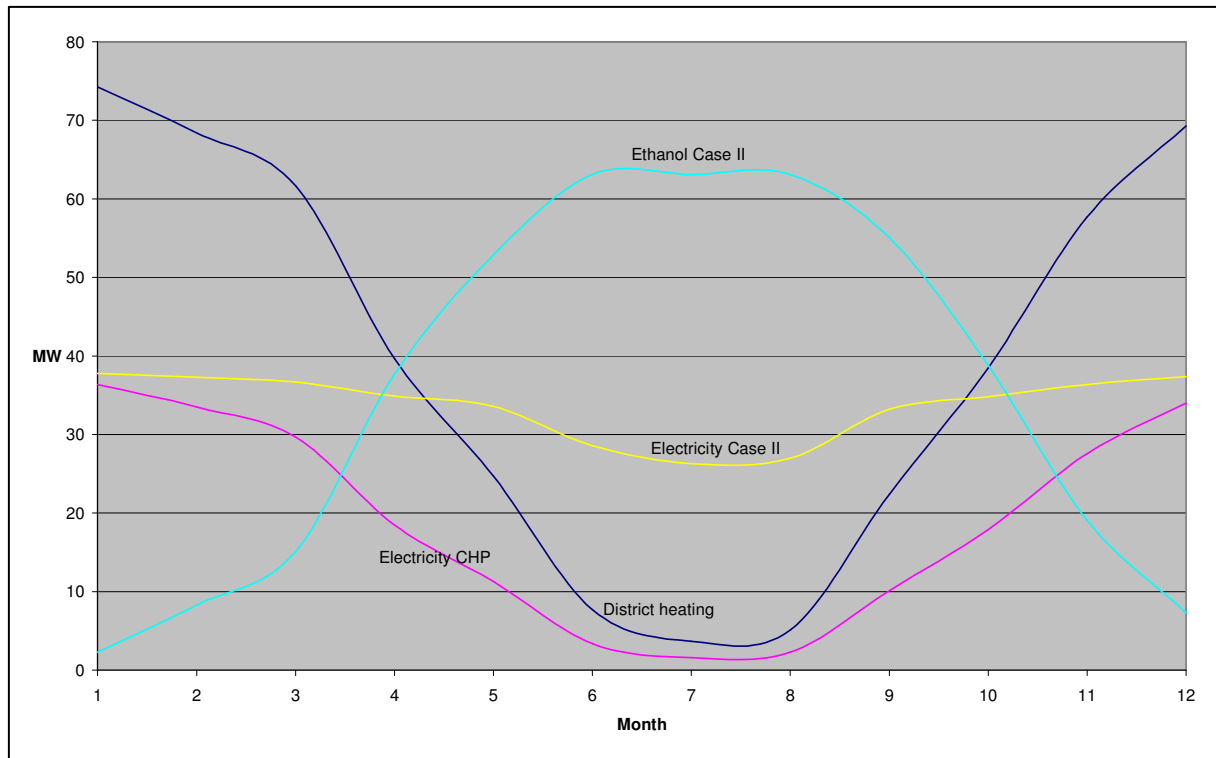
**Figure 9: Production costs for ethanol and differences between costs and incomes as a function of the steam price from the CHP plant of Case I and for a stand-alone CHP plant**

## 7.2 Results from Case II

Figure 10 and Table 12 show the district heating and steam demand for the ethanol plant supplied by the CHP plant over the year. The electrical output for the integrated plant and the stand-alone CHP plant are also represented there. As shown for Case II, the CHP plant can not supply the ethanol plant with its total effect need during the whole year. To be able to run the ethanol plant on its full capacity all the year, it is necessary to install a steam boiler of 60.8 MW for this case, which is used during October-May.

**Table 12: District heating, steam to ethanol plant and electrical output for a stand-alone CHP plant and Case II**

Month	District heating from CHP plant, MW	Electrical output, stand-alone CHP plant, MW	Electrical output, Case II, MW	Steam to ethanol plant, Case II, MW
Jan	74.3	36.4	37.8	2.3
Feb	68.4	33.5	37.3	8.3
Mar	61.7	29.7	36.7	15.1
Apr	39.7	18.5	34.9	37.7
May	24.7	11.3	33.6	52.9
Jun	7.7	3.4	28.6	63.1
Jul	3.7	1.6	26.3	63.1
Aug	5.1	2.3	27.0	63.1
Sep	22.4	10.1	33.2	55.1
Oct	38.5	17.9	34.8	38.8
Nov	57.7	27.6	36.4	19.1
Dec	69.3	34.0	37.4	7.3



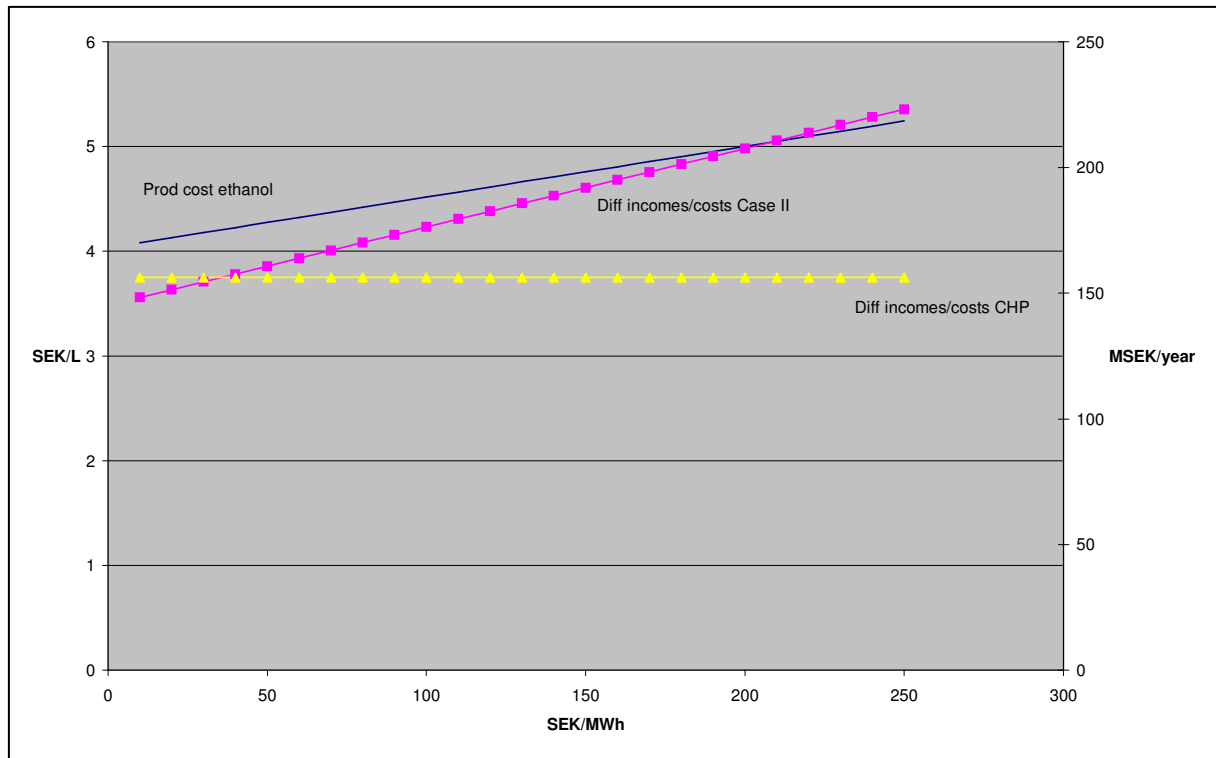
**Figure 10: Variations over the year of district heating, steam and electricity produced in the CHP plant of Case II, compared with a stand-alone CHP plant**

The integrated CHP plant of Case II can sell about 91 % more energy than the stand-alone CHP plant, but instead the fuel need increases with 96 %. Table 13 shows the annual production of district heating, electricity and steam as well as the fuel needed for the stand-alone CHP plant and the integrated plant of Case II.

**Table 13: Annual production of district heating, electricity, steam as well as fuel demand for a stand-alone CHP plant and for Case II**

	Electricity, GWh	District heating, GWh	Steam, GWh	Fuel, GWh
<b>Case II</b>	295	344	312	982
<b>Stand-alone CHP plant</b>	162	339		505

Figure 11 shows the difference between incomes and costs for the integrated plant of Case II and a stand-alone CHP plant as well as the production costs for ethanol, when the steam price from the integrated plant is varied from 10 to 250 SEK/MWh. The CHP plant of Case II results in a larger difference than the stand-alone CHP plant at a steam price of 36 SEK/MWh. The production costs for ethanol at a steam price of 36 SEK/MWh from the CHP plant is 4.21 SEK/litre, where 1.11 SEK/litre is the effective steam cost. The production costs for Case II are higher than for Case I as the energy need is higher.



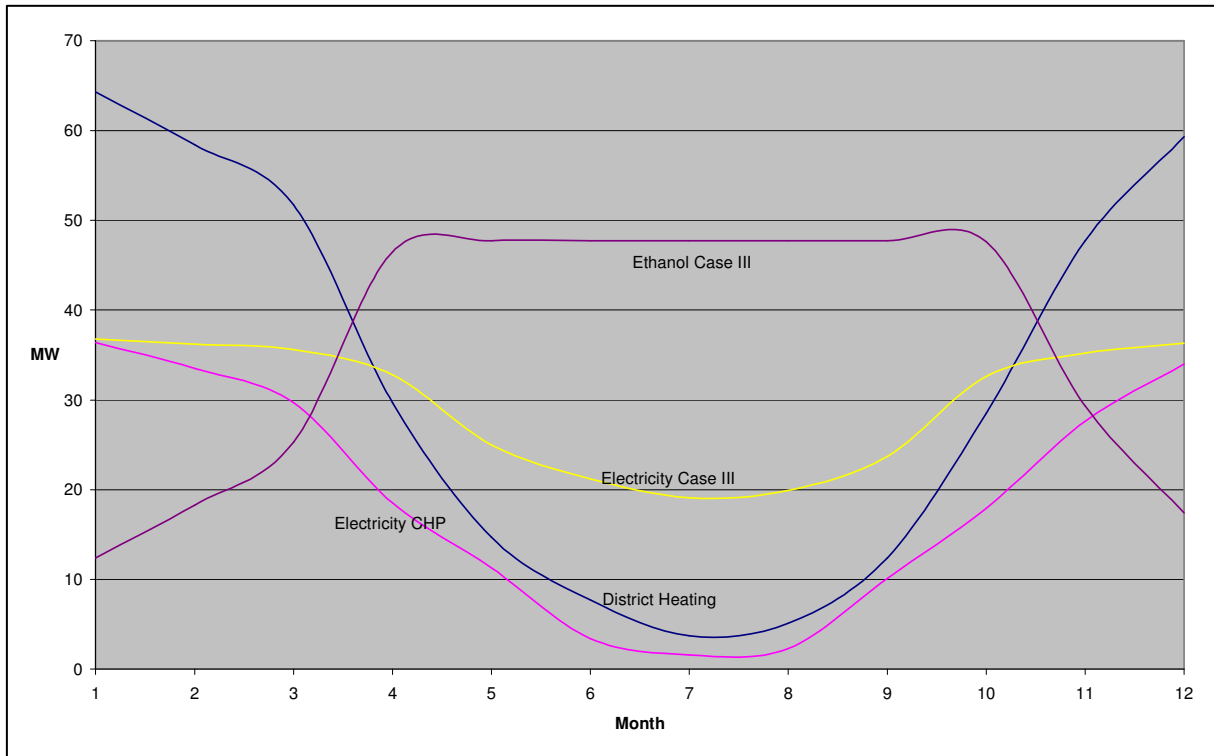
**Figure 11:** Production costs for ethanol and differences between costs and incomes as a function of the steam price from the CHP plant of Case II and for a stand-alone CHP plant

### 7.3 Results from Case III

Figure 12 and Table 14 show the district heating and steam demand for the ethanol plant supplied by the CHP plant over the year. The electrical output for the integrated plant and the stand-alone CHP plant are also represented there. The CHP plant can not supply the ethanol plant with its total effect need during the whole year, why it is necessary to install a steam boiler of 35.3 MW for Case III, which is used during October-April. The steam boiler is smaller than for Case I, as the waste heat from the ethanol plant replaces some of the district heating produced in the CHP plant, which leads to more steam being produced for the ethanol plant. Note that the district heating need lies 10 MW above the data presented in Table 14, except for the summer months (June, July and August) which have the same district heating need as for Case I. Waste heat of 10 MW from the ethanol plant is used to supply the district heating net during September-May, but not during the summer as the district heating need is less than 10 MW.

**Table 14: District heating and steam to ethanol plant and electrical output for a stand-alone CHP plant and Case III**

Month	District heating from CHP plant, MW	Electrical output, stand-alone CHP plant, MW	Electrical output, Case III, MW	Steam to ethanol plant, Case III, MW
Jan	64.3	36.4	36.8	12.4
Feb	58.4	33.5	36.2	18.3
Mar	51.7	29.7	35.6	25.3
Apr	29.7	18.5	32.8	46.5
May	14.7	11.3	25.0	47.7
Jun	7.7	3.4	21.2	47.7
Jul	3.7	1.6	19.1	47.7
Aug	5.1	2.3	19.9	47.7
Sep	12.4	10.1	23.7	47.7
Oct	28.5	17.9	32.6	47.6
Nov	47.7	27.6	35.2	29.3
Dec	59.3	34.0	36.3	17.4



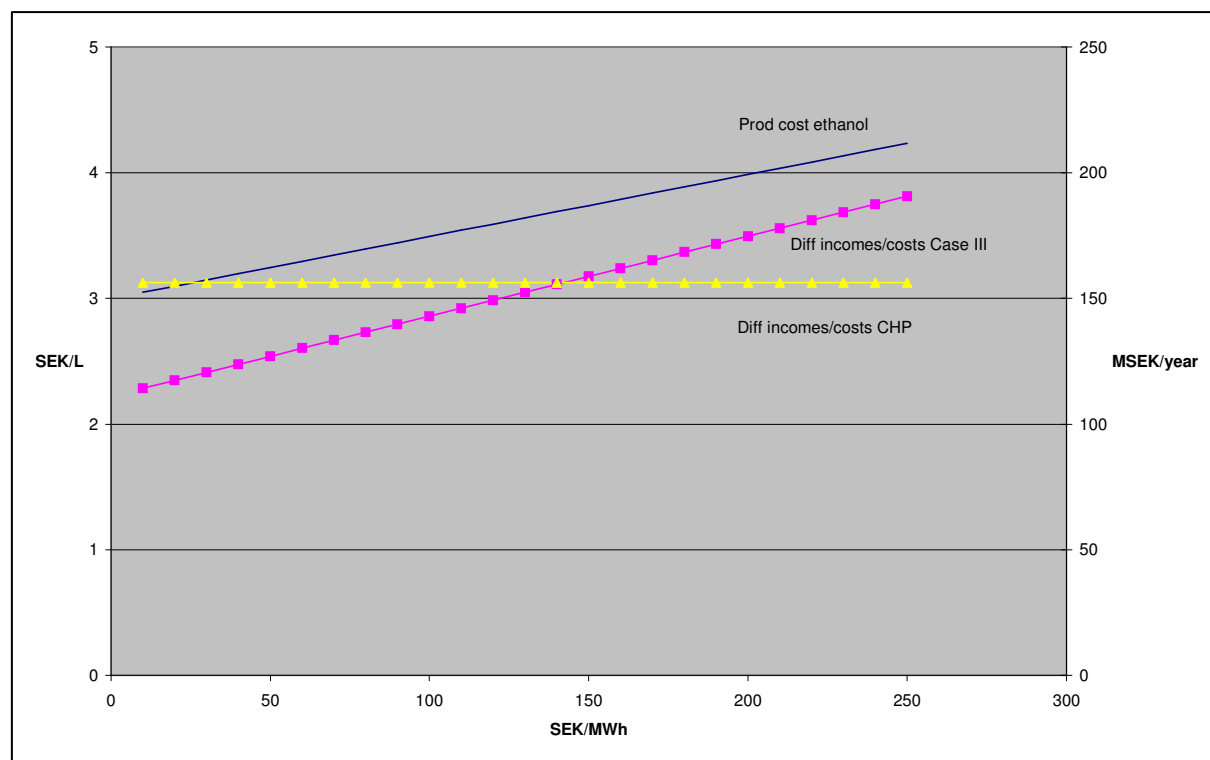
**Figure 12: Variations over the year of district heating, steam and electricity produced in the CHP plant of Case III, compared with the electrical output from a stand-alone CHP plant**

As shown in Table 15 the district heating production for Case III is less than for the CHP plant without extractions to an ethanol plant. Even though the integrated CHP plant can sell about 71 % more energy than the stand-alone CHP plant and the fuel need increases with 78 %.

**Table 15: Annual production of district heating, electricity, steam as well as fuel demand for a stand-alone CHP plant and for Case III**

	Electricity, MWh	District heating, MWh	Steam, MWh	Fuel, MWh
<b>Case III</b>	258	279	318	898
<b>Stand-alone CHP plant</b>	162	339		505

As the district heating production for Case III is decreased compared with the stand-alone CHP plant, the steam price must be higher to get a greater difference between incomes and costs for Case III. At a steam price of 143 SEK/MWh from the Case III CHP plant, the production cost for ethanol becomes 3.71 SEK/litre, where 1.09 SEK/litre is the average steam cost. The reason that the costs are lower for Case III than for Case I is that the waste heat from the ethanol plant substitutes some of the district heating produced in the CHP plant in Case I and II. Figure 13 shows the difference between incomes and costs for a stand-alone CHP plant and Case III as well as the production cost per litre for ethanol as function of the steam price.



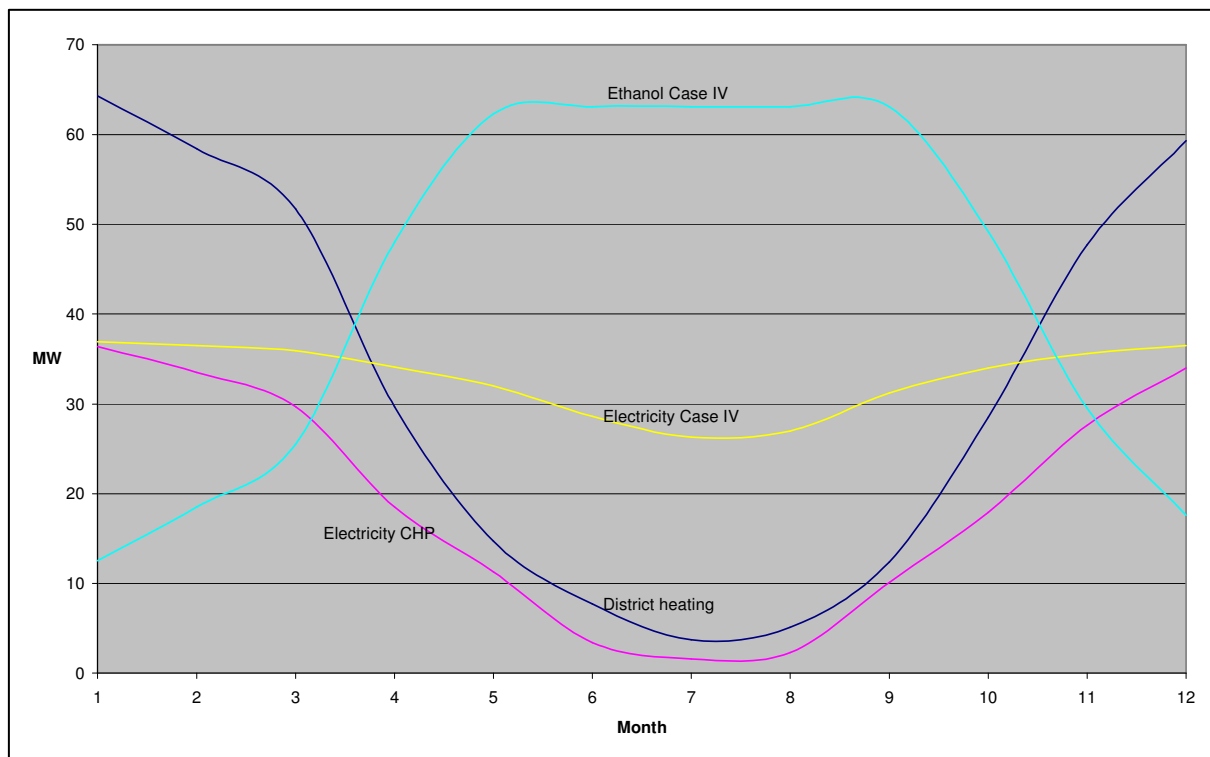
**Figure 13: Production costs for ethanol and differences between costs and incomes as a function of the steam price from the CHP plant of Case III and for a stand-alone CHP plant**

### 7.4 Results from Case IV

Figure 14 and Table 16 show the district heating and steam demand for the ethanol plant supplied by the Case IV plant over the year, as well as the electrical output for the integrated plant and the stand-alone CHP plant. The size of the steam boiler needed for the ethanol plant is 50.6 MW, which is used October-May. As for Case III the ethanol plant supplies the district heating net with 10 MW for all months except June, July and August.

**Table 16: District heating and steam to ethanol plant and electrical output for a stand-alone CHP plant and Case IV**

Month	District heating from CHP plant, MW	Electrical output, stand-alone CHP plant, MW	Electrical output, Case IV, MW	Steam to ethanol plant, Case IV, MW
Jan	64.3	36.4	36.9	12.5
Feb	58.4	33.5	36.5	18.5
Mar	51.7	29.7	35.9	25.5
Apr	29.7	18.5	34.1	48.0
May	14.7	11.3	32.0	62.3
Jun	7.7	3.4	28.6	63.1
Jul	3.7	1.6	26.3	63.1
Aug	5.1	2.3	27.0	63.1
Sep	12.4	10.1	31.2	63.1
Oct	28.5	17.9	34.0	49.2
Nov	47.7	27.6	35.6	29.5
Dec	59.3	34.0	36.5	17.6



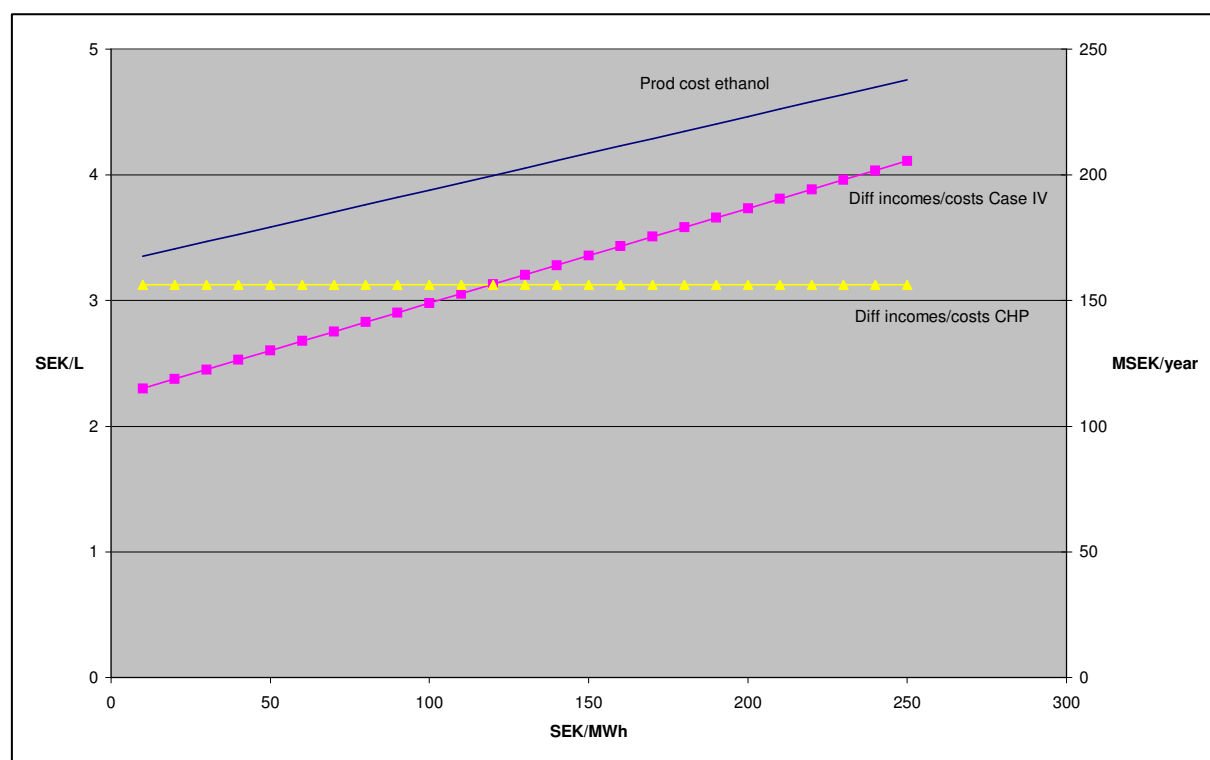
**Figure 14: Variations over the year of district heating, steam and electricity produced in the CHP plant of Case IV, compared with the electrical output from a stand-alone CHP plant**

As for Case III the district heating production for Case IV is less than for the CHP plant without extractions to an ethanol plant. But the total energy produced lies 70 % above the stand-alone CHP plant, which is somewhat better than Case III, as the steam demand for the Case IV ethanol plant is higher than for Case III.

**Table 17: Annual production of district heating, electricity, steam as well as fuel demand for a stand-alone CHP plant and for Case IV**

	Electricity, MWh	District heating, MWh	Steam, MWh	Fuel, MWh
<b>Case III</b>	288	279	284	981
<b>Stand-alone CHP plant</b>	162	339		505

Case IV gets a higher difference between incomes and costs than the stand-alone CHP plant when the steam price is 120 SEK/MWh or more. The production cost at that steam price is 4.23 SEK/litre, where the effective steam cost is 1.38 SEK/litre.



**Figure 15: Production costs for ethanol and differences between costs and incomes as a function of the steam price from the CHP plant of Case IV and for a stand-alone CHP plant**

## 7.5 Additional Simulations

Two additional simulations were carried out for the month of July in order to see the impact of using the secondary fuel steam from the fuel dryer within the ethanol plant instead of the CHP plant and the impact of integrating a larger ethanol plant. The ethanol plant produces a lignin fuel as co-product, which has a dry-matter content of 92 % after drying. One simulation was done for January, mixing the lignin fuel with wood.

### 7.5.1 Using the Secondary Fuel Steam within the Ethanol Plant

Table 18 shows the results from the simulation of the month of July, when the secondary fuel steam from the dryer is used within the distillation and evaporation processes (Case V) and when it is used for producing more steam to the low-pressure turbine. The condensate of 25 bar from the dryer is flashed down to 4 bar and is still connected to a low-pressure turbine, which produces 0.3 MW electric power compared with Case I, where the low-pressure turbine

connected to the dryer produces 0.5 MW. The total electric power decreases with 4.8 MW due to the decreased steam flow from the boiler. The secondary steam from the dryer is about 10.5 MW and when used within the distillation and evaporation the total heat demand for the ethanol plant is decreased from 47.7 MW to 37.2 MW.

*Table 18: Results from simulations for the month of July, Case I and Case V*

	Case V	Case I
District heating, MW	3.8	3.7
Electricity, MW	14.3	19.1
Distillation, MW	6.9	12.5
Evaporation, MW	6.0	10.9
Pretreatment, MW	12.9	12.9
Drying, MW	11.4	11.4
Fuel, MW	116.8	143.9
Steam flow, kg/s	22.2	28.4

### 7.5.2 Integrating a Larger Ethanol Plant to the CHP Plant

Table 19 shows the results from the simulation of the month of July, when a larger ethanol plant is integrated with the plant for combined heat and power production (Case VI). The annual ethanol production of the larger plant is supposed to be  $96.7 \cdot 10^6$  litres, compared with the smaller one that has a production of  $64.5 \cdot 10^6$  litres/year. The secondary fuel steam from the dryer is still used to produce more steam to the low-pressure turbine, which has an electrical output of 0.7 MW compared with 0.5 MW for Case I. The total electric power increases with 10.4 MW due to the increased steam flow. Increasing the production capacity from  $64.5 \cdot 10^6$  litres/year to  $96.7 \cdot 10^6$  litres increases the total heat demand for the ethanol plant from 47.7 MW to 71.6 MW.

*Table 19: Results from simulations for the month of July, Case I and Case VI*

	Case VI	Case I
District heating, MW	3.7	3.7
Electricity, MW	29.5	19.1
Distillation, MW	18.7	12.5
Evaporation, MW	16.4	10.9
Pretreatment, MW	19.4	12.9
Drying, MW	17.1	11.4
Fuel, MW	219.5	143.9
Steam flow, kg/s	45.6	28.4

### 7.5.3 Mixing Dry Lignin Fuel with Wood Fuel

One simulation was performed for the month of January, mixing the wood fuel with a dry-matter content of 50 wt-% and the lignin fuel with a dry-matter content of 92 wt-%. The total

amount of fuel needed when only wood was used was 66.9 ton/h and when the mixed fuel was used 52.5 ton/h, which means that 8.5 ton/h lignin (92 wt-% DM) substituted 22.6 ton/h wood (50 wt-% DM). If the costs are supposed to be 210 SEK/MWh for the lignin fuel and 140 SEK/MWh for the wood fuel, then the fuel costs for the CHP plant increases with about 2 MSEK for the month of January.

It is difficult to evaluate the impact of mixing the fuels, performing only one simulation. As mentioned in Chapter 7.7 the preheating of air could be improved in the process model. If the lignin fuel will be used in the CHP plant it will probably be used directly, without any drying. To get any profit of using dry lignin fuel in the CHP plant a higher amount will be necessary. If a dryer fuel is used the preheating of air will be less, which gives a lower fuel need. As the lignin fuel has a high content of dry matter it is better to pelletise it and sell it to external customers. The sulphur content in the lignin fuel is an important issue if it is sold to the private market.

## 7.6 Economic Evaluation of the Different Cases

For Case V (reusing the secondary fuel steam from the dryer and from the pretreatment within the ethanol plant) and Case VI (integration of a larger ethanol plant with the CHP plant) only one month was simulated. Based on the other simulations performed for Case I-IV the annual production of district heating, electricity and steam was estimated for these additional cases. Using the results from Case I-VI, the integration of a larger ethanol plant using both the secondary steam from the pretreatment and the dryer (Case VII) and using the waste heat for district heating (Case VIII), were estimated over the year as well. The waste heat is supposed to be 10 MW as for Case III and IV and was thus assumed to be independent of the plant size. The extra steam boiler needed for Case V-VIII and during which months they are used are shown in Table 20. All the cases are compared in Table 21 and 22.

*Table 20: Size of additional steam boiler for Case V-VIII and their respective usage*

	Steam boiler, MW	Usage over the year
Case V	35.4	November-March
Case VI	69.6	August-June
Case VII	54.1	September-May
Case VIII	44.1	October-April

Table 21 shows the results when the ethanol and the CHP plant have the same owner and different owners. When the integrated plant has the same owner the only steam paid for is the steam from the additional boiler, and the price of this steam was set to 250 SEK/MWh. The electricity price was set to 400 SEK/MWh and the district heating price to 479 SEK/MWh. The district heating production is 344 GWh for all the cases, but for Case III, IV and VIII 66 GWh is waste heat from the ethanol plant. The annual ethanol production is  $64.5 \cdot 10^6$  litres for Case I-V and  $96.7 \cdot 10^6$  litres for Case VI-VIII. The columns “1”, “2” and “3” show the production costs for ethanol when the integrated plant has the same owner (1), different owners (2) and the production costs for an ethanol plant which is not integrated with a CHP plant.

**Table 21: Comparison of Case I-VIII when the plants have the same owner (column “1”), different owners (column “2”) as well as production costs for a stand-alone ethanol plant (column “3”)**

	Electricity, GWh	Fuel, GWh	Steam CHP, GWh	Steam add boiler, GWh	1, SEK/L	2, SEK/L	3, SEK/L
<b>Case I</b>	272	921	267	151	3.68	3.86	4.72
<b>Case II</b>	295	982	312	241	4.03	4.21	5.24
<b>Case III</b>	258	898	318	99	3.49	3.71	4.23
<b>Case IV</b>	288	981	377	175	3.78	4.00	4.76
<b>Case V</b>	241	865	228	98	3.48	3.73	4.36
<b>Case VI</b>	300	1010	326	301	3.78	3.91	4.62
<b>Case VII</b>	284	960	295	195	3.50	3.63	4.27
<b>Case VIII</b>	278	945	348	142	3.37	3.50	3.94

Table 22 shows the minimum steam price for the different cases, when there are different owners, to get a larger difference between the incomes for electricity and heat (district heating and steam to the ethanol plant) than for the stand-alone CHP plant. For the integrated plant with same owners the ethanol price was varied in order to get a larger difference than for the stand-alone CHP, which is shown in Table 22 as well.

**Table 22: Steam price for different cases when there are different owners and ethanol price when the plants have the same owner**

	Steam price from CHP plant, SEK/MWh	Ethanol price, SEK/litre
<b>Case I</b>	43	3.86
<b>Case II</b>	36	4.21
<b>Case III</b>	143	3.71
<b>Case IV</b>	120	4.00
<b>Case V</b>	71	3.73
<b>Case VI</b>	40	3.91
<b>Case VII</b>	42	3.64
<b>Case VIII</b>	127	3.50

By using waste heat from the ethanol plant instead of producing it in the CHP plant, comparing for example Case I and III reduces the heat basis for the CHP plant. The increased steam production can not fully compensate for the lost district heating; furthermore the electricity production is reduced. However the CHP plant can compensate its lost income by increasing the steam price to the ethanol plant, and the production costs for ethanol will still be reduced as the waste heat is sold to the direct heating net. If the heat demand for the ethanol is higher it is more fortunate for the CHP plant as the heat and electricity production increase. But the ethanol production costs get higher, comparing Case I and II. Using the secondary steam from the steam dryer within the ethanol plant (Case V) is less fortunate for the CHP plant compared with Case I, but the production costs for the ethanol is reduced.

All the cases should be compared with a stand-alone CHP plant, which has an annual electricity production of 162 GWh and a district heating production of 339 GWh. Depending on which case the electricity production is increased with 49-85 % and the heat production with 80-98 % over the year.

### ***7.7 Evaluation of the Process Model***

Constructing and working with the process model for the integrated ethanol and CHP plant has been complicated, as it was hard to predict how the model should work for different flow rates of the steam. The electrical output from the CHP plant depends on the isentropic efficiency of the turbines and how different streams are mixed in the feed-water tanks. The isentropic efficiency as well as the outgoing steam pressure depends on the steam flow rate. To simplify the process model, the isentropic efficiency and the outgoing turbine pressure were set to be the same independently of the flow rate, which may have affected the results. Optimising the stream flows to the feed-water tanks in order to increase the electrical output could be done using a design specification or sensitivity analysis, but is complicated as there are four and five different streams respectively entering the two feed-water tanks in the model.

In a steam boiler the incoming air is preheated and the incoming feed-water heated to superheated steam by the flue gases. To get an effective combustion of wood the air should be preheated to 200-300°C, depending on the dry-matter content of wood. Wood with a dry-matter content of 50 wt-% was used in the process model for the boiler and the air was preheated to 250°C. To be able to find the optimal preheating temperature of air a design specification could be used to improve the model.

The steam boiler and the steam-turbine cycle are still separate models and the ethanol plant is only represented by four HEATER unit operation models. To complete the model the steam boiler and the steam-turbine process models should be connected to a model of the whole ethanol plant.

## 8. Conclusions

The results from the simulations completed with an economic evaluation indicate that integrating an ethanol plant with a CHP plant is profitable for both plants. The ethanol plant gets access to cheap energy and the electricity and heat production is increased for the CHP plant. In the calculations the stand-alone CHP plant was supposed to be closed down during 1.5-2 months in the summer, while the integrated plant was supposed to be run during the whole year. The integrated plant has to be closed down during a certain time as well for maintenance. The main reason that the CHP plant can produce more electricity and heat is not the running hours; it is the increased steam flow during April-October, when the district heating need is decreased. For Case I the ethanol plant reduces its production costs from 4.72 to 3.86 SEK/litre and the CHP plant increases the electricity production from 162 to 272 GWh and its heat production (district heating and steam to the ethanol plant) from 339 to 611 GWh per year.

To keep the production costs down the ethanol production process should be as effective as possible. The secondary steams from the pretreatment and the steam dryer should be reused within the ethanol process and the waste heat from the ethanol plant, if any, should be used for district heating purposes. Even though the possibilities for the CHP to produce heat and electricity are reduced, the energy production is still higher than for a stand-alone CHP plant. The possibilities of using waste heat from the ethanol plant should be investigated more thoroughly, see down below. It is preferable to sell the lignin fuel rather than use it in the steam boiler of the CHP plant.

The main purpose of the thesis project was to construct a process model to be able to evaluate the impact of integrating an ethanol plant with a CHP plant. But the aim was also to construct a model that could be used for integrated plants with other conditions, such as different sizes of the ethanol plant and/or the CHP plant, different pressures and temperatures of the superheated steam from the steam boiler etc. Regarding the constructed process models there are improvements to be made, but I consider that the two separate process models for the steam boiler and the steam cycle are good enough to be used for simulations of other integrated plants.

### 8. 1 Future Studies

As mentioned above the ethanol production process should be as effective as possible. The evaporation process step could be substituted by an anaerobic cleaning process, where biogas is produced. Compared with Case I, which has a heat demand of about 48 MW, the total heat demand could be reduced to 20 MW (Article IV, 2). Etek, among others, is developing and improving the technology for producing ethanol from lignocellulose.

In this project work, simulations were performed where the waste heat from the ethanol plant were supposed to contribute 10 MW. Studies should be done, verifying the quality of this waste heat as well as the possibilities of distributing it to the district heating net.

The electric power production in a CHP plant depends on, among other things, at what pressure extractions are done. By reducing the pressure for the steam dryer and some of the steam used for the pretreatment, the electrical output may be improved. The effects of reducing the pressure should be investigated.

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