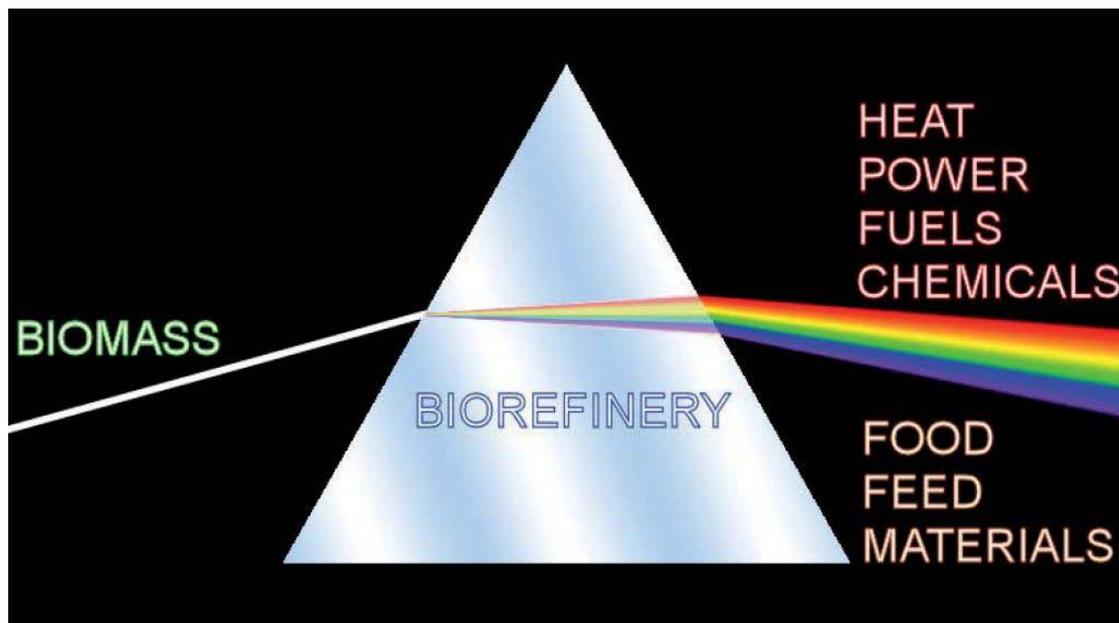


# CHALMERS



## Opportunities for Process Integrated Biorefinery Concepts in the Chemical Cluster in Stenungsund

Research Project Report

**ROMAN HACKL AND SIMON HARVEY**

Department of Energy and Environment

*Division of Heat and Power Technology*

CHALMERS UNIVERSITY OF TECHNOLOGY



PROJECT REPORT

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Cover:

Biorefinery: The sustainable processing of biomass into a spectrum of marketable  
products and energy (de Jong, van Ree, and Kwant 2009)

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## Summary

The energy and material needs of human society are increasing while at the same time fossil resources decline. Greenhouse gas (GHG) emissions are influencing the world's climate. The potential for substituting fossil hydrocarbons in synthetic products and liquid fuels by renewable raw materials is being discussed in order to fight climate change and decrease dependency on fossil resources. The biorefinery concept is a way to accomplish this transition. A wide range of renewable raw materials can be converted into value added products and therefore substitute fossil feedstocks.

High efficiency is very important in order to profitably implement biorefinery concepts. The interest for energy combines and eco-industrial parks is increasing nowadays as they offer the opportunity to exchange materials and energy between two or more industries and also the society. Therefore integration of biorefinery concepts into industrial cluster can be advantageous.

In this study suitable biorefinery concepts are identified and analysed with respect to integration opportunities in Sweden's largest chemical cluster in Stenungsund. Technical, economical and environmental consequences of integrating a biorefinery in the cluster compared to stand-alone operation are identified based on mass and energy balances, knowledge on the current energy situation in the cluster and the thermal characteristics of the different biorefineries.

Suitable biorefinery concepts for integration in the cluster include biomass gasification for syngas production, lignocellulosic ethanol production for conversion into ethylene and low temperature biomass drying for fuel upgrading. The current demand of steam produced in the cluster's boilers is 122 MW at pressure levels between 85 and 1 bar(g). Excess steam from a gasification unit with an assumed operation time of 8000 h/yr can be used for cogeneration to cover parts of this demand. By integration of a gasification unit producing 160 kt<sub>product</sub> gas/yr, 16 GWh<sub>el</sub>/yr and 128 GWh<sub>steam</sub>/yr can be delivered to the cluster. For a stand-alone unit it is assumed that all excess steam is used for electricity production in a condensing turbine, producing 47.4 GWh<sub>el</sub>/yr. This results in increased incomes between 18.3 and 47.4 MSEK/yr in the integrated case. CO<sub>2</sub> emissions reduction is 24.4 kt<sub>CO<sub>2</sub></sub>/yr higher with integration.

Ethanol production from lignocellulosic raw material yields substantial amounts of residual products which can be used for heat and power generation to cover parts of the clusters current energy demand and/or deliver heat and electricity to a downstream ethanol-to-ethylene dehydration plant. The results are obtained for a process that produces 100 kt ethylene/yr and has an operating time of 8000 h/yr. A lignocellulosic ethanol plant producing the feedstock (174 kt ethanol/yr) to an ethanol-to-ethylene plant has an energy surplus of 195.2 GWh/yr when all residues are combusted. In an integrated plant this yields 21.8 GWh<sub>el</sub>/yr and 168 GWh<sub>steam</sub>/yr to the cluster and/or the ethanol-to-ethylene plant, while in stand-alone operation (only production of electricity from excess steam) 64.3 GWh<sub>el</sub>/yr can be produced. Incomes by integration

are between 24.8 and 64.2 MSEK/yr higher and CO<sub>2</sub> emissions reduction is increased by 31.2 kt/yr by integration.

An improved utility system for maximum energy recovery was developed in a previous total site analysis (TSA) study. The residual waste heat is 498 MW at 99 °C to 27 °C. Utilising this heat for low temperature drying of biomass was compared to stand-alone dryer operation. This gave a total potential of  $4.3 \cdot 10^6$  tonnes dried biomass per year (15 wt-% moisture content). By integration 129 SEK/t<sub>dry mass</sub> less fuel costs and 234 kg/t<sub>dry mass</sub> less CO<sub>2</sub> emissions were found.

**Keywords:** Biorefinery, Chemical cluster, Process integration, Gasification, Bio-ethanol to ethylene, Biomass drying, Total Site Analysis.

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## **Preface**

This project was conducted at the Division of Heat and Power Technology at Chalmers University of Technology. Financing of the project was provided by the Swedish Energy Agency (Process Integration programme, project number P 32222-1). The project was conducted between June 16<sup>th</sup> 2009 and Sept 30, 2010 with Simon Harvey (Professor of Industrial Energy Systems at Chalmers University of Technology) as project leader. The study investigates opportunities for integration of different biorefinery concepts in the chemical cluster in Stenungsund.

We'd like to thank all the participants for their support and the Swedish Energy Agency for the financing of this project.

# Notations

## Abbreviations:

<b>ASU</b>	Air Separation Unit
<b>CC</b>	Composite Curves
<b>CCS</b>	Carbon Capture and Storage
<b>CFB</b>	Circulating Fluidised Bed
<b>CW</b>	Cooling Water
<b>DME</b>	Dimethylether
<b>EDC</b>	Ethane dichloride
<b>EMU</b>	Emulgol
<b>EO</b>	Ethylene oxide
<b>ETBE</b>	Ethyl Tertiary Butyl Ether
<b>EtOH</b>	Ethanol
<b>FTD</b>	Fischer-Tropsch Diesel
<b>GCC</b>	Grand Composite Curve
<b>GHG</b>	Greenhouse Gas
<b>HCl</b>	Hydrogen chloride
<b>HDPE</b>	High Density Polyethylene
<b>HP</b>	High Pressure steam
<b>HPPE</b>	High Pressure Polyethylene
<b>IEA</b>	International Energy Agency
<b>LDPE</b>	Low Density Polyethylene
<b>LLDPE</b>	Linear Low Density Polyethylene
<b>LP</b>	Low Pressure steam
<b>LPPE</b>	Low Pressure Polyethylene
<b>MEA</b>	Monoethanolamine
<b>MP</b>	Medium Pressure steam
<b>MSW</b>	Municipal Solid Waste
<b>MTO</b>	Methanol To Olefins
<b>MTP</b>	Methanol To Propylene
<b>NaOH</b>	Sodium hydroxide
<b>PE</b>	Polyethylene
<b>PVC</b>	Polyvinylchloride
<b>RME</b>	Rapeseed oil methyl ester
<b>SCN</b>	Steam Cracked Naphtha
<b>SEK</b>	Swedish Krona
<b>SSF</b>	Simultaneous Saccharification and Fermentation
<b>SSSP</b>	Site Source Sink Profiles
<b>STF</b>	Speciality Surfactants
<b>TSA</b>	Total Site Analysis
<b>TSC</b>	Total Site Composites
<b>TSP</b>	Total Site Profiles
<b>VCM</b>	Vinyl chloride

**Symbols:**

<b>Q</b>	Heat load
<b>Q<sub>cooling,min</sub></b>	Minimum cooling demand
<b>Q<sub>cooling,total</sub></b>	Overall cooling demand
<b>Q<sub>heating,min</sub></b>	Minimum heating demand
<b>Q<sub>heating,total</sub></b>	Overall heating demand
<b>Q<sub>rec</sub></b>	Heat recovered
<b>T<sub>start</sub></b>	Stream starting temperature
<b>T<sub>target</sub></b>	Stream target temperature
<b>ΔT<sub>min</sub></b>	Minimum temperature difference



# 1 Introduction

The energy and material needs of human society are increasing while at the same time fossil resources are in decline. Greenhouse gas (GHG) emissions are influencing the world's climate. The global demand for fossil fuel resources has to be reduced in order to mitigate climate change and to become less dependent on non-renewable resources. The biorefinery concept is a way to substitute fossil hydrocarbons in synthetic products and liquid fuels by renewable raw materials. A wide range of renewable raw materials can be converted into value added products and therefore substitute fossil feedstocks. The figure on the cover of this report illustrates this conversion. According to (Okkerse and Bekkum 1999) all the existing organic chemistry which is mainly based on fossil fuels can be replaced by biomass conversion into organic raw materials.

High efficiency is very important in order to profitably implement biorefinery concepts. Therefore their integration into an existing industrial cluster can be advantageous for many reasons, such as:

- Integration provides the opportunity to exchange heat which otherwise has to be discharged
- Biorefinery products can be delivered to a neighbouring site and used as feedstock, thereby avoiding long distance transportation
- Existing infrastructure can be used such as boilers, pipeline alignments, storage facilities etc.

In summary, the systematic integration of a biorefinery into an existing industrial cluster offers the opportunity to reach the goal of an eco-industrial park where companies and the local community among others share materials, energy and infrastructure to improve the economical and environmental situation of the total site.

## 2 Objective

This aim of this project is to identify interesting biorefinery concepts for integration within the chemical cluster in Stenungsund. An important input to this study is information about current and future energy systems of the companies involved already analysed in a project called “Energieeffektivisering i Stenungsundsklustret” (P 32223-1) (Hackl and Andersson 2010). There are a large number of different possible biorefinery concepts and a detailed analysis is necessary in order to identify the alternatives which give the best overall performance in combination with the cluster’s existing infrastructure.

The methodology together with goals of the study are presented below. In a first step a structured methodology is used in order to select suitable biorefinery concepts. Because of the large number of potential biorefinery routes the following two steps are conducted:

- qualified pre-selection of suitable biorefineries based on product and raw material flows existing in the cluster and
- general assumptions on the nature of the biorefinery processes (process has heat excess or deficit, process generates/consumes high or low temperature heat)

This is followed by the

- determination of general mass and energy balances for the selected biorefinery concepts
- determination of the biorefinery process source and sink temperature profiles relevant for exchange of heat with other processes
- assessment of energy integration potential of the selected biorefineries with the cluster in Stenungsund based on the results from the previously conducted Total Site Analysis (TSA) study
- assessment of consequences of biorefinery integration on CO<sub>2</sub> emission and profitability with help of “Scenarios for assessing profitability and carbon balances of energy investments in industry” (Harvey and Axelsson 2010)

## 3 Biorefinery

### 3.1 Definition

The role of biorefineries is according to the definition of the International Energy Agency (IEA) “the sustainable processing of biomass into a spectrum of marketable products and energy”. In other words, biomass is sustainably used as a resource for production of among others heat, power, fuels, chemicals and materials (de Jong, van Ree, and Kwant 2009).

The raw materials that can be used as feedstock for a biorefinery are, for example crops and residues, lignocellulosic material, municipal solid waste (MSW) and algae. They are provided by four different sectors: agriculture, forestry, industries and aquaculture. Biorefineries apply a wide range of technologies to separate the biomass inputs into their building blocks, such as hydrogen, carbohydrates and proteins, which then are converted into value added products (Cherubini 2010). Figure 1 gives an overview of different biorefinery routes showing the feedstocks, products and the technologies used to convert the raw materials and building blocks.

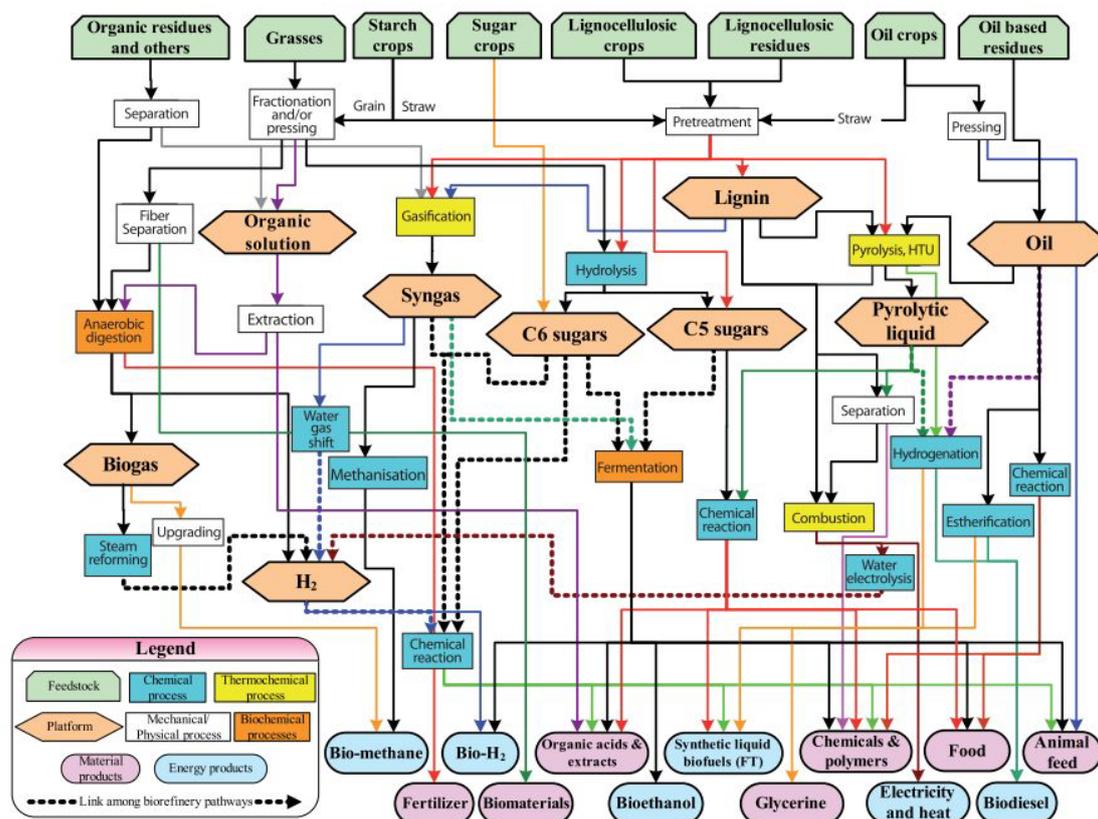


Figure 1 Overview of possible raw materials, technologies, building blocks and products used and produced in biorefinery (de Jong, van Ree, and Kwant 2009)

### 3.2 Processes

There are four main groups of processes applied in a biorefinery. Namely thermochemical, biochemical, mechanical and chemical processes. In order to separate the biomass input to the biorefinery into its constituting elements and convert

these into valuable products, different combinations of processes can be used. In the following the processes are described. Figure 1 shows the connection between material streams and processes.

### Mechanical processing

Mechanical size reduction of the incoming raw biomass is often the first processing step in a biorefinery. This is done by cutting and other processes resulting in a significant change in particle size, shape and bulk density. Other mechanical processes aim to separate and/or concentrate the substrates, intermediates or final products into their components. Examples for this are absorption and extraction processes (Huang et al. 2008).

### Thermochemical processing

This group mainly consists of three types of processes: gasification, pyrolysis and direct combustion. The first to yield energy and chemicals as products, while in direct combustion biomass is oxidized in order to produce heat and electricity. Biomass gasification occurs at high temperatures ( $>700^{\circ}\text{C}$ ) and in low oxygen environment. The raw material is converted into so-called syngas, consisting mainly of  $\text{H}_2$ ,  $\text{CO}$ ,  $\text{CO}_2$  and  $\text{CH}_4$ . This can be further processed into fuel or chemicals or used directly for heat and electricity production. By pyrolysis the biomass feedstock is converted into so-called pyrolysis oil, charcoal and a gaseous phase similar to syngas. The process operates at lower temperature than gasification ( $300\text{-}600^{\circ}\text{C}$ ) in an oxygen free environment. At the moment the pyrolysis oil and char can be used as fuel in stationary combustion operations, but processes to utilise the pyrolysis oil as e.g. transportation fuel are under development (Cherubini 2010).

### Biochemical processing

The main two processes in this group are fermentation and anaerobic digestion. Fermentation yields mainly alcohols and organic acids. Microorganisms or enzymes convert fermentable biomass components into products. Depending on the raw material different pre-treatment and special microorganisms are necessary (Kaparaju et al. 2009). The main product from anaerobic digestion is biogas. It consists depending on the substrate of 50-65 %  $\text{CH}_4$ , 35-50%  $\text{CO}_2$  and minor impurities. Typical raw materials are sewage, landfill waste or waste from the food industry. The raw biogas can be upgraded to biomethane by removing  $\text{CO}_2$  among others. Biomethane then can be delivered to the natural gas grid or used as vehicle fuel (Bruijstens et al. 2008).

### Chemical processing

This group contains a wide range of chemical reactions which change the chemical composition of the input materials. Hydrolysis and transesterification are common reactions applied in biorefineries, e.g. to depolymerise polysaccharides and proteins into their building blocks or to upgrade vegetable oils to biodiesel. Other examples for chemical processes are the conversion of syngas into  $\text{CH}_4$  (methanisation) (Gassner and Maréchal 2009), methanol production from syngas or Fisher-Tropsch synthesis to produce fuels (Gassner and Maréchal 2009; Naik 2010).

### 3.3 The role of industrial clusters

In order to become economically competitive a major challenge for biorefineries is to maximise their overall conversion efficiency (Bludowsky and Agar 2009). The interest for energy combines and eco-industrial parks is increasing nowadays as they offer the opportunity to exchange energy flows between industrial plant or with the surrounding energy system. Heat integration can help to increase overall thermal efficiency by distributing heat between sources of excess heat and heat consumers (T. Berntsson et al. 2006). Integrating a biorefinery in an industrial cluster can be advantageous from an energy point of view as the cluster can serve as both a source of excess heat to the biorefinery process (e.g. deliver low grade heat for biomass drying) and/or a sink for excess heat from the biorefinery (e.g. excess steam from biomass gasification).

Another advantage of integration into an industrial cluster is that the existing infrastructure (boilers, utility systems, air separation etc.) is already in place. Compared to a stand-alone biorefinery unit this can have positive impact on the process economics.

Biorefineries can produce a wide range of products including both final and intermediate products which can be used as feedstock to other chemical processes. Thereby it is possible to integrate a suitable biorefinery in an existing chemical cluster which produces the starting materials for the existing chemical processes. Among others e.g. ethanol from a lignocellulosic ethanol production plant can be converted into ethylene and used as feed to several processes at the chemical cluster in Stenungsund. Another example is a biomass gasification unit which can deliver syngas and steam to the cluster.

These advantages can be summarised by the definition of an eco-industrial park (EIP) which is “a community of businesses that cooperate with each other and with the local community to efficiently share resources such as information, materials, water, energy, infrastructure and natural habitat, leading to economic and environmental quality gains, and equitable enhancement of human resources for the business and local community” (Spitzer 1997).

In summary, a biorefinery integrated in an industrial process cluster can have the following advantages:

- Making use of existing infrastructure
- Serve as a user of excess heat (e.g. biomass drying)
- Serve as a heat source to partly cover the cluster's heat demand by excess heat from the biorefinery (e.g. excess steam from a gasification unit)
- Offer possibilities for direct heat integration
- Make use of biorefinery products that can be used directly as feedstock elsewhere in the cluster (syngas, ethanol etc.)
- Capitalize on existing knowhow on operation of chemical process

## 4 The chemical cluster in Stenungsund

In the following the cluster is described more in detail. The main raw materials and products used and produced by the different companies are shown, and an overview of the clusters energy generation and distribution systems is given. For a more detailed description of the processes and the energy systems the report on the total site analysis performed in 2009-2010 should be consulted (Hackl and Andersson 2010).

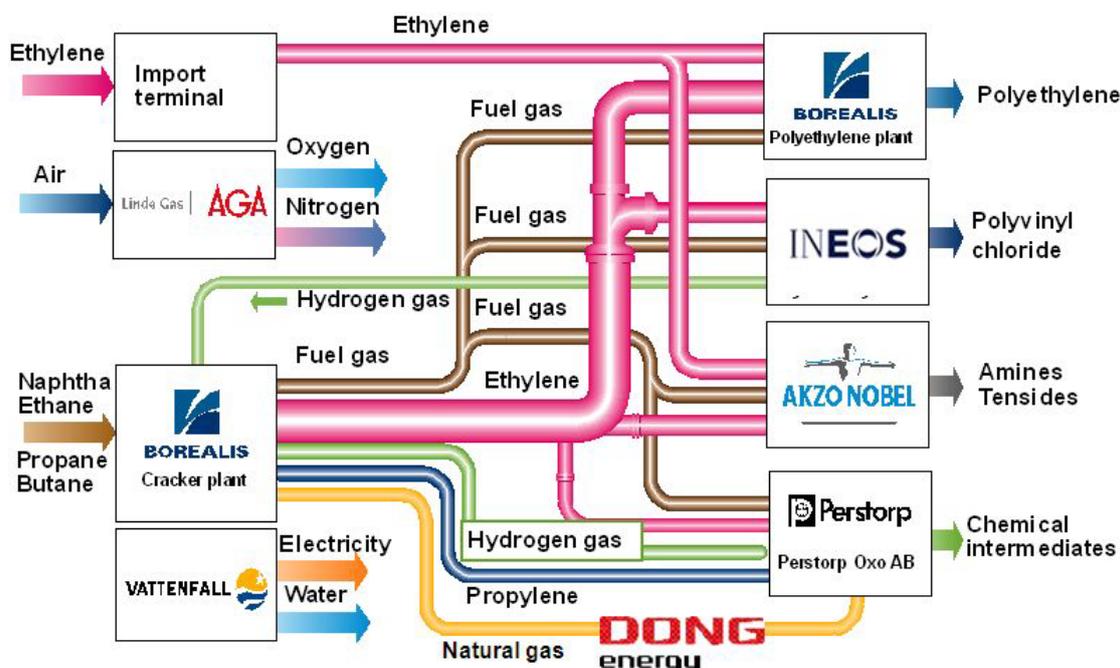


Figure 2 Major material and energy flows in the Stenungsund chemical cluster (Borealis AB 2009)

In this report advantages of integrating a biorefinery in the chemical cluster in Stenungsund, Sweden are investigated. The cluster is Sweden's largest agglomeration of its kind. The main companies involved are AGA Gas AB, Akzo Nobel Sverige AB, Borealis AB, INEOS Sverige AB and Perstorp Oxo AB. The heart of the cluster is a steam cracker plant run by Borealis. Figure 2 shows the material flows between the different companies and plants.

As can be seen in Figure 2 the companies already interact strongly with each other in terms of material exchange.

### 4.1 The cluster's processes, raw materials and products

**AGA Gas** is a manufacturer of industrial gases and operates a cryogenic air separation plant in Stenungsund. The plant produces oxygen ( $O_2$ ), nitrogen ( $N_2$ ), carbon dioxide ( $CO_2$ ) and argon (Ar) for the local companies and for sale on the market.

Table 1 Annual consumption and production of raw materials and products at AGA Gas in Stenungsund (Arvidsson 2010)

Raw materials	Annual consumption [kt/yr]
Air	990 <sup>1</sup>
Products and intermediates	Annual production [kt/yr]
O <sub>2</sub>	190 <sup>1</sup>
N <sub>2</sub>	116.5 <sup>1</sup>
CO <sub>2</sub>	25.6 <sup>1</sup>
Ar	8.3 <sup>1</sup>

**AkzoNobels'** site in Stenungsund consists of three plants. The heart of the site is the ethylene-oxide (EO) plant. EO is used as a raw material in the production of among others ethyloxylates, cellulose derivatives, ethanol- and ethylene amines. It is produced by partial oxidation of ethylene. In the surfactants plant surface agents are produced, which are used in e.g. disinfectants, textile softeners and detergents. Two units, the emulgol (EMU) and speciality surfactants (STF) unit, manufacture over 300 different end-products. The amine plant consists of two main processes. In one process ethanol amines are produced by a reaction between EO and ammonia.

Table 2 Annual consumption and production of raw materials and products at AkzoNobel in Stenungsund (Ekeröth 2010)

Raw materials	Annual consumption [kt/yr]
Ammonia	25
Ethylene	70
O <sub>2</sub>	80
Products and intermediates	Annual production [kt/yr]
CO <sub>2</sub>	60
Ethylene Oxide (EO)	85
Amines	80
Glycols	5

Monoethanolamine (MEA) is further converted with ammonia and hydrogen in a catalytic reaction under high pressure and high temperature into ethylene amines in the ethylene amine process.

**Borealis** is a supplier of plastic material for wires and cables, pipes, automotive and advanced packaging. In Stenungsund two Borealis sites are in operation. A cracker plant produces olefins and three polyethylene (PE) plants produce different qualities of PE used primarily for pipes, wires and cables. The cracker plant provides the feedstock to the Borealis PE plant, and to other companies present in the Stenungsund chemical cluster.

<sup>1</sup> Calculated from volume flows,  $\rho_{\text{air}}=1.25 \text{ kg/Nm}^3$ ,  $\rho_{\text{O}_2}=1.33 \text{ kg/Nm}^3$ ,  $\rho_{\text{N}_2}=1.165 \text{ kg/Nm}^3$ ,  $\rho_{\text{CO}_2}=1.83 \text{ kg/Nm}^3$ ,  $\rho_{\text{Ar}}=1.66 \text{ kg/Nm}^3$ , not the whole amount of the different air components is separated, e.g. most of the incoming nitrogen is sent back to the atmosphere, while 95 vol-% of the incoming oxygen is utilised.

Table 3 Annual consumption and production of raw materials and products at Borealis cracker plant in Stenungsund (Borealis AB 2009)

Raw materials	Annual consumption [kt/yr]
Naphtha	298
Ethane	321
Propane	286
Butane	268
Ethanol	7
Off-gas	3
Products and intermediates	Annual production [kt/yr]
Ethylene	565
Propylene	174
Other (incl. fuel gas)	424
ETBE	16

Table 4 Annual consumption and production of raw materials and products at Borealis PE-plant in Stenungsund (Borealis AB 2009)

Raw materials	Annual consumption [kt/yr]
Ethylene	444
Propylene	0.132
Co monomer	8.57
Additives	4.3
Master batch	17.3
Products and intermediates	Annual production [kt/yr]
High Pressure PE (LDPE), to be phased out in 2010	150
Low Pressure PE (HDPE)	93
PE3 Borstar, (HDPE)	206
New HPPE plant (LDPE), future production	350 <sup>2</sup>

**INEOS ChlorVinyls** in Stenungsund produces chlorine, sodium hydroxide (NaOH), hydrochloric acid (HCl), hydrogen, ethane dichloride (EDC), vinyl chloride (VCM) and polyvinylchloride (PVC) in three different plants. Table 5 shows the production of the different products in 2008. A chlorine plant converts sodium chlorine into chlorine, hydrogen and sodium hydroxide in an electrolytic amalgam-process. The chlorine is used as feed together with ethylene in a vinyl chloride plant to produce EDC and VCM. VCM is the monomer of PVC. The polymerisation takes place in batch reactors.

<sup>2</sup> Annual capacity, after the new HPPE plant is taken into operation and the old HPPE plant is phased out, the ethylene consumption will increase by app. 200 kt/yr. This is not taken into account in the table. Parts of the ethylene will be imported.

Table 5 Annual consumption and production of raw materials and products at INEOS in Stenungsund (Olsson 2010; Josefsson 2009)

Raw materials	Annual consumption [kt/yr]
Ethylene	79
NaCl	199
VCM	198
Cl <sub>2</sub>	115
Products and intermediates	Annual production [kt/yr]
NaOH	131
EDC	21
HCl	29
VCM	156
PVC	178
H <sub>2</sub> (sold)	0.3
Cl <sub>2</sub>	112

**Perstorp Oxo AB** is a manufacturer of speciality chemicals with production sites in Nol and Stenungsund. Synthesis gas is an important building block in processes at Perstorp Oxo which is produced in a methane steam reformer. Several plants produce a large variety of products. Table 6 shows annual consumption and production of the main raw materials, products and intermediates.

Table 6 Annual consumption and production of raw materials and products at INEOS in Stenungsund (Pollard 2010)

Raw materials	Annual consumption [kt/yr]
H <sub>2</sub>	5.7
Methane	95 <sup>3</sup>
Propylene	190
Ethylene	24
Oxygen	145
Rape seed oil	125
Methanol	14
Orthoxylene	10
Products and intermediates	Annual production [kt/yr]
Syngas (H <sub>2</sub> /CO = 1)	160 <sup>4</sup>
RME	125
Iso- and n-butyraldehyde	300
Octanoic acid	65
Propionic acid	50
2-ethylhexanol	75
DOP (dioctylphthalate)	25
Butanol	60

## 4.2 The cluster's current utility systems

In this section a brief overview of the cluster's utility systems is provided.

<sup>3</sup> all methane is used for syngas production

<sup>4</sup> Syngas purity = 99.8 %

Each of the five companies (except AGA Gas) has their own utility generation. Hot utilities used in the cluster are steam, fuel gas, hot oil and hot water. Cold utilities used are cooling water (CW), air, refrigerants and condensate for steam generation.

#### 4.2.1 Hot utility generation

Table 7 gives an overview of hot utility generation (steam, flue gases, hot oil and hot water) currently in operation in the cluster. It is also shown how the hot utility is generated (from recovered process heat, by-product incineration or by added fuel combustion). It can be seen that most of the hot utility is produced from excess process heat (280 MW<sub>hot utility</sub>). In order to satisfy the additional demand fuel is burned in steam boilers to produce 122 MW<sub>hot utility</sub>. Also 40 MW<sub>steam</sub> is produced by the incineration of by-products that can't be used in a better way and is therefore considered as utility from excess process heat in the following.

Table 7 Hot utility production in the chemical cluster in Stenungsund (Hackl and Andersson 2010)

Site	From excess process heat	From by-product incineration	From added fuel	Total hot utility generation
AGA			0	0
Akzo Nobel	35	6.4	25	66.4
Borealis Cr	159		35	194
Borealis PE	35	4	2	41
INEOS	14	2.6	33	47
Perstorp	39	27	27	93
Sum	280	40	122	442

#### 4.2.2 Excess and deficit of steam at different levels in the cluster

Figure 3 shows the steam deficit or excess at different steam levels and companies in the cluster. It is constructed from data gathered on recovered steam from process heat and steam demands at the clusters different steam pressure levels. The processes in the cluster have a high cooling demand. A significant amount of heat is recovered as steam and used for process heating. Heat recovery and use occurs often not at the same steam level, e.g. at Borealis a large amount of steam is recovered by cooling of the cracking products at 85 bar(g). At this pressure level there is only a small amount of heating demand at the cracker. Therefore the steam is expanded in turbines or throttle valves to a steam level where it is needed. The steam systems of the different companies are not connected and therefore excess steam from one company cannot be transferred to another. The diagram in Figure 3 does not include steam produced in the boilers. Therefore it can be used to determine the steam levels and the amount of steam which has to be covered by steam from the boilers or other sources, e.g. a biorefinery

It can be seen that above 40 bar(g) only Borealis and Perstorp have an excess of steam. Currently this steam is used to satisfy the demand at lower pressures. Akzo and INEOS have a deficit of steam at 40 bar(g) and below which currently is satisfied by steam from their boilers. Therefore a biorefinery which has an excess of steam could deliver it to the cluster at 40 bar(g)

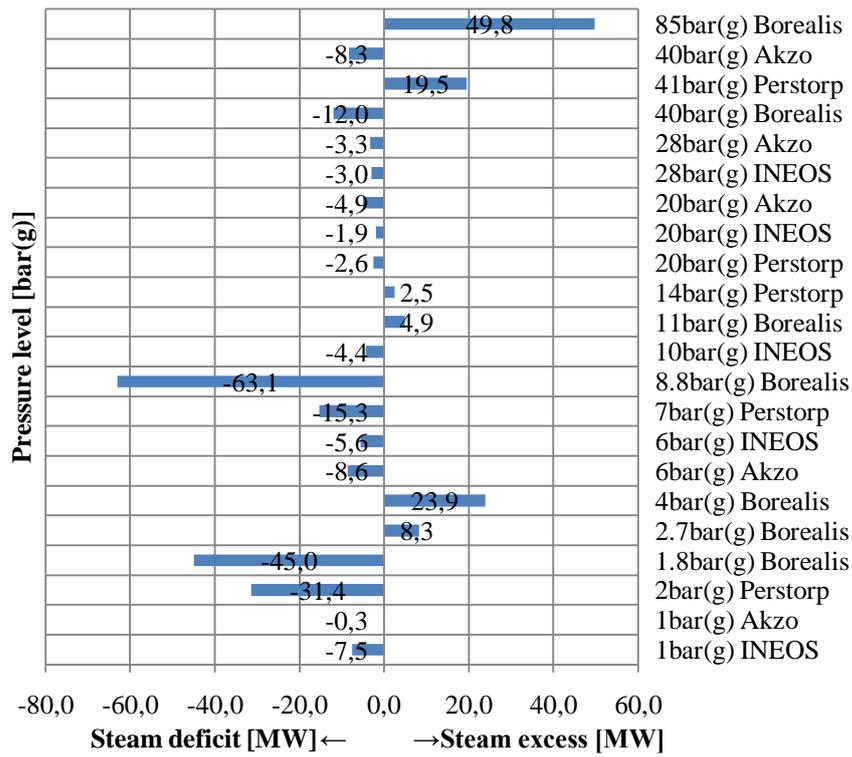


Figure 3 Representation of the steam excess/deficit at the total site pressure levels at the different plants (Steam produced in boilers by added fuel is not included) (Hackl and Andersson 2010)

## 5 Methodology

The study is performed in three steps:

- Qualitative selection of suitable biorefineries
- Determination of energy integration potential for the selected biorefinery concepts using pinch analysis and Total Site Analysis (TSA)
- Evaluation of economic performance and CO<sub>2</sub> emissions consequences of integration of the selected biorefinery concepts

The different steps are described more in detail in the following.

### 5.1 Identification of suitable biorefinery concepts

Due to the large number of potential biorefinery process concepts a first screening study was conducted in order to identify biorefinery concepts that are of particular interest for integration within the chemical cluster in Stenungsund. This selection was conducted in cooperation with the chemical companies situated in Stenungsund and based on literature research and data of a total site study of the chemical cluster (Hackl and Andersson 2010). The following criteria were considered in the selection:

- Material integration: The integration of material flows within the cluster is considered when selecting potential biorefineries. E.g. biorefineries yielding products which can be used as feedstock in existing processes in the cluster are preferred.
- Company interests: Several companies are currently investigating potential implementation of biorefineries in their processes. A site-wide view on the processes of interest is taken in this study.

Based on an overview of different biorefinery routes, suitable processes and products were determined. Biorefinery processes that are suitable for integration with the cluster were identified based on the following considerations:

- Biorefinery concepts with products that are currently used as feedstocks for different processes within the cluster, since these chemicals can be directly fed into existing processes. This type of biorefinery is of particular interest if a fossil feedstock previously imported to the cluster can be replaced by a biomass-based feedstock that is pre-processed and directly used at the cluster site.
- Biorefinery concepts with products that are currently produced and sold by the companies, since the existing product handling and sales structure can be used.
- Biorefinery concepts with potential for significant heat integration potential with the cluster, e.g. processes that consume low grade waste heat from the cluster or deliver steam.

Figure 4 illustrates the material and energy flows in the cluster and potential integration of biorefineries. Streams that can be replaced/complemented or utilised by integration of a biorefinery are also shown. As described above suitable biorefineries should ideally replace feedstock and/or heat and electricity input to the cluster and/or utilise excess heat from the cluster. Products of the biorefinery can even be used to

substitute intermediate products. By systematically replacing fossil-based feedstocks and intermediate products with renewable ones the share of “green”-products in the cluster can be increased.

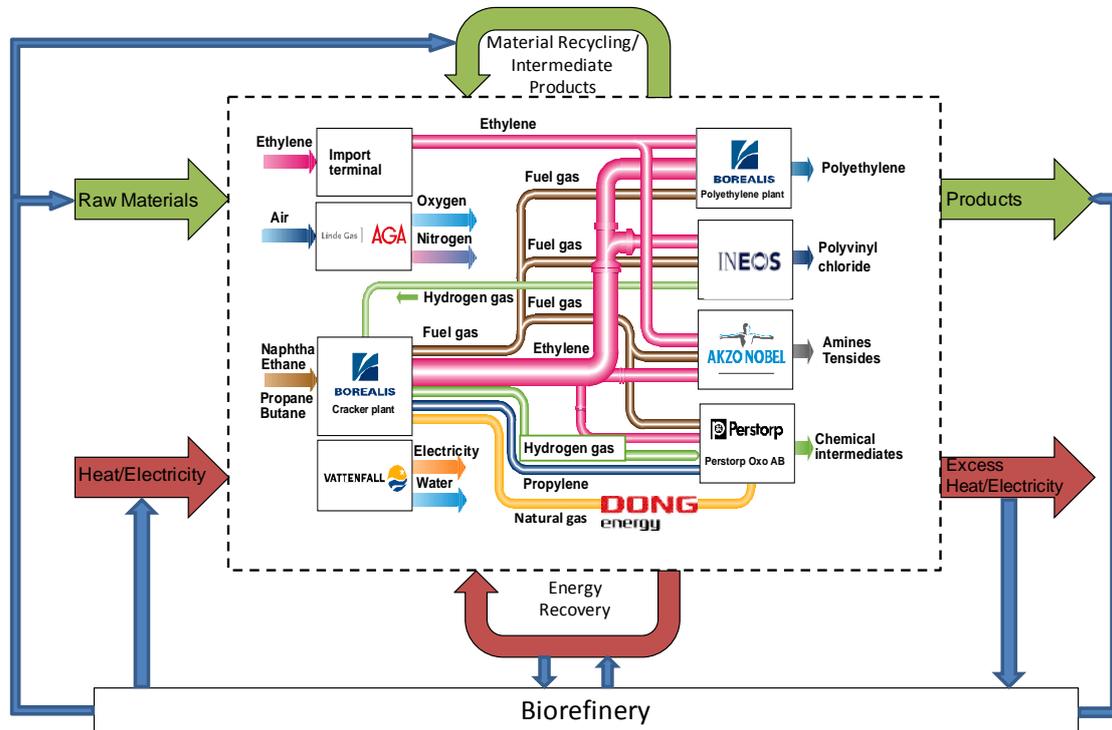


Figure 4 Illustration of material and energy flows within the chemical cluster and potential advantageous integration of biorefineries

In order to pre-select suitable biorefinery processes primary and secondary material outputs from a large number of biorefineries are compared to the raw materials and products used and produced in the cluster.

## 5.2 Total Site Analysis (TSA)

TSA is used to identify opportunities for heat recovery by integration of the individual heating and cooling demands of different processes at a total site based on Pinch Technology (Kemp 2007). Excess heat from one process plant can be transferred to a common utility (e.g. steam, hot water, hot oil) (Bagajewicz and Rodera 2001) and then delivered to processes with a heat deficit by the common utility system. The TSA method enables the amounts of hot utility generated and used by the combined individual processes, the amount of heat recovery in a common hot utility system, the steam demand from the boilers and the cogeneration potential to be determined (Perry, Klemes, and Bulatov 2008).

### 5.2.1 Data collection approaches for the total site analysis of the chemical cluster

Data collection for TSA studies is time consuming; therefore practitioners have defined different approaches that can be used for conducting studies at different levels of detail. These approaches are briefly discussed below and illustrated in Figure 5.

### White box approach or Detailed Pinch

A detailed pinch analysis of each plant is carried out and thereafter Composite Curves (CC) and the Grand Composite Curve (GCC) can be constructed for the total site and the minimum hot and cold utility demands are determined.

### Grey box approach:

For each plant, only the process-utility interface is considered and process-process heat exchange is ignored. Only process streams which are heated/cooled by utilities are considered in the analysis based on their starting temperature  $T_{\text{start}}$ , target temperature  $T_{\text{target}}$  and heating/cooling loads. The current level of integration within the single units is not changed, but it enables to identify opportunities for transferring heat between plants.

### Black box approach:

The process(-stream) is represented by its utility demand only and is in the analysis represented at the corresponding utility temperature. Other utility users such as steam tracing or tank heating are often represented as black boxes (Linnhoff March 2000).

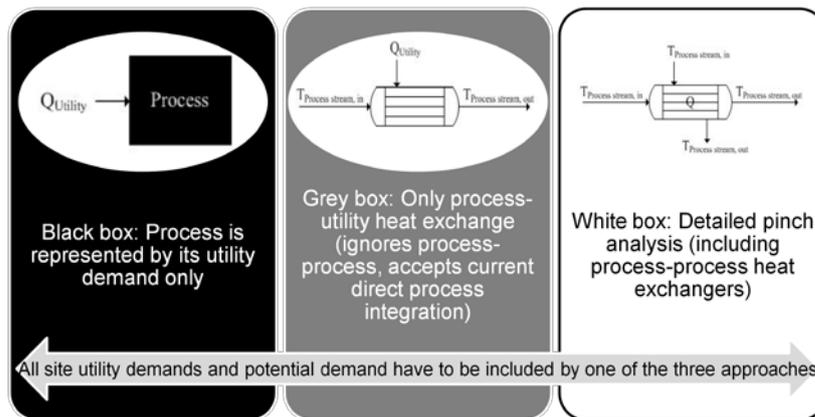


Figure 5 Illustration of the different approaches for data collection in TSA

It is important that all utility usage and potential demand is included in the study (Brown 1999). Most of the plants were analysed using the “grey box” approach since opportunities for implementing energy efficiency measures have been identified in previously conducted studies in the single plants, and the plants are therefore considered to be relatively energy efficient. The study is complemented by streams handled as “black box” to include utility consumers that are not included in the stream data gathered.

## **5.2.2 Total Site Profiles and Total Site Composite Curves**

From the data collected the process source/sink profiles and the utility profiles can be plotted. The so called total site profiles (TSP) are obtained, see left-hand side in Figure 6. This enables to analyse how heat is supplied to and discharged from the processes by different utilities. The site utility curves are developed from process stream lists by allocating the utilities used to cool/heat each process stream.

In order to identify the maximum amount of heat recovery for the total site by the utility system the total site profiles are moved towards each other until the hot and the cold utility curve intersect in one point, see Figure 6 to the right. This point is the so-called site pinch, which limits the amount of heat that can be recovered by the utility system. The overlapping curves in this figure are the so-called total site composites (TSC). They show the minimum amount of heat that has to be supplied to the processes externally as hot utility ( $Q_{\text{heating}}$ ). This is illustrated in Figure 6.  $Q_{\text{heating}}$  therefore directly relates to the fuel requirement.

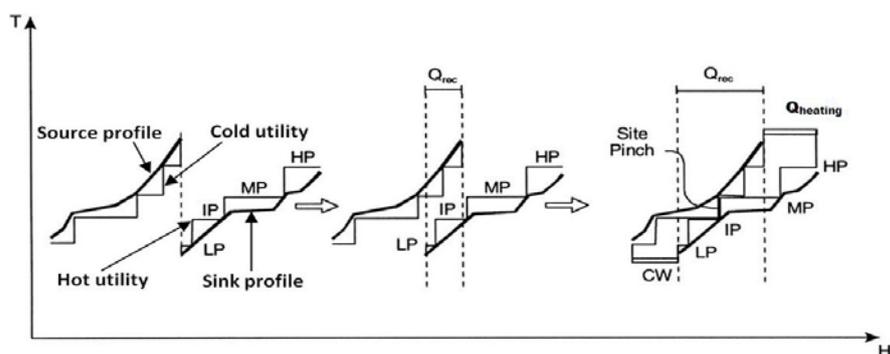


Figure 6 Total Site Profiles (TSP) and Total Site Composites (TSC)

The cooling demand, shown as cooling water (CW) in Figure 6 represents the amount of heat that has to be discharged from the processes. TSP and TSC can be used to identify changes to the utility system that improve the total site heat integration through the utility system. Utility system changes are e.g. replacing steam by introducing a hot water circuit (Bagajewicz and Rodera 2001), introduction of new steam mains, steam generation at higher levels or use at lower levels (Raissi 1994). The curves can be used to target for fuel consumption and cogeneration (Zhu and Vaideeswaran 2000).

### 5.2.3 Results from total site analysis of the chemical cluster in Stenungsund

Based on process stream data and their connection to the utility system the total site composites shown in Figure 8 can be developed. The curves represent the current utility system. The red full line represents the hot process streams, which require cooling, the green dashed line represents the cold utilities used to cool the hot process streams. The blue full line shows the cold process streams, which need to be heated and the orange dashed line illustrates the hot utilities used to heat these cold process streams.

In Figure 7 it can be seen that 122 MW<sub>heat</sub> from fuel added in boilers is needed to cover the cluster's current heating demand and 633 MW<sub>cooling</sub> are necessary to cool the processes. 320 MW<sub>rec</sub> of heat are recovered by the utility system.

It can be seen that around the site pinch the hot utility curve is placed at significantly higher temperature levels than the sink profile. This indicates that there is potential to use utilities at lower temperature to satisfy the processes heating demand. By lowering

the utility levels in certain heat exchangers low grade heat can be used for process heating which in return makes it possible to move the site pinch and thereby increase the amount of heat recovered by the utility system.

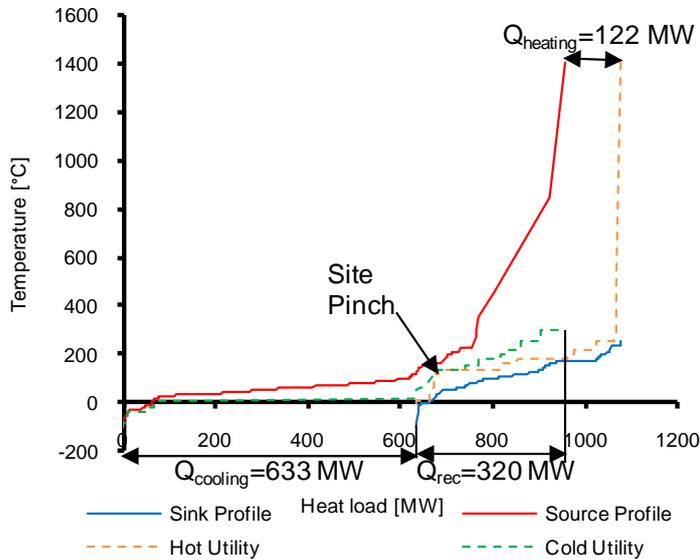


Figure 7 Total Site Composites (TSC) showing the heat integration in the Stenungsund chemical cluster with the current utility system (Hackl and Andersson 2010)

Figure 8 shows the cluster with an optimised utility system. The figure doesn't show the whole TSC. The temperature axis is cut to give a better view of the part in which the utility/process interface can be seen. Several measure to recover the maximum amount of heat from the processes and use it for heating process streams elsewhere in the cluster are applied in this optimised utility system. A hot water circuit is suggested which could enable recovery of 96 MW heat. Additionally 2 bar(g) recovery can be increased by 33 MW. By doing this it is possible to save all the current utility demand (122 MW<sub>heat</sub>). In addition, a surplus of 7 MW<sub>steam</sub> can instead be produced from excess process heat. The amount of heat recovered by the utility system increases to 449 MW and the cooling demand decreases to 506 MW respectively.

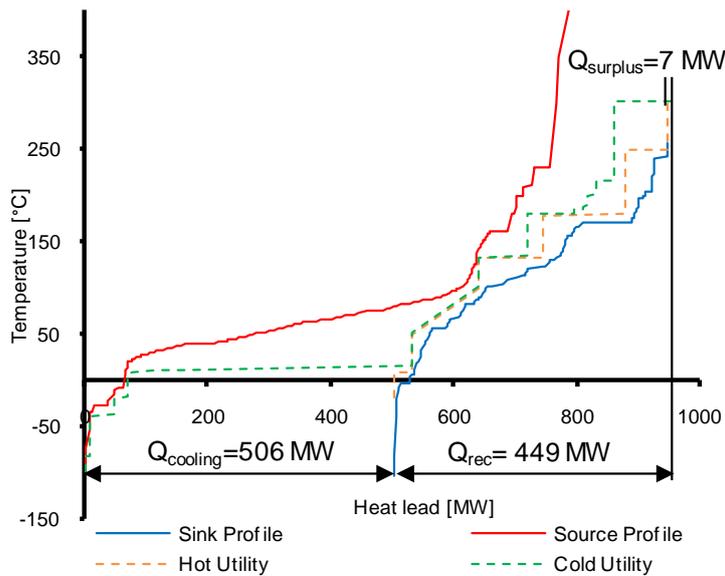


Figure 8 Total Site Composites (TSC) showing the heat integration in the Stenungsund chemical cluster with an optimised utility system (Hackl and Andersson 2010)

A qualitative evaluation of the feasibility of the suggested improvements to the utility system indicated that ca. 50 % of the savings can be achieved by implementing measures that only require moderate changes to the cluster’s utility system. More specifically it was found that with moderate changes to the utility system 56 MW of heat can be recovered in a hot water circuit and 2 bar(g) steam recovery can be increased by 4 MW. Figure 9 shows the TSC after implementation of these changes. Potential savings achievable with moderate changes are assumed to be implemented and therefore a remaining hot utility deficit of 62 MW<sub>heat</sub> still have to be covered by heat generated externally.

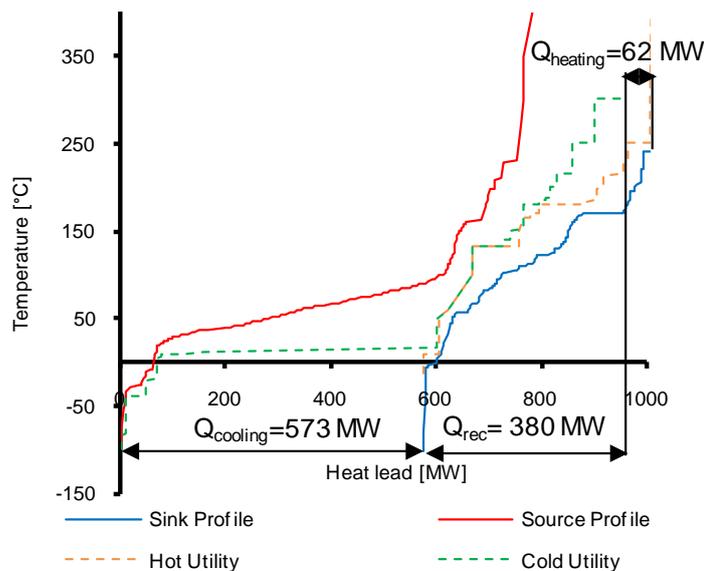


Figure 9 Total Site Composites (TSC) showing the heat integration in the Stenungsund chemical cluster after implementation of energy efficiency measures found in the feasibility study

Therefore in this study it is assumed that not all the heat demand of the cluster will be covered by internal heat exchange between the processes and that there will still be a demand of heat from boilers or other external sources of heat.

In the following the **main findings from the TSA** study conducted previously are presented:

From the stream data collected it can be seen in Figure 7 ( $Q_{\text{rec}} + Q_{\text{heating}} = Q_{\text{heating, total}}$ ) that the total demand of hot and cold utility of the cluster are 442 MW and 953 MW respectively.

The Total Site curves show a site pinch at the 2 bar(g) steam system temperature (132 C). The site pinch limits the potential for heat integration. To increase energy savings by heat integration it is necessary to change the position of the site pinch. It was shown that by introducing a site-wide hot water circuit, increased recovery of 2 bar(g) steam and adjustment of steam levels in several heaters, all the steam currently produced by purchased steam in the boilers can be replaced by internal heat exchange through the site utility system. It is even possible to produce an excess of 7 MW steam at 85 bar(g) from excess heat from the process. The potential shown is the maximum potential. In order to reach this all utility systems in the cluster have to be fully interconnected and the suggested measures (hot water, 2 bar(g) steam recovery) have to be implemented.

In the following practical measures to partly achieve the potential are described.

Only introducing a hot water circuit can save 51 MW of steam from purchased fuels, which corresponds to estimated savings of 122 MSEK/year<sup>5</sup>. It is possible to replace more steam by hot water, but the demand for 2 bar(g) steam is limited. Therefore a demand for low pressure steam must be created by adjusting steam levels in order to utilise more excess heat in a hot water circuit. The present delivery of heat to the district heating system is not affected by a site-wide hot water circuit.

There is a potential for increased recovery of 33 MW of 2 bar(g) steam from excess process heat. This would replace the production of the same amount of steam in the boilers, worth 79 MSEK/year<sup>5</sup>.

A qualitative assessment of the implementation of a hot water circuit shows estimated steam savings of 55.2 MW (132 MSEK/year<sup>5</sup>) with moderate changes (83.5 MW including more complex changes, 200 MSEK/year<sup>5</sup>). Technically the introduction of a hot water circuit includes hot water pipes between several plants, as most of the consumers of heat are situated at the cracker site and at Perstorp but the sources are spread out across the cluster. Also new piping is necessary to transfer the 2 bar(g) steam replaced by hot water to other plants with steam deficit.

The practical potential for increased 2 bar(g) steam recovery is estimated at 4.2 MW (10 MSEK/year<sup>5</sup>) with moderate changes and 26.6 MW including more complex changes (64 MSEK/year<sup>5</sup>). Increased 2 bar(g) recovery implies the construction of

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<sup>5</sup> see Appendix 3  
for underlying assumptions and calculations

new steam pipes from Borealis to Perstorp and INEOS, since most of the potential steam sources are located at Borealis but Perstorp and INEOS have a demand for 2 bar(g) steam

Currently there is one turboalternator installed in the cluster with a capacity of 10 MW<sub>el</sub>. The theoretical cogeneration potential in the cluster is 29 MW<sub>el</sub> including the 10 MW<sub>el</sub> generated today (additional revenue is 40 MSEK/year<sup>5</sup>). This assumes that all the steam currently throttled from higher pressure in let-down valves is expanded in back pressure turbines cogenerating electricity.

Some process streams below ambient temperature are heated with steam. 6.5 MW steam is used for heating streams well below ambient temperature. Heat sources at lower temperature can be used to save steam and eventually recover cooling energy if refrigerants are used for heating sub-ambient streams. This decreases the energy usage in the cooling system and also saves heating steam. Savings up to 48 MSEK/year<sup>5</sup> were estimated.

For more detailed information on the utility systems of the different companies and detailed results from the total site analysis of the cluster can be found in (Hackl and Andersson 2010).

### 5.3 Evaluation of economic and CO<sub>2</sub>-emissions consequences

Economic and CO<sub>2</sub> emissions consequences of integration of a biorefinery with a chemical cluster are assessed. This is done in the following steps:

- Mass and energy balances of the cluster and the selected biorefinery concepts are used to establish energy flows that can be exchanged between the existing and the new processes.
- Monetary flows for these streams are used to establish net cash flows<sup>6</sup>
- CO<sub>2</sub> emissions consequences are assessed by comparing the CO<sub>2</sub> emissions from alternative production of the established energy flows

The comparison is made between a biorefinery integrated in the cluster and a stand-alone unit. Investment costs for a stand-alone unit and a integrated unit are not taken into account, while the economic and CO<sub>2</sub> emissions consequences of integration are calculated based on the following assumptions. Economic consequences assess the cash-flow from sold/avoided purchase of energy of the two alternatives. For biorefinery concepts with a surplus of heat (e.g. gasification), integration makes it possible to produce both process heat and power while in stand-alone operation heat delivery is assumed to be not possible as there is no heat sink available. If low temperature excess heat from the cluster is used in the biorefinery this heat is assumed to replace heat that otherwise (in a stand-alone unit) has to be produced from fuel. This has consequences on both economics and CO<sub>2</sub> emissions.

If high temperature excess heat is rejected from the biorefinery steam can be produced, which can be further converted into electricity and heat. The assumptions taken are presented in Table 8.

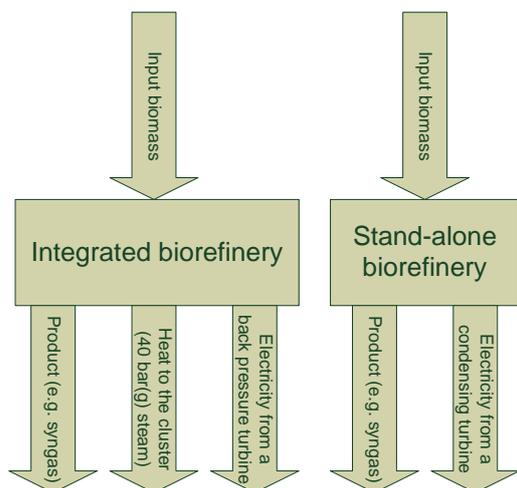


Figure 10 Illustration of an integrated and a stand-alone biorefinery producing an excess of heat

Figure 10 illustrates the comparison between an integrated and a stand-alone biorefinery. Both cases are assumed to produce the same amount of product. The

<sup>6</sup> Energy prices and CO<sub>2</sub> emissions data are taken from „Scenarios for assessing profitability and carbon balances of energy investments in industry“ (Harvey and Axelsson 2010)

integrated case assumes that excess heat at high temperatures from the biorefinery is converted into steam with the given parameters. The steam is then used for cogeneration thereby delivering heat and power to the cluster. The back pressure steam level after the turbine is assumed to be 40 bar(g), as there is a deficit of steam at this level and below in the cluster (see Figure 3). For a stand-alone unit it is assumed that all the steam is used for electricity production in a condensing turbine. The data for steam produced from biorefinery excess heat is taken from (Eriksson and Kjellström 2010) and assumed to be similar for an integrated and a stand-alone unit to achieve comparability.

*Table 8 Assumptions for utilisation of excess heat from an integrated biorefinery compared with a stand-alone unit*

Integrated biorefinery			Stand-alone unit		
Cogeneration of heat and power			Electricity production in a condensing turbine		
HP-steam generation	121	bar(a)	HP-steam generation	121	bar(a)
T <sub>superheated steam</sub>	540	C	T <sub>superheated steam</sub>	540	C
Isentropic efficiency turbine	0.8		Isentropic efficiency	0.8	
Generator efficiency	0.98		Generator efficiency	0.98	
Isentropic efficiency pump	0.8		Isentropic efficiency pump	0.8	
Steam delivered to the cluster at	40	bar(g) <sup>7</sup>	Expansion to	0.1	bar(a)
Running time	8000	h	Running time	8000	h

To evaluate the economic performance and consequences on CO<sub>2</sub> emissions of the integration vs. stand-alone operation data from “Scenarios for assessing profitability and carbon balances of energy investments in industry”. A description of the scenarios (Harvey and Axelsson 2010) is given below:

*The performance of future or long-term energy investments at industrial sites can be evaluated using consistent scenarios. By using a number of different scenarios that outline possible cornerstones of the future energy market, robust investments can be identified and the climate benefit can be evaluated. To obtain reliable results, it is important that the energy market parameters within a scenario are consistent. Consistent scenarios can be achieved by using a tool in which the energy-market parameters (e.g. energy prices and energy conversion technologies) are related to each other. Data underlying the scenarios is presented in Appendix 7.*

Table 9 shows the data for the scenarios. By doing so electricity and steam produced from waste heat from a biorefinery are evaluated. From the scenarios developed by Harvey and Axelsson two are chosen to assess the economic performance of integration vs. stand-alone operation. The prices are converted from € to SEK<sup>8</sup>. The price for saved steam is based on the price of natural gas and assumes a boiler efficiency of 80 %.

<sup>7</sup> To determine suitable levels the results from the “TSA Stenungsund” project are used. Currently there is a steam deficit at 40 bar(g) or lower in the cluster

<sup>8</sup> SEK/€=9.32

Table 9 Electricity and steam price data for economic evaluation of integrated vs. stand-alone biorefineries (Harvey and Axelsson 2010)

Scenario for	2020		2030		
	high fossil-fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	high fossil fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	
Price for electricity	540.56	428.72	652.4	419.4	SEK/MWh
Green certificates	186.4	186.4	186.4	186.4	SEK/MWh
Saved steam	431.05	302.9	594.15	349.5	SEK/MWh

Steam and electricity from biorefinery excess heat are produced from biomass and are therefore assumed to decrease fossil CO<sub>2</sub> emissions. To be able to compare the two alternatives based on their CO<sub>2</sub> emissions reduction the following values for heat and electricity from biorefinery excess heat are used, see Table 10. To assess the CO<sub>2</sub> emissions reduction when producing electricity from biomass Harvey and Axelsson assume that the electricity replaced is produced in coal power plants. The resulting CO<sub>2</sub> emissions are used for calculation of the CO<sub>2</sub> emissions reduction. When steam from the clusters boilers is replaced the lifecycle emissions for natural gas are used, which include emissions from combustion and production of natural gas.

Table 10 Consequences on CO<sub>2</sub>-emissions when delivering steam and electricity from a biorefinery (Harvey and Axelsson 2010)

	CO <sub>2</sub> -emission reduction <sup>9</sup>		Comment
Electricity from biorefinery excess heat	336	kg_CO <sub>2</sub> /MWh-el	Coal power plants assumed as marginal producer of electricity
Heat to the cluster from biorefinery excess heat	271	kg_CO <sub>2</sub> /MWh-steam	Lifecycle emissions for heat produced from natural gas <sup>10</sup>

Utilising low temperature excess heat from the cluster a biorefinery heat from the cluster heat is assumed to replace heat that otherwise (in a stand-alone unit) has to be produced from fuel (e.g. biomass in case of biomass drying). It is assumed that the same amount of dried biomass is produced.

<sup>9</sup> CO<sub>2</sub> emissions are the same in all four scenarios presented in Table 9

<sup>10</sup> assumed boiler efficiency: 0.8

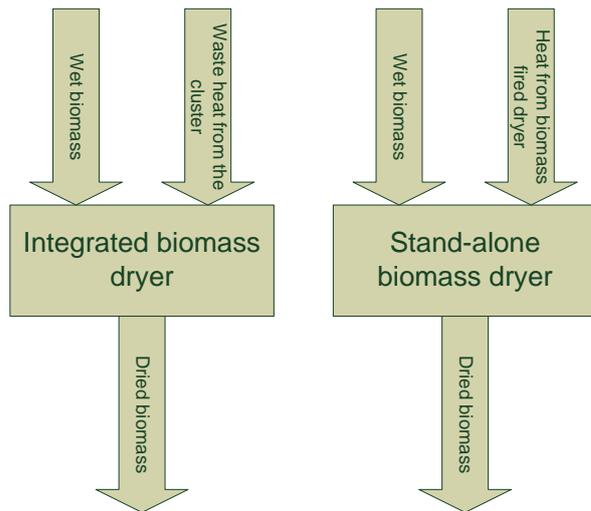


Figure 11 Illustration of an integrated (utilising waste heat) and a stand-alone biomass dryer (utilising heat from biomass combustion)

Economic and CO<sub>2</sub> emissions consequences are evaluated as described above. CO<sub>2</sub> emissions from biomass are calculated for a case where biomass is used for co-firing coal power plant and thereby replacing coal as fuel (336 kg CO<sub>2</sub>/MWh<sub>fuel</sub>). Table 11 shows the prices of biomass for the four scenarios chosen. They are used to calculate the cash-flow for saved biomass fuel if waste heat is used instead of heat from combustion of biomass for heating the drying process.

Table 11 Prices of biomass for scenarios in 2020 and 2030 (Harvey and Axelsson 2010)

Scenario for	2020		2030		SEK/ MWh
	high fossil-fuel price/CO <sub>2</sub> emission charge level 3	Low fossil fuel price/CO <sub>2</sub> emission charge level 1	high fossil fuel price/CO <sub>2</sub> emission charge level 3	Low fossil fuel price/CO <sub>2</sub> emission charge level 1	
Price of biomass	242.32	186.4	316.88	195.72	

## 6 Results

The results are presented in the following order:

- Results from the biorefinery selection process
- Description of the selected biorefinery processes
- Results from the integration of the selected processes

### 6.1 Selection of suitable biorefinery concepts

Figure 12 shows an overview of biorefinery concepts and potential products obtained by different conversion techniques.

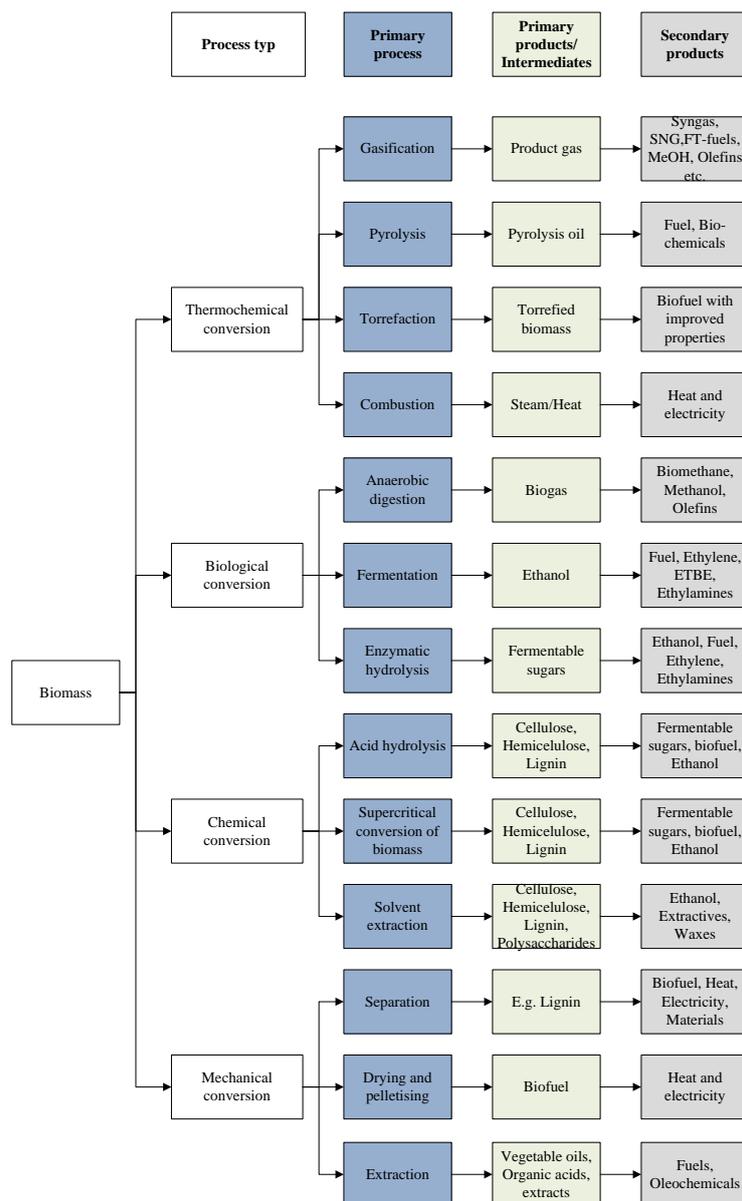


Figure 12 Overview of biomass conversion processes and potential products (Bludowsky and Agar 2009; de Jong, van Ree, and Kwant 2009; Naik 2010; Sadaka and Negi 2009; UOP LLC 2004)

The figure is divided in process type which represents the general kind of biorefinery process, primary processes which specify the main process step, primary products and

intermediates which are obtained from the primary processes and secondary products which can be derived from the primary products by further processing.

The overview in Figure 12 is used to determine suitable biorefinery concepts integrated in the chemical cluster in Stenungsund.

Table 12 shows the main feedstocks consumed and products and intermediate materials produced in the chemical cluster in Stenungsund.

*Table 12 Overview of the main raw materials used in the chemical cluster in Stenungsund (Arvidsson 2010; Ekeröth 2010; Josefsson 2009; Borealis AB 2009; Pollard 2010; Olsson 2010)*

Raw materials	Annual consumption [kt/yr]
Ethylene (incl new HPPE plant)	817
Butane	268
Cl <sub>2</sub>	115
Ethane	321
NaCl	199
Naphtha	298
Propane	286
Propylene	190
Rape seed oil	125
VCM	198
Methane	95
Products and intermediates	Annual production [kt/yr]
<b>Ethylene</b>	<b>565</b>
Propylene	174
PVC	178
<b>Syngas (H<sub>2</sub>/CO= 1)</b>	<b>160</b>
RME	125
VCM	156
Iso- and n-butyraldehyde	300
PE	649
Amines	80
Ethylene Oxid (EO)	85

As it can be seen in the table, ethylene is one of the most used and produced bulk chemical in the cluster. It can be also seen that there is a deficit between the annual ethylene production and consumption of 252 kt/yr. Also there is a strong interest among several companies to investigate alternative processes for the production of ethylene based on renewable feedstocks. Therefore biomass based production processes for the production of ethylene will be studied further from a process integration point of view.

Another intermediate which is produced and consumed to a large extent in the cluster is syngas. Syngas is currently produced at Perstorp from methane by steam reforming. 160 kt/yr of syngas consisting of equal volume fractions of CO and H<sub>2</sub> are consumed. Alternative (biomass based) syngas production routes exist. Therefore process integration of a biomass gasification process producing syngas will be studied in this report. The products of the gasification process can be further converted into biofuels (DME, FTD) or used as substitute other fuels and chemicals such as ethylene, fuel gas and H<sub>2</sub>. Biomass gasification in general yields an excess of high temperature heat, which can be used to cover parts of the clusters heat and electricity demand.

In the TSA study performed earlier (Hackl and Andersson 2010) it was found that there is a large amount of waste heat available, which is currently discharged. Low temperature vacuum dryers for biomass can make use of a fraction of this low grade heat. Figure 13 shows one result from the study. It can be seen that there are 498 MW of low temperature heat available which can potentially be used for drying biomass. The dry biomass can be further processed into pellets (Andersson, Harvey, and Thore Berntsson 2006) or used as feed to other biorefinery processes, e.g. gasification. Therefore integration of a biomass drying plant is investigated.

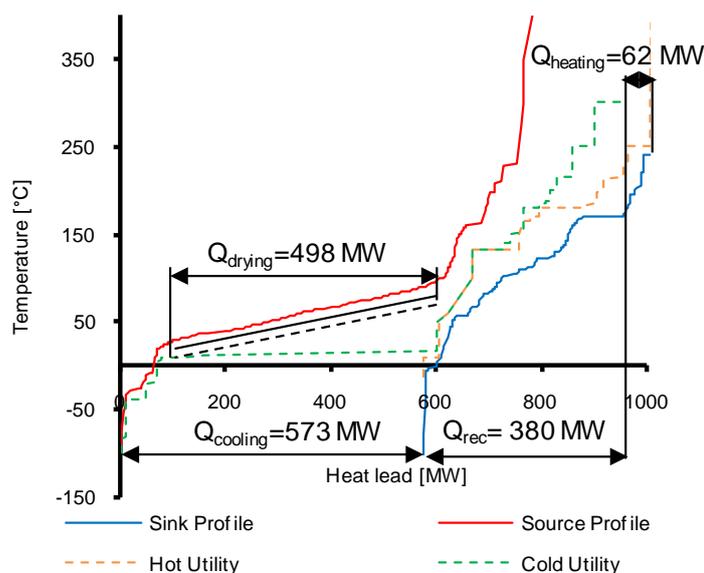


Figure 13 Results from the study „TSA Stenungsund” showing a large amount of waste heat available for drying of biomass with low temperature excess heat ( $Q_{drying}=498$  MW)

The biorefinery concepts chosen to further study process integration opportunities are:

- Biomass gasification for syngas production
- Ethylene production from biomass
- Biomass drying with low temperature excess process heat

The processes are further described as follows.

## 6.2 Description of the selected processes

### 6.2.1 Biomass to synthesis gas

Gasification is the conversion of carbonaceous fuel to a gaseous product with a usable heating value. The product gas consists mostly of syngas which is CO and H<sub>2</sub>. The syngas can be used as a fuel or as a raw material for fuels and chemicals (Higman and Burgt 2008). In Stenungsund syngas is currently produced by methane reforming at Perstorp’s site as an intermediate for the production of speciality chemicals.

There are a number of different gasifier types which are able to handle different raw materials and have different product outputs. Figure 14 shows the route for syngas production from biomass. Wet biomass is first dried before being fed to the gasifier. It also shows different types of gasifiers which are determined by the production capacity and process/product requirements. The gas leaving the gasifier mainly consists of CO, H<sub>2</sub>, CO<sub>2</sub>, CH<sub>4</sub> and water vapour (Heyne 2010). In order to make it

suitable for further processing it has to be cleaned in several steps. The remaining hydrocarbons are converted into CO and H<sub>2</sub> in a reforming step. Gas cleaning is performed by processes described in Figure 14 to remove tars, particulate matter, sulphur and nitrogen compounds. In the water gas shift reactor the ratio between CO and H<sub>2</sub> is adjusted to suit the requirements of the syngas application. In a final step CO<sub>2</sub> is removed from the syngas. The produced syngas can then be converted into fuel products such as FT-fuels, DME, Methanol or be used as gaseous feedstock to existing reaction processes in the cluster. Thereby syngas currently produced by methane steam reforming can be substituted by syngas produced from biomass by gasification.

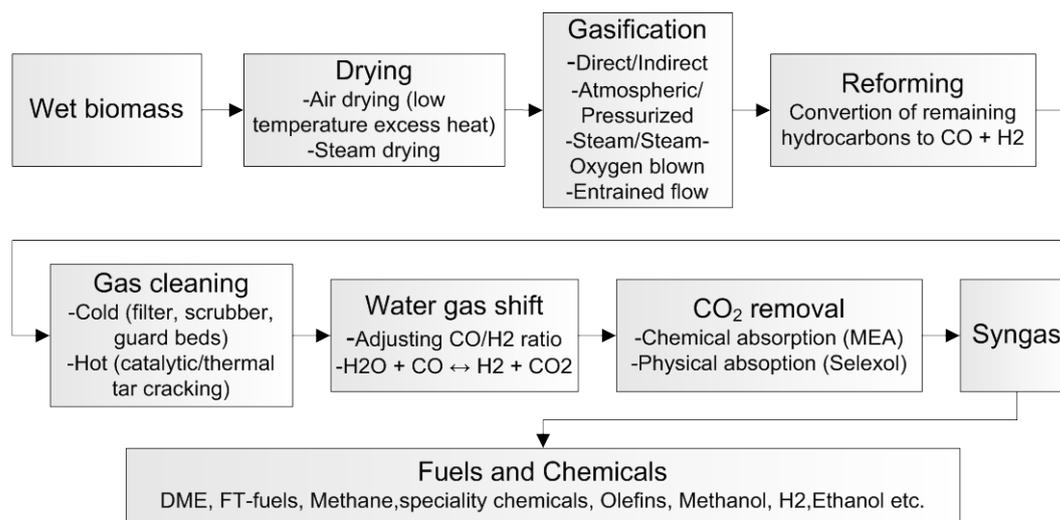


Figure 14 Different process options for biomass gasification

As an example on how integration of a gasification unit in an existing cluster can be done the flowsheet in Figure 15 shows an oxygen-blown circulating fluidized bed (CFB) gasifier. Other types of gasifiers and gas treatment processes can also be applied. Further investigation is necessary to determine to most suitable process.

In this example dried biomass, oxygen and steam enter the gasification chamber. In the high temperature filter fly ash is removed and in the reformer more oxygen and steam are added in order to catalytically convert methane and other hydrocarbons to syngas. Thereafter the gas is cooled and steam is added to adjust the CO/H<sub>2</sub> ratio in the water gas shift reactor. Finally the gas is further cooled and CO<sub>2</sub> is removed in a MEA absorption unit. Data for stream calculations is taken from (Eva K. Larsson 2010) and (Ahlgren et al. 2007).

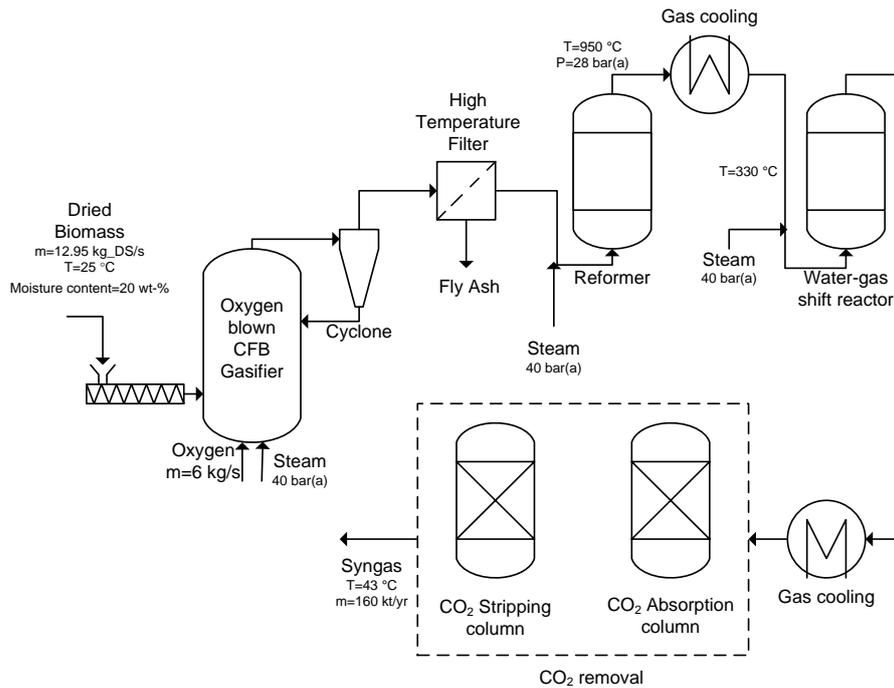


Figure 15 Flow sheet of an oxygen-blown CFB gasifier with high temperature filter, gas reformer shift reactor and CO<sub>2</sub>-removal

The final H<sub>2</sub>/CO ratio in this example is approximately 2:1 and is therefore not the same as used in the cluster (1:1). This is due to the fact that the data available for the gasification unit was intended for a gasifier targeting for high H<sub>2</sub> production. As not only direct utilisation of the syngas in the cluster is investigated but also potential production of other products, this process has been chosen as an example for integration of a gasification unit. Further investigation is necessary to determine the most suitable process design to meet the syngas specification in the cluster.

The gasification plant presented here needs oxygen which has to be produced in an air separation unit (ASU). AGA gas already operates an ASU in the cluster, which is also an advantage for integrating a gasification unit might be extended, the knowhow for air separation is already in place and surplus oxygen can be used to drive the gasification process<sup>11</sup>.

## 6.2.2 Biomass to Ethylene

In Figure 16 different processes for the production of ethylene from biomass are shown. In the following the routes are described more in detail.

<sup>11</sup> AGA produces at the moment 0.66 kg/s of oxygen more than the cluster needs.

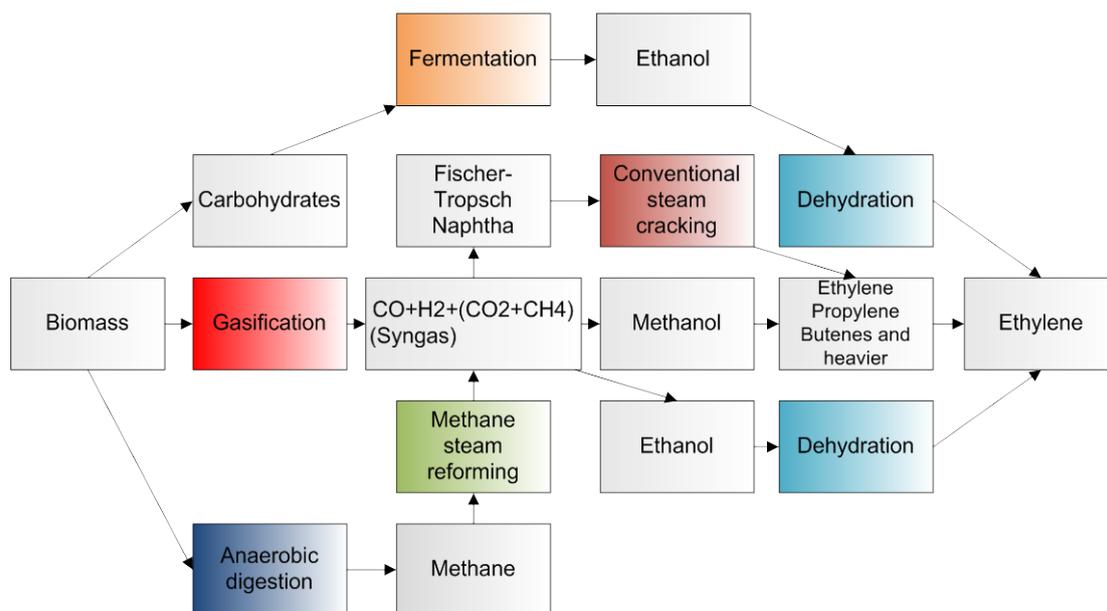


Figure 16 The biomass to ethylene production route: Different biomass conversion technologies to produce ethylene from biomass (Okkerse and Bekkum 1999; Ren et al. 2009; UOP LLC 2004; Jones, Kresge, and Maughon 2010)

### 6.2.2.1 The Fermentation route

One route to produce ethylene from biomass is a two step process. First the carbohydrates in the biomass are fermented to ethanol which then is dehydrated to form ethylene. Today most biomass derived ethanol is produced by hydrolysis/fermentation of sugar or starch from e.g. sugar cane, corn or wheat (Fornell and Thore Berntsson 2009) . Production of ethanol from lignocellulosic materials such as wood is still at the research stage. Today only pilot plants are in operation (Jones, Kresge, and Maughon 2010). In Figure 17 an example of an Ethanol plant for lignocellulosic raw materials is presented. The process consists of three steps, pre-treatment, saccharification and fermentation and distillation. In the pre-treatment wood chips are heated with steam to 160-220 °C and treated with H<sub>2</sub>SO<sub>4</sub> or SO<sub>2</sub>. By fast pressure expansion the wood structure is broken down and most of the hemicelluloses and part of the cellulose are hydrolysed. In the next step enzymes are added to hydrolyse the remaining into sugars which are simultaneously fermented to ethanol. The product leaving the simultaneous saccharification and fermentation (SSF) unit is distilled. This yields a 94 vol-% ethanol solution and solid (mainly lignin) and liquid (mainly pentoses and non-fermented hexoses) residues which mainly consist of lignin. In a filtration step the solid residues are removed after distillation. The liquid residues are concentrated to a dry matter content of 50 wt-%. The residues can then be used to produce energy for the process. Energy efficiency measures lead to a surplus of energy in the ethanol process<sup>12</sup> when utilising the residues for heat and power generation. This surplus can be used to deliver heat and electricity to other consumers, like industrial processes, the electricity grid or a district heating network. The solid

<sup>12</sup> as not all the heat from combustion of the residues is need in the process

residues can also be sold directly as fuel or further processed into pellets (Eriksson and Kjellström 2010).

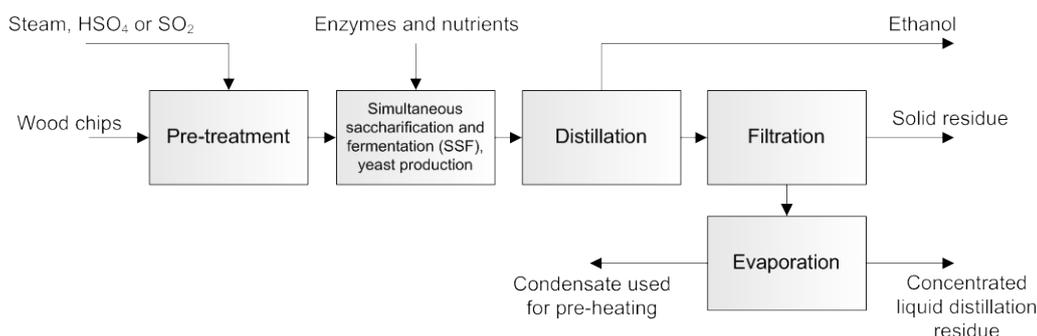


Figure 17 The ethanol production process from lignocellulosic raw materials (Eriksson and Kjellström 2010)

In the second step ethylene is produced by dehydration of ethanol. Ethanol dehydration processes are commercially available. The catalytic dehydration can be represented by the following chemical reaction:



The flowsheet of such a process is shown in Figure 18. The incoming ethanol is evaporated and then heated to reaction temperature in a furnace before it enters the reactor. Depending on the type of reactor used the operating temperature varies between 330°C and 500°C. After the dehydration reactor the raw ethylene is cooled, while part of the heat is recovered and then sequentially purified to reach chemical or even polymer grade (Morschbacker 2009). Usually a 99%+ conversion of ethanol is preferred, as ethanol recovery is energy and capital intensive.

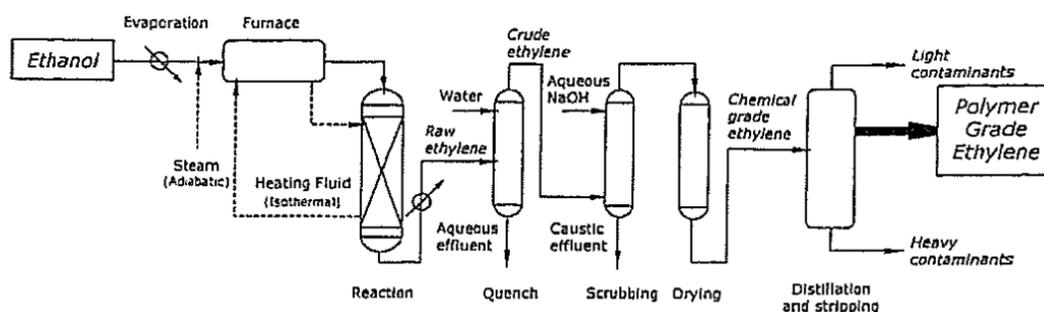


Figure 18 Process flow diagram for the production of polymer grade ethylene by ethanol dehydration (Morschbacker 2009)

Several companies have constructed or plan to construct plants to produce ethylene from ethanol. Braskem (Deligio 2010) and Dow Chemical Co. and Crystalsev (Ondrey 2007) e.g. are planning to start operating plants producing PE from sugar cane by fermentation of sugar to ethanol followed by ethanol dehydration. The planned capacities of the plants are 200 and 350 kt/year respectively. The investment costs for an EtOH-to-Ethylene plant are significantly less (per ton olefin) than for a conventional naphtha steam cracker. This can be seen in Figure 19. Investment costs for different ethylene production plants are shown. On the x-axis the annual production capacity in million lbs (1 lb = 0.454 kg) can be seen. Capital estimates on

the y-axis are escalated to fourth quarter 2008 using the SRI PEP index (Jones, Kresge, and Maughon 2010).

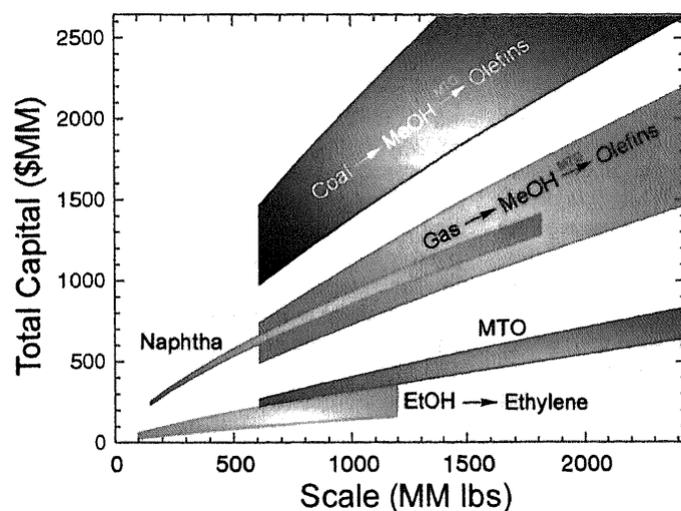


Figure 19 Total project capital as a function of total olefin production for different olefin production technologies. Total olefin production is the sum of ethylene and propylene. Uncertainty bands represent the range of literature values. (Jones, Kresge, and Maughon 2010)

It is most likely that in the short term that bio-ethanol (EtOH) dehydration for the production of ethylene will be established in regions with cheap access to bio-EtOH, e.g. Brazil where ethanol has achieved similar price as fossil-based fuel per unit energy (Jones, Kresge, and Maughon 2010). In Europe and the US this trend is expected after the introduction of lignocellulosic EtOH (Haveren, Scott, and Sanders 2008).

#### 6.2.2.2 Gasification for ethylene production

Gasification of biomass entails several possible routes to produce ethylene from the resulting syngas. In the following three possibilities are presented.

The most preferred option to produce ethanol from syngas is the Dow mixed alcohol technology. Quoted EtOH yields with this technology are higher than with fermentation (above 570 l-EtOH/ton dry biomass (Jones, Kresge, and Maughon 2010), fermentation yields app. 295 l-EtOH/ton dry biomass<sup>13</sup>). An integrated lignocellulosic EtOH plant with lignin gasification gives the opportunity to convert almost all the incoming biomass into fuel or chemicals. The ethanol can be then converted into ethylene as described above.

The methanol-to-olefin (MTO) process is another alternative for converting syngas from a gasification unit into ethylene (and other olefins). The total energy consumption of the most efficient process (UOP MTO) is 150 % higher than conventional state-of-the-art naphtha steam cracking plants.

Syngas can also be converted into FT-naphtha and afterwards used as feed to a conventional naphtha steam cracker. This has the advantage that existing equipment

<sup>13</sup> Calculations see Appendix

can be used to convert the FT-naphtha into olefins (Ren, Patel, and Blok 2008; Jones, Kresge, and Maughon 2010).

### **6.2.2.3 Methane from anaerobic digestion**

Currently the most common technology for syngas production is methane steam reforming. Natural gas is taken from the natural gas grid and the methane is converted into syngas. A renewable source for natural gas is by anaerobic digestion of biomass. The product gas from the digester is upgraded to meet the requirements for natural gas and then fed to the natural gas grid. As this process is likely to be applied in other locations and not close to the cluster it is not further considered for process integration in this study.

Competitive ethylene production technologies compared to steam cracking exist. EtOH dehydration gains more and more attention and is advantageous in certain regions. For Europe this technology is expected to become favourable after the introduction of lignocellulosic EtOH. According to the SRI process economics report (Cesar 2007) the production of ethylene from ethanol can be cost-competitive with steam cracking at small plant capacities. Considering historical price trends for ethanol and ethylene, a dehydration plant that purchases ethanol and sells ethylene at normal market prices would be only marginally profitable. Process integration can play an important role in improving the situation if e.g. the dehydration plant can be integrated with the upstream ethanol plant. Another possibility for improving economics of such a process is if bio-based ethylene could be sold at a higher price than petrochemical ethylene (Cesar 2007).

→ Therefore **in this study** ethylene production from lignocellulosic biomass fermentation and EtOH dehydration will be investigated more in detail.

### **6.2.3 Biomass drying**

Biomass drying is a very energy demanding process. There are several different dryers that utilise excess heat. (Fagernäs et al. 2010) presents a summary of several dryer options. A commercially available dryer utilising hot water or low pressure steam to heat the drying air is the band conveyor dryer.

It utilises waste heat at temperatures between 30 to 90 °C (but also other temperatures are possible). Figure 20 shows a schematic picture of a band conveyor dryer. The dryers can be operated under vacuum to accelerate the drying process. The incoming drying air is heated by waste heat in either a gas/gas or liquid/gas heat exchanger.

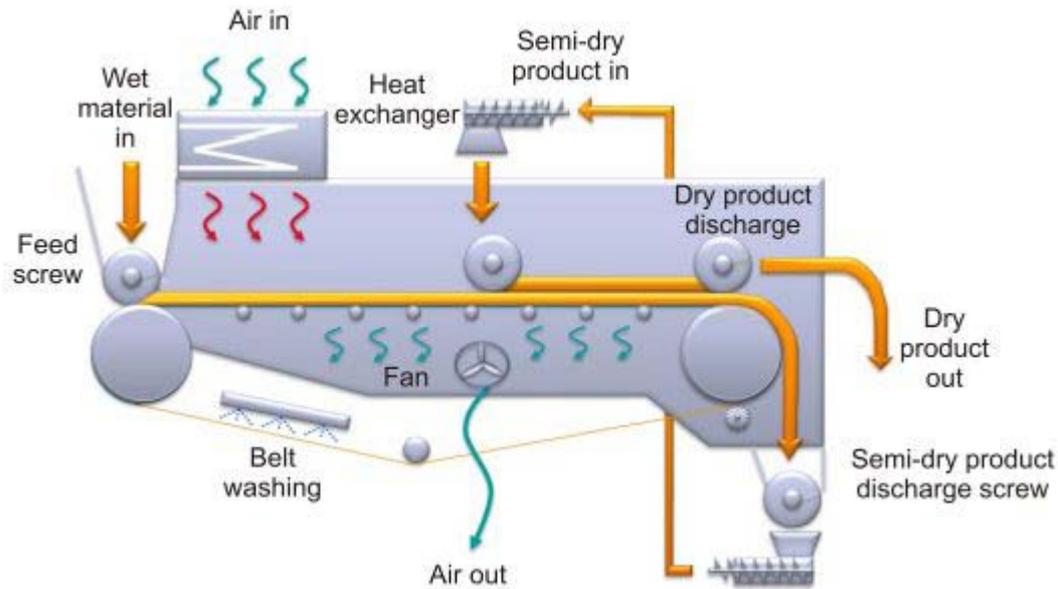


Figure 20 Schematic picture of a band conveyor dryer by Metso (Metso 2010)

The heat exchanger type depends on the heat source available. The drying material is distributed on conveyor belts and transferred between different drying zones. The drying medium flows through the conveyor belts and the wet material and is thereby cooled and saturated with moisture.

A reference plant installed at Agrior, Canada (Johansson, Sara Larsson, and Wennberg 2004) has a capacity of 1 ton-biomass/h with a moisture content of 55 wt-% and a final moisture content of 8 wt-% using low temperature heat at 42 °C.

## 6.3 Process integration of selected biorefineries in the chemical cluster

### 6.3.1 Biomass gasification

#### 6.3.1.1 Integration opportunities

Simulation of a biomass gasification plant consisting of the elements as shown in Figure 15 and input/output data as given in Table 13 and Appendix 2 was simulated using Aspen Plus with input data for the gas composition after the reformer from a study performed by TPS (Eva K. Larsson 2010). The size of the gasification plant was defined to cover the total consumption of syngas at Perstorp, which currently produces 160 kt/yr of syngas.

Table 13 Process parameters of biomass gasification unit for syngas production

Parameter	Value
Biomass input	12.95 kg_DS/s
Moisture content of incoming biomass	20 wt-%
Oxygen demand	6 kg/s
Gas flow after the reformer	28.84 kg/s
Gas temperature after the reformer	950 °C
Pressure after the reformer	29 bar(a)
Product gas flow	5.56 kg/s
Annual product gas production	160 kt/a

The full line in Figure 21 shows the GCC developed from the simulated process. It can be seen that all the heating demand of the process can be covered by internal heat exchange. The GCC shows two temperature regions where there is excess heat available. 18 MW of excess heat is available at high temperature levels (above 600 °C). This heat can be used to generate steam (dashed line in Figure 21) for process heating and/or cogeneration of electricity and power (to drive compressors) in the cluster.

In the cluster currently 122 MW of steam are produced in boilers by purchased fuel. Parts of this demand can be covered by high temperature excess heat from gasification. A turbogenerator is currently operated at part load. High pressure steam raised in the gasification unit can be used for cogeneration without the need for investing in a new turbine.

Another source of excess heat in the gasification process is at temperatures below 116 °C (22.5 MW). A potential use for part of this heat is low temperature drying of the incoming biomass to the gasifier or it can be used for process heating by a hot water circuit as suggested in the TSA study by (Hackl and Andersson 2010).

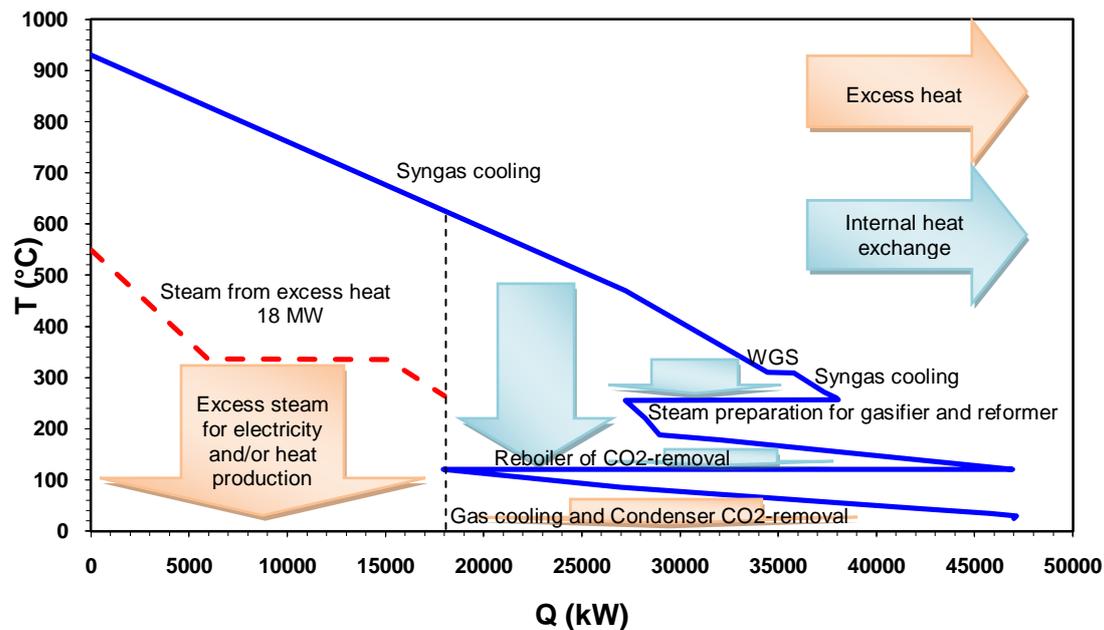


Figure 21 GCC of the gasification process described in 6.2.1, showing potential internal heat exchange and excess heat of the process

A large amount of heat is necessary for separation of CO<sub>2</sub> from the syngas (H<sub>2</sub>/CO ratio=2:1). This situation looks different, if a different H<sub>2</sub>/CO ratio (e.g. 1:1) is chosen. Here further investigation is necessary as discussed in 6.2.1. CO<sub>2</sub> removal from syngas with a H<sub>2</sub>/CO ratio of 1:1 is expected to be less energy demanding because less CO<sub>2</sub> is produced during gasification compared to syngas with higher H<sub>2</sub>/CO ratios.

### 6.3.1.2 Economic and CO<sub>2</sub> emissions consequences of integration compared to stand-alone operation

The background to this section is presented in 5.3. Excess heat from gasification can be utilised in the cluster. The previous TSA study of the cluster showed that there is a deficit of 40 bar(g) steam (and below)<sup>14</sup>. It is assumed that steam at 120 bar(g) is produced from excess heat from gasification. The steam expanded in a back-pressure turbine to 40 bar(g) cogenerating electricity and steam to be used in the cluster. For a stand-alone unit it is assumed that the excess heat is converted to electricity in a condensing turbine. The assumptions taken and the calculations performed are shown in Appendix 5. The results are presented in Table 14.

*Table 14 Heat and electricity produced from waste heat from an integrated gasifier and a stand-alone unit*

Integrated gasifier			Stand-alone			
	Excess heat from gasification	144	GWh/yr	Excess heat from gasification	144	GWh/yr
Electricity	$P_{el,turb}$	16	GWh/yr	$P_{el,turb}$	47	GWh/yr
Heat	$Q_{to\ cluster}$	124	GWh/yr	$Q_{waste\ heat\ to\ condenser}$	96	GWh/yr

Table 14 shows that in the case of an integrated gasification unit it is possible to produce 16 GWh/yr of electricity and to deliver 124 GWh/yr of steam at 40 bar(g) to the cluster. If all syngas at Perstorp is replaced by syngas from gasification, 161.6 GWh/yr of 40 bar(g) steam currently produced from methane steam reforming for syngas production also has to be replaced. Therefore it is important to deliver as much heat as possible from the gasification unit to the cluster. A stand-alone gasifier producing electricity with a condensing turbine can produce 47 GWh/yr. This means that with an integrated gasifier electricity from the grid and heat produced from fossil fuels is replaced by heat and “green” electricity from biomass. In the stand alone case “green” electricity is produced. By the production of “green” electricity it is assumed that subsidies are granted for the production. The economic value of these subsidies is also presented in the results.

The results of the evaluation of the economic performance of the two options is in Table 15 and Table 16.

*Table 15 Results from economic evaluation of an integrated gasifier*

Integrated case	2020		2030		
	High fossil-fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	High fossil fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	
Economic value of electricity	8.7	6.9	10.5	6.7	MSEK/yr
Economic value of "green" certificates	3.0	3.0	3.0	3.0	MSEK/yr
Avoided fuel purchase costs for steam production	53.5	37.6	73.7	43.3	MSEK/yr
<b>Total</b>	<b>65.1</b>	<b>47.5</b>	<b>87.2</b>	<b>53.1</b>	<b>MSEK/yr</b>

*Table 16 Results from economic evaluation of a stand-alone gasifier*

Stand-alone	2020		2030		
	High fossil-fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	High fossil fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	
Economic value of electricity	25,6	20,3	30,9	19,9	MSEK/yr
Economic value of "green" certificates	8,8	8,8	8,8	8,8	MSEK/yr
<b>Total</b>	<b>34.5</b>	<b>29.2</b>	<b>39.8</b>	<b>28.7</b>	<b>MSEK/yr</b>

Table 17 shows the CO<sub>2</sub> emissions reduction when replacing heat and electricity to the cluster by excess heat from a gasification unit. An integrated and a stand-alone unit are compared.

*Table 17 CO<sub>2</sub> emissions reduction by utilising waste heat from an integrated and a stand-alone gasification unit*

	<b>Integrated</b>	<b>Stand-alone</b>	
CO2 emissions reduction from replaced electricity	5403	15931	t/yr
CO2 emissions reduction from replaced steam	34688		t/yr
<b>Total</b>	<b>40091</b>	<b>15931</b>	<b>t/yr</b>

It can be seen that both economically and from a CO<sub>2</sub> emissions perspective it is advantageous to integrate a gasification unit in the cluster. Integration is advantageous compared to stand-alone operation in all scenarios. Depending on the scenario integration gives between 18.3 and 47.4 MSEK/yr higher incomes. The higher incomes can be used to cover the costs for integration. CO<sub>2</sub> emissions can be reduced by 24.4 kt/yr compared to stand-alone operation.

## 6.3.2 Biomass to Ethylene

### 6.3.2.1 Integration opportunities

As described in 6.2.2.1 ethylene production from biomass via ethanol involves two steps. The ethanol production from biomass and ethanol dehydration to ethylene.

#### Production of ethanol from lignocellulosic raw material

When producing ethanol from lignocellulosic raw material app. 35 % of the energy of the incoming biomass is converted into ethanol. 55 % of the energy is contained in the residues. In the ethanol production process most of the heat has to be supplied to the pretreatment reactor and for separation of the ethanol. This heat can be produced by utilising of unused raw materials, such as solid (mainly lignin and cellulose) and liquid (mainly pentoses and non-fermented hexoses) residues. Table 18 shows the energy content of the different material flows entering and leaving the ethanol plant per kt of ethanol produced. The following case assumes a plant running time of 8000 h/yr with a production capacity of 174 kt\_EtOH/yr, which is the amount of ethanol needed to supply a 100 kt\_ethylene/yr unit. Data was taken from a report by (Eriksson and Kjellström 2010), who assumed a 119 kt/yr plant to be a likely size for such a unit.

*Table 18 Energy content of the material flows in the ethanol production process (Eriksson and Kjellström 2010)*

Material flow	Energy content in [MW]
Feedstock, (50 wt-% moisture)	431.4
Ethanol (100 vol-%)	156.4
Solid residues (15 wt-% moisture)	170.2 <sup>15</sup>
Distillation residues (15 wt-% moisture)	69 <sup>15</sup>

Literature values of the energy demand of a lignocellulosic ethanol plant are given in Table 19. Steam to cover the ethanol plants energy demand is produced at 120 bar(g) by combustion of the residues.<sup>16</sup> The steam is expanded in a turbine to cogenerate heat and electricity to the process. Not all the energy from combustion of the residues is needed in the ethanol process. All this is illustrated in Figure 22. It contains the flue gas stream from combustion of the residues (full line), 120 bar(g) steam produced to cogenerate heat and power to the ethanol process (dashed line) the electricity demand of the ethanol production, steam demand of the ethanol process (dotted line) and excess heat (arrow).

*Table 19 Energy demand of a lignocellulosic ethanol plant (Eriksson and Kjellström 2010)*

Material flow	Energy demand in [MW]
Electricity	20.6
25 bar steam	37.4
12 bar steam	40.8
4 bar steam	67

<sup>15</sup> Can be utilised for heat and electricity production

<sup>16</sup> Assuming steam is produced by combustion of the residues in a boiler with an efficiency of 0.8

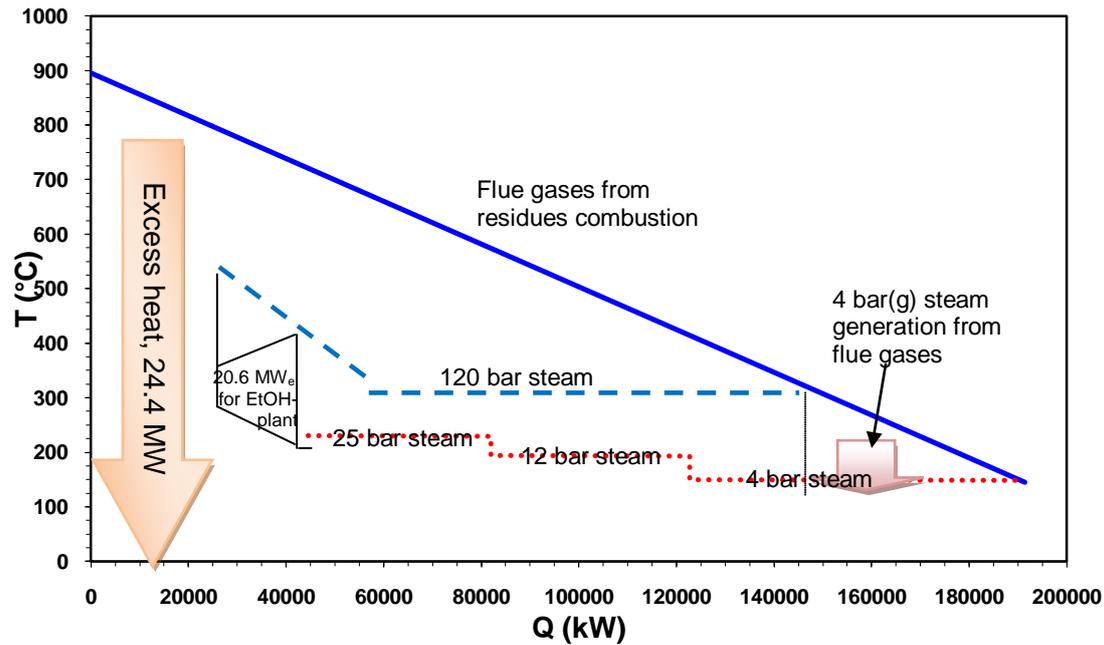


Figure 22 Illustration of the energy system of an ethanol production plant with residual material combustion, excess heat and steam and electricity production in a lignocellulosic ethanol production plant

For an ethanol plant with a capacity of 174 kt\_EtOH/a and 8000 operating hours per year, this means that considering a efficiency of the residues boiler of 0.8 a surplus of 24.4 MW can be exported as illustrated by the arrow in Figure 22.

As stated by (Eriksson and Kjellström 2010) a back-pressure steam plant connected to the ethanol production plant will need a large heat sink. This is the case in the cluster in Stenungsund as discussed in 5.2.3. It is possible to utilise excess heat from lignocellulosic ethanol production in the cluster. This collaboration between the cluster and a integrated lignocellulosic ethanol production plant leads to a high efficiency of the combined processes, as there is a potential use for all the energy products from ethanol production.

#### Summary of integration options for lignocellulosic ethanol production:

- Excess steam and electricity from the ethanol production plant can be used to cover:
  - part of the steam and electricity demand in the ethanol dehydration plant for converting the ethanol into ethylene
  - part of the cluster's heat and electricity demand
- Syngas obtained by gasification of the residue streams from the ethanol plant can be further processed to produce materials and energy to be utilised as described in 6.2.1.

#### Production of ethylene from ethanol

The ethylene dehydration process is shown in Figure 18. The process has a net energy demand. Literature values for the raw material and utility consumption of the process can be found in Table 20.

Table 20 Raw material and utility consumption of the ethanol dehydration process (Kochar, Merims, and Padia 1981)

Raw material and Utility	Consumption
Ethanol	1.74 kg/kg_ethylene
Steam (net)	1.21 kg/kg_ethylene
Cooling water CW ( $\Delta T=10K$ )	99.82 kg/kg_ethylene
Power	1116 kJ/kg_ethylene
Fuel (net), organic by-products used as fuel	1676 kJ/kg_ethylene
Process water	0.36 kg/kg_ethylene

Heat has to be added to the reactor as fuel or steam. After reactor heating the heat can be partly recovered and used in the process. The Ethylene purification also has a heat deficit. Detailed process stream data for the ethanol dehydration process was not considered in this study<sup>17</sup>. From the overall utility consumption the energy demand of a plant producing 100 kt\_ethylene/yr is calculated. Results are presented in Table 21.

Table 21 Steam, power and fuel consumption of an ethanol dehydration plant to produce 100 kt\_ethylene/yr, assumed operating hours 8000 h/yr

Utility	Consumption in [kW]
Steam (net)	10903
Power	3875
Fuel (net), organic by-products used as fuel	5819

#### Summary of integration options for ethanol dehydration:

- Necessary steam can be produced in boilers working at currently not operating at full
- Utilising excess steam and electricity from a lignocellulosic ethanol plant

#### **6.3.2.2 Economic and CO<sub>2</sub> emissions consequences of integration compared to stand-alone operation**

The background to this section is presented in 5.3. Excess heat from a lignocellulosic ethanol production plant can be utilised in the cluster and a downstream ethanol-to-ethylene plant. The energy demand of an ethanol-to-ethylene plant is 4 MW<sub>el</sub> and 11 MW<sub>steam</sub><sup>18</sup>. The previous TSA study of the cluster showed a deficit of 40 bar(g) steam (and below). As the steam levels necessary for the ethanol dehydration plant are unknown it is assumed that excess steam produced by incineration of residues from ethanol production is expanded in a back-pressure turbine to 40 bar(g) cogenerating electricity and steam. For a stand-alone unit it is assumed that the excess heat is converted to electricity in a condensing turbine. The assumptions taken and the calculations performed are shown in Appendix 6. The results are presented in Table 22.

<sup>17</sup> But will be gathered in a Master thesis project starting in September 2010

<sup>18</sup> the underlying assumption and the thorough analysis see above and Table 21, the steam levels are unknown

Table 22 Heat and electricity produced in an integrated and a stand-alone unit

Integrated				Stand-alone		
	Excess heat from gasification	195.2	GWh/yr	Excess heat from gasification	195.2	GWh/yr
Electricity	$P_{el,turb}$	21.8	GWh/yr	$P_{el,turb}$	64.3	GWh/yr
Heat	$Q_{to\ cluster}$	168	GWh/yr	$Q_{waste\ heat}$	128	GWh/yr

Table 22 shows that in the case of an integrated ethanol plant it is possible to produce 21.8 GWh/yr of electricity and to deliver 168 GWh/yr of steam at 40 bar(g) to the cluster and the downstream ethylene plant. A stand-alone unit producing electricity with a condensing turbine can produce 64.3 GWh/yr. This means that with an integrated plant, electricity from the grid and heat produced from fossil fuels is replaced by heat and electricity from biomass. In the stand-alone case “green” electricity is produced.

The results of the evaluation are shown in Table 23 and Table 24.

Table 23 Results from economic evaluation of an integrated ethanol production unit

Integrated case	2020		2030		
	High fossil-fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	High fossil fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	
Economic value of electricity	11,8	9,3	14,2	9,1	MSEK/yr
Economic value of “green” certificates	4,1	4,1	4,1	4,1	MSEK/yr
Avoided fuel purchase costs for steam production	72,4	50,9	99,8	58,7	MSEK/yr
Total	88.3	64.3	118.1	71.9	MSEK/yr

Table 24 Results from economic evaluation of a stand-alone ethanol production unit

Stand-alone	2020		2030		
	High fossil-fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	High fossil fuel price/CO2 emission charge level 3	Low fossil fuel price/CO2 emission charge level 1	
Economic value of electricity	34,7	27,6	41,9	27,0	MSEK/yr
Economic value of “green” certificates	12,0	12,0	12,0	12,0	MSEK/yr
Total	46.7	39.5	53.9	38.9	MSEK/yr

Table 25 shows the CO<sub>2</sub> emissions reduction when replacing heat and electricity to the cluster by excess heat from a gasification unit. An integrated and a stand-alone unit are compared.

Table 25 CO<sub>2</sub> emissions reduction by utilising waste heat from an integrated and a stand-alone gasification unit

	Integrated	Stand-alone	
CO <sub>2</sub> emissions reduction from replaced electricity	7325	21596	t/yr
CO <sub>2</sub> emissions reduction from replaced steam	45528		t/yr
Total	52853	21596	t/yr

It can be seen that both economically and from a CO<sub>2</sub> emissions perspective it is advantageous to integrate a ethanol production unit in the cluster. Integration is advantageous compared to stand-alone operation in all scenarios. Depending on the scenario integration gives between 24.8 and 64.2 MSEK/yr higher incomes. The higher incomes can be used to cover the costs for integration. By integration 31.2 kt/yr more CO<sub>2</sub> emissions are replaced by CO<sub>2</sub> from biomass.

### 6.3.3 Biomass drying

To calculate the amount of biomass which can be dried with waste heat from the cluster certain assumptions have been made:

- Moisture content of the wet biomass is assumed to be 50 wt-%, typical values are between 30-60 wt-% depending on type location time of harvest, storage period.
- Drying to a final moisture content of 15 wt-% is assumed as this is the regular value for dried biomass for pelletising or other purposes such as feedstock for gasification (Johansson, Sara Larsson, and Wennberg 2004; Fagernäs et al. 2010).
- Ambient air conditions are assumed at 7 °C and a relative humidity of 70 % (as the spring/autumn case in (Johansson, Sara Larsson, and Wennberg 2004)
- If indirect drying (heat carrier-drying medium-drying goods) is applied a temperature difference between the process streams/heat carrier and heat carrier/drying medium of 10 K is assumed

With data on the energy system in Stenungsund gathered in the total site analysis study previously performed (Hackl and Andersson 2010) the total heat available for biomass drying in the cluster can be determined. The case with an improved utility system is considered (see 5.2.3) as it represents the amount of excess heat which is available if feasible energy efficiency measure are assumed to be implemented.

In Figure 23 the amount of heat available for biomass drying with low temperature excess heat is shown. The part of the red full curve which is marked as  $Q_{\text{drying}}$  is currently cooled by cooling water. Cooling water can be replaced by a drying heat carrier and thereby excess heat from the processes (full red line) will be transferred to a drying heat carrier (represented by the full black line). This can be water. The water is then used to in the dryer to heat the drying air flowing through the dryer and taking up water from the wet biomass (represented by the dashed black line). Figure 24 illustrates the transfer of excess heat from the cluster to the drying air via a drying heat carrier.

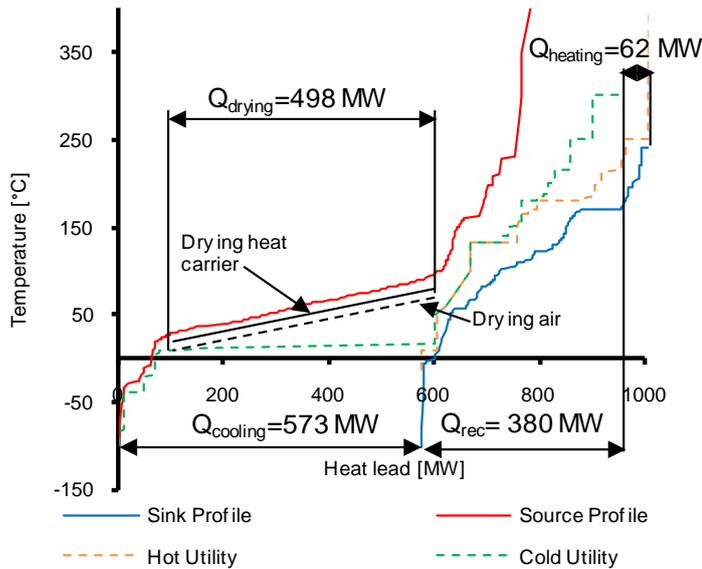


Figure 23 TSP showing heat available for biomass drying in the chemical cluster in Stenungsund if an optimised utility system is implemented

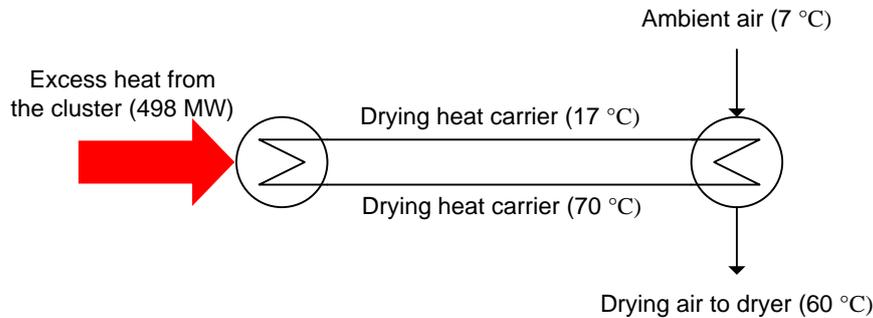


Figure 24 Illustration of excess heat transfer from the cluster to the biomass dryer by a heat transfer medium

As shown in Figure 23 the potential heat available for biomass drying is 498 MW of low temperature heat available in a temperature range suitable for biomass drying (between 98 °C and 27 °C). The hot water delivered to the dryer is heated by excess heat from the cluster from 17 to 70 C. This results in a air inlet temperature to the dryer of 60 °C.

The amount of excess heat which cannot be used for heating in cluster is used for estimating the potential for biomass drying. Analysis of the amount of biomass that can be dried with excess heat from the cluster gave the following results, see Table 26.

The heat available is  $Q_{\text{drying}}=498$  MW and the air inlet temperature to the dryer is  $T_{\text{drying air, in}}=60$  °C. The lower drying air temperature results in a higher heat demand for evaporation according to (Johansson, Sara Larsson, and Wennberg 2004, B-2). This results in a potential dry mass output of 150 kg<sub>dry mass</sub>/s with a moisture content of 15 wt-% (Incoming wet biomass: 255 kg<sub>wet biomass</sub>/s, moisture content 50 wt-%).

*Table 26 Amount of biomass possible to dry with the excess heat available in the cluster and the optimised utility system ( $Q_{drying}=498\text{ MW}$ ;  $T_{drying\ air, in}=60\text{ }^{\circ}\text{C}$ )*

Wet biomass input (50 wt-% H <sub>2</sub> O) in [kg/s]	255
Dried biomass output (15 wt-% H <sub>2</sub> O) in [kg/s]	150
Dry mass in [kg/s]	127
Water evaporated in [kg <sub>H2O</sub> /kg <sub>dry mass</sub> ]	0.82
Heat demand for evaporation in [kJ/kg <sub>H2O</sub> ]	4750
Water content (wet biomass) in [kg <sub>H2O</sub> /kg <sub>total biomass</sub> ]	0.50
Water content (after drying) in [kg <sub>H2O</sub> /kg <sub>total biomass</sub> ]	0.15

As shown in Table 26 there is a large potential for drying biomass utilising excess heat. Even when the processes are integrated as much as possible, as shown in Figure 23 there is a potential for producing **4.3\*10<sup>6</sup> tonnes dried biomass per year<sup>19</sup>** by using waste heat from the cluster.

It is of course not possible to deliver all the biomass to Stenungsund for drying, but it can be seen that there is a large potential for utilising industrial waste heat for drying of biomass.

### **Consequences of integration compared to stand-alone operation**

Typically in a stand-alone biomass drying unit some of the incoming biomass is combusted to produce the drying heat. In such a configuration the fuel requirements are 3050 kJ/kg<sub>H2O</sub> (Andersson, Harvey, and Thore Berntsson 2006). For the comparison of an integrated biomass drying with a stand-alone unit, the case with an improved utility system in the cluster is considered, see Table 26. An integrated dryer only uses waste heat from the cluster for drying of biomass.

The calculation of the amount of biomass necessary in a stand-alone unit for drying of wet biomass with a moisture content of 50 wt-% to 15 wt-% is shown in Table 27. The amount of wet biomass input is adapted from Table 26.

*Table 27 Biomass drying in a stand-alone unit: Calculation and results of energy and biomass fuel demand, assumed running time 8000 h/yr*

Heat demand for evaporation	3050	kJ/kg_H2O
Input of wet biomass	255	kg_wet biomass/s
Output of dry mass	127	kg_dry mass/s
Water evaporated	104.8	kg_H2O/s
Total heat demand for evaporation	319768	kW
Average heat content of wet biomass	9300	kJ/kg <sup>20</sup>
Amount of dry biomass for drying heat	34.4	kg_wet biomass/s
Percentage of incoming biomass for drying heat	13.5	%
CO <sub>2</sub> emission from biomass combustion	234	kg/t_dry mass

Table 28 shows that the costs for drying heat production in a stand-alone biomass drying unit are between 111 SEK/t\_dry mass and 188 SK/t\_dry mass, depending on

<sup>19</sup> Dried biomass with a moisture content of 15 wt-%

<sup>20</sup> from Engineering ToolBox (2010)

the scenario. This can be compared to the integrated case, where excess process heat is used which is assumed to be free as currently there is no alternative use for it. CO<sub>2</sub> emissions from biomass drying are calculated to 234 kg/t<sub>dry mass</sub>, as shown in Table 27. These emissions are calculated assuming that biomass is a limited resource. Biomass can be used in an alternative way to drying. Here it is assumed that the biomass can replace coal by cofiring in a coal power plant. These emissions can be avoided when using waste heat from the cluster.

*Table 28 Results from economic evaluation of a stand-alone biomass dryer*

Stand-alone dryer	2020		2030		SEK/t <sub>dry mass</sub>
	High fossil-fuel price/CO <sub>2</sub> emission charge level 3	Low fossil fuel price/CO <sub>2</sub> emission charge level 1	High fossil fuel price/CO <sub>2</sub> emission charge level 3	Low fossil fuel price/CO <sub>2</sub> emission charge level 1	
Economic value of biomass used for drying	144	111	188	116	

## 7 Conclusions

Advanced biorefinery concepts for integration with the chemical cluster in Stenungsund were qualitatively selected according to their potential for integration. This was followed by an evaluation of the consequences of integration with the cluster on energy efficiency, profitability and CO<sub>2</sub> emissions. Mass and energy balances for the selected biorefineries and the cluster were used to compare an integrated biorefinery with a stand-alone unit.

The selected biorefineries are:

- Gasification for syngas production
- Biomass to ethylene by fermentation of lignocellulosic raw material and ethanol dehydration
- Low temperature biomass drying for fuel upgrading

Syngas and ethylene are important raw materials for the cluster and the processes analysed show an excess of high temperature heat which can be used to cover the cluster's energy demand. Biomass drying was chosen as the cluster produces large amounts of waste heat, which in this way can be utilised.

The cluster has a steam deficit of 122 MW<sub>heat</sub> at pressure levels of 40 bar(g) and below. Gasification of biomass for syngas production yields excess heat which can be used to cover the clusters steam deficit. The gasification unit analysed had a capacity of 160 kt<sub>product</sub> gas/yr producing 144 GWh/yr excess heat. It was found that by integration with the cluster the excess heat can be used to produce 16 GWh<sub>el</sub>/yr and 124 GWh<sub>steam</sub>/yr for the cluster. In a stand-alone unit 47 GWh<sub>el</sub>/yr can be produced from the excess heat. Economic evaluation indicated increased incomes from heat and electricity of 18.3 to 47.4 MSEK/yr for the integrated case. CO<sub>2</sub> emissions reduction was calculated to be 24.4 kt<sub>CO<sub>2</sub></sub>/yr higher if the gasification plant is integrated.

When producing ethanol from lignocellulosic raw material by fermentation not all the incoming feedstock material is converted to ethanol. The process yields a substantial amount of by-products which can be used to supply energy to the process and for other purposes, e.g. supply heat and electricity to a downstream ethanol dehydration plant and the chemical cluster. The comparison of an integrated and a stand-alone unit showed that with 195.2 GWh/yr excess heat from the ethanol unit 21.8 GWh<sub>el</sub>/yr and 168 GWh<sub>steam</sub>/yr to the cluster and/or the ethanol-to-ethylene plant (integrated case) and 64.3 GWh<sub>el</sub>/yr (stand-alone) can be produced. This results in increased incomes of 24.8 to 64.2 MSEK/yr for the integrated case, depending on the scenario. Evaluation of CO<sub>2</sub> emissions reduction was 31.2 kt/yr higher when process integration is applied.

For biomass drying it was assumed that 498 MW waste heat from the cluster, which cannot be used for process heating, are available for drying purposes. The study showed that by utilising this heat 4.3\*10<sup>6</sup> tonnes dried biomass per year with a moisture content of 15 wt-% can be produced (moisture content of incoming biomass 50 wt-%). Utilising waste heat for the production leads to fuel costs savings depending

on the scenario between 111 SEK/t\_dry mass and 188 SK/t\_dry mass and 234 kg/t\_dry mass less CO<sub>2</sub> emissions compared to a stand-alone unit.

## 8 Future Work

In the gasification case only a CFB gasifier producing syngas at a CO:H<sub>2</sub> of ratio 1:2 was analysed. Other types of gasifiers with different product gas outputs should be analysed for integration. E.g. olefins can be produced by gasification with the MTO or methanol to propylene (MTP) processes.

When looking at biomass to ethylene by fermentation of lignocellulosic ethanol mostly literature data was used. In an ongoing Master thesis project at the division of Heat and Power Technology a more thorough investigation by applying detailed process simulation is conducted to improve the data for this process.

In the biomass drying part the overall potential for low temperature biomass drying is shown. Here the most interesting sources of waste heat should be identified and analysed economically.

In this study the advantages of integration of biorefinery concepts compared to stand-alone biorefineries are shown. Further economic investigations on the investments necessary for the biorefineries selected is necessary to be able to estimate the overall economic performance of the processes.

In order to get a better overview on the practical integration opportunities for biorefinery in the chemical cluster in Stenungsund, other integration aspects must also be discussed. Examples for those aspects are:

- To which degree is existing infrastructure sufficient
- Availability of feedstock such as forest residues, wood, straw, sugar plants, bio-ethanol etc.
- Availability of space
- Legislative issues, such as subsidies and site extension permission

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## 10 Appendix



# 10.1 Appendix 1

## Calculation related to biomass drying:

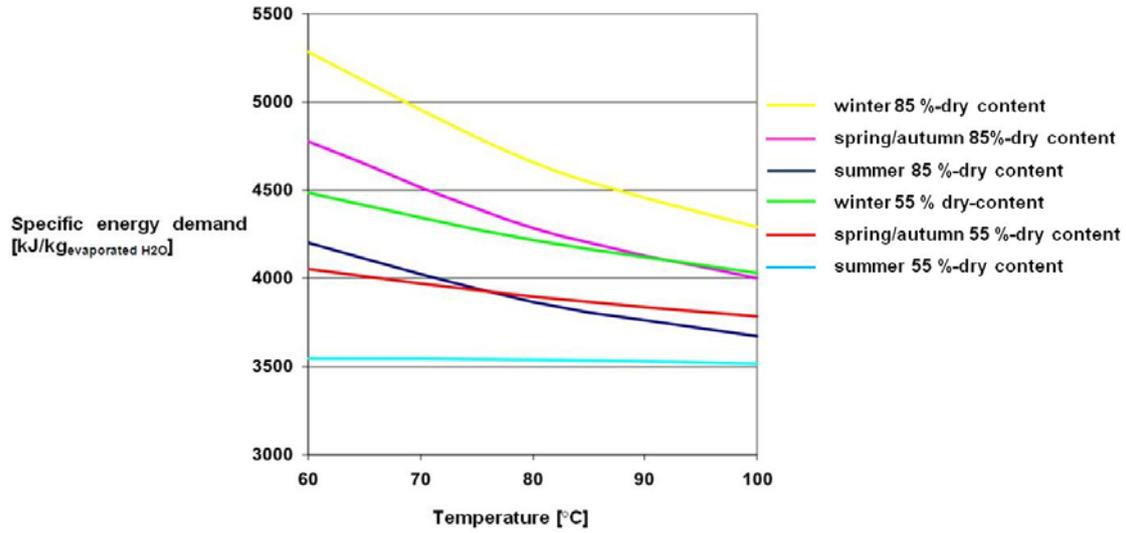


Figure 25 Specific energy demand depending on the drying temperature, the final dry content of biomass and the season (Johansson, Sara Larsson, and Wennberg 2004)

### Calculation of the amount of biomass possible to dry with low temperature heat:

- Water content on dry basis:

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- Amount of water evaporated during drying:

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- Dry mass of biomass:

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- Wet biomass input:

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- Dried biomass output:

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## 10.2 Appendix 2

Data from gasification simulations (Eva K. Larsson 2010)

*Table 29 Composition of incoming biomass*

Mass flow	22.27 kg/s
Composition	[wt-%]
C	50.68
H	6.02
O	40.64
N	0.5
S	0.04
Cl	0.01

*Table 30 Gas composition of product gas from gasification after the reformer*

Mass flow	22.27 kg/s
Temperature	950 °C
Pressure	29 bar
Gas composition	[mol-frac]
CH <sub>4</sub>	0,0047
CO	0,1177
CO <sub>2</sub>	0,0969
H <sub>2</sub>	0,1672
H <sub>2</sub> O	0,5967
N <sub>2</sub>	0,0167

*Table 31 Syngas properties after WGS-reactor*

Mass flow	22.27 kg/s
Temperature	330 °C
Pressure	28 bar
Gas composition	[mol-frac]
CH <sub>4</sub>	0,0047
CO	0,0949
CO <sub>2</sub>	0,1196
H <sub>2</sub>	0,1900
H <sub>2</sub> O	0,5740
N <sub>2</sub>	0,0167

*Table 32 Syngas properties after CO<sub>2</sub>-removal*

Temperature	40 °C
Mass flow	4.29 kg/s
Annual production	123.55 kt/yr
Gas composition	[mol-frac]
CH <sub>4</sub>	0.0153
CO	0.3076
CO <sub>2</sub>	0.0077
H <sub>2</sub>	0.6153
N <sub>2</sub>	0,0541

Table 33 Pinch analysis stream data for the gasification system

Stream	T-start in [°C]	T-target in [°C]	Q [kW]	$\Delta T_{\min}$ in [K]
Preheating demineralised water	15	250	4807	10
Steam preparation	250	251	8397	5
Steam superheating	251	450	2155	20
Syngas cooling HX1	950	330	28336	20
WGS-reactor	330	329	998	20
Cooling after WGS-reactor HX3	330	198,4	5615	20
Cooling/Condensation WGS-reactor HX4	198,4	40	35538	10
Oxygen heating	25	200	763	20
MEA reboiler (CO <sub>2</sub> separation)	115	116	22599	5
MEA condenser	90	40	4520	5

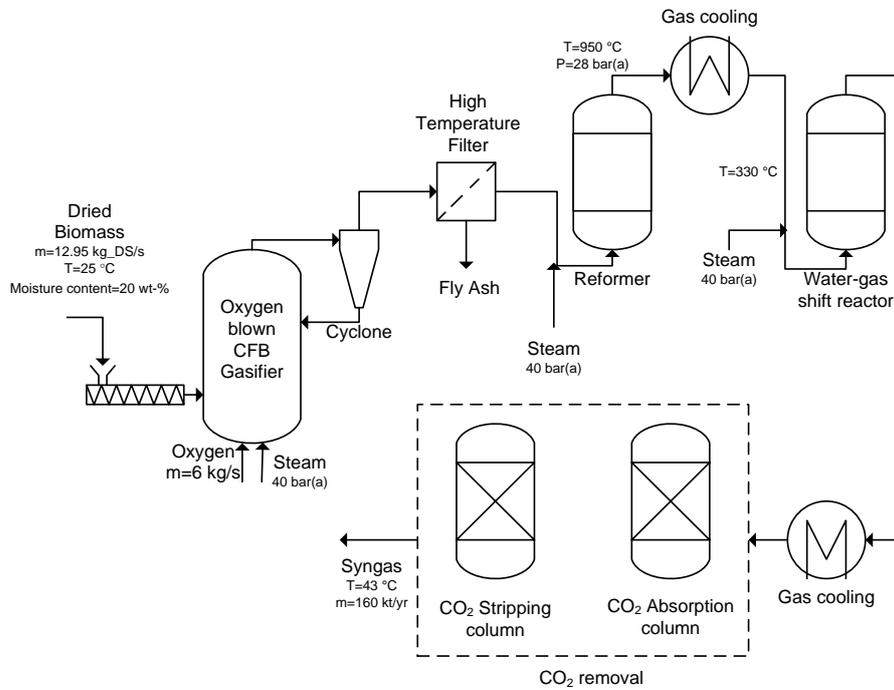


Figure 26 Flowsheet of the biomass oxygen blown CFB-gasification unit

### 10.3 Appendix 3

Table 34 Prices for steam, electricity and fuel assumed for economical evaluation of energy efficiency measures in the previous TSA study (Hackl and Andersson 2010)

<b>Assumed prices:</b>	<b>SEK/MWh</b>
<b>Saved steam</b>	300
<b>Electricity</b>	600
<b>Additional Fuel for steam production</b>	240

Table 35 Economical evaluation of energy efficiency measures suggested in the report with an assumed running time of 8000 h/year

<b>Cases</b>	<b>Steam savings [MW]</b>	<b>Electricity produced/saved [MW]</b>	<b>Value [SEK/h]</b>	<b>Incomes [MSEK/y]</b>
<b>Introducing only hotW,</b>	51		15300.0	122.4
<b>Increased steam recovery</b>	33		9900.0	79.2
<b>hotW moderate changes</b>	55.2		16560.0	132.5
<b>hotW moderate and more complex changes</b>	83.5		25050.0	200.4
<b>Steam recovery with moderate changes</b>	4.24		1272.0	10.2
<b>Steam recovery with moderate and more complex changes</b>	26.6		7980.0	63.8
<b>Utilising low-T evaporators at Borealis</b>		2.1	1260.0	10.1

## 10.4 Appendix 4

Calculation of ethanol yield from fermentation:

Input data from (Eriksson and Kjellström 2010)

*Table 36 Input data for the calculation of ethanol yield by fermentation of lignocellulosic raw material*

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Biomass input (moisture content 50 wt-%)	1024 t/yr
Biomass input (dry biomass)	512 t/yr
Ethanol production	119 t/yr
Density ethanol	0.789 g/cm <sup>3</sup>
Yield ethanol/dry biomass	295 l-EtOH/ton dry biomass

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## 10.5 Appendix 5

Table 37 Assumptions and calculations to compare the utilisation of excess heat from the gasifier for an integrated and a stand-alone unit

Integrated gasifier				Stand-alone		
Assumptions	Cogeneration of heat and power			Power production		
	HP-steam generation	121	bar(a)	HP-steam generation	121	bar(a)
	T superheated steam	540	C	T superheated steam	540	C
	Isentropic efficiency turbine	0.8		Isentropic efficiency	0.8	
	Generator efficiency	0.98		Generator efficiency	0.98	
	Isentropic efficiency pump	0.8		Isentropic efficiency pump	0.8	
	Steam delivered to the cluster at	41	bar(a)	Expansion to	0.1	bar(a)
	Running time	8000	h	Running time	8000	h
<hr/>						
Calculations	Excess heat from gasification	18000	kW	Excess heat from gasification	18000	kW
Turbine	$h_{120\text{bar}(g),540\text{ C}}$	3455	kJ/kg	$h_{120\text{bar}(g),540\text{ C}}$	3455	kJ/kg
	$s_{120\text{bar}(g),540\text{ C}}$	6.62	kJ/kgK	$s_{120\text{bar}(g),540\text{ C}}$	6.62	kJ/kgK
	$h_{40\text{bar}(g),6.62\text{kJ/kgK}}$	3121	kJ/kg	$h_{40\text{bar}(g),6.62\text{kJ/kgK}}$	2096	kJ/kg
	$h_{\text{real,after turb}}$	3188	kJ/kg	$h_{\text{real,after turb}}$	2368	kJ/kg
	$dh_{\text{is,turb}}$	333	kJ/kg	$dh_{\text{is,turb}}$	1359	kJ/kg
	$dh_{\text{turb}}$	266	kJ/kg	$dh_{\text{turb}}$	1087	kJ/kg
	$T_{\text{turbine outlet}}$	390	C	$T_{\text{turbine outlet}}$	46	C
	$P_{\text{mech,turb}}$	2051	kW	$P_{\text{mech,turb}}$	6047	kW
Electricity	$P_{\text{el,turb}}$	2010	kW	$P_{\text{el,turb}}$	5927	kW
Heat	$Q_{\text{to cluster}}$	16000	kW	$Q_{\text{condenser}}$	12000	kW
Pump	$h_{\text{condensateLP}}$	1095	kJ/kg	$h_{\text{condensateLP}}$	1912	kJ/kg
	$s_{\text{condensateLP}}$	2.81	kJ/kgK	$s_{\text{condensateLP}}$	0.649	kJ/kgK
	$h_{\text{is,comp}}$	1104.6	kJ/kg	$h_{\text{is,comp}}$	204.0	kJ/kg
	$dh_{\text{is,comp}}$	9.98	kJ/kg	$dh_{\text{is,comp}}$	12.17	kJ/kg
	$dh_{\text{comp}}$	12.5	kJ/kg	$dh_{\text{comp}}$	15.2	kJ/kg
	$h_{\text{after comp}}$	1117	kJ/kg	$h_{\text{after comp}}$	219	kJ/kg
	$T_{\text{after comp}}$	256	kJ/kg	$T_{\text{after comp}}$	49.9	kJ/kg
	$P_{\text{mech,pump}}$	74.73	kW	$P_{\text{mech,pump}}$	65.84	kW
Evaporator	$h_{\text{after preheat}}$	1495	kJ/kg	$h_{\text{after preheat}}$	1495	kJ/kg
	$Q_{\text{preheat}}$	2913	kW	$Q_{\text{preheat}}$	5522	kW
	$Q_{\text{evap}}$	9148	kW	$Q_{\text{evap}}$	5141	kW
	$Q_{\text{superheat}}$	5939	kW	$Q_{\text{superheat}}$	3337	kW
	$m_{\text{steam}}$ in kg/h	27720	kg/h	$m_{\text{steam}}$ in kg/h	20028	kg/h
	$m_{\text{steam}}$ in kg/s	7.70	kg/s	$m_{\text{steam}}$ in kg/s	5.56	kg/s

## 10.6 Appendix 6

Table 38 Assumptions and calculations to compare the utilisation of excess heat from a lignocellulosic ethanol production plant for an integrated and a stand-alone unit

Integrated gasifier				Stand-alone		
Assumptions	Cogeneration of heat and power			Power production		
	HP-steam generation	121	bar(a)	HP-steam generation	121	bar(a)
	T superheated steam	540	C	T superheated steam	540	C
	Isentropic efficiency turbine	0.8		Isentropic efficiency	0.8	
	Generator efficiency	0.98		Generator efficiency	0.98	
	Isentropic efficiency pump	0.8		Isentropic efficiency pump	0.8	
	Steam delivered to the cluster at	41	bar(a)	Expansion to	0.1	bar(a)
	Running time	8000	h	Running time	8000	h
<hr/>						
Calculations	Excess heat from gasification	24400	kW	Excess heat from gasification	24400	kW
Turbine	$h_{120\text{bar}(g),540\text{ C}}$	3455	kJ/kg	$h_{120\text{bar}(g),540\text{ C}}$	3455	kJ/kg
	$s_{120\text{bar}(g),540\text{ C}}$	6.62	kJ/kgK	$s_{120\text{bar}(g),540\text{ C}}$	6.62	kJ/kgK
	$h_{40\text{bar}(g),6.62\text{kJ/kgK}}$	3121	kJ/kg	$h_{40\text{bar}(g),6.62\text{kJ/kgK}}$	2096	kJ/kg
	$h_{\text{real,after turb}}$	3188	kJ/kg	$h_{\text{real,after turb}}$	2368	kJ/kg
	$dh_{\text{is,turb}}$	333	kJ/kg	$dh_{\text{is,turb}}$	1359	kJ/kg
	$dh_{\text{turb}}$	266	kJ/kg	$dh_{\text{turb}}$	1087	kJ/kg
	$T_{\text{turbine outlet}}$	390	C	$T_{\text{turbine outlet}}$	46	C
	$P_{\text{mech,turb}}$	2781	kW	$P_{\text{mech,turb}}$	8198	kW
Electricity	$P_{\text{el,turb}}$	2725	kW	$P_{\text{el,turb}}$	8034	kW
Heat	$Q_{\text{to cluster}}$	21400	kW	$Q_{\text{condenser}}$	16000	kW
Pump	$h_{\text{condensateLP}}$	1095	kJ/kg	$h_{\text{condensateLP}}$	1912	kJ/kg
	$s_{\text{condensateLP}}$	2.81	kJ/kgK	$s_{\text{condensateLP}}$	0.649	kJ/kgK
	$h_{\text{is,comp}}$	1104.6	kJ/kg	$h_{\text{is,comp}}$	204.0	kJ/kg
	$dh_{\text{is,comp}}$	9.98	kJ/kg	$dh_{\text{is,comp}}$	12.17	kJ/kg
	$dh_{\text{comp}}$	12.5	kJ/kg	$dh_{\text{comp}}$	15.2	kJ/kg
	$h_{\text{after comp}}$	1117	kJ/kg	$h_{\text{after comp}}$	219	kJ/kg
	$T_{\text{after comp}}$	256	kJ/kg	$T_{\text{after comp}}$	49.9	kJ/kg
	$P_{\text{mech,pump}}$	130	kW	$P_{\text{mech,pump}}$	115	kW
Evaporator	$h_{\text{after preheat}}$	1495	kJ/kg	$h_{\text{after preheat}}$	1495	kJ/kg
	$Q_{\text{preheat}}$	3949	kW	$Q_{\text{preheat}}$	9624	kW
	$Q_{\text{evap}}$	12400	kW	$Q_{\text{evap}}$	8959	kW
	$Q_{\text{superheat}}$	8051	kW	$Q_{\text{superheat}}$	5817	kW
	$m_{\text{steam}}$ in kg/h	37576	kg/h	$m_{\text{steam}}$ in kg/h	27149	kg/h
	$m_{\text{steam}}$ in kg/s	10.4	kg/s	$m_{\text{steam}}$ in kg/s	7.5	kg/s

## 10.7 Appendix 7

Table 39 Input data for the energy market scenarios (Harvey and Axelsson 2010)

Fossil fuel prices <sup>1</sup> (€/MWh)		2010	2020	2030	2040	2050
Oil	low	23	24	29	31	31
	high	23	39	49	55	57
Natural gas	low	18	19	22	24	25
	high	18	27	37	42	44
Coal	low	7,0	7,1	7,5	7,6	7,7
	high	7,0	8	10	11	11
<b>Policy instruments</b>						
CO <sub>2</sub> emission charge (€/ton)	level 1	20	15	15	15	15
	level 2	20	20	27	37	50
	level 3	20	30	45	68	101
	level 4	20	52	85	117	150
RES-E support <sup>2</sup> (€/MWh)		20	20	20	20	20