Methanol via biomass gasification

Thermodynamic performances and process integration aspects in Swedish chemical cluster and pulp and paper sites

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SUMMARY

This work was conducted within the Swedish Skogskemi (“forest chemicals”) project funded by Vinnova and aiming at investigating promising and competitive options for biomass based production of chemical intermediates such as olefins, methanol and butanol. The present report documents the contribution of Matteo Morandin to project work package on biomass gasification routes for methanol production and in particular focuses on mass and energy balances of three process concepts based on different gasification and gas cleaning technologies. In addition, the report discusses process integration opportunities for the biorefinery processes with industrial plants, such as the Stenungsund chemical cluster and the Värö and Iggesund pulp and paper mills, all of which were actively involved in the Skogskemi project. The characteristics of the three processes (gasification technology, location, size) were decided ex ante based on preliminary economic considerations such as maturity of technology and economy of scale aspects and represent an input to this work. The complete process layouts for the three plants were put together based on typical arrangements suggested in the literature for similar processes and on established engineering principles. The mass and energy balances of the three process flow-sheets were estimated with the help of Aspen plus and process models partially or completely available in the literature and partially developed at the Division of Industrial Energy Systems and Technologies, Chalmers. Following a rather common approach for preliminary design of chemical processes, the heating or cooling required to carry out the various thermochemical conversions from biomass to methanol were modelled as heaters and coolers. This allowed applying an energy targeting approach, Pinch Analysis, to estimate ideal heat recovery opportunities for the biomass based processes. As a result of the heat recovery analysis, the net heat available from the biomass based conversion processes (target) was obtained. Different opportunities for either exporting this heat to partially or fully replacing the steam deficit of the nearby industrial plants or for power generation by means of a heat recovery steam cycles are discussed. The thermodynamic performances of the three plants are then compared considering the combined effects of the material conversion from biomass to methanol and the fossil fuel savings in marginal heat and electricity producers as a consequence of the new heat and power balances at the industrial site once the biomass based processes are in place. Significant process synthesis and design variables are finally discussed and recommendations are provided for further investigations.

Keywords: Methanol, Biomass, Gasification, Simulation, Pinch Analysis

Contents

Preface

The present work was conducted within the Swedish Skogskemi ("forest chemicals") project funded by Vinnova and aiming at investigating promising and competitive options for biomass based production of chemical intermediates such as olefins, methanol and butanol (Joelsson, Engström et al. 2015).

The project involved major Swedish industrial stakeholders from the forest and chemical sectors, as well as research institutes and universities and was carried out from September 2012 to September 2014. The activities were grouped in three subprojects, each further subdivided in working packages (WP). Subproject 1 dealt with primary conversion of biomass into methanol via thermochemical gasification (WP1.1) and into ethanol via fermentation (WP1.2). Subproject 2 dealt with production of chemical intermediates such as: olefins via methanol to olefin technology or ethanol dehydration (WP2.1), butanol via ethanol and acetaldehyde (WP2.2), and chemical grade methanol by recovery of off-gases from black liquor evaporator strippers in pulp and paper mills (WP2.3). Subproject 3 dealt with analysis of system aspects of the proposed biomass to chemical value chains. Subproject 4 dealt with organization of discussion platforms and outreach.

The present report documents the contribution of Matteo Morandin to WP1.1 (methanol via biomass gasification) and in particular focuses on mass and energy balances of three methanol production concepts based on different gasification and gas cleaning technologies. In addition, the report discusses process integration opportunities for the biorefinery processes with industrial plants, such as the Stenungsund chemical cluster and the Värö and Iggesund pulp and paper mills, all of which were actively involved in the Skogskemi project.

This work is largely based on previous and parallel research projects at the div. of Industrial Energy Systems and Technologies, Chalmers, dealing with conceptual design and integration of biorefineries, especially concepts based on indirect biomass gasification. Project partners at the div. of Energy Technology, Luleå University, contributed with useful inputs about modelling of entrained flow gasifier. Project partners at the dept. of Applied Physics and Electronics, Umeå University, contributed with useful mass and energy balance data for biomass torrefaction. Eva Andersson at CIT Industriell Energi AB, project leader for WP1.1, also contributed with input about process concepts and by critically reviewing this report. Finally, Simon Harvey, Thore Berntsson and several Ph.D. students at the div. of Industrial Energy Systems and Technologies contributed with interesting discussion of results.

This is a revised version of the original report submitted to Vinnova.

Göteborg, June 2015

Matteo Morandin

Notations

Abbreviations

ASU	Air separation unit
CFB	Circulating fluidized bed
CHP	Combined heat and power
EtOH	Ethanol
HEN	Heat exchanger network
HHV	Higher heating value
LHV	Lower heating value
MTO	Methanol to Olefins
MeOH	Methanol
NG	Natural gas
SNG	Substitute natural gas
SR	Methanol synthesis ratio = $(H_2 - CO_2) / (CO + CO_2)$

Symbols

η	Conversion efficiency (biomass to methanol)
η_{total}	Total conversion efficiency
η_q	Boiler thermal efficiency
η_{el}	Electrical efficiency
η_{CG}	Gasifier cold gas efficiency
$\dot{Q}_{saved}$	Heat savings at industrial site by heat integration
$\dot{W}_{import}$	Electricity import at the industrial site
$\dot{m}$	mass flow rate

1 Introduction

Methanol is currently mostly produced via synthesis starting from hydrogen and carbon monoxide or carbon dioxide mixtures such as natural gas derived syngas.

Methanol synthesis technologies are rather well established with the research and development mainly focusing on reactor thermal management, lower synthesis pressures and synthesis from carbon monoxide or carbon dioxide rich syngas.

Methanol can be directly produced from biomass via pyrolysis and subsequent separation. However, much higher selectivity and larger process scale can today be obtained via biomass gasification, syngas upgrading, and subsequent conventional methanol synthesis. Major issues are in this case concentrated in the first steps of this route, biomass gasification and gas cleaning, which are well known processes but still expensive and less reliable compared to the fossil based syngas production technologies. To improve process economics, high conversion efficiency and process reliability are necessary.

In this work, the thermodynamic performance of three gasification systems for conversion of biomass into syngas and subsequent methanol synthesis are estimated based on process simulations. Considering the current status of the research in biomass gasification in Sweden, three types of biomass gasifiers were investigated: atmospheric indirect (dual bed) fluidized bed, pressurized direct oxygen steam blown circulating fluidized bed, and pressurized entrained flow (of torrefied biomass).

Since methanol is ultimately considered for use as a chemical intermediate at the chemical cluster in Stenungsund in this project (Joelsson, Engström et al. 2015), this location is considered of primary importance and is where two of the investigated systems are assumed to be placed. Although other possible locations in Sweden may provide better logistics for biomass harvesting and transportation, the chemical cluster in Stenungsund is a large steam consumer and therefore interesting heat integration opportunities may appear between the biomass based processes and the cluster. For comparison, a third system is assumed to be located nearby the Iggesund pulp and paper plant, in which case methanol is then transported to the Stenungsund cluster either by train or boat.

Another way to improve the logistics of the biomass based methanol production could be to convert biomass into a more transportable fuel. Two alternatives are therefore considered here: (a) production of substitute natural gas (SNG) collocated nearby a pulp production site (Värö) with subsequent SNG transportation to Stenungsund by means of the NG pipe on the Swedish West Coast: (b) decentralized biomass torrefaction and transportation of torrefied material to the Stenungsund cluster.

While it is clear that the ultimate comparison should be based on economic performance, a thermodynamic assessment of the three proposed solutions is of importance for conceptual design and for further adjusting the design parameters. It is clear, in particular, that the methanol yield is an important figure-of-merit for such comparisons. However, it is also important to conduct a careful evaluation of the opportunities for excess heat utilization for the three separate solutions. In fact, the methanol energy yield is inevitably less than 100% if based on biomass gasification since a part of the biomass energy is lost in form of heat, partially due to the intrinsic losses of thermochemical conversions, which can be recovered in different forms depending on the temperature levels.

The report is structured in three main chapters. The methodology is presented in chapter 2. The energy and mass balances of the three system alternatives are presented in chapter 3, each in separate sections, together with the discussion of their respective heat recovery opportunities. The performances of the three alternatives are compared in the chapter 4 and the main aspects of uncertainty for further implementation of the proposed processes are discussed. An appendix is also provided with detailed process layouts and estimated compositions and thermodynamic characteristics of major process streams.

2 Methodology

2.1 Definition of process configurations

The feasibility study of the production of methanol from gasification of biomass focuses on three process configurations. Important process characteristics such as type, size, number, and connections of equipment units were chosen based on heuristics due to time constraints which did not allow a rigorous comparison between a larger set of options.

The configurations were formulated considering some realistic options regarding feedstock, location, and gasification technologies that are commonly considered in the literature and that are of interest for the specific features of the Skogskemi project which aims at investigating practical routes for conversion of woody biomass (mainly Swedish forest feedstock) into chemical intermediates for further processing in the Stenungsund chemical cluster (Joelsson, Engström et al. 2015).

The following lists and paragraphs provide an overview of these options.

2.1.1 Feedstock

The following three lignocellulosic feedstock types are of major interest for methanol production via gasification.

- Wood chips:
    - 50% moisture.
    - Ultimate analysis: ash 0.156, C 51.9, H 6, N 0.12, Cl 0.015, S 0.009, O 41.8.
    - Higher heating value (HHV) 10.431 MJ kg^{-1}, lower heating value (LHV) 8.555 MJ kg^{-1}.
- Forest residues:
    - 40 % moisture.
    - Ultimate analysis: ash 2.869, C 51.6, H 6, N 0.48, Cl 0.015, S 0.036, O 39.
    - Higher heating value (HHV) 12.592 MJ kg^{-1}, lower heating value (LHV) 10.830 MJ kg^{-1}.
- Wood pellets:
    - 10 % moisture.
    - Ultimate analysis: ash 0.156, C 51.9, H 6, N 0.12, Cl 0.015, S 0.009, O 41.8.
    - Higher heating value (HHV) 19.984 MJ kg^{-1}, lower heating value (LHV) 17.573 MJ kg^{-1}.

2.1.2 Location

Three locations were considered relevant for methanol production and further upgrading into chemical intermediates.

- Near the Stenungsund cluster, due to possibilities of integration with the existing infrastructures and availability of harbour docks for feedstock delivery. This is also the place where the methanol to olefins (MTO) plant is assumed to be located and therefore where the integration with the methanol production process is the most straightforward to implement.
- Near a pulp and paper plant, where biomass delivery infrastructure is already present and integration is possible with the existing utility steam network, power network and district heating network. Access to natural gas network should be preferred.
- Central Sweden, where density of biomass is high and biomass supply chains are well developed.

2.1.3 Gasification technologies

Technical barriers for biomass gasification are well known (Heyne, Liliedahl et al. 2013).

Major issues are:

- The biomass feeding system for pressurized gasification and when grinding is required.
- Tar concentration in the product gas which can be detrimental for subsequent process equipment.
- Handling of the ashes which are highly corrosive and have lower melting points compared to coal.

Biomass atmospheric air gasification in fluidized bed is the most conventional technology at large scale. This technology is, however, mainly suitable for heat and power applications due the large amount of nitrogen in the product gas. The fluidized bed technology is very similar to that of the fluidized bed boilers which have been implemented in very large scales for coal combustion applications. The product gas from biomass air gasifiers is often used to revamp fossil fuel boilers and an application that can be regarded as an advance case of co-combustion. There are few cases of stand-alone integrated biomass gasification combined cycles (BIGCC).

More difficult is the use of biomass gasification for production of chemicals and fuels where high syngas purity is required and where pressurization and concentration of CO and H_2 is critical to guarantee reliability and optimal process economics.

This restricts the technological choices in this work to two major gasification concepts: indirect (dual bed) gasification and O_2-blown gasification. In the first case, the heat to support the endothermic gasification reactions is transported from a combustion zone to a gasification zone by recirculating bed material. The most known demonstration plants of this gasification technologies are: FICFB Güssing Wien (Hofbauer, Rauch et al. 2002), Rentech/Silvagas/Battelle Ferco, Vermont USA (Farris, Paisley et al. 1998), Milena (Deurwaarder, Boerrigter et al. 2005).

A broader range of options are available for O_2-blown gasification.

Although direct biomass gasification in circulating fluidized bed reactors has not been demonstrated at pilot scale, the concept appears equally promising as indirect gasification and more standard air gasification technologies could be converted to accommodate higher concentrations of O_2. Major technology providers investigating this option are ThyssenKrupp/Uhde (Ecotraffic 1997) and Foster&Wheeler (Bengtsson 2011).

Alternatively, two-stage concepts based on a first low temperature biomass devolatization and a subsequent high temperature entrained flow gasifier can be used. The two stages can be combined in a single plant (Greil, Hirschfelder et al. 2002, Althapp 2003) or decoupled. In this latter case, decentralized pyrolysis or torrefaction can take advantage of better feedstock supply logistics and allow increasing the energy density of the biomass into pyrolysis oil or torrefied biomass before transportation to the final syngas production site, e.g. in a centralized entrained flow gasifier (Weiland, Nordwaeger et al. 2014).

The concept is similar to that of black liquor gasification, where an entrained flow gasifier is used for syngas production and simultaneous recovery of Kraft chemicals. Woody biomass is pre-treated in the Kraft process for the separation of cellulose and hemicellulose and a lignin rich residue (black liquor) is obtained. Since black liquor gasification can be considered only as coupled to further processing of the cellulose part (e.g. pulp, paper, ethanol, etc.), it should be considered a special case of biomass gasification and is not analysed in depth in this work. Nevertheless, black liquor gasification represents one of the most interesting outcomes of gasification R&D in Sweden and large investigations were conducted in recent decades. This should be considered as an important case of comparison with the gasification based processes proposed here. The reader is referred to (Ekbom, Lindblom et al. 2003) for more details about this technology.

Another option for improving the logistics of the biomass conversion into fuels or chemicals is to produce synthetic natural gas (SNG) which has properties similar to natural gas and therefore can be transported by means of existing natural gas pipes (Heyne, Thunman et al. 2012). SNG production can be located in a different place than the methanol and olefins production thus allowing improved logistics. A reforming stage is used for syngas production prior to final methanol synthesis. This arrangement allows using natural gas as a backup in case biomass based SNG is not available and the NG pipe as a buffer system.

2.1.4 Size

The size of the biomass gasification plants in this study was decided based on the possibility of supplying a considerable part of the feedstock for a methanol to olefin plant (MTO) assumed to be located at the Stenungsund cluster in order to replace part of the olefins currently produced via naphtha cracking.

In order to produce 200 kt per year of olefins, more than 450 kt per year of methanol are required, which corresponds to a chemical energy rate of around 350 MW. With an average energy conversion efficiency from biomass to methanol of 55%, about 650 MW of biomass would be required, which was not considered as feasible given the current status of biomass gasification technology. A single investment in a plant with a capacity of a few hundreds of MW was considered to be more reasonable in a 2030 horizon for Sweden. As for the remaining share of methanol to produce the desired quantity of olefins considered in the Skogskemi project by means of the MTO technology, other sources of methanol must be considered, such as e.g. another biomass gasifier located somewhere else or methanol purchased in the global market (Joelsson, Engström et al. 2015).

To improve reliability and availability of the biomass based syngas, it is more convenient to split the overall biomass gasification operation into a parallel train of gasifiers. In this way, smaller units can be used that can be assumed to be commercially available within a short time horizon. In addition, the units can be slightly oversized so that they can operate in less demanding conditions. Equipment sizing and investment cost estimation are however beyond the scope of the present work and are not discussed in detail here but were object of parallel investigation within the Skogskemi project (Joelsson, Engström et al. 2015).

2.1.5 Selected process concepts

Based on the above recommendations, the following three system configurations were identified:

1. Indirect gasification and SNG production located in proximity of the Värö pulp mill on the Swedish West Coast (wood chips feedstock capacity 200 MW_{LHV}, 245 MW_{HHV}), plus indirect gasification located in Stenungsund (wood chips feedstock capacity 150 MW_{LHV}, 183 MW_{HHV}) for separate production of syngas to supplement the SNG derived syngas required for methanol synthesis.
2. Direct oxygen-steam blown pressurized CFB gasifiers and methanol synthesis located in proximity of the Iggesund pulp mill on the Swedish East Coast. (wood chips feedstock capacity 450 MW_{LHV}, 550 MW_{HHV}).
3. Two torrefaction plants each with a biomass input of 150 MW_{LHV} located in the centre and in the north of Sweden and transportation of torrefied biomass to Stenungsund. A third torrefaction plant with another 150 MW_{LHV} biomass input is also placed in Stenungsund. The torrefied biomass is then processed through entrained flow gasifiers to provide syngas for methanol synthesis in proximity of the Stenungsund cluster.

2.2 Definition of process layouts and mass and energy balances

The process layouts of the three system configurations were defined according to the main characteristics (gasifier type, size, location) discussed above and based on established designs proposed in the literature and engineering principles.

The methanol synthesis and methanol distillation layouts assume a conventional fixed bed reactor at medium/high pressure (90 bar) working at around 250°C, with a synthesis ratio SR at around 2.05, and carbon dioxide concentration at around 3% molar (Supp 1990). Methanol distillation is achieved with a sequence of two columns (first pressurized and second atmospheric) to obtain a water free methanol stream.

The upstream process layouts and their main design parameters were defined according to the following principles.

2.2.1 System configuration nr 1

In system configuration 1, an atmospheric indirect gasifier working at 850°C is considered.

Based on literature data (Milne, Abatzoglou et al. 1998), the tar content in the product gas of fluidized bed gasifiers is in the range 5 to 10 $g \cdot Nm^{-3}$, although lower content can also be obtained with a catalytic bed (Hofbauer, Rauch et al. 2002). The methane content can also be quite high (up to 10% dry gas).

The dual bed indirect gasifier has been successfully demonstrated and can be considered to be one of the most established technologies for production of biomass based syngas with low nitrogen content. A conservative approach was taken for the design of the gas cleaning and upgrading sections. A cold gas cleaning section based on an RME scrubber was assumed, and the RME plus the removed tar are thereafter burnt in the combustor required to reheat the circulating gasifier bed material. The tar free gas is then pressurized to reduce downstream equipment volumes.

Due to the large amount of methane in the product gas and since methane is inert to methanol synthesis, methane reforming stage is considered. A water gas shift reactor is then used to increase the hydrogen to carbon monoxide ratio thus increasing the carbon dioxide content in the gas.

Carbon dioxide is then removed prior to methanol synthesis by means of a Rectisol unit which uses methanol as solvent, and the gas is finally compressed to synthesis pressure levels.

This layout was slightly modified for the SNG plant located in Värö where methane reforming is avoided, carbon dioxide is removed by a conventional amine wash, and water gas shift is used to adjust the hydrogen to carbon monoxide content to the optimal level for methane synthesis. A triple fixed bed synthesis reactor is assumed. SNG is delivered at around 30 bar to the NG network.

In Stenungsund (S)NG is further reformed in a conventional NG reformer based on adiabatic pre-reformer and an autothermal reformer with carbon dioxide removal and partial carbon dioxide recycle.

The syngas from biomass gasification and from (S)NG reforming is then mixed, and sent to methanol synthesis.

2.2.2 System configuration nr 2

In system configuration 2, an oxygen steam blown pressurized CFB gasifier is assumed.

This is regarded as an advanced biomass gasification configuration since the technology has not yet been demonstrated at pilot scale. The design considered here is based on the concept proposed by the VTT Technical Research Centre of Finland (Hannula and Kurkela 2012).

Biomass wood chips are fed into the gasifier through a piston system. The gasifier works at around 25 bar with a steam to oxygen ratio of around 1.0 and with almost complete carbon conversion.

Tar and methane are also present in the product gas. An advanced CFB catalytic reformer is considered for tar conversion where part of the product gas is used for reheating the reformer bed.

An autothermal reformer is also considered to convert the residual methane followed by water gas shift and subsequent Rectisol carbon dioxide removal.

2.2.3 System configuration nr 3

In system configuration 3, torrefied material is gasified in a pressurized entrained flow gasifier in presence of oxygen.

Torrefied biomass is ground and injected into the gasifier through a lock-hopper system requiring inert gas for compression (here recycled carbon dioxide).

Due to the high gasification temperature, tar compounds are absent or present in negligible quantities in the product gas, thus no tar reformer or scrubber was considered in this system configuration. Additionally, very little methane is obtained in the gas and methane reforming is avoided.

Water gas shift and carbon dioxide removal are required prior to methanol synthesis. Part of the fresh syngas is diverted directly after the gasifier to quench the gas to prevent high temperature corrosion of downstream equipment.

2.3 Remarks on process modelling and simulation

Modelling of biomass gasification, syngas upgrading, reforming, methanol synthesis was done in collaboration with PhD students and the Div. of Industrial Energy Systems and Technologies within parallel research projects.

For sake of brevity, the mass and energy balances were estimated for wood chips only. A discussion is provided about the possible effects of other types of feedstock (forest residues, pellets) in Chapter 4.

A detailed description of all assumptions and energy and mass balances of the three system alternatives is provided in the next chapter. Detailed simulation results for the different process configurations can be found in the Appendix.

2.4 Heat integration analysis

The processes investigated in this project consist primarily of conversion of biomass feedstock streams into chemical intermediates or fuels. Most of these processes are based on thermochemical conversions which require or produce heat at different temperature levels, some very high such as combustion, others rather low such as biomass drying or methanol distillation.

The actual integration of these plants into existing industrial areas, such as a pulp mill or the chemical cluster, is primarily based on the material integration resulting from the flows of feedstock and products. However, due to the thermal features of the processes, the consumption or the availability of heat (e.g. steam) is also an important factor of integration that can affect the overall consumption of primary energy at the specific industrial site.

The optimal material and thermal integration typically follow separate principles. In this work, the size of new plants (biomass based process) was specified based upon the amount of intermediate product (e.g. methanol, ethanol, olefins, etc.) that optimally match the capability of existing plants to process such product / feedstock based on preliminary economic considerations. It is important to note that thermal integration (i.e. when the size of a new plant is specified based upon matching heat availability and heat requirements of the constituent plants) was not used as a sizing criterion with the consequence that the potential benefits of larger or smaller sizes of the new plant on the overall primary energy consumption are not fully exploited.

To estimate the thermal integration between the new plants analysed here and the existing industrial plants, a thermodynamic targeting approach was adopted using Pinch Analysis tools (Kemp 2007). The flowsheets of the biomass based processes were established by considering each single heating or cooling step as heaters and coolers, which are referred to as "thermal streams". Cold and hot thermal streams (respective those needing heating and cooling) are then aggregated into a “thermal cascade” by summing algebraically the heat contribution of each thermal stream at each relevant temperature interval provided that a temperature difference (here 20°C) exists between hot and cold thermal streams. In this way, the minimum amount of heat that must be provided by external hot utility (e.g. combustion, reaction heat, steam etc.) or the maximum amount of heat that can be exported to other plants can be estimated. The process thermal cascade can also be represented in a diagram with heat load and temperature axes, called “grand composite curve”, which provides an immediate reading of the opportunities for heat integration with other process thermal cascades (such as e.g. a steam cycle for power generation).

The great advantage of this energy targeting approach is that the detailed design of the heat exchanger network (HEN) is tackled in a subsequent design steps and therefore the thermodynamic characteristics of a process (here the conversion of biomass into methanol) are affected only by the intrinsic characteristics of the thermochemical conversions (i.e. by the technological features of the equipment units) and not by the design of heat exchangers and their arrangement.

The level of aggregation of thermal streams into one or more subsystems can however affect the estimation of the overall primary energy consumption. In principle, the larger the set of thermal streams, the larger the number of integration opportunities. Thus different choices are possible. One option is to combine all system hot and cold thermal streams into a single set to be integrated, e.g. all the petrochemical cluster thermal streams are included together with the thermal streams of the biomass conversion process into a single set and the overall heat integration potential is estimated. The result, although interesting from a theoretic point of view, is not indicative of the actual heat savings possibilities, which are constrained to much lower levels due to the existing heat exchangers that cannot be discarded unless a complete retrofit of the industrial area is pursued.

An alternative option is to investigate opportunities for exploiting the intrinsic heat integration possibilities with a utility network, typically steam at different pressure levels. Steam networks are common in industry, especially in energy intensive processes, and are a way to collect and distribute heat to the process in an efficient and relatively inexpensive way compared to the option of matching all the process streams through direct heat exchange. The net heat demand of a system translates in this way into a steam demand which can be supplied by a centralized boiler or CHP plant. This approach is often called "total site" as the system heat integration is ultimately achieved through a utility network (steam) interacting with separate locally integrated subsystems.

This latter is the approach used in this study. In particular we assumed that the new biomass based plant constitutes a subsystem (single set of thermal streams) where ideal heat integration is achieved (i.e. a single grand composite curve is generated). The potential heat integration between the biomass based plant(s) and the existing nearby industry is then estimated considering exchange of steam, e.g. generated by recovering the heat from gas cooling after biomass gasification and exported to the existing steam network, if possible at compatible steam pressure levels.

Two main sites offer interesting opportunities for steam exchange between biomass based processes and existing industries: the Stenungsund chemical cluster and the Iggesund pulp and paper plant. For these two sites, the steam demand at different pressure and temperature levels were collected with the help of other research partners involved in the Skogskemi project and colleagues at the Div. of Industrial Energy Systems and Technologies.

For a detailed overview of the processes and various utility demands of the Stenungsund chemical cluster the reader is referred to Ref. (Hackl, Anderson et al. 2010).

Data regarding the steam system of the Iggesund pulp and paper plant were obtained through internal communication with plant operators.

2.5 Evaluation of thermodynamic performance

The system configurations for biomass based methanol production were investigated with respect to their thermodynamic performance.

Energy and mass balances were evaluated by means of process modelling and simulation using Aspen Plus. Thereafter the biomass to methanol conversion efficiency was calculated, defined as the ratio between the chemical energy rates of the methanol output and the biomass input according to Equation (2.1).

$$\eta^{HHV} = \frac{\dot{m}_{\mathrm{MeOH}} \cdot HHV_{\mathrm{MeOH}}}{\dot{m}_{biomass} \cdot HHV_{biomass}} \qquad (2.1)$$

Note that higher heating values (“HHV”) are used in this report in order to avoid the bias introduced by the different water content of the various feedstocks considered. For completeness, corresponding values based on lower heating value (“LHV”) are also reported.

Note that similar conversion efficiency values can be calculated at different key process points as long as mass flowrate and heating value of a specific material stream other than methanol is used. In the case of a gasifier this is often referred to as the “cold gas efficiency” which is commonly used for comparing different gasifier concepts or plants.

By means of Pinch Analysis tools, the net excess heat available after ideal heat integration of the biomass based processes was estimated. Thereafter, the opportunities for excess heat utilization were estimated, including steam export to nearby industrial sites or power generation based on a heat recovery steam cycle (or combinations of these two options). These results were in turn used to calculate the fuel consequences in marginal energy conversion technologies used for heating or for power generation.

The net heat available from the biomass based processes is always positive due to the large amount of heat available from gas cooling after gasification and from exothermic methanol synthesis. Even though steam is required for the various reforming stages in the biomass based processes, large amounts of steam can nevertheless be exported to the nearby industrial sites, thus reducing the need for steam generation in existing boilers. As a consequence, less fuel (usually of the fossil type) is used. Fuel savings can be quantified by dividing the estimated amount of heat savings $\dot{Q}_{saved}$ by the average thermal efficiency of existing boilers at the site η_q.

As opposed to heat, electrical power is generally needed for driving the various gas compression operations in the biomass based processes and for oxygen production through air separation unit (ASU). Accordingly, the balance can be positive or negative depending on the amount of power that can be generated by means of a heat recovery steam cycle. Negative or positive power flows across the boundaries of the biomass based processes imply equivalent power generation or reduction in marginal electricity generation plants. Depending on the characteristics of these marginal plants, power can be generated at different efficiencies. There are significant differences between electric power generation technologies in different regions. Since a new biomass based methanol production plant will operate mostly at nominal capacity all the year around, it is reasonable to consider a build-margin approach for estimating the consequences of negative or positive power flows (Axelsson, Harvey et al. 2009). Reasonable assumptions for build margin technologies in European power grids are of the fossil fuel type. Fuel consumption in marginal electricity producers can be quantified by dividing the estimated amount of imported electricity $\dot{W}_{import}$ by the electricity production efficiency η_{el}.

Accordingly, a more complete figure of performance for comparing the different system configurations can be estimated according in Equation (2.2) which is denoted here as the “total biomass to methanol conversion efficiency”.

$$\eta_{total}^{HHV} = \frac{\dot{m}_{\mathrm{MeOH}} \cdot HHV_{\mathrm{MeOH}}}{\dot{m}_{biomass} \cdot HHV_{biomass} - \frac{\dot{Q}_{saved}}{\eta_q} + \frac{\dot{W}_{import}}{\eta_{el}}} \tag{2.2}$$

This efficiency is similar to the conversion efficiency from biomass to methanol defined in Equation (2.1), with the only difference the total input energy rate is accounted for, including the effect on marginal energy conversion technologies. This is done by deducing from the total energy rate of biomass input the fuel saved in boilers at the nearby industrial site due to the steam export from the biomass based by adding the fuel spent in marginal electricity production technologies to compensate for the electricity imported by the biomass based processes (or deducing this fuel in case of export of electricity). The total conversion efficiency therefore depends on the efficiency of existing boilers and of the values assumed for marginal electricity producers.

While it is relative easy to estimate the fuel savings in existing boilers (e.g. 80% boiler efficiency), marginal electricity may be produced with different technologies. A possible assumption is to consider that the electricity must be provided by a new plant (build-margin) using the technology and fuel that achieves the lowest levelised cost of electricity generation. In a harmonized European energy market, the corresponding marginal electricity producers are modern coal power plants or natural gas combined cycles. Another possible assumption is that the electricity is produced using the same fuel used for heating purposes today which is saved when steam is delivered from the biomass based processes. This fuel is natural gas which can therefore be used in a combined cycle power plant. A state-of-the-art combined cycle may reach an electrical efficiency around 60% on lower heating value basis.

If the biomass to methanol conversion efficiency is an intrinsic figure of performance of each specific process concept, the total conversion efficiency takes into account the integration effects with existing industrial sites and is therefore site dependent.

3 Results: Thermodynamic performance

3.1 System configuration nr 1

3.1.1 Description

The first system configuration for methanol production consists in converting biomass into syngas through indirect (dual bed) gasification in two plants: one located in proximity of Värö pulp mill on the Swedish West Coast and the other located within the Stenungsund cluster site. A representation of this system is shown in *Figure 3.1*.

At the gasification plant at Värö, 200 MW (LHV basis) of biomass is converted into substitute natural gas (SNG). The gas is compressed and delivered to the gas network.

Two syngas production lines are located in Stenungsund. One consists of an autothermal reforming for the production of syngas from an equivalent quantity of SNG produced at Värö and delivered through the natural gas network. The second line consists of an indirect biomass gasifier producing syngas of the same quality. The syngas streams produced by the two lines are merged and fed to the methanol synthesis plant. The methanol is then fed to the MTO plant. The biomass derived methanol is mixed with imported methanol to achieve a total olefin production of 200 kt·y^{-1} olefins. An optional quantity of ethanol can also be mixed with methanol to feed the MTO process, in which case a different ratio of C2 and C3 olefins is obtained.

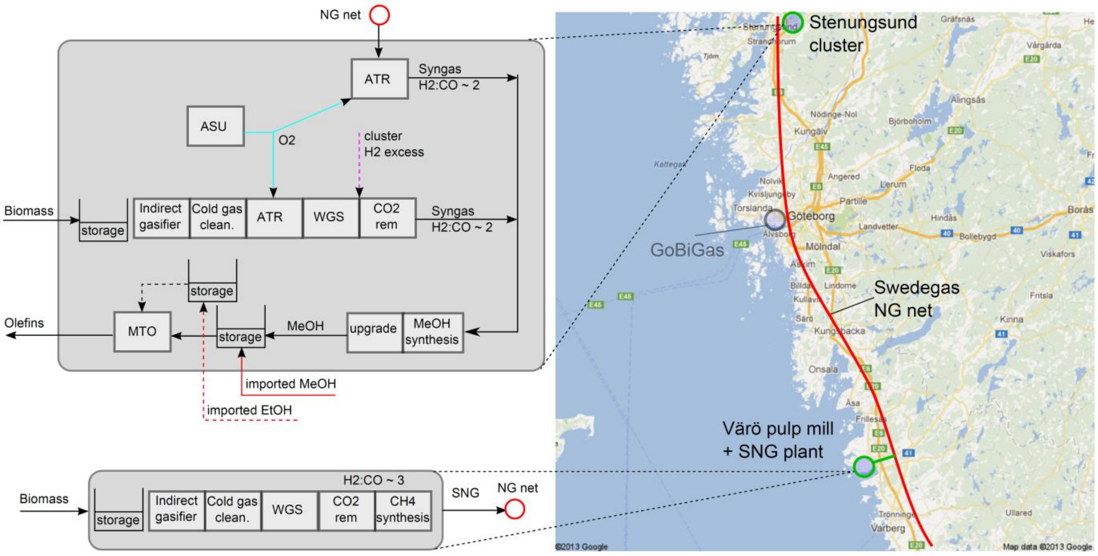

Figure 3.1: System configuration nr 1: biomass gasification and methanol synthesis based on indirect (dual bed) gasifier technology

By producing SNG at Värö and transporting it through the natural gas network to Stenungsund where it is reformed into syngas for the subsequent methanol synthesis, the natural gas network can be used as a backup or buffer in case of discontinuous operation of one or more plant units. The natural gas net can in fact be used to supply an additional quantity of natural gas to the Stenungsund cluster compared to that produced at Värö. This includes not only the possibility of importing extra fossil natural gas to substitute the SNG or to increase the throughput, but also of buying extra SNG in the future from other biomass-SNG plants located in Sweden (e.g. the GoBiGas phase 2 project planned to be built in Göteborg).

Since this first biomass conversion alternative is based on two large plants located into two different areas, namely close to Värö pulp mill and in the Stenungsund cluster, the processes and the heat integration opportunities of the two sites are discussed in separate sections.

3.1.2 SNG production at Värö pulp mill

Värö is a Kraft pulp mill with an average yearly production of 425 000 air-dried tonnes of pulp (Södra). The feedstock consists of wood chips from spruce (70%) and pine (30%). The Kraft process of conversion of wood into pulp is based on several chemical and thermal treatments of the feedstock and of the intermediate materials and by-products. Large quantities of steam at medium and low pressures are required and steam is mainly produced by the recovery boiler which, in addition to recovering the spent chemicals in the black-liquor, converts the lignin of the black liquor into heat through combustion.

3.1.2.1 CHP system at Värö pulp mill

In a recent master thesis work at the Division of Heat and Power Technology, a detailed energy analysis of the Värö plant was conducted (Bood and Nilsson 2013). The data shown in Table 3.1, which were mainly obtained from this work, were used to determine the integration opportunities between the existing pulp mill and a new biomass gasification and SNG production process.

The MP and LP steam headers are supplied by connection to the extraction and exhaust ports of a back-pressure steam turbine driven with HP pressure steam at 86 bar, 482°C which is primarily produced by the recovery boiler.

The steam network was modelled in order to obtain by a Pinch Analysis targeting procedure the real output of the existing plant (see Table 3.1). It was assumed that the recovery boiler runs at nominal capacity (395 MW), as the recovery boiler is commonly a bottleneck for increasing the mill production, that its efficiency is 74 % on LHV basis, and the turbine isentropic efficiency is 0.68. The combustion heat from the recovery boiler was considered divided into equal part, a radiative part at 1000°C and a convective part between 1000°C and 150°C (stack temperature). In addition, a 10 K temperature approach between gases and steam was assumed.

The resulting thermal profile of the CHP system at Värö is shown in *Figure 3.2.*

Table 3.1: Key data of Värö pulp mill, adapted from (Bood and Nilsson 2013)

Description	***Value***
Pulp production	1227 ADt per day
Raw material intake (wood chips)[1]	2824 ADt per day
Recovery boiler nominal capacity	395 MW
Recovery boiler average HP steam production	279 MW
LP (4.5 bar) steam consumption	161 MW
MP (11 bar) steam consumption	50 MW
Electricity production	51 MW
District heat delivery to Varberg	30 MW

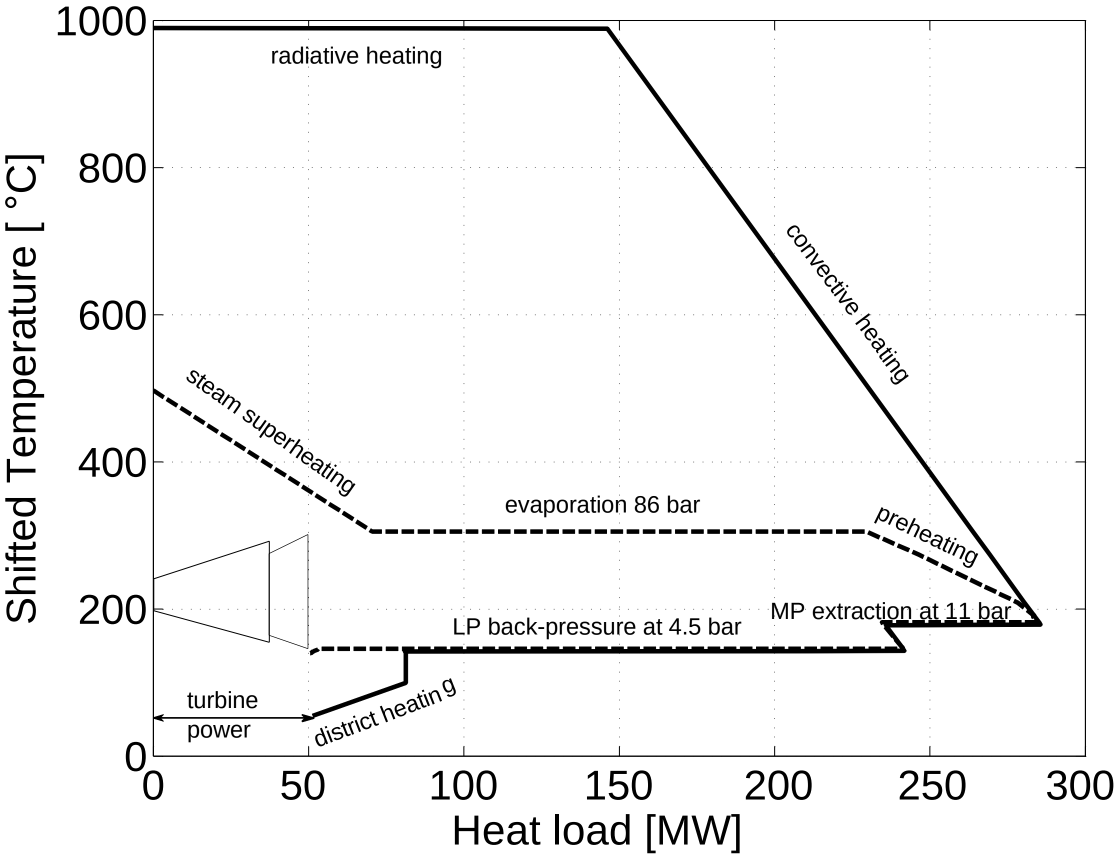

Figure 3.2: Grand composite curves of the CHP system at Värö pulp mill

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[1] According to the source this has been calculated considering a 43 % yield of a typical Kraft mill (FRAM model) although the actual yield at Värö is around 47 %.

3.1.2.2 Stand-alone SNG production via biomass gasification at Värö

For this configuration, a biomass intake of 200 MW_{LHV} wood chips at 50% moisture was considered, which corresponds to an average of 84 158 $kg\cdot h^{-1}$, that is around 673 $kt\cdot y^{-1}$ assuming 8000 hours operation.

Mass and energy balances of the biomass conversion to SNG process were calculated according to a published model developed at the Div. of Heat and Power Technology, Chalmers (Arvidsson, Heyne et al. 2012), for a process that is very similar to the process layout of the existing GoBiGas Bio-SNG plant in Gothenburg. *Figure **3.3*** depicts the process flowsheet and provides main process operating parameters.

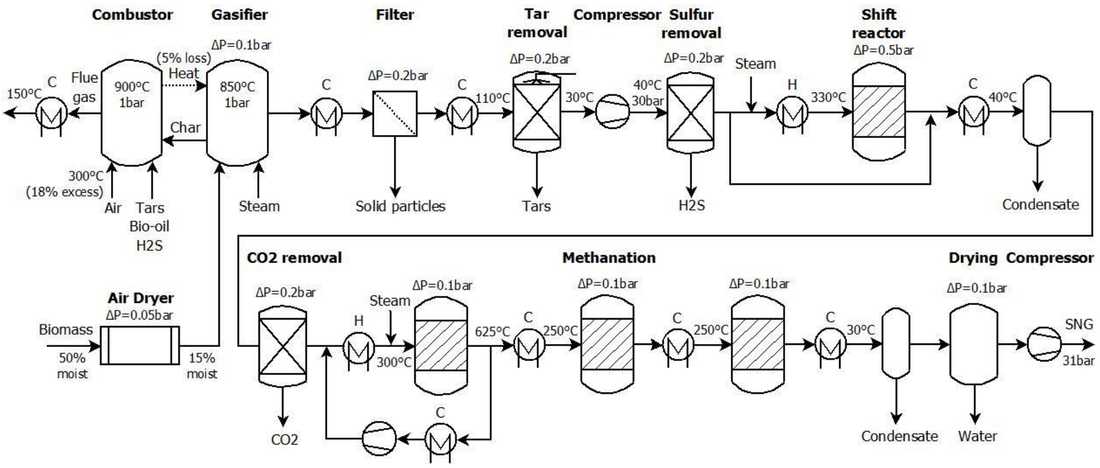

Figure 3.3: Layout of the biomass gasification and SNG production process, adapted from (Arvidsson, Heyne et al. 2012).

Biomass at 50% moisture is first dried and is then fed to the gasifier at 15% moisture where it undergoes gasification in presence of steam in the hot circulating fluidized bed. The bed is heated by means of char combustion in a separate combustor, thus avoiding mixing the nitrogen in the air with the product gas.

The biomass is converted into 51 243 $kg\cdot h^{-1}$ of gas with the composition reported in *Table 3.2*.

The product gas is first cooled and cleaned in order to remove particles and alkali metals. An oil scrubber is used to remove tar which is then recirculated to the gasifier combustor where it is burnt together with the biomass char. At this stage the main pressurization is obtained (up to around 30 bar). A selective H_2S and CO_2 removal is used which consists in two MEA stages where the H_2S is separated in the first and CO_2 is mainly separated in the second one. As the final product is a gas which should be as rich as possible in methane, no further reforming is required. Conversely, water-gas shift is necessary to adjust the H_2:CO ratio to around 3 which is the optimal methanation stoichiometry. Since the H_2:CO ratio in the product gas is already around 1.8, only a portion of the gas undergoes shift. Water-gas-shift is done in between the two MEA stages as the shift produces some CO_2 which has to be removed prior to methanation. A three step fixed bed methanation is then used to convert the syngas into methane, the reaction being substantially at full conversion. Water is finally separated and the gas is further compressed to the grid delivery pressure (around 31 bar).

Table 3.2: Composition of the product gas obtained from gasification of wood chips in a fast circulating indirect fluidized bed gasifier at 850°C and 1 atm.

Component	***Mol. Frac. (%)***
H_2	34.0
CO	18.5
CO_2	13.2
H_2O	26.3
CH_4	7.6
Toluene	0.0438
Naphthalene	0.0657
Phenol	0.2265
Ammonia	0.1265
H_2S	0.0041
HCl	0.0062

Based on the aforementioned model of the biomass based SNG process (more details can be found in the Appendix), hereafter denoted "bioSNG", the biomass is converted into SNG with the characteristics shown in *Table 3.3*.

Table 3.3: Characteristics of SNG produced from gasified biomass

Parameter	***Value***
Pressure	31 bar
Temperature	30°C
Composition (mol frac %)	
CH_4	95.46
H_2	3.61
CO_2	0.90
H_2O	0.004
others	0.026
Wobbe index	49.41 $MJ \cdot m^{-3}$

The size of the bioSNG plant corresponds to 200 MW_{LHV} biomass input (243.8 MW_{HHV}) which yields a production of 10 944 $\text{kg}\cdot\text{h}^{-1}$ of SNG[2] (166.2 MW_{HHV}, 149.6 MW_{LHV}), with a conversion efficiency of 68.2% on a HHV basis (74.8% LHV basis).

Assuming that ideal heat integration is achieved between process heat sinks and sources, the bioSNG plant is thermally self-sufficient. This means that no extra heating should be used for SNG production and even a certain amount of excess heat is available. Still, the work demand for compressing the syngas after gasification to the methanation pressure and the SNG up to the desired delivery pressure is around 1.77 kWh per kg of SNG which corresponds to around 19.4 MW of compression power for the 200 MW plant size.

An overview of the grand composite curve of the bioSNG process is shown in *Figure 3.4* together with the thermal profiles of an integrated steam cycle.

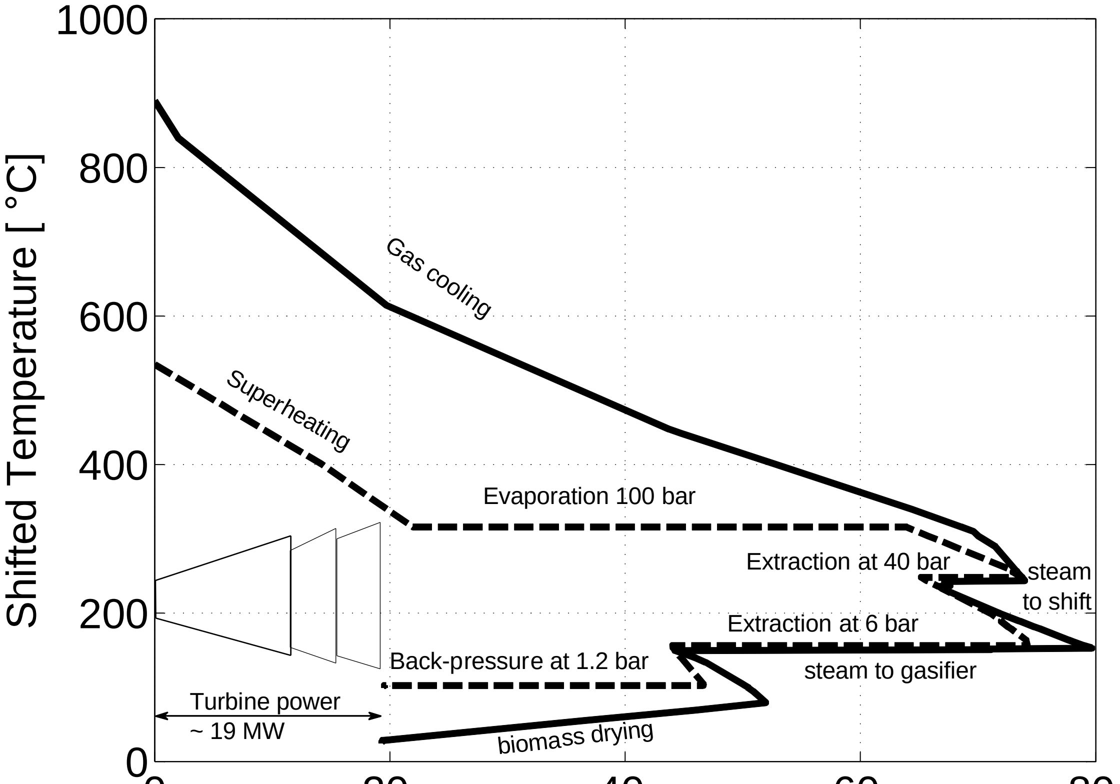

Figure 3.4: Grand Composite Curves of the stand-alone bioSNG process and of the integrated steam power cycle.

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[2] SNG lower heating value 49.201 MJ/kg, SNG higher heating value 54.673 MJ/kg

As the bioSNG is considered here as a stand-alone plant, the steam cycle parameters can be chosen independently of the existing steam network at the mill. Accordingly, a maximum steam pressure of 100 bar and maximum superheat temperature of 530°C were assumed. Two steam extractions were considered: at 40 bar (compatible with the shift reactor) and at 6 bar (suitable for the gasifier). In addition, the steam turbine back-pressure was assumed to be slightly over the atmospheric pressure to achieve a steam temperature compatible with the biomass dryer. The turbine isentropic efficiency was assumed at 0.78.

By maximizing the steam production to exploit all the available excess heat from the process, it is possible to produce around 19.3 MW of power[3].

By including the steam power cycle, power balance is achieved for the stand-alone bioSNG plant (19.3 MW produced by the steam cycle, 19.4 MW required for compression work) which can be therefore considered energy self-sufficient. Thus the system energy efficiency is equal to the biomass conversion efficiency.

3.1.2.3 Energy integration opportunities between the pulp mill and a biomass based SNG production plant at Värö

In the previous section it was highlighted that a stand-alone bioSNG plant of 200 MW biomass input (LHV basis) could be energy self-sufficient provided that a steam cycle is used to recover the process excess heat to produce power for the gas compression to the gas grid pressure.

As the bioSNG plant is assumed to be built in proximity of the pulp mill and as the pulp mill is already equipped with a large CHP plant mainly based on the recovery boiler, it is relevant to investigate if there is any benefit in integrating the mill steam network with the bioSNG steam network. Considering the extra complexity introduced by an integrated system which requires coordinating the simultaneous operation of the pulp mill recovery boiler and the bioSNG plant, a larger power production than the current mill power production (51 MW) is desirable.

To identify the possible benefits of building such an integrated CHP system, the grand composite curves of the recovery boiler and of the bioSNG plant were combined in a single grand composite curve and the integration of steam cycle with optimal extraction levels was investigated.

In particular, it was assumed that two steam production lines are implemented, one equivalent to the recovery boiler steam line (86 bar, 492°C) and one at 100 bar and 530°C, and that the amounts of steam extracted at 11 bar and 4.5 bar are sufficient to cover the mill requirements. A steam line at 40 bar was also included as it is needed for the shift reactor in the bioSNG plant. An optional 6 bar steam line was also investigated but it was found that this can be avoided while still maximizing the steam turbine power and the steam to the gasifier can be delivered instead with 11 bar steam line.

The resulting optimal CHP system is shown in *Figure 3.5.*

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[3] Note that this corresponds to the theoretical excess heat of the bioSNG plant under the condition of optimal heat recovery and conversion of biomass to SNG is maximized. Extra power can obviously be produced if part of the biomass or of the gas is burnt thus decreasing the biomass to SNG conversion.

As no extra steam than that provided through the CHP system based entirely on the recovery boiler is currently needed at the Värö plant, there is no incentive to investigate the option of steam export from the bioSNG plant for heating purposes only. The totality of the excess heat from the biomass based SNG production process should be used for power generation, thus minimizing the need for electricity import for gas compression.

It can be noticed that once again the power of the steam cycle can be maximized up to the theoretical value corresponding to the system excess heat, which amounts to around 70 MW. The resulting power production, corresponding to the net power requirements for gas compression in the bioSNG plant, is around 51 MW which is equivalent to the assumed power production by the recovery boiler for current conditions.

This result shows that an integrated CHP system is actually not of interest, since the extra operational complexity is not counterbalanced by any increase in power compared to the case in which no heat integration is achieved between the mill and the SNG plant.

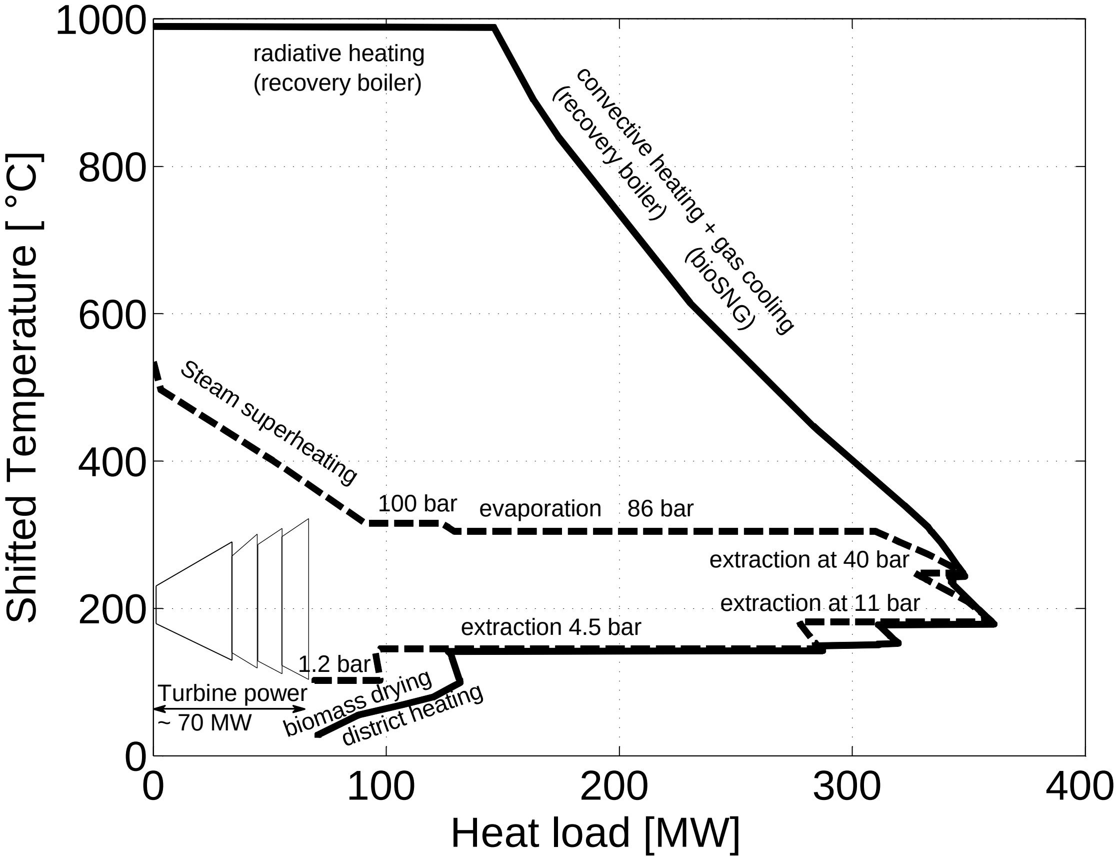

Figure 3.5: Grand composite curves of the mill recovery boiler and bioSNG plant and integrated steam cycle.

3.1.3 Syngas production through NG autothermal reforming in Stenungsund

A conventional methanol synthesis technology was assumed in this work. Typical values of the H_2, CO and CO_2 contents in the syngas prior to methanol synthesis were assumed for the syngas feed (Supp 1990) to achieve the synthesis ratio (SR) defined in equation (3.1) with 3 %mol of CO_2.

$$SR : \frac{H_2 - CO_2}{CO + CO_2} = 2.05 \qquad (3.1)$$

The production of high quality high purity syngas from natural gas is achieved by means of a state-of-the-art catalytic reforming. In principle, different types of reformer could be used, such as catalytic or non-catalytic partial oxidation and steam reforming. However, non-catalytic reforming is more suitable for heavy fuel reforming in which case large amounts of oxygen are required due to the high temperatures for carrying out reforming without catalyst. Steam reforming is instead usually used for hydrogen production, in which larger amounts of water are injected to promote the gas shift.

The Haldor Topsøe reforming technology was considered in this study which has been developed for gas to liquid applications (Aasberg-Petersen, Christensen et al. 2003). A process overview is given in *Figure 3.6.* The process consists of a combination of an adiabatic pre-reformer at relatively low temperature to promote the conversion of C2+ paraffins and a second autothermal reformer where methane is finally converted into H_2 and CO in presence of oxygen and steam.

After the reforming, the hot gas is cooled down prior to the final cleaning steps. The water is removed in a flash drum. A Rectisol process is used for CO_2 separation where methanol, which is produced in the subsequent synthesis process, is used as a physical solvent. As the solubility of CO_2 in methanol increases with decreasing temperatures, chilled methanol is normally employed which requires refrigeration.

Note that the steam to carbon ratio in the main reforming stage is around 0.6. Under such condition, the synthesis ratio SR is around 2 which would suggest to slightly increase the steam to carbon ratio to obtain the optimal 2.05 value for methanol synthesis. Some H_2 can be actually recovered from the purge after the methanol synthesis reactor and a lower value of the synthesis ratio can be therefore used since the syngas can be enriched with the H_2 recovery. The optimal synthesis ratio of the syngas before the H_2 recycle was estimated to be around 1.9, which is lower than the value obtained if only steam is used for reforming. To increase the CO content in the final syngas, some CO_2 obtained from separation is therefore recirculated into the reformer to obtain the desired value of the synthesis ratio.

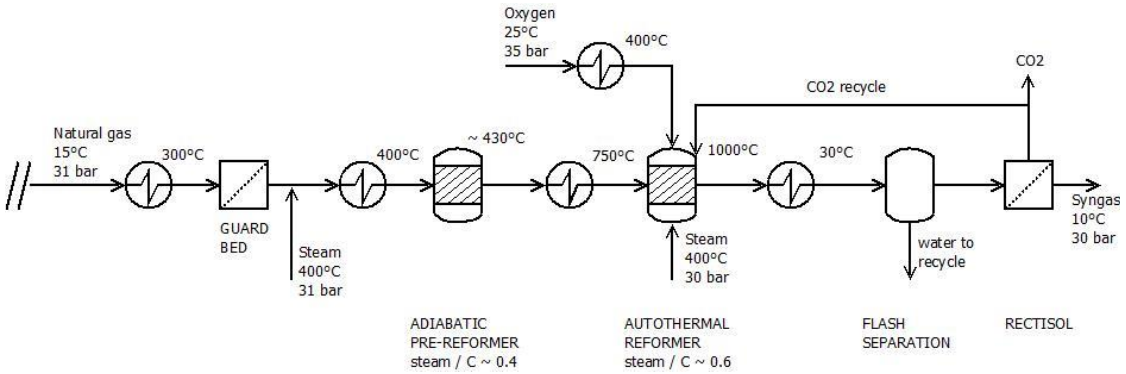

Figure 3.6: Layout of natural gas reforming process

The composition of the SNG is slightly different than the average composition of the natural gas available in the Swedish gas grid, which is shown in *Table 3.4*, although a similar value of the Wobbe index is obtained.

Table 3.4: Typical composition of the natural gas in the Swedish gas grid.

Compound	***Mol frac (%)***
Methane	88.83
Ethane	6.13
Propane	2.48
Butane	0.93
Pentane	0.21
Hexane	0.06
CO_2	1.04
N_2	0.32

The composition of the natural gas at the Stenungsund reforming plant is considered to be that of typical natural gas in the Swedish gas grid. The natural gas amount used for syngas production at the Stenungsund plant is therefore calculated assuming that the quantity of natural gas reformed at the Stenungsund site has the same chemical energy rate[4] as the SNG produced at Värö discussed in the previous section. The estimated nominal SNG production capacity at Värö is 10 944 kg h^{-1} which corresponds to a chemical energy rate of around 166.2 MW (HHV basis). This is equivalent to a mass flow rate of 11 269 kg h^{-1} of natural gas.

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[4] The chemical energy rate is calculated by multiplying the flow rate by the higher heating value. The natural gas lower heating value is 47.994 MJ/kg, and the higher heating value is 53.096 MJ/kg according to the composition reported in *Table 3.4*.

By reforming the above quantity of natural gas, a production of around 22 266 kg·h⁻¹ of syngas with the composition in *Table 3.5* was estimated (156.0 MW HHV basis, 139.4 MW LHV basis)[5]. This corresponds to a natural gas to syngas conversion efficiency of around 93.9% and of around 64.0% from biomass to syngas HHV basis (respectively 93.2% and 69.7% in LHV basis).

Table 3.5: Estimated composition of syngas at the output of the natural gas reforming.

Compound	***Mol frac (%)***
H_2	64.5
CO	29.0
CO_2	3.1
CH_4	2.4
N_2	1.0
H_2O	155 PPM

As shown in *Figure 3.6*, the reforming process was modelled considering that all the heating and cooling processes are operated separately (i.e. not matched in a heat exchanger network) and the overall thermal energy balance was studied after energy and mass balance using Pinch Analysis tools. The process grand composite curve was studied to investigate if the process is thermally self-sufficient.

As shown in *Figure 3.7*, the reforming process is thermally self-sufficient and a certain amount of excess heat is also available which highlights the opportunities either to produce power with an integrated steam cycle provided that steam required by reforming is also extracted at around 30 bar, or that steam is exported to the Stenungsund site. As the final utility system is studied considering the combination of all the processes for the production of methanol such as an additional gasifier and the methanol synthesis and separation steps, the site steam balance is discussed later on after introducing the remaining processes.

Note that the gas heating prior to the two reforming stages and part of the steam production are commonly achieved by means of a furnace (Aasberg-Petersen, Christensen et al. 2003). If so, the conversion of natural gas into syngas is partially lowered by the quantity of gas used in the furnace. The grand composite curve in *Figure 3.7* shows instead that gas preheating can be done in a more compact way by syngas cooling at the reformer outlet.

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[5] Syngas lower heating value 22.539 MJ/kg, higher heating value 25.227 MJ/kg according to the composition reported in *Table 3.5*.

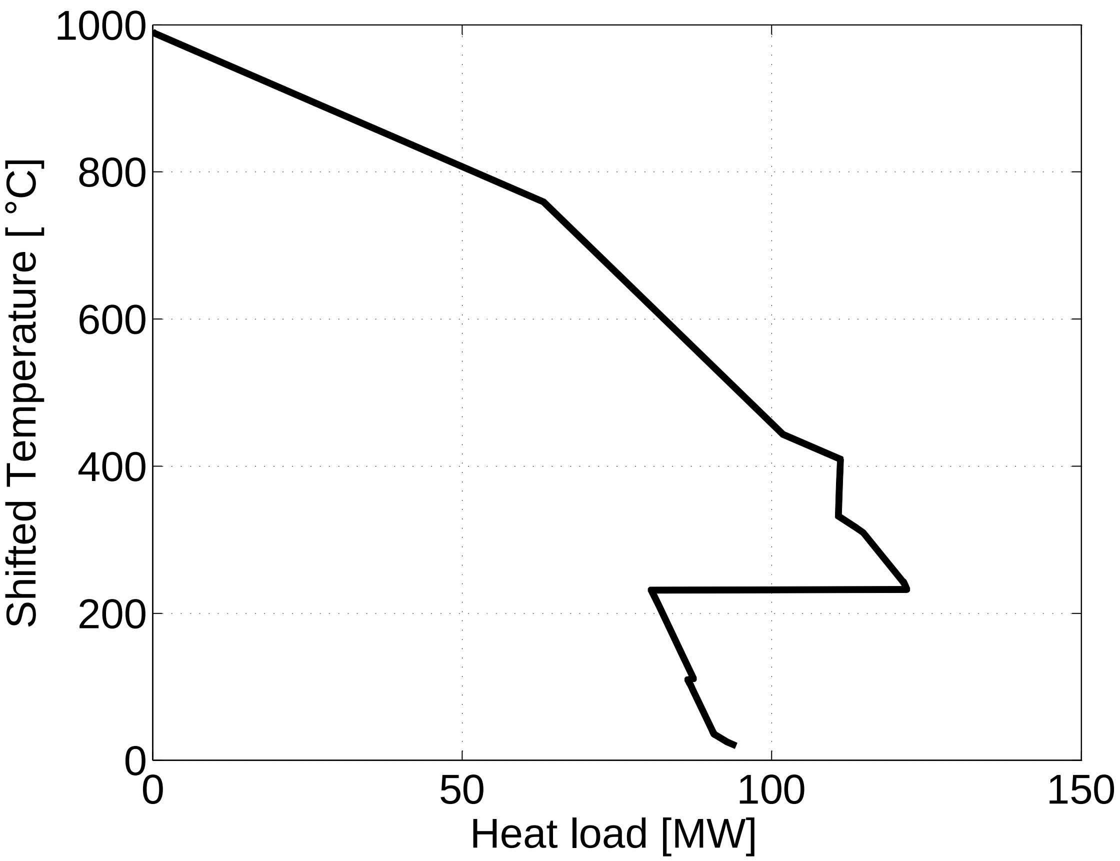

Figure 3.7: Grand composite curve of the natural gas reforming plant in Stenungsund.

3.1.4 Syngas production through biomass gasification in Stenungsund

Additional syngas production in Stenungsund through biomass gasification was investigated for direct use at the Stenungsund site for methanol synthesis. The process is referred hereafter to as "bio-syngas" plant and the layout is shown in *Figure 3.8*.

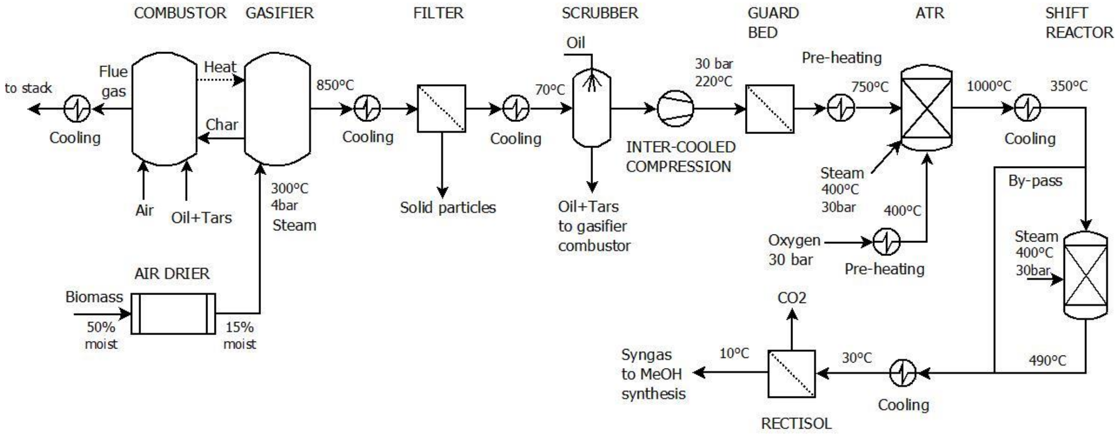

Figure 3.8: Layout of syngas production process based on indirect biomass gasifier

The same type of indirect gasifier considered for the Värö plant was assumed. Biomass at 50% moisture is firstly dried and is fed to the gasifier at 15% where it undergoes gasification in presence of steam in the hot circulating bed which is heated by means of char combustion in the combustor. The product gas is first cooled and cleaned to remove particles and alkali metals. An oil scrubber is used to remove tar which is then recirculated to the gasifier combustor where it is burnt together with the biomass char. The gas is subsequently compressed to 30 bar which is the same delivery pressure of the syngas from the natural gas reforming plant described in the previous section. As the product gas contains around 7.5 % of methane, a reforming step to convert as much as possible the methane into H_2 and CO is required. In addition, to adjust the value of the synthesis ratio to around 1.9 (the value of 2.05 is obtained after H_2 recovery from the methanol synthesis purge gas as explained later in this section), water gas shift and CO_2 separation are required. Only a portion of the syngas is sent to the water gas shift and the other part of the gas is by-passed to achieve the right syngas composition after mixing. The syngas is finally cooled prior the Rectisol process where CO_2 is removed by absorption in chilled methanol. The process model used for mass and energy balance is presented in more detail in the Appendix.

The plant was sized according to a biomass input of 150 MW LHV basis at 50% moisture (183 MW HHV basis). A production of 17 136 $kg \cdot h^{-1}$ of syngas with the composition shown in *Table* 3.6 is achieved (120.6 MW HHV basis, 107.8 MW LHV basis) corresponding to a biomass to syngas conversion efficiency of around 66% HHV basis (71.9% LHV basis)[6]. Note that this corresponds to around 4 efficiency points more than the conversion efficiency of the bio-SNG plant and subsequent natural gas reforming (62.2% HHV basis, 67.6% LHV basis). In addition, the oxygen demand of the natural gas reformer is much larger than that required by the reformer in the bio-syngas plant, thus the overall system efficiency of the bio-SNG option is lower than the biomass to syngas conversion efficiency due to the electricity requirements for ASU and oxygen compression.

Table 3.6: Estimated composition of syngas produced via biomass gasification.

Compound	***Mol frac (%)***
H_2	65.6
CO	29.6
CO_2	3.1
CH_4	1.5
N_2	0.2
H_2O	406 PPM

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[6] Syngas lower heating value 22.634 MJ kg^{-1}, higher heating value 25.335 MJ kg^{-1} according to composition in Table 3.6

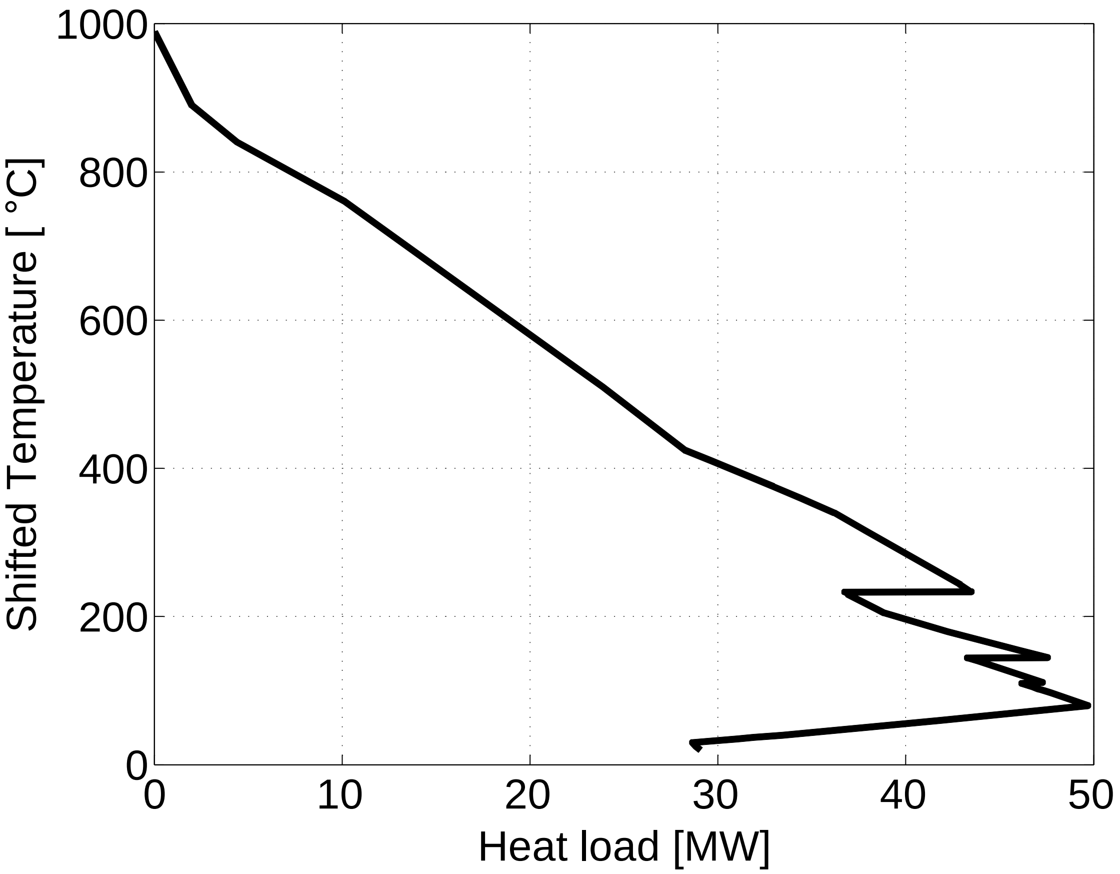

Figure 3.9: Grand Composite Curve of the bio-syngas process.

As done for the natural gas reformer process, a study was conducted to estimate the theoretical heat integration of the bio-syngas process. The grand composite curve of the process is shown in *Figure 3.9.*

The bio-syngas process is thermally self-sufficient and the theoretical amount of excess heat is around 28 MW. However, due to gas compression and Rectisol refrigeration units, the bio-syngas process consumes around 8.7 MW of electric power which must be either purchased from the grid or produced by means of an integrated steam cycle. As an optimal steam network configuration at the Stenungsund site can be integrated with the natural gas reforming plant and with the methanol synthesis plant, the design of such steam network is discussed later on in this Section after the methanol synthesis process is introduced.

3.1.5 Methanol synthesis in Stenungsund

The two syngas streams produced by means of biomass gasification and natural gas reforming in Stenungsund are subsequently merged and fed to the methanol synthesis process. A layout of the process is shown in *Figure 3.10.*

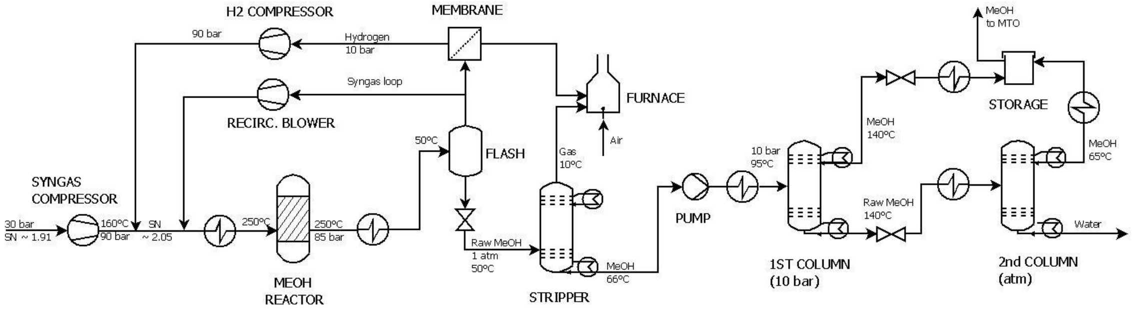

Figure 3.10: Layout of the methanol synthesis and upgrading process

The fresh syngas is compressed to the synthesis pressure of 90 bar after which it is mixed with a H_2 stream recovered from a synthesis purge stream to reach the final optimal gas quality requirement for methanol synthesis (synthesis ratio around 2.05 and 3% molar fraction of CO_2). The syngas is then mixed with the synthesis loop gas which is used to partially increase the overall conversion by increasing the number of passages of syngas through the synthesis and partially to increase the inert portion of the gas which prevents a dramatic increase of temperatures that would otherwise incur if only fresh syngas is used due to the highly exothermic reactions. The size of this loop is a crucial design factor which can be adjusted according to the further utilization of the purge gas and to the quality of the fresh syngas. Here the synthesis loop is assumed to be 3:1 in volume with respect to the fresh syngas, which is a typical value for the H_2 rich syngas considered here. The gas is further heated to the synthesis temperature of around 250°C and then fed to the methanol reactor where it is converted to methanol according to the following reactions:

Methanol synthesis from CO: $CO + 2H_2 \leftrightarrow CH_3OH$

Methanol synthesis from CO_2: $CO_2 + 3H_2 \leftrightarrow CH_3OH + H_2O$

Water gas shift: $CO + H_2O \leftrightarrow CO_2 + H_2$

The CO and CO_2 conversions to methanol are balanced by the shift reaction and the system is close to equilibrium. The methanol contained in the synthesis product is then separated by cooling and flashing. Syngas is recovered at the top of the flash and is partially recycled as synthesis loop and partially purged. This purge is still rich in H_2 and CO but a further increase of the loop would cause the flow through the reactor to become too large due to inert gases such as methane and nitrogen. H_2 recovery from the purge is done by means of membranes where the separation is favoured by the high pressures of the feed. This highly pure H_2 stream is therefore recycled into the reactor which allows use of a syngas feed slightly richer in CO therefore reducing the portion of the feedstock to be converted into CO_2 in the upstream processes or, in order words, to increase the overall conversion of the feedstock into methanol.

The raw methanol obtained at the bottom of the flash is then sent to a stripper after the pressure has been released to the atmospheric level. In the stripper the remaining light gases are removed at the top thus leaving a mixture of water and methanol at the bottom. The final separation of the methanol can be arranged in different ways. In this work a two pressure level separation is assumed where the first column operates under pressurized conditions thus allowing an optimal thermal integration with a second atmospheric column. The distillates of the two columns are mixed and the final methanol stream can be stored and subsequently used for olefin productions, while water is removed at the bottom of the second column.

Detailed mass and energy balances are shown in the Appendix.

A total production of 34 370 kg·h⁻¹ of grade AA methanol[7] is obtained (216.2 MW HHV basis, 189.8 MW LHV basis) which corresponds to around 78.1% conversion efficiency from syngas[8] to methanol on a HHV basis (76.7% LHV basis). Overall, a 50.8% conversion efficiency from biomass to methanol (HHV basis) is achieved (54.2% LHV basis). Note that the methanol flowrate at the bottom of the separation immediately after the reactor is around 34 640 kg·h⁻¹ so very little methanol is lost through separation.

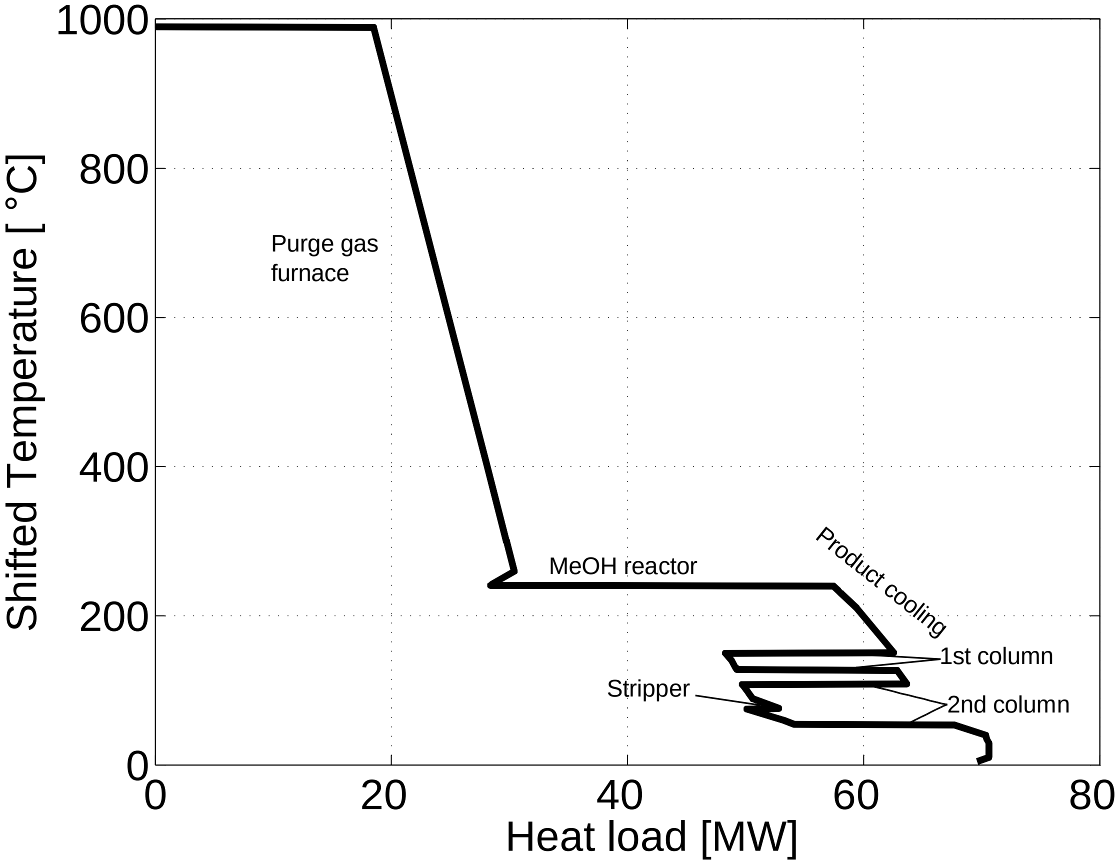

Figure 3.11: Grand composite curve of the methanol synthesis and upgrading process.

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[7] Mass fraction of MeOH is higher than 99.85 %. MeOH higher heating value: 22.65 MJ·kg^{-1}, lower heating value 19.88 MJ·kg^{-1}.

[8] Total syngas input 39 400 kg·h^{-1}, lower heating value 22.597 MJ/kg, higher heating value 25.292 MJ/kg.

The grand composite curve of the methanol synthesis and upgrading process is shown in *Figure 3.11*. It can be noticed that a considerable amount of heat is available from the purge gas furnace which can be used for steam production. It can also be noticed that an equal amount of heat is available from the methanol reactor which however operates at a lower temperature (250°C) thus making this heat suitable for MP steam production only[9].

3.1.6 System efficiency and opportunities for process integration at the Stenungsund site

3.1.6.1 Overview of heat and power requirements of the biomass based process

The production of syngas through natural gas reforming requires a large amount of high purity oxygen at high pressure. In addition, oxygen is required for the biomass to syngas process, in particular for methane reforming prior to the water gas shift unit in order to maximize the composition of H_2 and CO which can be subsequently used for synthetizing methanol.

Oxygen is assumed to be delivered by an air separation unit (ASU). Based on Aspen Plus simulation of an ASU model, the specific work consumption for delivering an oxygen stream at atmospheric pressure (95%mol O_2, 2%mol N_2, 3%mol Ar) is 0.263 $kWh \cdot kg^{-1}$, which is in agreement with the value of 0.23 $kWh \cdot kg^{-1}$ reported in Ref. (Fu and Gundersen 2011) and between 0.25 $kWh \cdot kg^{-1}$ and 0.35 $kWh \cdot kg^{-1}$ reported in Ref. (Hamelinck and Faaij 2002). Additional work is required for further compression of the oxygen stream up to the desired process pressure, here around 30 bar both for the natural gas reforming and for the reformer in the biomass to syngas process. The estimated total specific work consumption is 0.39 $kWh \cdot kg^{-1}$.

The total oxygen demand by the two syngas production processes considered in this first system configuration located at the Stenungsund site amounts to 14 935 $kg \cdot h^{-1}$, of which 11 206 $kg \cdot h^{-1}$ are used for natural gas reforming and 3 729 $kg \cdot h^{-1}$ for the reformer in the biomass to syngas process. This results in a total power demand for the ASU of around 5.8 MW.

While the syngas production from natural gas occurs in pressurized conditions, abundant syngas compression is needed in the biomass to syngas process when atmospheric gasification is used. In addition, further syngas compression is required to increase the syngas pressure from the 30 bar at which syngas is delivered from the two syngas production processes to 90 bar at which the methanol synthesis occurs. Power is also required by the recirculation blower in the methanol synthesis loop and the hydrogen recirculation blower.

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[9] The steam pressure should be low enough to guarantee adequate temperature control of the reactor and high enough for such steam to be useful for process heating.

Table 3.7: Power requirements of the methanol production process based indirect biomass gasification.

Process	***Equipment***	***Power (KW)***
Bio syngas	Water pump to gasifier	6
""	Water pump to ATR	38
""	Water pump to WGS	8
""	Syngas compressor	7 535
""	Rectisol refrigeration	1 075
NG reforming	Water pump to reformer	50
""	Rectisol refrigeration	535
ASU	O_2 production to 1 atm	3 930
""	O_2 compressor to 30 bar	1 897
methanol synthesis	Main gas compressor before synthesis	4 333
""	Synthesis loop recirc. blower	650
""	H_2 recirculation compressor	164
""	Distillation pumps	18
	Total	20 240

CO_2 removal is achieved through physical absorption and in particular by means of a Rectisol process which uses methanol as solvent. Absorption is favoured by higher pressures and low temperatures which are obtained through refrigeration. In addition, steam is required in the stripper for further separation of CO_2 from methanol in order to regenerate the absorbent.

Other ancillaries such as pumps and small compressors are also power consumers.

A breakdown of the process power requirement of the overall syngas production and methanol synthesis at the Stenungsund site is given in *Table 3.7*. Overall, around 20 MW of power are required for syngas production and methanol synthesis at the Stenungsund site.

In order to estimate heat recovery opportunities from the biomass based processes situated in Stenungsund, the grand composite curves of the three processes described above and namely the natural gas reforming (*Figure 3.7*), the biomass to syngas process (*Figure 3.9*) and the methanol synthesis process (*Figure 3.11*) are combined in a single grand composite curve which is shown in *Figure 3.12*.

Large amounts of heat appear to be available from the biomass based methanol production process, at sufficiently high temperature for use either for export of steam to the nearby chemical plants or to justify the use of heat recovery steam turbine cycle.

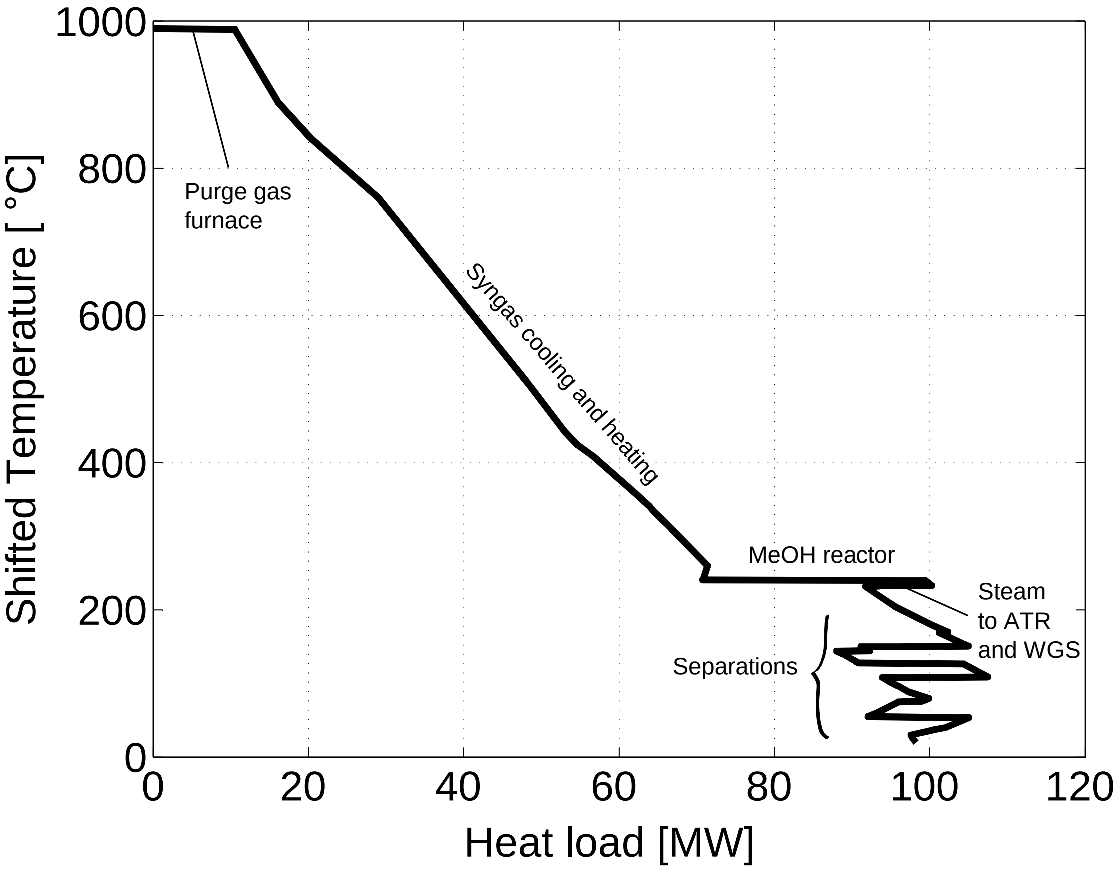

Figure 3.12: Grand composite curve of the syngas production and methanol synthesis at Stenungsund.

3.1.6.2 Overview of the steam demand at the chemical cluster

The Stenungsund chemical cluster consists of six major plants producing several chemical products, mainly starting from olefins (ethylene and propylene) currently produced by cracking of hydrocarbons (naphtha, ethane, butane, propane) at Borealis.

Although the chemical plants are highly integrated from a material point of view, the utility systems of the six plants are almost completely independent. Heat recovery and heat distribution are achieved primarily through steam networks where steam is produced in boilers and distributed at different pressure levels for process heating in each plant. A more advanced steam network exists at Borealis cracker where superheated steam is produced mainly by recovering the heat from cracker gases and thereafter expanded through turbines for power production (both for directly driving compressors and for electricity generation). Process heating is achieved with medium and low pressure steam at the turbine outlets.

According to data exchanged through internal communications with Borealis, the heat recovered from the cracker product gas cooling is less than the current steam need at cracker plant, and an additional 75 t·h^{-1} of high pressure steam must be provided by boilers fired with natural gas.

The biomass to methanol processes located in Stenungsund are considered in this project for the subsequent production of olefins through the MTO technology, and these olefins are intended to substitute part of the current fossil based production. It is therefore necessary to take into account possible effects that the substitution of olefins from hydrocarbon cracking with olefins from MTO has on the steam balance at the cracker. This type of analysis was the object of a parallel investigation within the Skogskemi project (Joelsson, Engström et al. 2015). According to internal communications with Borealis[10], the steam deficit can increase from 75 t·h^{-1} to around 140 t·h^{-1} as a result of implementing an MTO plant for a total olefin production of around 200 kt·y^{-1}.

The steam balances of the other plants within the cluster are essentially unaffected by the implementation of MTO process, so the current steam requirement can be retained for the purpose of the current analysis. Due to process operation, the steam demand can vary significantly during the year. However, average steam demands can be used to obtain reasonable indications of possible integration opportunities between the existing plants and the new biomass based processes (Hackl, Anderson et al. 2010).

Table 3.8: Overview of the cluster steam requirement for process heating purposes, adapted from (Hackl, Anderson et al. 2010). (: estimated considering MTO for around 200 kt·y $^{-1}$ olefins, high propylene yield case, based on internal communication with Borealis)*

Plant	***Pressure (barg)***	***Steam quality***	***Mass flow (t·h $^{-1}$)***	***Corresponding fuel in boilers (MW) – 80% thermal eff.***
Borealis	85	485 °C	140(*)	144.4
Akzo	40	Saturated	17.52	14.8
Akzo+INEOS	28	Saturated	12.58	10.6
Akzo+INEOS	20	Saturated	13.03	11.0
INEOS	10	Saturated	7.92	6.6
Akzo+INEOS	6	Saturated	24.75	20.5
Perstorp	2	Saturated	45.43	37.1
Akzo+INEOS	1	Saturated	12.76	10.4
Total Cluster	-			255.4

Overall, the steam demand at different pressure levels is shown in *Table 3.8.* Note that, with exception of the superheated steam at Borealis cracker, process heating is obtained

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[10] Erika Johansson, Lars Pettersson, Borealis, March 2014.

by steam condensation (latent heat). Due to heat losses and let down to lower pressures levels, condensate is returned to the boiler at temperature levels below saturation. Accordingly, steam generation must include also condensate preheating up to the saturation temperatures at the different pressure levels. Although some differences may appear between the steam systems in different plants, condensate return is assumed here at 90°C and at 1 atm, independently of the actual pressure at which steam is delivered.

3.1.6.3 Maximum steam export to the chemical cluster

The excess heat from the biomass based processes shown in *Figure 3.12*, is about half of the heat demand at the cluster (steam demand shown in *Table 3.8*). One possible option for process integration is to export as much steam as possible to the chemical plants.

Assuming that all the cluster boilers have more or less the same efficiency and are all fired with natural gas[11], export of steam from the biomass based processes to the cluster has equal effect (i.e. substitution of steam boilers) independently of the pressure levels. The pressures levels at which the steam should be exported must therefore be chosen in order to maximize the amount of steam export from the biomass based processes only. This depends on the availability of heat from the biomass based processes at different temperatures which can be read from the GCC in *Figure 3.12*.

Two relevant scenarios of steam export are considered here based on the following criteria:

- Prioritize export of steam at lower pressure levels, *Figure 3.13*.
- Prioritize the export of steam to the Borealis cracker (85 barg, 485°C), *Figure 3.14*.

In the first case (*Figure 3.13*), all the steam demand from 40 barg to 1 barg can be covered, that is all the current steam demand in cluster plants other than the Borealis cracker[12]. In addition, 5 $t{\cdot}h^{-1}$ of high pressure superheated steam can also be exported to the Borealis cracker. To satisfy the steam demand at the cracker plant around 135 $t{\cdot}h^{-1}$ of steam must be provided by the boilers.

In the second case (*Figure 3.14*), the excess heat from the biomass based processes is almost sufficient to cover all the high pressure steam demand of cracker plant. Around 112 $t{\cdot}h^{-1}$ of high pressure steam can be exported. In this case, to satisfy the steam demand of the cracker plant, around 28 $t{\cdot}h^{-1}$ of high pressure superheated steam must still be provided by boilers. No steam is exported to the other cluster plants, which therefore continue to operate the current steam boilers.

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[11] Some fuel gas is also obtained as by-product of the hydrocarbon processing at the cluster. Currently this fuel gas is burnt in boilers so the natural gas imported is only a portion of the actual fuel used for process heating. Fuel gas can be used for other purposes than combustion (e.g. hydrogen production) so that all the fuel used in boiler can be considered as imported natural gas.

[12] Steam should probably be generated at one medium and one low pressure levels and let down valves should be used to match the demand pressure levels.

In both cases, the amount of steam exported to the cluster is sufficient to substitute around 116 MW of natural gas in actual steam boilers which therefore shows that the two options are equally interesting from a primary energy saving point of view.

However, when looking at the possible implementation of the two options, there are different aspects that suggest that the export of high pressure steam to Borealis cluster is of greater interest. First of all, the biomass based processes are meant for the production of methanol which should be used for production of olefins via the MTO technology. Integrating the steam network of the biomass based processes with the cracker steam network appears a natural choice since the methanol production and steam delivery are simultaneous and so is the methanol and steam demand at the cracker plant. In addition, the amount of high pressure steam is sufficient to almost cover the estimated net steam demand at the cracker when the MTO technology is in operation. The current boilers at Borealis cracker can provide around 75 $t \cdot h^{-1}$ and can be used only in case of discontinuity of the biomass to olefin processes. An MTO plant with production capacity larger than 200 $kt \cdot y^{-1}$ can also be installed without severely compromising the steam balance at the plant if the same amount of methanol is produced by means of biomass gasification.

If steam is exported from the biomass based processes at lower pressure levels to cover the steam demand at the other cluster plants (*Figure 3.13*), the steam network must extend to a much larger area with inevitable increase of investment and no increased substitution of natural gas boilers with respect to the case where steam is only delivered to Borealis cracker. In addition, the remaining high pressure steam that should be generated at the cracker plant to balance the plant steam demand (135 $t \cdot h^{-1}$) would exceed the current boiler capacity at Borealis cracker.

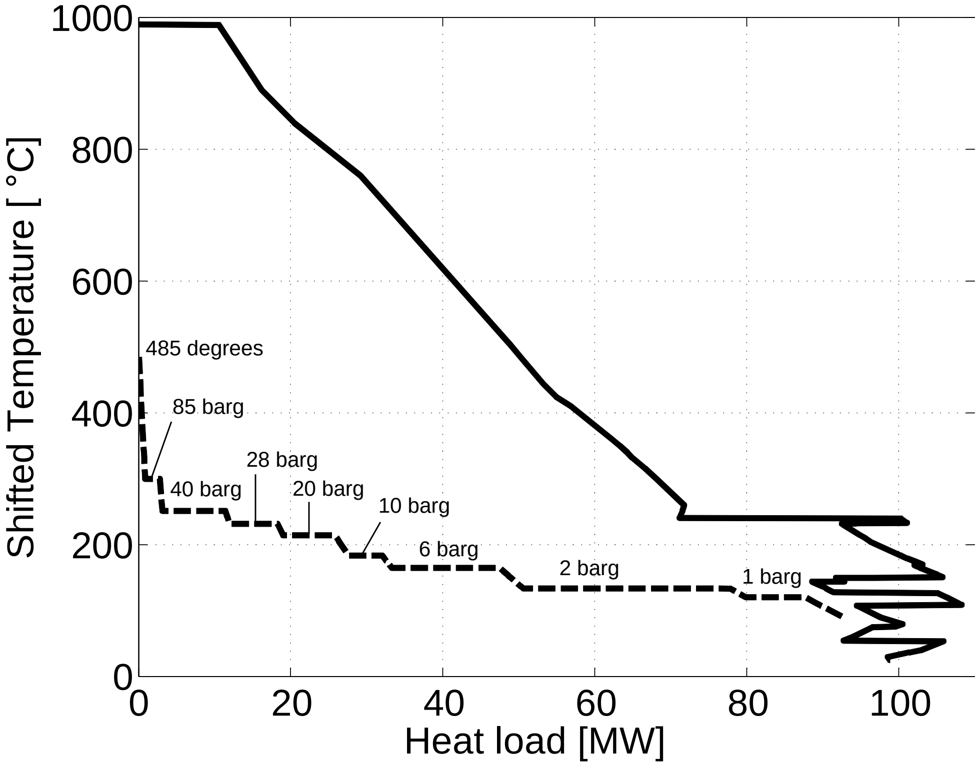

Figure 3.13: Maximum export of low pressure steam to all cluster plants.

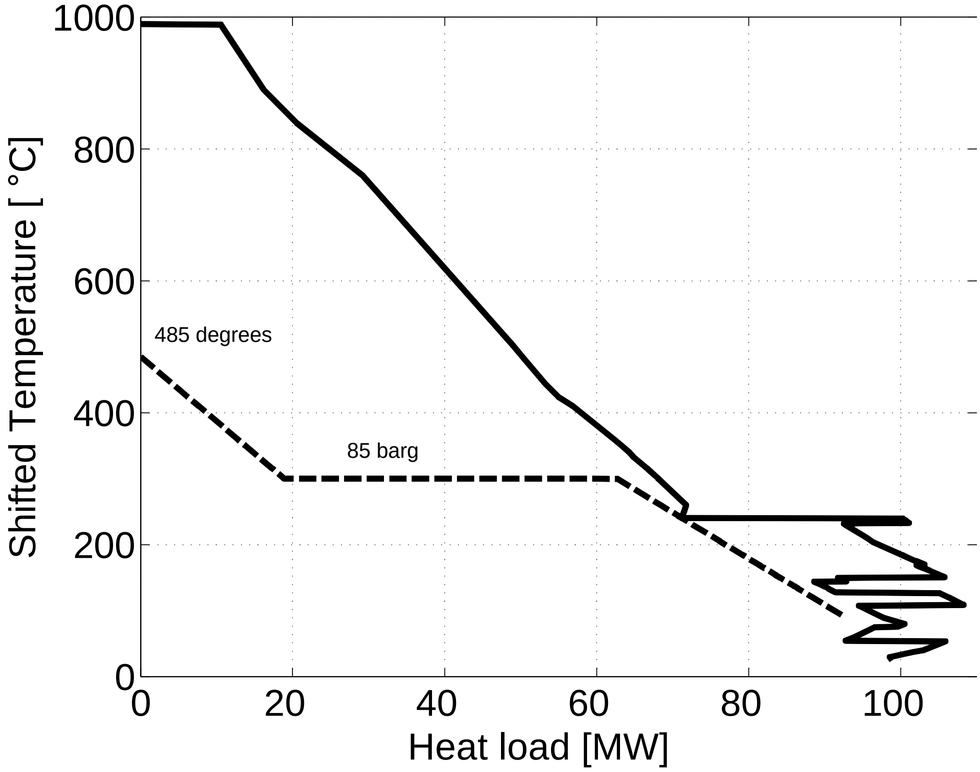

Figure 3.14: Maximum export of high pressure superheated steam to Borealis cracker.

Furthermore, it must mentioned that recent studies show that an integrated hot water network between cluster plants could largely reduce the current demand of low pressure steam at the cluster. Delivering low pressure steam from a biomass based process is not desirable if such energy efficiency measures are implemented.

A third configuration of steam export can also be considered. A quantity of high pressure steam sufficient to cover the steam deficit while still using the installed boilers (140 - 75 = 65 t·h^{-1}) could be exported to the Borealis plant and the remaining excess heat from the biomass based processes could be exported in form of medium pressure steam. This option appears very similar to the case in which the export to the cracker is maximized and is therefore not discussed in detail.

3.1.6.4 Heat recovery steam cycle for maximum power generation

The excess heat from the biomass based processes could be used to generate superheated steam at very high pressure which is then used to drive steam turbines and generate power. Power can be made available in form of shaft power to drive process compressors or sold to the electricity grid.

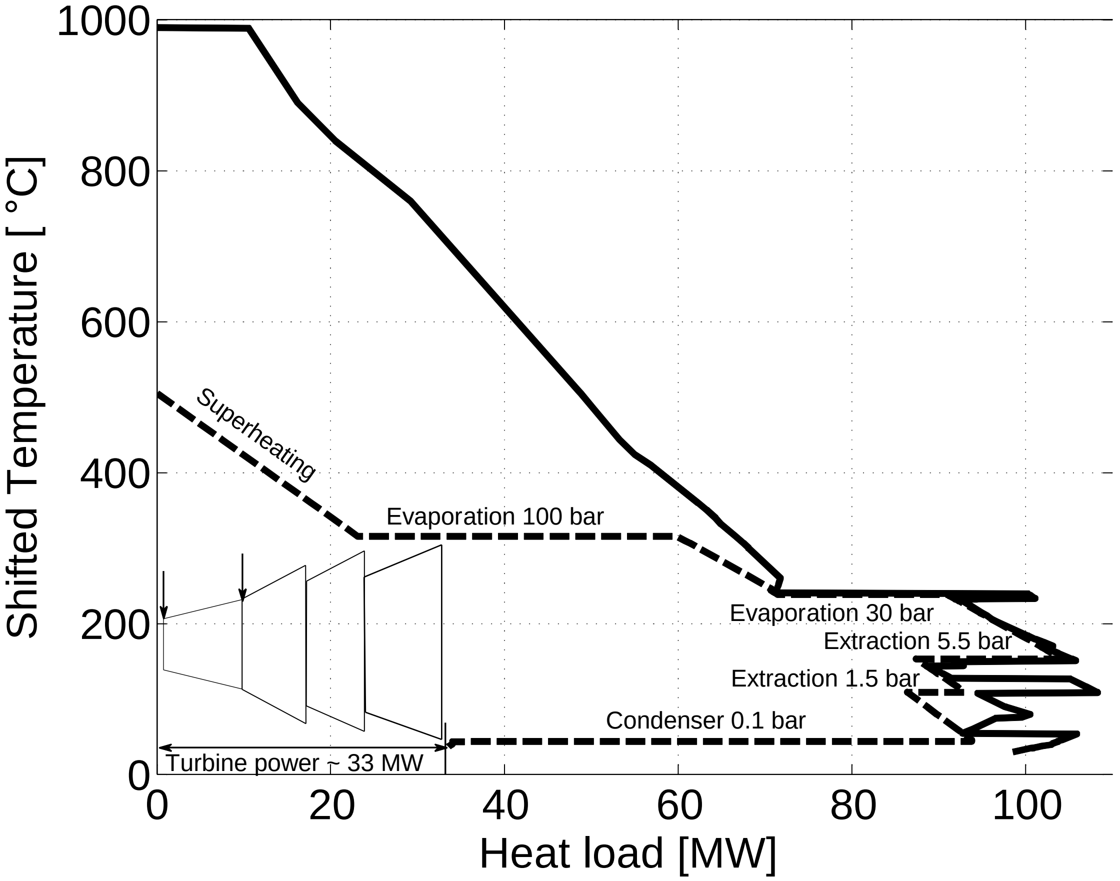

Figure 3.15: Integration of a condensing extraction steam turbine (2 pressure levels) for maximum power production.

The biomass based processes actually require some heat at medium and low temperature levels. Even if the thermal cascade in *Figure 3.12* indicates that this heat can be cascaded from high temperature sources, it can be convenient to make this heat available in the form of steam for two main purposes: (a) steam is the most common way to recover and distribute heat in large plants, (b) the high temperature heat can be recovered in form of high pressure steam to be expanded in turbines whereas low pressure steam can be used for heating purposes. Accordingly a combined heat and power system is considered here which is based on a condensing steam turbine with few steam extractions to deliver some heat to the biomass based processes.

For estimating the potential power generation with such a steam turbine system, a maximum pressure of 100 bar and a maximum temperature of 500°C were considered. The shape of the thermal cascade in *Figure 3.12* suggests that some high pressure steam can also be generated from the methanol reactor at a pressure around 30 bar. A limit of 0.1 bar pressure was considered for the steam condenser. A set of optimal pressures level of steam extraction was considered: 5.5 bar and 1.2 bar.

As shown in *Figure 3.15*, the maximum power generation is around 33 MW which corresponds to a net production of around 13 MW after the power needed for the biomass based processes is covered (gas compression, ASU, Rectisol, etc.).

3.1.6.5 Combined heat and power system

As previously discussed, maximum steam export to the cluster could result in a potential saving of around 116 MW of natural gas. Among the different pressure levels at which steam can be exported, the export of high pressure superheated steam to Borealis cracker appears to be the simplest solution, with equally high natural gas substitution effects (*Figure 3.14*). However, it should be noticed that production of electric power has also significant substitution effects in marginal electricity producers. The production of electricity by means of a condensing turbine is a poor solution (*Figure 3.15*) due to the low efficiency compared to state-of-the-art power plants. A way to maximize the efficiency of excess heat recovery is to generate high pressure superheated steam to be expanded in a counter-pressure steam turbine and use the steam at the turbine outlet for process heating.

The cluster steam demands at low pressure levels can be covered by steam extractions and by counter-pressure steam. This allows shifting from a current steam network based on fossil fuel boilers for process heating only to a CHP system based on excess heat from biomass conversion processes.

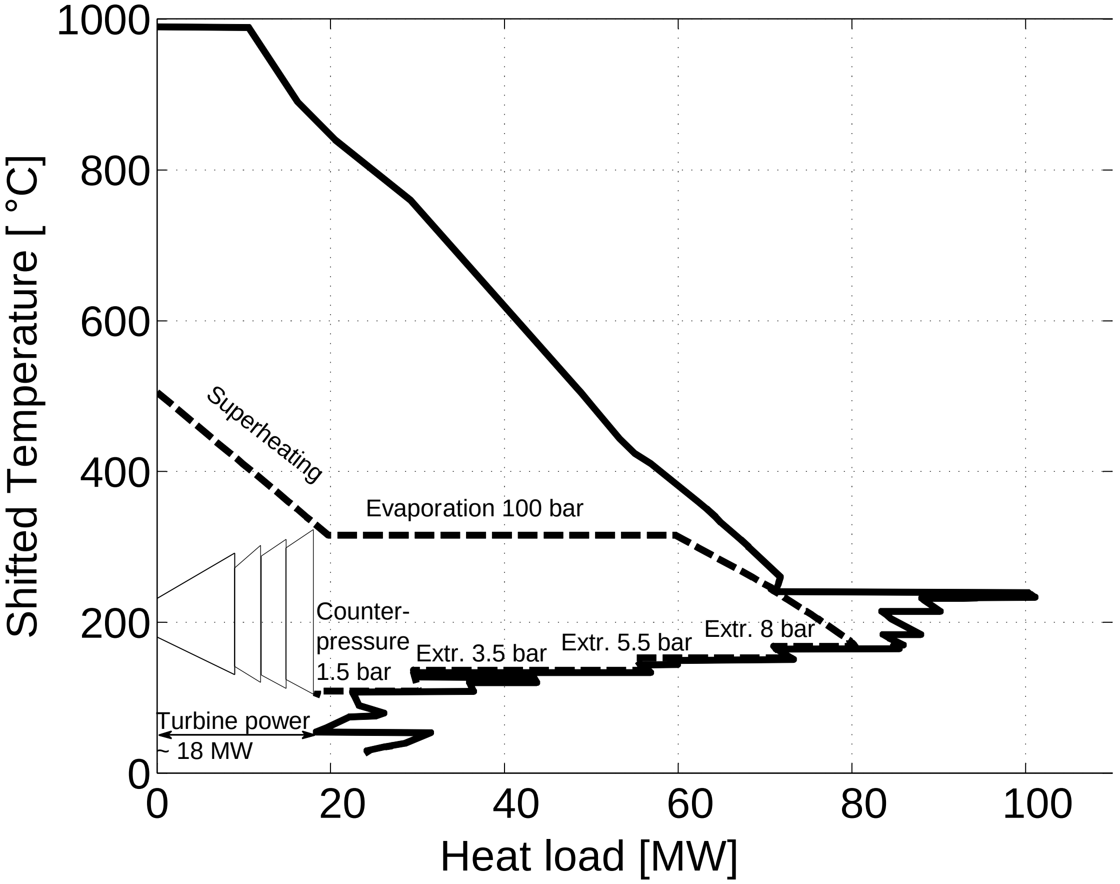

Figure 3.16: Optimal combined heat and power production with integrated counter-pressure extraction steam turbine.

Table 3.9: Steam export to cluster plants from the suggested integrated CHP system.

Steam pressure (barg)	***Mass flow rate (t h $^{-1}$)***
28	9.00
20	13.03
10	7.92
6	24.75
2	45.43
1	12.76

The suggested system is shown in *Figure 3.16*, with the same maximum steam data values as those considered in previous sections (100 bar and 500°C). With respect to *Figure 3.15*, no steam is generated at 30 bar due to a lower margin for steam expansion to lower pressure levels. In addition, a different set of steam extraction was considered in order to maximize the low pressure steam export to cluster plants while still allowing for power generation by steam expansion: 8 bar, 5.5 bar, 3.5 bar. The turbine back-pressure is fixed at 1.5 bar.

The suggested CHP system generates around 18 MW of power which corresponds to a net electricity demand of the biomass based processes of around 2 MW. The potential steam export to the cluster is sufficient to cover all the demands below 28 barg and around 70% of the steam demand at 28 barg whereas no steam is exported at 40 barg pressure level (*Table 3.9*). As all the superheated steam is used here to power production through steam expansion, no superheated high pressure steam is exported to the Borealis cracker.

The steam export to the cluster plant allows about 93 MW of natural gas savings in cluster steam boilers. Due to the more complex arrangement of the steam network that has to be able to deliver heat to multiple cluster plants, this system may require a large investment. In addition, as already underlined, the export of low pressure steam may be less interesting in the case increased heat recovery between cluster plants is made possible in the future by means of an integrated hot water network.

3.1.6.6 Summary of the process integration opportunities and system consequences

Locating the bioSNG plant in the proximity of the Värö pulp mill introduces a substantial logistic advantage compared to stand-alone SNG plants as biomass transport and storage and, to a large extent, pre-treatment facilities can be upgraded to accommodate the extra biomass intake required for SNG production. However, no extra steam than that provided by the current steam network at the mill based on recovery boiler is required, and process heating at the mill is essentially based on steam turbine extractions already on site. Even an integrated steam network between the bioSNG process and the pulp mill does not show any benefit compared to separate steam systems. The excess heat from the bioSNG process should therefore be used for power generation to counterbalance the power required by syngas and SNG compression.

At the Stenungsund chemical cluster site, the integration opportunities are much more apparent. The large steam demand currently supplied through natural gas combustion in steam boilers can be covered by exporting the excess heat in form of steam from the biomass based processes (biomass gasification, natural gas reforming, methanol synthesis and upgrading).

An overview of the total energy balance of the first system alternative is given in *Table 3.10.*

Under the condition that the imported electricity is generated by a fuel-based production technology, it is possible to give a more general picture of the performance of the proposed biomass to methanol system by introducing the “total conversion efficiency” of biomass into methanol.

Table 3.10: Overview of steam export and electricity import of the first system alternative under different conditions of process integration.

Configuration	***Maximum steam export to Borealis***	***Maximum steam export to other plants***	***Maximum power generation***	***Combined power and steam export***
Steam export to Borealis cracker (t·h $^{-1}$)	*112*	*5*	-	*0*
Steam export to other plants (t·h $^{-1}$)	0	134	-	113
Fuel saving in cluster boiler (MW NG)	116	116	0	93
Power generation (MW)	-	-	*33*	*18*
Steam turbine configuration	-	-	Condensing / extraction - 2 press.levels	Back-press. / extraction
Electricity import (MW)	*20*	*20*	*-13(export!)*	*2*

As shown in the methodology chapter (Equation 2.2), this is similar to the conversion efficiency from biomass to methanol (i.e. the ratio between the energy rates of produced methanol and biomass input) with the only difference that the total input energy is accounted for. This is done by deducing from the total energy rate of biomass input the fuel saved in boilers at the Stenungsund site due to the steam export from the biomass based processes to the chemical plants and by adding the fuel spent in marginal electricity production technology to compensate for the electricity imported by the biomass based processes (or deducing this fuel in case of export of electricity). Note that a certain quantity of RME is also used in the oil scrubber for tar removal (6.9 MW HHV basis, 6.5 MW LHV basis in total), and the corresponding chemical energy flowrate is also added to the denominator in the above equation since it represents an input to the system.

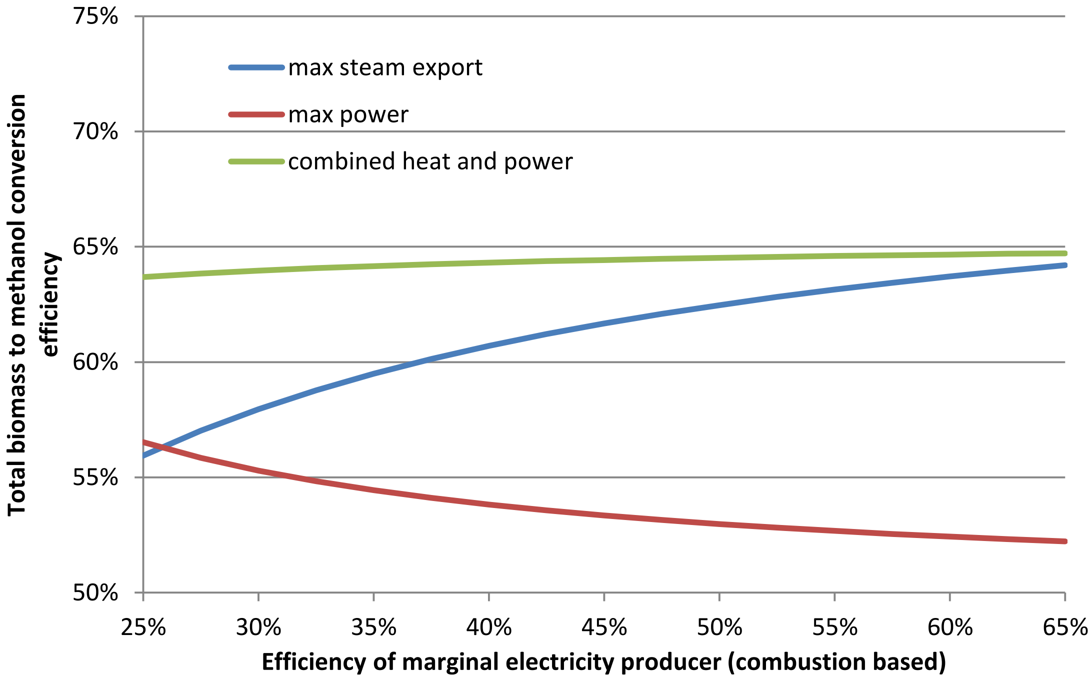

Figure 3.17: Effect of electrical efficiency of combustion based marginal electricity producer on the total biomass to methanol conversion efficiency for the first system alternative and for different process integration options.

Accordingly, the total conversion efficiency depends on the efficiency of the boilers currently used for steam production and on the efficiency of marginal electricity producers.

It is interesting at this stage to show the influence of efficiency of marginal electricity producer on the total conversion efficiency of biomass to methanol while a more thorough analysis of the consequences on carbon dioxide balances and process economics was the object of parallel investigations within the Skogskemi project (Joelsson, Engström et al. 2015). Results are shown in *Figure 3.17*, while a breakdown of main energy flows around the biomass based methanol production processes is given in *Table 3.11* considering two efficiency levels for marginal electricity producers.

It is possible to conclude that process integration options and in particular the export of steam to the cluster plants allows a large increase in the overall conversion efficiency.

Power production with an integrated condensing steam turbine is not of interest as the efficiency is significantly lower than that of the marginal electricity producers (only if the marginal efficiency is lower than 25% can it be of interest to produce power locally to reduce the penalty introduced by marginal fuel consumption). The opportunities to save natural gas in cluster steam boilers are larger compared to the marginal effect of power production.

From a purely thermodynamic point of view, a CHP system results in the highest substitution of fuel in boilers and marginal electricity producers. However, electricity production has a minor impact on the total conversion efficiency compared to the impact of export of steam to the cluster steam network. If the latter is maximized an equally high total conversion efficiency is obtained, especially if the efficiency of marginal electricity producers is high. Considering the big difference of capital investment between building a CHP system delivering steam to all the other cluster plants and exporting high pressure superheated steam to the nearby Borealis cracker, the latter option is the most attractive.

Table 3.11: Overview of performances of the biomass to methanol system based on indirect gasifier technology, for different process integration options with the existing chemical cluster plants.

Configuration	***Maximum steam export to Borealis cracker***	***Maximum steam export to other plants***	***Maximum power generation***	***Combined power and steam export***
Biomass input energy rate (MW) HHV basis	426.7 (+6.9MW RME)	426.7 (+6.9MW RME)	426.7 (+6.9MW RME)	426.7 (+6.9MW RME)
Methanol output energy rate (MW) HHV basis	216.15	216.15	216.15	216.15
Biomass to methanol conversion efficiency	*50.8%*	*50.8%*	*50.8%*	*50.8%*
NG[13] saving in cluster (MW), 80% eff.	116 (LHV basis) 128 (HHV basis)	116 (LHV basis) 128 (HHV basis)	0	93 (LHV basis) 103 (HHV basis)
Electricity import (MW)	*20*	*20*	*-13(export!)*	*2*
Marginal fuel to electricity (MW)				
... 40 % electrical efficiency	50.0	50.0	- 32.5	5
... 60 % electrical efficiency	33.67	33.67	- 21.33	3.67
Total conversion efficiency (HHV basis)				
... 40 % marg. el. eff.	*60.7%*	*60.7%*	*53.8%*	*64.3%*
... 60 % marg. el. eff.	*63.7%*	*63.7%*	*52.4%*	*64.7%*

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[13] The natural gas lower heating value is 47.994 MJ/kg, and the higher heating value is 53.096 MJ/kg.

3.2 System configuration nr 2

3.2.1 Description

The second system configuration for methanol production is based on an oxygen-steam blown biomass gasifier and subsequent methanol synthesis located in proximity of the Iggesund pulp and paper mill on the Swedish East Coast. A representation of the investigated system is shown in *Figure 3.18*.

The biomass derived methanol is transported to the Stenungsund MTO plant where it is mixed with other imported methanol in order to achieve a total olefin production of 200 kt per year of olefins. An optional quantity of ethanol can also be mixed with methanol to feed the MTO process, in which case a different ratio of C2 and C3 olefins is obtained.

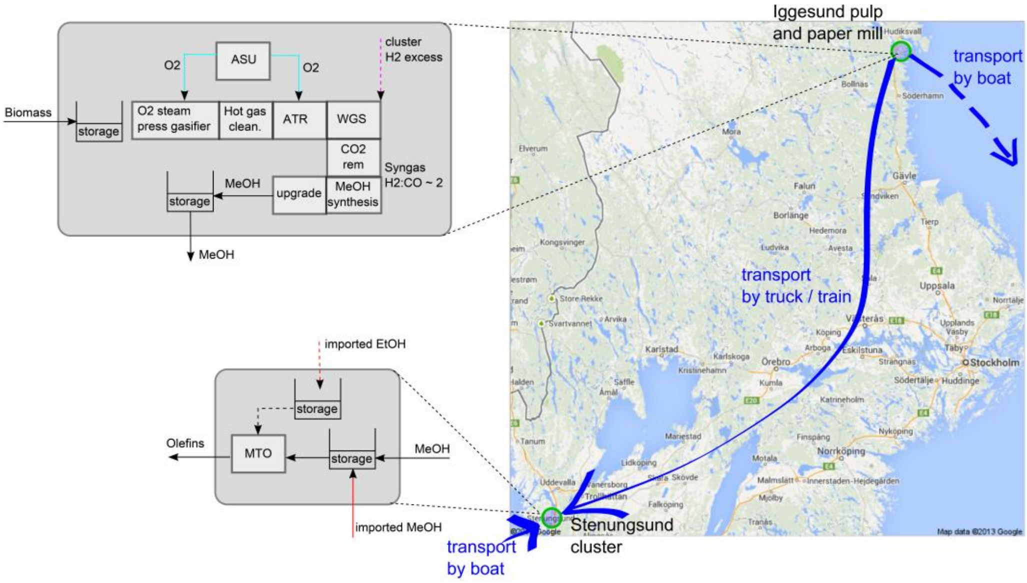

Figure 3.18: Overview of system configuration nr 2: biomass gasification and methanol synthesis based on oxygen steam blown pressurized gasifier.

This second configuration differs from the first one in several important ways. The whole gasification and methanol synthesis is located far from the Stenungsund cluster and close to the Iggesund pulp and paper plant. The energy integration opportunities are therefore related to the possibility of improving the steam balance and/or the electricity production at the mill. Changes in fuel and steam balance at the Stenungsund cluster are related to the integration of the MTO plant only. Methanol must be transported from the Iggesund site to the Stenungsund cluster either by truck or train or boat which contributes to an increase of the final methanol cost and reduces the offset of CO_2 emissions of the biomass feedstock if we assumed that transportations run on fossil fuel.

3.2.2 Iggesund pulp and paper mill

The pulp and paper mill at Iggesund is own by Holmen. Around 355 kt of paper and 40 kt of pulp are produced yearly in average. The steam and power requirement at the mill are such that the recovery boiler is not sufficient to cover the whole mill steam demand and additional fuel is required. This consists partly of bark directly produced at the mill and of fossil fuel import, partially heavy oil.

*Table 3.12: Key data of Iggesund pulp and paper mill (*according to internal communication with Holmen based on 2013 data; ** according to internal data at Chalmers).*

Description	***Value***
Yearly paper production **	350 000 Adt
Yearly pulp sales to market ** ("avsalu")	58 000 Adt
Recovery boiler maximum fired capacity *	270 MW
Additional boiler maximum fired capacity *	115 MW
Yearly fossil fuel consumption **	350 GWh
Yearly bark consumption in additional boiler **	482 GWh
Turbine nominal power **	70 MW
Turbine typical power *	57 MW
Average district heating export *	13 MW

3.2.3 Oxygen steam blown pressurized gasifier and methanol synthesis

This section presents the mass and energy balance of a methanol production process based on oxygen-steam blown pressurized gasification with 450 MW_{LHV} biomass input consisting of wood chips at 50% moisture. This corresponds to a biomass intake of 189 360 $kg \cdot h^{-1}$, corresponding to around 1 514 $kt \cdot y^{-1}$ assuming 8000 hours of operation per year.

Mass and energy balances of the biomass conversion to methanol were estimated by process modelling using published models developed at the Div. of Heat and Power Technology Chalmers (Isaksson, Pettersson et al. 2012). The main assumptions for the gasification system were taken from published literature data and in particular from the results of activities conducted by the Finnish VTT Technical Research Center (Hannula and Kurkela 2010, Hannula and Kurkela 2013). A picture of the process is shown in *Figure 3.19* where the main process operating parameters are also given.

This system for biomass based methanol differs from the concept based on indirect gasification investigated in the first system configuration only for the gasification technology and on the gas cleaning section.

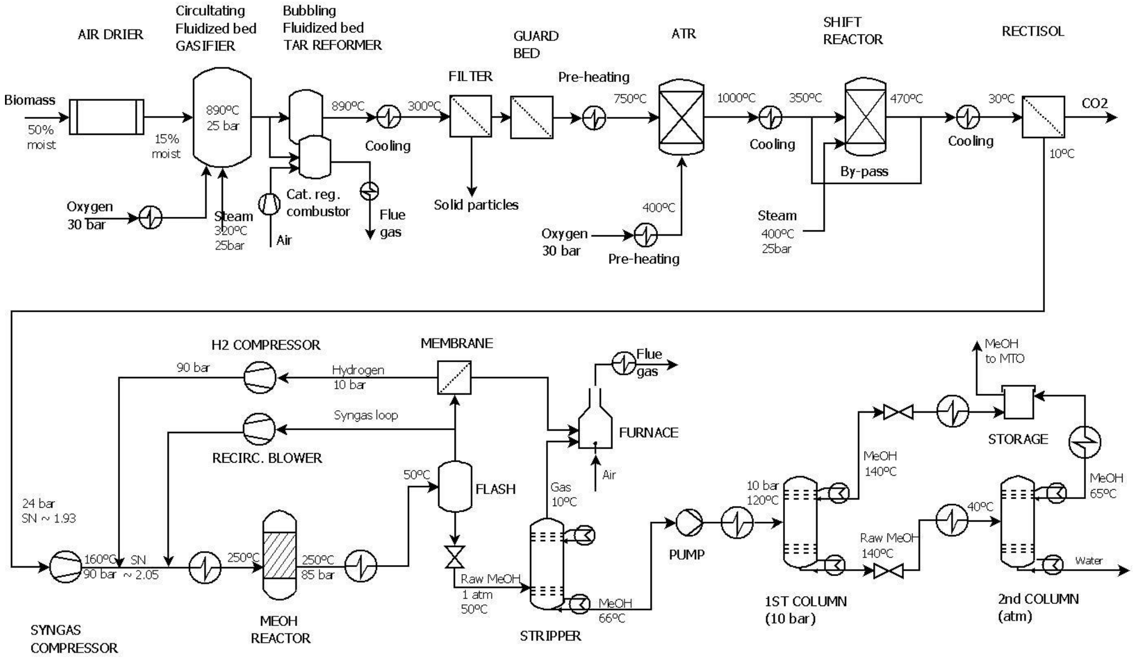

Figure 3.19: Layout of the second system configuration for a biomass gasification and methanol production process.

Here biomass is gasified in pressurized conditions (at 25 bar) and in presence of oxygen and steam. Pressurization of the whole gas production process through a pressurized biomass feeding system introduces significant savings in the subsequent gas compression required to achieve the optimal synthesis pressures both in terms of capital and operating costs. However, since combustion and gasification occur in the same reactor, to avoid large amount of inert nitrogen the direct gasification concepts necessitates a pure oxygen stream. Gasification occurs also in presence of steam which is required as a reforming agent. These latter endothermic reactions require energy to be provided by combustion and therefore the consumption of oxygen increases with the steam input. As reported in Ref. (Hannula and Kurkela 2012) the optimal ratio between oxygen and steam for a temperature around 886°C and for a pressure of around 25 bar is around 1:1.

The biomass is converted into around 176 981 $kg{\cdot}h^{-1}$ of gas with the composition reported in *Table 3.13*. The higher heating value (HHV) of this gas is 9.191 $MJ{\cdot}kg^{-1}$, the lower heating value (LHV) is 7.790 $MJ{\cdot}kg^{-1}$.

The gasifier cold gas efficiency is therefore:

$$\eta_{CG}^{HHV} = \frac{\dot{m}_{gas} \cdot HHV_{gas}}{\dot{m}_{biomass} \cdot HHV_{biomass}} = 82.4\%; \qquad \eta_{CG}^{LHV} = \frac{\dot{m}_{gas} \cdot LHV_{gas}}{\dot{m}_{biomass} \cdot LHV_{biomass}} = 85.1\%.$$

Table 3.13: Composition of the product gas obtained from gasification of wood chips in a oxygen steam blown gasifier at 890°C and 25 bar.

Component	***Mol. Frac. (%)***
H_2	20.1
CO	15.8
CO_2	21.9
H_2O	34.1
CH_4	5.4
N_2	0.027
O_2	-
Acetylene	0.035
Ethylene	1.476
Ethane	0.377
Propane	0.017
Benzene	0.328
Naphthalene	0.364
Ammonia	0.050
H_2S	0.003
HCl	0.005

The gasification is followed by an isothermal (890°C) catalytic bubbling fluidized bed in which the tar compounds are reformed in presence of the abundant steam content of the product gas. Data about such a reformer were obtained from publications by the US NREL (Spath, Aden et al. 2005). Tar compounds such as Naphthalene, light hydrocarbons such as propane and ethane as well as ethylene and acetylene are found in the product gas from the gasifier and are largely reformed into H_2 and CO by catalytic cracking. Methane is also partially cracked although about half of that still remains in the gas at the reformer outlet. The heat for reforming is provided by circulating the bed and the catalyst from a side combustor which is fuelled by a certain quantity of product gas (about 10% of the total product gas) that is by-passed prior the reformer and therefore does not contribute to methanol production.

At the outlet of the tar cracker, particulate matter, alkali materials, and sulphur compounds are still present in the gas and must be removed prior to the gas upgrading and synthesis reactions. In this second system configuration, this is done by hot gas cleaning technologies and in particular by candle filters. The sulphur, assumed here completely in the form of H_2S, is removed through a guard bed based on metal oxides.

In order to finally convert the remaining hydrocarbons that would otherwise remain as inert in the methanol synthesis process, a catalytic autothermal reformer was considered as in the first system alternative. Here, however, enough steam is present in the gas after the gasification and the tar cracker, and steam injection is not required as the steam to carbon ratio is already higher than 1.

The syngas at the reformer outlet is then cooled to a temperature around 350°C and sent to a high temperature water gas shift reactor where the final ratio between H_2 and CO is obtained. A certain amount of gas is also bypassed so that only the required amount of gas is shifted. The gas is subsequently cooled to ambient temperature and most of the CO_2 is removed by a Rectisol process where methanol is used as a physical absorbent. A final CO_2 concentration of 3% in the dry syngas is obtained as it is the optimal concentration for subsequent methanol synthesis. The syngas is finally compressed up to the synthesis pressure of 90 bar.

A certain amount of hydrogen is recovered from a purge stream after the methanol synthesis which is mixed to the fresh syngas feed to reach the final synthesis ratio[14] of around 2.05 with a 3% CO_2 molar fraction. The syngas is then mixed with the synthesis loop gas obtained at the top of the flash drum right after the reactor. Here the synthesis loop is assumed to be 3:1 in volume with respect to the fresh syngas.

Methanol synthesis occurs in a fixed bed reactor where steam is produced by reactor cooling. The reaction products are subsequently cooled leaving a two-phase stream. The crude methanol is obtained at the bottom of the flash drum and is then sent to distillation after the pressure has been released to atmospheric values. A stripper is used to evaporate the remaining gases entrained in the raw methanol. The methanol rich liquid still present a large quantity of water which is removed in a two column system.

The purge gases are burnt thus producing additional heat that can be used for steam production.

More details about the modelling assumptions and stream composition and characteristics in the different process steps are given in the Appendix.

Overall, a biomass input of 450 MW_{LHV} (548.7 MW_{HHV}) yields a production of 53 534 $kg \cdot h^{-1}$ of methanol (336.8 MW_{HHV}[15], 295.6 MW_{LHV}), which corresponds to a conversion efficiency of 61.3% HHV basis (65.7% LHV basis).

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[14] $(H_2 - CO_2)/(CO + CO_2)$

[15] MeOH higher heating value: 22.65 MJ kg^{-1}, lower heating value 19.88 MJ kg^{-1}.

3.2.4 System efficiency and opportunities for process integration with nearby Iggesund pulp and paper mill

3.2.4.1 Overview of heat and power requirements of the biomass based process

Heat recovery opportunities from the biomass to methanol process were estimated based on the results of process simulation and by including the hot and cold thermal streams (cooling and heating process steps) in a single GCC curve as according to Pinch Analysis, see *Figure 3.20*.

Most of the heat from the methanol synthesis reactor is recovered in form of steam at around 25 bar which is used for gasification and water gas shift. The remaining heat can be used for producing steam to run the first distillation column.

Abundant heat is available instead from syngas cooling, which can be used for high pressure steam production.

Note that heat is required by the first distillation column which is run at 10 bar as well as for biomass drying. Still, no other large heat demand appears in the biomass to methanol process which opens the opportunity for condensing steam turbine in case power shall be maximized or for steam export to the nearby pulp and paper plant.

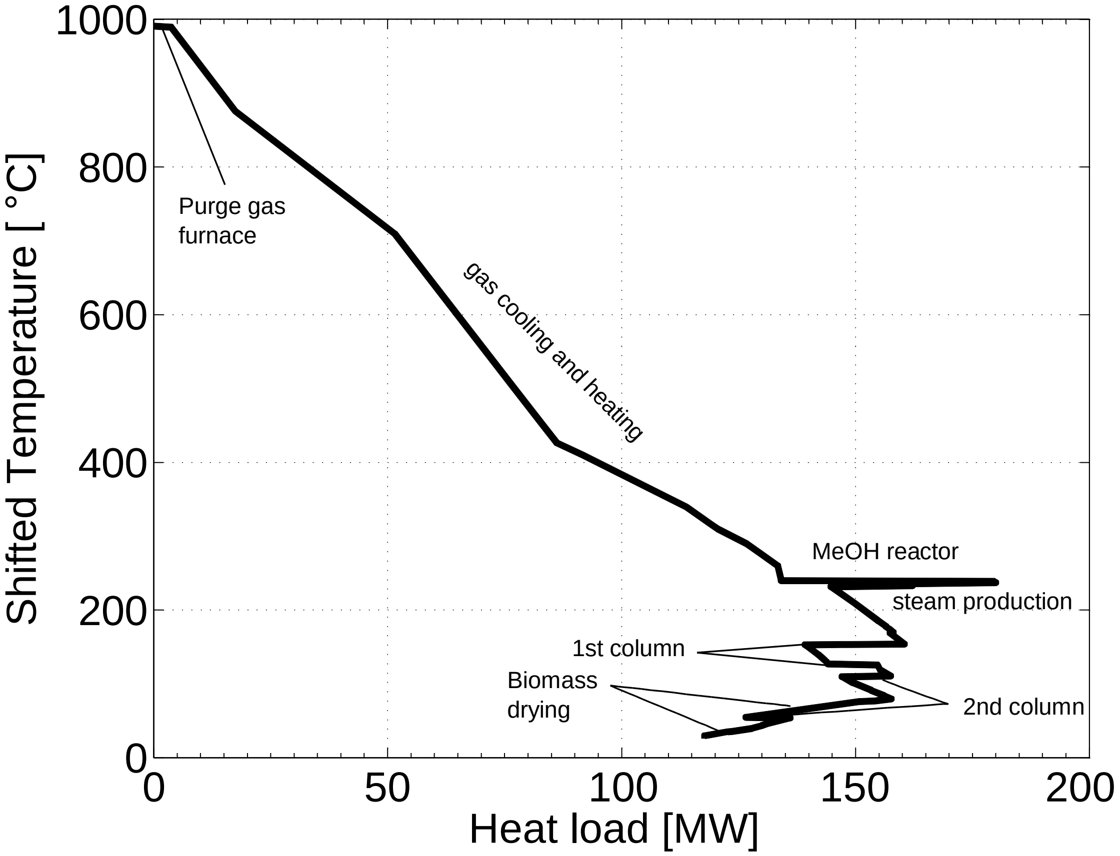

Figure 3.20: GCC of the methanol production process based on oxygen steam blown biomass gasification.

The production of methanol from biomass through an oxygen steam blown gasifier requires a large amount of high purity oxygen at high pressure. The biggest share of the oxygen is used for the gasification itself. Some oxygen is also used for the catalytic reforming of hydrocarbons (mostly methane) prior to the water gas shift, in order to further recover them in form of H_2 and CO which can be subsequently used for synthetizing methanol.

Oxygen is considered to be delivered by an air separation unit (ASU). Based on Aspen Plus simulations, the specific work consumption for delivering an oxygen stream at atmospheric pressure (95%mol O_2, 2%mol N_2, 3%mol Ar) is 0.263 $kWh \cdot kg^{-1}$, which is in agreement with the value of 0.23 $kWh \cdot kg^{-1}$ reported in Ref. (Fu and Gundersen 2011) and 0.25 to 0.35 $kWh \cdot kg^{-1}$ reported in Ref. (Hamelinck and Faaij 2002) . Additional work is required by further compression of the oxygen stream up to the desired process pressure, here around 30 bar. The total specific work consumption has been estimated to be around 0.43 $kWh \cdot kg^{-1}$ kg^{-1}oxygen.

The total oxygen demand in this second system alternative amounts to 42 747 $kg \cdot h^{-1}$, of which 32 835 $kg \cdot h^{-1}$ for the gasifier and 9 914 $kg \cdot h^{-1}$ for the autothermal reforming. This results in a total power demand for the ASU and further oxygen compression of around 18.3 MW.

Although pressurized gasification helps to reduce the gas compression (at least for operating pressures up to 25 to 30 bar) compared to the indirect gasification concept, power is still required for compressing the gas up to the synthesis pressure of 90 bar. In addition, power is required by the recirculation blowers in the methanol synthesis loop.

Table 3.14: Power requirements of the methanol production process based on oxygen steam blown biomass gasification.

Equipment	***Power (KW)***
Biomass feeding system (piston)	2 250
Pumps	146
Main gas compressor before synthesis	9 120
Air compressor to tar reformer combustor	8 783
Synthesis loop recirculation blower	1 021
H_2 recirculation compressor	735
Rectisol refrigeration	3 804
ASU	11 255
O_2 compressor	7 053
Total	44 167

Compression is also required for pressurizing the air needed for the catalyst regenerator combustor and bed reheating for the tar reformer.

A piston-type biomass feeding system is assumed for biomass injection in pressurized gasification conditions and 5 kW electricity consumption per 1 MW fuel (LHV basis) is considered.

CO_2 removal is achieved through physical absorption in a Rectisol process with methanol as solvent. Absorption is favoured by high pressures and low temperatures which are obtained through refrigeration. In addition, steam is required in the stripper for further separation of CO_2 from methanol in order to regenerate the solvent.

Other ancillaries such as pumps and small compressors are also power consumers.

A breakdown of the process power requirement is given in *Table 3.14*. Overall, a total power consumption of around 44.2 MW is estimated.

3.2.4.2 Steam deficit at the Iggesund pulp and paper mill

According to the data available from internal communication with Holmen[16], the steam produced in the recovery boiler is not sufficient to satisfy the mill steam demand and extra steam needs to be produced in additional bark and oil boilers. There are currently two large additional boilers of around 115 MW fired capacity each. However one of these boilers is not equipped for bark combustion and is used only as a back-up. The bark boiler can also accommodate an extra quantity of fuel than bark and oil is also burnt in this boiler when the mill production is at the maximum.

In the primary additional boiler, steam is produced at 60 bar and 450°C and injected in the second turbine stage (the turbine primary inlet pressure is 115 bar which is the pressure of the steam produced by the recovery boiler). The bark and oil boilers were assumed to have 87% efficiency (LHV basis).

Considering a total fuel firing capacity of 115 MW of which 75 MW are bark and remaining 40 MW are oil, the steam production capacity is around 100 MW. Based on the declared specific enthalpy of the return condensate of 525 $kJ \cdot kg^{-1}$, the total steam production capacity[17] was estimated to be around 130 $t \cdot h^{-1}$.

3.2.4.3 Maximum steam export to Iggesund pulp and paper mill

The GCC curve of the biomass based methanol production process (*Figure 3.20*) shows a large heat excess that could be recovered for steam generation.

A first process integration option between the biomass based process with the nearby pulp and paper mill consists in producing steam at pressure and temperature levels that

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[16] Klas Simes, Holmen, april 2014.

[17] An analysis of recent plant data revealed that only around 480 GWh of bark are burnt per year, corresponding to an average of 56 MW of fuel supply rate, which is lower than the maximum bark firing capacity (75 MW). Assuming a proportional share of oil based on the nominal boiler capacity, the average load of the additional boiler amounts to around 85 MW (30 MW extra oil co-fired). The steam produced in this boiler amounts to around 96 $t \cdot h^{-1}$, which is considerably lower than the nominal value of 130 $t \cdot h^{-1}$ calculated here.

are compatible with the steam that is produced by the additional bark and oil boiler at the mill.

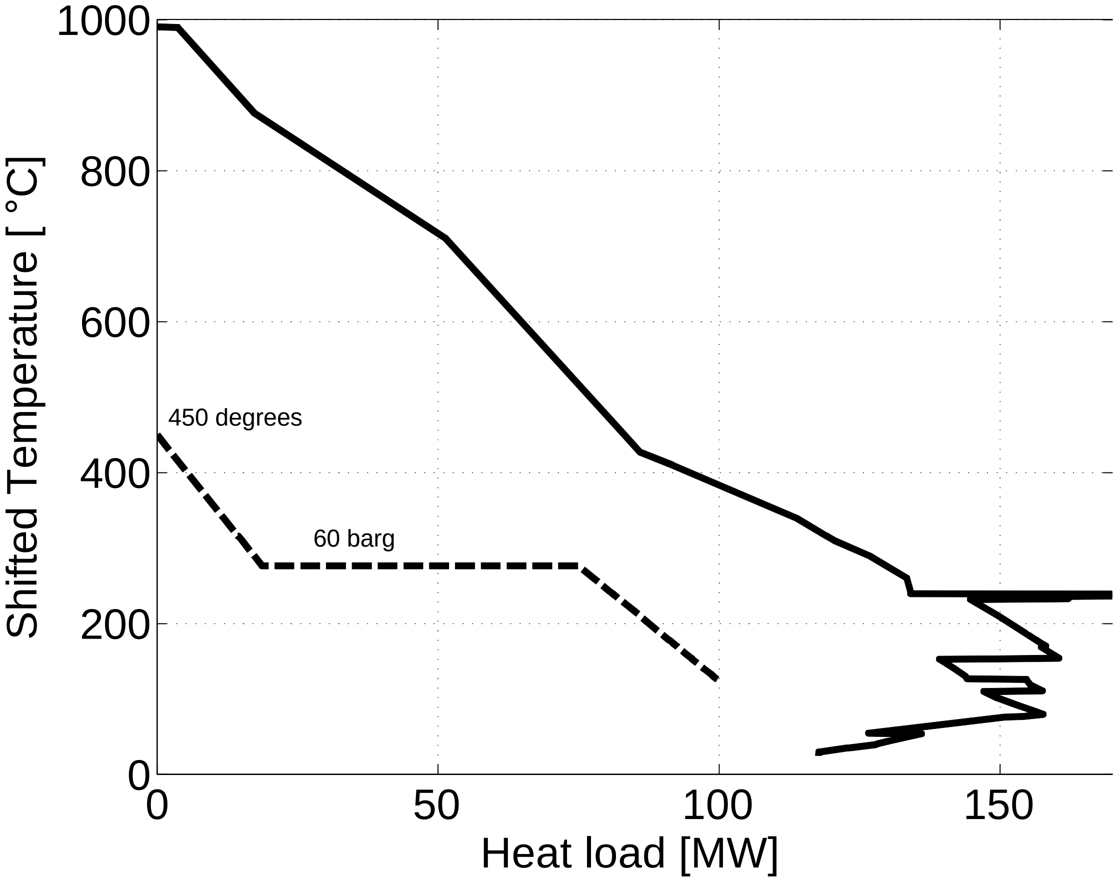

Figure 3.21: Maximum steam export to the Iggesund pulp and paper mill.

Figure 3.21 illustrates the heat flows corresponding to the maximum steam production that can be accomplished by recovering the excess heat from the biomass based processes. The excess heat from the biomass based methanol production process exceeds the heat corresponding to the nominal steam production capacity of the oil and bark boiler (100 MW, 130 $t{\cdot}h^{-1}$ of steam at 60 bar 450ºC). The lower cost of steam from the biomass based methanol process compared to steam produced by oil combustion could in fact justify a larger steam flow to the turbine and an increased mill production. In absence of detailed data about the steam turbine and the mill steam system, the assumed maximum steam export was limited to the nominal capacity of the bark and oil boiler currently installed at the mill. In such case, a total fuel energy rate of around 115 MW to the bark and oil boilers would be replaced by steam export from the biomass based methanol process.

Based on nominal data of the bark and oil boiler, the nominal fired capacity of 115 MW is covered by a maximum bark share of 75 MW which leaves 40 MW to be provided through oil combustion. Thus 75 MW bark could be used to partially covered the total biomass intake for methanol production, reducing the total intake from 450 MW to 375 MW_{LHV}.

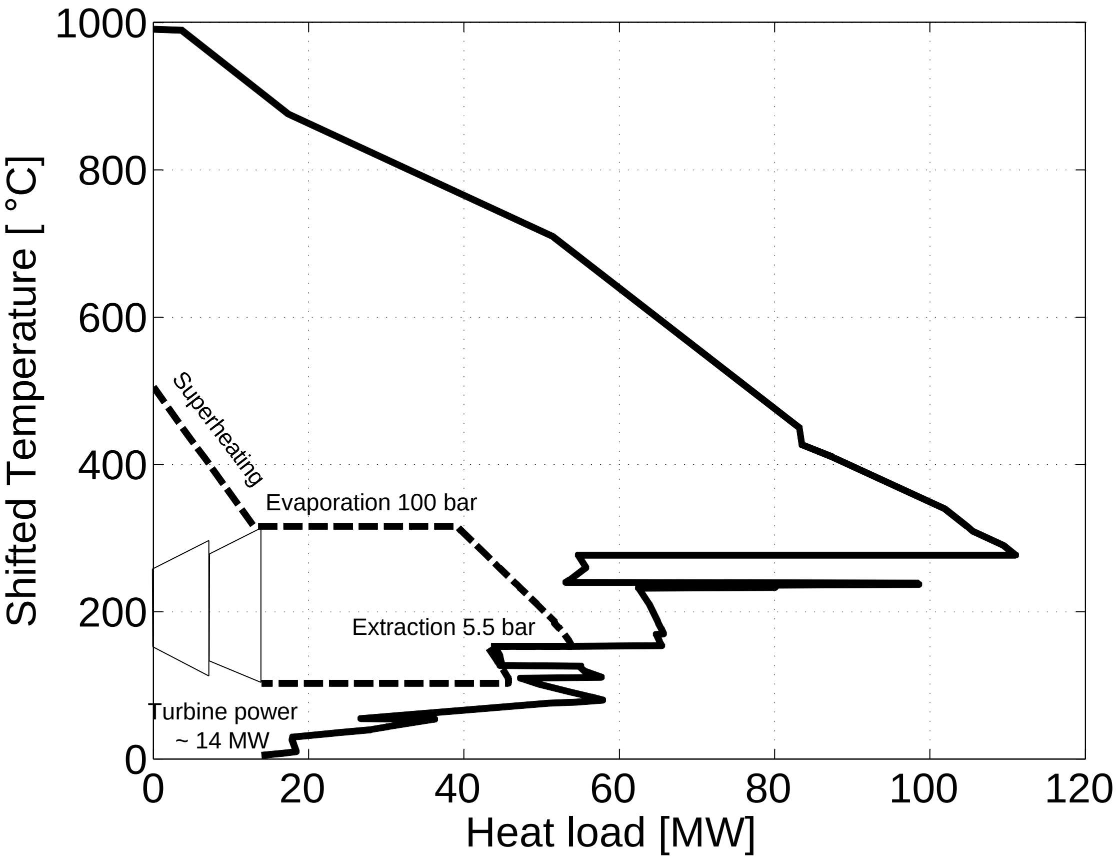

Figure 3.22: Integration of a back-pressure steam turbine for recovering the remaining excess heat at the net of the maximum steam export to the Iggesund mill.

The remaining excess heat can be recovered for generating extra steam that can be expanded in a new steam turbine as shown in *Figure 3.22*. The following steam data was assumed: 100 bar and 500°C at the turbine inlet; extraction at 5.5 bar, back-pressure at 1.2 bar.

The power production potential amounts to around 14 MW which reduces the total import of electricity to the biomass based methanol production process from 44 to 30 MW. Note that the envisioned steam turbine system is used for combined heat and power production since the extraction and back-pressure steam flows are used for heating purposes such as for distillation and biomass drying.

3.2.4.4 Heat recovery steam cycle for maximum power generation

An alternative option for using the excess heat from the biomass based methanol production process is to expand the steam in a new turbine set to produce extra power as shown in *Figure 3.23*.

For estimating the power production potential, the following steam data was assumed: 100 bar and 500°C at the turbine inlet; condenser pressure of 0.1 bar, steam extraction at 5.5 bar and 0.6 bar.

The potential for power generation amounts to around 52.5 MW which is higher than the power requirements for compression and auxiliaries at the biomass based methanol production plant (44.2 MW), thus leaving around 8.4 MW for export to the power grid.

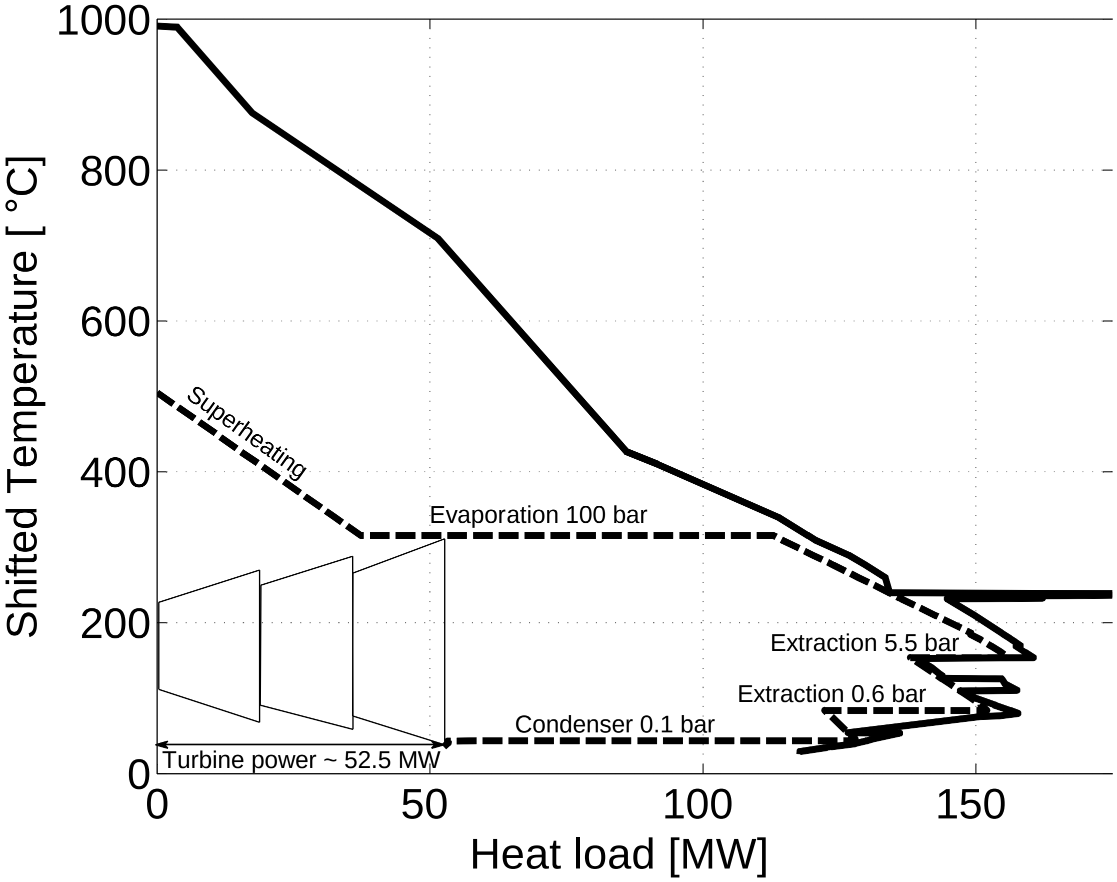

Figure 3.23: Integration of a condensing extraction steam turbine for maximum power production.

3.2.4.5 Summary of the process integration opportunities and system consequences

An overview of the energy balances of the different process integration options is given in *Table 3.15*.

Under the condition that the imported electricity is generated by a fuel-based production technology, it is possible to give a more general picture of the performance of the proposed biomass to methanol system by introducing the "total conversion efficiency" of biomass into methanol.

As shown in the methodology chapter (Equation 2.2), this is similar to the conversion efficiency from biomass to methanol (i.e. the ratio between the energy rates of produced methanol and biomass input) with the only difference that the total input energy is accounted for. This is done by deducing from the total energy rate of biomass input the fuel saved in boilers at the mill due to the steam export from the biomass based methanol process and by adding the fuel spent in marginal electricity production technology to compensate for the electricity imported by the biomass based methanol process (or deducing this fuel in case of export of electricity).

Table 3.15: Overview of steam export and electricity import of the second system alternative under different conditions of process integration with the Iggesund pulp and paper mill.

Configuration	***Maximum steam export to mill***	***Maximum power generation***
Steam export to mill (t·h $^{-1}$)	*130*	-
Fuel saving at the mill (MW) … of which oil (EO3) … of which bark	115 40 75	-
Power generation (MW)	*14*	*52.5*
Steam turbine configuration	-	Condensing / extraction
Electricity import (MW)	*30.2*	*-8.4 (export!)*

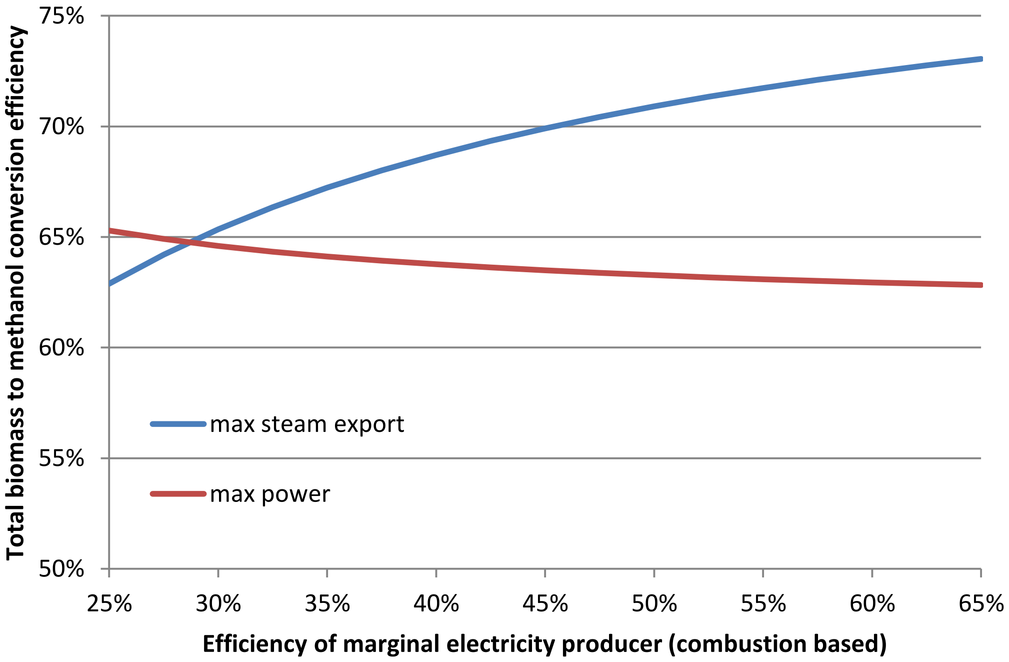

Figure 3.24: Effect of electrical efficiency of combustion based marginal electricity producer on the total biomass to methanol conversion efficiency for the second system configuration and for different process integration options.

Accordingly, the total conversion efficiency depends on the efficiency of the boilers currently used for steam production and on the efficiency of marginal electricity producers.

The fuel savings in existing boilers is around 115 MW_{LHV} [18] of which 75 MW are bark and 40 MW are oil (EO3 type) in the case where all the theoretical excess heat from the biomass based methanol production is recovered to produce steam for export to the mill.

It is interesting at this stage to show the influence of efficiency of marginal electricity producer on the total conversion efficiency of biomass to methanol. Results are shown in *Figure 3.24*, while a breakdown of the main energy flows is given in *Table 3.16* for two levels of efficiency for the marginal electric power generation technology.

Table 3.16: Overview of performances of the biomass to methanol system based on oxygen steam blown pressurized gasifier technology, for different process integration options with the Iggesund pulp and paper mill.

Configuration	***Maximum steam export to mill***	***Maximum power generation***
Biomass input energy rate (MW) HHV basis	548.7	548.7
Methanol output energy rate (MW) HHV basis	336.8	336.8
Biomass to methanol conversion	*61.3%*	*61.3%*
Fuel saving in mill boiler (MW) HHV basis, 87% eff.	134.3	-
Net electricity import (MW)	30.3	-8.2 (export!)
Marginal fuel to electricity (MW)		
... 40 % electrical efficiency	75.8	-20.5
... 60 % electrical efficiency	50.5	-13.7
Total conversion efficiency (HHV basis)		
... 40 % marg. el. eff.	*68.7%*	*63.8%*
... 60 % marg. e. eff.	*72.4%*	*62.9%*

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[18] For simplicity bark heating value is assumed equal to wood chip heating value: HHV 10.43 MJ·kg^{-1}, LHV 8.55 MJ·kg^{-1}. Oil higher heating value 45.7 MJ·kg^{-1}, lower heating value 42.7 MJ·kg^{-1}. Fuel savings at the mill HHV basis: 91.5 MW bark, 42.8 MW oil, 134.3 MW total.

It can be concluded that process integration options and in particular the export of steam to the cluster plants allows a large increase in the overall conversion efficiency. Steam export to the mill offers greater advantages: the whole nominal steam production capacity of the extra boiler at the mill can be replaced by steam export; the bark currently burned for steam production can be used for methanol production; the total fuel savings are larger than the corresponding fuel savings in marginal electricity producer in case all the excess heat is used for local electricity production.

In particular, when electricity is produced in marginal electricity producers with efficiency lower than 30%, it is more suitable to locally produce part or the totality of this power to reduce the penalty of larger fuel intake in marginal electricity generation plants. It should be mentioned that for instance state-of-the-art coal power plants are able to exceed such efficiency level. If natural gas combined cycles are assumed as marginal electricity producers, efficiencies in the range between 55 to 60 % should be considered. If, in addition to that, the costs for an additional steam turbine set is taken into account for the case power where is to be generated with the steam from recovered excess heat, the option of exporting the totality of the steam to the nearby pulp and paper mill to substitute the bark and oil usage is the most interesting process integration option.

3.3 System configuration nr 3

3.3.1 Description

The third system configuration for methanol production is based on an oxygen blown entrained flow gasifier and subsequent methanol synthesis located in proximity of the Stenungsund petrochemical cluster. The gasifier is fed with torrefied biomass which is provided by one nearby torrefaction plant and two decentralized torrefaction plants located in the centre and north of Sweden. A representation of the investigated system is shown in *Figure 3.25*.

Similarly to the previous alternatives, the biomass derived methanol is then mixed with other imported methanol and processed through a MTO plant in order to achieve a total olefin production of 200 kt·y^{-1} olefins.

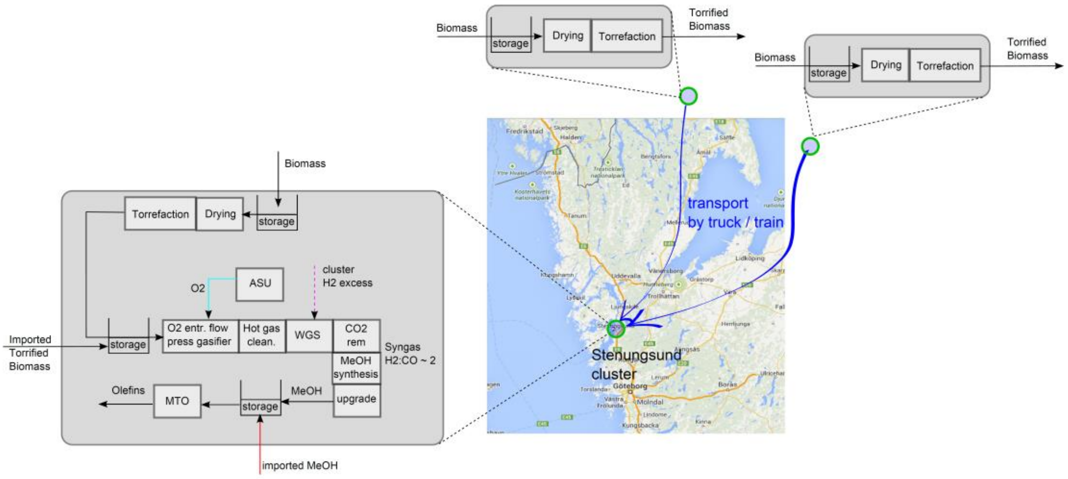

Figure 3.25: System configuration nr 3: biomass gasification and methanol synthesis based on decentralized biomass torrefaction and oxygen blown pressurized entrained flow gasifier.

3.3.2 Torrefaction

Entrained flow gasification is a well-established technology for processing coal due to the considerably cleaner product gas obtained compared to fluidized bed gasifiers. For the gasification reaction to occur in entrained flow conditions, a very small particle size is required which is obtained through fine grinding of the solid feedstock unless liquid or gaseous feedstock is used. This is rather simple for coal but more complicated when it comes to fibrous feedstock such as biomass, although this has also been demonstrated (Weiland, Hedman et al. 2013). The solid feedstock can also be mixed with water or other liquids to form a slurry which simplifies the feeding of the fuel into the gasification reactor especially when gasification is pressurized. This is however not a mandatory procedure, injection of pulverized material being also a well-established technology in coal boilers for examples.

There are essentially two different pre-treatment options prior to the final entrained flow gasifier, which are often considered when biomass or waste-derived fuel are used:

- Gasification or pyrolysis can be used to convert the feedstock into raw gas or liquid which is subsequently fed into the entrained flow gasifier, possibly in combination with a side-injection of char also leaving the first reactor.
- Biomass can be torrefied thus leaving a coal-like material (torrefied biomass) which can be ground and even pelletized in a much easier way than raw biomass. This option is considered here.

Torrefaction is a thermal decomposition of the biomass that occurs between 200 and 350°C and with a few minutes up to a half hour of residence time. The typical products are solid biomass with higher carbon content and very low to no moisture, and a gas containing water, acids, alcohols and phenols which are the primary products of decomposition of the biomass building blocks, referred to as torrefaction gas.

The kinetics and experimental data of wood torrefaction are discussed in detail in Ref. (Prins 2005).

Three torrefaction plants are considered for the third system configuration discussed here, each with an input of 150 MW_{LHV} biomass. Two of these plants are located somewhere in the centre and north of Sweden while a third plant with the same capacity is located in proximity of the gasification process (Stenungsund).

Although the benefits of biomass torrefaction are somewhat related to its coupling with the entrained flow gasification technology and therefore with the advantage of a tar-free product gas, the high gasification temperature also causes a large portion of the biomass substrate to be combusted, and consequently a large oxygen demand, in order to provide sufficient heat to sustain the gasification reactions. Thus, this configuration is not expected to show significant advantages in terms of methanol yields compared to the other configurations investigated in this study, but only minor investment savings due to simplified gas cleaning. Conversely, the logistics of feedstock transportation and storage are significantly improved due to the large weight reduction obtained by drying and torrefaction in decentralized plants compared to the case in which raw biomass, at best with 40 to 50 % moisture, is transported. This is accomplished by burning the torrefaction gas and a part of the torrefied product for heat production. This in turns eliminates the opportunity to thermally integrate the biomass pre-treatment, which is rather energy intensive, with the gasification and methanol synthesis processes, thereby further reducing the overall efficiency of the biomass to methanol conversion. In the system configuration investigated here, one third of the total raw biomass input (150 MW_{LHV} of the total 450 MW_{LHV}) is dried and torrefied in the proximity of the gasification plant in order to improve the logistics of biomass harvesting and transportation in the south of Sweden and partially mitigate the reduced heat integration opportunities of biomass pre-treatment.

The design of the torrefaction plants for the relevant biomass feedstock and adjustment to the appropriate plant scale was carried out at the Dept. of Applied Physics and Electronics, Umeå University and is discussed in a parallel project report.

Due to the heat demand of drying and torrefaction, the net yield of dry solids for the stand-alone plants is lower than for the Stenungsund plant where heat for drying and torrefaction is provided by steam or hot gases produced by downstream processes. This however does not significantly alter the characteristics of the torrefied biomass as the torrefaction step itself is not changed (yield is kept constant at 76.5% of input dry solids) and the only difference is that part of the torrefied product and the torrefaction gas are used for heating in the stand-alone plant.

The main data concerning the overall yields and composition of torrefaction products are summarised in the following for wood chips.

TORREFIED WOOD CHIPS

- Moisture: 1.8 %
- Ultimate analysis (%): ash 1.4, C 56.7, H 5.8, N 0.3, Cl 0, S 0.02, O 35.78
- HHV ($MJ \cdot kg^{-1}$): 22.48
- LHV ($MJ \cdot kg^{-1}$): 21.63

STAND-ALONE BIOMASS TORREFACTION

Biomass input
- energy rate: 150 MW_{LHV} (183 MW_{HHV})
- mass flow rate: 63 648 $kg \cdot h^{-1}$
- of which dry solids: 31 824 $kg \cdot h^{-1}$

Torrefied biomass output
- mass flow rate: 23 756 $kg \cdot h^{-1}$
- of which dry solids: 23 330 $kg \cdot h^{-1}$ (72.2% of input dry solids)
- energy rate: 140 MW_{LHV} (149 MW_{HHV})

Conversion: 0.93 LHV basis, 0.81 HHV basis

STENUNGSUND BIOMASS TORREFACTION

Biomass input
- energy rate: 150 MW_{LHV} (183 MW_{HHV})
- mass flow rate: 63 648 $kg \cdot h^{-1}$
- of which dry solids: 31 824 $kg \cdot h^{-1}$

Torrefied biomass output
- mass flow rate: 24 791 $kg \cdot h^{-1}$
- of which dry solids: 24 345 $kg \cdot h^{-1}$ (76.5% of input dry solids)
- energy rate: 145 MW_{LHV}, 155 MW_{HHV}

Conversion: 0.96 LHV basis, 0.85 HHV basis

Torrefaction gas output
- mass flow rate: 10 568 $kg \cdot h^{-1}$

- of which[19]:
    - acetic acid 684 $kg\cdot h^{-1}$
    - lactic acid 977 $kg\cdot h^{-1}$
    - furfural 977 $kg\cdot h^{-1}$
    - formic acid 977 $kg\cdot h^{-1}$
    - hydroxy acetone 684 $kg\cdot h^{-1}$
    - methanol 977 $kg\cdot h^{-1}$
    - CO 104 $kg\cdot h^{-1}$
    - CO_2 521 $kg\cdot h^{-1}$
    - water 4 667 $kg\cdot h^{-1}$

- HHV ($MJ\cdot kg^{-1}$): 8.82
- LHV ($MJ\cdot kg^{-1}$): 6.95
- energy rate: 25.89 MW_{HHV} (20.40 MW_{LHV})

Overall the biomass intake of this third system alternative is 450 MW_{LHV} (550 MW_{HHV}) which corresponds to around 190 942 $kg\cdot h^{-1}$ (around[20] 1 530 $kt\cdot y^{-1}$).

The total torrefied material at the gasifier input corresponds to 72 307 $kg\cdot h^{-1}$. In addition 10 568 $kg\cdot h^{-1}$ of torrefaction gas are also available at the Stenungsund site.

3.3.3 Entrained flow gasification and methanol synthesis

At the Stenungsund site, entrained flow gasifiers are used to gasify 72 307 $kg\cdot h^{-1}$ torrefied biomass. The total input chemical energy rate is around 452 MW_{HHV} (426 MW_{LHV}).

A picture of the process is shown in *Figure 3.26* where the main process operating parameters are also given.

Compared to the previous two system configurations, the process layout is essentially unchanged from the water gas shift reactor until methanol synthesis and upgrading. With respect to the first and second configurations where different concepts of fluidized bed gasifier were considered, an oxygen blown entrained flow gasifier is used here which operates at considerably higher temperatures, above the ash-melting point. This promotes complete conversion of the biomass substrate into light gases and a tar free product gas is obtained. For this reason, thermodynamic equilibrium can be reached with the only exception for methane (around 1% mol. in dry gas) (Weiland, Hedman et al. 2013). Still, the methane content is considerably smaller than in the case of a fluidized bed gasifier, and its reforming into lighter gases can be avoided without reducing significantly the methanol production and undesired build-up of methane in

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[19] These yields were calculated based on mass balance and data regression according to Ref. Prins, J. M. (2005). *Thermodynamic analysis of biomass gasification and torrefaction*. Ph.D. thesis, Eindhoven Technical University.

[20] Considering 8000 hours of operation per year.

the synthesis loop. Gas cleaning and upgrading is therefore simplified and gas reheating before a reformer is also avoided.

A temperature of 1300°C and a pressure of 30 atm were chosen here for gasification of torrefied biomass.

Note that gas leaving the torrefaction reactor could theoretically be converted into lighter gases and used to boost the syngas product. Due to pressurized gasification conditions, however, the gas must be compressed. This could be done in practice by cooling the gas from the torrefaction temperature (around 275°C) and separating water and condensable gases from lighter gases such as CO. Then the liquid phase could be pumped while the gas phase could be compressed, both being injected at a slight overpressure than the gasifier (30 atm). In fact, this appears a rather complicated process arrangement which does not seem to be counterbalanced by the relatively low chemical energy rate of the torrefaction gas (around 26 MW_{HHV}) compared to the main torrefied material (452 MW_{HHV}). Accordingly, the torrefaction gas is sent to combustion together with the purge gas from the methanol synthesis reactor.

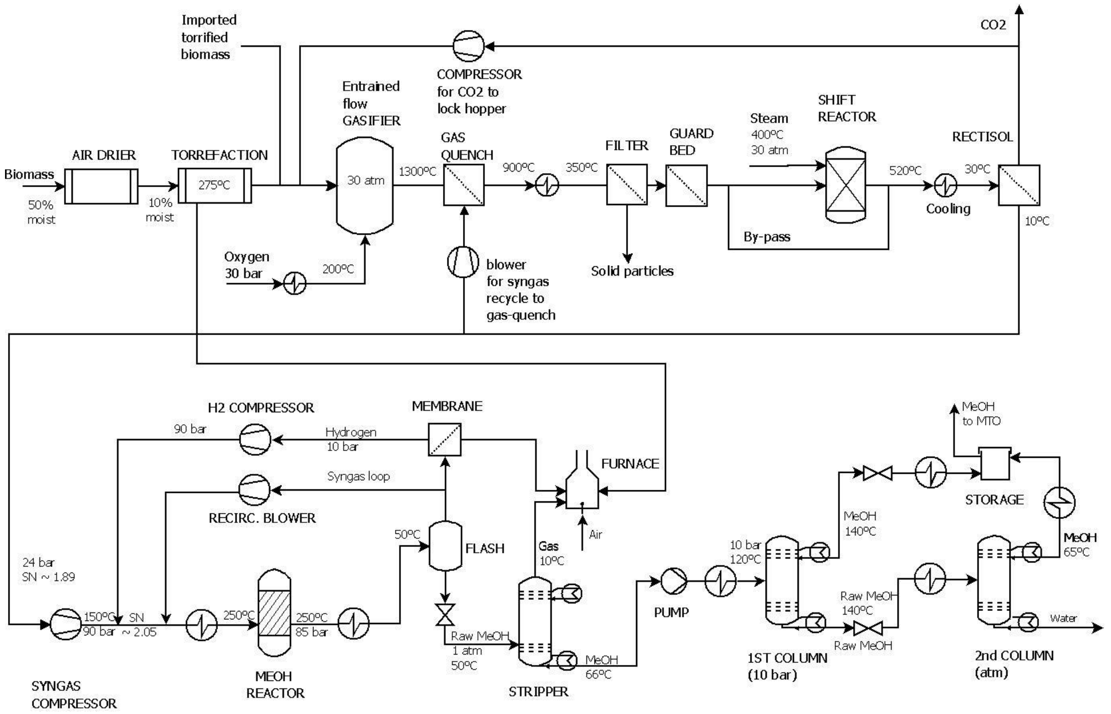

Figure 3.26: Layout of the third system configuration for biomass gasification and methanol production.

Due to pressurized conditions, the torrefied biomass powder must be injected through a special lock-hopper system working with an inert gas (Van Der Drift, Boerrigter et al. 2004). CO_2 is well suited for this purpose since it is anyway removed prior to methanol synthesis, and plenty is available. This gas should be around 1 m^3 per tonne of fuel (Van Der Drift, Boerrigter et al. 2004).

The biomass is converted into around 114 593 $kg \cdot h^{-1}$ of gas with the composition reported in *Table 3.17*. The higher heating value (HHV) of this gas is 11.446 $MJ \cdot kg^{-1}$, the lower heating value (LHV) is 10.626 $MJ \cdot kg^{-1}$.

The gasifier cold gas efficiency is therefore:

$$\eta_{CG}^{HHV} = \frac{\dot{m}_{gas} \cdot HHV_{gas}}{\dot{m}_{biomass} \cdot HHV_{biomass}} = 80.6\%; \quad \eta_{CG}^{LHV} = \frac{\dot{m}_{gas} \cdot LHV_{gas}}{\dot{m}_{biomass} \cdot LHV_{biomass}} = 79.4\%.$$

Table 3.17: Composition of the product gas obtained from entrained flow gasification of torrefied wood chips at 1300°C and 30 atm.

Component	***Mol. Frac. (%)***
H_2O	9.63
H_2	27.21
N_2	1.34
H_2S	0.01
CH_4	0.90
CO	54.24
CO_2	6.66

Gasification is followed firstly by a gas quench to cool the product gas to a temperature below 900°C since heat exchange for steam production would be subject to severe fouling at higher temperatures due to hot particles and the reducing behaviour of the product gas. Cooling is done by injecting a portion of the syngas after CO_2 removal. After the quench, gas cooling is done by a heat exchanger. The raw gas leaves the heat exchanger at an appropriate temperature (here 350°C) for solid particles removals which can be achieved for instance by means of candle filters.

Due to the high temperature of gasification, the product gas is made almost entirely of hydrogen and CO. The H_2:CO ratio of the product gas is however around 0.5 which is considerably below the ratio required for methanol synthesis and a water gas shift reaction is therefore needed for adjusting the gas composition to appropriate levels (H_2:CO ratio should be around 2). Although the concentration of CO_2 in the gasifier product gas is quite low, large amounts of CO_2 are produced by shift reaction, and a Rectisol section is then required to remove CO_2 before methanol synthesis. A portion of the CO_2 is also redirected to the gasifier fuel feeding system, while a portion of the syngas is redirected to the gas quench.

A net flow of 53 515 $kg \cdot h^{-1}$ of syngas[21] is then used for methanol synthesis. A membrane system is used to recover part of the hydrogen leaving the synthesis step in the purge. This allows conditioning the syngas to lower gas synthesis ratio since the recovered hydrogen is used to adjust this ratio to the optimal value considered for

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[21] Syngas heating value: 24.128 $MJ \cdot kg^{-1}$ HHV basis, 21.559 $MJ \cdot kg^{-1}$ LHV basis.

methanol synthesis. As the remaining part of the process is similar to the second system configuration, the reader is referred to the previous sections for a thorough description.

Process details are summarized in the Appendix.

A methanol production of around 45 949 kg·h^{-1} is obtained which corresponds to a chemical energy rate of 289.1 MW$_{HHV}$ (253.7 MW$_{LHV}$).

The conversion efficiency from torrefied biomass to methanol therefore corresponds to around 64.0% HHV basis (59.6% LHV basis).

Considering the energy conversion of the upstream torrefaction step (82.3% HHV basis, 94.0% LHV basis), the overall conversion efficiency of this third system configuration is 52.6% HHV basis (56.0% LHV basis).

3.3.4 System efficiency and opportunities for process integration with a nearby chemical cluster

3.3.4.1 Overview of heat and power requirements of the biomass based process

Heat recovery opportunities from the biomass to methanol process were estimated based on the results of process simulation and by including temperatures and heat loads of the various cooling and heating process steps in a single GCC curve according to Pinch Analysis. This curve is shown in *Figure 3.27* and represents the net excess or deficit of heat at various temperature levels.

By looking at the process GCC curve in *Figure 3.27*, it is possible to see that heat from the methanol synthesis is fully used for generation of steam for the water-gas-shift reactors. Such a large steam demand is a consequence of the poor H_2:CO ratio of the product gas which must be adjusted through the water-gas-shift to optimal levels for methanol synthesis.

Abundant heat is available instead from product gas and syngas cooling, which can be used for high pressure steam production. In addition, the purge gas boiler delivers a considerable amount of heat since torrefaction gas is combusted together with the purge gas from methanol synthesis. The radiative portion of this combustion heat is represented as a heat source at a constant temperature of 1000°C for simplicity in order to reduce the temperature scale of the diagram (this is the convention used in all the other diagrams presented in this report).

Note that heat is required for methanol distillation and biomass drying. Furthermore, no other large heat demand appears at lower temperatures and around 100 MW of the heat is available in the temperature range 1000°C to 400°C. This opens the opportunity for including a condensing steam turbine if the objective is to maximize electric power production or, alternatively, for steam export to a nearby industrial process plant.

Entrained flow gasifiers operate with oxygen under pressurized conditions. A large amount of oxygen is required due to the high gasification temperatures.

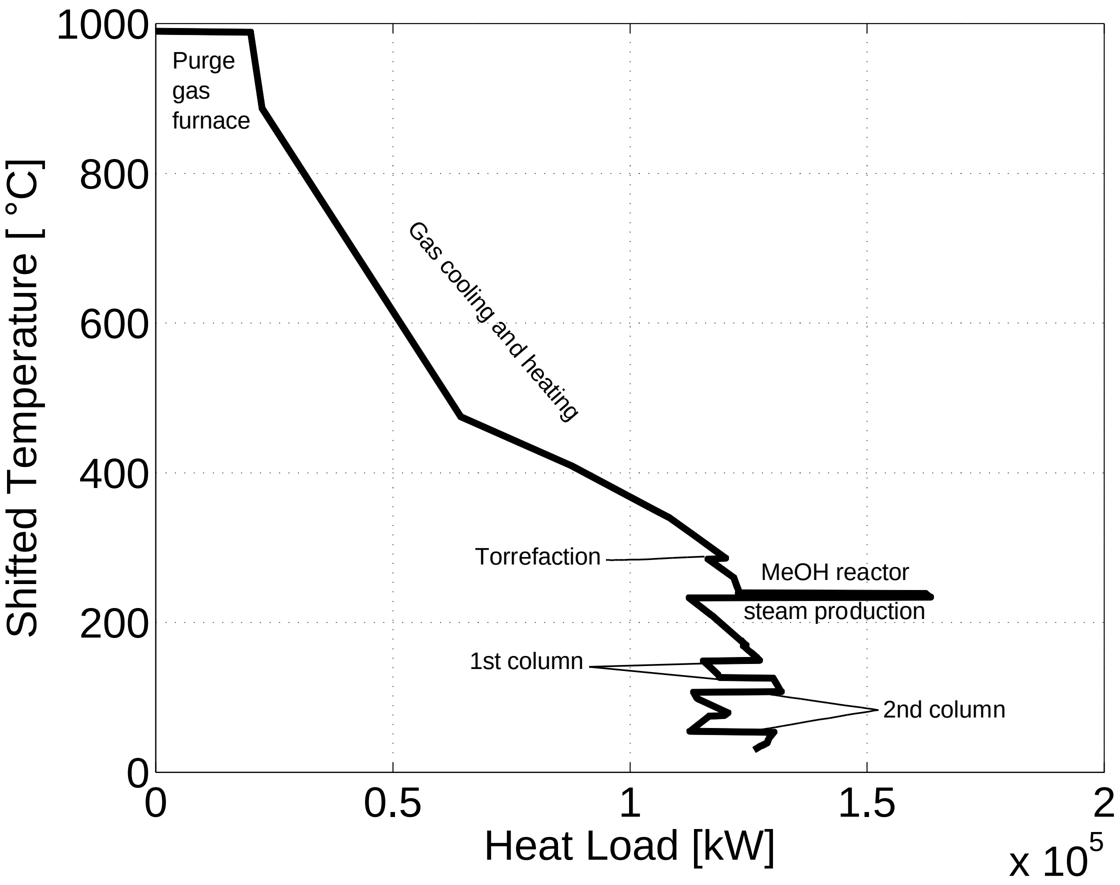

Figure 3.27: GCC curve of the methanol production process based gasification of torrefied biomass in an entrained flow gasifier.

Oxygen is delivered by an air separation unit (ASU). Additional work is required to further compress the oxygen up to the gasification pressure of around 30 atm. The total specific work consumption was estimated to be around 0.39 kWh·kg^{-1} oxygen. The total oxygen demand in this third system configuration amounts to 41 903 kg·h^{-1}. This results in a total power demand for the ASU and further compression of around 16.4 MW.

The gasifier fuel feeding system consists of a lock-hopper which also needs some inert gas, here CO_2. Electricity consumptions ranging between 0.01 and 0.02 MW per MW fuel have been reported for such type of feeding system (Bergman, Prins et al. 2005). A compressor is required to compress the CO_2 at the outlet of the Rectisol (here assumed to operate at around 1 atm) up to the gasifier pressure (a slight overpressure is considered). In addition, grinding of the torrefied wood pellets is necessary prior to injection into the gasifier. An electricity consumption equal to 10 kW of electricity per MW of fuel (LHV basis) is assumed, which results in a total electricity need of 4.3 MW for grinding.

Power is also required for compressing the syngas up to the methanol synthesis pressure (90 bar), by the syngas recirculation blower for the gas-quench, by the gas blower in the methanol synthesis loop and by the hydrogen recycle blower. In addition, power is needed to drive the compression refrigeration machines in the Rectisol process which uses methanol as solvent. Other ancillaries such as pumps and small compressors are also power consumers.

A breakdown of the process power requirement is given in *Table 3.18*. Overall, a total power consumption of around 40.8 MW is estimated.

Table 3.18: Power requirements of the methanol production process based on oxygen steam blown biomass gasification.

Equipment	***Power (kW)***
Fuel feeding (lock-hopper + CO_2 compress.)	6 913
Syngas rec. blower to gas-quench	214
Water pump to water gas shift	136
Main gas compressor before synthesis	6 103
Synthesis loop recirc. blower	870
H_2 recirculation compressor	726
Rectisol refrigeration	3 598
ASU	11 033
O_2 compressor	5 351
Torrefaction	2 000
Grinding	4250
Total	*40 841*

3.3.4.2 Overview of the steam demand of the chemical cluster

The Stenungsund chemical cluster consists of six major plants producing several chemical products mainly starting from olefins (ethylene and propylene) currently produced by cracking of hydrocarbons (naphtha, ethane, butane, propane) at Borealis.

Although the chemical plants are highly integrated from a material point of view, the utility systems of the six plants are almost completely independent. Heat recovery and heat distribution are achieved primarily through steam networks where steam is produced in boilers and distributed at different pressure levels for process heating in each plant. A more advanced steam network exists at Borealis cracker where superheated steam is produced mainly by recovering the heat from cracker gases and expanded through turbines for power production (both for direct driving of compressors and for electricity generation). Process heating is achieved with medium and low pressure steam available at the turbine outlets.

According to data exchanged through internal communications, the steam generated by recovering heat from the cracker product gas is not sufficient to cover the steam need at the Borealis plant and an additional 75 $t\cdot h^{-1}$ of high pressure steam must be produce in boilers fired with natural gas.

Since the biomass to methanol processes located in Stenungsund are considered in this project for the subsequent production of olefins through the MTO technology and these olefins are intended to partially substitute the current fossil based production, it is necessary to take into account possible effects that the partial substitution of the current olefin production has on the steam balance at the cracker. This type of analysis was the object of a parallel investigation within the Skogskemi project (Joelsson, Engström et al. 2015). As a result of implementing an MTO plant for a total olefin production of around 200 $kt\cdot y^{-1}$, it was estimated that the steam deficit should increase from 75 $t\cdot h^{-1}$ to around 140 $t\cdot h^{-1}$ when the MTO plant is in operation[22].

Table 3.19: Overview of the cluster steam requirement for process heating purposes, adapted from (Hackl, Anderson et al. 2010). (: estimated considering MTO for around 200 kt·y $^{-1}$ olefins, high propylene case, based on internal communication with Borealis)*

Plant	***Pressure (barg)***	***Steam quality***	***Mass flow (t/h)***	***Corresponding fuel in boilers (MW) – 80% thermal eff.***
Borealis	85	485°C	140(*)	144.4
Akzo	40	Saturated	17.52	14.8
Akzo+INEOS	28	Saturated	12.58	10.6
Akzo+INEOS	20	Saturated	13.03	11.0
INEOS	10	Saturated	7.92	6.6
Akzo+INEOS	6	Saturated	24.75	20.5
Perstorp	2	Saturated	45.43	37.1
Akzo+INEOS	1	Saturated	12.76	10.4
Total Cluster	-			255.4

3.3.4.3 Maximum steam export to the chemical cluster

The excess heat from the biomass based processes shown in *Figure 3.27*, is about half of the corresponding steam generation heat demand at the cluster shown in

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[22] Erika Johansson, Lars Pettersson, Borealis internal communication, March 2014.

Table 3.19. One possible option for process integration is to export as much steam as possible to the chemical plants.

Assuming that all the cluster boilers have more or less the same efficiency and are all fuelled with natural gas[23], export of steam from the biomass based processes to the cluster has equal effect (i.e. substitution of steam boilers) independently of the pressure levels. The pressures levels at which the steam should be exported must therefore be chosen in order to maximize the amount of steam export only. This depends on the availability of heat from the biomass based processes at different temperatures.

Two relevant scenarios of steam export were considered here based on the following criteria:

- Prioritize export of steam to lower pressure levels first, *Figure 3.28*.
- Prioritize the export of steam to Borealis cracker (85 barg, 485°C), *Figure 3.29*.

In the first case (*Figure 3.28*), all the steam demand from 40 barg to 1 barg can be covered, that is all the current steam demand in cluster plants other than Borealis cracker[24]. In addition, around 29 $t{\cdot}h^{-1}$ of high pressure superheated steam can be exported to Borealis cracker, so that around 111 $t{\cdot}h^{-1}$ must still be provided by the boilers at the cluster plant.

In the second case (*Figure 3.29*), the excess heat from the biomass based processes is almost sufficient to cover all the high pressure steam demand of cracker plant. Around 136 $t{\cdot}h^{-1}$ of high pressure steam can be exported so that around 4 $t{\cdot}h^{-1}$ of steam must still be provided by boilers at the cluster plant.

In both cases, the amount of steam exported to the cluster is sufficient to substitute around 140 MW of natural gas in existing steam boilers which therefore shows that the two options are equally interesting from a primary energy saving point of view.

However, when looking at the possible implementation of the two options, there are different aspects that suggest that the export of high pressure steam to Borealis cluster is of greater interest. First of all, the biomass based processes are meant for the production of methanol which should be used for production of olefins via the MTO technology. Integrating the steam network of the biomass based processes with the cracker steam network appears a natural choice since the methanol production and steam delivery are simultaneous and so is the methanol and steam demand at the cracker plant. In addition, the amount of high pressure steam is sufficient to almost cover the estimated net steam demand at the cracker when the MTO technology is in operation. The current boilers at Borealis cracker are capable of delivering around 75 $t{\cdot}h^{-1}$ of steam and can be used only in case of disruption of the biomass to olefin processes. This result also shows that an MTO capacity larger than 200 $kt{\cdot}y^{-1}$ can be installed

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[23] Some fuel gas is also obtained as by-product of the hydrocarbon processing at the cluster. Currently this fuel gas is fired in boilers so the natural gas imported is only a portion of the actual fuel used for process heating. Fuel gas can be used for other purposes than combustion (e.g. hydrogen production) so that all the fuel used in boiler can be considered as imported natural gas.

[24] Steam should probably be generated at one medium and one low pressure levels and let down valves should be used to match the demand pressure levels.

without severely compromising the steam balance at the plant if the same amount of methanol is produced by means of biomass gasification.

In the case where steam is exported from the biomass based processes at lower pressure levels to cover the steam demand at the other cluster plants (*Figure 3.28*), the steam network must extend to a much larger area, which requires a significant investment and no increased substitution of natural gas boilers with respect to the case where steam is only delivered to Borealis cracker. In addition, the high pressure steam that must be generated at the cracker plant to balance the steam demand (111 t·h^{-1}) would exceed the current boiler capacity at Borealis cracker.

It should be noted that recent studies indicate that an integrated hot water network between cluster plants could largely reduce the current demand of low pressure steam at the cluster. Delivering low pressure steam from a biomass based process is not desirable if such energy efficiency measures are implemented.

A third configuration of steam export can also be considered. A quantity of high pressure steam sufficient to cover the steam deficit while still using the installed boilers (140 - 75 = 65 t·h^{-1}) could be exported to the Borealis plant and the remaining excess heat from the biomass based processes could be exported in form of medium pressure steam. This option appears very similar to the case in which the export to the cracker is maximized and is therefore not discussed in detail.

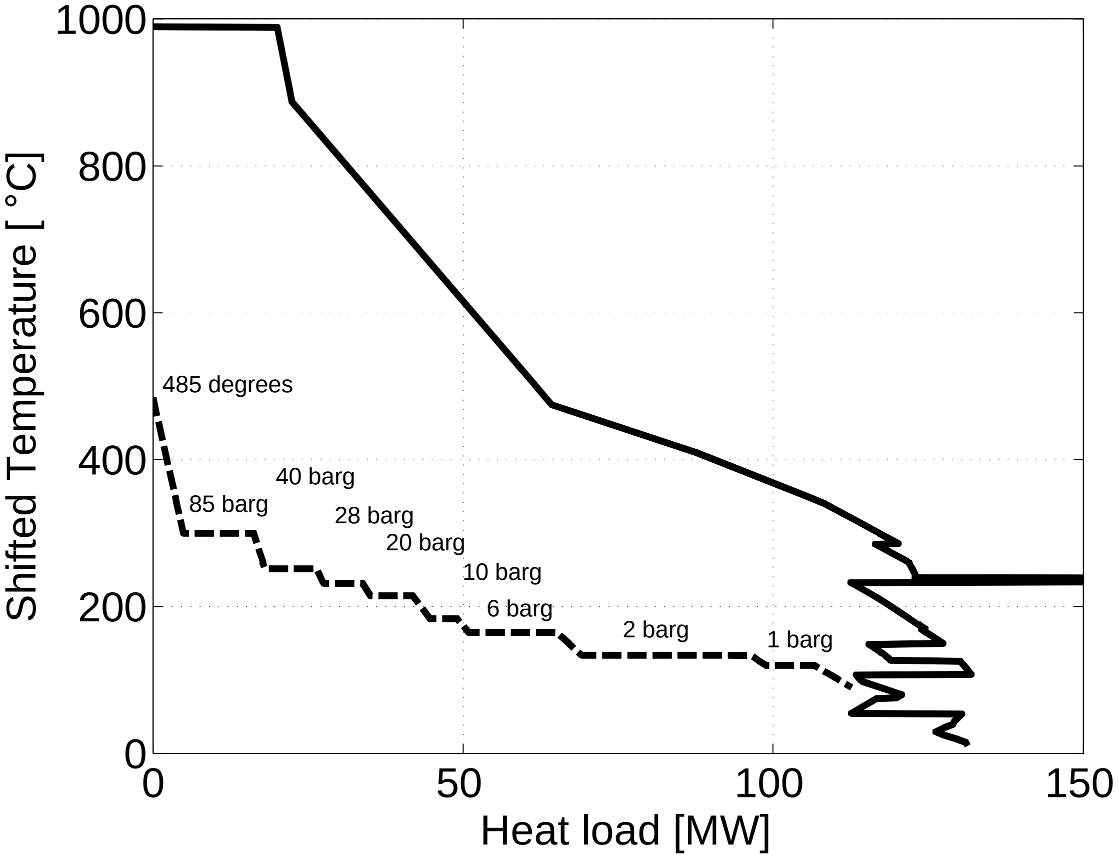

Figure 3.28: Maximum export of low pressure steam to all cluster plants.

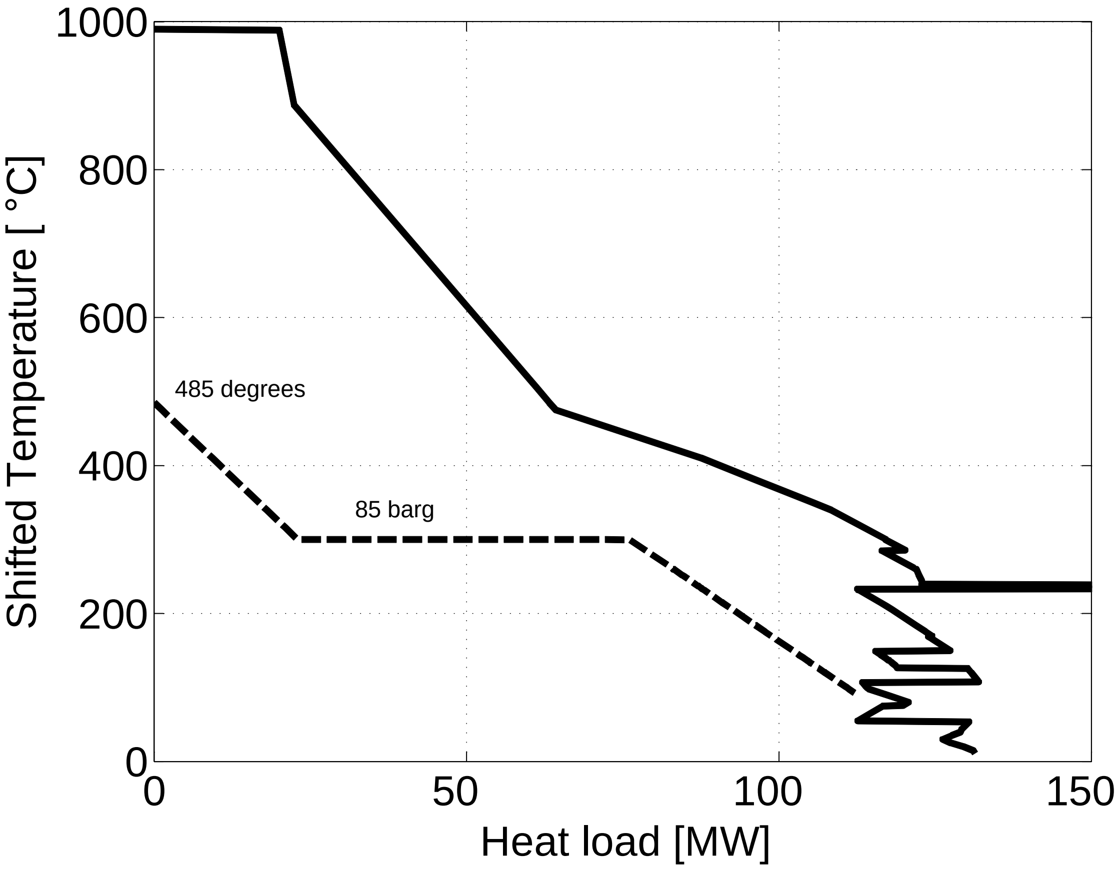

Figure 3.29: Maximum export of high pressure superheated steam to Borealis cracker.

3.3.4.4 Heat recovery steam cycle for maximum power generation

The excess heat from the biomass based processes could be used to generate superheated steam at very high pressure which can then be used to drive steam turbines and generate power. Power can be made available in form of shaft power to drive process compressors or sold to the electricity network.

The biomass based processes actually require some heat at medium and low temperature levels. Even if the thermal cascade in *Figure 3.27* shows that this heat can be cascaded from high temperature sources, it can be convenient to make this heat available in form of steam for two main purposes: (a) steam is the most consolidated way to recover and distribute heat in large plants, (b) the high temperature heat can be recovered in form of high pressure steam while low pressure steam can be used for heating purposes at lower temperatures and this steam can be expanded in turbines. Accordingly a combined heat and power system is considered here which is based on a condensing steam turbine with few steam extractions to deliver some heat to the biomass based processes.

For estimating the potential power generation with such steam turbine system, turbine inlet steam data of 100 bar, 500°C were considered. A limit of 0.1 bar pressure was considered for the steam condenser. A set of optimal pressure levels for steam extraction were considered: 33 bar (mainly for water gas shift), 5.5 bar and 1.5 bar (mainly distillation columns).

As shown in *Figure 3.30*, the maximum power generation is around 44 MW, corresponding to a net production of around 8 MW after the power needed for the biomass based processes is covered (gas compression, ASU, Rectisol, etc.).

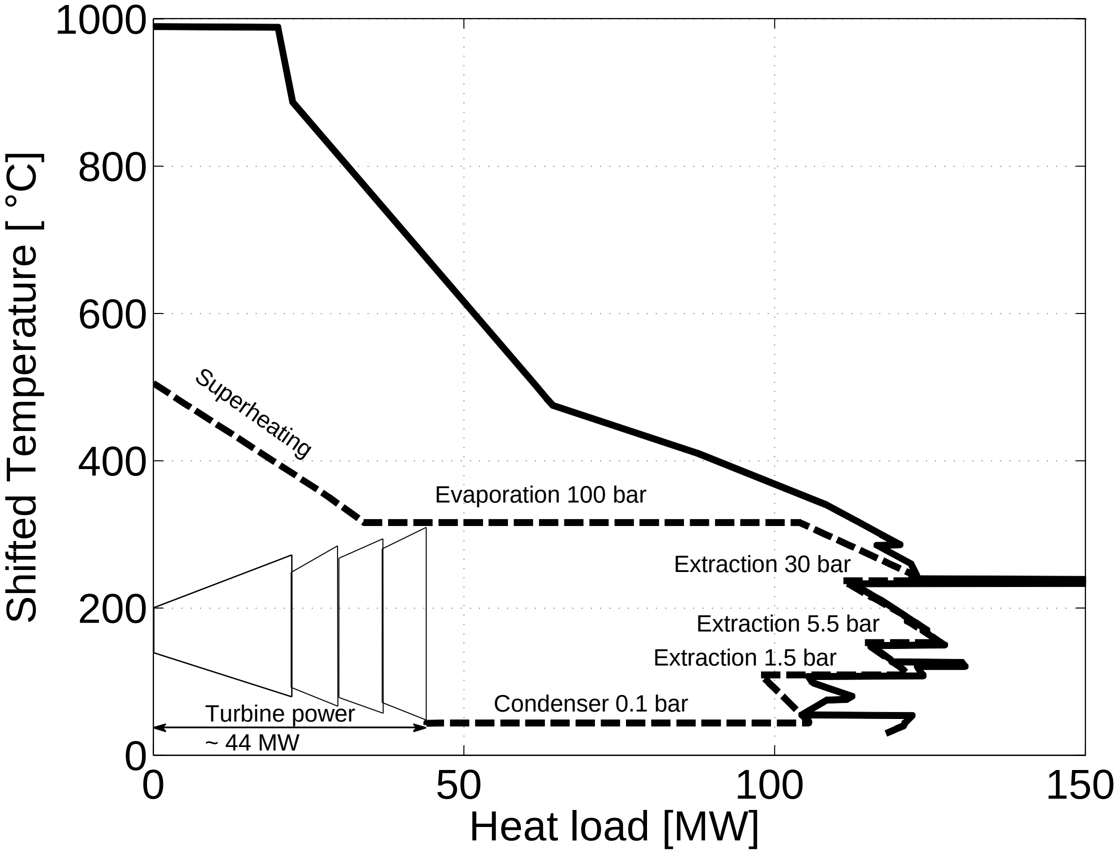

Figure 3.30: Integration of a condensing extraction steam turbine for maximum power production.

3.3.4.5 Combined heat and power system

As previously discussed, the maximum steam export to the cluster can introduce a potential saving of around 140 MW of natural gas. Among the different pressure levels at which steam can be exported, the export of high pressure superheated steam to Borealis cracker appears to be the simplest solution with equally high natural gas substitution effects (*Figure 3.29*). However, it should be noticed that production of power has also great substitution effects in marginal electricity producers. The production of electricity by means of a condensing turbine is a poor solution (*Figure 3.30*) due to low efficiency compared to state-of-the-art power plants. A way to maximize the efficiency of excess heat recovery is to generate high pressure superheated steam to be expanded in a counter-pressure steam turbine and using the steam at the turbine outlet for process heating.

The cluster steam demands at low pressure levels can be covered by steam extractions and by back-pressure steam. This allows shifting from a current steam network based on fossil fuel boilers for process heating only to a CHP system based on excess heat from biomass thermo-chemical conversion processes.

The suggested system is shown in *Figure 3.31*, where same maximum steam values as in previous sections were considered: 100 bar and 500°C. With respect to *Figure 3.30*, a different set of steam extraction was considered in order to maximize the low pressure steam export to cluster plants while still allowing for power generation by steam expansion: 43 bar, 33 bar, 8 bar, 3.5 bar and back-pressure at 1.5 bar.

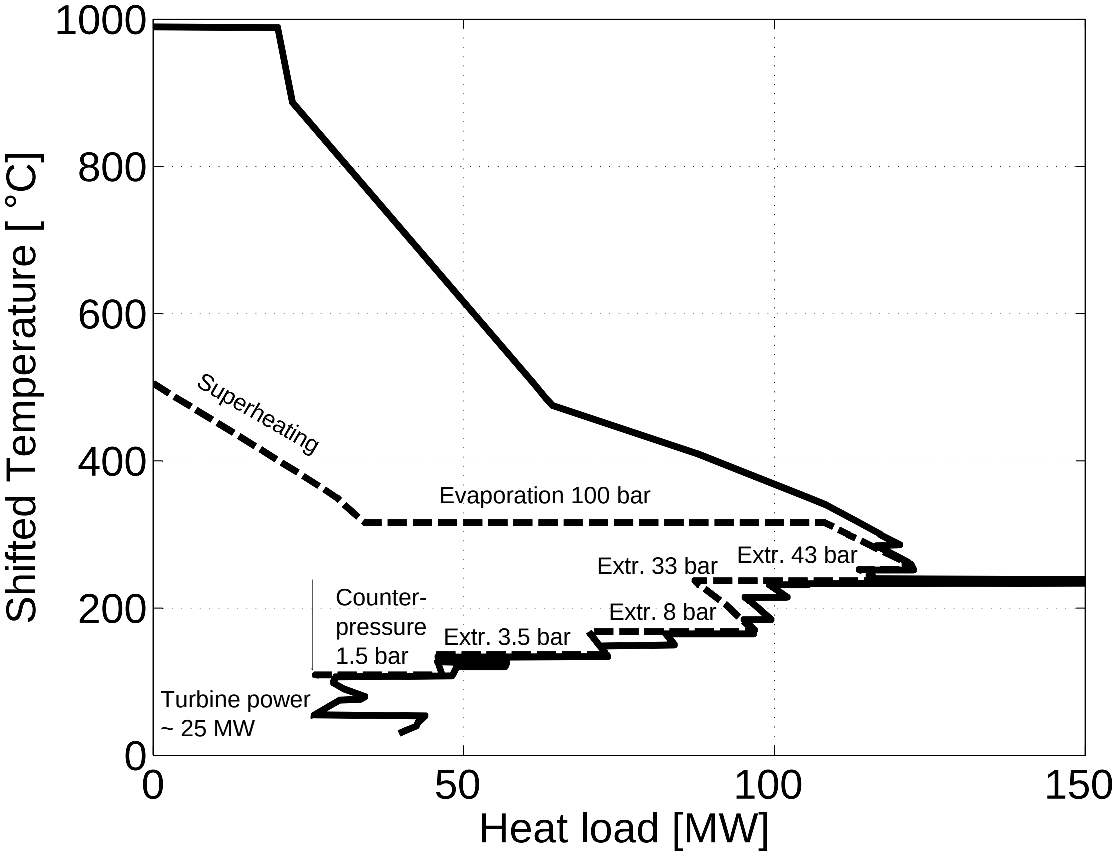

Figure 3.31: Optimal combined heat and power production with integrated back-pressure extraction steam turbine.

The suggested CHP system allows generating around 25 MW of power which corresponds to a net electricity demand of the biomass based processes of around 11 MW. The potential steam export to the cluster is sufficient to cover all the demands below 40 barg and around 80% of the steam demand at 40 barg. No superheated high pressure steam is exported to the Borealis cracker in this case since all the superheated steam is used here for power production through steam expansion.

The steam export to the cluster plant allows about 108 MW of natural gas savings in cluster steam boilers. Due to the more complex arrangement of the steam network that has to be able to deliver heat to multiple cluster plants, this system may require a large investment. In addition, as already underlined, the export of low pressure steam may be less interesting in the case increased heat recovery between cluster plants is made possible through an integrated hot water network.

Table 3.20: steam export to cluster plants from the suggested integrated CHP system.

Pressure (barg)	*Mass flow rate ($t \cdot h^{-1}$)*
40	14.6
28	12.58
20	13.03
10	7.92
6	24.75
2	45.43
1	12.76

3.3.4.6 Summary of the process integration opportunities and system consequences

An overview of the energy balances of the different process integration options is given in *Table 3.21*.

Under the condition that the imported electricity is generated by a fuel-based production technology, it is possible to give a more general picture of the performance of the proposed biomass to methanol system by introducing the "total conversion efficiency" of biomass into methanol.

As shown in the methodology chapter (Equation 2.2), this is similar to the conversion efficiency from biomass to methanol (i.e. the ratio between the energy rates of produced methanol and biomass input) with the only difference the total input energy is accounted for. This is done by deducing from the total energy rate of biomass input the fuel saved in boilers at the Stenungsund site due to the steam export from the biomass based processes to the chemical plants and by adding the fuel spent in marginal electricity production technology to compensate for the electricity imported by the biomass based processes (or deducing this fuel electricity is exported).

Table 3.21: Overview of steam export and electricity import of the third system alternative under different conditions of process integration at the Stenungsund petrochemical cluster.

Configuration	***Maximum steam export to Borealis***	***Maximum steam export to other plants***	***Maximum power generation***	***Combined power and steam export***
Steam export to Borealis cracker (t·h ⁻¹)	*136*	*29*	-	*0*
Steam export to other plants (t·h⁻¹)	0	134	-	131
Fuel saving in cluster boiler (MW NG)	140	140	0	108
Power generation (MW)	-	-	*44*	*25*
Steam turbine configuration	-	-	Condensing / extraction	Counter-press./ extraction
Electricity import (MW)	*40.8*	*40.8*	*-3.2(export!)*	*15.8*

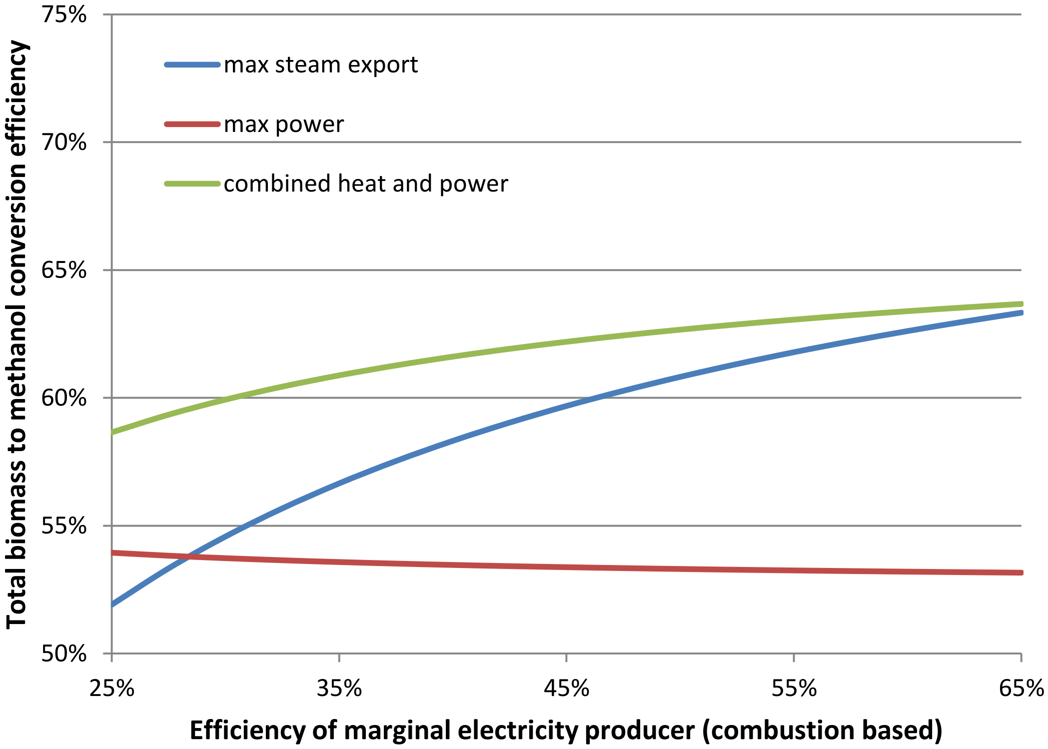

Figure 3.32: Effect of electrical efficiency of combustion based marginal electricity producer on the total biomass to methanol conversion efficiency for the second system alternative and for different process integration options.

It is interesting at this stage to show the influence of efficiency of marginal electricity producer on the total conversion efficiency of biomass to methanol.

Results are shown in *Figure 3.32*, while a breakdown of relevant energy flows around the biomass based methanol production is shown in *Table 3.22* for two efficiency levels of marginal electricity producers.

It is possible to conclude that process integration options and in particular the export of steam to the cluster plants allows a large increase in the overall conversion efficiency. Power production with an integrated condensing steam turbine is not of interest as the efficiency is significantly lower than the marginal electricity producers (only if marginal efficiency is lower than 30% does it become interesting to produce power locally). The opportunities to save natural gas in cluster steam boilers are larger compared to the marginal effect of power production.

From a purely thermodynamic point of view, a combined heat and power system allows to the highest substitution of fuel in boilers and marginal electricity producers. However, electricity production has a minor impact on the total conversion efficiency than the export of steam to cluster boilers. If the latter is maximized, an equally high total conversion efficiency is obtained, especially in presence of high efficiency electricity producers.

Considering the big difference of capital investment between building a combined heat and power system delivering steam to all the other cluster plants and exporting high pressure superheated steam to the nearby Borealis cracker, it is possible to conclude that the latter option is the most interesting one.

Table 3.22: Overview of performances of the biomass to methanol system based on torrefaction and entrained flow gasifier technology, for different process integration options with the existing chemical cluster plants.

Configuration	***Maximum steam export to Borealis***	***Maximum steam export to other plants***	***Maximum power generation***	***Combined power and steam export***
Biomass input energy rate (MW) HHV basis	548.7	548.7	548.7	548.7
Methanol output energy rate (MW) HHV basis	289.09	289.09	289.09	289.09
Biomass to methanol conversion efficiency	*52.7%*	*52.7%*	*52.7%*	*52.7%*
NG saving in cluster boilers (MW), 80% eff.	140 (LHV) 155 (HHV)	140 (LHV) 155 (HHV)	0	108 (LHV) 119 (HHV)
Electricity import (MW)	*40.8*	*40.8*	*-3.2(export!)*	*15.8*
Marginal fuel to electricity (MW)				
... 40 % electrical efficiency	102	102	-8	39.5
... 60 % electrical efficiency	68	68	-5.3	26.3
Total conversion efficiency (HHV basis)				
... 40 % marg. el. eff.	*58.3%*	*58.3%*	*53.5%*	*61.6%*
... 60 % marg. e. eff.	*62.6%*	*62.6%*	*53.2%*	*63.4%*

4 Discussion of performances and technological aspects

The thermodynamic performances of the three system configurations for production of methanol via biomass gasification have been discussed in the previous sections separately. In this chapter, the performances and key technological aspects of the three configurations are compared and discussed.

4.1 Tracking the conversion losses through the processes

A first comparison is made here by tracking the conversion of the biomass into gas and methanol along the process paths adopted for the three system configurations. This is done by calculating the chemical energy rate (flow rate times heating value) of the material streams in some key process points and calculating the ratio between such chemical energy rate and that of the input (biomass at 50 % moisture). This approach is similar to calculating the cold gas efficiency of a gasifier with the only difference that, for other key process points, the chemical energy rate of the gas or liquid streams downstream from the gasifier is used in the numerator instead of that of the gasifier product gas. The higher heating value is used as basis in order to avoid the bias otherwise introduced by the water content.

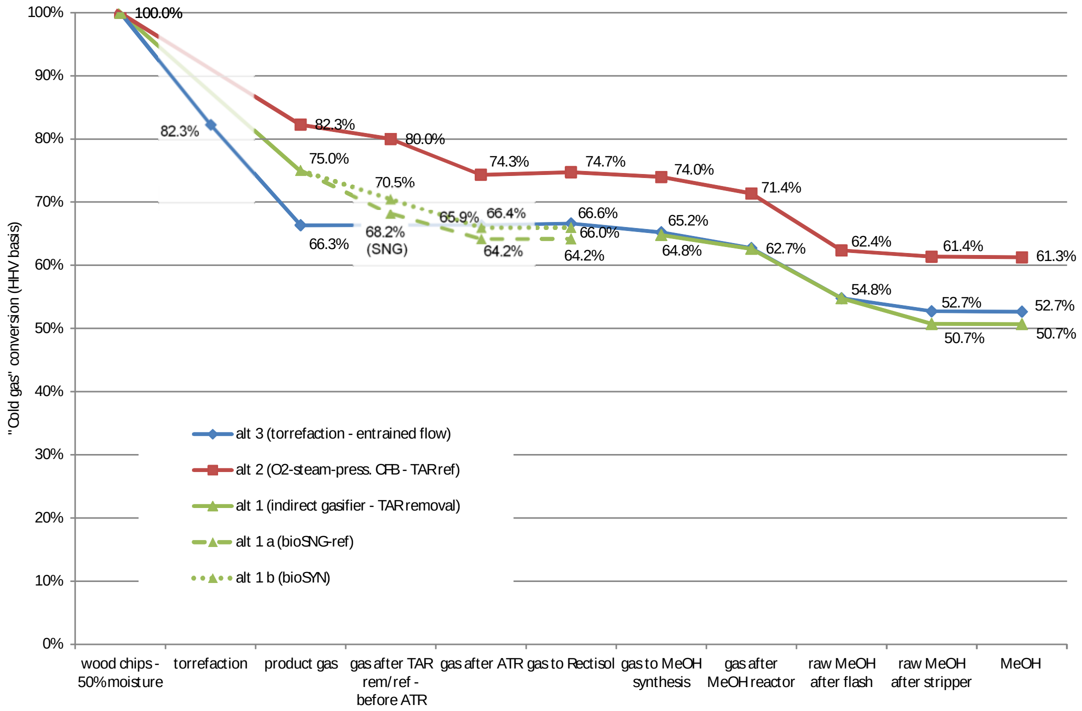

Figure 4.1: Comparison of trend of conversion of biomass into syngas and methanol in various process stages of the three system configurations investigated.

As shown in *Figure 4.1*, the conversion of biomass into methanol is characterised by a continuous decrease of the chemical energy flowrate of the gas along the subsequent process conversion path. The major degradation steps are the gasification unit and separation of the produced methanol from the purge gases after methanol synthesis. A substantial degradation is also introduced by torrefaction in the third system configuration which is itself responsible for about 40% of the conversion losses (82.3% torrefaction energy conversion, 52.7% biomass to methanol conversion).

The biomass to methanol conversion of the first and third system configurations are similar (between 51% and 53% HHV basis) while the second system configuration shows a substantial advantage over the others (about 61% HHV basis).

The same assumptions for autothermal reforming, water gas shift, methanol synthesis and distillation were made for the three system configurations. This is confirmed in *Figure 4.1*, where similar trends can be observed for energy conversion along the process path. Conversely, large differences in terms of conversion efficiency can be observed in the very first steps of biomass conversion due to the different assumptions regarding feedstock pre-treatment, gasification technology and gas cleaning sections.

The major reasons behind the much higher conversion of the second system configuration are the better conversion of both the gasifier and the tar reformer.

It is important to note that the oxygen steam blown pressurized gasifier considered in the second system configuration has not been demonstrated more than at the lab scale, as opposed to the indirect gasifier (first system configuration) and the entrained flow gasifier (third configuration), the latter being very similar to a coal gasifier. There are two important aspects of concern. The almost complete carbon conversion declared for the gasifier (Hannula and Kurkela 2012) are seldom reached in other type of direct fluidized bed gasifiers. Carbon conversion is normally limited by residence time of the biomass in the fluidized bed, so a certain amount of carbon exits with the ashes thus requiring downstream combustion of this residual carbon in the ashes in another boiler. Secondly, the empirical correlations for estimating the content of methane and heavier hydrocarbons were developed for lower pressurized conditions (2.5 bar) than the levels considered in this study (25 bar) which are believed to underestimate the hydrocarbon content in the product gas.

In addition, the heating value of the gas after the reformer increases due to the conversion of all hydrocarbons (including tar) and water into carbon monoxide and hydrogen. While the heat for reforming must be provided by burning some of the product gas in a side combustor from which the heated bed and catalyst are recirculated into the reformer, the net loss in the total conversion amounts to about 2 efficiency points only (from 82.3% at the gasifier outlet to 80% at the reformer outlet). The estimated loss after tar removal is minor compared to the first system configuration in which the product gas contains larger amount of tar which are all removed and burnt in the gasifier combustor. This advantage of the second configuration is achieved at the expense of a considerable investment (the reformer can cost up to 50% of the gasifier cost) compared to the first system configuration in which oil scrubbing constitutes a relatively inexpensive solution. In the third system configuration, no tar are produced because of the high gasification temperature.

Compared to the first system configuration, the advantage in terms of conversion in the second system configuration is about 7 percentage points at the gasifier outlet (config 1: 75%; config 2: 82%) and increases to about 9 percentage points at the end of the last reforming stage (ATR and natural gas reformer). Compared with the third system configuration, the advantage in terms of conversion in the second configuration is instead about 16 percentage points at the gasifier outlet and decreases to 8 percentage points before the shift stage. This indicates that the third configuration is somewhat penalized by the torrefaction step but this penalty progressively reduces because of the much cleaner product gas and the low methane content so that subsequent reforming stages and their intrinsic losses can be avoided.

4.2 Topological choices

The layouts of the three system configurations for production of methanol via biomass gasification were selected based on heuristic choices. The main criterion used was to compare three gasification technologies: (1) atmospheric indirect fluidized bed, (2) pressurized oxygen steam blown fluidized bed, (3) entrained flow.

These three different choices of gasification technology together with specific design parameters (e.g. temperature and pressure) were partially responsible for the choices of subsequent process steps required to complete the conversion route from the gasifier product gas to pure methanol.

4.2.1 Bio-SNG

A specific topological choice was introduced in the definition of the first system configuration which must be discussed separately as a specific criterion was considered. Here, a portion of the total biomass intake is firstly converted into SNG and an (energy) equivalent quantity of NG is then reformed into syngas prior to methanol synthesis. The natural gas grid is used in this case as a buffer system that mitigates the effects of possible disruptions of the biomass gasification unit on the downstream methanol production. This choice implies a particular process layout for the bioSNG process part consisting of water gas shift, carbon dioxide removal and methanation. Natural gas is then reformed in a conventional autothermal reformer and carbon dioxide is removed prior to compression and methanol synthesis. The penalty introduced by methanation and subsequent reforming appears can be read in *Figure 4.1*, and is around 2 percentage points (the conversion efficiency after the methane autothermal reformer is 64.2% versus 66% respectively with and without the intermediate SNG synthesis). The overall impact on the final conversion efficiency of the first system configuration is therefore less than 2 percentage points since only 200 MW of the total 350 MW biomass follow the SNG route, the remaining 150 MW being converted to syngas for methanol synthesis at the Stenungsund site.

4.2.2 Syngas conditioning and compression

A conventional methanol synthesis technology based on syngas with a synthesis ratio[25] of around 2.05 with 3% of CO_2 was assumed for all the three system configurations. The gasifier product gas has in general a lower H_2:CO ratio and much larger quantity

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[25] $(H_2 - CO_2)/(CO + CO_2)$

of CO_2. This excess of carbon must be removed prior to methanol synthesis. A conventional layout based on water gas shift and Rectisol carbon dioxide removal was considered for this purpose. Since methanol synthesis occurs under moderate / high pressurized conditions, gas compression is required after gasification. In addition, the content of methane in the product gas justifies a reforming stage (with the exception for the entrained flow gasifier) in order to avoid large inert recycles in the methanol synthesis loop and to increase the overall conversion by making available methane in form of hydrogen and carbon monoxide. To reduce the volume of the gas cleaning and upgrading process steps and to increase the absorption of carbon dioxide in methanol, it was therefore chosen to compress the gas before reforming when gasification occurs at atmospheric conditions, as in the first system configuration which, in turn, implies that gas must be cooled and cleaned of condensable vapours prior to compression. In the second and third configurations instead, gasification occurs in pressurized conditions (respectively around 25 and 30 bar) which allows avoiding the intermediate compression and severe cooling of the product gas.

4.2.3 Tar removal and reforming

Ultimately, the only significant topological choice, after the choice of gasifier, is on how tar and other pollutants such as particulates, alkali metal, sulphur and chlorine shall be removed. In the third system configuration based on an entrained flow gasifier, no tar are produced and no reforming or scrubbing stage is in principle required. Tar must be removed if a fluidized bed gasifier (either direct or indirect) is selected. The principle followed for the choice of tar removal or reforming was to choose a tar reformer if it is possible to fully exploit its benefits, not least the fact that the gas exits at high temperatures which calls for a complete train of hot gas cleaning and upgrading. This is the case of the second system configuration in which a pressurized steam oxygen blown gasifier is chosen. Additionally, certain amounts of acetylene, ethylene, ethane and propane are produced in this gasifier which could not be removed with a conventional scrubbing stage so that a catalytic pre-reformer would in any case be considered prior to the subsequent methane reformer. In the first system configuration, the cooling stage prior to compression requires cold gas cleaning where tar is removed by oil scrubbing and burnt in the combustor reactor of the indirect gasifier.

It is interesting therefore to quantify the penalty introduced by tar removal by oil scrubbing compared to reforming in terms of conversion efficiency. The loss is about 5 percentage points by tar scrubbing whereas fluidised bed catalytic reforming introduces a loss of around 2 percentage points for the reasons discussed above. It is reasonable to conclude that similar effects could be obtained independently of the gasifier used. If a tar reformer was also used in the first system configuration, this would probably reduce the difference between the first and second system configurations in the overall biomass to methanol conversion to about 7 percentage points (against the 10 points estimated here).

Note that a loss of about 2 percentage points is also introduced by a larger amount of unconverted methane (partially due to intermediate SNG synthesis) in the first configuration which contributes to the whole syngas chemical energy rate but is inert to the methanol synthesis.

If, in addition, a more conservative assumption is used for carbon conversion in the oxygen steam blown CFB gasifier, the differences between the direct and indirect fluidized bed gasifiers and subsequent process routes could substantially level out as far as the conversion of biomass into methanol is concerned.

4.3 Design choices

Some choices have been made about the values of some design parameters. For time constraints, it was not possible to repeat the above analyses for different values of main design parameters so a discussion is included here a posteriori to highlight possible room for improvement of the investigated processes.

4.3.1 Gasification pressure

As already outlined above, the major differences between the three configurations are related to the choice of gasification and gas cleaning technologies. In general, temperature levels are constrained by technological reasons. For instance the temperature of the gasifier product gas is limited by ash melting for fluidized bed gasification technology and reaches an optimal level for obtaining the maximum feedstock conversion in case of entrained flow gasifier. The temperature profile of the gas cleaning sections also depends on specific equipment units such as tar reformer, oil scrubber, candle filter, etc. Temperature levels of ATR, water gas shift and Rectisol are also somewhat determined by the specific thermochemical reactions or vapour-liquid equilibrium.

On the other hand, the pressure level at which gasification occurs is to a certain extent the result of a trade-off between costs for feedstock feeding and pressurization and reduced sized of equipment due to volume reduction. There are reasons to believe that the indirect gasification technology considered in the first system configuration can also operate under pressurized conditions since no significant differences appear between the indirect and direct gasifier as far as biomass feeding and pressurization are concerned. In such case, it is also reasonable to consider an increase in methane and tar content in the product gas as an increase of pressure favours thermodynamically the chemical species of higher density. However, a pressurized gasification somehow favours the adoption of hot gas cleaning section with possibly a tar reformer with the benefits already outlined in the previous section.

4.3.2 Catalyst in gasifier bed

A less measurable but still meaningful assumption is whether catalyst such as olivine, dolomite or zeolite are used in the gasifier bed to reduce the tar content in the gasifier product gas. The quantity of such catalyst and the influence on tar production has not been modelled in this study and conclusions are difficult to make. As a general assumption, the tar content in the indirect fluidized bed gasifier (first configuration) has been assumed to levels that can be achieved in a non-catalytic fluidized bed gasifier (Milne, Abatzoglou et al. 1998), while catalyst is used in the tar reformer in the second system configuration. From experimental evidences reported in the literature, it appears possible to significantly decrease the tar content in indirect fluidized bed gasifier by adding catalyst in the gasification bed (Hofbauer, Rauch et al. 2002). This could help further increase the overall biomass conversion into desired products.

4.3.3 Methanol synthesis pressure and synthesis loop ratio

A separate analysis should be conducted regarding the temperature and pressure level of the methanol synthesis step. As usual, higher pressure favours thermodynamically the reactor conversion since methanol density is higher than that that of syngas. On the other hand, the development of the methanol synthesis technologies has been pushed towards progressively lower pressure levels in order to reduce the cost for compression while overall conversion is somehow optimized by adjusting the amount of syngas loop. The temperature level is somewhat dependent on kinetic factors and could possibly be adjusted depending on operating and investment costs.

This leads to the discussion of another important design parameter which is the amount of synthesis loop in relation to the amount of fresh syngas. The loop ratio has been fixed here to around 3:1 (molar) according to established literature (Supp 1990). This might however significantly affect the amount of inert gases such as traces of nitrogen and methane in the fresh syngas which accumulates in the loop and therefore reduces the use of the active surface of the synthesis reactor that must be therefore oversized to guarantee the maximum conversion of the reactants into methanol.

It should be mentioned at this stage that the outer loop of hydrogen recovered from the purge by membrane separation somewhat mitigates the effect of a reduced synthesis loop since when less synthesis loop is considered, larger amount of hydrogen ends in the purge and larger amount of hydrogen can therefore be recovered (at the expense of larger membranes). This additional loop of hydrogen must be therefore regarded as a another design parameter which can be optimally chosen based on costs for membrane separation and compression and based on the cost of the upstream water gas shift. Still, with lower loop ratio, progressively larger amounts of carbon monoxide and carbon dioxide (plus hydrogen) end up in the purge and cannot participate to methanol synthesis, thereby reducing the overall conversion. As an indication, it has been estimated that with the same fresh syngas composition, the methanol production decreases by 10% when the synthesis loop ratio decreases from 3:1 to 1:1 (molar). Indeed, it also appears that when a synthesis loop ratio of about 2:1 is considered, the conversion loss is minimal, while the total flow of syngas through the reactor reduces by about 30% (which can translate in an equal reduction of investment costs for the methanol synthesis section).

Note that when larger amount of syngas ends up in the purge, this is combusted and therefore the energy is recovered as heat and ultimately as steam which can be exported to a nearby industrial process.

4.3.4 Methanol upgrading, distillation

The methanol obtained after the synthesis product is cooled and flashed still has a considerable amount of gases which are subsequently removed in a stripper. At the bottom of the stripper crude methanol is obtained which contains around 7% vol. of water. When methanol is transported over long distances, it is preferable to remove such water in order to minimize transportation costs. However, when methanol is produced in proximity of the MTO plant, this portion of water does not represent in principle a problem since some water is desired prior to the subsequent conversion of methanol into olefins. It is interesting here to estimate the possible benefits in terms of process operating costs of avoiding methanol distillation while the investment savings were discussed by another project partner.

The advantage in terms of operating costs of avoiding distillation consists essentially in saving the steam that is otherwise required by column reboilers. Since steam is produced by recovering excess heat available from biomass based processes at high and medium temperatures, steam savings due to avoided distillation represent an increased opportunity for steam export to a nearby industrial process. To exemplify this effect, the GCC curves of the third system configuration with and without methanol distillation are shown together in *Figure 4.2*. The effect of avoided distillation can be quantified as an increase of about 10% of steam export which could potentially increase by the same amount the fuel savings in the nearby industrial process plants. The revenues related to fuel savings can represent about 25% of the total revenues (the major part being methanol sales) depending on the location and on the processes, so the overall effect on cash flow of avoided distillation can be estimated to be about at maximum 2 to 3% positive increase. This should be added to the avoided investment in distillation to obtain the overall impact on the process economics.

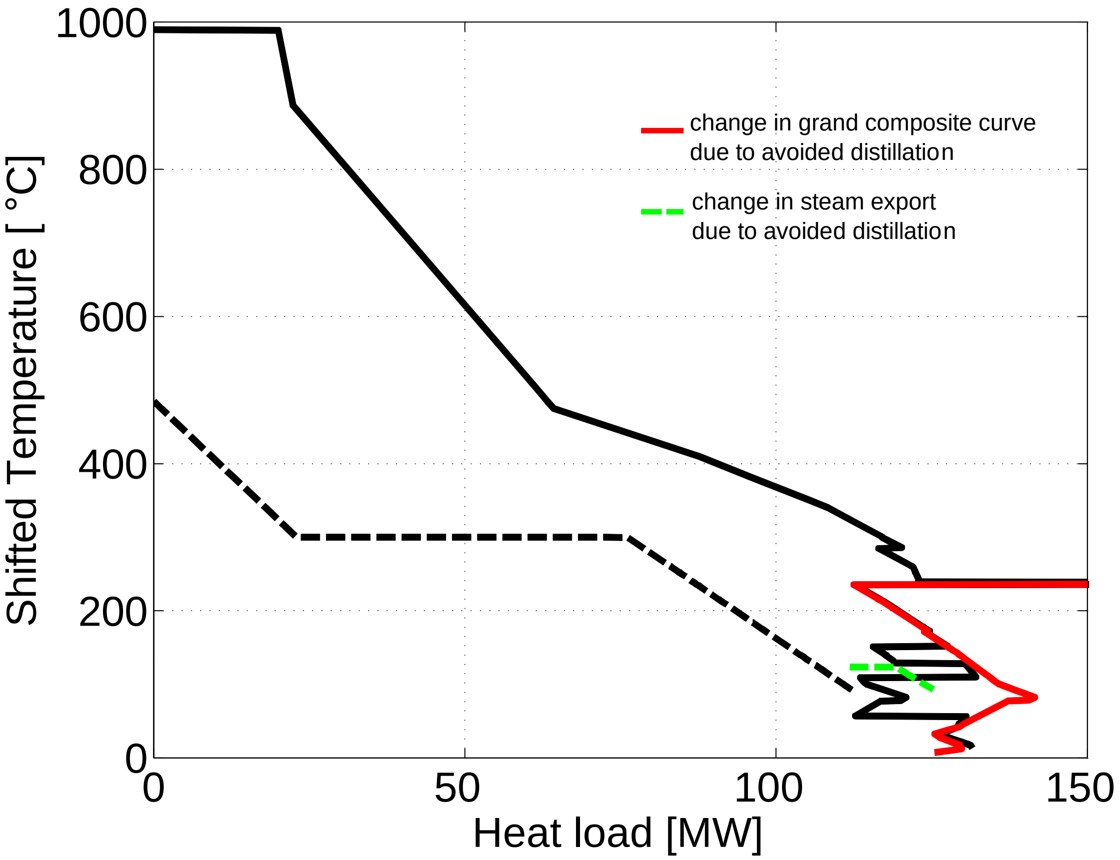

Figure 4.2: Effect of avoided methanol distillation on thermal profile of the gasification based methanol production (configuration 3) and on steam export opportunities.

4.4 Types of feedstock

The mass and energy balances and the results of process integration with the nearby industrial sites shown above were established for wood chips only.

Since different types of biomass may have substantially different prices, one of the aims of the project was to estimate the economics of the same process concepts for different types of biomass, as outlined in the introduction.

For time reasons it was not possible to recalculate in detail the mass and energy balances of the other two relevant feedstock types: forest residues and pellets. The main differences with respect to the results obtained with wood chips are instead discussed in this section.

The feedstock differences are substantially of two types: (a) ultimate composition, that is the percentage of chemical elements (C,H,N,O,Cl,S) and ashes in the dry biomass, (b) the quantity of moisture. The characteristics assumed for the three biomass types are summarized in the introduction (section 2.1.1).

The ultimate analysis of wood chips and pellet are assumed to be the same, while a larger amount of ashes and sulphur and less oxygen are considered for forest residues. This leads to slightly higher heating value of dry forest residues. Nevertheless, the impact on the thermochemical conversion of biomass into methanol appears to be minor.

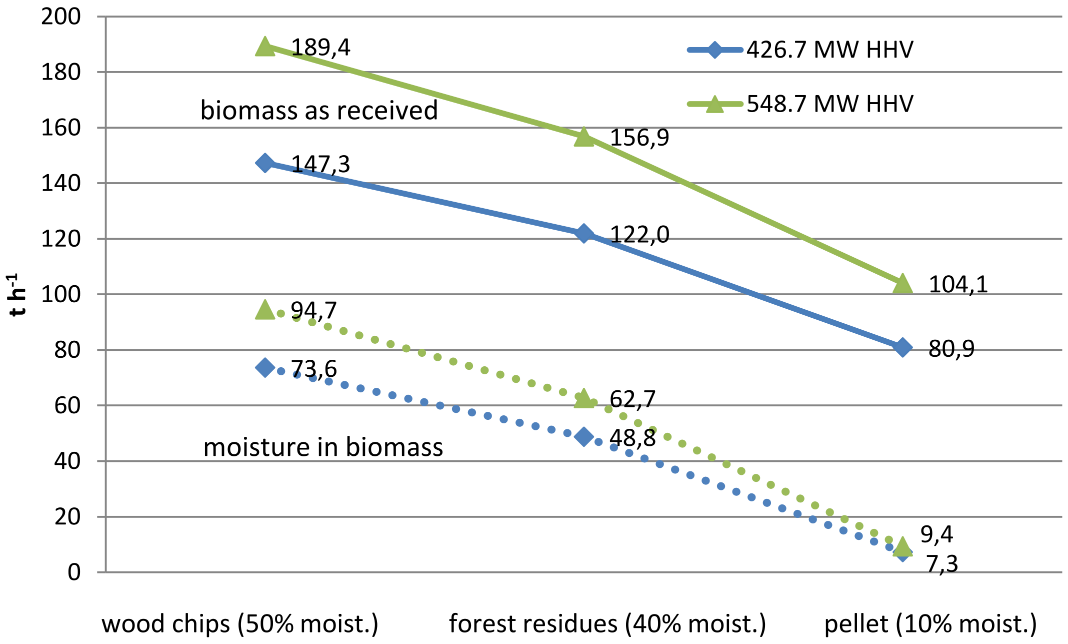

Figure 4.3: Difference in biomass intake and total water content between three relevant feedstock types for constant chemical energy rate on HHV basis.

Conversely, the major difference is in terms of moisture content which causes the quantity of biomass needed to obtain the same amount of methanol to vary considerably depending on the type of biomass as shown in *Figure 4.3*. In this figure the total biomass intake is shown for the two sizes considered in this project (426.7 MW_{HHV} in configuration 1 and 548.7 MW_{HHV} in configurations 2 and 3).

If the small difference in ultimate analysis of forest residues with respect to wood chips and pellet is neglected, it is reasonable to assume that the total chemical energy rate of the input biomass must be the same for constant methanol production. With respect to wood chips, the biomass intake for forest residue and pellets is 17% and 45% less, respectively, which directly translates into equal savings in transportation and storage costs.

Additionally, lower moisture implies reduced need for drying prior to gasification. Indeed, this is true only for forest residues, since drying is in any case necessary prior to pelletizing. However drying in case of pellets is not necessary at all, since the 10% moisture content of pellets is below the required moisture level assumed for gasification in this study.

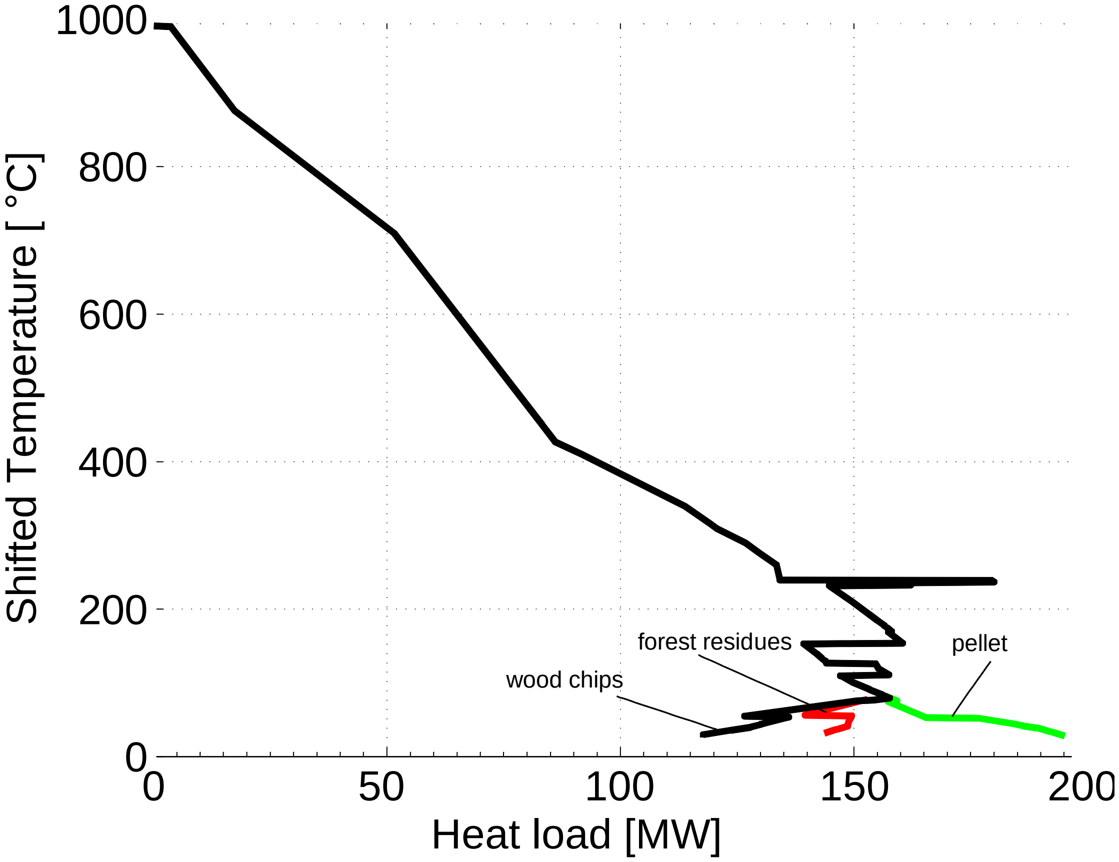

Figure 4.4: Influence of different feedstock drying on the GCC curve of the biomass based methanol production (configuration 2).

In *Figure 4.3*, the reduction in moisture is also highlighted. In the case of forest residue the reduction in water to be removed by drying is about 33.8% less than for wood chips. If the drying load is assumed proportional to the quantity of water removed, less water to be removed translates in an equal increase in heat excess from the biomass based methanol production. To illustrate this aspect, the impact of reduced drying on the GCC curve of system configuration 2 is shown in *Figure 4.4*.

In this example, the excess heat increases from around 115 MW for wood chips to about 140 MW for forest residues, and to about 195 MW for pellets, corresponding to increases of 21% and 73%, respectively. Still, it should be noticed that the effect on process economics depends on the use of this excess heat. If steam is exported, the increase can be assumed equal to the increase in the excess heat. However, the opportunities for steam export might be limited to values lower than the total available excess heat as for the case of Iggesund, or already saturated as in the Stenungsund case, already when wood chips are used, so the potential increase in steam export cannot be accommodated, thus cancelling the actual benefit of reduced drying heat load (for the process sizes considered in this work). The effect of reduced drying heat demand on power generation is in addition quite modest due to the low temperature level assumed for drying.

It can be concluded that the benefit of reduced drying heat load due to feedstock with lower moisture content compared to wood chips is mainly a reduction or the total absence of the wood drier.

4.5 Location and process integration opportunities

Different locations were considered for the different configurations. Since the total biomass intake for each configuration was decided ex ante, the integration opportunities in terms of steam export with the nearby industrial plants vary for the different configurations and impact the overall economics of methanol production in different ways.

In particular it was shown that for both configurations 1 and 3, the excess heat can be fully exploited for the production of high pressure high temperature steam which can be exported to Borealis cracker.

The steam deficit at the cracker is in fact already large and natural gas is currently imported for steam production in boilers. This steam deficit is expected to almost double if an MTO process with an olefin production capacity of around 200 $kt \cdot y^{-1}$ is implemented, thus making the steam import from the gasification based methanol process a vital process integration opportunity.

Another interesting aspect of integration with the Stenungsund cluster that has not been investigated in detail is the possibility of using the hydrogen from hydrocarbon cracking for methanol production by mixing it with the syngas from biomass. From internal communications with Borealis, the amount of hydrogen that could be made available today is about 360 $kg \cdot h^{-1}$.

This is about 5% of the hydrogen obtained from biomass gasification before water gas shift in the third system configuration. The increase in methanol production can be considered essentially equivalent to this extra hydrogen available so it appears definitely an option to consider. Still, it must be noted that other plants at the cluster need hydrogen, so the use of hydrogen for methanol production could lead to extra hydrogen production from natural gas in other plants.

The opportunities of steam export to the Iggesund mill cannot be fully exploited since the capacity of the current steam system at the mill appears lower than the maximum level of high pressure high temperature heat that can be exported from the biomass based methanol production process. Still, it should be noted that the fact that similar biomass is used for pulp and paper production, the logistics of biomass handling, the opportunities of using bark for methanol production, the possibility of retrofit of the auxiliary bark and oil boiler into a gasification plant, the redundancy of the boiler system, may be equally interesting reasons for choosing the Iggesund mill or an equivalent mill as optimal location for a gasification plant. In such case, the size of the methanol production process could be chosen in a more appropriate way to allow fully exploiting the excess heat for steam export to the pulp and paper plant.

References

Aasberg-Petersen, K., T. S. Christensen, C. Stub Nielsen and I. Dybkjær (2003). Recent developments in autothermal reforming and pre-reforming for synthesis gas production in GTL applications. *Fuel Processing Technology* 83(1-3): 253-261.

Althapp, I. A. (2003). Kraftstoffe aus Biomasse mit dem Carbo-V - Vergasungsverfahren. *Regenerative Kraftstoffe*: 107.

Arvidsson, M., S. Heyne, M. Morandin and S. Harvey (2012). *Integration Opportunities for Substitute Natural Gas (SNG) Production in an Industrial Process Plant*. In Proc. of 15th Conference on Process Integration, Modelling and Optimization for Energy Savings and Pollution Reduction (PRES 2012), Prague (CZ).

Axelsson, E., S. Harvey and T. Berntsson (2009). A tool for creating energy market scenarios for evaluation of investments in energy intensive industry. *Energy* 34(12): 2069-2074.

Bengtsson, S. (2011). The CHRISGAS project. *Biomass and Bioenergy* 35: S2-S7.

Bergman, P. C. A., J. M. Prins, A. R. Boersma, K. J. Ptasinski, J. H. A. Kiel and F. J. J. G. Janssen (2005). *Torrefaction for entrained-flow gasification of biomass.* Energy Research Centre of the Netherlands (ECN), Petten (NL).

Bood, J. and L. Nilsson (2013). *Energy analysis of hemicellulose extraction at a softwood Kraft pulp mill*. M.Sc. thesis, Chalmers University of Technology.

Deurwaarder, E. P., H. Boerrigter, H. Mozaffarian, L. Rabou and A. van Der Drift (2005). *Methanation of MILENA product gas for the production of bio-SNG*. 14th European Biomass Conference & Exhibition.

DOE (1998). *Economic analysis LPMEOH process as an add-on to integrated gasification combined cycle (IGCC) for coproduction.* United States Department of Energy, Washington, D.C.

Ecotraffic (1997). *Feasibility phase project for biomass-derived alcohols for automotive and industrial uses.* Ecotraffic R&D AB, Nykomb Synergetics AB, Stockholm (SE).

Ekbom, T., M. Lindblom, N. Berglin and P. Ahlvik (2003). *Technical and Commercial Feasibility Study of Black Liquor Gasification with Methanol/DME Production as Motor Fuels for Automotive Uses-BLGMF.*

Farris, M., M. Paisley, J. Irving and R. Overend (1998). *The Biomass Gasification Process by Battelle/Ferco: Design, Engineering, Construction, and Startup*. 1998 Gasification Technologies Conference.

Fu, C. and T. Gundersen (2011). Power reduction in air separation units for oxy-combustion processes based on exergy analysis. 29: 1794-1798.

Greil, C., H. Hirschfelder, O. Turna, T. Obermeier and S. P. Sekundärrohstoffverwertungszentrum (2002). *Operational Results From Gasification Of Waste Material And Biomass In Fixed Bed and Circulating Fluidised Bed Gasifiers*. 5 European gasification conference, th.

Hackl, R., E. Anderson and S. Harvey (2010). *Total Site Analysis (TSA) Stenungsund.* CHALMERS, Göteborg, Sweden.

Hamelinck, C. N. and A. P. C. Faaij (2002). Future prospects for production of methanol and hydrogen from biomass. *Journal of Power Sources* 111: 1-22.

Hannula, I. and E. Kurkela (2010). A semi-empirical model for pressurised air-blown fluidized-bed gasification of biomass. *Bioresour Technol* 101(12): 4608-4615.

Hannula, I. and E. Kurkela (2012). A parametric modelling study for pressurised steam/O2-blown fluidised-bed gasification of wood with catalytic reforming. *Biomass and Bioenergy* 38: 58-67.

Hannula, I. and E. Kurkela (2013). *Liquid transportation fuels via large-scale fluidised-bed gasification of lignocellulosic biomass.* VTT Technical Research Centre of Finland, Espoo, Finland.

Heyne, S., T. Liliedahl and M. Marklund (2013). *Biomass gasification - a synthesis of technical barriers and current research issues for deployment at large scale.* F3 Swedish Knowledge Centre for Renewable Transportation Fuels,

Heyne, S., H. Thunman and S. Harvey (2012). Extending existing combined heat and power plants for synthetic natural gas production. *International journal of energy research* 36(5): 670-681.

Hofbauer, H., R. Rauch, G. Loeffler, S. Kaiser, E. Fercher and H. Tremmel (2002). *Six years experience with the FICFB-gasification process*.

Isaksson, J., K. Pettersson, M. Mahmoudkhani, A. Åsblad and T. Berntsson (2012). Integration of biomass gasification with a Scandinavian mechanical pulp and paper mill – Consequences for mass and energy balances and global CO2 emissions. *Energy* 44(1): 420-428.

Joelsson, J. M., C. Engström and L. Heuts (2015). *From green forest to green commodity chemicals: evaluating the potential for large scale production in Sweden for three value chains (Final report from Skogskemi: a project with the programme Challenge-Driven Innovation).* VINNOVA,

Kemp, I. C. (2007). *Pinch analysis and process integration: a user guide on process integration for the efficient use of energy*, Butterworth Heinemann.

Milne, T. A., N. Abatzoglou, R. J. Evans, I. Kemestrie and N. R. E. Laboratory (1998). *Biomass gasifier" tars": their nature, formation, and conversion.* National Renewable Energy Laboratory, Golden (CO), USA.

Prins, J. M. (2005). *Thermodynamic analysis of biomass gasification and torrefaction*. Ph.D. thesis, Eindhoven Technical University.

Spath, P., A. Aden, T. Eggeman, M. Ringer, B. Wallace and J. Jechura (2005). *Biomass to Hydrogen Production Detailed Design and Economics Utilizing the Battelle Columbus Laboratory Indirectly-Heated Gasifier.* NREL National Renewable Energy Laboratory, Golden (CO), USA.

Supp, E. (1990). *How to produce methanol from coal*, Springer-Verlag Berlin.

Södra. "Södra Cell Värö." 2014, from www.sodra.com.

Van Der Drift, A., H. Boerrigter, B. Coda, M. K. Cieplik and K. Hemmes (2004). *ENTRAINED FLOW GASIFICATION OF BIOMASS - Ash behaviour, feeding issues, and system analyses.* ECN Biomass,

Weiland, F., H. Hedman, M. Marklund, H. Wiinikka, O. Öhrman and R. Gebart (2013). Pressurized Oxygen Blown Entrained-Flow Gasification of Wood Powder. *Energy & Fuels* 27(2): 932-941.

Weiland, F., M. Nordwaeger, I. Olofsson, H. Wiinikka and A. Nordin (2014). Entrained flow gasification of torrefied wood residues. *Fuel Processing Technology* 125: 51-58.

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Appendix

Main modelling assumptions

General assumptions

Biomass input: 15°C, 1 atm, moisture 50% at the drier inlet, reduced to 15% before gasifier.
Wood chips ULTIMATE ANALYSIS (% mass):
ash 0.156, C 51.9, O_2 41.8, H_2 6, N_2 0.12, Cl 0.015, S 0.009.

Property models: Redlich Kwong Soave, NRTL is used for vapour liquid equilibria when necessary.
10% excess of air in gas combustions, 20% excess of air in solid fuel combustions.
Exhaust gas stack temperature: 150°C
Pumps, compressors, turbine eff.: 0.75
Mechanical and electrical driver eff.: 0.98
Compression intercooling end temp.: 50°C (water is separated with flash)
Pressure drops in reactors: 0.2 bar unless specified
No pressure drops in heat exchangers

Air-drier

Inlet biomass moisture: 50%
Outlet biomass moisture: 15%
Max air temperature: 70
Out air temperature: 20

Indirect biomass gasifier

Temperature: 850°C
Pressure: 1 atm
Steam to Biomass ratio: 0.5
CH_4 molar fraction in product gas: 10% of dry gas (Hofbauer, Rauch et al. 2002)
Total tar yields in product gas: 3 $g \cdot Nm^{-3}$ dry gas (Milne, Abatzoglou et al. 1998)
tar modelled as Phenol, Naphthalene and Toluene, and yields fixed as follows (Milne, Abatzoglou et al. 1998):
Phenol yield in product gas: 55% mass total tar
Naphthalene yield in product gas: 20% mass total tar
Toluene yield in product gas: 25% mass total tar
Biomass S is fully converted in H_2S.
Biomass Cl is fully converted in HCl.
Biomass N is fully converted in Ammonia.
Carbon conversion is set so that the unreacted C is burnt in the combustor together with the tar removed in downstream scrubber and the combustion heat above 900°C balances the gasification heat demand.
CO, H_2O, CO_2, H_2, reacting C, are all considered at equilibrium.

Gas cleaning for indirect gasification system

Particle filter is modelled as a pressure drop.

Tar, Ammonia are removed through an oil scrubber and sent to combustor, heating value of oil is neglected from energy balance.

Oxygen steam blown pressurized fluidized bed gasification

The model has been taken from (Hannula and Kurkela 2012).

Temperature T_{gas}: 886°C
Pressure: 25 bar
Steam to Oxygen ratio: 1
Carbon conversion: 99%

Steam injection is adjusted to obtain extra heat (loss) equal to 1% of the input biomass chemical energy rate in HHV which corresponds to a steam to biomass ratio of around 0.5 (biomass moisture excluded).

Methane molar fraction in product gas: $7.074\ mol\cdot kg^{-1} - 0.0155\ T_{gas}$
Acetylene molar fraction in product gas: $0.06454\ mol\cdot kg^{-1} - 0.00004\ T_{gas}$
Ethylene molar fraction in product gas: $2.987\ mol\cdot kg^{-1} - 0.002\ T_{gas}$
Ethane molar fraction in product gas: $1.196\ mol\cdot kg^{-1} - 0.001\ T_{gas}$
Propane molar fraction in product gas: $0.150921\ mol\cdot kg^{-1} - 0.000155\ T_{gas}$
Benzene molar fraction in product gas: $0.27\ mol\cdot kg^{-1}$
Naphthalene molar fraction in product gas: $0.3\ mol\cdot kg^{-1}$
Ammonia molar fraction in product gas: $0.04154\ mol\cdot kg^{-1}$

Biomass S is fully converted in H_2S.
Biomass Cl is fully converted in HCl.
Biomass N_2 not yielding Ammonia is inert.
CO, H_2O, CO_2, H_2, reacting C, are all considered at equilibrium.

Catalytic tar reformer

Tar reforming temperature is set equal to temperature of incoming gasifier product gas (here 886°C). Methane reforming and water gas shift reaction are assumed at equilibrium (WGS with 40 K of temperature approach).

Hydrocarbon conversion higher than methane is set according to (Spath, Aden et al. 2005).

The following conversions are assumed:

Acetylene 50%, Ethylene 50%, Ethane 90%, Propane 90%, Benzene 70%, Naphthalene 95%, Ammonia 70%.

Entrained flow gasification

Temperature: 1300°C
Pressure: 30 atm

H_2, CO, CO_2, H_2O are considered at equilibrium with the exception for methane which concentration is set at 1%-mol in dry product gas.

All biomass sulphur ends up in H_2S, all biomass chlorine ends up in HCl, all biomass nitrogen ends up in NH_3.

Autothermal reforming (for biomass syngas)

Adiabatic reactor with steam and O_2 injection.

Steam to carbon (in hydrocarb.) ratio: 1 (this is achieved by steam injection, if enough steam is present in the gas, no steam is injected)

Exiting temperature: 1000°C

All species are considered at equilibrium.

Water-gas-shift reactor
Adiabatic reactor with steam injection to avoid carbon deposition.
Steam to CO ratio: 3
Inlet temperature: 330 / 350°C
water-gas-shift at equilibrium with 10°C temperature approach.

Methanation (SNG plant)
Adiabatic reactor with steam injection to avoid carbon deposition.
H_2O molar fraction in fresh syngas: 20%
All species are considered at equilibrium.
MDEA CO_2 removal (SNG plant)
Modelled as a 99.5% CO2 selective separation.
Stripper heat demand assumed to be 3.3 MJ/kg at 150°C.

Rectisol CO_2 removal
Modelled as a CO_2 selective separation to achieve the desired CO_2 content in the syngas.
Syngas inlet temperature: 30°C
Syngas outlet temperature: 10°C
Refrigeration work: 0.55 $kWh \cdot kmol^{-1}$ CO_2 removed
Steam demand at 6 bar: 3.4 $kg \cdot kmol^{-1}$ CO2 removed

Methanol Synthesis
Temperature: 250°C
Pressure: 90 bar
Pressure drop: 5 bar
Synthesis loop ratio: 3
Stoichiometric equilibrium reactor with temperature approach of 27°C according to data found in (DOE 1998) for both reactions:
Methanol synthesis of CO $CO + 2H_2 \leftrightarrow CH_3OH$
Water gas shift $CO + H_2O \leftrightarrow CO_2 + H_2$
Methanol separation and upgrading
Stripper is modelled as equilibrium column with 10 stages, 1.6 reflux ratio, bottom to feed ratio around 0.9 (depending on the composition of the incoming raw methanol stream).
Distillation of methanol from water is simulated using a shortcut distillation model (Winn, Underwood, Gilliland methods) considering 20 stages and limiting the water concentration in distillate to 0.1%. Methanol recovery is imposed in the first column to obtain a balanced heat demand between the condenser of the first column and the reboiler of the second column, while in the second column it is imposed to obtain 99.9% recovery.

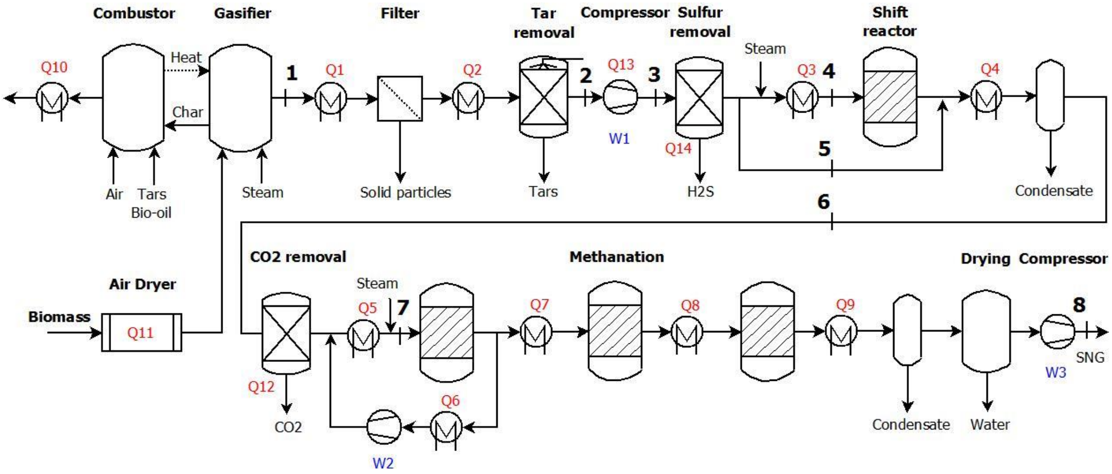
Mass and energy balances

Note on nomenclature:

Material streams are shown in black bold numbers
Heat sinks / sources are shown in red with prefix "Q"
Power demands are shown in blue with prefix "W"

System configuration n 1

SNG process



Biomass input (50% moisture, drier inlet): 84 158 kg·h^{-1} (200 MW LHV basis, 243 MW HHV basis)

Stream	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**
Mass flow (kg h^{-1})	51243	50284	37265	25776	19055	35100	75403	10944
Temp (°C)	850	110	40	330	40	40	300	55
Pressure (atm)	0.9	0.9	29	28	27	26	26	31
Molar flow (kmol·h^{-1})	2849	2840	2118	1583	1291	2333	5773	693
Mol frac. (%)								
H_2	34.0	34.1	45.7	27.0	45.7	54.9	37.2	3.6
CO	18.5	18.6	24.9	15.6	24.9	18.3	8.4	-
CO_2	13.2	13.2	17.7	4.4	17.7	16.1	3.0	0.9
H_2O	26.3	26.4	1.4	46.9	1.4	0.3	20.0	-
CH_4	7.6	7.6	10.2	6.1	10.2	10.4	31.3	95.5
N_2	-	-	-	-	-	-	-	-
O_2	-	-	-	-	-	-	-	-
Toluene	0.0	-	-	-	-	-	-	-
Naphthalene	0.1	-	-	-	-	-	-	-
Phenol	0.2	-	-	-	-	-	-	-
Ammonia	0.1	-	-	-	-	-	-	-
H_2S	0.0	0.0	0.0	-	-	-	-	-
HCl	0.0	-	-	-	-	-	-	-

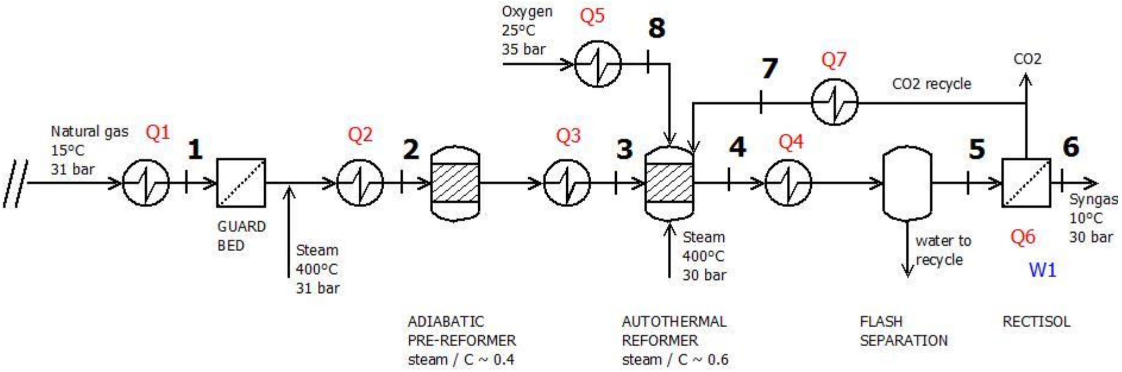
Heat sink/source	**Q1+Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q7**	**Q8**
Tin (°C)	850	189	458	231	625	625	452
Tout (°C)	110	330	40	300	300	250	250
Heat load (kW)	30706	2595	12375	4220	15776	6657	3305

Heat sink/source	**Q9**	**Q10**	**Q11**	**Q12**	**Q13**	**Q14**
Tin (°C)	314	900	20	140	244	140
Tout (°C)	30	150	70	141	40	141
Heat load (KW)	10545	30084	34854	11981	12056	15570

Power demands	**tot**
Power (kW)	19.4*

*this power is completely covered by steam power generation (calculations are reported in the respective section)

Natural gas autothermal reforming



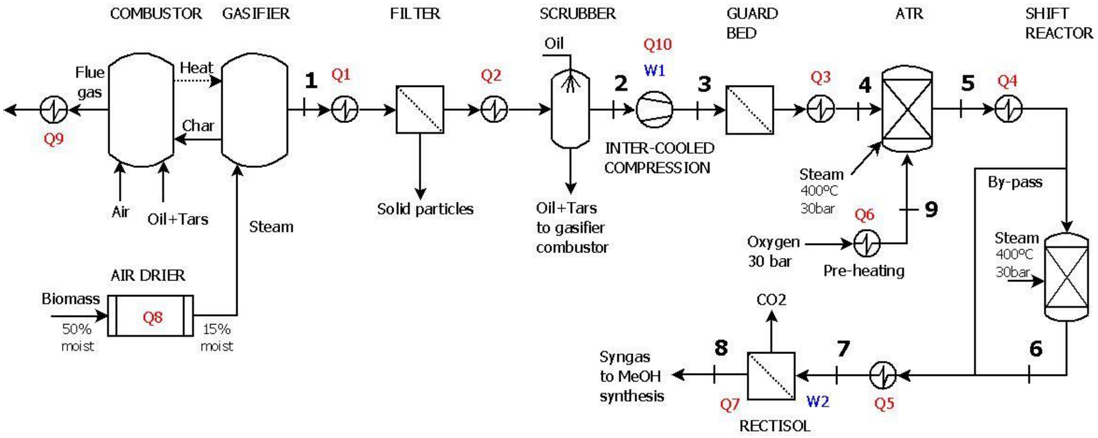
NG input: 11 270 kg·h^{-1} (Swedish natural gas)

	1	2	3	4	5	6	7	8
Mass flow (kg·h^{-1})	11269	15604	15604	31979	24199	22266	1218	11206
Temp (°C)	300	430	750	1000	30	10	400	400
Pressure (atm)	30.6	29.6	29.6	29.6	29.6	29.6	29.6	30
Molar flow (kmol·h^{-1})	609.84	850.45	850.45	2407.38	1989.44	1943.91	28.68	352.40
Mol frac. (%)								
H_2	-	-	3.9	52.1	63.0	64.5	-	-
CO	-	-	0.1	23.6	28.3	29.0	-	-
CO_2	1.0	0.7	4.1	4.6	5.2	3.1	94.0	-
H_2O	-	28.3	19.4	16.9	0.2	0.0	6.0	-
CH_4	88.8	63.7	72.3	2.0	2.3	2.4	-	-
N_2	0.3	0.2	0.2	0.8	1.0	1.0	-	5.0
O_2	-	-	-	0.0	-	-	-	95.0
Ethane	6.1	4.4	-	-	-	-	-	-
Propane	2.5	1.8	-	-	-	-	-	-
Butane	0.9	0.7	-	-	-	-	-	-
Pentane	0.2	0.2	-	-	-	-	-	-
Hexane	0.1	0.0	-	-	-	-	-	-

Heat sink/source	**Q1**	**Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q7**
Tin (°C)	15	322	434	1000	25	232	233
Tout (°C)	300	400	750	30	400	233	400
Heat load (kW)	2354	1427	4787	26576	1162	156	103

Power demand	**W1**
Power (kW)	536

Syngas production through biomass indirect gasification



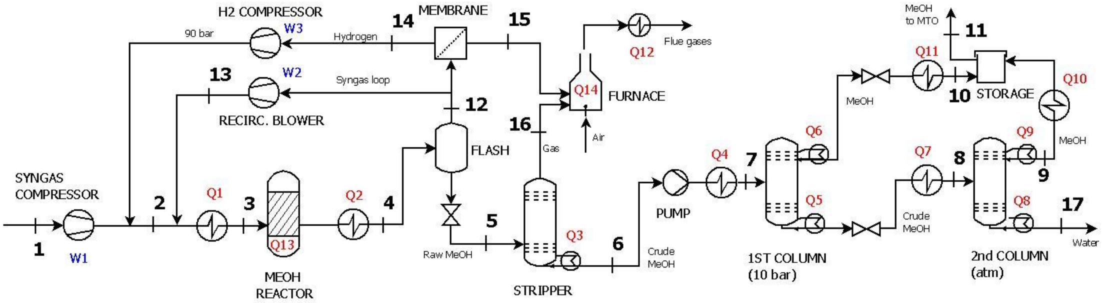
Biomass input (50% moisture, drier inlet): 63 108 kg·h^{-1} (150 MW LHV basis, 182.7 MW HHV basis)

Stream	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**
Mass flow (kg·h^{-1})	38432	37707	27944	27944	34192	23400	44784	17136	3729
Temp (°C)	850	70	220	750	999	493	30	10	25
Pressure (atm)	0.9	0.9	29.6	29.6	29.6	29.6	29.6	29.6	29.6
Molar flow (kmol·h^{-1})	2137	2130	1588	1588	2007	1340	2595	1521	117
Mol frac. (%)									
H_2	34.0	34.1	45.7	45.7	40.1	36.9	38.4	65.6	-
CO	18.5	18.6	24.9	24.9	32.0	3.6	17.4	29.6	-
CO_2	13.2	13.2	17.7	17.7	8.7	19.2	14.1	3.1	-
H_2O	26.3	26.4	1.4	1.4	17.9	39.6	29.1	0.0	-
CH_4	7.6	7.6	10.2	10.2	1.2	0.6	0.9	1.5	-
N_2	-	-	-	-	0.1	0.1	0.1	0.2	5.0
O_2	-	-	-	-	0.0	0.0	0.0	-	95.0
Toluene	0.0	-	-	-	-	-	-	-	-
Naphthalene	0.1	-	-	-	-	-	-	-	-
Phenol	0.2	-	-	-	-	-	-	-	-
Ammonia	0.1	-	-	-	-	-	-	-	-
H_2S	0.0	0.0	0.0	-	-	-	-	-	-
HCl	0.0	-	-	-	-	-	-	-	-

Heat sink/source	**Q1+Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q7**	**Q8**	**Q9**	**Q10**
Tin (°C)	850	220	1000	428	25	232	20	900	215
Tout (°C)	70	750	350	30	400	233	70	150	50
Heat load (kW)	17429	8702	12838	19092	385	630	27619	21933	11987

Power demands	**W1**	**W2**
Power (kW)	7535	1075

Methanol synthesis



Syngas inlet: (17 136 + 22 266 =) 39 402 kg·h^{-1} (MW LHV basis, HHV basis)

Stream	1	2	3	4	5	6	7	8	9
Mass flow (kg·h^{-1})	39402	39663	108160	108160	38607	35830	35830	21386	19926
Temp (°C)	10	17	78	250	45	66	120	40	65
Pressure (atm)	29.6	29.6	88.8	83.9	83.9	1.0	9.9	1.0	1.0
Molar flow (kmol·h^{-1})	3465	3595	14405	12241	1263	1153	1153	702	622
Mol frac. (%)									
H_2	65.0	66.2	78.1	73.5	0.0	0.0	0.0	-	-
CO	29.3	28.2	9.5	3.0	0.4	0.0	0.0	-	-
CO_2	3.1	3.0	1.8	1.5	2.0	0.0	0.0	0.0	-
H_2O	0.0	0.0	0.0	0.7	6.3	6.9	6.9	11.4	0.0
CH_4	2.0	2.0	4.3	5.1	4.9	0.0	0.0	0.0	-
N_2	0.6	0.6	5.8	6.8	0.7	0.0	0.0	-	-
O_2	-	-	-	-	-	-	-	-	-
MeOH	-	-	0.5	9.4	85.6	93.1	93.1	88.6	100.0

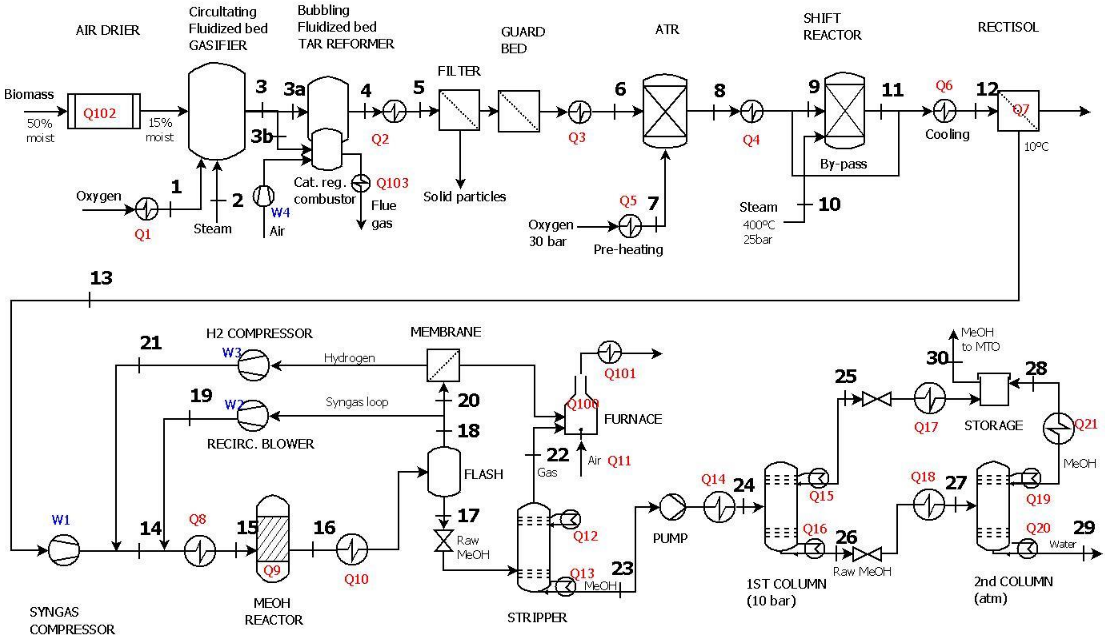
Stream	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**
Mass flow (kg·h^{-1})	14444	34370	69553	68499	261	793	2776	1460
Temp (°C)	137	40	45	52	200	45	10	99
Pressure (atm)	9.9	1.0	83.9	88.8	29.6	83.9	1.0	1.0
Molar flow (kmol·h^{-1})	451	1073	10977	10811	130	37	110	81
Mol frac. (%)								
H_2	-	-	82.0	82.0	100.0	18.6	0.0	-
CO	-	-	3.3	3.3	-	15.1	4.8	-
CO_2	0.0	0.0	1.4	1.4	-	6.5	23.2	-
H_2O	0.0	0.0	0.0	0.0	-	0.1	0.0	99.2
CH_4	0.0	0.0	5.1	5.1	-	23.1	56.4	-
N_2	-	-	7.5	7.5	-	33.9	8.5	-
O_2	-	-	-	-	-	-	-	-
MeOH	100.0	100.0	0.6	0.6	-	2.8	7.1	0.8

Heat sink/source	**Q1**	**Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q7**	**Q8**	**Q9**
Tin (°C)	78	250	65	66	140	135	137	99	65
Tout (°C)	250	50	66	120	141	134	40	100	64
Heat load (kW)	21066	34538	2877	1995	14303	13584	1412	13987	13395

Heat sink/source	**Q10**	**Q11**	**Q12**	**Q13**	**Q14**
Tin (°C)	65	137	1000		
Tout (°C)	40	40	150		
Heat load (kW)	457	1412	8273	29153	10510

Power demands	Power (kW)
W1	4332
W2	650
W3	164

System configuration n.2



Biomass input (50 % moisture, drier inlet): 189 360 kg h⁻¹ (450 MW LHV basis, 548.7 MW HHV basis)

Stream	**1**	**2**	**3**	**3b**	**3a**	**4**	**5**	**6**
Mass flow (kg·h^{-1})	32 833	32 833	176 981	18 583	158 398	158 398	158 398	158 377
Temp (°C)	200	228	886	886	886	886	300	700
Pressure (atm)	24.7	24.7	24.7	24.7	24.7	24.5	24.0	24.0
Molar flow (kmol·h^{-1})	1 026	1 822	7 775	816	6 959	8 180	8 180	8 180
Mol frac. (%)								
H_2	-	-	20.1	20.1	20.1	32.0	32.0	32.0
CO	-	-	15.8	15.8	15.8	27.2	27.2	27.2
CO_2	-	-	21.9	21.9	21.9	14.6	14.6	14.6
H_2O	-	100.0	34.1	34.1	34.1	23.4	23.4	23.4
CH_4	-	-	5.4	5.4	5.4	2.6	2.6	2.6
N_2	-	-	0.027	0.027	0.027	0.042	0.0	0.0
O_2	100.0	-	0.000	-	-	-	-	-
Acetylene	-	-	0.035	0.035	0.035	0.003	0.003	0.003
Ethylene	-	-	1.476	1.476	1.476	0.126	0.126	0.126
Ethane	-	-	0.377	0.377	0.377	0.003	0.003	0.003
Propane	-	-	0.017	0.017	0.017	0.000	0.000	0.000
Benzene	-	-	0.328	0.328	0.328	0.003	0.003	0.003
Naphthalene	-	-	0.364	0.364	0.364	0.000	0.000	0.000
Ammonia	-	-	0.050	0.050	0.050	0.004	0.004	0.004
H_2S	-	-	0.003	0.003	0.003	0.003	0.003	-
HCl	-	-	0.005	0.005	0.005	0.004	0.004	-
Methanol	-	-	-	-	-	-	-	-

Stream	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**
Mass flow (kg·h^{-1})	9 914	168 291	74 754	36 311	111 065	204 575	59 644	60 141
Temp (°C)	200	1 001	350	400	503	50	59	264
Pressure (atm)	30.0	23.6	23.6	29.6	23.5	23.5	23.5	88.8
Molar flow (kmol·h^{-1})	311	8 616	3 828	2 016	5 843	10 631	5 301	5 547
Mol frac. (%)								
H_2	-	31.2	31.200	-	34.200	32.900	65.900	67.416
CO	-	28.0	28.000	-	4.500	15.100	29.700	28.390
CO_2	-	14.2	14.300	-	23.100	19.100	3.100	2.936
H_2O	-	26.3	26.300	100	37.900	32.700	0.800	0.804
CH_4	-	0.2	0.200	-	0.100	0.200	0.300	0.281
N_2	2.0	0.1	0.100	-	0.100	0.100	0.200	0.173
O_2	98.0	0.0	-	-	-	-	-	-
Acetylene	-	0.0	-	-	-	-	-	0.000
Ethylene	-	0.0	-	-	-	-	-	0.000
Ethane	-	-	-	-	-	-	-	-
Propane	-	0.0	-	-	-	-	-	0.000
Benzene	-	0.0	-	-	-	-	-	-
Naphthalene	-	0.0	-	-	-	-	-	-
Ammonia	-	0.0	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-	-
Methanol	-	-	-	-	-	-	-	-

Stream	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**
Mass flow (kg·h⁻¹)	130 599	130 599	58 950	71 649	70 456	1 192	497	2 475	56475
Temp (°C)	250	250	50	50	57	50	408	18	66
Pressure (atm)	88.8	83.9	83.9	83.9	88.8	83.9	88.8	1.0	1
Molar flow (kmol·h⁻¹)	22 190	18 830	1 904	16 925	16 643	281.67	246.5	70.5	1834
Mol frac. (%)									
H_2	85.936	82.791	0.000	92.109	92.109	92.109	100	0.001	0.0
CO	9.191	2.543	0.328	2.792	2.792	2.792	-	8.855	0.0
CO_2	1.823	1.514	2.064	1.452	1.452	1.452	-	55.781	0.0
H_2O	0.219	0.892	8.611	0.023	0.023	0.023	-	0.003	8.9
CH_4	0.657	0.774	0.702	0.782	0.782	0.782	-	18.962	0.0
N_2	1.621	1.910	0.195	2.103	2.103	2.103	-	5.259	0.0
O_2	-	-	-	-	-	-	-	-	0.0
Acetylene	0.000	0.000	0.000	0.000	0.000	0.000	-	0.000	0.0
Ethylene	0.000	0.000	0.000	0.000	0.000	0.000	-	0.000	0.0
Ethane	-	-	-	-	-	-	-	-	0.0
Propane	0.000	0.000	0.000	0.000	0.000	0.000	-	0.000	0.0
Benzene	-	-	-	-	-	-	-	-	0.0
Naphthalene	-	-	-	-	-	-	-	-	0.0
Ammonia	-	-	-	-	-	-	-	-	0.0
H_2S	-	-	-	-	-	-	-	-	0.0
HCl	-	-	-	-	-	-	-	-	0.0
Methanol	0.554	9.575	88.100	0.738	0.738	0.738	-	11.139	91.1

Stream	**24**	**25**	**26**	**27**	**29**	**28**	**30**
Mass flow (kg·h^{-1})	56475	38564	17911	17911	2941	14970	53534
Temp (°C)	120	137	143	40	98	65	40
Pressure (atm)	10	10	10	1	1	1	1
Molar flow (kmol·h^{-1})	1834	1204	630	630	162	468	1672
Mol frac. (%)							
H_2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO_2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H_2O	8.9	0.1	25.8	25.8	99.1	0.3	0.2
CH_4	0.0	0.0	0.0	0.0	0.0	0.0	0.0
N_2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O_2	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Acetylene	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethylene	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ethane	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Propane	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Benzene	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Naphthalene	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Ammonia	0.0	0.0	0.0	0.0	0.0	0.0	0.0
H_2S	0.0	0.0	0.0	0.0	0.0	0.0	0.0
HCl	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Methanol	91.1	99.9	74.2	74.2	0.9	99.7	99.8

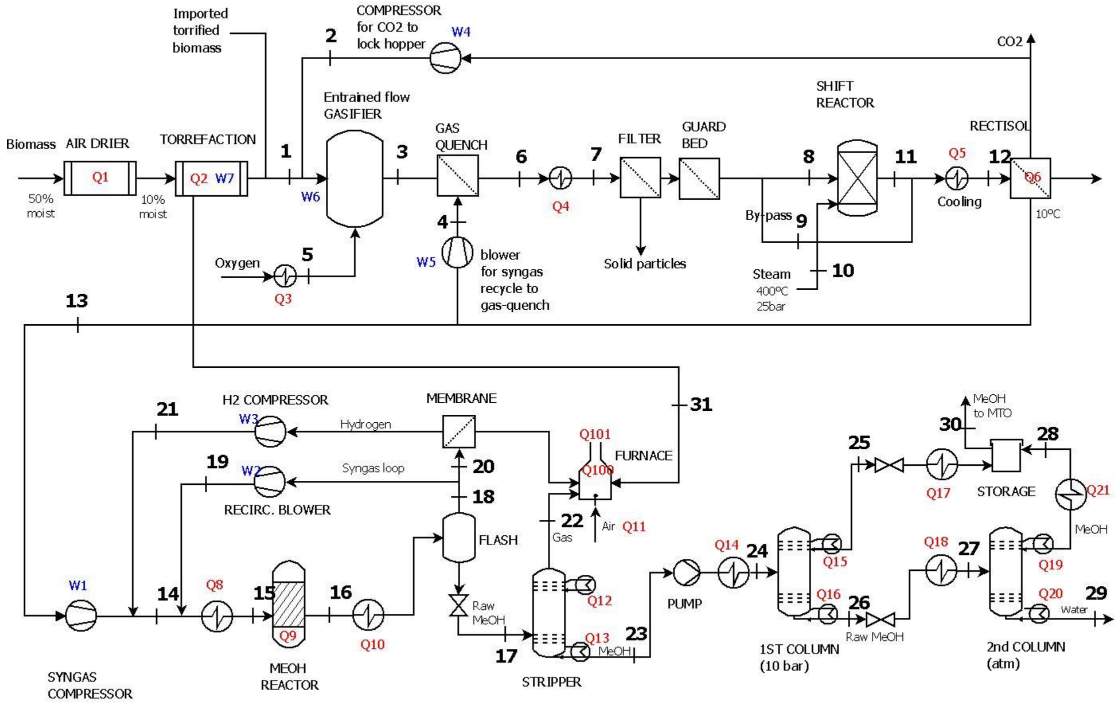
Heat sink/source	**Q1**	**Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q7**	**Q8**	**Q9**
Tin (°C)	15	886	300	1000	25	437	159	109	250
Tout (°C)	200	300	700	350	200	50	160	250	249
Heat load (kW)	1629	49620	33146	57798	466	83191	1008	26215	45141

Heat sink/source	**Q10**	**Q11**	**Q12**	**Q13**	**Q14**	**Q15**	**Q16**	**Q17+Q21**	**Q18**
Tin (°C)	250	25	54	66	91	137	143	65	143
Tout (°C)	50	200	18	66	120	136	144	40	40
Heat load (kW)	52315	460	1249	2279	3206	10478	21482	13957	2083

Heat sink/source	**Q19**	**Q20**	**Q100**	**Q101**	**Q102**	**Q103**
Tin (°C)	66	100	1000	1000	20	1000
Tout (°C)	65	101	999	150	70	150
Heat load (kW)	10095	10687	3651	3638	76978	23066

Power demands	Power (kW)
W1	9120
W2	1021
W3	735
W4	8783
Rectisol	3804
Biomass feeding	2250
ASU	11255
O2 compress to 30 bar	7053
Ancillaries	146

System configuration n.3



Biomass input:

WOOD CHIPS, 50% moisture, drier inlet: 63 648 kg·h^{-1} (150 MW LHV basis, 183 MW HHV basis)

Imported torrefied biomass, 1.8% moisture: 47 515 kg·h^{-1} (291 MW HHV basis)

Gasifier input (stream 1, torrefied biom.): 72 307 kg·h^{-1} (426 MW LHV basis, 452 MW HHV basis)

Stream	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**
Mass flow (kg·h⁻¹)	1 307	114 593	33 984	41 903	148 507	148 492	73 512	74 980
Temp (°C)	405	1 300	30	200	897	350	350	350
Pressure (atm)	35	30	35	30	30	30	30	30
Molar flow (kmol·h⁻¹)	30	5 471	2 951	1 318	8 423	8 422	4 169	4 253
Mol frac. (%)								
C	-	-	-	-	-	-	-	-
O_2	-	-	-	95.00	-	-	-	-
H_2O	-	9.63	0.03	-	6.27	6.27	6.27	6.27
H_2	-	27.21	64.93	-	40.43	40.43	40.43	40.43
N_2	-	1.34	1.54	5.00	1.41	1.41	1.41	1.41
S	-	-	-	-	-	-	-	-
Cl_2	-	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-	-
H_2S	-	0.01	-	-	0.01	-	-	-
CH_4	-	0.90	0.83	-	0.88	0.88	0.88	0.88
CO	-	54.24	29.63	-	45.62	45.62	45.62	45.62
CO_2	100.00	6.66	3.04	-	5.39	5.39	5.39	5.39
NH_3	-	-	-	-	-	-	-	-
Methanol	-	-	-	-	-	-	-	-

Stream	**10**	**11**	**12**	**13**	**14**	**15**	**16**
Mass flow ($kg \cdot h^{-1}$)	98 098	171 610	141 841	53 531	53 974	171 204	171 204
Temp (°C)	400	536	49	10	175	250	250
Pressure (atm)	30	30	30	30	89	89	84
Molar flow ($kmol \cdot h^{-1}$)	5 445	9 615	8 855	4 649	4 877	19 046	16 136
Mol frac. (%)							
C	-	-	-	-	-	-	-
O_2	-	-	-	-	-	-	-
H_2O	100.00	43.42	0.40	0.03	0.03	0.02	0.64
H_2	-	33.47	55.76	64.96	66.61	73.69	68.34
N_2	-	0.61	1.31	1.53	1.46	11.76	13.89
S	-	-	-	-	-	-	-
Cl_2	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-
CH_4	-	0.38	0.69	0.81	0.76	1.77	2.09
CO	-	3.85	25.41	29.61	28.22	10.13	3.55
CO_2	-	18.27	16.42	3.06	2.92	2.03	1.78
NH_3	-	-	-	-	-	-	-
Methanol	-	-	-	-	-	0.59	9.71

Stream	**17**	**18**	**19**	**20**	**21**	**22**	**23**
Mass flow (kg·h^{-1})	51 367	119 838	117 230	2 608	459	3 556	47 810
Temp (°C)	39	50	57	50	432	22	66
Pressure (atm)	1	84	89	84	90	1	1
Molar flow (kmol·h^{-1})	1 651	14 484	14 169	315	228	116	1 536
Mol frac. (%)							
C	-	-	-	-	-	-	-
O_2	-	-	-	-	-	-	-
H_2O	6.05	0.02	0.02	0.02	-	0.00	6.51
H_2	0.00	76.13	76.13	76.13	100.00	0.00	0.00
N_2	1.38	15.31	15.31	15.31	-	19.76	0.00
S	-	-	-	-	-	-	-
Cl_2	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-
CH_4	1.84	2.12	2.12	2.12	-	26.27	0.00
CO	0.45	3.91	3.91	3.91	-	6.38	0.00
CO_2	2.33	1.72	1.72	1.72	-	33.24	0.00
NH_3	-	-	-	-	-	-	-
Methanol	87.95	0.79	0.79	0.79	-	14.35	93.49

Stream	**24**	**25**	**26**	**27**	**28**	**29**	**30**
Mass flow (kg·h^{-1})	47 810	19 326	28 484	28 484	26 623	1 861	45 949
Temp (°C)	120	137	140	40	65	96	30
Pressure (atm)	10	10	10	1	1	1	1
Molar flow (kmol·h^{-1})	1 536	603	933	933	831	101	1 434
Mol frac. (%)							
C	-	-	-	-	-	-	-
O_2	-	-	-	-	-	-	-
H_2O	6.51	0.02	10.71	10.71	0.12	97.54	0.08
H_2	0.00	-	-	-	-	-	-
N_2	0.00	-	-	-	-	-	-
S	-	-	-	-	-	-	-
Cl_2	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-
H_2S	-	-	-	-	-	-	-
CH_4	0.00	0.00	0.00	0.00	-	-	0.00
CO	0.00	-	-	-	-	-	-
CO_2	0.00	0.00	0.00	0.00	-	-	0.00
NH_3	-	-	-	-	-	-	-
Methanol	93.49	99.98	89.29	89.29	99.88	2.46	99.92

Heat sink/source	**Q1**	**Q2**	**Q3**	**Q4**	**Q5**	**Q6**	**Q8**	**Q9**	**Q10**
Tin (°C)	20	250	25	897	485	159	88	251	250
Tout (°C)	70	251	200	350	25	160	250	250	50
Heat load (kW)	27 696	3 930	1 989	42 705	117 783	774	26 265	39 325	44 707

Heat sink/source	**Q11**	**Q12**	**Q13**	**Q14**	**Q15**	**Q16**	**Q17+Q21**
Tin (°C)	25	54	65	90	137	139	65
Tout (°C)	200	22	66	120	136	140	30
Heat load (kW)	2 403	2 028	3 054	2 698	11 120	12 126	2 931

Heat sink/source	**Q18**	**Q19**	**Q20**	**Q100**	**Q101**
Tin (°C)	140	65	97	1 000	1 000
Tout (°C)	40	64	98	999	150
Heat load (kW)	3 023	17 922	18 678	19 991	20 002

Power demands	Power (kW)
W1	6 103
W2	870
W3	726
Rectisol	3600
W4	139
W5	214

Grinding : 4.3 MW