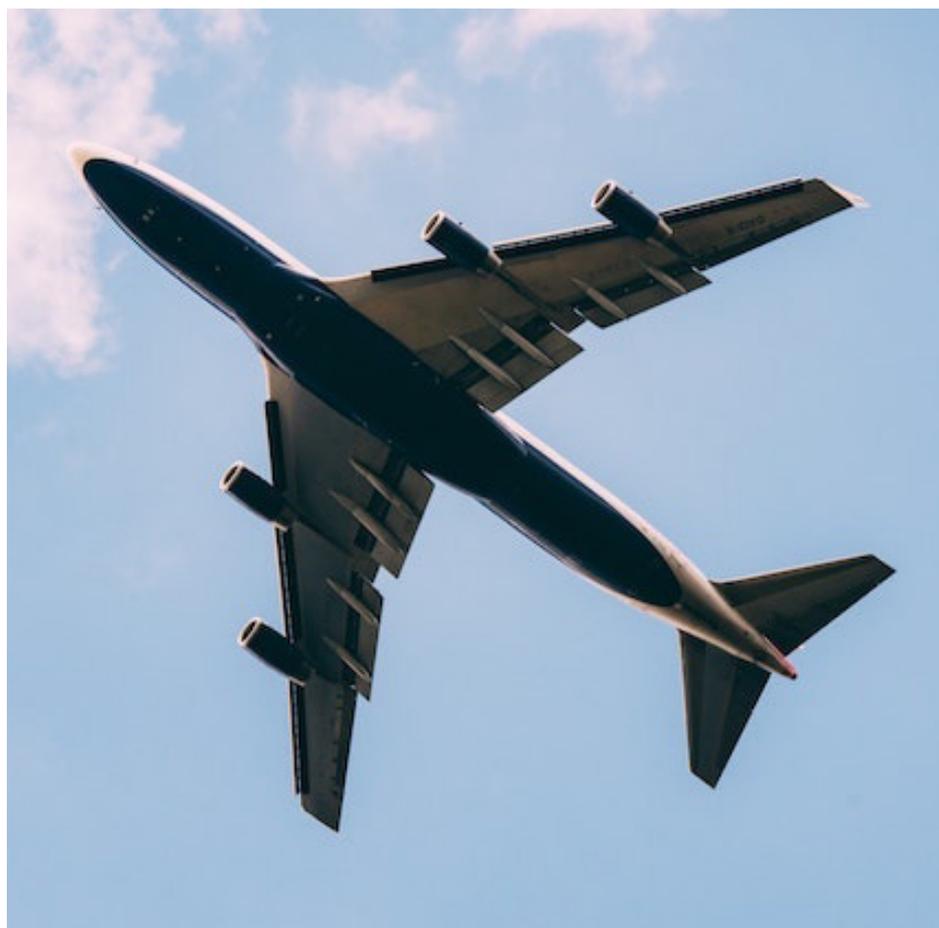
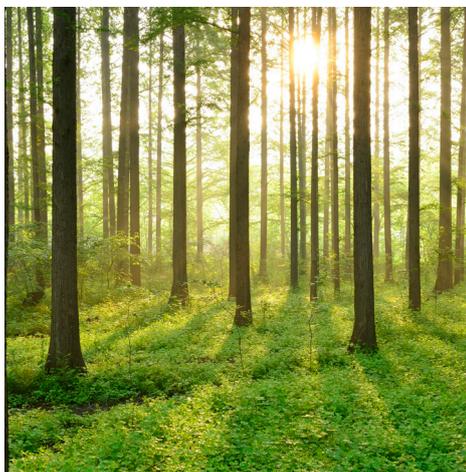


CO-GENERATION OF BIOJET IN CHP PLANTS

REPORT 2020:664



Co-Generation of BioJet in CHP Plants

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Foreword

The project has been conducted within the Energiforsk programme Biofuels for Sweden 2030 (Biodrivmedel för Sverige 2030), with the goal to contribute to the development of biofuels for the transportation sector and a fossil free transportation fleet by 2030.

The programme has been financed by EON Gas Sverige AB, Gasnätet Stockholm AB, Göteborg Energi AB, Neste AB and the Region Skåne.

Within this project, the feasibility of integrating a production of bio-based FT-crude, a possible intermediary to aviation fuel, with an existing CHP plant, has been investigated. The case study has been based on Karlstads Energi's biofuel based CHP-plant Hedenverket. In addition, it has been explored how the introduction of electricity as an additional feedstock can increase the efficiency and capacity of the proposed setup. The feasibility of using ash-rich particles from the combustion process in a separate guard bed has also been investigated, and the findings from this experimental study can be found in a separate report (Energiforsk 2020:665: "Removal of Sulphur species using industrial residual streams").

Anton Larsson, Gabriel Gustafsson and Christer Gustavsson (project leader) have conducted the project presented in this report. Emil Gammelgaard and Christian Hulteberg have conducted the experimental part of project, reported in the Energiforsk 2020:665.

The reference group for the project had the following members: Raza Naqvi, Karlstad University, Jan Brandin, Linnaeus University, and Bertil Wahlund/Anton Fagerström (Energiforsk). The reference group is gratefully acknowledged for invaluable contribution to the project.

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Bertil Wahlund Energiforsk AB

These are the results and conclusions of a project, which is part of a research programme run by Energiforsk. The author/authors are responsible for the content.

Sammanfattning

Kraftvärmeindustrin är en väl etablerad del av det svenska energisystemet. På grund av ett varmare klimat och hård konkurrens från värmepumpar upplever dock många aktörer i kraftvärmesektorn en situation där efterfrågan på fjärrvärme är stagnerande eller sjunkande. Dessutom har lägre intäkter från elproduktion gjort att lönsamheten för kraftvärme sjunkit ytterligare, både i fjärrvärmesektorn och skogsindustrin. Att etablera biodrivmedelsproduktion är en möjlighet för kraftvärmeindustrin att bredda sitt utnyttjandet av biomassa och addera en ny högvärdig produkt till sitt produktportfölj. Förnybar flygbränsleproduktion tilldrar sig stort intresse och kraftfulla incitament för att stödja sådan produktion är under framtagande.

I denna studie undersöks genomförbarheten i ett projekt för att bygga om en existerande fjärrvärmeanläggning för samproduktion av FT-vax, ett möjligt mellansteg för förnybart flygbränsle. Det föreslagna konceptet bygger på att bygga om en av anläggningens fluidiserade bädd pannor till en tvåbädds förgasare och sedan använda den producerade gasen som råmaterial till Fischer-Tropsch syntes. Utöver detta har studien även undersökt hur användningen av elektricitet som extra råvara kan öka kapacitet och verkningsgrad på den tänkta uppställningen.

En fallstudie har gjorts för Hedenverket ägt av Karlstads Energi. Där har en cirkulerande fluidiserad bädd panna, vanligtvis körd på mestadels grot undersökts. Produktionsmålet har varit 30 MW FT-vax med oförändrad fjärrvärmeleverans.

Ett flertal processupplägg har undersökts genom simulering. Det slutliga upplägget är baserat på FT-syntes vid låg temperatur med koboltkatalysator, återcirkulering av restgaser och en elektriskt värmd metanreforming. Omfattande värmeintegrering har genomförts för att ta vara på så mycket värme som möjligt för fjärrvärmeproduktion.

Som en separat del av projektet att analysera genomförbarheten för kraftvärmeintegrerad flygbränsleproduktion genomfördes en experimentell studie av att använda industriella biprodukter, som flygaska, grönlutslam och mesa för att ta bort de svavelämnena som bildas vid förgasning av biomassa. Resultat från dessa experimentella försök rapporteras separat i Energiforsk 2020:665: "Removal of Sulphur species using industrial residual streams."

Den slutliga anläggningen har en marginalverkningsgrad för FT-vax på 56 -75 % och en totalverkningsgrad på 78 – 100 % baserat på LHV beroende på pannlasten. Denna marginalverkningsgrad är avsevärt högre än vad som rapporterats för fristående anläggningar för FT-produktion [1]. Den totala investeringskostnaden har beräknats till 980 MSEK och den totala produktionskostnaden, exkl. kapitalkostnader, till 750-800 SEK/MWh. Den specifika investeringskostnaden såväl som produktionskostnaden är jämförbar med de hos avsevärt större anläggningar (200MW) [1]. I en LCA studie som utförts kopplat till projektet har klimatpåverkan hos produkten uppskattats till c:a 10 g CO₂-eq/MJ.

Summary

Biomass based Combined Heat and Power (CHP) is a well-established industry in Sweden. However, due to warmer climate and strong competition from heat pumps, many actors in the district heating sector are facing stagnating or declining demand for heat. Furthermore, falling revenues from electricity production has decreased the CHP profitability in both the district heating and forest industry. A growing demand and planned incentives for renewable aviation fuel forms an opportunity for the CHP industry to add a high value product to their present product portfolio and broaden their utilization of biomass.

This study investigates the feasibility of integrating a production of bio-based FT-crude, a possible intermediary to aviation fuel, with an existing CHP plant. The proposed route is through retrofitting one of the plant's boilers to a Dual Fluidised Bed (DFB) gasification setup in combination with Fischer-Tropsch synthesis. In addition, the study has explored how the introduction of electricity as an additional feedstock can increase the efficiency and capacity of the proposed setup.

A case study has been based on the Hedenverket CHP-plant owned by Karlstads Energi. The boiler studied is a Circulating Fluidised Bed boiler (CFB), operating on a fuel mix consisting mostly of forestry residue. The target production after the retrofit is 30 MW FT-crude with unaltered production of district heating.

Several process configurations for production of FT-crude has been investigated through simulation. The final process configuration proposed in this work is based on a low temperature Fischer-Tropsch synthesis with a cobalt catalyst, tail gas recirculation and an electrically heated steam reformer setup. Extensive heat integration has been performed to recover as much heat as possible for use as district heating.

As a separate part of the feasibility study, an experimental study of using industrial residual streams such as fly ash, green liquor dregs and lime for the removal of sulphur species derived from biomass gasification have been carried out. Results from these experiments are reported in Energiforsk report 2020:665. "Removal of Sulphur Species using Industrial Residual Streams"

The final plant has a marginal efficiency for FT-crude of 56 – 75% and a total efficiency of 78 -100% on LHV basis depending on the boiler load. This marginal efficiency is significantly higher than the energy efficiency of a standalone plant for FT-production [1]. The total investment cost has been estimated to 980 MSEK and the total production cost, excl. costs for capital, 750-800 SEK/MWh, these estimates are on par with those of significantly larger standalone plants (200MW) [1]. According to an LCA study made in conjunction with the project, the global warming potential (GWP), of the produced FT-crude is below 10 g CO₂-eq/MJ.

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1 Introduction

The importance of business development is highlighted by the currently stagnating or even declining revenues in the heat and power industry. A growing demand for renewable aviation fuel, supported by strong incentives, poses an opportunity for the CHP industry to broaden their utilization of biomass and add a high value product to their present product portfolio.

The concept of biomass firing for combined heat and power (CHP) production is well established in the Swedish district heating and forest industry sectors. In 2003, renewable power production became financially supported by means of green certificates. The already established concept then gained even broader industrial application in Sweden. However, in recent years the market conditions for biomass-based CHP have become less favorable due to low electricity prices. In parallel, the time-limited incentives for bio-based power production have expired for many actors and they have seen the revenue from power production drop significantly.

In addition to the falling revenue from power production, several actors in the Swedish DH business are facing stagnant or declining demand for heat. This is a consequence of warmer climate, an increase in energy efficient buildings and strong competition from alternative solutions such as heat pumps.

In the transport sector, biofuels are seen as an important tool for fulfilment of the climate goals in combination with other measures such as electrification. Especially in the aviation sector, where electrification is less likely to provide a comprehensive solution, biofuels are seen as key option. This is highlighted by the strong economic incentives supporting biofuels that are proposed to be implemented in the near future [2]. The incentives are tentatively aiming for a Swedish domestic Biojet utilization of 424 000 m³/a by 2030.

A growing demand for renewable aviation fuel in combination with a need for innovating the existing business models in large parts of the CHP sector brings the need for a wider perspective on biomass utilization. This purpose of this work is therefore to investigate the feasibility of co-production of biofuels by means of Fischer-Tropsch (FT) technology in existing CHP plants. The aims off current project are:

- A1. To investigate different technical solutions for co-production of FT-crude with CHP and find a viable solution for which the following should be described:
 1. Process outline
 2. Technical performance of the retrofitted plant including:
 - a. Production capacity and product composition
 - b. Consumption of feedstock and other consumables
 - c. Impact on power production
 - d. Life cycle assessment

3. Aggregated investments costs
 4. Operational and maintenance costs
- A2. To analyze the potential for utilization of the CO₂ produced by the process to enhance the production of FT-products. CO₂ can be utilized for production of electro- or electro-chemical fuel, and an aim is to investigate how it can improve the carbon utilization in this type of process. Furthermore, it will be investigated how such an implementation can affect the production cost.
- A3. To investigate how this type of process integration can affect the degree of utilization of CHP plants on a Swedish national scale.

Keywords

Biofuels, Co-production, Polygeneration, Fischer-Tropsch, Gasification, BioJet

2 Combined heat and power

Combined heat and power (CHP) operation is a mature, well established business. In Sweden as well as in other countries and regions rich in biomass, bio-based CHP plants significantly contributes to a sustainable energy system. Solid fuel boilers are extensively used in Swedish district heating systems and for industrial heat supply, especially in the forest industry. Presently about 50 TWh/a of unprocessed solid biomass, corresponding to some 10 million tons/a, is combusted in Sweden every year. The solid fuel boilers are spread over the entire country with most of them in densely populated areas and in regions with extensive forest industry operation.

Plants based on fluidized bed boilers are of special interest in the context of this report and they constitute the vast majority of the large-scale solid fuel boilers in Sweden, with approximately 100 units currently in operation. The total capacity of these boilers is approximately 6 GW_{th}. The majority (appr. 80%) of the boilers are of Bubbling Fluidized Bed (BFB) type whereas the remaining 20% are of Circulating Fluidized Bed (CFB) type.

The asset utilization, i.e. the actual output compared to the theoretical maximum for a full year, is generally low. In the district heating sector, where the utilization is often as low as 30-40%. In forest industry it is generally somewhat higher, but with large variations between individual mills.

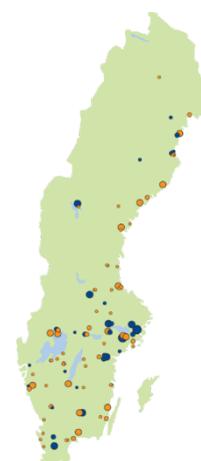


Figure 1: Fluidized bed Boilers in Sweden (2016)

2.1 FLUIDIZED BED BOILERS

Fluidized bed boilers are based on the principle of letting air percolate through a bed of granular material; commonly silica sand. When the gas velocity is high enough that the drag of the air counteracts the weight of the sand, this bed will start behaving in a fluid-like manner and it is then commonly referred to as fluidized. The fluidized bed can be divided into two separate phases, emulsion and bubbles. The concentration of bed material in the emulsion phase is very high, while in the bubbles it is very low. The paths made by bubbles rising through the bed cause continuous mixing of the particles. A bed material with a high thermal capacity improves mixing and provides a thermal inertia to the combustion process. Some of the particles will be entrained into the freeboard of the boiler, how many depends on the superficial gas velocity through the bed and fluidized bed boilers can be divided into two categories based on this: bubbling fluidized bed (BFB) boilers and circulating fluidized bed (CFB) boilers.

In a BFB the gas velocity is limited to keep the majority of the particles from leaving the reactor. In a CFB the velocity is higher to have the bed material entrained and transported by the gas to a higher extent. In a CFB particles and flue

gas are separated using a cyclone external to the main combustion chamber. Figure 2 illustrates a BFB and a CFB boiler.

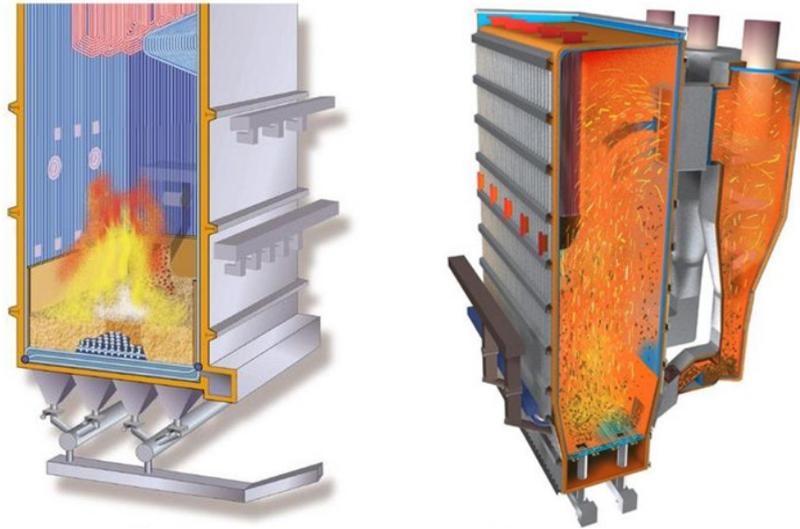


Figure 2: Illustration of a BFB and a CFB combustion chamber. Illustration by Valmet (former Metso)

3 Thermochemical conversion

Thermochemical conversion can be used for processing biomass into a variety of products. The outcome of this process depends on variables such as temperature, heating rate, reactants, active components, reactor design and pretreatment of the fuel. A thermochemical conversion process can be categorized based on the main product of the process as summarized in Table 1.

Table 1: Overview of a number of thermochemical conversion processes, indicating the main product for each of them.

	Solids	Liquids	Gas	Heat
Torrefaction	Main prod.	low	low	Required
Pyrolysis	low	Main prod.	low	Required
Gasification	low	low	Main prod.	Required
Combustion	low	low	high	Main prod.

As a simplification, thermochemical conversion of biomass can be divided into 3 steps:

1. Drying.
2. Pyrolysis (also referred to as devolatilization).
3. Secondary reactions of the char and volatiles.

When biomass is heated it will first dry as water is evaporated. When the biomass has dried the temperature starts to increase further, initiating pyrolysis. The pyrolysis and secondary reaction of the biomass are illustrated in Figure 2.

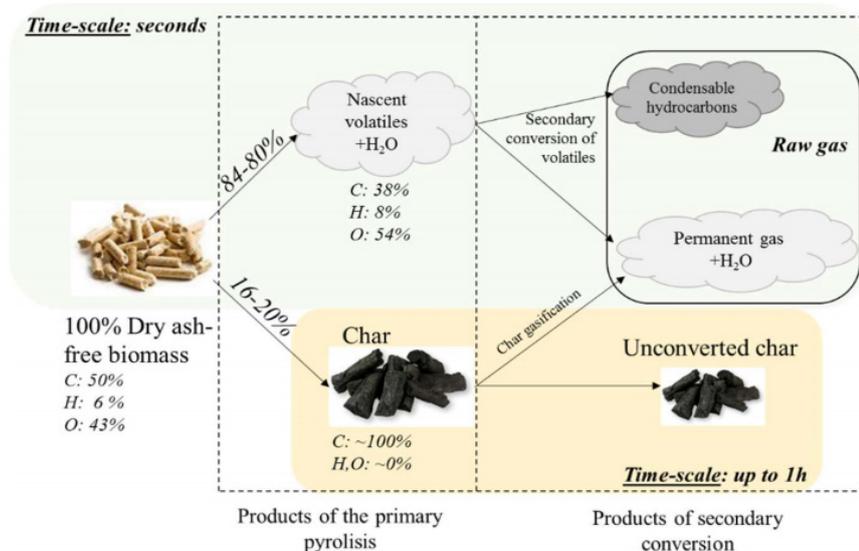


Figure 3: Thermochemical conversion of biomass, from [3]

Pyrolysis requires no reactants or active components, only heat, and forms char and volatiles. The yield of volatiles and char depends mainly on the peak temperature and heating rate but also on the type of biomass [4]. The yield of char, which can also be referred to as fixed carbon, decreases with higher temperature until about 600 °C. The heating rate also has a strong impact on the yield of char and with at low heating rate and low temperature the char yield can be maximized (as in a torrefaction process). The heat needed to dry and pyrolyze biomass can be provided through either secondary reactions or external heat sources. Depending on the reactants provided the secondary reactions can produce or require further heat.

The present work focuses on fluidized beds where the heating rate is high and, therefore, the char yield is limited to about 16-20% at temperatures above 600 °C. Increasing the temperature further will have no significant effect on the char yield but it affects the composition of the volatiles. The volatiles can be composed of hundreds of different components and these can be grouped into condensable components and permanent gases, referring to their state at ambient conditions. At high temperatures the yield high for permanent gases and low for condensable components, see Figure 3. However, a higher temperature also affects soot formation and the composition of the organic condensable components, commonly referred to as “tar” in gasification applications and “pyrolysis oil” in pyrolysis applications, further described below.

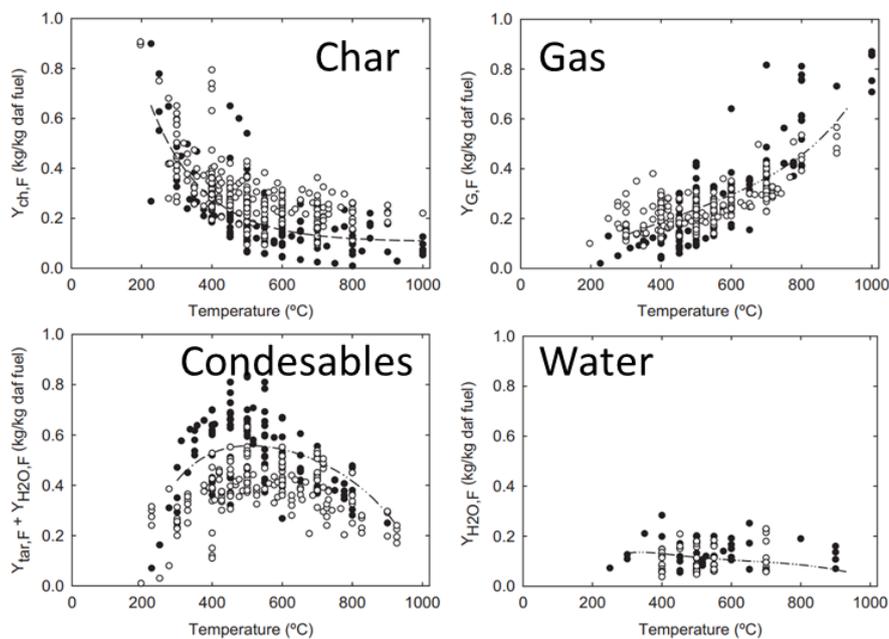


Figure 4: Yields during pyrolysis as a function of temperature, adapted from review by Neves et al [4]. Filled symbols indicate fast heating rate and open symbols indicate slow heating rate.

Both the solid char and the gaseous volatiles are further affected by secondary reactions. In combustion processes oxygen is introduced to oxidize both volatiles and char. The volatiles burn as a flame through gas-gas reactions while the char burns without a flame through a gas-solid reaction. In combustion high heat

production and complete conversion into CO_2 is desired. In gasification, as opposed to combustion, it is desired to produce a calorific gas that can be further utilized in downstream processes. The char can be converted through partial oxidation with O_2 , or through gasification by H_2O or CO_2 . Introduction of oxygen reduces the calorific value of the gas and should therefore be limited to amounts just sufficient to provide the heat required by the process.

The volatiles are comprised of hundreds of different components that can interact with each other and with added reactants. An important process is the reformation and polymerization of the organic components. Starting with oxygen rich compounds in the nascent volatiles, originating mainly from the cellulosic structure of the biomass, moving to less oxygen rich compounds and finally polyaromatic hydrocarbons (PAH), see Figure 5. Note that at high temperatures the PAHs continue to polymerize to also form soot particles.

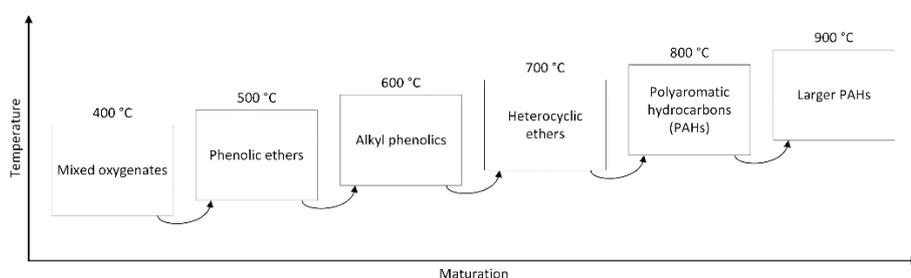


Figure 5: Maturation scheme for tar components in the volatiles from biomass pyrolysis as described by Elliott [5].

An important aspect of the change in yield and composition of the tar is the change in dew point, a crucial parameter for downstream process equipment. The downstream equipment should be designed to avoid fouling by tar and it is therefore important to understand how the operation of the thermochemical process affects the tar yield and composition.

In pyrolysis applications it is desired to have a high yield of pyrolysis oil, but it is also desired to limit the amount of oxygen in the oil to improve its stability and to decrease the complexity of the refining process. Thus, there is a trade-off between maximizing the yield and reducing the oxygen content.

In gasification applications, the focus of the present work, it is desirable to limit the tar yield and thereby increase the gas yield. However, as a high dew point can complicate downstream gas treatment polymerization may be undesirable even in this application. The dew point of different tar components is illustrated in Figure 5, showing that it is much higher for PAHs with 4 rings or more than for components with 2-3 rings. Thus, also in gasification applications there is a trade-off on how to operate the thermochemical conversion part of the plant.

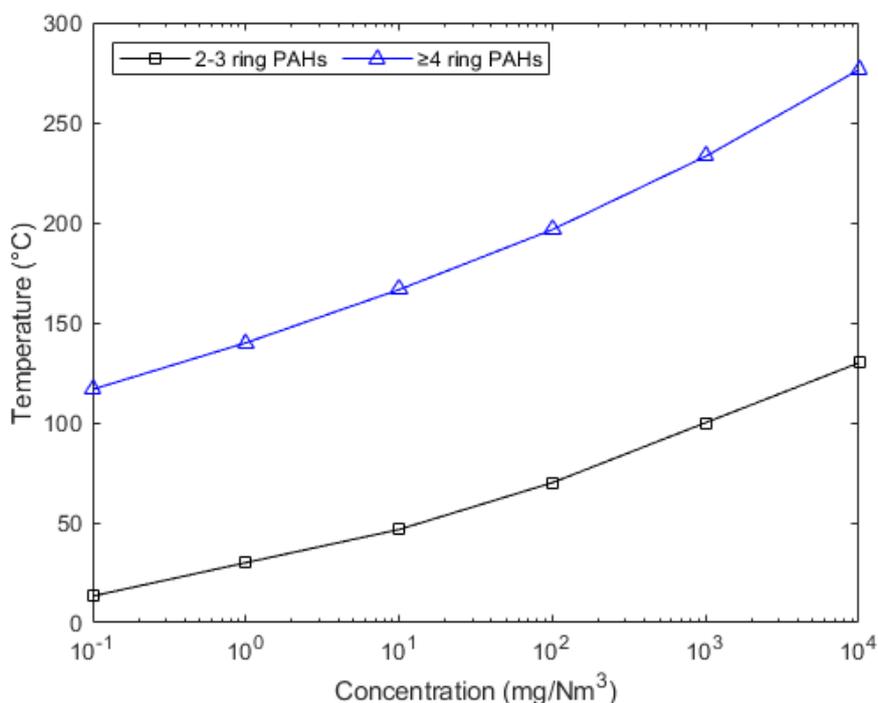


Figure 6: Examples of dew point of tar components as a function of the gas phase concentration for tar components with 2-3 aromatic rings compare to tar components with 4 or more aromatic rings. Based on data from [6].

Apart from the temperature and heating rate, several process specific variables will impact the yields of the thermochemical conversion process. These include reactants, active components and reactor design, which is further discussed in the following section.

3.1 DFB GASIFICATION

Present work is focused on the so-called dual fluidized bed (DFB) gasification technology and a concept for retrofitting FB boilers into DFB systems developed by BioShare AB. The basis of a DFB system is illustrated in Figure 7 where the thermochemical conversion is conducted in two interconnected FB reactors: a combustor and a gasifier. The gasification reactor is fluidized with steam and operated at about 800 °C. It produces a nitrogen lean, high calorific gas suitable for downstream synthesis processes. Part of the char is converted through reaction with the steam and part is transported with the bed material to the combustion chamber where it is burnt to produce heat. The heat is then transported back to the gasification reactor by the bed material, to provide the heat required for the fuel conversion in the gasifier. In this way the DFB technology omits the need for the char to be completely converted through the slow reaction with steam, as it is instead burnt to cover part of the heat demand. The technology also gives the means to provide a nitrogen lean gas without the need for pure oxygen. The benefit of a low N₂ concentration in the gas is further highlighted in section 7.

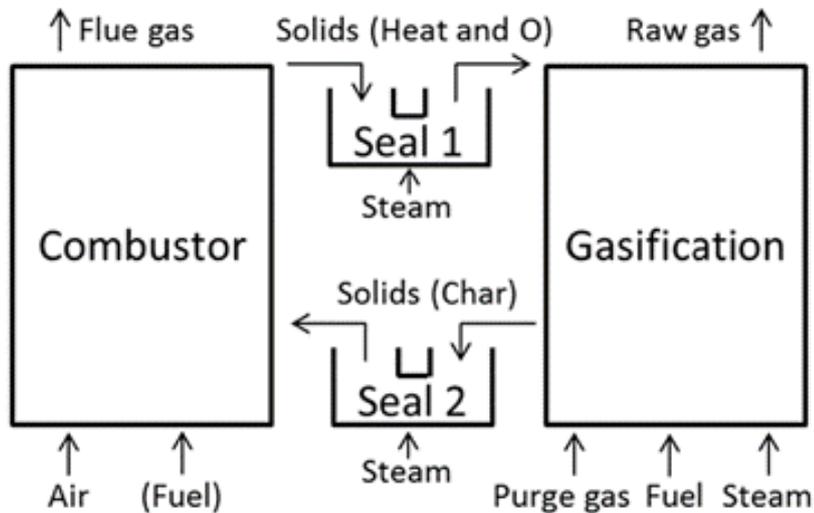


Figure 7: Schematic of a DFB gasification process [7].

The DFB gasification technology has been used commercially for CHP in several plants where the plants in Güssing [8], Oberwart [9] and Senden [10], with thermal inputs of 8-14MW, are the most well documented. It has also been demonstrated in an advanced biofuel production application in the GoBiGas project [11], which had a thermal input of 32 MW. All these plants were built as stand-alone units.

The feasibility of retrofitting an existing CFB boiler has previously been shown with the Chalmers 2-4 MW pilot gasifier [12].

Extensive research and technology development have been conducted at the above-mentioned plants and extensive amounts of data are available on fuel conversion and gas quality in these gasifiers, as well as how to operate them. Figure 8 illustrates an example of the energy balance from the GoBiGas gasifier operating with shredded bark as fuel. This shows that for a thermal input to the gasifier normalized to 100 by the LHV of the dry part of the fuel, about 19 % of this need needs to be transported with the bed material from the combustor to the gasifier. This is to be compared to a case where the GoBiGas plant was operated with wood pellets with 8% moisture and where the heat demand of the gasifier was less than 10% of the fuel input. This illustrates how much extensive drying of the fuel benefits a DFB gasifier.

The fuel moisture content also impacts the required product gas recirculation (PG rec) amount, which is much lower in the pellet case, contributing to a significantly higher chemical efficiency of biomass to cold gas. Note that the need for recirculation of gas is required in a stand-alone plant like GoBiGas, but not in a retrofitted CHP-plant, where instead a separate fuel feed is included. This also allows regulation of the heat production without changing the gas production.

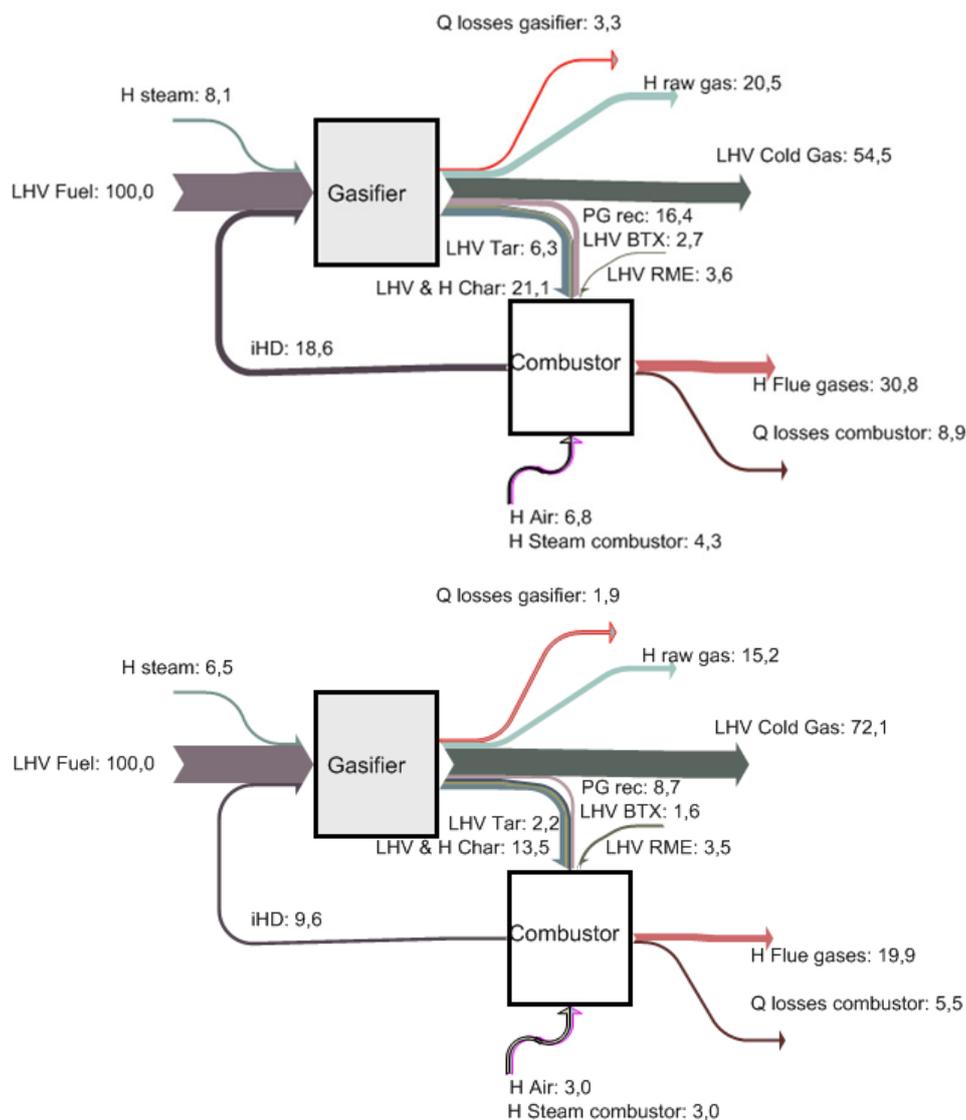


Figure 8: Sankey diagram of the GoBiGas DFB gasifier operating on bark dried to 25% moisture (upper) and wood pellets with 8% moisture (lower), from [13]. Number are normalized to a thermal input based on the LHV of the dry part of the fuel of 100.

Another important aspect of a DFB gasification system is the type of bed material used, and a range of bed materials have been tested and described in literature. A comparison of the char yield using different bed materials in the Chalmers gasifier was made by Berdugo [3], Figure 8. The tar yield is significantly higher when using silica sand than for the other investigated bed materials. The difference can be explained by the catalytic activity of the particles and their interactions with catalytic ash components. Olivine is the only bed material that has been applied in commercial DFB gasifiers so far, while silica sand is the bed material normally used in FB boilers.

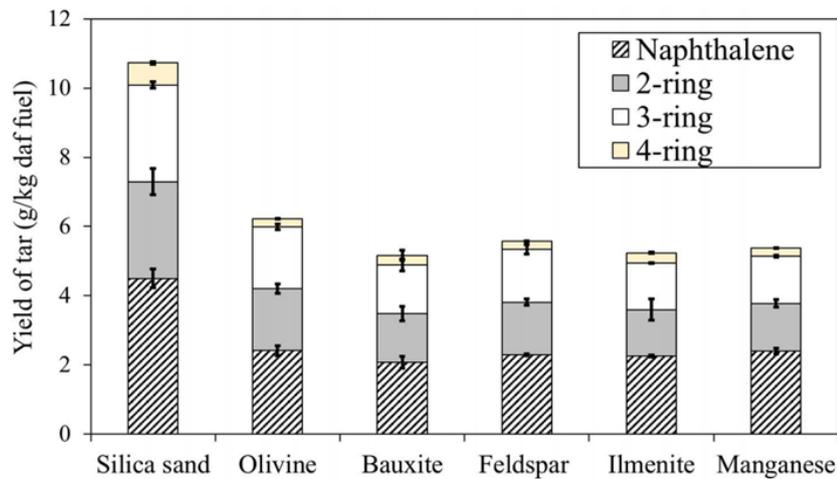


Figure 9: Tar yield from the Chalmers gasifier using different bed materials as summarized by Berdugo [3]

Another important aspect of the different bed materials is their ability to transport oxygen from the combustion section of the process to the gasifier, where it will combust part of the gas. This has been illustrated through the O/C and H/C ratio of the syngas components (H_2 , CO and CO_2) for the otherwise same case from the Chalmers gasifier [3], Figure 9. In this type of illustration high H/C and low O/C is attractive for downstream process and a very high O/C and low H/C ratio are the consequence of high oxygen transport. Both silica sand and olivine have a poor ability to transport oxygen and yield attractive gas compositions for the current application.

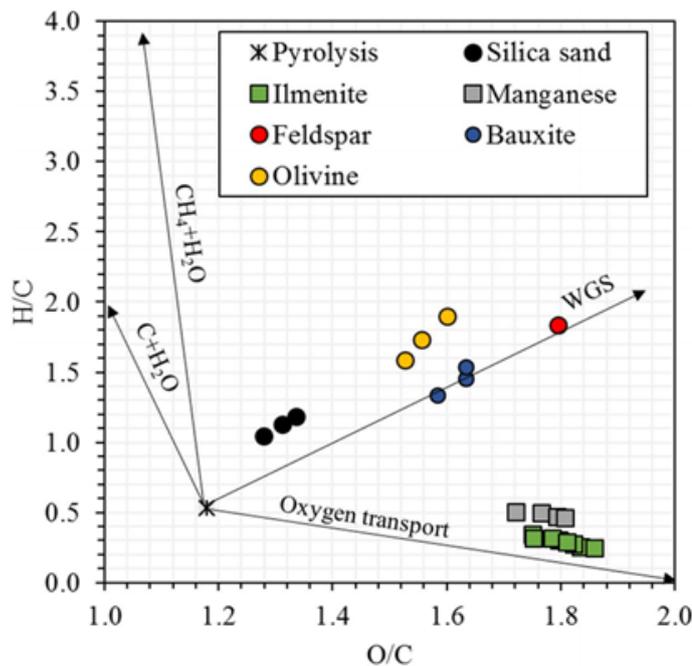


Figure 10: A Van Krevelen diagram based on the syngas components H_2 , CO and CO_2 with indications of how different reactions affects the coordinates of a pyrolysis gas reference. Results using different bed materials in the Chalmers gasifier are shown, from [3]

Which is the most beneficial bed material will be site specific, depending on aspects such as the technical solution for the retrofit, the need for heat production, the potential offset for tar components, and the selected downstream cleaning system.

4 Advanced biofuels

Biofuels produced from a lignocellulosic feedstock such as agricultural and forestry residues excluding food crops, or industrial waste and residual streams can fulfill the required GHG reduction as defined by EU's renewable energy directive (RED II). A full list of possible feedstocks for the production of advanced biofuel can be found in Part A of Annex IV of REDII.

Through gasification, biomass can be converted into a product gas which can be further treated to produce a gas that can be utilized for synthesis, commonly referred to as *syngas*. Through different catalytical processes and separation a syngas can be used to produce a range of different advanced biofuels, materials or chemicals. Figure 11 shows a number of alternatives of end-products and intermediaries where the production routes in focus of the present work are highlighted with bold text.

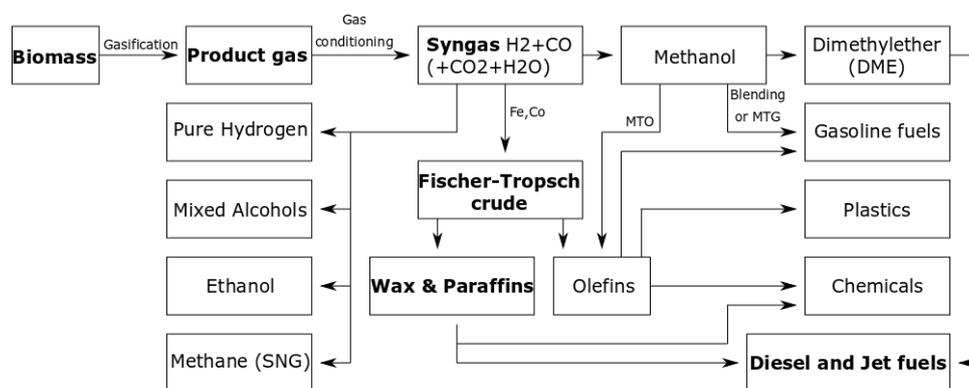


Figure 11: Examples of different production routs based on gasification of biomass, adapted from [14]

Production of advanced biofuels based on the DFB gasification technology was demonstrated in the GoBiGas project where 20 MW of CH₄ was produced from biomass. The GoBiGas plant was built as a stand-alone plant which caused a high capital and operational cost, finally causing the shutdown of the plant.

Co-producing advanced biofuels with heat and power by retrofitting existing CHP plants gives a potential to bring down both the capital and operational costs for this type of processes. Figure 12 shows how a block diagram of what such a plant could look like and indicates different control volumes (CV) used in the present work to describe the performance of the plant. CV1 incorporates the whole plant with all the incoming and outgoing streams. CV2 incorporates the gasification and primary gas cleaning. This boundary is the basis for the commonly used term *cold gas efficiency*. CV3 incorporates gas conditioning and synthesis and is used here to investigate the *cold gas to FT-crude performance*.

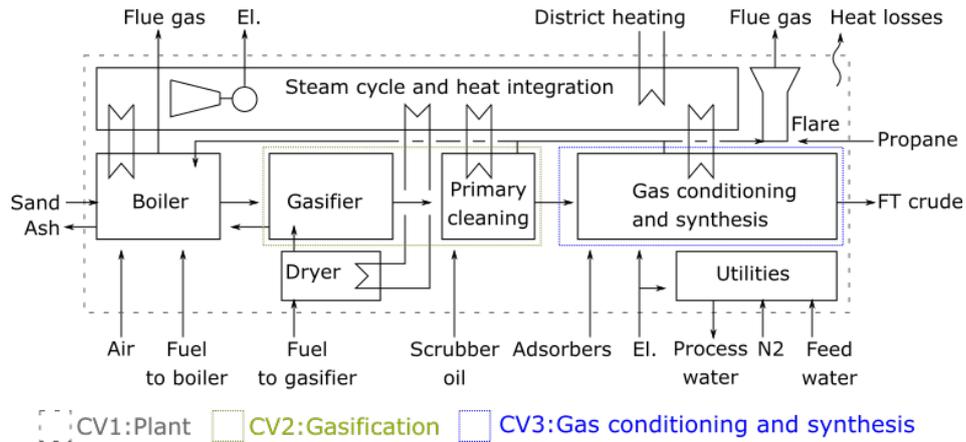


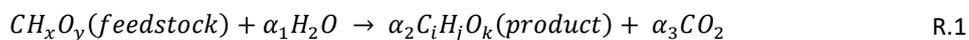
Figure 12: Block diagram of a co-production plant for CHP and FT-crude.

To describe and discuss the performance of this type of plant several different key indicators will be used, see Table 2:

Table 2: Terminology used to describe the performance of co-production of FT-crude with CHP plants.

Term	Notation	Description
“Carbon utilization”	μ_c	The degree of carbon reallocated into the desired stream in a CV
“Chemical efficiency”	η_{chem}	Describes the amount of chemically bound energy reallocated in a CV based on LHV.
“Carbon efficiency”	η_c	Carbon utilization compared to the theoretical max as described below
“Marginal efficiency”	η_{marg}	Used to describe the synergistic effect of adding another energy stream to the main process stream
“Electro-Chemical efficiency”	$\eta_{el-chem}$	The amount of chemical and electrical energy reallocated into chemical energy in a CV
“Total energy efficiency”	η_{tot}	Includes all forms of energy of a CV

The conversion of biomass or gas from gasification of biomass into advanced biofuels can be described by the following global reaction:



where α_{1-3} are the molar numbers required to balance the reaction, x , y , i , j and k are molar concentrations in the feedstock and product. The reaction is normalized with the carbon concentration in the feedstock.

The theoretical degree of carbon utilization, $\mu_{C,theo}$, given by reaction R.1 can be estimated based on the composition of the feedstock and product [7]:

$$\mu_{C,theo} = \left(1 - \frac{1}{2} \left(\frac{O}{C} \right)_{feed} + \frac{1}{4} \left(\frac{H}{C} \right)_{feed} \right) / \left(1 - \frac{1}{2} \left(\frac{O}{C} \right)_{prod} + \frac{1}{4} \left(\frac{H}{C} \right)_{prod} \right) \quad Eq.1$$

where (O/C) is the oxygen to carbon molar ratio and (H/C) is the hydrogen to carbon molar ratio. The abbreviation *feed* stands for feedstock and *prod* for product.

A theoretical chemical efficiency, $\eta_{chem,theo}$, can be estimated based on the theoretical degree of carbon utilization:

$$\eta_{chem,theo} = \mu_{C,theo} \frac{LHV_{C,prod}}{LHV_{C,feed}} \quad \text{Eq.2}$$

where LHV_C is the lower heating value per unit of carbon [MJ/kgC].

Assuming that sufficient heat and H₂O are available in all steps, and a cold gas with the properties (O/C)_{feed} = 1.01, (H/C)_{feed} = 1.84, and $LHV_{C,feed} = 37.24$ as feedstock the theoretical performance over CV3 estimated as stated in Table 2. It shows that, based on this product gas composition about 50-65% of the carbon can theoretically be reallocated as product while about 35-50% of the carbon will form CO₂. This corresponds to a theoretical chemical efficiency of about 90% for all the included end products.

Table 3: Estimate of the theoretical carbon utilization and efficiency for different products based on Eq.1 and Eq.2

	FT	CH ₄	MeOH	DME
H/C	2.08	4	4	3
O/C	0	0	1	0.5
$\mu_{C,theo}$	0.63	0.48	0.64	0.64
LHV_C	43.4	67	53	55
$\eta_{chem,CV2,theo}$	0.98	0.86	0.91	0.94

In a real process there are several aspects that will affect the carbon utilization and chemical efficiency, including:

- Combustion of some of the gas to cover a heat demand, by addition of O₂ or extraction of gas.
- Bleed of some of the gas to avoid accumulation of inert gases.
- Losses due to imperfect selectivity in separation steps such as H₂O and CO₂ separation.
- Removal of undesired components such as trace tar components.
- Addition of hydrogen or hydrogen rich streams to the process.

These aspects can be incorporated in Eq.1 as a variable for streams lost from the main process, ϑ_{lost} , and a variable of added streams, ϑ_{add} :

$$\mu_{C,theo} = \frac{\left(1 - \frac{1}{2} \left(\frac{O}{C}\right)_{feed} + \frac{1}{4} \left(\frac{H}{C}\right)_{feed} - \vartheta_{lost} + \vartheta_{add}\right)}{\left(1 - \frac{1}{2} \left(\frac{O}{C}\right)_{prod} + \frac{1}{4} \left(\frac{H}{C}\right)_{prod}\right)} \quad \text{Eq.3}$$

where the variable for the streams lost or added to the main process can be quantified as:

$$\vartheta_j = \frac{\sum \left(\dot{n}_{C,i} - \frac{1}{2} \dot{n}_{O,i} + \frac{1}{4} \dot{n}_{H,i}\right)}{\dot{n}_{C,feed}} = \frac{\sum \left(1 - \frac{1}{2} \left(\frac{O}{C}\right)_i + \frac{1}{4} \left(\frac{H}{C}\right)_i\right) \dot{n}_{C,i}}{\dot{n}_{C,feed}} \quad \text{Eq.4}$$

where j represents the *lost* or *add* term, and i represents the components lost or added.

Comparing the actual carbon utilization with the theoretical one can express a carbon efficiency of the process:

$$\eta_C = \frac{\mu_C}{\mu_{C,theo}} \quad \text{Eq.5}$$

When adding a stream to the main process it is interesting to quantify a marginal efficiency that illustrates the kind of leverage one can expect from synergistic effects. The marginal efficiencies are estimated here as:

$$\eta_{marg} = \frac{\dot{Q}_{prod} - \dot{Q}_{prod,ref}}{\dot{Q}_{add}} \quad \text{Eq.6}$$

where \dot{Q}_{add} is the energy in streams added to the reference case. When electricity is added to the process it is also of interest to quantify the electro-chemical efficiency:

$$\eta_{el-chem} = \frac{\dot{Q}_{prod}}{\dot{Q}_{feed} + P_{el}} \quad \text{Eq.7}$$

The total efficiency includes all valuable outputs of a co-production plant including the liquid product as well as electricity and district heating, but it also includes all mass and energy input (including the main feedstock) as well as electricity and auxiliary feedstock, $\dot{Q}_{aux,feed}$, such as scrubbing oil and fuel fed directly to the combustion reactor:

$$\eta_{tot} = \frac{\dot{Q}_{prod} + P_{el} + \dot{Q}_{DH}}{\dot{Q}_{feed} + P_{el} + \sum \dot{Q}_{aux,feed}} \quad \text{Eq.8}$$

5 Unit operations in production of FT-crude in retrofitted CHP-plants

This section is focused on unit operation aspects and synergies that arise from different technology choices when incorporating the production of FT-crude with a CHP plant.

5.1 BIOMASS DRYING

Previous work has shown that using low-grade heat for drying the biomass is an efficient measure to improve the chemical efficiency of a stand-alone advanced biofuel production unit based on DFB gasification [15]. Drying the fuel greatly reduces the heat demand in the gasification and thus also reduces the required bed material flow between the reactors in a DFB system. Further, a dryer can help amend variations in the process by giving the fuel a more consistent moisture content, which is beneficial for this type of process.

Belt driers are a mature technology and well suited for utilizing low temperature heat. They cause low gaseous emissions, have low fire hazard and high robustness [16][17] and were therefore the one of the evaluated technologies for biomass drying in this work. Under certain conditions, especially in the district heating sector, other drying configurations with better conditions for heat recovery may be a more energy efficient solution and hence a strong alternative to the previously described bed drying solution.

To achieve a very low moisture content, the size, heat demand, complexity and thereby cost of the dryer and downstream equipment becomes higher. The most cost-efficient moisture content will therefore be site specific and depend on the area and low-grade heat available at the site. For this specific project 15 % moisture in the fuel was chosen.

5.2 DFB GASIFICATION

Fluidized bed combustion plants commonly use silica sand as bed material and using silica sand in the retrofitted setup minimize the changes required for the retrofit. However, the choice of silica sand puts constraints on the temperatures in the system and 800 °C can be assumed for the gasifier bed, while 850 °C can be assumed for the bed of the combustor. Further, silica sand cannot be activated to any large extent by the fuel ash. The limits in activation level and temperature limits the fuel conversion in the gasifier with low char conversion and high tar concentrations.

In stand-alone DFB gasifiers olivine has most commonly been used as bed material. The olivine reduces the risk of agglomeration and enables gasification temperatures of up to about 870 °C. Olivine can be activated by catalytical ash components to improve the fuel conversion, reducing the tar yield and increasing the char conversion.

Depending on the feedstock operating parameters, bed material and activation level the cold gas efficiency, commonly used to describe the chemical efficiency of the conversion of fuel to cold and tar-free gas, $\eta_{chem,CG}$ is in the range of 50 - 75 %. DFB gasifiers are usually operated at slight sub-atmospheric pressure to limit the risk of any gas leakage. An estimated gas composition is summarized in Table 4.

Table 4: Summary of the estimated concentrations in the raw gas from the gasifier using silica sand as bed material.

Component	Concentration	unit	comment
H ₂	23.2-26.7	mole%	
CO	17.4-19.8	mole%	
CO ₂	14.5-17.6	mole%	
CH ₄	6.7-7.2	mole%	
C _x H _y	2.9-3.4	mole%	Mainly C ₂ H ₄
H ₂ O	27.7-29.1	mole%	
BTX	3500-7000	ppm	Mainly Benzene
Tar	3300-7500	ppm	Naphthalene and larger components
N ₂	700-7000	ppm	N ₂ formed from fuel plus and assumed amount from purge gas and other sources.
Other trace components	1900-6500	ppm	Further described in section 5.3

5.3 GAS COOLING AND CLEANING

The raw gas from thermochemical conversion of biomass contains a number of impurities or undesired by-products that can cause problems in downstream equipment. This is a brief summary regarding different impurities and by-products and the methods considered here for their removal. Commercial processes are available for this type of gas treatment, in plants for Fisher Tropsch (FT) synthesis based on coal gasification. However, these plants operate at a scale not relevant for biomass-based production. Besides the difference in scale and fuel, different fuel conversion technologies are used. These differences need to be considered for the design of the production of FT-crude via gasification of biomass. In this section, the gas cleaning and conditioning process steps following a dual fluidized bed (DFB) reactor for syngas production are considered.

Which process that is the most convenient method for handling the by-products/impurities also depends on the set-up of the refinery in terms of downstream technology choices and integration level of the gasification reactor. There are three categories of biorefineries for production of FT-crude based on BioShare technology:

- Gasification with aromatics separation
- Gasification with Partial Oxidation (POX)
- Activated gasification with aromatics separation

Gas cleaning methods can be categorized based on the operational temperature as well as if it is dry or wet. Here, processes operated below about 250 °C are referred to as “cold” gas cleaning while processes operating at higher temperatures are

referred to as “hot” gas cleaning. “Wet” gas cleaning steps employ spray and wash towers, while “dry” gas cleaning is performed without any addition of water or solvents.

5.3.1 Solids

The raw syngas from a gasifier will contain a variety of particles originating from the bed material and fuel including ash, soot and char particles. The ash comes mainly from the fuel but can also be transported between the fluidized beds with the bed material.

Partial hot removal of particles can be achieved with cyclone separation or other particle disengagement at the gas exit from the gasification reactor. These methods can remove the bulk of larger particles but will require further particle removal downstream. These methods would probably be applied at a temperature close to the freeboard of the gasifier, at which alkaline metal components can have significant partial pressures and therefore not be separated. The risk of clogging the equipment with either ash or tar is less than for filters at lower temperature. Further, no purging with inert gas (N_2 or CO_2) is required.

Hot gas filters achieve satisfactory removal of particles and by operating them at temperatures of 400-500 °C the ash components can be removed as solid compounds. If present in the gas, some trace components such as KOH, HCl can be in gas phase while metal and alkaline earth metal components can be assumed to be satisfactorily removed. An exception is mercury, but this is rarely present in any significant amount in the feed. Disadvantages with hot gas filters are that they are bulky and expensive and not commonly applied.

Cold gas filters, such as the ones used with the Repotech gasifiers are bag filters with limestone precoating operated at about 200 °C. This mode of operation may be possible for applications with a rather low dew point of the tar such as with activated olivine or low temperature operation with concentrations of large PAHs. However, the risk of clogging the product gas cooler is significant and forms a common challenge for these plants. The benefit of cold gas filters is a more compact solution with lower risk of slip of ash component traces. In this setup the filter also exhibits an initial tar removal where heavy tar components are captured together with the particles. However, this type of particle-removal should mainly be considered in combination with an activated gasification process.

5.3.2 Gas cooling

The gas leaving the gasifier is hot and it is desirable to recover some of the heat. The technology used for cooling the gas also needs to be considered in combination with the particle removal. The technology options can be categorized as:

- Water-tube boiler (radiative)
- Coils (radiative and convective)
- Tube heat exchanger (mainly convective)
- Quench

A water-tube boiler has cooled walls and a low gas velocity where the gas-surface contact is minimized, and the heat mainly transported through radiation. This is a bulky solution but one that reduces the risk of clogging due to ash or tar fouling. This option is commonly used for the initial cooling of flue gas from biomass-based combustion, where fouling from fuel ash is a challenge.

Coils where the cooling media is transported inside the tubes and heated by the passing gas are less bulky than the pure radiative solution. However, there is also a higher risk of fouling from ash components. Such coils are usually used in biomass-based combustion after radiative cooling has taken the temperature down to around 250-500 °C depending on the amount and composition of the ash in the fuel.

In a tube heat exchanger, the hot gas passes through a multitude of narrow (vertical) tubes, cooled on the outside where heat transfer through convection is significant. This enables an efficient and compact design, but it is also sensitive to fouling and is best suited for clean gases. With particles in the gas (or forming in the gas during cooling) a high gas velocity has to be maintained to avoid that particles settle on the cold surface of the tubes.

The temperature of the gas can also be reduced by quenching. This can be effective and low risk, but it also means that the heat cannot be utilized at its highest temperature. In a water quench, liquid water is sprayed into the gas so that the heating and evaporation of the water draws heat from the gas. The quench also dilutes the tar which could reduce the dew point slightly but increase the amount of process water produced.

In a gas quench, cooled gas is recycled and mixed with the warm gas. This is less effective than water and increases the volume flow through parts of the process significantly, increasing the investment cost. Furthermore, the concentrations of tars are reduced.

5.3.3 Tar removal

Tar and carbon-containing solids can either be removed or converted in a secondary conversion step. The tar can either be handled with a cold cleaning process, scrubbing, or a hot secondary conversion process.

A secondary conversions step can be a catalytic process such as used at the Skive gasifier or a process based on thermal cracking, such as a partial oxidation (POX) unit. Catalytical processes requires low concentrations of especially sulfur in the feedstock and is generally not for other feedstocks than stem wood. In a POX unit, oxygen is added to burn some of the gas and raise the temperature to thermally crack or steam reform tar and lighter hydrocarbons. Just heating the gas will enhance polymerization and soot formation but by introducing oxygen for partial oxidation and with the right process design and operation, a significant part of the tar can be cracked[18].

Water is commonly used to scrub the gas from coal gasification, but it has a poor removal efficiency, especially for heavy tar components. The removal can be as low as 50% for large refractory tar components, while heterocyclic tar components are

removed more efficiently [19]. The performance can be improved by including a wet electrostatic precipitator (WESP) downstream of the scrubbing system, which increase the removal efficiency of especially heavy tar significantly [19].

More efficient scrubbing can be achieved by using a solvent such as scrubbing oil instead of water. With partial particle removal upstream of the scrubber, particles will have to be removed from the solvent before recirculated. Still a WESP can also separate particles if any gets past the scrubber.

The scrubbing system should be designed to avoid that tar components turn sticky or solid. This depends on a range of design and operating parameters such as the temperature, solvent, packing material and make-up flow or regeneration of the solvent.

5.3.4 Nitrogen trace components

The nitrogen compounds in the gas comes mainly from the nitrogen in the biomass. In the reducing atmosphere of the gasifying process mainly ammonia (NH_3) will form, as well as low concentrations of hydrogen-cyanide (HCN).

Based on investigation of the fuel nitrogen conversion in a dual fluidized bed gasifier by Wilk and Hofbauer [20] more than 98 Wt% of the fuel nitrogen in the gasifier are present in the gas.

N_2 can be considered inert to downstream processes, however as recycling of gas is commonly used in FT-processes N_2 will accumulate, diluting the reactants and swelling the gas flows. Therefore, some gas always needs to be bled of, and depending on the reforming technology the concentration of N_2 might be the limiting factor for how low the gas bleed may be. This is further discussed below.

Nitrogen containing tar components can be found in low concentrations and are assumed to be removed with the other tar components.

A conventional technology to remove remaining nitrogen compounds is water scrubbing as they have good solubility in water. A separation in the order of 95% of the NH_3 can be achieved [19].

The high solubility of the nitrogen compounds in water also means that these components will dissolve in the condensed water (as much as 90% of the NH_3), complicating the treatment of the process water. Many tar compounds are toxic for nitrifying organisms commonly used for water treatment and if required to use such organisms tar and water should preferably be removed from the gas separately. An option is to spend heat to evaporate or strip some of the water, nitrogen compounds and light aromatic compounds from the process water and introduce this stream to the combustion process for destruction. Note that it is important to find an appropriate temperature in the combustion process for where to add this stream, to avoid increasing the level of NO_x or NH_3 in the flue gas.

Activated carbon could also be used to adsorb both NH_3 and HCN from the gas. As the gas also contain sulfur components and hydrocarbons there will be a competition for the active sites on the carbon. A detailed study in collaboration with the manufacturer is required to determine an appropriate operational

temperature and if chemical modification of the activated carbon is required. Activated carbon beds are however rather bulky and expensive.

HCN can also be converted through hydrolysis to NH_3 but might not be appropriate to do in the same reactor as for hydrolysis of COS as the catalyst for that can be sensitive to NH_3 .

There are investigations looking at hot treatment of NH_3 based on catalytical decomposition, but these are generally based on Ni-catalysts which are sensitive to sulfur poisoning. This is therefore not viewed as a relevant option at present.

5.3.5 Halide trace components

Halides like Cl is prone to form salts with K or Ca if present in the gas, the salts can then be removed as solids. Trace components of HCl could however form and would end up in the water or in guard beds. ZnO beds is sensitive to Cl but can be protected from Cl by including an Al-based bed upstream of it.

Most of the chlorine in the fuel is not volatilized. Wilk and Hofbauer [20] estimated that 6.1% of the Cl in the fuel ended up as HCl while the rest ended up as ash.

5.3.6 Sulphur trace components

The sulfur in the biomass mainly forms H_2S and traces of COS and perhaps sulfur containing hydrocarbons such as thiophene. Sulfur conversion in gasification is not well documented in literature where the common statement is that H_2S dominates but there are also minor amounts of COS, Thiophene(s), Mercaptan(s), and C_2S .

Several studies on FT production via gasification neglect COS, for example NREL [21] assumed that it was so low that it did not need to be considered when designing the gas cleaning. At GoBiGas, Hydrogenation, MDEA based scrubber and sulfur guard bed was installed to remove sulfur from the gas. However, the concentration of sulfur containing component upstream of the MDEA-scrubber, downstream of the active carbon beds used there to remove BTX was very low, raising the question if the H_2S scrubber was excessive for that plant [11].

Sulfur components can be removed by dry or wet methods. Wet methods are usually restricted to cold conditions while dry methods can be used at elevated temperatures. Wet methods enable for a continuous regeneration and are therefore preferred in processes with a relative high concentration of sulfur in the gas. A drawback of wet solutions to remove sulfur is that it may require such low temperatures that a refrigerated process is required. Commercial refrigerated wet sulfur removal has been applied at very large scale, such as Rectisol or Selexol. However, these are complex processes and are not considered for biomass-based systems.

Amine based solvents can be used for bulk removal at about ambient temperatures. To reach very low concentrations the amine removal system can be combined with a downstream dry method in the form of guard beds. An amine system can also be designed for co-adsorption of COS and CO_2 .

Dry removal can be achieved using activated carbon, metal impregnated activated carbon or metallic beds such as ZnO. H₂S can also be adsorbed by ash-coated particles in the process or filter pre-coat and thereby partially removed in the particle removal. In the present project the feasibility of using ash-rich particles from the combustion process in a separate guard bed has been investigated. The findings from this experimental study can be found in a separate report (Energiforsk 2020:665: "Removal of Sulphur species using industrial residual streams").

ZnO is optimally used at temperatures of 400 C where its capacity is highest but can also be used at lower temperatures. The ZnO is a catalyst for the WGS reaction and it is therefore appropriate to operate the guard bed at a temperature where the WGS reaction is limited. Note that if there is organic sulfur in the gas hydrolysis should be applied upstream of this type of guard beds to convert COS and other organic sulfur containing components into H₂S. Further, to polish the gas after the ZnO additional guard material, for example used FT-catalyst or Ni-based material, can be used to achieve very low concentrations.

Calcium components such as CaO, CaCO₃, dolomite or ash rich particles can also be used to remove H₂S which means that calcium containing compounds captured by particle filters or added to the filter as pre-coat can also remove some of the H₂S as well as HCl from the gas inside of the gasification process, particle filters or in separate guard beds.

5.3.7 Selected approach for gas cleaning

Based on the estimated level of contaminants the gas cleaning steps summarized in Figure 12 was assumed for present work. Potential simplification of the sulfur removal is discussed in section 6. The present setup offers an efficient removal of N, Cl and especially S contaminants to protect downstream catalysts while remaining BTX components and olefins are handled by the pre-reformer as described below.

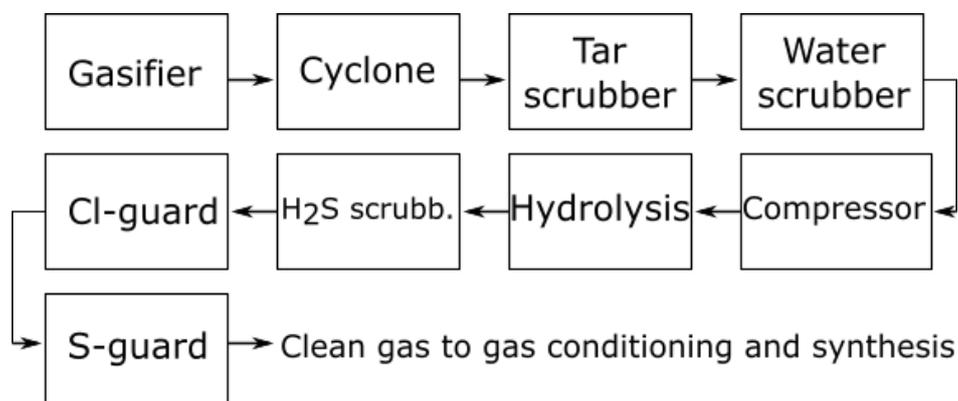


Figure 13: Gas cleaning steps included in present analysis.

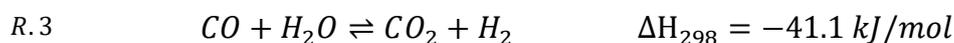
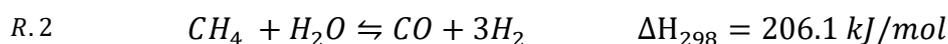
5.4 GAS CONDITIONING

Before the cleaned gas can be used for synthesis its composition has to be adjusted further. More specifically, the methane and higher hydrocarbons in the incoming gas should be broken down into hydrogen and carbon monoxide while still preserving a suitable ratio between the two. In addition, excess carbon dioxide and water is selectively removed to minimize the size of the FT-reactor. This process part is also needed to take care of the byproducts formed in the FT-synthesis.

5.4.1 Steam Methane Reforming (SMR)

Steam reforming is a process used to produce a hydrogen rich syngas from a hydrocarbon feedstock such as natural gas or naphtha together with steam. It is the most common process for hydrogen production over several different industries. When speaking about a “hydrogen production unit” in a process plant, what is referred to is often a subprocess with a steam reformer at its heart. In this project steam reforming is used to convert methane present in the cold gas and byproducts of the FT- synthesis into useable syngas.

The reaction mechanisms in a steam reformer are relatively complex, for a more detailed description the reader is referred to a book chapter by J.R Rostrup-Nielsen [22] and articles by Xu and Froment [23], [24]. For this report a simplified description is made, considering only two global equilibrium reactions.



The first reaction (R.2) is the methane steam reforming reaction, its reverse being the methanation reaction. As visible from the reaction enthalpy, ΔH_{298} , the steam reforming reaction is a highly endothermic reaction. The second reaction, (R.3) is the water gas shift reaction, being slightly exothermic in its forward direction. Note however that, under the conditions found in a reformer, the WGS reaction can proceed in both directions depending on the feed composition. The combined equilibrium of these two reactions can be said to govern the resulting conversion and product composition out of the reformer. Figure 14 shows the conversion at equilibrium as a function of temperature. As visible from Figure 14, to achieve a high conversion of methane the gas must leave the reformer at a relatively high temperature. The exact exit temperature for a given process will vary depending on feedstock and desired conversion. An example from an ammonia plant where the reformer is run on natural gas is ca 1073 K [22]. However, in a process such as for producing reducing gas, exit temperatures as high as 1223 K are observed [22]. Based on the curve in Figure 14, with the feed in this project, an exit temperature above 1073 K is needed to achieve a conversion above 50%.

With the high temperatures involved the reactions reaches equilibrium relatively fast. This means that the exit composition of a reformer can be approximated relatively straight forward from thermodynamic quantities. A concept that is sometimes used in this context is the “approach to equilibrium”, often quantified as a temperature difference such as:

$$\Delta T_{approach} = T(Q_R) - T_{exit} \quad \text{Eq.9}$$

where $T(Q_R)$ is the temperature the exiting gas would have if it was at equilibrium and T_{exit} being the actual exit temperature. The quantity Q_R represents the reaction quotient of the exiting gas. In this project, a $\Delta T_{approach}$ between 20K and 10K has been used when estimating compositions, depending on design parameters. This approach to equilibrium is within the range usually observed of 5 to 20 K [22] [25].

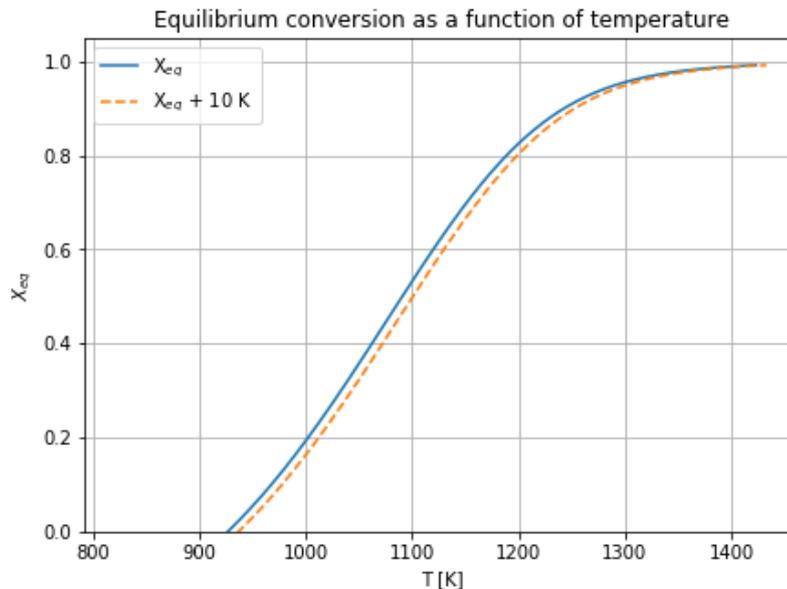


Figure 14. The equilibrium conversion as a function of temperature. The orange line denotes the conversion that could be expected at an approach to equilibrium of 10K.

Traditionally the high temperatures are achieved by conducting the reaction inside tubes filled with catalyst inside of a gas fired furnace. In order to provide the heat necessary to sustain R.2, the heat transfer to the catalyst bed must be good, leading to a high temperature difference needed across the tube wall [26]. These conditions obviously put quite strict requirements on the tube materials. In this work, when referring to actual process equipment, the acronym SMR refers to a traditional gas fired reformer.

An alternative to a gas fired reformer that has become widely used is the Autothermal Reformer (ATR). Instead of conducting the reactions in externally heated tubes, in an ATR the catalyst is placed as an adiabatic bed. To provide the heat necessary for the steam reforming reactions the feed is preheated by partially burning a part of it in pure oxygen just before it is to enter the bed [26]. In addition to providing heat at a high temperature, the addition of oxygen also alters the gas composition before the gas meets the catalyst, shifting the equilibrium somewhat. What could be seen as a variation of an ATR is to use air instead of pure oxygen. In this process, often called partial oxidation, (POX), instead of ATR, the reactions usually proceed uncatalyzed, this leads to potentially needing very high temperatures, around 1300 °C, to achieve a high conversion with a feasible residence time [25]. While a setup like this can lead to capital cost savings it often

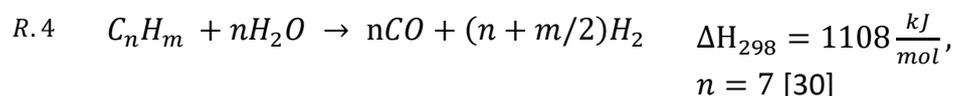
leads to a loss of energy efficiency due to the heat needed to get to the required temperatures.

The last alternative to a traditional SMR that will be discussed here is the possibility of using electrical heating to achieve the temperatures needed. This concept, here referred to as eSMR, while not a very mature technology, has a few potential benefits over the traditional setup if done correctly. Researchers at DTU and Haldor Topsoe recently published and patented a design for an inductively heated reformer [27]. One of the major advantages presented in their article is the size reduction possible with their design, stated to be up to a factor 100 times smaller than current reformer platforms. Furthermore, since no part of the reformer feedstock or other fuel gas needs to be combusted, an electrically heated process can achieve greater carbon utilization than a traditional setup.

Besides managing the temperature, another factor to consider is the possibility of side reactions leading to coking on the catalyst, and thus deactivating it. Simply explained, the way coking is usually handled is to conduct the process with an excess of steam and/or carbon dioxide. However, any extra steam beyond what is necessary is costly due to the energy required to produce it. There exists therefore significant incentive to optimize this parameter. Commercial reformer manufacturers like Haldor Topsoe has empirical data on the critical steam to methane ratio needed for different feeds [22]. Values seen for the steam to methane ratio is between 1 to 5 on molar basis [28]. However, any higher hydrocarbons except methane will drastically increase the risk for coking, thus leading to a higher steam to methane ratio needed [28] [29]. A way to sidestep this issue somewhat is by use of a pre-reformer before the feed enters the reformer [29]

5.4.2 Pre-reforming

The purpose of the pre-reformer is to convert any hydrocarbons with two carbon atoms or more into methane and syngas, and by doing this prevent them from coking in the reformer. This process also increases the hydrogen concentration in to the reformer, which offers further protection against coking [28]. The reactions in a pre-reformer can be described by an irreversible decomposition of the higher hydrocarbons into syngas, as per reaction R.4 below followed by the establishment of equilibrium of the WGS reaction, R.3 and the methanation reaction, the reverse of R.2 [30].



While the breaking of bonds in the higher hydrocarbons requires energy, as evident by the reaction enthalpy of reaction R.4, the overall combination of R.2-4 can be both endothermic and exothermic depending on the state of the feed. Since the reactions are usually carried out adiabatically at a feed temp of around 400 to 550 °C, significant methanation can occur depending on the feed composition, thus offsetting the energy needed for reaction R.4 [30]. In general, for a feed containing mostly methane, the process tends to be endothermic resulting in a lower exit temperature than the feed temperature, while for a feed containing more higher hydrocarbons, such as naphta, the situation is the reverse [30].

Just as in the main reformer, coking is an issue in a pre-reformer as well and care has to be taken as to supply a sufficient amount of steam. However, due to the lower temperature in the pre-reformer the total amount of steam needed for both the reformer and the pre-reformer is usually lower than if the pre-reformer had been omitted [29].

5.4.3 Water-Gas-Shift

As evident from the descriptions of the pre-reformer and reformer, the water gas shift reaction, R.3, is an ever-present factor when working with these syngas mixtures. Since the FT-synthesis requires a certain H₂ to CO ratio it can be beneficial to make use of a separate shift reactor after the steam reformation to adjust the final composition.

The shift reactor is usually operated either in high temperature regime, around 400 °C, or in the low temperature regime, around 200 °C, sometimes both are used. For the high temperature case an Fe based catalyst is used and for the low temperature case a Cu based one [26]. Depending on the feedstock the amount of shift needed will vary. It is possible that additional steam must be added to adjust the equilibrium further.

After the shift reactor, the gas can be cooled and most water contained within condensed off as any still left will tend to inhibit the following FT-reaction.

5.4.4 CO₂ removal

As seen in section 4, a certain amount of CO₂ will always have to be removed from the process due to the oxygen content of the biomass. By selectively removing this CO₂ the volume flows in the process goes down while a high efficiency of the process is maintained.

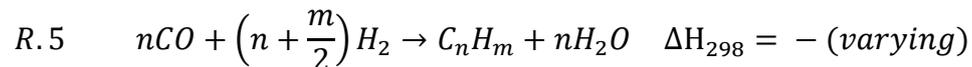
Removal of CO₂ is performed in many different processes and as such many different alternatives exist. A common route is wet treatment, washing the CO₂ out of the gas in scrubber columns. Here there is considerable overlap with the wet processes used for H₂S removal, both CO₂ and H₂S are considered “acid gases” and are often removed together [31], [32]. Thus, the refrigerated, physical absorption processes like Rectisol mentioned before as well as the chemical absorption processes with alkanolamines like MDEA are all used for CO₂ removal. In addition there are several membrane based solutions as well as processes based on hot potassium carbonate solutions [32]. For a more detailed overview the reader can consult a summary by C. Higman and M. van der Burgt [31].

In this project chemical absorption with an activated MDEA solution was chosen. For the scale considered, this alternative was deemed to be the most suitable. Since MDEA is the same amine as the one used in the H₂S scrubber the storage and make up can be simplified. Note however, that for MDEA to be suitable for CO₂ removal an activator must be added, without the activator the absorption rate for CO₂ is significantly lower. For H₂S scrubbing this lower CO₂ absorption rate is desired since this makes the solvent more selective towards H₂S, but for CO₂ removal it obviously needs to be improved [32]. The activator in this case is

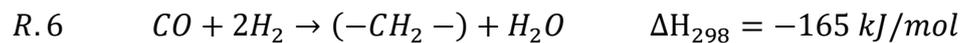
piperazine and the activated formulation is often referred to as aMDEA, being BASF's trademark name for it [31].

5.5 FISCHER TROPSCH SYNTHESIS

In a general sense, Fischer-Tropsch, (FT) synthesis can be thought of as a polymerization reaction where hydrocarbon chains are formed from carbon monoxide and hydrogen with the help of a metal catalyst [33]. A global formula for the reaction can be written as



A practical way of thinking of it is as a chain building process where methylene groups, made from CO and H₂, are added to an already existing chain. This description has the added benefit of having a more easily definable heat of reaction [33].



While these reaction formulas are obvious simplifications, and the reaction mechanism even today is not fully known, they serve to illustrate the fundamental characteristics of the process. The first thing to note is that the overall stoichiometry depends on the expected product distribution as seen in R.5. However, for a process targeting longer chain lengths, the CO to H₂ ratio will tend towards 2. Secondly, while the overall heat of reaction also depends on the product distribution, it is obvious from R.6 that this is a highly exothermic process. Efficient cooling will have to be implemented for the reaction to be conducted in a stable fashion and to avoid sintering of the catalyst. Furthermore, the reaction temperature has impacts on the product distribution as well, as will be discussed below.

5.5.1 Product distribution

While, as noted above, the reaction mechanism for the FT reaction is not fully understood yet and the product distribution obviously to some degree dependent on this mechanism, there are still simplified approaches to estimating it. The most encountered approach is that of the Anderson-Schulz-Flory (ASF) distribution [29], [30].

In the ASF approach, the process is characterized by a statistical parameter, that is "The probability of chain growth" commonly denoted by the greek letter alpha (α). The basic concept is that CO adsorbs to the catalyst surface and reacts as in R.6 above; at this point the process can either continue by the continued adsorption of CO or the chain is terminated by the formed species leaving the catalyst surface as either an olefin (C_nH_{2n}) or a paraffin (C_nH_{2n+2}). Since the α is the probability that the process continues as opposed to terminates, a higher α leads to products with longer chain lengths. By assuming the α constant for all chain lengths an expression for the probability for any one species to be formed can be obtained and from this a product distribution derived analytically. The final expression can be

seen below. For a more complete derivation, the reader can consult the works of R. Chadeesingh (2011) [26] or M.E. Dry (2004) [34].

$$w_n = n * \alpha^{n-1} * (1 - \alpha)^2 \quad \text{where } \sum_{n=1}^{n_{max}} w_n = 1 \quad \text{Eq.10}$$

Here w_n denotes the mass fraction of species with n carbon atoms and n_{max} the longest chain length observed. Figure 15 shows three examples of product distributions by this approach.

Given that the ASF distribution can be considered acceptable, which it has been seen to be in some cases but not all [34], the next step becomes to provide the process conditions that result in the alpha value that is deemed necessary. This in turns depends on a number of parameters, the most notable being temperature, pressure, feed gas composition, catalyst type and reactor design [26]. While there exists a few correlations for the alpha value's relation to some of these parameters [26], there is no magic bullet. In general, the alpha value is seen to increase with lower temperatures and lower H_2 to CO ratios in the feed gas. Furthermore, cobalt catalysts have shown a higher selectivity towards higher hydrocarbons than iron catalysts, i.e a higher alpha value [26], [35].

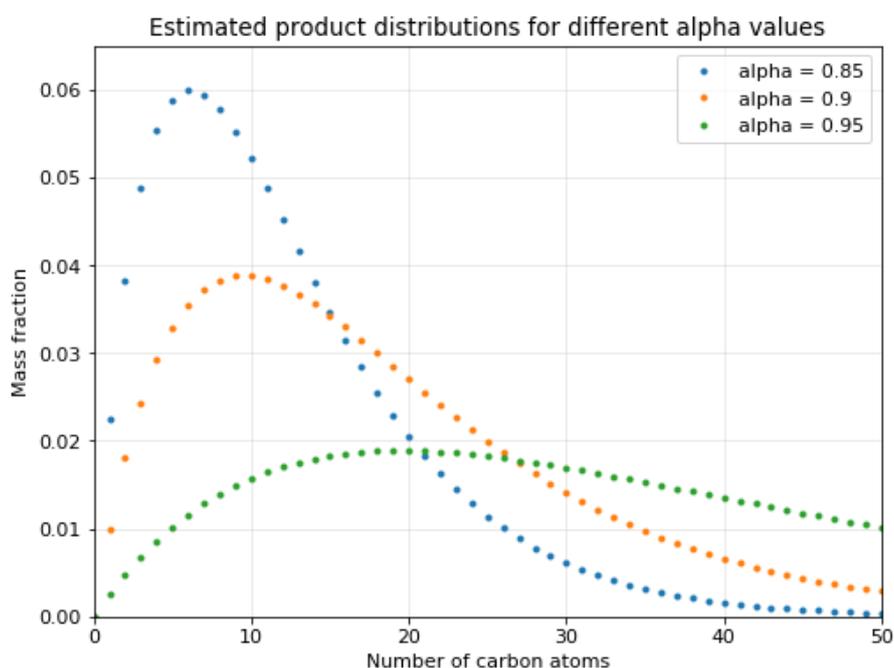


Figure 15. Examples of product distributions at 3 different alpha values.

However, neither the simplified reaction formulas above nor the ASF distribution show that for certain process conditions appreciable amounts of oxygenates like various alcohols tend to form as well. This once again highlights the need for further understanding of the FT reaction process. Here cobalt catalysts are known

to give rise to less oxygenates than iron catalysts, especially at lower temperatures [26][35].

In this project several alpha values have been examined but the final configuration is based on a value of 0.9. A low reaction temperature and a cobalt catalyst is thought to correspond to this value.

5.5.2 Choice of Catalyst

There are in theory a multitude of candidates for the metal catalyst. The metals, Ni, Rh, Ru, Fe, Co, Pd, and Pt are all active in the FT reaction [35]. However, in practice, the two catalysts used in industry are Fe and Co [26], [34]. The reason behind this is that most of the other metals are more expensive. Nickel being the possible exception, is also disregarded since it is too selective towards methane over higher molecular weight hydrocarbons [34]. Co catalysts are considerably more expensive than Fe but are still an alternative since they are active at lower partial pressures [35] and are more suitable for production of heavier hydrocarbons. Also, as noted above Cobalt catalysts tend not to produce as much oxygenates as Fe catalysts [34].

Perhaps the biggest difference between Fe and Co however is that Fe also catalyzes the WGS reaction [26], [33]. This makes iron less suitable for operation with high per pass conversions since, unless CO₂ is co-feed to the reactor, a significant part of the incoming CO will tend to form CO₂ instead of the intended FT-product. At low temperatures where a higher alpha value can be expected the WGS reaction is also highly skewed, requiring larger volumes of additional CO₂ to counteract.

Typical conditions for iron are high temperature synthesis above 300 °C, aiming for light to intermediate products such as gasoline [34]. It is especially used in fluidized beds where the iron can be made more resistant to abrasion compared to cobalt [36]. Applications with iron catalysts also tend to go for lower per pass conversions and larger amounts of recirculation to compensate [37] [34]. Iron could also be suitable when the feed to the process lacks hydrogen, such as when coal gasification is used to produce the necessary syngas, and the WGS reaction is welcome to some extent [34].

Cobalt tends to be used more often in low temperature synthesis, under 250 °C, where it can be used to achieve a high per pass conversion with a possibly high alpha, maximizing the production of heavier hydrocarbons (waxes) [37]. In synthesis where the syngas is produced from natural gas, and tends to have more hydrogen than necessary, cobalt catalysts could also be preferable, since the WGS reaction wouldn't be beneficial in this case [34].

In this project both types of catalysts were investigated but, in the end, the final configuration uses a cobalt catalyst. The target product is towards the higher end of the spectrum and thus a low temperature was needed in order to achieve a high alpha value. In our simulations, the flows of CO₂ needed to balance the WGS reaction in case of an iron catalyst were simply too large to be feasible. However, it should be noted that more detailed models might have lessened this effect somewhat.

5.5.3 Reactor design

Designing a reactor for Fischer-Tropsch synthesis is not to be taken lightly. The high exothermicity of the reaction and the adverse effects on the product distribution with rising temperatures makes for strict requirements on the heat removal from the reactor as well as internal heat transfer to avoid hotspots. Furthermore, the reaction being a heterogeneously catalyzed gas-liquid reaction means that at least for low temperature synthesis, the reactor tends to be a three-phase system. The additional phase of course complicates the analysis of these systems. Additionally, good mass transfer characteristics are desired as to make for a high reaction rate and a corresponding lower reactor volume.

Since the invention of Fischer-Tropsch technology in the early 20th century many different reactor types have been conceived and tested. In the time between 1935 and 1955 German scientists investigated more or less every reactor configuration possible for this kind of reaction [38]. However, in commercial use today for low temperature wax production there are two main alternatives, the fixed multitubular bed reactor and the slurry bubble column reactor. For high temperature production, fluidized bed reactors, both circulating and fixed are used. More information about reactors for high temperature synthesis can be found in [26], [36].

In a multitubular fixed bed reactor the catalyst is placed in parallel tubes cooled on the outside by boiling water. The reactor can be thought similar to a tube and shell heat exchanger with the tubes filled with catalyst and boiling water on the shell side. An example of multitubular reactor technology are the so called ARGE reactors first designed in 1955 and still used to this day [36][38]. The main advantages of these reactors are their ease of handling and their ease of design. Since the parallel tubes behave very similarly scale up is made easier since only a single tube has to be considered [38], [39]. There are disadvantages of course, in the dimensions used industrially radial heat transfer can be insufficient leading to hotspots unless the temperature of the cooling as well as the feed is kept substantially lower than the maximum temperature [38]. Furthermore, to avoid too high pressure drops the size of the catalyst particles cannot be made too small [34]. Smaller sizes would be beneficial for the catalyst activity and the product distribution [36]. Relative to other technologies the cost of multitubular fixed bed reactors can tend to be higher [38].

In a slurry reactor the catalyst is dispersed in a liquid phase, often consisting of the formed FT-products, and feed gas is bubbled through the reactor. The cooling is achieved by way of internal cooling tubes running throughout the reactor. Unreacted gas is usually recirculated to keep the suspension fluidized and to increase the overall conversion [36][38]. Since the catalyst particles flow with the FT-product, a separation step is needed to remove the catalyst from the product and return the catalyst to the reactor. The catalyst particles are usually in the form of powders with particle diameters on the order of micrometers instead of millimeters like those in a multitubular bed reactor [38]. This decreased particle size greatly decreases internal mass transfer limitations, increases catalyst utilization and allows for greater control of the selectivity and activity [38]. Furthermore, with well-designed cooling surfaces the heat removal is enhanced,

allowing for a more even temperature distribution than that of a fixed bed [38]. The main challenges of a slurry reactor design are to ensure satisfactory particle separation, sufficient gas-liquid mass transfer, and scale up [36], [38]. The reactor being a three-phase system with complex fluid dynamic interactions between all phases makes a slurry reactor harder to design. Scale up is made highly difficult for the same reasons, and suitable pilot reactors can be costly to develop [39]. However, once the design is set, a slurry reactor is usually cheaper to manufacture than its fixed bed counterpart [36].

In this project the multitubular design is the one chosen. Even though the slurry reactor could present several advantages with regards to productivity and manufacturing cost, the aforementioned difficulty of design and scale up makes the cost estimation of such a reactor at this stage unfeasible.

In addition to the reactors mentioned above, more recent concepts such as membrane reactors, microchannel reactors, honeycomb reactors might prove good alternatives in the future [38]. However, in this project it was chosen to pursue more well-established technologies when the FT reactor was concerned.

5.5.4 Synthesis setups

To generalize, two different setups can be identified in terms of how to run the synthesis process, either a “once-through” setup or a setup with tail gas recirculation. Which one of these will be preferred depends on the feed to the FT-process, the desired product composition and to some extent also the reactor specifications.

In a once-through process the produced syngas is feed through the FT-reactor and after this any unwanted FT-products or unconverted methane in the syngas are discarded as fuel gas [37]. In this setup the limiting factor for the carbon efficiency of a once-through process is the degrees of conversion over the FT reactor and the reformer. Furthermore, having a high selectivity in the FT-reactor for the desired product is key [37]. Generally speaking, the only product besides methane that can be produced with high selectivity is high molecular mass waxes, corresponding to a high alpha [34]. These two factors, the need for a high alpha and a high conversion means that cobalt catalysts tend to be favored for once through applications. Benefits of a once through process is decreased complexity and decreased operating costs. Disadvantages are a lower maximum efficiency than a recycled process and possibly stricter demands on the equipment performance.

In a setup with tail gas recirculation unreformed methane and the unwanted products of the FT-reaction are sent back to the gas conditioning step to be reformed back into syngas [37]. A recycled approach will not need an as high per pass conversion in the reformer and offers the possibility of targeting other cuts than hard waxes [37]. This is the approach taken in this project. For iron catalysts and especially for HTFT targeting lighter to intermediate fractions like gasoline this approach is also preferred. In in a recycled setup the limiting factor to the carbon efficiency is the amount of gas that must be bled from the recirculation loop. The purpose of the bleed is to prevent any inert species entering with the syngas, such as N₂ used for purging for instance, from accumulating within the

looped system leading to inflated volume flows and equipment sizes. However, since the bleed is not operating in a selective manner to remove the inerts a certain fraction of hydrocarbons and syngas will also be bled.

Regardless of whether tail gas is being sent for reformation or not, a common feature is to include an “internal recirculation” around just the FT-reactor [36]. By doing this the FT-reactor itself does not need to have a very high per pass conversion and the gas velocity through the reactor can be regulated. Furthermore, the increased flow of inert species through the reactor can help mitigate the temperature increase. In certain reactor setups cooling is also performed on this recirculation flow [36], [39]. In this project the internal recirculation is regulated to make sure that a certain fraction of inert thermal ballast is present.

6 Bio-FT Process Configuration Options

To identify an attractive combination of technologies for the reformation and FT process a screening was performed to investigate the performance of different combinations using ASPEN plus by Per Tunå (LTH). The investigated cases were evaluated based on the efficiency based on cold gas to FT-crude (CV_2), $\eta_{chem,CG-FT}$, and the electro-chemical efficiency for the FT section of the plant, $\eta_{el-chems,CG-FT}$. Further this is compared with the efficiency based on the theoretical carbon utilization, $\eta_{chem,theol}$. See section 4 for definitions.

6.1 REFORMING TECHNOLOGIES

The values presented in this section should be taken as qualitative indication and focus here is to provide comparable cases. Figure 9 shows the estimated efficiencies for cases based on low temperature FT-process (LTFT) and different technologies for reformation:

- SMR: Gas fired steam reformer where tail gas from the FT is used to heat the reformer instead of being recycled to the reformer.
- ATR+elektrolysis: Partial oxidation and catalytic reformation where oxygen is assumed to be produced through electrolysis as this enables parallel production of hydrogen, which can also be added to the process to limit the carbon loss of the ATR. In this case only a small fraction of the tail gas is bled of to avoid N_2 accumulation.
- eSMR: Electrically heated steam reformer. In this case only a small fraction of the tail gas is bled of to avoid N_2 accumulation.
- POX: thermal non-catalytic cracking of tar and hydrocarbons by addition of oxygen.

The different cases have different impacts on the heat integrations. Heat integration and total efficiency of the process is further investigated in section 7.

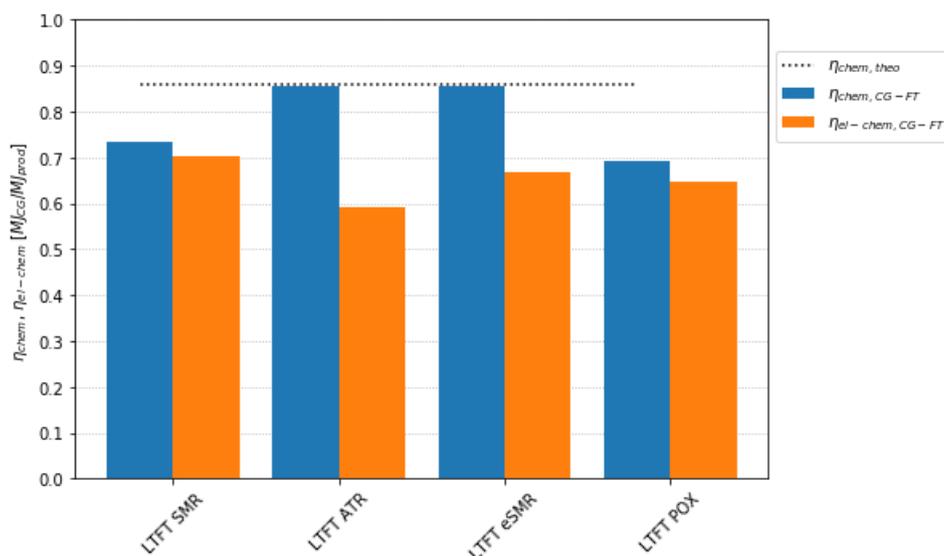


Figure 16: Estimated chemical and electro-chemical efficiencies using different technologies for reformation of the gas conditioning and synthesis section (CV3).

The results of the investigated cases are also compared with the theoretical chemical efficiency based on the stoichiometric carbon utilization (Eqs. 1-5), which indicates how much of the cold gas which could not be utilized in the FT-crude production for different reasons. For instance, some of the gas always needs to be bleed of to avoid accumulation of N_2 or to be burnt for heat production but can also be restricted by the equilibrium of the WGS reaction in the reformer, as further discussed in section 7.

Results in Figure 16 show that the chemical-efficiency, and thereby the utilization of the gas produced in the gasifier, can be significantly improved by also adding electricity to the process through an ATR or SMR. Increasing the carbon utilization is very attractive as it enables a higher production rate with a specific gasification capacity. For biomass-based processes the gasifier can generally be assumed to be the limiting factor for how much a plant can be designed to produce. This makes the ATR+elektrolysis and eSMR options very attractive. Note that even though the electro-chemical efficiency is lower for these cases than for the investigated SMR case based on product gas (CV3), it is higher based on biomass (CV1).

In the POX case tar is converted, which could improve the chemical efficiency, but results show that due to the high heat demand of the POX the chemical efficiency is still lower than for the SMR case. Further, the electro-chemical efficiency decreases as well as the volume flow increase before the compressor, requiring more electricity for the compression.

6.2 FT-TECHNOLOGY

Simulations were conducted to investigate the impact of applying a high temperature FT-process (HTFT) compared to a low temperature FT-process (LTFT). The temperature of the process has a strong impact of the selectivity where

production of larger molecules is favored at lower temperature (high α -value). For HTFT, which can be operated at about 350 °C Fe-catalyst is used. For LTFT, Co-catalyst is more common and can be operated at temperatures below 250 °C. For the following comparison $\alpha = 0.7$ was assumed for the HTFT (Fe-cat. and 350°C) case and $\alpha = 0.85$ was assumed for the LTFT (Co-cat. and 250 °C). Another important aspect to be consider is that the Fe-catalysts are active towards the WGS-reaction while Co-catalysts are not.

Figure 13 shows the simulation results for LTFT and HTFT combined with the reforming technologies SMR and eSMR. Results shows that the previously described benefit of the eSMR process over the SMR process hold also when combined with a HTFT. The comparison also shows that a significantly lower chemical efficiency is achieved with HTFT then with LTFT. This is due to a higher selectivity towards lighter components with HTFT which needs to be recirculated and reformed ones more to be utilized for FT-crude production. The penalty is large in combination with an SMR as more tail gas needs to be extracted to fuel the SMR. With an eSMR the penalty is more limited but are still significant.

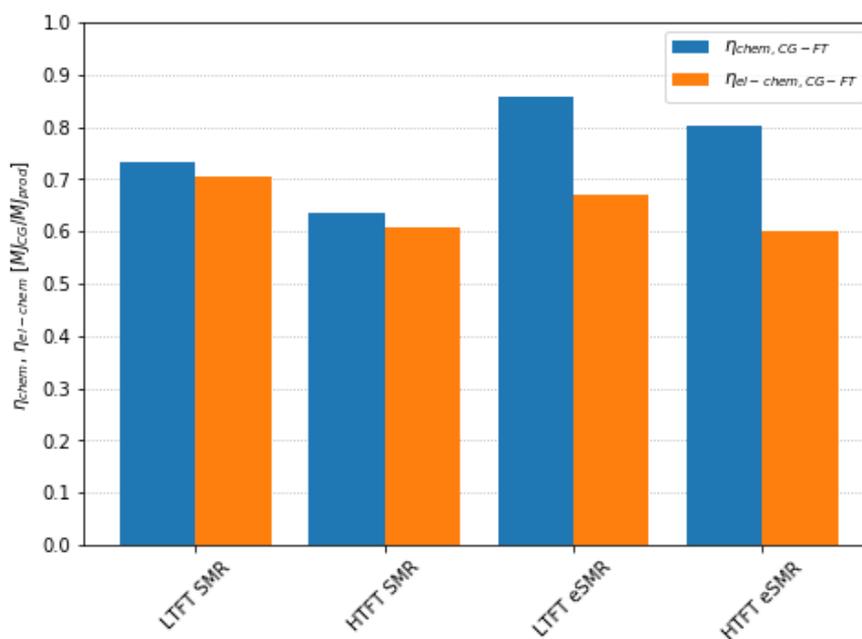


Figure 17: High temperature FT (HTFT) compared with low temperature FT (LTFT) based on the chemical and electrochemical efficiency of the gas conditioning and synthesis section (CV3).

6.3 ELECTRO-FUELS AND RAW BIOGAS INTEGRATION

A number of cases where electricity or biogas was added to the process was simulated to investigate the marginal efficiency of electricity or biogas to this type of process. The marginal efficiency was quantified according Eq. 6, (see section 4) and represents the increased production of FT-crude (MW, LHV) compared to the selected reference case based on LTFT with SMR divided by the additional energy added as electricity or biogas. The investigated cases are:

- LTFT ATR: electrolysis is used to produce the oxygen needed for the ATR process and where simultaneously produced hydrogen is introduced to the process.
- LTFT eSMR: electricity is used to directly heat the reformation process.
- LTFT SMR+EI: electricity is used to produce enough hydrogen to adjust the H₂/CO ratio without a WGS reactor.
- LTFT SMR+Biogas: An arbitrary amount of raw biogas of 6.5 MW (65% CH₄ and 35% CO₂) was added to the process prior to the reformation.

Simulation results are presented in Figure 14, which indicates that marginal efficiency of about 30-60% can be reached for electricity while a marginal efficiency of as high as 80% can be reached for raw biogas. The highest marginal efficiency for the electricity was achieved with the eSMR for which the FT-crude production can be increased with about 0.5 MW per MW of electricity as compared with the LTFT SMR case with same feed of cold gas. This indicates that electricity could be used rather efficiently to produce electro-chemical biofuels. Results also indicate that if local streams of raw biogas are available it could be utilized with very high marginal efficiency in this type of process. Note however that the CO₂ concentration will have a strong impact on the marginal efficiency of the raw biogas. The raw biogas also needs to be cleaned from contaminants such as H₂S either before addition or by adding or by adding the raw biogas prior to the cleaning steps in the main process.

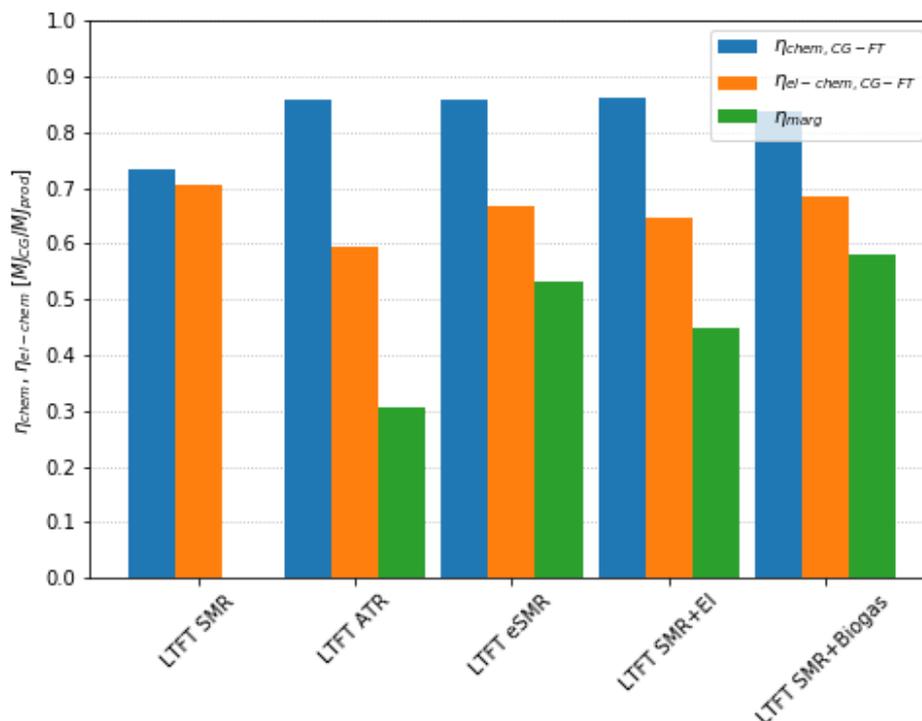


Figure 18: Chemical and electrochemical efficiency of the gas conditioning and synthesis section (CV3) for different cases with addition of electricity or raw biogas. Also includes the marginal efficiency of the electricity or biogas using the LTFT SMR as reference case, see Eq.6.

7 Case study - Karlstads Energi

In this section, project and site-specific features are described for the conducted case study of a co-production of 30 MW FT-crude at Karlstads Energi site "Hedenverket" by retrofitting an existing boiler.

7.1 PRESENT PLANT CONFIGURATION AND OPERATION

The studied boiler at Hedenverket is a circulating fluidized bed (CFB) boiler, originally built to fire both coal and wood chips, though these days operating solely on biomass, mostly forestry residues. Due to the previous coal firing, the boiler is set up to operate with flue gas recirculation if needed. The boiler has three superheaters with condensate injections in between, the steam exiting the last superheater is at 500 °C and 66 bar. To further increase the plant efficiency and amount of district heating produced, the boiler is also fitted with a flue gas condenser. In addition to the condenser, as the last recovery step before the stack, the flue gases also pass a rotating heat exchanger and humidifier to preheat and increase the moisture content of the incoming air.

At high load the steam produced amounts to around 80 MW, of this around 20 MW ends up as electricity through expansion through the turbine and around 55 MW is utilized as district heating in the turbine condenser. An additional 20 MW is obtained in the flue gas condenser at high load under normal conditions.

In later years the studied boiler has been used more sparingly. At the site Karlstads Energi also operates a newer boiler for electricity production as well as a smaller boiler firing waste. The waste boiler is usually in full operation followed by the newer boiler with the studied boiler taking more of a supportive role. The limiting factor is of course the heat demand of the district heating grid.

7.2 RETROFITTING OF THE CHP PLANT

The existing combustion reactor is a CFB boiler and can therefore be retrofitted by rebuilding the loop seal of the boiler to include a BFB gasifier. The gasifier and sand transport are designed to enable start-up and operation of the boiler without the gasifier in operation. In the current process silica sand is used as bed material and to minimize the required retrofit same bed material will be kept for the co-production. Redundant fuel feeding systems to the boiler can be adapted for fuel feeding to the gasifier. The temperature in the gasifier will be controlled by the solid circulation through the gasifier.

The primary gas cleaning should be placed as close as possible to the gasifier and a cyclone separator, product gas cooler, tar scrubber, water scrubber and fan should be located close to the process. Material removed in the primary cleaning step will be feed to the combustor for destruction and heat recovery.

Seen from the perspective of the existing furnace, the main difference after retrofitting is a changed fuel composition as char from the gasifier and byproducts from the gas cleaning now will be burnt. Secondly, the boiler will have to back down its production a bit since district heating will be produced by the FT-plant that should be taken up by the district heating loop. Together these two changes means that both elemental composition as well as the average moisture content of the fuel will have changed. Because of this change, the flue gas from the boiler should have a higher quota of CO₂ compared to H₂O after the retrofit.

7.3 BIOJET PROCESS CONFIGURATION

After considering several different layouts, as discussed in section 5.6, the chosen configuration of the process from biomass to FT-crude is now based on LTFT combined with eSMR the details of which will be seen below. Note however, that the estimated performance and dynamics is case specific, and depends on a number of external parameters, notably the raw gas composition. Also, note that the estimation is based on simplified models of the reformation and synthesis reactor that should be further elaborated during detailed engineering.

The final configuration can be seen in Figure 19. The process is divided into three parts, the first two parts, A and B, being reasonably linear with the last part involving several loops. To provide a better understanding about part C, the looped part, this part is expanded further in Figure 20.

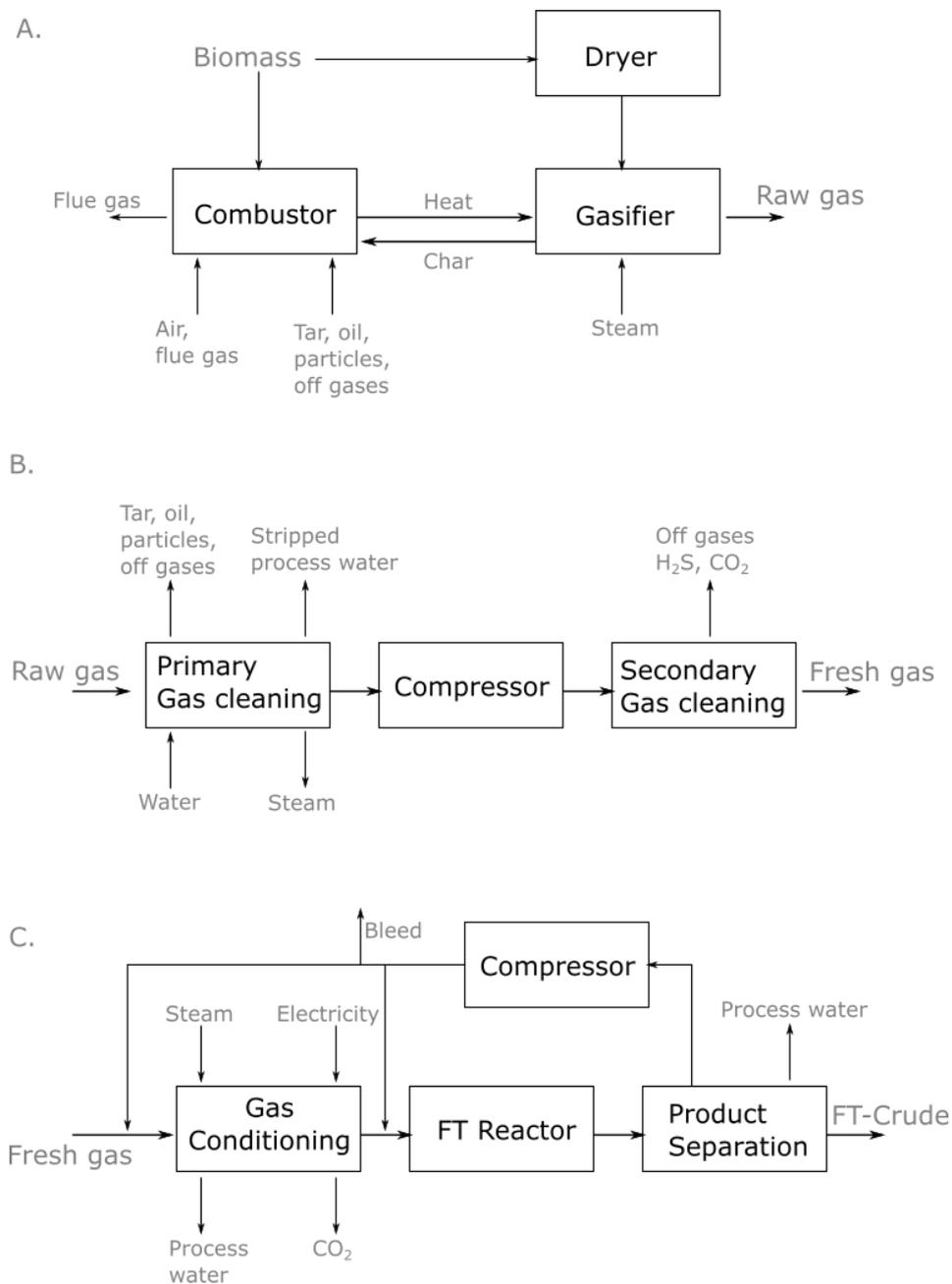


Figure 19: An overview of the chosen process layout. The three parts A-C are consecutive. The combustor block refers to the already existing furnace with which the gasifier will be integrated.

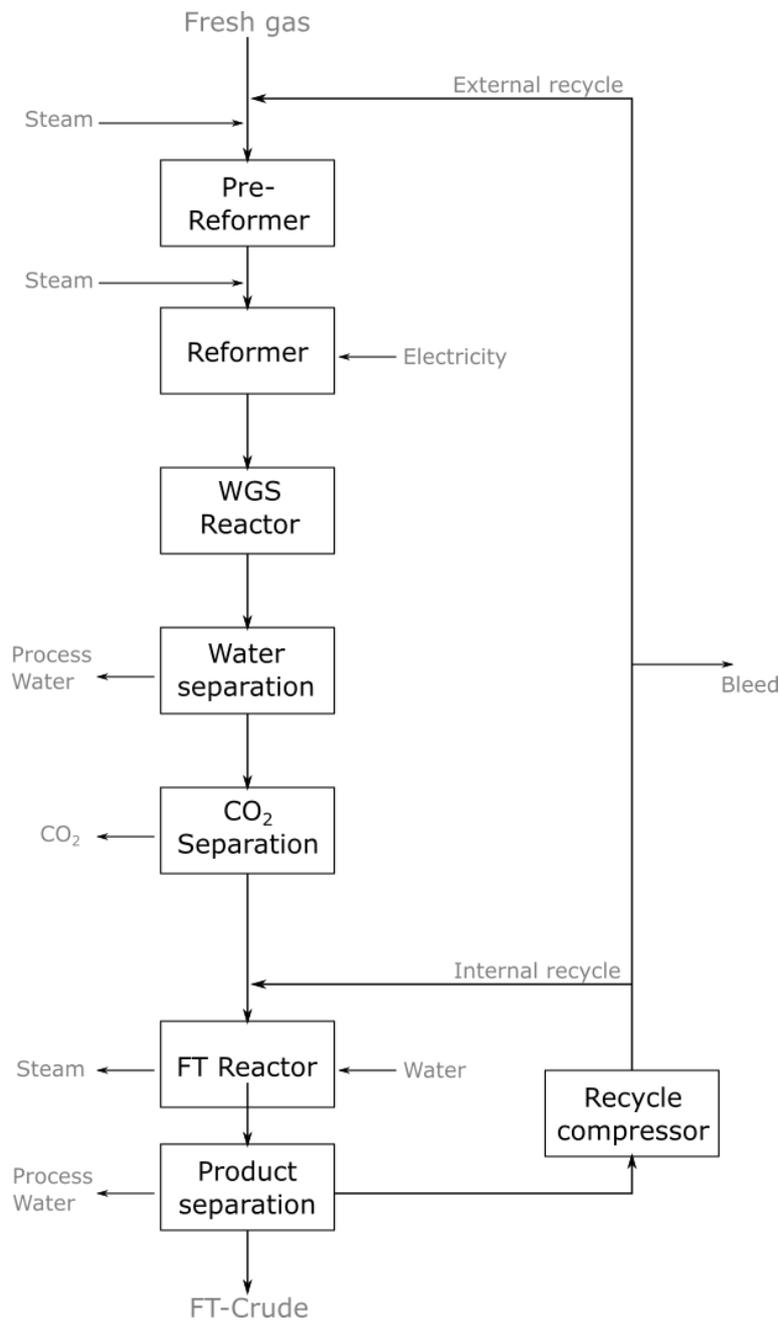


Figure 20: Zoom in on the gas conditioning and FT synthesis parts. Note the 2 recycle loops and the sequence of operations leading up to the FT-reactor.

In the FT- synthesis part, part C, due to the many points where the flow is split there are many ways in which the process could be run. In order to find an operating point, besides sizing the equipment, these split ratios must be determined. The main parameters that must be set are:

- The size of the internal recirculation
- The fraction of the reformed gas sent to the WGS reactor
- The size of the bleed from the system
- The amount of CO₂ that is removed in the CO₂ separation

Regarding the internal circulation, in this project this split is set by considering that the composition of the feed to the FT reactor should contain a certain fraction of gases inert in the FT reaction to act as thermal ballast. For this project a mole fraction of inert of 0.4 was chosen.

A similar reasoning can be made for the utilization of the WGS reactor. The FT-feed should have a H_2 to CO ratio of around 2. As long as the ratio is below this value when it exits the reformer, the split before the WGS reactor can be used to ensure that just enough is reacted to land the FT-feed on this mole ratio.

With these two constraints, a given conversion and the desired amount of FT-crude out from the plant, the feed to the FT reactor is fully determined. The remaining two parameters then have to be chosen in order to minimize the raw gas needed to produce the FT-feed while at the same time making sure to avoid accumulation of any one species in the loops. Accumulation would lead to increased volumetric flows and thus increased capital costs.

In general, there are three kinds of accumulation in this system.

- Accumulation of CO_2
- Accumulation of H_2
- Accumulation of inert species such as N_2 and unreformed CH_4 .

Accumulation of CO_2 happens, quite simply, when not enough CO_2 is removed in the CO_2 separation step. Taken too far this form of accumulation also leads to a difficulty maintaining the desired inert fraction in the FT-reactor.

On the other hand, if too much CO_2 is removed accumulation of H_2 occurs. Without a certain amount of CO_2 to balance the WGS reaction in the reformer, (R.3), more hydrogen than is needed to get the correct H_2 to CO ratio will be formed. Since the current system does not include the means to selectively remove hydrogen any excess hydrogen will tend to accumulate. Not considered, this form of accumulation could lead to very large volume flows in the reformers and especially in the FT-reactor. In the FT-reactor the effect on the volume flow is most severe since more hydrogen also means that the internal circulation must be increased if the specified inert concentration is to be achieved. A possible workaround for this kind of accumulation is to increase the temperature in the reformer until the WGS equilibrium quotient is balanced by the amount of CO_2 entering with the fresh gas. However, due to the inherent difficulties of working at high temperatures the next best solution is to regulate the amount of CO_2 being removed. A third workaround would be to put a bypass on the reformer instead, though that would have an impact on the efficiency of the process to some degree due to a higher amount of energy containing species in the bleed stream.

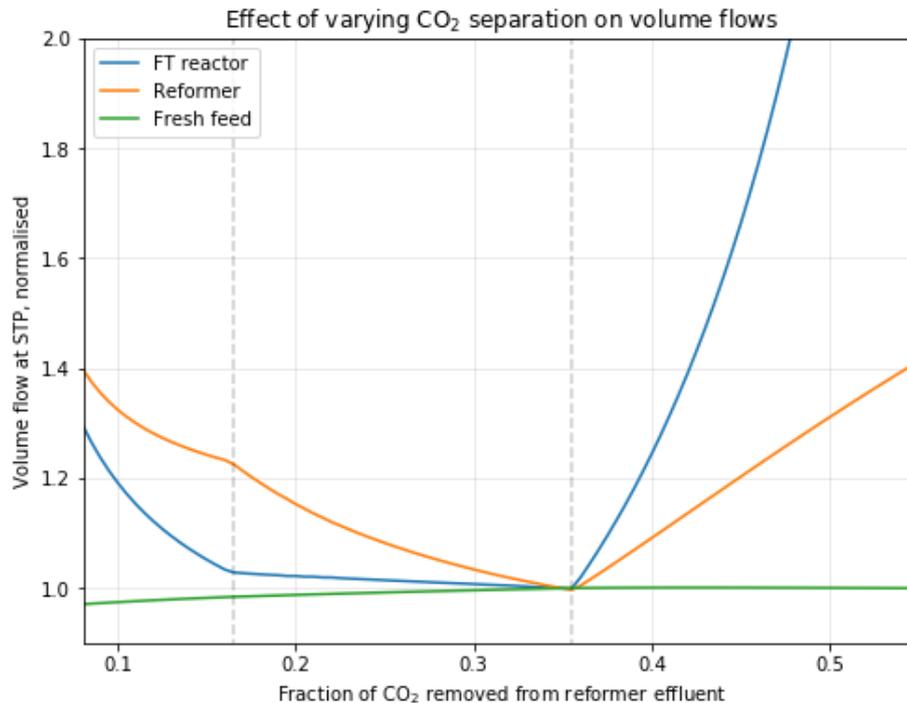


Figure 21. The effect of varying CO₂ removal while the bleed flow is kept constant. The flows are normalized around the chosen operating point. Note the three different regions visible. In the first region from the left there is accumulation of CO₂ in the loops due to insufficient removal. The middle region represents the feasible window where both the internal circulation and the WGS split is able to regulate as to give the FT-feed its desired composition, note the almost constant volume flow to the ft reactor in this region. The third region is where hydrogen accumulates since the amount formed in the reformer is above the desired value.

The third form of accumulation, that of N₂ and unreformed methane, happens when the bleed is set too low. It is assumed that a certain amount of N₂, although very small, will always be present in the raw gas. Since N₂ isn't selectively removed in any part of the process, the only possible outflow for it is in the bleed stream. However, at least if CO₂ is not being accumulated, the bleed stream is tied strongly to the efficiency of the process, as seen in Figure 22. Thus, this kind of accumulation forms a limit for the feasible overall efficiency of the process. While a higher reformer temperature would help somewhat in this case as well, the only real solution if a lower bleed is desired is to minimize the amount of N₂ entering with the raw gas.

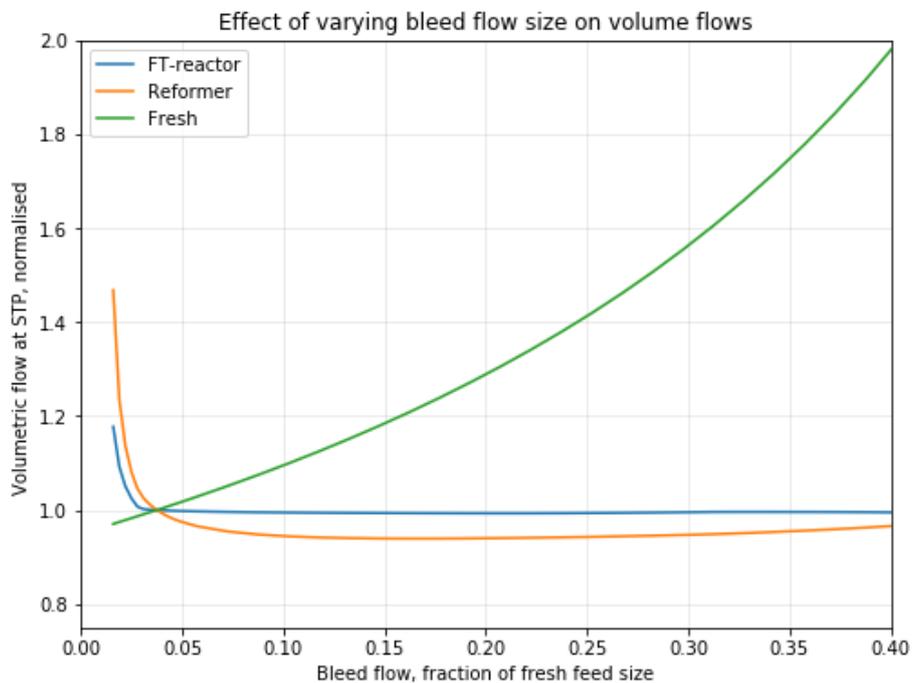


Figure 22. The effect of varying the bleed flow, normalised at the chosen operating point. Note how the volume flows for both the reformer and the FT reactors rise sharply at low bleed flows indicating an accumulation of inert material in the loops. Note also how the fresh feed grows in order to produce the same amount of FT-crude for higher bleed flows, indicating a loss of efficiency.

Finding the optimal amount of CO₂ to remove analytically is not trivial, the loops and the bleed complicates things. A similar argument can be made for the optimal bleed flow. However, by numerical simulation, a 2D sweep over both parameters can be made. Some results from such a sweep is shown in Figure 23 to 25 below.

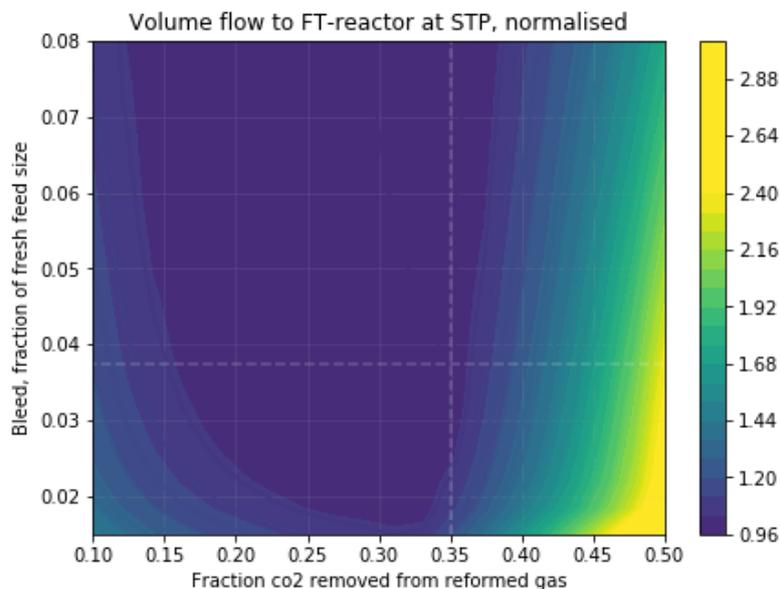


Figure 23. A 2D sweep of the FT reactor volume flow as affected by bleed and CO₂ removal. The colors are normalised by the value at the operating point. The dashed lines follow the univariate sweeps shown in Figure 11 and Figure 12 with the operating point at their intersection.

In Figure 23 the volume flow through the reactor is visualized, here there is a plateau visible with near constant volume flows providing the 2D analog of the window of operation seen in Figure 21. Moving to the right of this plateau the volume flows increase drastically due to accumulation of H₂, and to the left due to accumulation of CO₂. Moving downward in the figure shows how the limited bleed leads to accumulation of inerts and methane as seen in Figure 22 before.

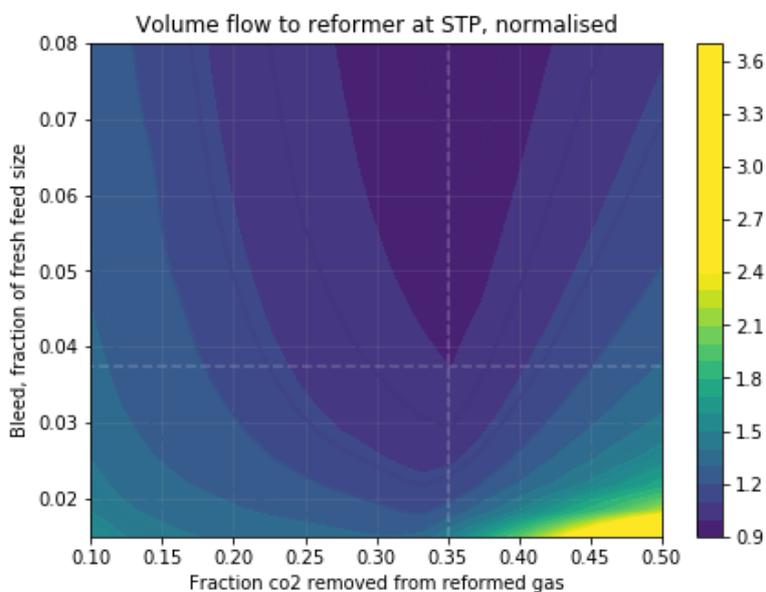


Figure 24. A 2D sweep of the reformer volume flow as affected by bleed and CO₂ removal. The colors are normalised by the value at the operating point. The dashed lines follow the univariate sweeps shown in Figure 11 and Figure 12 with the operating point at their intersection.

The effects of the same phenomena are visible in Figure 24, but this time their effect on the reformer volume. Since, in contrast to the FT-reactor, no attempt to control the composition of the reformer feed is made, its volume changes more with changing parameters. No plateau is visible in this figure, instead it seems that the reformer has a preferred degree of CO₂ removal along which the volume is minimal, possibly getting a bit skewed on the lower end of the bleed spectrum. Not surprisingly the reformer volume decreases with increased bleed flow, at least in the region investigated in this sweep. However, looking back at Figure 22 it would seem that this trend reaches a turning point somewhere around 0.15, outside of the sweep.

Note that as the bleed flow decrease, the gradients of the volume flows for both reactors go from almost horizontal, to diagonal to vertical. This transition indicates a shift between the dominating parameter. As long as the gradients are not entirely vertical or entirely horizontal both parameters have to be chosen in tandem.

The last 2D sweep in Figure 25, shows the effect on the chemical efficiency. Here it is clear that the bleed is the most important parameter in most of the region. The exception is in cases corresponding to accumulation of CO₂ where both parameters seems to have an influence. This is reasonable since in those cases the bleed contains mostly CO₂ and as such contributes less to the chemical energy lost than in the rest of the region.

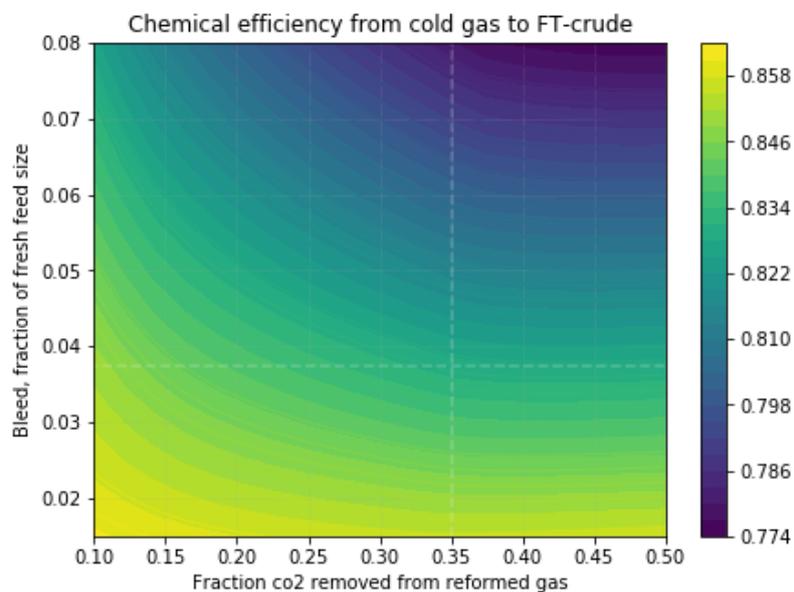


Figure 25. A 2D sweep showing how the chemical efficiency is affected by changing the bleed flow or fraction of CO₂ removed. The dashed lines follow the univariate sweeps shown in Figure 11 and Figure 12 with the operating point at their intersection.

Given this overview, the operating point is mainly determined by the minimum reformer volume and a desire to maximize chemical efficiency. Care is taken as to place it in the window of operation for the FT reactor, the blue plateau in Figure 23. In the 2D sweeps the operation point can be seen by the intersection of the two

dashed lines. With this operating point, the efficiencies for the two internal control volumes described in section 4 can be seen in Table 5 below.

Table 5: Performance characteristics of the two control volumes defined in section 4, CV2 and CV3 and the combination of the two. The definitions of these efficiencies can be seen in section 4.

	Biomass to cold gas, CV2	Cold gas to FT-Crude, CV3	Biomass to FT-crude, CV2 and CV3
Carbon utilization [%]	58.8	59.4	34.9
Carbon efficiency [%]	-	95.8	-
Chemical efficiency [%]	58.1	82.9	48.2
Electro-chemical efficiency [%]	58.1	61.3	35.6

7.4 PROCESS INTEGRATION

In this section the physical integration as well as the heat integration are described and discussed.

7.4.1 Physical integration

The gasifier and primary gas cleaning close to boiler while FT-Process can be built separate from existing units. The gasifier has to be located in connection to the boiler and can be adapted to fit the site-specific space. The gasifier is placed in connection to the boiler cyclones and bed material can either be circulated through the gasifier or as in the original boiler straight back to the combustor. The primary gas cleaning is also placed in close proximity to the boiler to facilitate recycling of tars and particles back to the combustor.

Process equipment downstream of the primary gas cleaning can be located separate from the existing boiler. Figure 27 shows a printout from a tentative 3D-model of the process located at Karlstad Energi's Hedenverket. Roughly 3500 m² was required for the plant including all process equipment, tank farm and housing for the electrical and instrumental equipment as well as the fuel dryer.

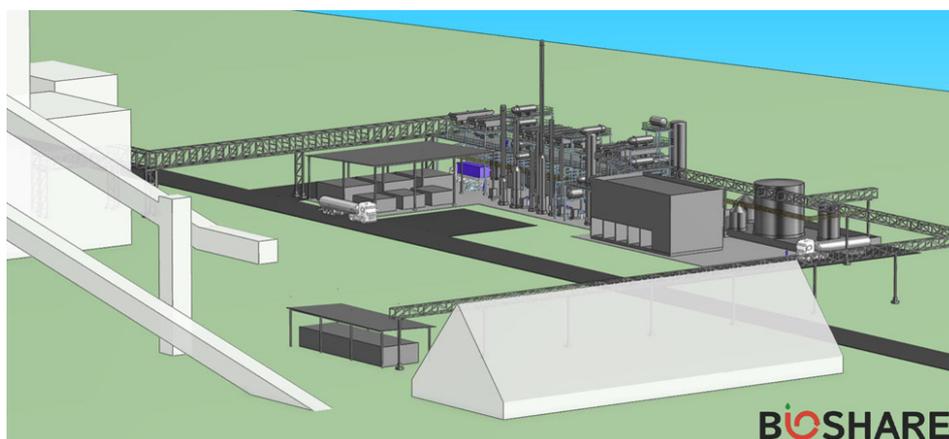


Figure 26: 3-D model of the simulated process, excluding pipes.

7.4.2 Heat integration

As with any process, the suggested co-production plant has several steps that both demand and produce heat. A heat exchanger network must be constructed to satisfy these demands and recover a fair deal of the heat for internal use. Since district heating is a product from the plant, good heat recovery is encouraged.

To provide some visual aid for the discussion about heat integration Figure 27 shows a grand composite curve (GCC) incorporating the major heating and cooling demands of the process. In a GCC the heating and cooling demands at all given temperature intervals of the process is summed up. The GCC can be used to gain some insight into what heat recovery is possible to gain from heat integration and what different utility loads that is to be expected. The theoretical minimum hot and cold utility is read from a GCC by the values at the top and bottom of the curve respectively. Temperature regions of the process where potential savings can be made through integration is visible as “pockets” in the GCC. The concept of a GCC is derived from the method of Pinch Analysis, for more background on this the reader is referred to the book “Pinch Analysis and Process Integration” by Ian C. Kemp [40].

Analyzing the GCC starting from the highest temperature, the first thing to note is that most of the hot utility needed, around 10 MW, must be supplied at a temperature level above 800 °C. This is the heat needed in the reformer to sustain the reforming reactions and to preheat its feed. The next thing to note is the horizontal line just beneath the 400 °C mark. This is the heat contained in the high temperature, high pressure, process steam supplied to the reformer. Optimization of the steam to methane ratio in the reformer is of interest since this heat demand is what constitutes most of the additional hot utility. However, note the pocket over the process steam line as well, this is an indication that a significant amount of the process steam can be generated internally, in this case through the radiant syngas cooler and reformer effluent.

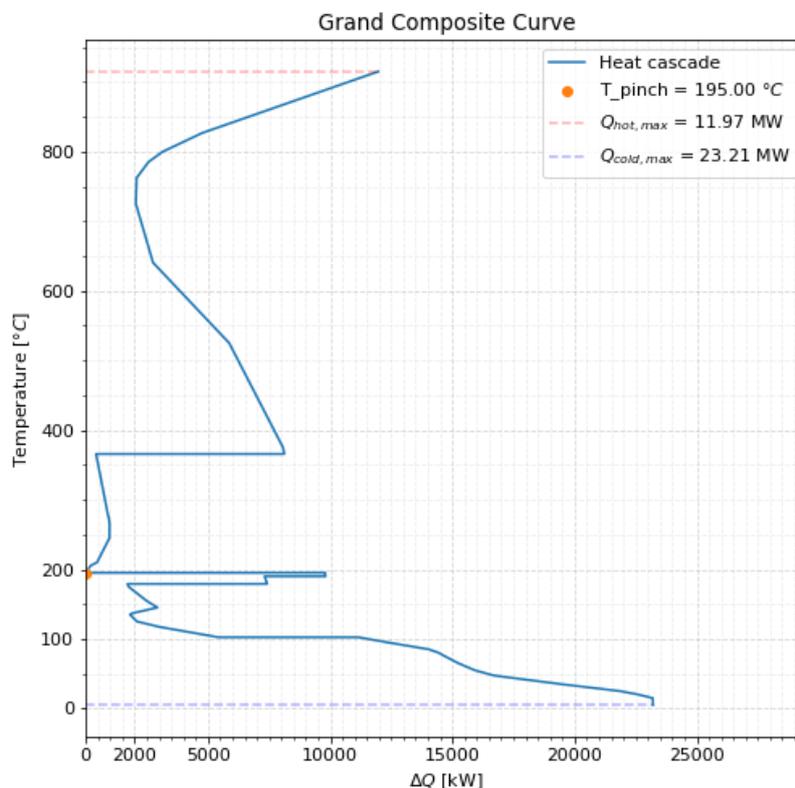


Figure 27. Grand Composite Curve over the process from raw gas to FT crude. In the heat cascade a minimum temperature difference of 50 deg C was assumed. The quantities Q_{hot} and Q_{cold} represent the theoretical minimum heating and cooling demands of the process.

The next part to note in the GCC is the lower horizontal line at around 195°C, this heat load is the reaction heat that must be removed from the FT-reactor. It is obvious that successful recovery of the heat from the FT-reactions has great benefits for any FT-plant. In this setup there are demands for steam that can be met by the steam from the reactor cooling. These steam demands are both fluidization steam and steam to power the dryer, they are the next horizontal lines below the reactor cooling, the dryer steam being the larger of the two. Heat recovery from the dryer is seen as the last horizontal segment around 100 °C in the diagram.

The final point to note is that of the cold utility, around two thirds can be supplied at a temperature between 65 and 90 °C, which can be accepted by the district heating loop, a very beneficial aspect of integrating with a CHP plant. The rest of the cold utility is too low grade and will have to be provided by air cooling.

However, while the GCC lets us reason about the potential for heat integration, one must remember that the utility loads derived from a GCC are under the assumption of a perfectly integrated network where all available heat is recovered. Such a network is for the most part not feasible due to physical constraints, such as where the equipment is located within the plant, as well as for economic reasons, the most energy efficient network is not necessarily the most cost effective one. Secondly, one has to remember that the GCC depends on what streams that are included in the analysis, while the GCC above includes most of the heated or

cooled streams in the FT process, this subsystem has to be integrated with the existing plant as well.

To provide a more detailed description of the actual network that was devised is outside the scope of this report, however in Table 5 the total hot and cold utility of the final network is presented and broken down.

Table 6. Summary of the external utilities used by the FT-process.

Heating [MW]		Cooling [MW]	
Electrical heat	11.78	District heating water, after flue gas condensers.	18.32
Superheated high pressure steam	2.04	Air cooling	5.6
District heating return water.	0	District heating return water.	2.59
Total	13.813	Total	26.51

As discussed in relation to the GCC, most of the hot utility comes from the electrical heating and pre-heating of the reformer, but a certain amount of high-pressure steam from the boiler is still used. Note that, like what was shown in the GCC, the process makes use of more steam than what is shown in Table 5 but since a considerable amount of HP steam is produced internally by the syngas cooler the net amount necessary to draw from the boiler is reduced. Furthermore, as already discussed, the integration with the CHP plant allows for utilization of district heating water at temperatures lower than what is usually seen. In addition, the low temperature district heating return can be used to capture some additional low-grade heat.

7.4.3 Overall Plant balances

Given the FT - process characteristics above and the flows around the gasifier being known, what remains is to decide a new operating point of the now retrofitted CHP plant and examine the differences between an integrated and an unaltered plant. Figure 29 shows the net energy flows before and after the integration.

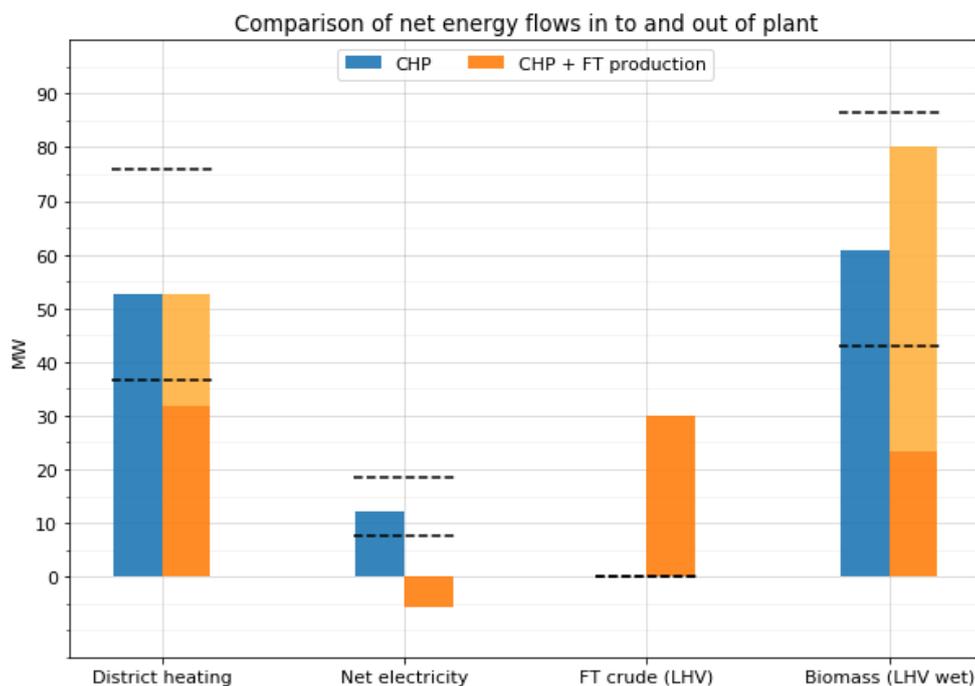


Figure 28: The net energy flows involved before and after the proposed retrofit. The dashed lines indicate an approximate span between low and high load conditions. The lighter orange hue shows the portion of the district heating that comes from the FT-process as well as the portion of the fuel that goes to the gasifier.

As already noted, the boiler must back down its production in order to leave room on the district heating loop for the heat given off by the FT-process. Furthermore, the HP steam required by the FT-process leads to a slight decrease in electricity production since this steam would otherwise have gone through the turbine. The larger change when it comes to the net electricity production is of course the high consumption used for heating the reformer though. For low or medium loads this makes the integrated system a net consumer of electricity. Note also from Figure 29, that while a sizeable portion of the total fuel is fed to the gasifier, a significant part of the energy in this stream ends up in the boiler due to the transfer of char from the gasifier as well as the combustion of tar, particles and off gases from the gas cleaning step. Evidence of this transfer of fuel from the FT-process to the combustor can be seen by noting the disproportionate amount of fuel feed to the combustor in relation to the amount of district heating still produced.

The comparison is made against a case where the district heating produced is in the middle of the previous load span. However, since this time a significant amount of heat is produced from outside the boiler, the amount of steam produced in the integrated case corresponds to the lower end of the load spectrum. It was decided not to decrease the load of the boiler in the integrated case further since this would put the model of the boiler outside the range of the data it was produced from. If necessary, to operate with a lower amount of district heating produced, parts of the heat from the FT-plant can be cooled off to the air. This would of course mean that the efficiency of the process is decreased but might still be feasible for parts of the year.

Table 7 shows a breakdown with five different load cases, of these the case “Summer” corresponds to such a case where cooling is needed in order to get down to the amount of district heating produced at low load before integration. The case “Low” is the same case as shown in Figure 29 with the steam production still at its lowest but with no cooling applied. The following cases have increasing steam production all the way up to in the “Winter” case where instead of comparing to an unintegrated case with the same amount of district heat produced, the integrated plant actually exceeds the previous district heating production at high load. However, to ensure that the boiler model still operates within the range of data it is produced from, the “Winter” case is capped by having the same amount of heat released in the furnace as the present boiler has during high load. To possibly make the relationship between the cases clearer Figure 30 shows the efficiencies of the 5 cases as a function of the district heating produced in each.

Table 7. Final breakdown of the key figures for the integrated plant. *Estimated values, corresponding to a constant district heating load before and after. ** Based on wet fuel with 45% moisture content. The omitted efficiencies in the Winter case is due to having no suitable “before” case to compare to as it lies outside the present district heating load range.

Quantity	Unit	Load condition				
		Summer	Low	Mid	High	Winter
Consumption of biomass, (\dot{Q}_{bio})	MW (LHV**)	80.2	80.2	93.2	106.1	117.3
Consumption of RME, (\dot{Q}_{RME})	MW (LHV)	2.8	2.8	2.8	2.8	2.8
Electricity produced, (P_{el})	MW	7.2	7.2	10.3	13.5	16.3
Net electricity consumption, ($P_{el,net}$)	MW	-5.6	-5.6	-2.4	0.7	3.5
District heating produced, (\dot{Q}_{dh})	MW _{th}	37.1	52.8	64.3	75.9	86.0
FT-crude produced, (\dot{Q}_{FT})	MW (LHV)	30.0	30.0	30.0	30.0	30.0
$\eta_{chem,marginal}$	%	81	155	154	154	-
$\eta_{el-chem,marginal}$	%	56	75	75	75	-
$\eta_{tot,before}^*$	%	104	107	108	109	-
$\eta_{tot,after}^*$	%	78	94	96	98	100
$\eta_{chem,marginal} = \frac{\dot{Q}_{FT}}{\dot{Q}_{bio} - \dot{Q}_{bio,before}}$						
$\eta_{el-chem,marginal} = \frac{\dot{Q}_{FT}}{(\dot{Q}_{bio} - \dot{Q}_{bio,before}) + (P_{el,net} - P_{el,net,before}) + \dot{Q}_{RME}}$						
$\eta_{tot} = \frac{\dot{Q}_{FT} + P_{el} + \dot{Q}_{dh}}{\dot{Q}_{bio} + \dot{Q}_{RME} + P_{el,demand}}$						

As evident from Table 7, the overall efficiency of the plant is the highest at high load. This trend stems partly from the boiler efficiency being the highest at high

load, but also from the fact that at high load a larger proportion of the total fuel is fed to the CHP-process. Since the CHP process has a higher efficiency than the FT-process the total efficiency is increased in proportion. Furthermore, a point worth noting is the sign change of the net electricity consumption between high and medium load, signifying that somewhere a bit before the high load case, the process starts producing more electricity than it consumes. The marginal efficiency reached of 75% is significantly higher than the energy efficiency reported for large scale standalone plants for FT production, which is around 40 - 55 % according to a recent review [1].

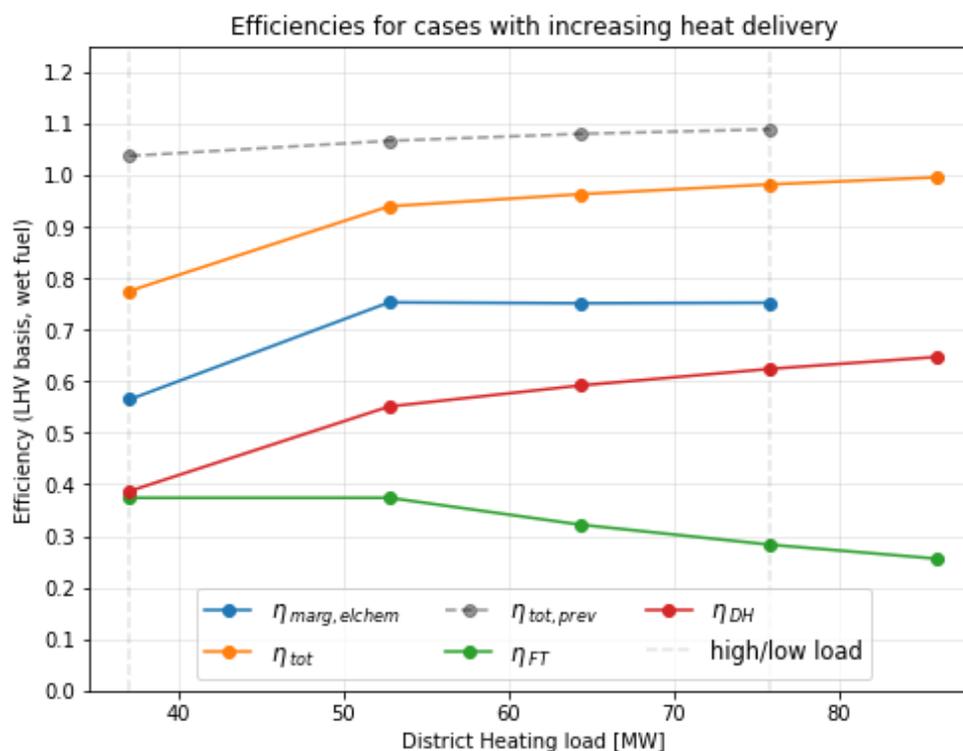


Figure 29. Shows efficiencies for the 5 load cases described earlier as a function of their heat delivery. Note the the light grey dashed lines, these are the limits for district heating produced during high and low load for the boiler before integration. The efficiencies η_{FT} and η_{DH} denote the energy in the FT-product and the district heat respectively divided by the total energy input to the plant.

7.5 ECONOMY

With the proposed incentive for renewable aviation fuel that currently is considered for implementation, a market segment with significant revenues will emerge. The incentive system is proposed to be based on a reduction quota system with a CO₂-fee of 6 SEK/kg. This fee that is higher than those for diesel and gasoline in the already established reduction quota system for road transport and will result in a high value for BioJet with low climate impact.

Tentative volumes and price levels as foreseen by the governmental committee 2019 are found in Table 8.

Table 8: Foreseen volumes and estimated price for biojet (SOU 2019:11)

	2021	2025	2030
Total volume of biojet (m3)	13 500	70 000	424 000
Total Energy in biojet (TWh)	0,13	0,67	4,07
Biojet price (SEK/liter)	18	14	12

For the case study where integration of 30 MW FT-crude with an existing CHP plant was examined, the Investment cost (Capital Expenditure, CAPEX) and Operational Cost (Operational Expenditure, OPEX) were estimated.

- Total investment cost: 980 MSEK
- Total production cost (excl. capital cost): 750-800 SEK/MWh

Significant investment costs are allocated to gas compression, cleaning and conditioning whereas costs for biomass supply system and raw gas production were found to be moderate thanks to the utilization of existing infrastructure and equipment. A significant portion of the CAPEX is allocated to utility systems, civil costs and different services, implying that the potential for cost savings following technology developments in the core process should not be overestimated.

Costs for catalysts and scrubber oil represent a significant part of the OPEX in addition to the costs for biomass and electrical power that, as can be expected, constitute the bulk of the operational costs.

7.6 BIOJET LIFE CYCLE ASSESSMENT

Based on the production and consumption figures for the studied co-production case, an independent LCA analysis was carried out by IVL: B 2377. The attributional LCA was performed in accordance with ISO 14040:2006 and ISO 14044:2006 with 1 MJ FT crude as the functional unit.

The LCA resulted in a calculated global warming potential (GWP) around 10 g/MJ FT crude depending on the boiler load. A somewhat comparative value for fossil aviation fuel would be 89 g/MJ [2]. However, remember that the FT-crude has some further refining needed before it can be considered aviation fuel. Figure 30 shows the calculated GWP for 3 different cases, corresponding to the “Summer”, “Low” and “High” cases above. As can be seen from the figure, the amount of district heating produced carries strong weight in the LCA. Furthermore, significant contribution to the GWP-value was found to be caused by transport of biofuel to the plant as well as by the consumption of tar scrubbing oil. The latter was also found to significantly contribute to the Eutrophication Potential as well as the Acidification Potential. Hence, identifying alternative scrubber oils can contribute to further reduced environmental impact from the Biojet fuel.

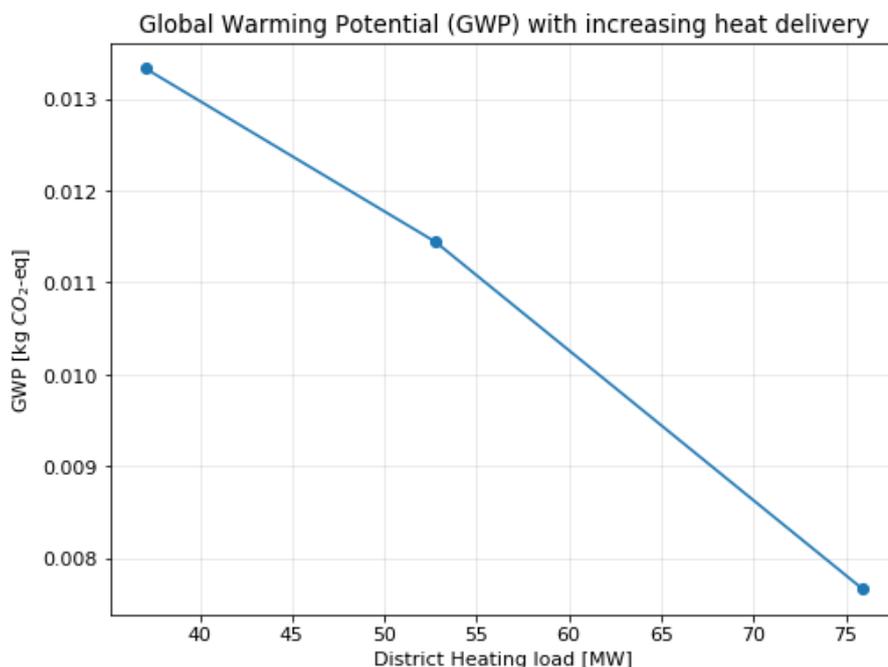


Figure 30. Global warming potential for three different load conditions. The three points from left to right correspond to the Summer, Low and High case respectively. The GWP is calculated excluding biogenic carbon emissions.

7.7 CO₂ UTILIZATION AND POTENTIAL FOR ELECTROFUELS

The potential of utilizing CO₂ produced in the process to enhance the FT-production was analyzed. In the co-production process described in this report, CO₂ will be present in different gas mixtures throughout the plant, including:

- Flue gases.
- CO₂ separated from synthesis gas.
- Product or synthesis gas.

The flue gas is produced from combustion of biomass and off-streams from the FT-process. The CO₂ in the flue gas is diluted with N₂, H₂O and other flue gas emissions. Therefore, rather extensive measures are required capture and utilize the CO₂ from the flue gas. Despite this, there is currently a strong project and technology development to enable feasible CO₂ capturing from flue gases. A good example is the development by Stockholm Exergy which are aiming for negative CO₂ emissions by capturing and storing CO₂ from the flue gas from biomass combustion (BECCS). In the proposed process outline roughly 81-87% of the biomass carbon will form CO₂ in the flue gas through combustion of char, off-gases and byproducts from processes downstream of the gasifier, and additional biomass feed to the boiler.

By integrating the biofuel production with the heat and power production in the manner suggested here, a single system for separation can be used to separate CO₂ from the flue gas from both the CHP and the biofuel production process, potentially producing a biofuels with negative climate impact.

Biomass is rich in oxygen and poor in hydrogen compared to FT-crude. This means oxygen needs to be removed from the process as CO_2 , thus sacrificing some of the carbon from the biomass to get rid of the oxygen. CO_2 forms in the process through the water gas shift reaction and with an activated amine scrubbing system it can be separated from the process with very high selectivity and very low concentrations of contaminants. This stream corresponds to about 4-6% of the carbon in the biomass and with minor treatment can be used as purge gas in the system, utilize in other processes or stored.

The oxygen could also be removed as H_2O , but this would require extensive addition of hydrogen and would at the same increase the utilization of the carbon in the product gas. Hydrogen can be added to the process as pure H_2 or through steam reforming of hydrocarbons with a H/C ratio higher than the syngas, such as CH_4 . By reforming hydrocarbons in the process the carbon in the hydrocarbons can be utilized in the production, but the high hydrogen content also increase the utilization of the carbon in the syngas components formed during gasification as more of the oxygen in the syngas can be removed as H_2O instead of CO_2 . How potent this synergy effect is will depend on the H/C- and O/C-ratio of the syngas and is strongest if the H/C-ratio is low and the O/C-ratio is high. With a high H/C-ratio adding more hydrogen requires a reversed water gas shift to be utilized but could even cause accumulation of hydrogen in the process which is not desired. Thus, the impact of reforming hydrocarbons for enhanced utilization of the syngas will be case specific depending on product gas composition, reforming technology and gas recirculation strategies among other factors.

As described in section 4 and 5.4 different reforming technologies can be applied to reform the hydrocarbon in the product gas such as a regular gas fired SMR or an electrically heated eSMR. If applying eSMR, electricity is used to keep the temperature of the reforming process instead of burning part of the gas. Thus the gas can be used for production instead of forming additional flue gas. This is a highly efficient alternative for electrofuel production as electricity is used to maximize the utilization of the gas produced from the biomass, this alternative was therefore selected in the present work. Another way of introducing electricity to the process is to produce H_2 and add it to the gas conditioning section of the plant, this was further discussed in section 4.

Results from present study shows that about 9-13% of the carbon in the total flow of biomass to the process will be reallocated into FT-crude with the suggested co-production. This corresponds to about 18% of the carbon in the fuel fed to the gasifier and as much as 50% of the carbon in the additional biomass feed to the co-production plant compared to a standalone CHP with same district heating production. This can be compared with results from a stand-alone plant for production of advanced biofuel, GoBiGas, where CH_4 was produced from biomass and about 30% of the carbon in the biomass formed reallocated to the product[11].

This shows how a co-production plant can be implemented with a very high marginal utilization of carbon in the biomass reallocated to FT-crude based on a to a CHP-plant with the same district heating production. It also shows that the BECCS technology could be implemented at a co-production plant to store large amounts of CO_2 . Further, present work has identified several implementations for

electro-fuel production which can contribute to an efficient utilization of the biomass and bring down both production and investment products. Further investigations should be conducted to identify the most techno-economical implementations for electro-fuels.

7.8 POTENTIAL IN SWEDEN

A large number of solid fuel fired combustion plants are found in Swedish industry- and district heating sector. Approximately 10 million tons of non-refined woody biomass are fire annually in several hundred boilers to fulfil the demand of heat in buildings and industrial processes. Among the largest boilers, FB technology is the most widely employed technology and since 1980 more than 100 boilers have been installed, Figure 1.

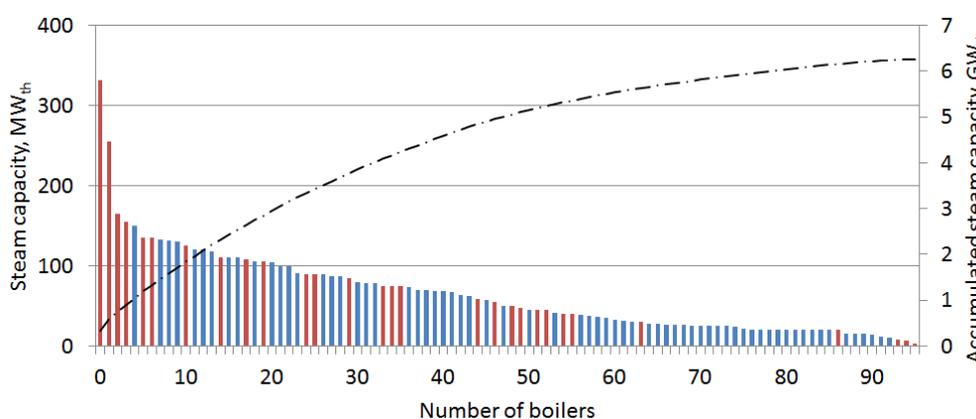


Figure 31: Boilers of FB type in Sweden installed 1980-2015 (Red=CFB, Blue=BFB) [41]

Many of these boilers can be retrofitted for co-production of biofuels. Such a broad implementation of the co-production concept would enable significant volumes of biofuels to be produced. As described in this report, the efficiency of the biofuel production is dependent on a utilization of the excess heat and residual streams from the process. Hence, even if very high accumulated production capacity can be achieved from a strict capacity-perspective, the excellent efficiency and GWP-performance found in the studied case can only be achieved during the periods when the boiler operates above a certain minimum load. In plants with several boilers, the biofuel production should be integrated with the boiler having the highest number of operating hours over the year.

Industrial boilers generally have a higher average boiler load than district heating boilers, and this emphasizes the relevance of co-production for these plants.

Many plants in the district heating sector utilizes waste incineration in combination with biomass combustion. Due to practical reasons and the favorable price of different waste assortments, the waste incineration boilers are normally operated as base load while the biomass boilers at these plants are operating at part load during the fire-season. Hence, combined plants operating both waste-fired and biomass-fired boilers might face difficulties in realizing highly efficient

co-production of biofuels within their present strategic operations. However, the recently decided tax on waste incineration is aiming to reduce the amount of incinerated waste and might hence improve the long-term viability of biofuel co-production in these plants. Furthermore, seasonal energy storage that has gained interest in the district heating sector during the last years is a technology that can further strengthen the co-existing of waste incineration and biofuel co-production.

8 Concluding Remarks

One major conclusion from this work is that introduction of electricity as an additional feedstock besides biomass has the potential to increase the efficiency and capacity of advanced biofuel production significantly. Furthermore, the electricity does not necessarily need to be added in the form of hydrogen from electrolysis, but simply replacing energy demands that traditionally are satisfied through fired heaters can have a big impact. Further research and development into this crossover between bio- and electro-fuels is justified.

The costs associated with the proposed plant are in line with those presented for significantly larger facilities. The estimated capital costs for large scale (200MW) standalone plants are in the range 46 – 76 EUR/MWh, according to a recent IEA report [1]. The specific capital cost for the significantly smaller plant studied in this project was estimated to be within this range however highly dependent on assumptions regarding interest and investment subsidies. From the same IEA report the total production cost including feedstock- and operating costs is estimated to 75 – 144 EUR/MWh. Once again, the corresponding cost for this facility of 75-80 EUR/MWh is well within the estimated range for large scale plants.

By integrating biofuel production like this with a district heating plant, byproducts and low-grade heat can be utilized to great extent. The value of this synergy can now be seen in that the marginal efficiency of the proposed plant of 75% is significantly higher than energy efficiency seen in standalone facilities for FT production, around 40 – 55 % [1]. Even during a low load case, when the plant has to operate at reduced efficiency, the obtained efficiency of 56% is in the higher end of the range presented for standalone facilities.

The studied concepts can be implemented in many CHP plants in Sweden and elsewhere, and can be considered commercially viable given a price for FT-crude exceeding approximately 1300 SEK/MWh. In order to secure such a product price the planned and already decided reduction quotas for air- and road transport should be kept and developed.

9 References

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CO-GENERATION OF BIOJET IN CHP PLANTS

Co-production of biofuels poses an opportunity for the CHP sector, where falling revenues for electricity in combination with stagnating or falling demand for heat in the district heating sector emphasizes the need for business development.

In this report, co-production of FT-crude for use as an BioJet intermediary has been examined. The results show that a process based on a gasifier integrated with a biomass boiler followed by subsequent gas cleaning and synthesis, can produce FT-crude with a marginal efficiency as high as 75 percent thanks to utilization of excess heat and residual streams from the FT-process in the CHP plant. The specific investment costs and operational costs calculated in the project are on par with recently published costs for significantly larger plants. Co-production can therefore enable commercial biofuel production in a moderate scale.

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