Biomass is an important energy source and can provide sustainable materials, fuels and chemicals. CHP plants are expected to play a key role in further increasing the use of renewables in the Finnish energy system. This thesis provides the thorough investigation of three biomass upgrading processes integrated with a municipal CHP plant considering energetic, environmental and economic performance. The thesis also answers the question of how the energetic and environmental performance of future polygeneration plants can be assessed in an objective way. It was shown that all three processes can be integrated with CHP plants and that all improve the energetic and environmental performance. Under the assumptions made in this work, the biomass fast pyrolysis process shows the highest economic potential. The thesis also concludes that for system level analysis primary energy should be the preferred tool for process assessment because it allows fair process comparison by including the global impact of the applied technologies in a simple manner.
Improving Municipal CHP Production Efficiency by Integrating Biomass Upgrading

Thomas Kohl

A doctoral dissertation completed for the degree of Doctor of Science in Technology to be defended, with the permission of the Aalto University School of Engineering, at a public examination held in Auditorium 216 at Aalto University on 25th of November 2016 at noon (12 o'clock).

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Abstract

Reducing CO2 emissions in order to abate climate change is one of the biggest challenges currently facing societies. The efficient and sustainable use of biomass conversion technologies for energy purposes is one important means of mitigating CO2 emissions while simultaneously reducing the need to import energy. However, compared to fossil fuels, biomass conversion per se suffers from lower energetic conversion efficiencies. New well-designed, deeply heat and material-integrated processes can help to overcome this obstacle. As an abundant but widely scattered resource, biomass should primarily be used locally or else upgraded before transport to central installations. Biomass-based, combined heat and power (CHP) production is an established and efficient technology that has proved it can compete with the conventional methods of energy generation under current tax regimes. However, increased use of biomass raises the price for limited biomass feedstock, and, in the future, it might hamper the affordability of municipal biomass-fired CHP plants. In order to react to the expected increase in competition for restricted biomass resources, communal CHP plants should be integrated with biomass upgrading processes that add valuable products to the portfolio.

The objective of the thesis is to investigate the influence of integrating three different biomass upgrading processes, pelletising, torrefaction and fast pyrolysis, with a municipal CHP plant. In particular, their influence on important operational parameters, the energetic and environmental performance of the plant as well as the plant’s finances, should be taken into account. With respect to the latter aspects of system’s performance, a concise and significant assessment methodology that makes it easy to compare the processes yielding multiple products is developed and assessed in this study.

The study shows that all three integration options are quite possible within the operational limits of the CHP plant. By utilising free boiler capacity during times of low district heating demand, high bio-product yields and greater district heating output can be realised, which in turn leads to improved primary energy (PE) efficiency and reduced CO2 emissions. Despite its lower energetic and exergetic efficiency, the integration of fast pyrolysis demonstrated the best economic performance.

The thesis also concludes by arguing that the results obtained via thermal efficiency analysis do not add any new information to the results obtained via exergy and PE analyses. Exergy and PE analyses should be the preferred means for process assessment.

Keywords Biomass, Combined Heat and Power, District Heating, Fast Pyrolysis, Torrefaction, Wood Pellets, Energy Efficiency, Exergy

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“Change is the only constant in life.”
(Heraclitus)

Preface

This work was carried out in the research group of Energy Technology and Environmental Engineering at the Helsinki University of Technology which later became Aalto University. During this period I gratefully received project funding and several grants from Fortum Oy, Ekokem Oy, Gasum Oy, the Academy of Finland, Tekes and Aalto University’s School of Engineering without which this work would not have been possible.

Many people helped to shape and develop my dissertation and I owe them my deepest gratitude. To name three specifically: my former, now retired supervisor Carl-Johan Fogelholm who gave me the opportunity to start my work and research career in Finland in a liberal, free and respectful working environment. My current supervisor and former colleague Mika Järvinen who helped me a great deal to focus my work and finally bring it to a conclusion. Thirdly I thank my instructor Timo Laukkonen who was probably most involved with my work and always had time to answer and discuss my sometimes not so thought-through questions patiently and humorously.

Special thanks to my co-authors Mari Tuomaala, Kristian Melin and Moises Teles for accelerating my learning by sharing their knowledge and experience. The same accounts for the numerous BSc and MSc candidates I had the pleasure to work with. Thanks also to Henrik Thunmann and Esa Vakkilainen for their quick and thorough pre-examination and invaluable comments. However, work isn’t all but nothing without coffee breaks. We had great coffee breaks in which many work-related problems were solved—or dissolved?; mostly because we did not talk about problems but forgot about them for a while. Thank you all for having me at the coffee table!

Living abroad is not always easy: Thank you to my parents, family and friends in Germany, who were always there for me when I needed someone to share the oddities, curiosities, normalities and sophistications of living and working in Finland; my friends in Finland did a great deal too! The most rewarding thing after long working days was, is and will be to see the growth and development of my beloved children. I thank Erik, Lisa and Mio for their great ability to distract me from the problems and challenges of working life and for showing me every day that there are by far more important things in life than work! Last but not least, the deepest bows, thanks and gratitude go to my wife, Jenni, who always supported and backed me up during this long journey to reach my goal. I love you!

Helsinki, 31st October 2016

Thomas Kohl
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List of Publications

The doctoral thesis consists of a summary of the following publications:

**Publication I:**
Kohl T, de Moya CG, Fogelholm CJ.

**Publication II:**
Kohl T, Laukkanen TP, Järvinen MP.
*Integration of biomass fast pyrolysis and precedent feedstock steam drying with a municipal combined heat and power plant.*

**Publication III:**
Kohl T, Laukkanen T, Järvinen M, Fogelholm CJ.
*Energetic and environmental performance of three biomass upgrading processes integrated with a CHP plant.*

**Publication IV:**
Kohl T, Teles M, Melin K, Laukkanen T, Järvinen M, Park SW, Giudici R.
*Exergoeconomic assessment of CHP-integrated biomass upgrading.*

**Publication V:**
Kohl T, Laukkanen T, Tuomaala M, Niskanen T, Siitonen S, Järvinen M, Ahtila P.
*Comparison of Energy Efficiency Assessment Methods: Case Bio-SNG process.*
Author’s Contribution

Publication I:
**Primary Energy Efficiency Consideration of Biomass-to-Liquid Systems Yielding Fischer-Tropsch Fuel.**
The candidate did all calculations and the writing based on the references collected by author de Moya. Co-author Fogelholm contributed by discussing and commenting on the paper.

Publication II:
**Integration of biomass fast pyrolysis and precedent feedstock steam drying with a municipal combined heat and power plant.**
The candidate developed all of the simulation models as well as the assessment method. He also analysed the results and wrote the paper. Co-authors Laukkanen and Järvinen contributed by discussing and commenting on the paper.

Publication III:
**Energetic and environmental performance of three biomass upgrading processes integrated with a CHP plant.**
The candidate developed all of the simulation models as well as the assessment method. He analysed the results and wrote the paper. Co-authors Laukkanen, Järvinen and Fogelholm contributed by discussing and commenting on the paper.

Publication IV:
**Exergoeconomic assessment of CHP-integrated biomass upgrading.**
Co-authors Teles and Melin provided exergy stream data and cost data for the upgrading processes and wrote the parts of the paper related to their work. The candidate developed all of the simulation models and was responsible for calculating the exergoeconomic data as well as for analysing it. All of the co-authors contributed by discussing and commenting on the paper.

Publication V:
**Comparison of Energy Efficiency Assessment Methods: Case Bio-SNG process.**
Co-author Laukkanen provided parts of the simulation model (air-separation, acid gas removal) and carried out the exergy analysis. Co-author Tuomaala conducted the general literature research and sharpened the objectives of the paper. The basic process layout was based on co-author Niskanen’s MSc. Thesis, which also provided the technical references. The candidate carried out energy and primary energy analysis and wrote most of the paper. All co-authors contributed by discussing and commenting on the paper.
1. Introduction

1.1. Background

“Anthropogenic greenhouse gas emissions have increased since the pre-industrial era, driven largely by economic and population growth, and are now higher than ever. This has led to atmospheric concentrations of carbon dioxide, methane and nitrous oxide that are unprecedented in at least the last 800,000 years. Their effects, together with those of other anthropogenic drivers, have been detected throughout the climate system and are extremely likely to have been the dominant cause of the observed warming since the mid-20th century.” [1]

According to the International Energy Agency, global CO₂ emissions in 2030 will need to be cut by 40% compared to 2012 levels. This would allow the world’s energy consumption to grow by approximately 17% (based on the 2012 level) and would still limit the increase in the average global temperature to 2°C (the so-called 450-policy scenario). Experts assume that meeting this target will help us avoid the most severe and irreversible effects of global warming on our environment, society and economy. The World Energy Outlook 2009 [2] defines the most important measures for reaching the required CO₂ levels to include improved energy efficiency and the increased use of renewable energy sources, which account for approximately 57% and 23% of emission reductions, respectively. [3]

In the European Union (EU), such considerations led to the formulation of the so-called “20–20–20” targets (European Council and Parliament, 2008). They have established three key objectives for the year 2020: a 20% reduction in EU greenhouse gas emissions from their 1990 levels; raising the share of EU energy consumption generated from renewable resources to 20%; and a 20% improvement in the EU’s energy efficiency. The targets were legally implemented by, e.g. EU directives on the promotion of energy from renewable sources (2009/28/EC) and the promotion of cogeneration (2004/8/EC), which call for sustainably increased and efficient conversion of local biomass for transport and energy applications. The goal is to reduce CO₂ emissions and import dependency of the energy supply by sustainably activating more renewable, local energy sources.

Biomass as an energy source is carbon-neutral due to its CO₂ uptake while growing by means of photosynthesis. However, biomass-based technologies are distinguished by lower energy conversion efficiencies when compared to technologies that apply fossil fuels. Energy conversion efficiency is the ratio of the useful energy output of a conversion process compared to the energy input. In the case of biomass, the lower energy conversion efficiencies can primarily be attributed to the following three aspects. First, biomass has a lower heating value due to its low carbon and high oxygen content, which are usually in the range of 50 wt% and 40 wt%, respectively (coal and oil show carbon contents of 70 wt% and above). Second, the heating value is further reduced by high as-received water contents of up to 60 wt% for solid biomass. Third, the ash often contains larger amounts of low-melting compounds with a high corrosion and
clogging potential (silica, chloride, alkali metals and their compounds), which makes it necessary to reduce the temperatures of the heat exchange equipment [4, 5].

Thus, the goal of increasing the use of biomass for energy purposes on the one hand and increasing energy efficiency on the other make using abundant but restricted biomass resources as efficiently as possible a challenge.

In Finland and Sweden, apart from being used in the wood and pulp & paper industry, woody biomass is mainly used for heat and power production. Biomass-based combined heat and power (CHP) production, or cogeneration, is an established and efficient technology that has proved to be competitive with conventional energy generation under current tax regimes that account for the high CO₂ emissions related to fossil fuels. However, the increased use of biomass raises the prices for limited biomass feedstock (see Fig. 1) and might hamper the economy of municipal biomass-fired CHP plants in the future.

![Figure 1. Recent net energy price development in Finland (data extracted from Refs [6, 7]).](image)

As CHP plants can provide both the source of heat needed for thermal conversion of biomass as well as the district heating network (DHN) as a sink for process waste heat, it would be beneficial to integrate basic biomass upgrading processes like fast pyrolysis, torrefaction and pelletisation with CHP plants. It is anticipated that products, such as wood pellets (WP), torrefied wood pellets (TWP) and pyrolysis products could be produced with higher energy conversion efficiency when integrated with CHP plants as when produced in stand-alone facilities. We further would expect that the new products added to the plant portfolio would create more income and profit. Further, the upgraded products would have a higher energy density and, hence, allow for more efficient transport to central installations. The technologies applied for such upgrading are relatively simple and robust, and thus investment costs and system complexity both are expected to be at a reasonable level. This would also make
it interesting for local small to medium scale solutions of up to 50 MW thermal (MWth) [8]. With respect to such integrated usage approaches, there is the further challenge of evaluating complex systems that yield more than one product in terms of whether or not they increase energy efficiency. The upgraded products can further be processed to form, e.g. biodiesel, ethanol, methanol, hydrogen or other chemicals [9], but they can also be directly combusted for power and heat production (see e.g. [10, 11]). This would make the CHP plant a supplier to the currently establishing bioeconomy, which focuses on producing new materials, chemicals and transport fuels from biological resources. At the same time, the new product portfolio would stay independent of developments in the final product markets which tend to be influenced heavily by societal and political decisions.

1.2. Outline

Chapter 2 provides a literature review on the use of biomass and on the conversion technologies and assessment methods discussed in this work. With that background, the objectives and the scope of the thesis are defined in Chapter 3, as are the major restrictions and assumptions of the study. Chapter 4 describes the developed simulation models and the applied assessment methods.

Chapter 5 presents and discusses the major results of the thesis and, finally, in Chapter 6, conclusions are drawn and the scientific significance of the study as well as future work to be done are highlighted.
2. Technology and Literature Review

2.1. Lignocellulosic Biomass

According to the definition of the European Directive 2009/28/EC on the promotion of the use of energy from renewable sources, “biomass” means the biodegradable fraction of products, waste and residues from biological origin from agriculture (including vegetal and animal substances), forestry and related industries including fisheries and aquaculture, as well as the biodegradable fraction of industrial and municipal waste.”

Biomass is abundantly available and its use currently contributes approximately 10% [2] and 7.3% [12] to the world’s and EU’s primary energy (PE) consumption, respectively. Experts estimate that the 2030 potential will be approximately 15% of the EU’s PE demand [13] and 25% of the world’s 2050 PE demand [14], respectively.

There are two major advantages of using biomass for energy purposes: first, it is considered carbon-neutral when used sustainably and, second, it is the only renewable carbon source with the potential to replace products currently produced by refining fossil-based hydrocarbons, such as crude oil, coal and natural gas. In addition, biomass energy can be stored and used on-demand for base-load energy generation. This stands in contrast with other renewables, which fluctuate greatly and depend on the micro- and macro-climate, such as wind and solar power [4, 5, 13].

The major environmental concerns related to the extended, large-scale use of biomass resources are the rather large collection areas, which also cause high transport costs, typically 20–40% of the fuel costs [15]. In addition, if the sustainability of biomass use is not taken into consideration, extended biomass usage can lead to soil erosion and compaction, decreased biodiversity, nutrient leaching and land-use competition [5, 13]. However, this discussion does not pertain to the focus of this thesis and so henceforth, it is assumed that the biomass used in the described conversion processes could be harvested and supplied on a sustainable basis.

As the name indicates, ligno-cellulosic biomass consist of celluloses (40–50 wt%), hemicelluloses (20–40 wt%) and lignin (10–25 wt%) [4, 5]. This definition excludes aquatic biomass, manures and sewage sludge as well as those parts of plants that contain high amounts of sugar, starch or oil; thus, lignocellulosic biomass can be described as the fundamental entity in all forestry (i.e. woody) and agricultural herbaceous biomass. It further indicates that lignocellulosic biomass is not utilised for food and feed production and, thus, it also comprises the feedstock for the so-called 2nd generation biofuel production [9]. For biofuel production and energy generation, special focus is on energy crops and biomass residues or wastes, which cannot be used for or form by-products of material or food production processes, such as industry wood, pulp and paper, and grain production. A considerable share of such residues is currently unused, but a specific value for the global or EU potential could not be found in the existing literature.
With respect to energy utilisation of biomass, the following properties play a crucial role: ultimate and proximate composition, heating value, moisture content, heating value, ultimate and proximate composition, ash and alkali content. The moisture content of the as-received lingo-cellulosic biomass can be well above 50 wt%, which is usually also the upper boundary for efficient thermochemical conversion. Higher moisture contents greatly impede the energy conversion efficiency, and wet biochemical or hydrothermal processing should be the preferred means of conversion [4, 16].

The ultimate composition of biomass derives from the weight fractions of the major chemical compounds: carbon, hydrogen, oxygen, nitrogen and sulphur. Table 1 shows the typical ranges for the weight fractions. The chemical composition determines the products of any applied conversion process, the oxygen demand for combustion and the heating value. The heating value defines the energy content of the biomass and is usually given on a dry, ash-free basis. It can be further distinguished between the lower heating value (LHV) and the higher heating value. The latter also contains the latent heat of the water vapour. Often, this latent heat is not utilised, and thus, the LHV is often the appropriate value for comparing different fuels and calculating energy efficiency. The higher heating values of lignocellulosic biomass vary between 17 and 21 MJ/kg. When given on a wet basis, the moisture reduces the heating value proportionally. The proximate analysis defines the reactivity of a fuel. The volatile matter is released upon heating at 950 °C, while the fixed carbon remains in place. Further combustion gives the ash content. The typical values given in Table 1 show that the biomass constitutes a highly reactive fuel. For biomass processing, ash and alkali content, although small, are important values because they interact with one another when forming low-melting, sticky compounds, which tend to block and corrode heat exchange equipment [4].

Table 1. Typical range for the ultimate and proximate analysis of lignocellulosic biomass [4].

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<th>Carbon [wt%]</th>
<th>Hydrogen</th>
<th>Oxygen</th>
<th>Nitrogen</th>
<th>Sulphur</th>
<th>Fixed Carbon</th>
<th>Volatile Matter</th>
<th>Ash [wt%]</th>
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<td>Range</td>
<td>41–55</td>
<td>5–7</td>
<td>38–43</td>
<td>0–0.7</td>
<td>0.1</td>
<td>10–19</td>
<td>66–82</td>
<td>0.8–4.5</td>
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2.2. Cogeneration of Heat and Power from Biomass

Cogeneration, or CHP production, is the conversion of chemical or thermal energy sources into electricity and heat in a simultaneous process. Sources can be fossil or renewable solids, liquid or gaseous fuels, and solar or geothermal heat. Primary energy conversion technologies consist of combustion, gasification and fuel cell technologies, to name a few. Secondary equipment can be complemented by steam and gas turbines or steam and internal combustion engines, which all serve the same purpose of converting high-temperature heat into mechanical and finally electrical energy. The remaining steam enthalpy is extracted for heating industrial processes or for to provide heating networks.

Cogeneration has been known for long to be an efficient energy conversion technology because the joint generation of power and heat requires less fuel input than separate generation processes [17]. Thus, biomass-based
cogeneration is an explicitly mentioned technology that can be used to achieve the EU’s 20–20–20 goals [18].

Salomón et al. have provided a good review of biomass cogeneration technologies [19]. Currently, the Rankine cycle, which consists of a boiler and steam turbine, is the technology applied most frequently for solid biofuels, such as lignocellulosic material [17, 19], in Sweden and Finland. Boiler equipment has been used since the beginning of industrialisation in the 18th century, and hence, the Rankine cycle is a well-developed, reliable and fairly inexpensive technology that is generally available throughout Europe and the rest of the world.

Gasification-based cogeneration and the benefits it offers in terms of increased power generation is currently being intensively studied because of its potential to also produce liquid or gaseous fuels. However, it is still at the early stages of commercialisation [17, 20–22]. The technology’s tremendous potential to produce synthetic natural gas (SNG) has been analysed in Publication V, and consequently, it will be briefly explained in the following paragraphs along with conventional cogeneration based on combustion.

For gasification-based biofuel processes, the lignocellulosic biomass is dried to approximately 15–20 wt% of moisture, and subsequently, gasified in a steam/oxygen atmosphere. An air separation unit usually provides oxygen. The heat for the reaction is generated via partial oxidation of the raw material. The product gas consists of hydrogen, carbon monoxide, CO₂, water, methane and some impurities, such as tars and hydrogen sulphide. As the impurities will harm the downstream equipment, further treatment is necessary. Tars are cracked inside the gasifier and/or in an additional catalytic tar reformer. Tar-free gas is filtered to remove any particles and heavy metals. The filtered gas enters a water scrubber to remove any ammonia, hydrogen cyanide and chlorides. Depending on the desired product, after removing the acid gas, well-known synthesis processes can be applied to form, for example, methane, ethanol or Fischer-Tropsch diesel [23, 24]. In addition, it is also possible to achieve cogeneration via gas turbines and internal combustion engines [19].

For industrial cogeneration via combustion, two boiler technologies are for the most part available: grate-fired and fluidised bed boilers. The latter boiler technology can be divided into bubbling and circulating fluidised bed boilers. The major advantages and disadvantages of each technology have been described quite well in the existing literature (e.g. [25]). Whereas grate-fired boilers are usually applied for a firing power below 20 MWth, in Finland and Sweden the most commonly applied combustion technology for lignocellulosic biomass fuels in the size range between 20 and 50 MWth is the bubbling fluidised boiler. It is characterised by a high degree of fuel and moisture flexibility along with a high level of combustion efficiency; also, the easy to control, relatively low combustion temperature of approximately 850 °C is well suited for biofuels with a high amount of critical inorganic compounds [19, 25, 26].

Fig. 2 presents a flowsheet of a typical small-scale CHP plant. Apart from the flowsheet, Fig. 2 provides necessary stream information and a list of the
equipment used. In addition, the highlighted areas define the functional units used for exergy analysis (see Section 4.2.3).

The boiler (3) walls are usually equipped with evaporator and superheater tubes. Together with the superheater tubes (4), which are situated in the freeboard, they transfer the radiative heat of the hot combustion gases to the steam side. The flue gases are then led over the convective heat exchange areas of the economiser (8) and air-preheater (2), where they cool down to approximately 120-140 °C before they enter the stack or are condensed in cases in which flue gas condensation is applied [27]. The hot steam leaving the superheater is expanded in a steam turbine (5). Typical steam values for biomass cogeneration plants are 70–80 bar and 400–500 °C. In the turbine, the steam is expanded and its enthalpy is converted into the rotor’s rotational energy and further converted into electricity within an alternator. The enthalpy still remaining in the steam exiting the last stage of the backpressure turbine which is condensed (6) at the outside temperature-dependant pressure is used to provide district heating (DH) at varying temperatures. If attached to an industrial process, steam can be extracted at higher pressures; process steam can then be provided.

Figure 2. Typical layout of a small-scale (<20 MWth) biomass cogeneration plant. The highlighted areas describe the functional units as defined for the exergy analysis.

Usually the economics of such CHP plants depended heavily on the power-to-heat ratio, which is the ratio of the generated electricity to useful heat. This is because, for most of the time, electricity had a higher market value than heat. Since municipal CHP plants are driven according to the heat demand, it has been considered advantageous that they would produce as much electricity as possible at given heat load; consequently the electrical efficiency of the process should be as high as possible. Savola [28] found that steam reheating, feedwater preheating and double-stage DH exchangers are the most economical means of achieving this target. However, likely because of the related costs, such measures only seem to be applied in new installations; older plants are not often retrofitted. On the other hand CHP plants should also be able to react flexibly to changing electricity prices in order to ensure most profitable operation [29, 30]. This is especially the case in times of exceptionally low electricity prices of
approximately 30 €/MWh as currently observed on the Nordic electricity spot market Nordpool\(^1\). However, for CHP production from biomass in Finland, a prime calculated based on the current price of CO\(_2\) emission certificates and the peat tax, is paid on top of the spot price. For plants smaller than 6 MW\(_{el}\) a feed-in-tariff of 83.5 €/MWh is granted. Nevertheless, such subsidies are subject to change and the CHP plant should preferably be able operate profitably without the need of subsidies in any form.

It is also important to notice that the operation of the CHP plants as well as the expansion of their use strongly depends on effectively using the operation time [31]. Especially in the Nordic countries, which have large differences in the seasonal heat demand, biomass-fired municipal CHP plants operate for long amounts of time at only part-load and are often shut off for a considerable amount of the year. **Fig. 3** shows a typical heat duration curve of a district heating network (DHN) under Nordic conditions. It is evident that the full-load operation of a plant is only possible approximately 70 days of the year; the plant will be operated under less beneficial part-load conditions approximately 145 days per year and will be eventually shut down for approximately 150 days, which clearly exceeds the time required for yearly maintenance. In other words, such plants offer a huge potential for available unused boiler capacity that could supply heat for thermo-chemical conversion processes, such as gasification or pyrolysis. In such a scenario, the heat required for thermochemical conversion would be supplied by the fuel boiler and low-grade heat potentially rejected at the cold end of the conversion process could be utilised in the DHN, thus increasing the overall efficiency of the plant. Such integration could potentially increase the capacity utilisation of the boiler and add additional biomass product to the plant portfolio, which could increase competiveness and allow CHP plants to be a part of the currently establishing bioeconomy.

![Figure 3. Typical heat duration of a DHN under Nordic conditions.](image-url)
2.3. From Cogeneration to Polygeneration

Based on the large number of recent publications on the topic, there is great research interest in converting biomass into high-grade fuels and chemicals. It seems clear that integrating such processes with CHP plants (forming so-called polygeneration plants) is very beneficial because it increases the overall energetic efficiency and mitigates CO₂ emissions [20–22, 32–37].

Other studies [38–40] have found that integrating the gasification of biomass for syngas production with a CHP-based DH system is technically feasible and economically advantageous.

Eriksson and Kjellström [41] concluded that wood-based ethanol production is most efficient when waste heat is used in an integrated CHP plant, but CO₂ mitigation costs were lowest for electricity-only generation in a combined cycle process. One concern also has to do with the amount of heat generated in an economical-sized ethanol plant requiring a DHN with a large amount of free capacity.

Starfelt et al. [42] stated that retrofitting a CHP plant by integrating it with bioethanol refining would increase its operation hours, electricity generation and system efficiency, even in a short-term perspective [43]. Song et al. [44] integrated biomass drying with the polygeneration system investigated by Starfelt et al. [42] and found that it increases overall efficiency and electricity generation by 3.1% and 5.5%, respectively. Other work of Song et al. [45] found that integrating pellet production with an existing CHP plant increases power production by up to 97% and significantly reduces global CO₂ emissions.

Wahlund et al. [10] showed that retrofitting an existing communal CHP plant by integrating it with pellet production results in increased local biomass utilisation as well as increased operation hours, electricity production and efficiency and, along with decreased transport costs and CO₂ emissions of the considered energy system, greater income for the DH company. Researchers often suggest that advanced thermochemical biomass conversion and biofuel production processes can be integrated with power and heat production in order to achieve greater energetic efficiency. Although it would be possible in most cases to introduce co-generated renewable electricity to a well-liberated and -unbundled market, for heat, due to its thermo-physical properties, there is only a local market with a well-defined demand available. Therefore, it remains doubtful whether the, often considerable, amount of base-load heat produced in large-scale biofuel production facilities can be utilised, and hence, it should most likely not be accounted for in efficiency calculations. Vice versa, integrating existing DHNs with biomass conversion and biofuel production processes would ensure that the heat rejected at the cold end of the various processes would for the most part be used.
2.4. Upgrading of Biomass

It is generally acknowledged that complex, large-scale biorefineries will need to be built due to the economics of scale. However, bio-based refineries do not strictly follow the economics of scale. This is because, with increasing scale, the transport cost of feedstock increases disproportionately as the required collecting area increases. This leads to a certain optimum size for such facilities. If the optimal capacity is exceeded, the production cost would rise due to the increasing transport cost. Optimally sized lingo-cellulosic biorefineries have been suggested to be in the capacity range of 1–2 GWth [46]. It has further been demonstrated that the biomass transport cost (and hence related CO₂ emissions) can be reduced by local or regional upgrading which leads to reduced feedstock cost in plants that depend on the improved properties of the upgraded biomass such as lower moisture content, better feeding and easier storage [8, 47–50]. The purpose of such pre-treatment efforts is to increase the energy density on a mass or volumetric basis by densification. A major part of the ballast is naturally the water content of the received biomass, which can be reduced by drying. However, further energetic densification can be achieved by pelletising, torrefaction and fast pyrolysis [8, 47–51].

2.4.1. Drying

Biomass drying is the removal of moisture by evaporation. The advantage is that for combustion processes, the level of efficiency can be increased by up to 15% due to the increased heating value, thus resulting in higher combustion temperatures. In addition, the flue gas volume is reduced, which makes it possible to use combustion equipment with smaller dimensions. Furthermore, for other upgrading processes, such as gasification or pyrolysis, drying reduces the energy consumption (water is not heated) and increases the product concentration. Targeted biomass moisture contents for gasification and torrefaction as well as for pelletisation and pyrolysis are 10–20% and 5–10%, respectively [52].

For biomass, three technologies can be distinguished based on the drying medium applied: air drying, flue gas drying and steam drying. In principal, all of the technologies can be applied either directly or indirectly. In practice, only steam drying is applied indirectly. Depending on the technology, the level of energy consumed, usually given in kJ/kg water evaporated (kJ/kg\text{H₂O}), varies widely from 500 kJ/kg\text{H₂O} for direct steam dryers with heat recovery to as much as 5000 kJ/kg\text{H₂O} for air and flue gas dryers [53].

However, the advantage of air dryers is that heat can be used at considerably lower temperatures [54]. On the other hand, steam dryers make it possible to recover the condensation heat, provided that a heat sink is available. With steam dryers, the higher temperature gradient results in higher drying rates and, consequently, drying occurs faster; thus, drying can be done using smaller equipment [52]. While all types of dryer emit volatile organic compounds and odours, they are worse at higher temperatures. Thus, appropriate abatement technologies must be considered before installation. For direct steam dryers, condensate impurities must also be considered, which may make wastewater
treatment necessary. Also, when considering considerable cost variations for different types of dryers, Pang et al. [52] concluded that, “there are advantages and disadvantages to each (drying) solution; therefore, comprehensive technology evaluation, cost analysis, and assessment of environmental impact are needed to select a feasible technology in a practical application.”

2.4.2. Pelletising

Pelletising is the mechanical densification of biomass to form cylindrical pellets with a diameter of approximately 6–10 mm and length below 50 mm. It allows for easier handling and transport and offers more uniform fuel properties. For instance, pelletising increases the bulk density of wet wood chips from roughly 300 kg/m³ [55] to approximately 650–700 kg/m³ [8].

Pelletising consists of milling, drying, densifying and fan cooling the product. The pelletising process requires an initial moisture content of approximately 10%, and thus wood must be dried prior to pelletising. The amount of energy consumed for pelletising is relatively low at roughly 2.5% of the biomass’s LHV, but it can increase to up to 20% if drying is included [25].

Pelletising is an established technology and has been applied industrially for decades.

2.4.3. Torrefaction

Torrefaction is a thermal treatment at temperatures in the range of 200–300°C at atmospheric pressure in an inert atmosphere for long residence times of 10–60 minutes. It is also referred to as mild pyrolysis or roasting. By means of near total water evaporation and the decomposition of hemicellulose, wet wood loses approximately 30% of its initial weight but only 10% of its energy [56]. The resulting moist gas consists mainly of carbon monoxide, carbon dioxide, water and some organic impurities (mainly acetic acid) [57]. The solid product is also referred to as torrefaction char. Relatively more oxygen and hydrogen are converted into a gaseous state, which results in a higher heating value for the product. Especially when pelletised, the heating value and the energy density of torrefied wood (18–23 MJ/kg and 15–18 GJ/m³, respectively) are higher than for virgin wood and considerably improve its transport properties [50]. Other advantages include the fact that torrefied wood is easier to store due to its hydrophobic nature and biological stability, and that it is more brittle so the grinding energy can be reduced by 80% [11]. It has been reported, that torrefaction increases the pelletising energy due to higher friction in the press channels and also forms less stable pellets [58, 59]. However, those issues can be overcome by adding some water and increasing the die temperature [60, 61]. When considered together with the fact that the thermal efficiency of subsequent gasification or combustion processes can be increased due to the higher heating value, torrefied wood becomes an interesting alternative for centralised CO₂-neutral power, heat and/or biofuel production [62–64]. In addition, using torrefied biomass as reducer in blast furnaces is the most beneficial way to reduce CO₂ emissions of steel production [65].
The torrefaction process consists of drying the wood to approximately 15 wt%, the actual torrefaction reactions and solid cooling [11]. As mentioned previously, torrefaction is also considered a high-temperature drying process, thus possible reactor technologies are usually adapted from drying technologies and reviews of the process are provided in Refs [66, 67]. There are several advantages and disadvantages for each type of reactor, and we can conclude that choosing optimal equipment will depend on many factors, as is the case for drying.

Torrefaction is an emerging technology and has not yet been proved on an industrial scale. However, existing plants in Sweden² and The Netherlands³, with annual capacities of 16 ktons and 60 ktons, respectively, suggest it will be in the near future. In addition, the Energy Centre of the Netherlands successfully carried out co-firing experiments at a co-firing rate of 25% without adverse effects on fuel handling, milling and combustion⁴.

Batidzirai et al. [67] and van der Stelt et al. [56] have discussed the torrefaction process and its economic feasibility. Both affirmed the improved fuel and transport properties as well as the strong economic potential of TWP. Adams et al. [50] compared WP with TWP production when produced in Norway and delivered to a power station in UK. They found that the system based on TWP has a lower environmental impact. Starfelt et al. [68] studied the possibilities of integrating TWP production with a CHP plant and found that it is economically feasible to replace coal demand of the considered DH system’s coal boilers with torrefaction char.

### 2.4.4. Fast Pyrolysis

Biomass fast pyrolysis (BFP) is the thermal decomposition of preferably dry biomass in the absence of oxygen at temperatures of approximately 500 °C and under atmospheric pressure. It yields a mixture of approximately 13 wt% for non-condensable gases (mainly carbon monoxide), 75 wt% for pyrolysis oil (incl. water) and 12 wt% for solid char. The fast pyrolysis process requires approximately 15% of the feedstock energy to provide the heat needed for the endothermic fast pyrolysis process (BFP requires heating up the biomass to the reaction temperature as well as heat for the actual decomposition reactions). Thus, it has a high level of energetic efficiency since all products can be energetically used during the process [51].

Researchers have recently assessed the available reactor technologies for BFP and found that research facilities are mainly of the fluidised bed type (bubbling and circulating). However, several scholars also applied screw feeder-driven reactors when studying their industrial applications [51, 69].

In 2014, Fortum Oy⁵ started operation of the world’s first industrial BFP facility in Joensuu, Finland. It produces 50 kt/year of pyrolysis oil. At this plant,

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the BFP process has been integrated with a municipal CHP plant. Fast pyrolysis occurs in a circulating fluidised bed reactor with the heat transferred by sand removed from an interconnected bubbling fluidised bed combustor. The hot pyrolysis gases are separated from sand and char in a hot cyclone, and char and sand are recycled to the combustion reactor. After quenching the gases with cooled pyrolysis oil, the non-condensable gases are combusted in the boiler. Some of the oil will be used in the company’s own heating plants and some of it will also be sold to another energy supplier to replace heavy and light fuel oil. A recent study found that it is possible to combust unblended, acid pyrolysis oil in industrial heating boilers equipped with specially designed burners as long as all parts of the burner in contact with the acid pyrolysis oil are made out of acid-resistant stainless steel [70].

Another successful application of the BFP process is the bioliq® process patented by the Karlsruhe Institute of Technology (KIT). BFP takes place in a twin-screw reactor where the biomass and hot sand are mixed. The hot sand provides the heat for the pyrolysis reaction. Thereby the sand cools down from approximately 550 °C to 450–500 °C depending on the sand-to-biomass ratio, which is typically in the range of 1:10–1:20 [personal communication R. Stahl, KIT, 2009]. The hot pyrolysis gases and the majority of the char leave the reactor and are separated in a hot cyclone, and the sand is recycled to a re-heater that could be fired with the non-condensable pyrolysis gases [47]. Pyrolysis of straw yielded 55 wt% of pyrolysis oil (incl. water) [47], but higher values of approximately 68 wt% for liquid have been reported earlier for wood feedstock [71]. Within the bioliq® approach, the pyrolysis oil is re-mixed with the char product (forming so-called pyrolysis slurry) and supplied to an entrained-flow gasifier to produce syngas and, subsequently, dimethylether. Dimethylether can be further synthesised to form methanol or gasoline [47]. According to the bioliq® website6, the entire process chain, which consists of BFP, gasification, gas cleaning, dimethylether synthesis and gasoline synthesis, is already in operation, but detailed process data have not been reported yet. In addition, Green Fuel Nordic Oy is starting to build a stand-alone BFP facility in Iisalmi, Finland, with an output of 90 kt/year of pyrolysis oil.7 Two additional plants are also being planned. In The Netherlands, EMPYRO has started to build a 25 kt/year BFP demonstration plant at Hengelo.8

Apart from using pyrolysis oil in fuel oil boilers, it could also be used in gas turbines and internal combustion engines for CHP production but so far, long-term operation has not yet been achieved [72]. As described above, gasoline has successfully been produced after the thermal gasification of pyrolysis slurry (PS) on a pilot scale by KIT. According to Bridgwater [51], no research activities have been reported since 2010 for other upgrading technologies aiming to produce hydrocarbons, chemicals and improved bio-oils. This is likely related to some of the disadvantageous properties of pyrolysis oil, such as mineral matter and solid-liquid separation. For this reason, Bridgwater concluded that much of the

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6 http://www.bioliq.de, August 2014
research is still at a fundamental scale and that the most interesting activities would require larger-scale applications in order to prove the feasibility and viability of this strategy. That said, Woolf et al. [73] have recently noted the potential for CO₂ mitigation when pyrolysis char is sequestered.

When stored, pyrolysis oil shows some ageing due to chemical and thermal instability, which in turn is due to its high content of reactive oxygenated hydrocarbons. Diebold has provided an overview of the different ageing mechanisms [74], concluding that ageing mainly leads to higher viscosity and lower LHV and that adding ethanol or methanol would stabilise pyrolysis oil under ambient conditions. Likewise, lowering the storage temperature could put a stop to the ageing problem [75]. KIT [47] states that after the separation of char and two-stage condensation, which separates the pyrolysis oil into a water-rich and tar-rich phase, ageing does not impede the quality of the produced bio-slurry.

### 2.5. Process Sustainability Assessment

With respect to the background to the EU's 20–20–20 targets (see Section 1.1), it is important to increase the use of biomass resources (20% more by 20202) when developing new pathways. The EU simultaneously provided targets to ensure that this is done in the most efficient way possible. Only then will it be possible to exploit the full CO₂ mitigation potential of the biomass resources (EU, 20% reduction) and achieve the goal of increased energy efficiency (20% target set by the EU). The expected lower energy conversion efficiencies for biomass-based processes pose a challenge in terms of meeting the EU's energy efficiency improvement target of 20%. This makes it crucial to analyse the energy consumption of the entire process chain, including plantation, harvesting, transport and processing. It also makes it crucial to analyse how the final product is used. However, no standard methods can be found in the open literature for evaluating the energy efficiency of the biomass conversion processes. In the following chapter, the different assessment methods are briefly explained because they are used to analyse and compare the integration concepts discussed in Section 4.

The most common method used to evaluate the efficiency of energy conversion processes is thermal efficiency (TE). TE refers to the ratio between the work and/or the useful heat output and the energy input needed for an energy conversion process. More generally, TE can be defined as the ratio between the useful energy output and energy input of the process. This approach is based on calculating the energy conversion efficiencies that can be applied to all energy systems. However, in biofuel-based energy systems it is important to broaden the system boundaries and, in particular, include more upstream processes in the analysis. This is routinely done when conducting, e.g., life-cycle assessments, since the processes that utilise biomaterials involve several transportation and pre-treatment steps.

PE analysis is another method used to evaluate the energy efficiency of energy conversion systems at broader system boundaries. It takes into account all of the PE inputs into a production system that are necessary to produce a certain
product at the system boundary. It is the sum of all the PE inputs into the system divided by the useful energy delivered at the system border, thus yielding a primary energy factor (PEF), with the reciprocal being called the PE efficiency). While PE is a general concept, it can be analysed based on, e.g. EN 15603 [76]. The PE analysis done according to this standard is an integral part of the Energy Performance of Buildings Directive (European Council, 2010). In Publication I–III, the PE analysis based on EN 15603 was used to evaluate biomass pretreatment systems. More generally, it has been applied in different fields of process engineering ranging from the evaluation of carbon capture and storage [77] to power, heat and cooling generation [78—81], vehicle powertrains [82] and cement plants [83].

Exergy analysis is used to compare, improve and optimise processes. It provides efficiencies that measure how far the process being studied is from the ideal case and indicates the parts of the process in which exergy losses occur. Exergy by definition is the maximum useful work that can be obtained from a system at a given state in a given environment. It is based on the second law of thermodynamics and thus takes into account the irreversibility that occurs in all real processes. Exergy is a combination property of a system and its environment because, unlike energy, it depends on the state of both the system and the environment. Exergy is a state property for a fixed environment and the exergy of a system in equilibrium with the environment is zero. Good introductions to the concept of exergy analysis have been provided, for instance, by Dincer and Rosen [84] and Bejan, Tsatsaronis and Moran [85]. Exergy analysis has been applied in many fields, but with respect to the focus of this thesis, it was previously applied to the Bio-SNG process [21, 86] and fast pyrolysis [87, 88].

The concept of sustainability also has to do with the process being economically feasible, i.e. profitable. If thermodynamic data on a particular process are available, it can be combined with the cost data for fuel, commodities, products, investment and operations. Based on such a combination, a cost balance can be obtained that reveals the profitability of the process. A major challenge is posed by allocating the cost to the products in order to calculate specific product costs. Exergoeconomic analysis is one way to allocate process costs to multiple products. This combined approach is able to reveal the major cost and thermodynamic inefficiencies of the process under consideration. When combined with information on the fuel and product market prices the overall economic feasibility of the process can easily be obtained. Having its origin in the evaluation of combined gas turbine cycles [85], the method has been applied widely to energy systems and the application is well defined when applying, e.g. the specific exergy costing (SPECO) method [89].
3. Objectives and Scope

3.1. Objectives
The objective of the thesis is to investigate the effect of integrating different biomass upgrading processes with a municipal CHP plant. In particular, their influence on important operational parameters and energetic and environmental performance as well as the plant’s economics will be highlighted. With respect to the latter aspects of system performance, a concise and significant assessment methodology will be developed that makes it easy to compare processes yielding multiple products. Furthermore, the following aspects will be considered as well:

- the chosen processes must be able to convert various lignocellulosic biomass resources;
- the processes must be suitable for application in small- to medium-scale plants;
- a mostly homogenous product with a high energy density must result in order to assure suitability for better handling and more viable long-distance transport;
- the product must be a valuable and versatile inter-stage or final product.

An integration approach was sought for the highest possible product yield within the CHP plant’s operational limits while simultaneously retaining the demands of the adjunct DHN.

Adding new products to a municipal CHP plant’s portfolio is expected to increase its profitability and to make it less sensitive to fuel and product price fluctuations. This would be the case because the plant would benefit from new markets currently opening up within the emerging bioeconomy. By using unexploited boiler capacity during times of low DH heat demand, the degree of capacity utilisation would improve. Likewise, the CHP plant would financially benefit by yielding more products. In addition, CHP-integrated upgrading would also increase the energy conversion efficiency of biomass compared to conversion at stand-alone plants, which potentially would allow them to benefit from CO₂ trading schemes.

Biomass resources are versatile but restricted, and it is difficult to predict how they will be used in the future. This means that the applied processes must be as flexible as possible with respect to fuel so that they can respond to changes on the feedstock supply side. Thus, they should be able to handle various feedstocks.

The retrofitting of heat-only appliances into new small- to medium-scale CHP plants and the construction of such plants has the potential to significantly mitigate CO₂ levels, but the economics of such plants are still considered unstable. By using the existing biomass infrastructure of local plants for increased biomass utilisation, local biomass resources could be sustainably exploited and the plants could contribute to CO₂ mitigation while simultaneously improving the economics of the plant.
The newly generated products must be valuable and versatile with respect to their future use in large-scale applications. Thus, a high level of energy density for viable transport as well as homogenous fuel properties are desired.

To summarise, an evaluative methodology needs to be developed that makes it possible to objectively assess new multi-product processes and the upgrading processes must prove their suitability for integration from a technical standpoint. Furthermore, the new process concepts must prove to be beneficial with respect to the system’s energetic, environmental and economic performance. Within this thesis, five original papers describe the investigative process and prove the correctness of the hypotheses. The contribution of each paper is listed in Table 2.

<table>
<thead>
<tr>
<th>Original Paper</th>
<th>Contribution to the research goal</th>
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<tr>
<td>Publication I:</td>
<td>The study demonstrates the general importance of biomass densification prior to large-scale biofuel production with respect to the use of fossil resources. It shows that the primary energy concept can be easily applied to processes other than power and heat production and that it gives correct results.</td>
</tr>
<tr>
<td>Publication II:</td>
<td>A suitable model for BFP yield estimation is presented and successfully applied in order to show the influence of integrating BFP with a municipal CHP plant based on important operation parameters. Important modelling assumptions have been analysed. The study shows that the DHN’s PE efficiency and CO₂ mitigation potential can clearly be improved.</td>
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<tr>
<td>Integration of biomass fast pyrolysis and precedent feedstock steam drying with a municipal combined heat and power plant.</td>
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<tr>
<td>Publication III</td>
<td>The process concept of the CHP plant integrated with BFP that is analysed in Publication II is challenged with the concepts WP and TWP production integrated with the same CHP plant. The study demonstrates that all integration options are possible within the CHP plant’s operational limits and that they increase DH generation, PE efficiency and CO₂ mitigation potential. The results indicate that WP is the most beneficial integration option.</td>
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<tr>
<td>Energetic and environmental performance of three biomass upgrading processes integrated with a CHP plant</td>
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<tr>
<td>Publication IV:</td>
<td>Based on the simulation results presented in Publication III, an exergoeconomic analysis was conducted. Combined with projected future product market prices, the study demonstrates that the least efficient integration option, fast pyrolysis, has the greatest economic potential. This could be attributed to reasonable investment costs along with high expected prices for the product on the market.</td>
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<tr>
<td>Exergoeconomic assessment of CHP-integrated biomass upgrading</td>
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<tr>
<td>Publication V:</td>
<td>Based on a case study of Bio-SNG production, the informative value of thermal efficiency calculations were compared to exergy analysis and PE efficiency calculations. The paper confirms that exergy analysis has advantages when it comes to detailed process analysis due to a reduced need for assumptions, but that on a system level, PE efficiency yields objective results, thereby demonstrating the influence of the process on the global PE resources.</td>
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<tr>
<td>Comparison of Energy Efficiency Assessment Methods: Case Bio-SNG process</td>
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</tbody>
</table>
3.2. Scope

The study uses thermodynamic CHP simulation models that are based on the composition of virgin wood, as shown in Table 3. With respect to the simulation models used to obtain the mass and energy balance, the influence of the possible variations in the composition of other possible feedstock on the results can be considered small and is of even less importance when comparing processes. However, when really building the suggested plants based on other feedstock, the ash content and properties in particular need to be considered in more detail.

The thesis is based on a case study of one specific CHP plant, with the design data obtained from the literature. The plant was integrated with a DHN; detailed demand data of the DHN were available, as explained in Publication II. In order to transfer the data to other DH systems and make generalisations about the results, assumptions were challenged by means of a sensitivity analysis. In particular, the DH demand was varied in order to represent substantially colder and warmer years. In addition, structural changes were made to the plant’s layout and the influence of the applied boiler technology (bubbling versus circulating fluidised bed versus grate furnace) was investigated. Based on the sensitivity studies documented in Publication II, the results can be transferred to smaller plants based and/or plants based on circulating fluidised bed and grate technology. Based on the results presented in Publication II, the scope was narrowed to bubbling fluidised bed technology and a comparison was carried out for one typical operation year.

The part-load operating characteristics of the upgrading equipment were neglected, i.e. they were simulated using a constant specific energy demand. No detailed data on the part-load behaviour of the equipment described in this study could be found in the existing literature. Even though it was assumed that the required equipment could operate at somewhat lower loads, the author is aware of the fact that for very low loads (less than 50%), the exact physical size of the various pieces of equipment would need to be determined and that a two-line approach in which two units would work in parallel to best represent the required load might be necessary. However, as the focus of the work is on comparing processes, no detailed component sizing was carried out.

Based on the sensitivity analyses, the author is confident that the results obtained are of a general nature. However, a more detailed technology review would be necessary before real-life implementation, especially a review of fuel properties and the upgrading equipment.
4. Process Simulation and Assessment

4.1. Process Simulation

Computer-aided design tools, such as simulation software, play a crucial role in the development of new processes, both in industry and academia. After the preliminary process layout has been synthesised (e.g. by means of a block diagram), simulation tools are used to design a flowsheet of the process, including all equipment and its interconnections. The model is used to compute the mass and energy balances of the process. This makes it possible to determine energy flows and utility requirements as well as to estimate the size of the physical equipment. Once the basic process layout has been established and evaluated, the input parameters can easily be changed and structural optimisation (design improvement) possibilities can be implemented. Heat integration in particular is facilitated which can considerably increase the energetic and economic efficiency of the processes. [90, 91]

There are two major subgroups of simulation software, which use either a sequential-modular approach or a simultaneous or equation-oriented approach. With the latter approach, the entire process is described by a set of equations, which are solved simultaneously. This usually requires more computing power and the model is more dependent on good starting values in order to achieve convergence. The set-up for sequential-modular simulators is easier and requires less computing power. This is because the process is split into different units, which are solved one by one in a sequential manner by applying iterative solving strategies. Ordering routines for the unit calculations are often implemented in the software in order to achieve rapid convergence. Sequential-modular approaches do not allow for dynamic modelling (which is important in estimating a plant’s start-up and shutdown behaviour); however, usually steady-state simulation is sufficient for process design. [90]

Simulations can either be conducted in design mode or in analysis/off-design mode. When using the simulation mode, the output is fixed and the software calculates the required inputs. [91]

For instance, when designing a CHP plant, the DH load is fixed. By leaving the power output as a free variable (only the performance curve of the turbine is given), the required fuel input can be determined. In contrast, the input is specified and the output is calculated when using off-design mode. In this study, for instance, the CHP plant is defined by its design operation point, but part-loads and the integration of upgrading equipment have been simulated using the off-design mode, i.e. with fixed physical dimensions for the turbine and heat-exchange equipment. Information on the most important process simulation software is provided in Ref [90], along with recommendations for further reading.

In this study, the thermal power plant simulator Prosim® was used. It is commercially available software with demonstrated reliability for the preliminary design of natural gas and solid-fuel-fired power plants [92]. Prosim® is a sequential-modular process simulator and its simulation models
are built from process modules and streams. Once the process design is complete and all user data have been inputted, the software iterates mass and energy balances by applying the Newton-Raphson method. In design case calculations, it also calculates the physical sizes of the selected process modules. The sizes can be fixed for off-design calculations, i.e. when simulating part-load operation [93].

4.1.1. Multi-period Model of the District Heating Network

Real yearly DHN data for a small town in southern Finland has been used on a daily average basis to represent the load of the adjunct DHN. In order to match the data with the chosen CHP plant capacity, the data were scaled using a constant factor so that the pre-defined CHP plant provides 60% of the hourly peak demand of the DHN when at full load. It was assumed that the CHP plant would be shut off at 50% of its designed DH load, which corresponds to a 30% demand in the DHN. As stated in Ref [28], those are common operating parameters for municipal solid-fuel-fired CHP plants. Under those conditions, the CHP plants provide 75% of the total yearly DH load, which is a typical value [19].

In order to represent the yearly generation of the base-case, stand-alone CHP plant, a multi-period DH load model was developed following a method used by Savola [28]. Applying a linear correlation, the DH supply and return temperatures varied between 110 °C and 80 °C and 83 °C and 36 °C, respectively, depending on the load. Here, the highest and lowest monthly average loads are assumed to represent the highest and lowest supply- and return temperature profiles of 110/83 °C and 80/36 °C, respectively. As can be seen in Fig. 3, two full-load and five part-load levels (90%, 80%, 70%, 60% and 50%) represent the heat duration curve. The operating hours of the part-load points have been set at equal time and, together with the two full-load periods, they provide the CHP plant’s annual DH load. The full-load periods are also of the same duration, but they differ in their supply and return temperatures in order to represent different outside temperatures during full-load operation. Therefore, the full-load periods have been split into two loads of equal duration, with one having a supply and return temperature profile of 110/83 °C (later referred to as 100%) and the other a reduced temperature profile of 107/79 °C (later referred to as 100% r.t.). Simulation models for each load point were run and the fuel input was adjusted in order to match the required DH output, while leaving electricity generation as a free variable.

All integration options allow for longer operation hours. This fact was taken into account by first determining the lowest possible heat load (later referred to as min) and the related operation hour extension for the integrated CHP plant. Second, the average DH load representing the respective operation time and heat production was computed and added as an additional load level to the base-case scenario (see Fig. 4), later referred to as the subload.

For calculating the annual production, it was assumed that the DH demand not provided by the CHP plant was generated in oil-fired, heat-only boilers with a TE of 0.93 [94]. Allowing time for yearly maintenance, the annual maximum operation hours of the CHP plant have been restricted to 8000 hours.
Figure 4. Representation of the extended operation hours in the multi-period load model.

4.1.2. Simulation Model of the Combined Heat and Power Plant

Process parameters and performance data for the medium-sized CHP plant based on bubbling fluidised bed technology have been adapted from Savola [28] and are presented in Table 3. An illustrative example of the process is given in Fig. 2. In order to adjust the simulation model to match the demand of the DHN, the fuel input was determined in order to achieve a DH output of 16.5 MW. The temperature level of the DH supply and return streams was set as 83/110 °C and matched the full load temperature values observed for the DHN, as explained in Section 4.1.1. Electricity generation was left as a free variable. For sizing the turbine and calculating the power output, the standard turbine curves implemented in the software’s database were used. They represent typical turbine performance curves for different turbine capacities, including part-load behaviour. In order to assess steam extraction at reduced pressure for drying, as explained in detail in Publication II, the turbine was divided into four stages.

Ultimate analysis of the wood fuel assumed in this study was derived from the Phyllis database [95], with the fuel moisture set at 50 wt%, a typical value for woody biomass [96] (see Table 3). The LHV was calculated using the Boie correlation [97] with the heat of evaporation of water being 2253 kJ/kg.

It was assumed that the operation of the bubbling fluidised boiler is restricted to a DH load of 50% in order to assure stable combustion at a reduced fuel flow. The part-load operation of the modelled CHP plant was simulated in the off-design mode by adjusting the fuel mass flow in order to reach the user-defined DH load points of 90%, 80%, 70%, 60% and 50% capacity. No further model validation or optimisation was carried out for the stand-alone CHP plant since detailed operation data of the CHP plant could not be obtained due to confidentiality reason. However, based on the suitability of the simulation software, the model forms a solid basis for comparing processes under part-load conditions since it responds realistically to load variations.

The CHP plant simulation model described above forms the base case for comparing all subsequent process variations discussed in the following sections.
Table 3. Wood fuel properties and CHP plant design data (M: moisture)

<table>
<thead>
<tr>
<th>Wood fuel properties</th>
<th>CHP plant – design data</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ultimate analysis: carbon [wt%]</td>
<td>50.64</td>
</tr>
<tr>
<td>sulphur [wt%]</td>
<td>0.08</td>
</tr>
<tr>
<td>hydrogen [wt%]</td>
<td>6.10</td>
</tr>
<tr>
<td>ash [wt%]</td>
<td>0.80</td>
</tr>
<tr>
<td>Oxygen [wt%]</td>
<td>42.22</td>
</tr>
<tr>
<td>nitrogen [MJ/kg]</td>
<td>0.16</td>
</tr>
<tr>
<td>LHV [MJ/kg]</td>
<td>7.9</td>
</tr>
</tbody>
</table>

4.1.3. Simulation of Biomass Drying

All of the upgrading processes considered here require low biomass moisture contents of 10–15 wt%. Thus, precedent biomass drying must be integrated as well. The drying technology involved indirect steam drying by means of a rotary steam-tube dryer. The reason for the choice was the high level of efficiency of the technology and the availability of both hot steam as a heat source and the boiler feedwater as a heat sink for the dryer condensate. An advantage of indirect steam drying is that the steam is not contaminated and thus can be recycled to the steam cycle with no wastewater treatment required. Furthermore, since live steam is used for drying, no changes to the turbine are necessary, which will reduce costs in the case of retrofit installation. Steam extraction also provides an easy means of providing load control for the DH exchanger. When more steam is used for drying, less DH is generated.

The process layout is illustrated in Fig. 5. The amount of heat consumed by the steam-tube dryer was estimated at 3000 kJ/kgH₂O [96]. Wet biomass (wood chips) and hot flue gases are fed to the dryer (9). If heat is available from the flue gases, they are cooled down to 143°C, which is the minimal flue gas temperature at the base-case CHP plant. Lower flue gas temperatures (approximately 120 °C) might be possible, but 143 °C was chosen in order to allow for fair comparison with the CHP stand-alone process. The remaining heat demand was provided by steam extracted before entering the turbine (5). While paying attention to the retrofit situation, live steam was extracted (13), throttled to 10 bars (14), and further cooled to 190°C by spraying in the saturated water leaving the dryer (15). The dryer condensate was throttled to 2 bars (16), then sent to the feedwater tank (7).

The dryer load for each load point is given by the steam extraction rate required to match the DHN’s heat demand. Thus, the maximum possible product yield requires the highest possible boiler fuel input since heat must also be provided for the drying process. Conversely, this means that the steam enthalpy exceeds the demand of the DHN because the boiler temperature is controlled by means of the evaporator and superheater tubes in the boiler walls and in the freeboard, respectively. If the heat input in the boiler is kept at a higher level than usual, the water amount needed to dissipate the heat from the boiler walls increases, and consequently, this heat must now be “dissipated” in the dryer so that it matches the DH load.

The DH load was matched by adjusting simultaneously the dryer load and fuel input. With this setup, the dryer load constantly increases at the same time as the DH load levels decrease to 60–70% at full boiler load. The maximum
condensate flow off the dryer (and thus its capacity) is restricted by the pressure of the steam extracted from the turbine (5) for feedwater pre-heating as this pressure defines the pressure prevailing in the feedwater tank (9). As the extraction pressure, the pressure in the feedwater tank decreases along with the declining steam mass flow and the thus related steam parameters. Hence, there is a pressure-dependent maximum heat flow that can be fed into the feedwater tank until a saturation state is reached for the condensate mixture from the DH exchanger (6) and dryer (9). In order to overcome this restriction, the pressure in the feedwater tank has been load-dependently increased to a maximum of 2 bars, thereby matching its design pressure (it should be noted that for increased pressures, steam is no longer extracted from the turbine). However, for the reason mentioned above, the heat that would need to be “dissipated” in the dryer for realising DH loads below 60–70% would result in such a high heat flow for the dryer condensate that the feedwater would exceed the saturation state. Hence, the boiler load was decreased gradually for those cases. With this control strategy, the lowest possible DH load decreases considerably, which explains the increase in operation hours compared to the base case.

**Figure 5.** CHP plant integrated with WP production. The highlighted areas describe the functional units as defined for exergy analysis.

### 4.1.4. Simulation of Wood Pellets Production

Pelletising consists of drying the biomass to a moisture content of 10 wt% for the milling, actual pelletising and fan cooling of the pellets. The simulation of the drying step is explained in Section 4.1.3, while the additionally required process steps have been only considered with respect to the electricity consumed: estimated at 260 kJ/kg of wet wood based on data provided in Ref [98]. This was taken into account when calculating the CHP plant’s net power output. The process layout is shown in **Fig. 5.** Biomass and hot flue gas enter the dryer (9), and subsequently, dried biomass is crushed (10), pelletised (11) and cooled (12).
4.1.5. Simulation of Torrefied Wood Pellets Production

Based on a prior study [11], the enthalpy for the torrefaction of wood (20°C, moisture 15%) at 280°C has been estimated as 714 kJ/kg. Based on the LHV of the input wood, energy yields were set at 98.2% for TWP and 5.1% for gases, respectively. The gas is thought to be co-fired in the boiler and its energy is subtracted from the boiler fuel input. The power consumed by the dryer, torrefaction reactor and subsequent milling and pellet production was estimated to be 340 kJ/kg of wet wood [11, 98].

The process was integrated as follows (Fig. 6): prior to torrefaction, the biomass was dried (14), reducing the moisture content from 50% to 15% when using the same dryer layout described in Section 4.1.3. Heat for the actual torrefaction process was provided by a flue gas split-off (9) after the economiser (8).

Flue gases have a load-dependent temperature of 400–460°C and are cooled to 265°C, thus providing heat indirectly to the torrefaction reactor (10), and they are mixed back into the main flow of the flue gas before reaching the air-preheater (2). This leads to a reduced flue gas temperature after the air-preheater (minimum 106.5°C), which calls for a critical evaluation of potential corrosion. In this case, flue gases are reheated in the dryer to the minimal temperature of the base case in order to obtain comparable results and account for potential corrosion risks.

As the process simulator does not make it possible to simulate complex chemical reactions, the heat extraction for the torrefaction reaction is modelled with the help of an additional evaporator representing the heat required for torrefaction. The DH load is matched by simultaneously adjusting the dryer input (as explained in Section 4.1.3.), the split-off required for torrefaction of the dried biomass and the reduction in the boiler load. In continuation the torrefied biomass is crushed (11), pelletised (12) and cooled (13).
4.1.6. Simulation of Pyrolysis Slurry Production

A crucial parameter for modelling BFP is the heat for pyrolysis, which consists of the heat required to increase the biomass’s temperature from the ambient temperature to the reaction temperature, and is complemented by the heat of pyrolysis, which is required for the actual (usually) endothermic pyrolysis reaction. In order to estimate the heat for pyrolysis of the woody biomass (the specific composition is given in Table 3), a mathematical model was recently developed primarily by the author of this thesis [99]. The main assumption of the model is that BFP forms certain chemical groups consisting of carboxylic acid, aldehyde, phenol and sugar species along with char, pyrolytic lignin and water. With the model, the heat for pyrolysis was calculated at 1829 kJ/kg and the water content of the pyrolysis oil was calculated at 23.5 wt%, which is well in range of the values given in Ref. [100] and Ref. [101], respectively. The mass and energy balance of the BFP process (including drying) are illustrated in Fig. 7.

Based on the LHV of the wet biomass, the energy yields for the BFP process are 78% for BFP oil, 25% for char and 12% for gases, respectively. Publication II provides a detailed discussion of the technical feasibility of the integration concept: hot flue gases are split off (9) after the actual boiler section (3) and used as a heat source for the BFP reactor (10), as depicted in Fig. 8.

The cooled flue gas is mixed at 480 °C back into the main flue gas stream before reaching the economiser (8). The BFP reactor (10) could either be of the circulating fluidised bed type or the screw type. Both reactors offer a wide capacity range [69].

After the BFP reactor (10), char needs to be separated from the pyrolysis gas by means of a hot cyclone (11). After the separation process, the gases are usually quenched in a spray cooler (12) with cool pyrolysis oil in order to achieve rapid cooling for high liquid yields. The permanent gases are then co-fired in the boiler (3). Char and pyrolysis oil can be re-mixed with a colloid mixer (13) to form PS before transport or further upgrading. The benefit of such an integration concept is that it can be applied to all types of common boilers (bubbling fluidised bed, circulating fluidised bed and grate–fired boilers).
The influence of the different types of boilers is described and discussed in Publication II.

Since the major interaction between BFP and the CHP plant takes place when heat is extracted from the flue gas, the heat extraction process itself is modelled with the help of an additional evaporator.

All other equipment required for pyrolysis has been considered based solely on power consumption. BFP requires a dryer, a BFP reactor and a cyclone to separate solids. Based on the bioliq® concept developed by the KIT [102], a colloid mixer is needed to prepare the PS. The power consumed by the upgrading equipment has been estimated to be 280 kJ/kg of wet biomass based on values found in Refs [11, 98, 103, 104], with more details provided in Publication II. The power consumption of the upgrading equipment is subtracted from the net power generation that is calculated by the simulation software.

After integrating BFP and steam drying (14) until a moisture content of 10 wt% is achieved, the DH load can be controlled by simultaneously adjusting the flow of the flue gas to the BFP reactor, the steam flow to the dryer and, if required, by reducing the fuel input to the boiler. The restriction for the dryer load, as set by the feedwater tank, has been dealt with in Section 4.1.3. The heat demands of the steam dryer and BFP reactor are linked such that all dried biomass can be pyrolysed. The combustion of the pyrolysis gas has not been simulated here. Though, the energy content of the pyrolysis gas was subtracted from the boiler’s fuel input. The influence of the combustion of the pyrolysis gas is described and discussed in Publication II.
4.2. Process Assessment

Basically three different methods were used to assess the process efficiency in this study, with the method for assessing PE efficiency also being applicable for evaluating CO₂ emissions. Exergy analysis was complemented with economic data in order to carry out an exergoeconomic analysis based on the SPECO approach. **Publications I–III** apply TE and PE efficiency, whereas **Publication IV** provides an exergy and exergoeconomic analysis. **Publication V** compares the efficiency assessment methods based on a case study on Bio-SNG production in order to reveal the advantages and disadvantages of the various methods in a comparative manner. The methods and the necessary input data are explained in the following section.

4.2.1. Thermal Efficiency

The overall TE of the integration concepts, \( \eta_{\text{tot}} \), was calculated using the following equation:

\[
\eta_{\text{tot}} = \frac{Q_{\text{DH}} + P + Q_{\text{BP}}}{Q_{\text{BM}} + Q_{\text{FO}}} \quad \text{(Eq. 1)},
\]

where \( Q_{\text{DH}} \), \( P \) and \( Q_{\text{BP}} \) are the integrated plant’s output in terms of DH, net power and bioproduct. \( Q_{\text{BM}} \) and \( Q_{\text{FO}} \) account for the chemical energy contained in biomass and fuel oil, which are fed to the integrated CHP plant and the DHN’s heat-only boiler, respectively. In addition, specific efficiencies, \( \eta_{\text{DH}} \), \( \eta_{\text{P}} \) and \( \eta_{\text{BP}} \), for DH, net power and bioproduct were calculated by relating each of them separately to the total fuel input. The power-to-heat ratio, \( \alpha \), was calculated using the following equation:

\[
\alpha = \frac{P}{Q_{\text{DH}}} \quad \text{(Eq. 2)}.
\]

4.2.2. Primary Energy Efficiency and CO₂ Emissions

PEFs and CO₂ emission coefficients, the definition for which was derived from EU standards EN15603 [76] and EN 15316-4-5 [105], were used to compare the different process layouts. European standard EN15603 [76] defines the calculations for the energy performance indicators for buildings in general, whereas standard EN 15316-4-5 [105] provides definitions for DH systems in particular. Based on the annual generation data for the DHN, the PEFs and CO₂ emission coefficients are calculated by applying the power-bonus method [105]. In brief, the PEF represents the sum of all PE inputs into a DHN divided by the amount of DH delivered. It can, hence, be understood as a reciprocal efficiency of the DHN. Similarly, all CO₂ emitted in order to deliver one unit of DH is expressed by the CO₂ emission coefficient. By applying the power bonus method, as defined by the European Committee of Standardization [105], the PEF of the DHN (\( \text{PEF}_{\text{DHN}} \)) can be calculated as follows:

\[
\text{PEF}_{\text{DHN}} = \frac{Q_{\text{BM}} \times \text{PEF}_{\text{BM}} + Q_{\text{FO}} \times \text{PEF}_{\text{FO}} + P \times \text{PEF}_{\text{P}}}{Q_{\text{DH}}} \quad \text{(Eq. 3)},
\]

where \( Q_{\text{BM}} \), and \( Q_{\text{FO}} \) are the LHV-based annual fuel input of wood to the CHP plant and fuel oil to the heat-only-boiler and \( P \) and \( Q_{\text{DH}} \) are generated net power and DH, respectively. \( \text{PEF}_{\text{DHN}}, \text{PEF}_{\text{BM}}, \text{PEF}_{\text{FO}} \) and \( \text{PEF}_{\text{P}} \) are the corresponding PEFs with their values given in Table 4. The electricity generated in the CHP plant, by definition, replaces electricity generated elsewhere and \( \text{PEF}_{\text{P}} \) is defined
as the PEF of the replaced electricity. Scholars have not yet come to a general agreement on which value to choose at this point, but values for the power generated in a CHP plant may vary from those that marginally replace coal-condensing power to national and regional efficiencies for power production. The respective standard does not provide a guideline, and thus, in this study the average efficiency value for electricity generation in Finland was used. The considerable influence of different PEF\(_P\) values has been discussed in Publication III.

Standard EN 15316-4-5 [105] only pertains to the generation of heat and electricity; thus, the power bonus method was extended in this thesis by regarding the bioproducts as a “bonus” too. The PEF of the integrated systems can then be calculated as follows:

\[
P_{EF_{DIN}} = \frac{(Q_{BM} \cdot PEF_{BM} + Q_{FO} \cdot PEF_{FO} - P \cdot PEF_{P} - Q_{BP} \cdot PEF_{BP})}{Q_{DH}} \tag{Eq. 4}
\]

where \(Q_{BP}\) is the chemical energy contained in the bioproducts (WP, TWP, PS) and \(PEF_{BP}\) is the PEF of the replacement. In this study, it was assumed that PS and BFP oil products would replace fuel oil (Publication II) or coal (Publication III) and that pyrolysis char, TWP and WP would replace coal. Values and references for the PEFs are provided in Table 4. They describe the “energy history” of the fuels (extraction, transport, pre-treatment, etc.). The CO\(_2\) emission coefficients were calculated accordingly by replacing \(PEF_i\) in Eq. 4 with the corresponding CO\(_2\) emission coefficients, which are given in Table 4.

Table 4. The PEFs and CO\(_2\) emission coefficients used in the study

<table>
<thead>
<tr>
<th>PEF [-]</th>
<th>CO(_2) emission coefficient [kg/MWh]</th>
</tr>
</thead>
<tbody>
<tr>
<td>Biomass</td>
<td>1.09 [76]</td>
</tr>
<tr>
<td>Fuel oil</td>
<td>1.35 [76]</td>
</tr>
<tr>
<td>Coal</td>
<td>1.19 [76]</td>
</tr>
<tr>
<td>Power (coal condensing)</td>
<td>4.05 [76]</td>
</tr>
<tr>
<td>Power (Finnish average)</td>
<td>3.11 [106]</td>
</tr>
<tr>
<td>Power (Norwegian average)</td>
<td>1.33 [106]</td>
</tr>
</tbody>
</table>

### 4.2.3. Exergy Analysis

Although energy is always conserved in any process, its useful part — exergy — is not conserved because of irreversible phenomena, such as heat transfer with a finite temperature gradient, mixing, chemical reactions and expansion. Exergy analysis identifies and quantifies these irreversibilities by taking into account the entropy that is generated. Exergy analysis can reveal the margin available to design more efficient energy conversion systems. Although the maximum efficiency in energy conversion systems corresponds to an ideal process (no entropy generation and reversibility), all real processes are to some extent entropy generators. Exergy analysis can indicate how far from ideality a process is, and thus, indicate the potential for improvements.

As for energy, different forms of exergy exist. For thermochemical processes nuclear, magnetic, electrical and interfacial effects can be excluded. For material streams in steady-state systems also the kinetic and potential exergy usually can
be neglected, and the exergy of a system can be described as the sum of its physical and chemical exergy according to Eq. 5:

\[ e_{\text{tot}} = e_{\text{ph}} + e_{\text{ch}} \]  
(Eq. 5).

Physical exergy takes into account the departure of the system from environmental temperature and pressure \((T_0, p_0)\), while chemical exergy takes into account the differences between the system’s chemical composition and the standard chemical composition of the environment. The specific physical exergy is calculated based on the following equation:

\[ e_{\text{ph}} = h_1 - h_0 - T_0 \cdot (s_1 - s_0) \]  
(Eq. 6),

where \( h_1 \) and \( s_1 \) are the specific enthalpy and entropy, respectively, which are evaluated at the present \((p, T)\) and at the reference \((p_0, T_0)\) state.

Chemical exergy, which is based on the standard chemical exergy approach described by Szargut et al. [107], can be derived as follows:

\[ e_{\text{ch}} = \sum_i x_i e_{\text{ch},i} + R T_0 \sum_i x_i \ln x_i \]  
(Eq. 7),

with \( x_i \) representing the molar fraction of the component \( i \).

The overall exergy balance for any control volume can then be expressed as follows:

\[ e_{\text{tot,in}} = e_{\text{tot,destroyed}} + e_{\text{tot,losses}} \]  
(Eq. 8).

The main characteristic of Eq. 8 is its non-conservative nature. In other words, even though mass and energy are conserved in a real process, their usefulness (exergy) is not.

The exergetic efficiency (EXE) of each process (combustion, drying, turbine stages, etc.) and the overall EXE of the thermal plant at each load point were calculated based on the definition for products and fuels provided by Lazzaretto and Tsatsaronis [89].

4.2.4. Exergoeconomic Analysis

The exergoeconomic analysis carried out in Publication IV requires economic data input. The means for calculating such input are explained below.

**Investment cost estimation for the CHP plant:**

The total capital investment (TCI) costs of the power plant were estimated based on data from a power plant recently built in the city of Kerava, Finland. The plant in question has a thermal capacity of 81 MW and it generates 21 MW, 10 MW and 48 MW of power, process steam and DH, respectively. The TCI of the Kerava plant was approximately 65 M€\(^9\).

By applying the so-called six-tenth rule, the cost of the smaller unit describing the base case can be calculated as follows:

\[ C_{\text{new}} = C_{\text{old}} \cdot \left( \frac{S_{\text{new}}}{S_{\text{old}}} \right)^f \]  
(Eq. 9),

where \( C_{\text{new}} \) is the cost of the scaled equipment, \( C_{\text{old}} \) is the known cost, \( f \) is the scaling factor and \( S_{\text{new}}/S_{\text{old}} \) is the ratio of two characteristic dimensions for equipment. In the case of power plants, either the power output or fuel input is used as a characteristic dimension and the scaling factor was set at 0.6. By

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applying Eq. 9 to both characteristic dimensions, the cost can be estimated at between 25.6 M€ and 28.9 M€ at 2008 prices. Using the Chemical Engineering Plant Cost Index (CEPCI), the known prices for reference year 1 can be adjusted to match the prices for reference year 2 according to Eq. 10:

\[
C_{\text{year}2} = C_{\text{year}1} \cdot \frac{CEPCI_{\text{year}2}}{CEPCI_{\text{year}1}} \quad \text{(Eq. 10)}
\]

When considering the CEPCI, the TCI was set at 26 M€ in 2013 prices. Therefore, the calculated equipment cost considers the influence of inflation and price escalation. In order to determine the detailed equipment cost, the cost contributions of a single equipment unit were adapted based on a prior study \[108\]. The boiler cost has been broken down into costs for the evaporator, the superheaters, the economiser and the air pre-heater by rating them against the transferred heat in the specific unit. The cost for effluent treatment has been distributed to the other equipment based on the weighted average of their cost contribution. The purchased equipment cost (PEC) was then calculated to represent 39% of the TCI. A detailed cost breakdown for the CHP plant is given in Publication IV (Table 2) and a summary is provided in Table 5.

**Investment cost estimation for WP, TWP and PS:**

Two steps were used to estimate the TCI costs for the production of WP, TWP and PS. First, the various sizes of the equipment were estimated. The equipment was designed for the maximum flow rate, which is reached at 60–70% of the DH-load as explained within Section 4.1.3. No overdesign was included. Second, the PEC for a specific unit was extracted from the references. If needed, an exchange rate of 1.35 US$/€ was used. All costing details, including the references used for the estimations, can be found in Publication IV (Tables 3a–c). A summary is provided in Table 5. If the PEC for the required capacity was not directly available from the literature, the PEC was scaled by applying Eq. 9. The scaling factors are given in Publication IV (Tables 3a–c). In order to determine the total onsite costs (TOCs), it was estimated that the installation cost factors range from 10% to 25% based on the PEC. All prices and costs have been adjusted to match 2013 values by applying the CEPCI (see Eq. 10). The CEPCI for the reference year 2013 is 567.6.

As no detailed cost data for the processes described are available, it was assumed that the TOCs represent 35% of the TCI costs, which is a typical value for chemical plant design \[109\]. In order to consider potentially higher costs for connecting the CHP plant with the upgrading equipment, 10% of the TCI cost was added as a contingency.

**Table 5.** Investment costs for the CHP plant and upgrading equipment (in M€)

<table>
<thead>
<tr>
<th></th>
<th>CHP</th>
<th>WP</th>
<th>TWP</th>
<th>PS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Equipment Cost</td>
<td>10.2</td>
<td>5.6</td>
<td>12.0</td>
<td>7.4</td>
</tr>
<tr>
<td>Total Capital Investment</td>
<td>26.0</td>
<td>17.6</td>
<td>37.7</td>
<td>23.3</td>
</tr>
</tbody>
</table>

**Exergoeconomic analysis:**

Exergoeconomic analysis makes it possible to calculate the production costs of multi-product processes. The so-called specific exergy costing (SPECO) approach, as described in Ref \[89\], was applied here. It allows for the allocation
of cost and exergy destruction once the fuel and product for each process component under consideration have been defined. More detailed information on exergy costing in general can be obtained from the literature, e.g. the book by Bejan, Tsatsaroni and Moran [85]. In general, exergoeconomic analysis divides the total costs of the process into process material stream costs and equipment costs (so-called Z-factors). The exergy cost balance of one process unit on an hourly basis can be written as follows:

$$\sum_{out}(c_{out}\dot{e}_{out})_k + (c_w \cdot \dot{W})_k = (c_q \cdot \dot{e}_q)_k + \sum_{in}(c_{in}\dot{e}_{in})_k + \bar{Z}_k$$

(Eq.11).

Compared to the energy balance, the exergy cost balance describes the sum of all exergy cost streams entering the unit, $k$, and the appropriate cost for unit operation, $Z_k$ (i.e. capital investment, operation and maintenance), which equals the sum of all cost streams (material and work) leaving unit $k$. The fuel and product rules defined by Lazzaretto and Tsataronis [89] clearly establish that for every exergy stream that undergoes change or conversion in a specific equipment unit, the specific equipment cost is charged to the product flow. This means that for all equipment units, so-called fuel and product streams need to be defined; in the case of multiple products, all product flows will be charged at the same specific cost. In total, all additional costs required for the unit operation and/or caused by exergy destruction will be charged to the product. In turn, this means that the specific costs of all fuel flows remain unchanged.

For the exergy analysis, the analysed processes were divided into functional units. They are highlighted and capitalised in Figs. 2, 5, 6 and 8. The definitions for fuel and products used for all functional units are given in Publication IV (Table A1). Exergy loss costs, which are accounted for at the system boundaries, will be allocated to the products weighted by their exergy content.

The SPECO analysis is based on the annual levelised cost approach. Therefore, all cost positions mentioned in Publication IV (Tables 3a–c) have been multiplied by an annuity factor in order to determine the annual levelised cost. Fixed-cost annuities have been calculated by multiplying the TCI cost with the constant cost escalation levelisation factor (CELF), which considers price inflation as well as price escalation. The CELF is calculated according to Eq. 12 with the lifetime of the investment ($n$) set to 20 years:

$$CELF = \frac{k(1-k^n)}{1-k} \cdot \frac{(1+i)^n}{(1+i)^n-1}$$

(Eq.12),

with

$$k = \frac{1+r}{1+i}$$

(Eq.13),

where $i$ is the interest rate and $r$ the rate of general inflation. Based on recent statistic data for Finland [6], $i$ and $r$ were set conservatively to 6% and 2%, respectively. With this data, the CELF was calculated as being 1.19. In addition, the plant operation cost which consists of maintenance and personnel costs have been considered as follows: The yearly maintenance cost was assumed to be 1.5% of the fixed capital investment. The personnel cost for the CHP plant has been estimated at 3 shifts per day with each of them requiring 6 persons (3 for fuel handling and 3 for plant operation). The yearly man salary was estimated at 30000€/year and a salary overhead and administration overhead
was added at 1.5 and 1.3, respectively [89]. For the operation of the upgrading equipment, the personnel cost has been estimated at 3 shifts per day, with each of them requiring 5 persons for TWP and PS and 2 persons for WP. In all cases: 2 persons would be needed for fuel handling and 2 or 3 for operating the equipment. The yearly man salary, the personnel and administration overhead was estimated with the same assumptions as given above.

Different prices and escalation rates can be expected for the different products (electricity, DH, WP, TWP and PS), as well as for the wood fuel. It was assumed, based on their fuel properties that TWP and PS could be sold at the same price per thermal unit as coal and heavy fuel oil, respectively. In order to estimate the changes in the price of wood, WP, coal and heavy fuel oil throughout the investment period, historical price developments in Finland based on data from Statistics Finland [6] and the Finnish Customs [7] were used. The price increase rate of the different fuels was recorded and averaged with respect to the available data history (a common practice in the energy business). This average rate of price increase was then used to make future price projections for the years 2014–32 by applying a linear correlation. Based on that step, the average price throughout the whole investment period (comparable to the annuity) was used in the economic assessment. All of the input data used, as well as the estimated time-averaged unit price for future commodities are supplied in Table 6 and further information on the data extraction process is given below.

• Power: The Finnish yearly average net prices (obtained from Ref. [6], excluding all taxes and charges) for private users (provided in 4 different categories based on the user’s total annual consumption) were averaged for the years 2004–13. Based on this data the average price increase over this 10-year period was calculated and used for price prediction with a linear model (constant yearly price increase). This scenario would constitute a best-case-scenario and thus give the highest possible earning from power sales. Potentially lower prices that result from selling electricity to key account customers or the Nordic options or spot market will be accounted for in a sensitivity analysis.

• DH: It is assumed that all heat produced can be sold to small private consumers and hence the Finnish yearly average net prices (obtained from Ref. [6], excluding all taxes and charges) for private users (provided in 5 different categories -after 2010 in three- based on the users’ total annual consumption) were averaged for the years 2004–13. Based on this data the average price increase over this 10-year period was calculated and used for price projection with a linear model.

• Wood chips / fuel: Statistical data on wood chips net price development in Finland is available for the years 2007–13 [6] and was used in order to calculate the average annual price increase rate. This rate has been used for the future price projection.

• WP: Statistical data of the WP net price development in Finland is available for the years 2008–13 [6] and was used in order to calculate the average annual price increase rate. The rate has been used for the future price projection.
• TWP: It is assumed that TWP is sold as a replacement for hard coal and that it can be sold at the same price per thermal unit. Data can be obtained from Ref. [6] for the years 2004–13. The final coal price in Finland consists of the coal import price which is subject to taxation by means of energy content tax, carbon dioxide tax and stockpile fee. We assumed that none of those are charged on TWP (TWP taxing was treated in the same way as with WP), and thus that TWP could be sold at the final market price of coal (excl. VAT). The future prices have been calculated with the 2013 tax-level.

• PS: It is assumed that PS is sold as a replacement for HFO and that it could be sold at the same price per thermal unit. Statistical data on the HFO price history for the years 2004–13 has been obtained from Ref [7]. In Finland HFO is subjected to tax and hence the final price includes the energy content tax, CO₂ tax and stockpile fee. For PS it is assumed that no CO₂ tax needs to be paid (as is the case for other lignocellulose-based biofuels in Finland).

Fig. 9 shows the recorded and projected price changes for the commodities considered.

<table>
<thead>
<tr>
<th>Table 6. Price projection for energy commodities in €/MWh</th>
</tr>
</thead>
<tbody>
<tr>
<td>Power</td>
</tr>
<tr>
<td>---</td>
</tr>
<tr>
<td>Price in 2013</td>
</tr>
<tr>
<td>Future price 2013–32</td>
</tr>
</tbody>
</table>

Figure 9. Price history and projection of applicable energy commodities (continuous line: statistical price data; dotted line: projected prices).
5. Results and Discussion

5.1. Combined Heat and Power Plant Operation

Naturally, the CHP plant integrated with the different upgrading processes showed substantially changed mass and energy balances. The following results are mainly based on the findings presented in Publication III, which discusses how WP, TWP and PS are integrated with the CHP plant and then compares them with the base case stand-alone CHP plant.

Due to the heat consumed during the upgrading processes, it is possible to realise lower DH demands. That means a 25–38% increase in operation hours, as shown in Fig. 10. PS integration leads to maximum operation hours, since the large amount of heat taken from the flue gases decreases the influence of the restriction set by the pressure of the feedwater tank (see Section 4.1.3). For the case involving PS integration at increased feedwater tank pressures of 3 bars, which is described in Publication II, the plant can operate at an upper limit of 8000 hours per year. The production of upgraded biomass increases the amount of biomass fuel consumed by the integrated CHP plant by a factor of approximately 2.3 when compared to the base case.

![Figure 10. Comparison of yearly operation hours and fuel input for the stand-alone CHP plant (base case) and the integration options.](image)

With longer operation hours, more DH is generated and the share of DH provided by the CHP plant increases from 75% to 87% for PS integration (see Fig. 11). Apart from greater capacity utilisation for the boiler, the amount of fossil fuel consumed by the heat-only boilers can also be reduced since less of the base load needs to be provided by those boilers. The electricity output is decreased by up to 8.7%, of which 88–95% can be attributed to the electricity consumed by the additional equipment (dryer, reactors, etc.). If not considered, the power output would increase slightly as a result of longer operation hours. As discussed in Publication II, it is possible to compensate for the loss in the amount of power generated by the CHP plant if the steam used for drying is extracted from the turbine at lower pressure. However, such a change to the
turbine would mean additional costs, especially in a retrofit situation, and thus it would need to be considered carefully. Finally, Fig. 11 also shows that a high amount of valuable product can be co-produced, which, with respect to its energy content, clearly exceeds the DH and power output. Since WP consumes the lowest amount of conversion energy, it has the highest product yield.

![Graph](image.png)

**Figure 11.** Yearly product output of the (integrated) CHP plant.

Because of the regulation strategy of adjusting the DH load by simultaneously adjusting the dryer load and heat extracted for pyrolysis or torrefaction, it is possible to operate the boiler at a 100% load with decreasing DH output. Fig. 12a shows how the boiler operates at full load until the DH supply is at 60–70%. The higher boiler load increases the amount of steam generated by the boiler (Fig. 12b). However, for PS and TWP the lower combustion temperatures, which are caused by lower combustion air temperatures, impede a constant rate of steam. The lower combustion air temperatures are a result of the heat extracted from the flue gas for the torrefaction and pyrolysis reaction, respectively. This effect is more distinct for PS, since the pyrolysis reaction requires more heat than the torrefaction reaction. The steam mass flow drops sharply below a DH load of 60–70% due to the reduced flow of fuel, and hence, of flue gas, which also reduces the efficiency of the heat exchange equipment.

At a continuous full boiler load below 60–70%, the dryer would reach such high loads that the recycled condensate would raise the feedwater beyond the saturation point (Fig. 12c) at maximum feedwater tank pressure (2 bar). Thus, we decreased the boiler load. For PS integration, the boiler can be kept at higher loads for a longer period of time because a larger amount of heat is extracted from the flue gas for the pyrolysis reaction. Consequently, torrefaction ranks between pyrolysis and pelletizing, which is controlled solely by the dryer load. The required reduction in the fuel input to the boiler is the reason that the product output decreases for loads below 60–70% (Fig. 12d). The simulation results also show that all processes, pelletisation and production of TWP and PS are quite possible within the operational limits of the CHP plant. Integrating PS with the heat extracted from the flue gas before the first superheater affects the steam and flue gas temperatures the most.
Results and Discussion

Due to the heat extracted for fast pyrolysis, the live steam temperatures decrease by up to 36 °C (Fig. 13a) at the point where we see the highest pyrolysis product output (Fig. 12d). This relates to the decreased mass flow and heat content of the flue gas, which reduces the amount of heat exchanged in the respective superheater 1. Since no structural changes occur in the high temperature equipment with respect to pelletisation and torrefaction, the live steam temperature can be maintained. The increased amount of steam generated in the more heavily loaded boiler almost compensates for the amount of steam required for drying. The steam enthalpy flow to the turbine (after extraction to the dryer) remains nearly constant (Fig. 13b), and thus we can expect that the amount of gross power generated will be similar to that generated for the stand-alone CHP for the integrated cases. This indicates, as mentioned above, that power generation is not significantly impeded by the dryer operation; rather, it is significantly impeded by power demand of the upgrading equipment.

When integrating TWP and PS, attention must also be paid to the flue gas temperature. Fig. 13c shows that the temperature can drop far below the minimum temperatures at the stand-alone CHP plant. Since we adjusted those temperatures to 143 °C in this study to adequately compare the alternatives, this means that there is a potential for increased product yields at lower flue gas temperatures provided that corrosion risk can be ruled out. The flue gas
temperatures drop the most for TWP integration because the heat needed for torrefaction is extracted just before the air-preheater, which in turn is solely penalised by the reduced heat rate. For PS, the reduced heat rate of the flue gas is accounted for in the air-preheater and the economiser, which explains the lower boiler approach temperatures (Fig. 13d). Due to higher feedwater temperatures for the economiser, the approach temperatures increase for WP and TWP, reaching the saturation point at low loads with harmless levels of maximum steam content of 1.5%.

Figure 13. Critical process parameters (r.t., reduced temperature, see section 4.1.1).
5.2. Sensitivity Analysis

In Publication II, we conducted a sensitivity analysis in order to assure that the results obtained for the case study could be transferred to other DH systems as well and that we made the correct assumptions.

One major assumption in the study has to do with the DHN’s heat duration curve. In order to represent substantially colder and warmer years, we fixed the respective minimum and maximum loads and varied the median value by +/- 30%. We adjusted the remaining values via a quadratic function. As a consequence, the total DH demand changed by +/- 19%. The resulting heat duration curves with varied slopes are presented in Fig. 14.

Table 7 shows that the variation in the boiler load for the base case leads naturally to a 14% increase and a 27% decrease in the operation hours for the higher and lower DH demand, respectively. PS integration in both cases allows for a 31% and 66% increase in operation hours. The net power output for the higher DH demand (CHP&PS DH +19%) is slightly lower, just as it is for operations pertaining to the DH design load (CHP&PS), due to the amount of power consumed by the additionally installed equipment. However, this penalty can be overcome and the losses caused by the additional equipment can be compensated for as the result of a sharp increase in the number operation hours in the case of low DH demand.

<table>
<thead>
<tr>
<th>Case</th>
<th>DH output [GWh/y]</th>
<th>Power output [GWh/y]</th>
<th>PS output [GWh/y]</th>
<th>Operation hours [d/y]</th>
</tr>
</thead>
<tbody>
<tr>
<td>CHP</td>
<td>70.8</td>
<td>22.9</td>
<td>-</td>
<td>212</td>
</tr>
<tr>
<td>CHP DH + 19%</td>
<td>85.6</td>
<td>27.5</td>
<td>-</td>
<td>242</td>
</tr>
<tr>
<td>CHP DH - 19%</td>
<td>49.0</td>
<td>15.8</td>
<td>-</td>
<td>154</td>
</tr>
<tr>
<td>CHP&amp;PS</td>
<td>82.8</td>
<td>22.2</td>
<td>103.2</td>
<td>293</td>
</tr>
<tr>
<td>CHP&amp;PS DH +19%</td>
<td>96.9</td>
<td>27.3</td>
<td>87.4</td>
<td>318</td>
</tr>
<tr>
<td>CHP&amp;PS DH -19%</td>
<td>63.9</td>
<td>15.8</td>
<td>109.3</td>
<td>256</td>
</tr>
</tbody>
</table>

In summary, PS integration improves the performance of the CHP plant in all operational situations and it is particularly beneficial in years when DH loads
are lower than the design load. This might be important for the future, when the DH load can be expected to decrease due to improved building insulation (e.g. net-zero energy buildings).

Not every biomass-fired CHP plant is equipped with a bubbling fluidised bed boiler. Often applied technologies consist of circulating fluidised bed boilers and grate-fired boilers, with the latter type of boiler often applied in plants with scales below 20 MWth. Publication II demonstrates that the boiler type and the adjusted integration approach only influence the mass and energy balance of the system to a minor degree. However, smaller plants often generate steam at a lower pressure and temperature to reduce investment costs. The results obtained from the same simulation model with reduced live steam values of 16.5 bar and 400 °C are presented in Table 8. The increase in the number of operation hours and fuel input was in the range of the larger plant (35% and 120%, respectively). However, due to the lower live steam pressure and temperature, the loss in power was higher, accounting for 10%. In general, PS integration is also beneficial for smaller plants with lower steam parameters.

Table II also shows that using a minimum boiler load of 30% had only a negligible influence on the results. In addition, it is shown that the simplified process of mathematically subtracting the heat generated by permanent torrefaction and pyrolysis gases from the boiler fuel input did not affect the results.

Table 8. Influence of lower steam values on the plant’s performance

<table>
<thead>
<tr>
<th></th>
<th>CHP 60 bar</th>
<th>PS 60 bar</th>
<th>CHP 16.5 bar</th>
<th>PS 16.5 bar</th>
</tr>
</thead>
<tbody>
<tr>
<td>Operation time [days]</td>
<td>212</td>
<td>293</td>
<td>212</td>
<td>288</td>
</tr>
<tr>
<td>District Heat [GWh]</td>
<td>70.8</td>
<td>82.8</td>
<td>70.8</td>
<td>82.3</td>
</tr>
<tr>
<td>Fuel In [GWh]</td>
<td>106.1</td>
<td>241.7</td>
<td>99.5</td>
<td>219.7</td>
</tr>
<tr>
<td>Product Out [GWh]</td>
<td>-</td>
<td>103.2</td>
<td>-</td>
<td>90.7</td>
</tr>
<tr>
<td>Net Power [GWh]</td>
<td>22.9</td>
<td>22.2</td>
<td>15.2</td>
<td>13.6</td>
</tr>
</tbody>
</table>

5.3. Thermal Efficiency

Fig. 15 shows that the overall TE represents the energy conversion- or product efficiencies of the individual upgrading processes. Compared to the base case they were somewhat higher for WP and TWP, but decreased slightly in the PS case. Compared to WP and TWP longer operation hours allow for more heat and power production, which results in higher efficiency levels for PS integration. In addition, the power-to-heat ratio benefits from PS integration too. Likewise, because of the longest operation hours and additional DH generation, PS can reduce the amount of fossil fuel consumed in the heat-only boiler the most.

Based on the results presented above, an evaluation of the process only gives a few indications of which process performed better from an energetic standpoint and further evaluation tools need to be applied in order to obtain more conclusive results.
5.4. Primary Energy Efficiency and CO₂ Emissions

As mentioned in Section 4.2.2, calculating the DHN’s PEF requires the input of the PEF for the generated electricity that is thought to be replaced. Since no general guideline is provided regarding which value to choose, three different PEF scenarios for electricity have been compared in Publication III, with the values given in Table 4. The values show that cogenerated electricity could replace the coal condensing power (PEFₚ = 4.05), the power generated using the average generating efficiency for Finland (PEFₚ = 3.11) and the power generated using the average generating efficiency for Norway (PEFₚ = 1.33). The latter represents a highly renewable energy system. Keeping in mind that the PEF describes a reciprocal efficiency figure, with the low values indicating high efficiency levels, Fig. 16 shows that, depending on the replaced electricity’s PEF, the DHN’s PEF can be improved by up to 35% via WP integration. This is due to the high product yield of the WP. For the case involving a high share of renewable electricity (PEFₚ = 1.33), the improvement potential was lower, but still considerable at 25%. The lower PEF improvement potential for PS integration remained constant at 17% regardless the PEFₚ.
**Fig. 17** also shows that the CO₂ emission coefficients are lowest for WP integration due to the high production rate. The results further indicate that integrating the upgrading processes can still decrease CO₂ emissions even in energy systems that are already highly dependent on renewables (CO₂ emission coefficient equals 9). It should be noted that EN 15603 specifies that a PEF below one will be set to one. The same reasoning applies to CO₂ emission coefficients below zero – they, too, should be set to zero. Still, the reduction potential was considerable and integrated WP production performed best with respect to the DHN’s CO₂ emission coefficients and PEFs.

**Publication II** deals in more detail with integrating different product usage scenarios for PS. PS can either be sold as a replacement for fuel oil. Additional oil and char can be separated and the oil can be primarily used in the DHN’s heat-only boiler. Excess oil would be sold as a replacement for fuel oil. The char could then be either used in-situ, sold to the market as a replacement for coal or sequestered as a carbon sink, as suggested in a recent study [73]. It was shown that, owing to the higher PEF of coal, separating the PS and selling the char as a replacement for coal would increase PE efficiency and decrease CO₂ emissions the most. The in-situ combustion of char and its sequestration would apparently be less beneficial.

**Figure 17.** Comparison of different CO₂ emission coefficient (CCO₂) scenarios.

### 5.5. Exergy Analysis

**Fig. 18** shows that the overall EXE can be increased considerably from 24% to 55%, 50% and 46% for WP, TWP and PS, respectively. This clearly reveals the high EXE levels of the additional functional process units (which are explained in Section 4.2.3, units K and I), which were usually well over 90%. Comparably high EXE levels for the torrefaction and fast pyrolysis process have been reported in Refs [62] and [88]. There are two main reasons for the high efficiency levels. First, the biomass fuel is broken down chemically to a much less extent than when it is combusted. No chemical reaction occurs in the case of WP, which basically conserves the chemical exergy. Second, the process temperatures are much lower than they are for combustion, which means that
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less high temperature heat is converted to low-temperature heat, as was the case in the boiler. This heat “quality conversion” from high to low temperatures naturally decreased both the EXE levels and the Carnot efficiency of the cycle. When taking into account the lower temperature levels at which heat transfer occurred, low EXE levels could only be found in the evaporator, the superheaters and the economiser.

Figure 18. Exergy efficiency of the various conversion units.

This fact is also represented by the exergy destruction rates presented in Fig. 19. The relative exergy destruction rates represent the exergy destruction in each unit relative to the whole process. We calculated them as a ratio of the exergy destruction rate observed in a specific unit to the exergy destruction rate for the entire process. The calculation was based on the exergy stream data provided in Tables A2–5 (Publication IV) and by applying the fuel and product rule, as defined in Table A1 of Publication IV. The relative exergy destruction rate for the different units diminished slightly because of the additional (small) exergy destruction within the additional equipment. In general, the units’ relative exergy destruction rates, with the exception of superheater 2 in the PS case, did not show significant changes. This can be interpreted as a sign that the integration process does not hamper the plants’ operability. In the PS case, superheater 2 experienced heavily reduced mass flows, which decreased the absolute amount of exergy destruction, and hence, its relative exergy destruction rate. Publication IV provides a more detailed discussion of the exergoeconomic analysis; a detailed analysis is presented there in Tables 5a–d.

In summary, we suggest that the exergetic analysis also indicates that WP production is more favourable compared to TWP and PS production. The only remarkable difference appeared in the calculation for the EXE levels, and here integrated WP production clearly performed best, as demonstrated in the PE efficiency assessment provided in Publication III. Thus, the exergy analysis provides more insight into the sources of exergy destruction and reveals the points that further process optimisation should focus on.
5.6. Exergoeconomic Analysis

The results of the exergy analysis do not yet give concise enough information to base an investment or development continuation decision on. As the exergoeconomic analysis yields production costs for all of the allocated products based on their exergy content, those costs can be compared to the potential turnover obtained on the product markets. Those results are presented in Table 9. The results are based on the projected future average price for the fuel and products, as explained in Section 4.2.4. As a result of the increased operation hours, the operation costs of the CHP plant would increase by 15%, 17% and 22% for WP, TWP and PS, respectively. As a result, the annual fuel costs would increase, too. Naturally, the total cost for the upgrading equipment would vary based on the investment cost, the linked operation cost and the throughput-dependent fuel cost. The overall annual values would roughly double for the integrated cases, but interestingly they would vary by only 4.8% and 12.8% between WP and PS and WP and TWP, respectively. Based on the TCI for the upgrading equipment (see Table 5), higher variations could have been expected,
but the results undermine the importance placed on the fuel costs, which is a common finding for biomass processing plants.

When comparing the exergy-allocated production unit cost with the market prices, we found that power production has a negative profit margin, whereas DH and PS are always positive, i.e. profitable. The production of TWP and WP is not viable from an exergoeconomic standpoint due to the projected low prices and slow price development projected for coal and WP. Product markets for PS and TWP are currently far from being established and thus the prices used in this analysis only can give a rough estimation. However, in the opinion of the author, the assumption that TWP and PS could be sold at the same price per thermal unit (€/MWh) as their fossil competitors (coal and heavy fuel oil) constitutes a most neutral and objective manner for the time being. TWP are a premium product since the usage of TWP is considered carbon-neutral and TWP has the potential to replace fossil coal at high co-firing rates without modification of existing coal-fired boilers. The fact that the TWP price is estimated to be lower than the price of WP might appear peculiar; however, this is because it relates to the price of coal, which has been traded at lower prices than WP. It must be pointed out that the coal price used has been corrected with the Finnish carbon tax, i.e. the price projected is carbon tax-free since TWP originate from biogenic raw material. One reason for the higher price of WP might be that WP are not a typical fuel for medium- to large sized biomass power plants in Finland and Sweden. They usually use wood chips as a fuel, which is much cheaper. WP are rather sold to small-scale applications and private customers that are subjected to higher prices. Energy companies might be willing to pay a higher price if they were compensated, for instance by means of the EU Emission Trading System (EU ETS).

In summary, integrating WP and PS would increase the yearly profit by 12% and 110% respectively. However, exergoeconomically, the increased profit of WP is related to the increased DH production and not to the WP sales. Also the small profit increase of WP integration of 0.38 M€ p.a. would hardly justify an investment of 17.6 M€, due to a long payback period. We found that integrating TWP would decrease the profit due to the high investment costs and the low price for coal. For the sake of completeness, also results were presented for cases in which the plant owner would need to pay a transmission fee for constructing and maintaining the DHN. We assumed that a value of 30% of the DH turnover would be reasonable. However, we did not take into account the transmissions losses. Then, the profits would certainly be lower, and in the case of TWP integration even negative.

With respect to the price development, we observed that the PS-integrated system will provide the same profit as WP when the average price over the investment period is 65.72 €/MWh, whereas it will remain the same for TWP when the average price is 41.08 €/MWh. This corresponds with the PS price being 31% and 57% lower than projected. In turn, in order to make TWP competitive with PS, the cost of using coal as fuel (thereby considering the actual fuel and the cost for the emission trading certificates) would need to be as high 97.22 €/MWh, which is double the price projected in this work.
Table 9. Results of the economic evaluation

<table>
<thead>
<tr>
<th></th>
<th>CHP</th>
<th>WP</th>
<th>TWP</th>
<th>PS</th>
</tr>
</thead>
<tbody>
<tr>
<td>Annual payment CHP plant [M€]</td>
<td>1.54</td>
<td>1.54</td>
<td>1.54</td>
<td>1.54</td>
</tr>
<tr>
<td>Operation Cost CHP [M€]</td>
<td>1.27</td>
<td>1.46</td>
<td>1.49</td>
<td>1.55</td>
</tr>
<tr>
<td>Fuel Cost CHP [M€]</td>
<td>3.48</td>
<td>4.51</td>
<td>4.49</td>
<td>4.66</td>
</tr>
<tr>
<td>Annual Payment Upgrading [M€]</td>
<td>-</td>
<td>1.04</td>
<td>2.24</td>
<td>1.38</td>
</tr>
<tr>
<td>Operation Cost Upgrading [M€]</td>
<td>-</td>
<td>0.99</td>
<td>1.54</td>
<td>1.28</td>
</tr>
<tr>
<td>Fuel Cost Upgrading [M€]</td>
<td>-</td>
<td>3.51</td>
<td>3.44</td>
<td>3.27</td>
</tr>
<tr>
<td>Overall Annual Production Cost [M€]</td>
<td>6.30</td>
<td>13.05</td>
<td>14.73</td>
<td>13.68</td>
</tr>
<tr>
<td>Power Unit Cost [€/MWh]</td>
<td>178.57</td>
<td>173.67</td>
<td>184.36</td>
<td>178.22</td>
</tr>
<tr>
<td>DH Unit Cost [€/MWh]</td>
<td>31.26</td>
<td>28.35</td>
<td>29.30</td>
<td>29.39</td>
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<tr>
<td>Bio-product Unit Cost [€/MWh]</td>
<td>-</td>
<td>57.66</td>
<td>73.25</td>
<td>70.55</td>
</tr>
<tr>
<td>Annual Power Generation (energy) [GWh]</td>
<td>22.86</td>
<td>21.60</td>
<td>20.88</td>
<td>22.29</td>
</tr>
<tr>
<td>Annual DH Generation (energy) [GWh]</td>
<td>70.85</td>
<td>79.60</td>
<td>80.60</td>
<td>82.80</td>
</tr>
<tr>
<td>Annual Bio-product Generation (energy) [GWh]</td>
<td>-</td>
<td>122.30</td>
<td>116.40</td>
<td>103.20</td>
</tr>
<tr>
<td>Wood Chips Market Price [€/MWh]</td>
<td>32.60</td>
<td>32.60</td>
<td>32.60</td>
<td>32.60</td>
</tr>
<tr>
<td>Power Market Price [€/MWh]</td>
<td>137.00</td>
<td>137.00</td>
<td>137.00</td>
<td>137.00</td>
</tr>
<tr>
<td>DH Market Price [€/MWh]</td>
<td>89.30</td>
<td>89.30</td>
<td>89.30</td>
<td>89.30</td>
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<tr>
<td>Bio-product Market Price [€/MWh]</td>
<td>-</td>
<td>53.40</td>
<td>48.80</td>
<td>95.70</td>
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<tr>
<td>Power Turnover [M€]</td>
<td>3.13</td>
<td>2.96</td>
<td>2.86</td>
<td>3.05</td>
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<tr>
<td>DH Turnover [M€]</td>
<td>6.33</td>
<td>7.11</td>
<td>7.20</td>
<td>7.40</td>
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<tr>
<td>Total Turnover</td>
<td>9.46</td>
<td>16.59</td>
<td>15.74</td>
<td>20.32</td>
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<tr>
<td>Total Profit [M€]</td>
<td>3.16</td>
<td>3.54</td>
<td>1.00</td>
<td>6.64</td>
</tr>
<tr>
<td>Total Profit [M€] 30% of DH turnover for piping</td>
<td>1.26</td>
<td>1.41</td>
<td>-1.16</td>
<td>4.42</td>
</tr>
</tbody>
</table>

The results presented above are heavily dependent on the price assumptions, and hence, we independently varied the projected fuel and product market prices by +/- 50%. The TCI cost, especially for the upgrading equipment, also contains a high degree of uncertainty; we varied it as well by +/- 50%. The results are presented in Fig. 20.

All of the cases show that the plant’s dependence on the electricity market price is minor compared to the level of variation in DH and bio-product prices, respectively. Profit variations are approximately +/- 50% for the CHP plant, but they can be reduced to +/- 23% for the PS plant (Fig. 20d), which makes it clearly less dependent on changes in the electricity prices. In the background of current exceptionally low electricity prices, it was calculated that the CHP plant would reach zero profit (at constant DH price) if the electricity price was as low as -1.38 €/MWh. In comparison the PS and the WP (Fig. 20b) cases (at constant DH and product prices) would reach zero profit if the electricity price was as low as -161 €/MWh and -27 €/MWh, respectively. Hence we can state that such integrated systems are a powerful tool to decrease the CHP plants’ economic vulnerability in times of low electricity price.

While the plant’s dependence on the DH price is generally high, possible profit variations can be decreased by 44 percentage points when integrating PS. Due to the negative profit margin of TWP, the plant is extremely sensitive to variations in the DH price (Fig. 20c) and the profit varies by a factor of 3.6. at 50% DH price variation.

Being the major product, the integrated systems’ profit naturally depends also heavily on the bio-product prices. Changes in the PS and WP profits react more distinctly, varying by -8 and +18 percentage points, compared to changes in the DH price. Again, for the TWP system profit variation are highest.
Regarding the investment costs, the profit varies by +/- 31% with a 50% variation in the investment cost of the stand-alone CHP plant (Fig. 20a). For the integrated options, we kept the cost of the CHP constant and only varied the price of the upgrading equipment. The figure shows that TWP reacts most sensitively due to high FCI and low profit margins. In general, with constant CHP plant investment costs, the influence of the investment costs for PS and WP equipment would only affect the profitability of the plant to a small extent (below 20%). We calculated that even at zero equipment cost, the calculated profit for TWP would only reach half of the profit of PS. If the PS equipment were 2.7 times more expensive, then the profit would be the same as in the WP case.

For increased feedstock prices, the figure shows that the integrated systems react more sensitively mainly due to the greater amounts of fuel used in the various processes. Except in the case of the highly sensitive TWP, WP and PS react only moderately to the increase in the wood chip price when considering that they consume 2.3 times more fuel compared to the CHP case. WP is more sensitive to variations in the fuel price because changes in the bio-product costs have a greater influence on the profit margin.

We can conclude that integrating PS reduces the uncertainty and the risk of the plant operation considerably. On the other hand, the DH market is quite stable and price variations are less likely than in, e.g. the quite volatile oil market; this is due to its local nature and high infrastructure costs, which restrict the possibilities for market liberalisation. However, the future price for WP increased only moderately compared to wood chip prices (23% versus 57%, see Table 6). This price difference defines the potential profit margin from input to product, and if a steeper price increase occurs, then the WP’s price-profit dependence might become more beneficial.

To summarise, integrating PS with a municipal CHP plant has the potential for achieving the greatest profitability under the assumed future market scenario. The production of WP seems to be the second-most profitable option. Replacing coal by producing TWP is penalised due to high investment levels and a low fuel-product profit margin. In addition the integration has the potential to keep the profitability on a higher level in times of low electricity prices.

Interestingly, it appears that the least efficient (both exergetically and energetically) process has the potential to be most profitable due to expected steep price increases for heavy fuel oil. We should mention that removing char from the PS would yield a product that might be comparable to light-fuel oil, which is traded at a much higher price but could be produced only in smaller volumes. In addition, PS and pyrolysis oil form an interesting inter-stage product that can be further upgraded to transport fuel either by de-oxygenation or gasification and subsequent Fischer-Tropsch synthesis. This could further increase market demand and price as soon as technical and political barriers are overcome, and 2nd generation biofuels would then enter the transport fuel market. Finally, we should mention that all of the products discussed here can be produced at competitive prices under current tax and subsidy regimes.
Figure 20. Results for the price sensitivity analysis.
6. Concluding Remarks

The following section presents the main conclusions, discusses the scientific significance of the thesis and provides an outlook for future work.

6.1. Conclusions

Publication I demonstrates that the PE method derived from EN 15603 can be adapted with relative ease to give a general evaluation method that can be applied to biofuel processes other than CHP generation. It gives clear results and good insight into the actual impact of the conversion systems. The advantages and disadvantages of fossil PE use and CO₂ emissions then become evident. The saving potential for PE and the CO₂ mitigation potential can be calculated in a simple manner. The results are more in line with the final user since the method is product oriented (“how much do I need to produce one unit”) rather than input oriented (“how much can I produce with one unit raw material”). This approach is also easier for technical laypersons to understand, which might be important when it comes to deciding whether a technology should be further developed or not. It lacks the input of economic data, which is a disadvantage when it comes to evaluating the economic feasibility of the approach. However, it has the advantage of potential estimation with respect to developing the most efficient biomass conversion systems for the future. Most likely, the method can be applied to other energy- and material conversion systems as long as basic process knowledge is available. The PE method then makes it possible to compare systems at a relatively early stage of development. Furthermore, it should make it possible to compare different multiproduct energy systems. The main advantage of the method is the clearly defined system boundary, which often seems to be the bottleneck when discussing energy efficiency.

Publication II studied the feasibility of integrating PS production within a CHP plant. We concluded that integration is possible for all operation points. Attention should be given to the exhaust temperatures after the air-preheater, since they can drop to below the temperatures reached in a stand-alone operation. The steam dryer should be designed for low exhaust temperatures so that a large amount of the available heat can be utilised from the flue gases which are used as a carrier medium. There is still room for improvement when it comes to adapting the turbine extraction nozzles. This would help compensate for the power losses caused by the additionally installed equipment, but cost-effectiveness of such improvements still needs to be established. Due to lower possible DH loads, integration makes it possible to increase the operation hours of a plant by as much as 57%. If possible, the feedwater tank pressure should be as high as possible, as this clearly improves the energetic and environmental performance the most. In order to increase the pressure after the dryer, an additional direct mixer could be installed as well so that the integrated process could be operated within the safety and material restrictions set by the original
feedwater tank. Due to the lower efficiency of the PS process compared to the stand-alone CHP process, the overall efficiency as an average of the efficiencies of the various processes will drop slightly. Regarding the uses of the product, we conclude that as much product as possible should be sold to the market since it clearly improves the DHN’s PE efficiency and CO₂ savings. However, this result depends to a large extent on the assumptions made regarding the PEF of the replaced energy form. In total, an integrated small- to medium sized plant has the potential to mitigate CO₂ emissions by up to 60kt annually, which is equal to 0.11% of all Finnish emissions. We further conclude that the benefits of integrating PS production within a CHP plant can also be accounted for in years with considerably higher and lower DH demand. We also demonstrated that different boiler concepts and sizes can be integrated as well, with none of them exhibiting critical or unbeneficial behaviour. To summarise, we conclude that integrating PS production with CHP plants is thermally possible and beneficial with respect to the energetic and environmental performance. If indirect steam drying is applied, the condensate should be fed back at pressure levels that are as high as possible. With this strategy, future CHP plants will be able to provide valuable products at greater levels of efficiency and help mitigate global CO₂ emissions.

Publication III compares the integration of PS production with the competing biomass upgrading technologies of pelletisation and torrefaction. Integrating both processes is possible within the operational limits of a CHP plant. The results show that yearly operation hours were extended by 25–38%, resulting in 17.4–21.6% more DH generated in the CHP mode. When we considered upgraded biomass as well, the total product output was 2.2–2.4 times higher. Conversely, annual power output decreased by 2.5–8.7%, mainly due to the power requirements of the additionally installed equipment. WP production resulted in the highest product yield and, for the evaluation method used, it performed best both energetically and environmentally. The study also demonstrated that the choice of the PEF applies for the produced electricity and the CO₂ emission coefficients greatly impact the results. Hence, the choice of the PEF and CO₂ emission coefficient can influence future DH and CHP strategies. Further critical discussion is crucial with respect to this issue. In summary, the study shows that integrating WP, TWP and PS biomass production with a CHP plant by increasing the yearly boiler workload leads to improved PE efficiency, reduced CO₂ emissions, and, compared to stand-alone generation, also to substantial fuel savings.

In Publication IV, we conducted an exergoeconomic analysis on integrating WP, TWP and PS production with a CHP plant. We compared the results for each process with one another and to the results for the CHP stand-alone plant. Integrating the processes causes the annual cost to rise considerably. The highest level of investment is required for the TWP integration process and it mainly results from the large reactor volume required due to the long residence time of approximately 20 minutes. The combustion process in the boiler and the flue gas-water heat exchangers cause the highest rates of exergy destruction. We attribute this to the high rate of irreversible chemical breakdown during
Combustion and to the heat exchange from hot flue gas to moderate-tempered steam. Conversely, the integrated processes show very low rates of exergy destruction, which results in high EXE levels. The total EXE of the systems compared to the stand-alone CHP plant demonstrate that the systems' efficiency levels can be improved substantially from 24% to 55%, 50% and 46% for WP, TWP and PS, respectively. This is well in line with results we obtained when assessing the PE efficiency in Publication III. Furthermore, in order to evaluate the economic feasibility of the systems studied, Publication IV provides future cost projections for fuel and products based on historical price data. The economic results reveal that the investment and operational costs represent only a relatively small share of the total annual production costs; i.e. all systems are fuel-cost dominated. By basing the cost allocation on the exergy content of the products, we demonstrated that the cost of generating electricity and producing TWP and WP has a negative profit margin, but that this is balanced out by the other products DH and / or PS, respectively. With the projected future prices, PS integration allows for the highest profit, a profit that is 2.1 times higher than for the stand-alone CHP plant. We calculated that WP integration would yield 47% less profit than PS, whereas TWP would lead to a three-fold reduction in the profit of the stand-alone CHP plant. If we assume a transmission fee constituting 30% of DH turnover, then the PS would yield a profit that is 3.5 times higher than for the stand-alone CHP plant and perform 3 times better than WP.

The findings highlight the importance of the assumed product and fuel market prices and investment costs, and consequently, we carried out a sensitivity study in which the various prices and costs were varied individually by +/- 50%. We found that integrating WP and PS decreases their dependence on the power and heat market price. Even though it represents the least efficient upgrading technology, with a moderate investment cost and a high future potential at high bio-oil demand and price, PS integration constitutes the best integration option from an economic standpoint.

Finally, in Publication V we compared the process evaluation methods, thermal efficiency (TE), exergetic efficiency (EXE) and primary energy (PE) analysis, used to assess CHP-integrated biomass upgrading, by applying them to a case study on Bio-SNG production. We modelled the Bio-SNG production chain using Aspen Plus and Microsoft Excel. With the advantages and disadvantages of each method being summarised in Fig. 21, we can state that the TE method gives a good indication of the general efficiency of the process. It is relatively simple to use and widely applied in practice, but it lacks the ability to reflect the use of global energy resources. Furthermore, it does not consider the thermal quality of the in- and outputs, and energy allocation may pose challenges. The PE efficiency describes quite well the influence of the process on the use of global PE resources if the PEFs are available for inputs and outputs. By means of the power-bonus method, it is possible to allocate PE to the products since the PEFs add thermal quality information. With TE, all products can be considered equally valuable. Compared to TE and EXE, PE efficiency is a product-oriented assessment
method. However, the results depend to a large extent on the methodological assumptions, which can lead to improved or impaired PE efficiency for the SNG process.

The exergy method reveals the potential for improving the process units. Based on the measures obtained in the study, we can clearly specify which process units account for the main thermodynamic losses. The method sums up different forms of exergy, all of which include thermal quality information. Exergy losses are “additive”, which means that a process can be divided into sub-processes and the sum of the exergy losses resulting from these sub-processes will equal the losses resulting from the overall process both quantitatively and qualitatively. The main drawback of the method is the fact that the calculations are relatively complex and the data necessary for conducting the analysis might sometimes be difficult to obtain.

Figure 21. Comparison of the energy efficiency assessment methods.

### 6.2. Scientific Significance

For this thesis, we developed simulation models of three different biomass upgrading processes that could be integrated with a communal CHP plant and would yield valuable high-grade energy products. The upgrading technologies can be applied to versatile biomass feedstock. The study demonstrated that integrating biomass upgrading with a communal CHP plant does not impede the plant’s operation; in fact, it is highly beneficial energetically, environmentally and economically.

We were the first to model and analyse in detail the mutual influences of the CHP plant part-load operation with the integration of various biomass upgrading processes. The applied multi-period model enabled us to analyse in detail how the parameters both, on the flue gas side and on the steam side are influenced by the integration. Using the integration approach with a fixed DH demand assures that all of the heat being generated can essentially be utilised
Concluding Remarks

within the DHN. Often the existing literature has described the opposite type of integration approach, i.e. an approach in which a CHP plant integrated with a biofuel process (e.g. SNG production) and the heat demand is left as a variable without assuring that the heat can really be utilised.

As a novelty, we evaluated different internal and external usage pathways for the fast pyrolysis product fractions (pyrolysis oil and char) with respect to their energetic and environmental performance. We concluded that it is less beneficial to use the products internally with respect to PE consumption and CO₂ emissions.

The work further provides a detailed exergy analysis of the three upgrading processes, which we also compared under similar conditions. We did the same when it came to the economic evaluation. The work clearly indicates the economic feasibility of the integration of PS with small- to medium-scale plants under current tax regimes.

Since the PE analysis was based on the existing legal framework, we successfully developed a general method for evaluating and comparing biomass-based processes. We compared the method objectively to two other commonly used assessment methods, namely thermodynamic efficiency analysis and exergy analysis. We addressed the benefits of each method and concluded that process evaluation should not be based on thermodynamic efficiencies; rather, it should be based on PE. For detailed process analysis, exergy analysis makes it easier to address any allocation issues. The findings are supported by the combination of results presented in Publications III and IV, which show that it is crucial to include economic data when evaluating the final process. This indicates that economic analysis should be included in any process analysis at as early a stage as possible.

This thesis defines a pathway of how existing and future CHP plants can be integrated with biomass upgrading processes in order to form economic viable biorefineries. We demonstrated that such integrated CHP plants can play a major role in reaching the EU’s 20–20–20 goals by utilising more biomass at reduced CO₂ emissions and increased primary energy efficiency.

6.3. Future work

The following sections point out possible future work. Future work can be divided into two aspects. First, we discuss the points where the process design and the technologies applied should be further analysed. Second, it is discussed how the applied evaluation methods can be generalised for application in a broader context.

6.3.1. Technological Aspects

With the choice of using drying technology at an early stage of this study, future studies should consider advanced low-temperature drying concepts, such as water-heated belt dryers. Allowing all steam to expand in the turbine should make it possible to generate more power, if desired, and at least partly compensate for the lower amounts of power generated by the systems discussed
here. Low-temperature drying could also open up possibilities to recover heat when cooling the pyrolysis and torrefaction products. If such heat recovery is possible, the efficiency of the aforementioned processes should also improve because those particular cooling processes account for the major heat losses. However, sufficient control of the DH load and bioproduct quality must be assured at all operation points.

Compared to the stand-alone CHP plant, where the DH load is controlled solely by the fuel input, the integrated systems require more advanced control algorithms since the fuel input, pressure in the feedwater tank and the load of the upgrading equipment all need to be simultaneously adjusted in order to control the overall load. Such control strategies might need to be developed.

Due to a lack of availability of part-load performance data for the upgrading equipment, we made the assumption that it can be operated at decreasing loads with a constant energy conversion efficiency. However, in reality this will unlikely be the case. If the concepts described in this study become operational, it will be especially important to obtain part-load performance data on the dryer and the torrefaction and pyrolysis reactors, and experimental work will be crucial in order for this to be successfully done.

6.3.2. Methodological Aspects

We have demonstrated that it is possible to assess new multi-product bioprocesses using the PE method. It can be used to assess the system’s influence on global PE resources and also makes it possible to straightforwardly assess the potential for mitigating CO₂ levels. However, it is crucial to develop general and mandatory guidelines for PEF that can be used to make fair and generally acceptable comparisons. Therefore, databases on consensual PEF and CO₂ emission factors need to be developed. The databases should be updated constantly in order to assure that changes in future energy systems will always lead to reductions in the amount of PE consumed and in CO₂ emissions.

In this study, the amount of energy allocated to products was based on the power-bonus method, as suggested in the related EU standard. To better understand and improve the method, other allocation methods should be analysed, for instance those based on exergy or economic value. Such allocation methods might make it possible to apply the PE concept to non-energy industries (e.g. steel making), and thus help to develop a universal process efficiency indicator.
Nomenclature

Abbreviations

<table>
<thead>
<tr>
<th>Abbreviation</th>
<th>Description</th>
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<tbody>
<tr>
<td>BFP</td>
<td>Biomass Fast pyrolysis</td>
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<tr>
<td>CELF</td>
<td>Constant Cost Escalation Factor</td>
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<td>CEPCI</td>
<td>Chemical Engineering Plant Cost Index</td>
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<tr>
<td>CHP</td>
<td>Combined Heat and Power</td>
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<td>DH</td>
<td>District Heating</td>
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<tr>
<td>DHN</td>
<td>District Heating Network</td>
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<tr>
<td>EU</td>
<td>European Union</td>
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<tr>
<td>EXE</td>
<td>Exergetic Efficiency</td>
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<tr>
<td>KIT</td>
<td>Karlsruhe Institute of Technology</td>
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<tr>
<td>LHV</td>
<td>Lower Heating Value</td>
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<tr>
<td>PE</td>
<td>Primary Energy</td>
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<td>PEC</td>
<td>Purchased Equipment Cost</td>
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<td>PEF</td>
<td>Primary Energy Factor</td>
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<tr>
<td>PS</td>
<td>Pyrolysis Slurry</td>
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<tr>
<td>SNG</td>
<td>Synthetic Natural Gas</td>
</tr>
<tr>
<td>TCI</td>
<td>Total Capital Investment</td>
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<tr>
<td>TE</td>
<td>Thermal Efficiency</td>
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<tr>
<td>TWP</td>
<td>Torrefied Wood Pellets</td>
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<tr>
<td>WP</td>
<td>Wood Pellets</td>
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Symbols

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<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
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<tbody>
<tr>
<td>C</td>
<td>Cost [€]</td>
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<tr>
<td>c</td>
<td>Cost per unit of exergy [€/kWh]</td>
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<tr>
<td>e</td>
<td>Exergy rate [kW]</td>
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<tr>
<td>e</td>
<td>Specific exergy [kJ/kg]</td>
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<tr>
<td>f</td>
<td>Scaling factor [-]</td>
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<td>h</td>
<td>Specific enthalpy [kJ/kg]</td>
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<tr>
<td>i</td>
<td>Interest rate [-]</td>
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<tr>
<td>k</td>
<td>Variable used in CELF calculation [-]</td>
</tr>
<tr>
<td>n</td>
<td>Number of years [-]</td>
</tr>
<tr>
<td>P</td>
<td>Annual power production [kJ]</td>
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<tr>
<td>Q</td>
<td>Annual heat production or demand [kJ]</td>
</tr>
<tr>
<td>r</td>
<td>Inflation rate [-]</td>
</tr>
<tr>
<td>R</td>
<td>Specific gas constant [kJ/kg·K]</td>
</tr>
<tr>
<td>S</td>
<td>Characteristic dimension [case-dependent, in here MW]</td>
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<tr>
<td>s</td>
<td>Specific entropy [kJ/kg·K]</td>
</tr>
<tr>
<td>T</td>
<td>Temperature [K]</td>
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<tr>
<td>W</td>
<td>Work rate [kW]</td>
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<tr>
<td>x</td>
<td>Molar fraction [-]</td>
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<tr>
<td>ź</td>
<td>Z-factor, cost of unit operation [€/h]</td>
</tr>
<tr>
<td>α</td>
<td>Power-to-heat ratio [-]</td>
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<tr>
<td>η</td>
<td>Thermal efficiency [-]</td>
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### Subscripts

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<tr>
<th>Subscript</th>
<th>Description</th>
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<td>BM</td>
<td>Biomass</td>
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<tr>
<td>BP</td>
<td>Bio-product</td>
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<td>ch</td>
<td>Chemical</td>
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<td>District Heating</td>
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<td>DHN</td>
<td>District heating network</td>
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<td>FO</td>
<td>Fuel oil</td>
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<td>i</td>
<td>running index</td>
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<td>k</td>
<td>Process unit</td>
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<td>P</td>
<td>Power</td>
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<td>Total</td>
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<td>w</td>
<td>Work</td>
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References


65


References


[95] Phyllis, database for biomass and waste, Energy research Centre of the Netherlands [selected subgroups: untreated wood --> birch and fir/pine/spruce]; 2009.


Biomass is an important energy source and can provide sustainable materials, fuels and chemicals. CHP plants are expected to play a key role in further increasing the use of renewables in the Finnish energy system. This thesis provides the thorough investigation of three biomass upgrading processes integrated with a municipal CHP plant considering energetic, environmental and economic performance. The thesis also answers the question of how the energetic and environmental performance of future polygeneration plants can be assessed in an objective way. It was shown that all three processes can be integrated with CHP plants and that all improve the energetic and environmental performance. Under the assumptions made in this work, the biomass fast pyrolysis process shows the highest economic potential. The thesis also concludes that for system level analysis primary energy should be the preferred tool for process assessment because it allows fair process comparison by including the global impact of the applied technologies in a simple manner.