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TAMPERE UNIVERSITY OF TECHNOLOGY

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DEVELOPING A TOOL FOR TECHNO-ECONOMIC ANALYSIS OF PULP MILL INTEGRATED BIOREFINERIES

Master's thesis

Examiner: Professor Risto Raiko
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ABSTRACT

JUHO-ANTTI KASURINEN: Developing a Tool for Techno-Economic Analysis of Pulp Mill Integrated Biorefineries
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A kraft pulp mill forms an attractive platform for integrated biorefining due to the availability of biomass residues and access to low cost process heat. Integrating a biorefinery to a pulp mill aims to improve the overall efficiency of raw material utilization and to offer new revenue opportunities besides pulp production. Because the field of pulp mill integrated biorefining is still relatively unexplored, it is necessary to develop methods for assessing the feasibility of alternative technologies.

The purpose of this thesis project was to design a techno-economic analysis tool (TEA-tool) for Valmet's offering of pulp mill integrated biorefineries. The tool was intended to evaluate the feasibility of four different biorefinery processes from the customer point of view. The general motivation for building the tool was to improve the accessibility of techno-economical methods for users with different backgrounds and to provide an unbiased profitability model for cross-technology comparisons. The biotechnologies included in the tool were lignin extraction from black liquor by LignoBoost, black pellet production by steam explosion, bark gasification and bio oil production by integrated fast pyrolysis. The thesis project consisted of building the tool and performing a rough feasibility comparison between the included technologies.

The priority task of developing the TEA-tool succeeded well, receiving a positive overall reception. The tool allowed quick and effortless comparison between the technologies in a wide range of investment scenarios. The new TEA-tool will offer a flexible platform for Valmet's future techno-economic evaluations.

In general, the analysis boosted confidence on the economic potential of biorefining. The profitability model was discovered being the most sensitive to production capacities, end product values and substitute fuel prices. The selection of process parameters and feedstock properties had significantly lower impact on the profitability estimates.

From the four biotechnology alternatives, LignoBoost and gasification processes were observed being the most profitable investments. The steam explosion process was shown to be competent with these technologies, but would require large production capacities to reach the same level of returns. The integrated pyrolysis process was shown to be theoretically highly profitable in favourable operating conditions. However, the incompleteness of bio oil markets slightly lowers the attractiveness of the particular pyrolysis process. Distributing the produced bio oil to multiple mill sites would reduce the dependency on external markets.

TIIVISTELMÄ

JUHO-ANTTI KASURINEN: Teknistaloudellisen kannattavuuslaskentatyökalun suunnittelu sellutehtaaseen integroitaville bioteknologioille
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Sulfaattisellutehdas tarjoaa houkuttelevan ympäristön integroiduille bioteknologiaratkaisuille saatavilla olevan bioraaka-aineen ja prosessilämmön ansiosta. Bioteknologioiden integroinnin tavoitteena on tehtalle tuotavan puubiomassan tehokkaampi hyödyntäminen ja uusien tulovirtojen luominen selluntuotannon lisäksi. Koska biokonversioprosessien hyödyntämisestä sellutehdasympäristössä ei ole toistaiseksi kattavaa kokemusperäistä tietoa, on teknologioiden kannattavuutta arvioivien menetelmien kehitys tarpeellista.

Tämän diplomityön tarkoituksena oli laatia teknistaloudellinen kannattavuusvertailutyökalu (TEA-tool) Valmetin bioteknologiatarjoomalle. Työkalun perusajatuksena oli mahdollistaa neljän erityyppisen bioteknologian vertailu sellutehdasasiakkaan näkökulmasta. Työkalun rakentamisen päämääränä oli madaltaa kynnystä teknistaloudellisten analyysien suorittamiseen käyttäjän taustoista riippumatta, sekä tarjota objektiivinen laskentamalli eri teknologioiden välisiin vertailuihin. Työkaluun sisällytettäviä teknologioita olivat ligniinin erotus mustalipeästä LignoBoost –prosessilla, höyryräjäytyspelletin tuotanto steam explosion –menetelmällä, kuoren kaasutus sekä bioöljyn tuotanto integroidulla pyrolyysiprosessilla. Diplomityö koostui työkalun rakentamisesta ja työkalulla tehtävästä, suuntaa antavasta kannattavuusvertailusta teknologioiden välillä.

Työkalun rakentaminen onnistui hyvin ja se sai yrityksessä positiivisen vastaanoton. Työkalu mahdollisti nopeiden ja vaivattomien kannattavuusvertailujen tekemisen monipuolisille investointiskenaarioille. Työkalu tulee tarjoamaan joustavan laskentaympäristön Valmetin tuleville teknistaloudellisille tarkasteluille.

Yleisesti ottaen bioteknologiat osoittautuivat taloudellisessa mielessä lupaaviksi. Herkkyystarkastelussa havaittiin, että investointien kannattavuus riippuu pääasiassa tuotantokapasiteetista, lopputuotteen arvosta ja korvaavan polttoaineen hinnasta. Prosessiparametrien ja raaka-ainesyötteiden ominaisuuksien merkitys kannattavuuden kannalta todettiin vähäiseksi.

Tarkastelluista prosesseista LignoBoost ja kaasutus olivat selvästi kannattavimmat. Höyryräjäytysprosessin todettiin olevan kilpailukykyinen näiden teknologioiden kanssa suurilla tuotantokapasiteeteilla. Integroitu pyrolyysiprosessi näytti tuottavan vertailuparametreilla korkeaa tuottoa, mutta bioöljyn markkinoiden kehittymättömyyden todettiin vähentävän investoinnin houkuttelevuutta. Investoinnin riippuvuutta ulkoisista markkinoista voitaisiin pienentää tuottamalla bioöljyä keskitetysti usean tehtaan meesauuneille.

PREFACE

This thesis was made as a partnership project in Valmet Technologies Oy during the second half of 2015. The project was funded by a grant from The Industrial Research Fund of Tampere University of Technology. I want to thank Professor Risto Raiko for supervising and reviewing the thesis.

Working with the TEA-tool project has been full of joy. The project has taught me a lot from myself and from the topic in question. The interesting topic and tremendous coworkers at the Valmet office made the time pass incredibly fast.

Thanks to Erkki Välimäki for providing me a very engaging research subject and giving lots of encouraging feedback along the project. I really appreciate the creative freedom entrusted to me with the TEA-tool design.

I also want to sincerely thank every single individual who contributed to the work at Valmet by giving feedback about the tool development. Without your invaluable advice the project would not have succeeded so well.

Finally, thanks to my family for all the support and caring. Special thanks to my girlfriend who made the studying years the best time of my life.

Tampere, 18.12.2015

Juho-Antti Kasurinen

CONTENTS

LIST OF SYMBOLS AND ABBREVIATIONS

ADt		Air dry ton
BFB		Bubbling fluidized bed
CAPM		Capital asset pricing model
CFB		Circulating fluidized bed
CHP		Combined heat and power
et al.		et alii (and others)
i.e.		id est (that is)
IRR		Internal rate of return
MIRR		Modified internal rate of return
MS		Microsoft
NCG		Non-condensable gas
NPV		Net present value
O&M		Operations and maintenance
PI		Profitability index
tDS		Dry solids ton
TEA		Techno-economic analysis
VTT		Valtion teknillinen tutkimuskeskus (Technical Research Centre of Finland)
WACC		Weighted average cost of capital
CaO		Calcium oxide
$CaCO_3$		Calcium carbonate
CO_2		Carbon dioxide
H_2SO_4		Sulfuric acid
$NaOH$		Sodium hydroxide
Na_2S		Sodium sulfide
Na_2SO_4		Sodium sulfate
Na_2CO_3		Sodium carbonate
CH_4		Methane
CO		Carbon monoxide
η_E	[%]	Average growth rate of the company's equity
η_s	[%]	Turbine isentropic efficiency
η_g	[%]	Generator efficiency
C	[€]	Cash flow (in general)
C_{depr}	[€]	Annual depreciation
C_{LK}	[€/h]	Lime kiln fuel savings
$C_{net,current}$	[€/a]	Annual net cash flow of the investment
$C_{sales,current}$	[€/a]	Annual revenue from product sales
C_{sold}	[€/h]	Cash flow from sold bioproduct (per hour)
D	[€]	Company's debt value
D_n	[-]	Discount factor for n:th year of investment's life time
E	[€]	Company's equity value

h_1	[kJ/kg]	Steam enthalpy before expansion
h_2	[kJ/kg]	Steam enthalpy after expansion (actual)
h_{2s}	[kJ/kg]	Steam enthalpy after expansion (ideal)
$H_{bio,wet}$	[MJ/kg]	Lower heating value of wet bioproduct
k	[-]	Sum index
K_{bio}	[%]	Biofuel-induced increase in lime kiln heat demand
$\dot{m}$	[kg/s]	Steam mass flow
$\dot{m}_{ds}$	[t/h]	Feedstock dry solids mass flow (pyrolysis feed)
$\dot{m}_{product}$	[t/h]	Product mass flow (pyrolysis oil)
n	[-]	Sum index
N	[a]	Economic life time
p_1	[bar]	Steam pressure before expansion
p_2	[bar]	Steam pressure after expansion
P	[MW]	Power generated by the generator
P_{bio}	[€/MWh]	Bioproduct market price
$P_{bio,tDS}$	[€/tDS]	Bioproduct market price (alternative unit)
P^*_{bio}	[€/MWh]	Biofuel market threshold price
$P^*_{bio,tDS}$	[€/tDS]	Biofuel market threshold price (alternative unit)
$P_{break-even}$	[€/MWh]	Break-even price
P_{LK}	[€/MWh]	Default lime kiln fuel price
q_{bio}	[MJ/t]	Specific lime kiln heat demand per lime ton for biofuel
$q_{default}$	[MJ/t]	Specific lime kiln heat demand per lime ton for default fuel
$q_{LK,total}$	[MJ/t]	Actual lime kiln heat demand per lime ton
Q	[MW]	Power transferred from steam
Q_{bio}	[MW]	Product flow of a bioproduct
Q_{LK}	[MW]	Lime kiln heat load with default fuel
Q_{sales}	[MW]	Product flow of a bioproduct to markets
r	[%]	Real discount rate
r_D	[%]	Company's debt rate
r_i	[%]	Inflation rate
r_n	[%]	Nominal discount rate
r_{tax}	[%]	Company's tax rate
t	[a]	Full years with negative cumulative discounted cash flow
T_{oper}	[h/a]	Annual operating hours
T_p	[a]	Discounted pay-back period
$y_{organics}$	[%]	Pyrolysis organics yield of dry feedstock
y_{water}	[%]	Pyrolysis water yield of dry feedstock
x_{feed}	[%]	Feedstock moisture content (pyrolysis feed)
x_{water}	[%]	Water content (in general)
X_{bio}	[%]	Bioproduct fraction of lime kiln heat load

1. INTRODUCTION

The purpose of this thesis work is to develop a tool for techno-economic feasibility analysis of Valmet's new biorefinery technologies. The analysis is performed from the standpoint of a pulp mill customer and is framed to four alternative technologies. The biotechnologies included in the tool (later referred as the TEA-tool) will be lignin separation by LignoBoost, black pellet production by steam explosion, bark gasification and integrated pyrolysis. The thesis project benefits Valmet by bringing understanding to the company's own offering portfolio, which in turn eases the sales department's task of delivering the biorefinery concepts to customers. The tool will also work as a flexible platform for preliminary feasibility evaluations of new biotechnologies.

The scope of the project includes building the tool, documenting the tool and testing the finished tool. The testing phase will focus on comparing the feasibility of a number of pre-selected investment scenarios in a case pulp mill. The written part of the thesis work concentrates on introducing the examined biorefinery processes, explaining the tool design and presenting a summary of the results obtained from the scenario analysis. The TEA-tool will not be available for public distribution due to confidentiality issues. However, the thesis is structured to be readable as its own entity without having access to the actual tool.

This chapter introduces the concept of biorefining and defines the frame of reference for the thesis. The thesis frame of reference is explained through presenting the project goals and methods used in the project.

1.1 The concept of biorefining

During the past decades, industrial businesses have been facing a growing pressure to migrate from non-renewable fuels and raw materials to more sustainable solutions. This results from increased awareness of environmental issues raised by fossil fuel use and political regulation related to these issues. One way of adapting to this trend is to replace non-renewable raw materials with bio-based products. Biorefining is a concept of converting organic biomass into higher valued bioproducts that can be used as substitutes for conventional raw materials. These bioproducts span a range of combustible fuels, food products and intermediate products for further refining.

A conventional pulp mill forms an attractive platform for biorefinery implementation as the process generates excess heat and disposable side product streams that can potentially be utilized in a more efficient way. The aim with pulp mill integrated biorefining is to add extra value to the mill production portfolio and to offer alternative uses for residue biomasses. In addition to profitability, biorefining generally increases the overall efficiency of pulping wood utilization from roundwood to end products. The traditional way of combusting the residue biomass in a boiler generally wastes refining potential of the biomass. The future markets for pulp and fossil fuels also involve uncertainty, which incites further motivation for expanding the production portfolio of existing pulp mills.

Some of the refined bioproducts can be utilized locally at the mill, thus reducing the plant operating costs and improving the self-sufficiency of the mill. Some products can alternatively be sold to markets. Although the markets for some of the products are still incomplete, growing demand for environmentally friendly fuels and raw materials may open new market opportunities for bioproducts in the near future. Defining the acceptable price levels and other market requirements is essential for the initial decision of developing and distributing a specific biotechnology.

1.2 Project goals

Before starting the TEA-tool project, desired effect goals that the project was expected to fulfill were evaluated. The main function of the tool was defined to be the ability to compare the feasibility of different biorefining processes and their effects to a kraft pulp mill. The tool aims to give a general idea of attractiveness of the biorefining possibilities from the customer perspective. Another function of the tool is to find the decisive factors affecting the investment profitability. For actual investment decisions and detailed process engineering, more accurate models would be needed.

Although this kind of tools have already been implemented, the lacking documentation and varying level of complexity set unnecessary challenges for the tool users. The new tool should enable the comparison between the investments in a balanced and understandable manner, while maintaining the level of detail that is required to obtain reliable results.

A complete overview of the project goals is shown in figure 1.1. The main effect goal was phrased to be the development of a simple tool for pulp mill customer benefit visualization. Other recognized sub-goals consisted of user experience related features such as a simple user interface and accessibility to users with varying backgrounds of technical and financial knowledge. In addition to the listed goals, a general hope was that the tool would later be expandable to other technologies as well.

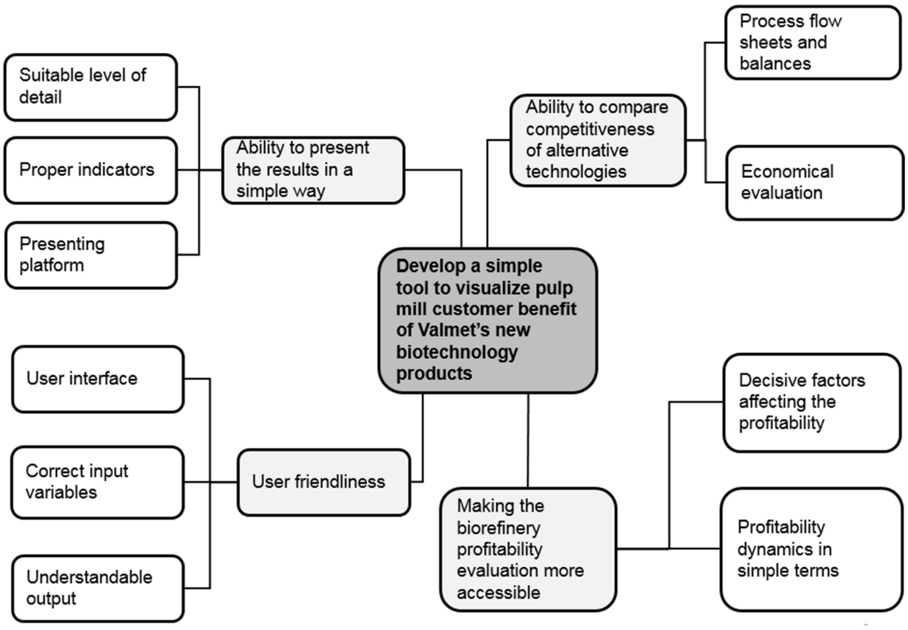


Figure 1.1 *A mind-map of TEA-tool project goals.*

Ultimately, the successfulness of TEA-tool implementation is evaluated by the degree that these goals are achieved. For a tool serving such wide user segment, compromises are unavoidable. The challenge of the project is to find a way of representing the results in a way that satisfies all the process departments and is relevant for customers.

1.3 Methods

The mass and energy balance model of the tool will be based on a similar tool for the LignoBoost process built by Technical Research Centre of Finland (VTT) in cooperation with Valmet. The old tool has been constructed as a spreadsheet balance model on Microsoft (MS) Excel. Based on this spreadsheet version, a separate graphical interface has been built on Adobe Flash platform. The Flash implementation provides visual and user friendly interface for the tool user. The new TEA-tool is intended to expand the analysis to cover a wider selection of biorefineries.

Because of the all-in-one-approach, the new TEA-tool model will be re-built from scratch. The LignoBoost tool will be used as a reference for the pulp mill and LignoBoost process balances. The old balance models have to be slightly modified in order to allow connections to the new calculation modules and to improve customizability. Essentially this means decoupling the existing balances into a modular structure and building a more flexible tool logic. Simultaneously, the calculations are expanded to include additional modules for the other three reviewed biotechnologies. The process balances of the new biotechnologies are constructed using similar level of detail as in the LignoBoost model.

The process flow model of the tool will be created using a so called black box method to reduce the model complexity. This means that the individual process parts are modeled as simple input/output components with known correlations. The correlations will be constructed according to predicted and known performance data of the biotechnology processes.

The TEA-tool model will be built on a set of calculation spreadsheets using Microsoft Excel. The spreadsheets will include all the input data and mill balance calculations required in the economical evaluation. The spreadsheet version will also include its own user interface with complete tool functionality.

The finished tool will be used to analyze the sensitivity of the profitability against various input variable changes and to find the boundaries in which the individual biorefinery investments would be profitable. In addition to this analysis, the core principles of the tool functioning logic will be presented along this thesis.

Regarding the scenario analysis carried out in this study, the TEA-tool features and the presented scenario assumptions have to be separated from each other. The TEA-tool will allow modifying of process parameters and market conditions after user preference. The scenario analysis is made by utilizing the tool but it only scratches the surface of what the tool has to offer. For this reason, the tool design and user input will be presented separately from the scenario analysis.

The source material of the thesis spans a range of internal documents of Valmet's technologies and finances. In addition to this, some of the missing data needed for the tool is gathered from interviews with key personnel. The theoretical background will be mainly based on public references.

2. TECHNICAL BACKGROUND

To understand the biorefinery processes and the significance of correct input data needed in the tool, some of the technical principles related to the pulping process have to be explained. The next sub-chapters introduce the kraft process as the operating environment for the biorefinery integrates. Additionally, the properties of pulping wood and biomass dryer performance have been identified as concepts of high importance to the biorefinery process analyses.

After familiarizing the reader with the kraft pulping process, the effects of the mill operating environment are briefly assessed. Most of the fluctuations in operating conditions can be related to the geographical location of the mill. These factors have to be taken into account to understand the need for customization possibilities of the TEA-tool model.

2.1 Kraft process

The kraft process, also known as the sulfate process, is a chemical pulping process where cellulose rich pulp is produced out of wood by separating lignin from the wood biomass. A simplified flow chart of the kraft process is shown in figure 2.1. The wood arriving at the plant is utilized with high efficiency by turning it into pulp, steam, electricity and other by-products. Roughly half of the wood biomass is refined into exportable products – mainly bleached pulp. The other half of the wood is combusted locally at the mill in a form of cooking liquor and process reject streams.

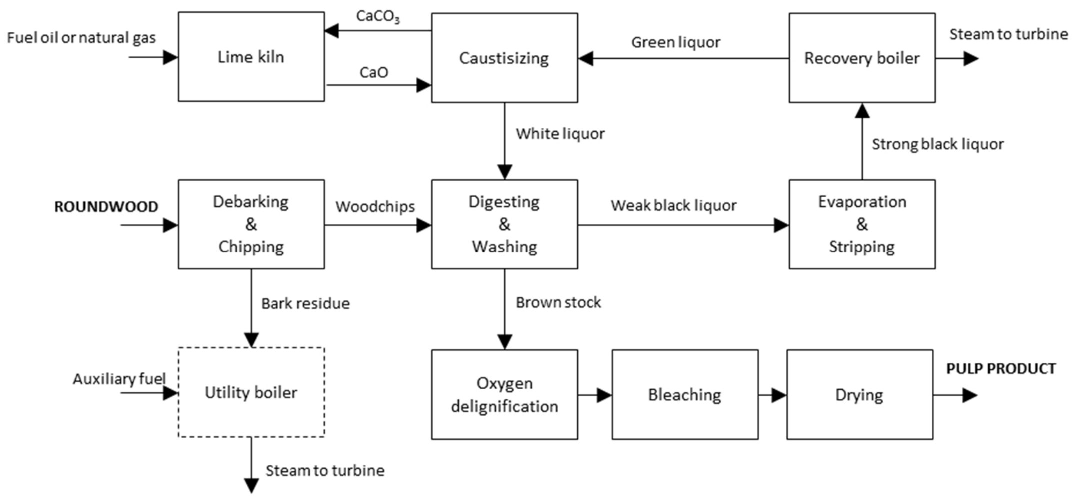


Figure 2.1 *The flow chart of a kraft pulp mill emphasizing process inputs and outputs. Process chemicals and other consumables such as steam and electricity have not been drawn into the chart.*

The primary lignin separation is performed chemically in the digesting step. The produced pulp is then washed, bleached and dried. The cooking chemicals are recovered in a recovery process where the black liquor produced in cooking is turned back into white liquor. The energy produced by combustion of black liquor in a recovery boiler is sufficient enough to fully cover the mill heat and electricity demand. To take advantage of the available process steam, a paper mill is often integrated to the pulping process. Excess electricity can be sold outside of the plant.

2.1.1 Pulping

The pulping wood may arrive at the mill site either as roundwood, debarked roundwood or pre-chipped woodchips. Whether the wood is pre-treated in-site or off-site, it has to be debarked and chipped before the pulping process. The wood reject generated in in-site woodhandling has to be utilized locally or disposed of. Most of this biomass reject consists of bark. A common way of disposing the residue biomass is incinerating it in a utility boiler to produce additional process steam and electricity. The biomass residue has a great potential to be used in various biorefinery processes in a more efficient way. The biofuels could then be used locally at the plant or sold to markets.

Wood biomass consists mainly of celluloses, hemicelluloses and lignin [1, p. 2]. Lignin is the component that holds the cellulose fibres together and is an unwanted component in chemical pulp. In the digesting step (also called the cooking step) the wood chips are cooked in a white liquor suspension under high temperature and pressure in order to dissolve the lignin from the wood biomass.

The main cooking chemicals in white liquor are sodium hydroxide (NaOH) and sodium sulfide (Na_2S). The digested wood biomass or pulp is washed and fed to the bleaching line. The residue suspension consists mainly of cooking chemicals (NaOH and Na_2S), sodium carbonate (Na_2CO_3), sodium sulfate (Na_2SO_4), lignin and water. This so called weak black liquor (solids content approximately 15 %) is directed to the recovery cycle, where the process chemicals are recovered to produce white liquor for cooking. [1, pp. 153-155] [2, pp. 522-543]

After digesting, pulp still has a high lignin and impurity content, resulting in a coloured tint in the pulp. For this reason, bleaching is required to generate a desired pulp product. The bleachability of pulp can be expressed with the Kappa number. The Kappa number is defined as the amount of needed potassium permanganate solution consumed in pulp bleaching [3]. In practice, the Kappa number is closely related to the residue lignin content in pulp [1, pp. 73-74]. While there is a variety of different techniques for bleaching, a usual way is the removal of excess lignin by oxygen delignification. Bleached pulp is then either sold to markets or used in an integrated paper mill to produce various paper-based products.

A common way of expressing the process flows of a kraft pulp mill is to relate them with the amount of produced pulp. Usually the amount of pulp is expressed as air dry ton (*ADt*). Air dry ton of pulp is defined as a pulp product with a dry solids content of 90 wt-%.

2.1.2 Recovery cycle

The weak black liquor from cooking is driven through evaporators, where excess water is evaporated from the black liquor stream. The purpose of this step is to prepare the black liquor to be combusted in the recovery boiler. The heating energy needed in the evaporation step is taken from the steam turbine low pressure section. The product from evaporation is called strong black liquor and has a solids content of 60-85 %. [2, pp. 522-543]

The strong black liquor is combusted in the recovery boiler to generate electricity and process steam in conjunction with a steam turbine and a generator. After being injected to the recovery boiler the black liquor droplets burn in a multi-stage process involving drying, pyrolysis, gasification and char burning [2, pp. 535-539]. The inorganic compounds formed by the cooking chemicals create a smelt flow from the bottom of the boiler. The smelt is dissolved in water to form green liquor, which is cycled through causticizing process to produce white liquor for cooking. [1, pp. 159-163]

In the causticizing cycle, green liquor consisting mainly of sodium sulfide (Na_2S) and sodium carbonate (Na_2CO_3) is causticized using lime (CaO). The reaction products from causticizing are sodium hydroxide (NaOH) and calcium carbonate ($CaCO_3$). Sodium sulfide does not react in the causticizing process. Calcium carbonate precipitate or lime mud is sent to the lime kiln, where the CaO needed in causticizing is recovered by calcination. The sodium hydroxide and sodium sulfide form white liquor. [1, pp. 165-170] [2, pp. 522-543]

The heating and calcination energy needed in causticizing is commonly provided by either natural gas or heavy fuel oil. Other possible lime kiln fuels are tall oil, tall oil pitch, sawdust and coal gas [4]. Due to increased environmental awareness, the regulation and taxation around these fuel types is expected to change in the near future. This creates an incentive to find alternative fuels for the lime kiln. The heating power of a lime kiln varies between 10-100 MW depending on the pulp mill scale.

Some of the bioproducts reviewed in this study can be used as a partial replacement fuel in the lime kiln, thus providing savings in fuel purchases. The use of these biofuels, however, require certain properties and burner modifications. The total heat rate requirement in the lime kiln is also dependent on the used fuel [5] [4] [6]. A summary of heat demands for different lime kiln fuels is presented in table 2.1. The data is estimated from empirical data of lime kiln performance. Actual heat rate requirements vary depending on the lime kiln parameters.

Table 2.1 *Fuel-specific lime kiln heat rate (MJ/t lime) requirements. [5] [7]*

Fuel oil & Bio oil	Natural gas	Wood fuel & Lignin	Biogas
6000 - 6300	6300 - 6600	6300 - 6600	6600 - 7200

In co-firing cases the gross heat demand can be roughly estimated as the weighted average of the specific heat demands [4]. Because the amount of lime stays constant, this assumption leads to the conclusion that the relative amount of auxiliary fuel per lime ton changes when a fraction of heat input is replaced with another fuel type. Furthermore, the amount of fuel savings are not necessary equal to the amount of used replacement fuel. Methods for calculating the change in gross heat demand and profit from fuel savings will be further addressed in chapter 3.3.

2.2 Softwood and hardwood

The contents and the microstructure of wood biomass vary between different wood types. The three main elements common for all wood biomasses are cellulose, hemicellulose and lignin. Cellulose is formed of polymerized carbohydrates that form the primary fibre structure of wood. Hemicellulose resembles cellulose but consists of smaller polysaccharide molecules. Cellulose and hemicellulose are the main ingredients in pulp production. The cellulose-hemicellulose balance and the chemical composition of these molecules define the properties of the produced pulp. The cellulose and hemicellulose fibres are bound together by lignin, which is a complex mix of polymerized organic compounds. The main principle of pulping is to separate the cellulose fibres from lignin that is generally an unwanted component in the final pulp product. In addition to the three main components, dry wood also contains ash and various extractives, most of which can be recovered in pulping. [1, pp. 1-7]

The wood types used in chemical pulping can be categorized into softwood and hardwood. Softwood is characterized by long fibres and high lignin content. Hardwood has a significantly higher cellulose content and the fibres are shorter than in softwood. In Europe, pine and spruce are commonly used in softwood kraft pulping because of their availability. The most common hardwood species suitable for pulping are birch, aspen, eucalyptus and oak. [8]

Due to the strength offered by long fibres, softwood pulp is usually used in quality packing materials whereas hardwood pulp suits better for printing papers and tissue papers. The differences between softwood and hardwood properties, however, vary by case and the pulp types can be used in various mixtures to achieve a desired product. [1, pp. 15-17] [8]

Although the exact composition of wood cannot be accurately estimated even within the same wood species, distinctive properties affecting the pulping process can be identified for each wood type respectively. The gross pulp yield for hardwood species is usually higher than that of softwood pulp [1, p. 15]. The Kappa number is also typically higher for softwood pulp due to the higher overall lignin content in raw wood. A rough comparison between the wood type dependencies on cooking conditions has been compiled into table 2.2. It should be noted, however, that the cellulose-hemicellulose-lignin balance may vary greatly between different species within the same type of wood. [9]

Table 2.2 *Example values of hardwood and softwood properties. [1, pp. 1-2] [9] [10]*

	Cellulose (wt-%)	Hemicellulose (wt-%)	Lignin (wt-%)	Extractives (wt-%)	Pulp yield (wt-%)	Kappa number -
Hardwood	43-47	25-35	16-24	2-8	50-54	16-18
Softwood	40-44	25-29	25-31	1-5	44-48	24-29

Instead of forecasting accurate microcontent-dependent product yields, this study focuses on exploiting known correlations for the bulk wood arriving at the pulp mill. This is a valid approach as the gross yields can be assumed to stay fairly constant within a single wood type. The future case mills are also expected to have the required balance and correlation data available.

2.3 Biomass drying

Water is generally an unwanted component in biorefinery processes and refined bioproducts. The most convenient solution of reducing the bioproduct moisture content is to remove water prior to processing. Pre-drying offers a more economical solution compared to post-drying because the excess water content does not have to be heated in the actual biorefining process.

In the TEA-tool, the biorefinery feedstock is assumed to be dried using a belt-dryer that utilizes heat from in the mill. In a belt dryer, heat is transferred to the biomass by air that is blown through a conveyor belt carrying the biomass. The drying air is heated with a secondary ethylene glycol loop, which in turn is heated with low pressure steam and hot water from the mill. In a warm climate, water can be used instead of ethylene glycol.

The most obvious reason for the use of a belt dryer in pulp mill scenarios is the availability of low cost heat. The operating temperature is also lower than in flue gas dryers and therefore less volatile losses are induced during the drying phase. The drawback with this type of dryer is the size that grows along the biomass flow rate and required evaporation rate.

In practice, the optimum target moisture content of biomass after pre-drying is 8-10 % for all the processes reviewed in this study [4] [11] [12]. This value stays fairly constant regardless of process scaling or other parameters. The dry solids portion of the feedstock is assumed to behave the same way regardless of the moisture content. This is a fair assumption as the practical moisture content stays always within this narrow range. [4] [11] [12]

Because the aim of the TEA-tool is to simplify the process balance calculations, the moisture-dependent drying steam consumption is expressed through a pre-calculated correlation. A reference value of 1.26-2.5 MJ/kg of evaporated water for the belt dryer heat requirement can be obtained from literature [13]. A more valid value for the specific gross heat consumption in average operating conditions would be closer to 3-5 MJ/kg-evaporation [7], including heat losses.

The dryer energy consumption is significantly affected by ambient air temperature. In the tool, the specific heat consumption is expressed as a function of this temperature. The correlation between the specific heat demand and temperature was predicted using known performance data from existing dryers. This correlation can later be changed to correspond to the performance of actual case dryers. Reference boundary curves for the specific heat consumption of Valmet's belt dryers have been presented in figure 2.2. The dashed lines enclose the range, in which the specific heat consumption is expected to fluctuate.

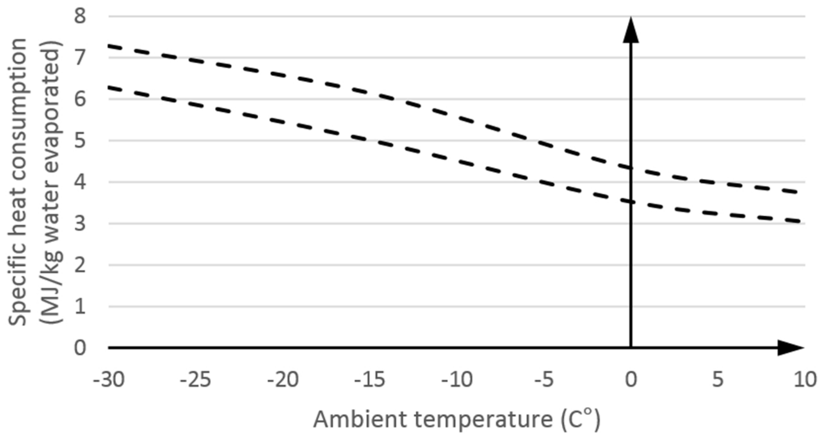


Figure 2.2 *Approximate boundaries for belt dryer specific heat consumption (MJ/kg of water evaporated). The reference data is evaluated using empirical data gathered from Valmet's dryers. [7]*

When the biomass temperature drops below 0 °C, the feedstock moisture starts to appear in the form of ice. The latent heat required to melt the ice (0.334 MJ/kg) can be seen in the above figure as an incremental increase in heating energy demand when the temperature decreases. A cold temperature of the drying air is not a problem as long as surplus hot water is available. Otherwise low pressure steam needs to be used instead.

In addition to the heat demand, the temperature difference between the heating medium and wet biomass has to be taken into account. Although this may sound trivial, the temperature levels have a major impact on what heat sources can effectively be used for biomass drying. A decent temperature difference between the hot water supply and the desired drying air temperature is important, as too small or negative temperature difference neglects the energy transfer potential of the heat exchanger. Insufficient temperature difference results in high steam/water mass flows and increases the dryer size.

The TEA tool calculates the process balance as steady state scenario with a single heat consumption value. Therefore the average effective air temperature has to be estimated over the operating period when calculating the dryer predicted performance. This temperature can vary depending on local climate. By default, the TEA-tool will assume effective temperatures between 2-5 °C that are typical average yearly temperatures for the Finnish climate [14].

Although the specific heat consumption can fluctuate heavily, it is not expected to make a major difference in the ultimate profitability figures. This issue is further discussed in the sensitivity analysis. The steam demand can still become a constraint in mills where the supply of surplus heat is limited.

2.4 Pulp mill operating environment

The pulp mill operating environment may have a major impact on the distinctive process parameters and the cash flow structure of a potential biorefinery investment. The contrast between local infrastructure, political regulation, climate and available feedstocks fluctuate considerably between different geographical locations. In case the biorefinery investment cannot be motivated by environmental benefits, the attractiveness of biotechnology implementation leans purely on increased revenue. Investment support funding may also be more openly available in certain areas. [15]

The wood type used in the process affects the kraft process through the mechanics introduced in chapter 2.2. Especially the yield differences between softwood and hardwood species are noticeable. The wood type also affects other bioproduct yields and their quality. In addition to these factors, the raw wood state on arrival may vary in different mill sites. In addition to the moisture content fluctuations, the wood arriving at the mill may be already debarked upon cutting. Especially eucalyptus may arrive at the mill debarked due to its relatively easy on-site debarking characteristics. [15] [16, p. 21]

The heat and electricity demand of a kraft pulp mill varies greatly between different operating environments. It is dependent on local infrastructure, climate and the possible presence of an integrated paper mill. In geographically remote areas the ability to sell electricity or excess heat to markets may be limited. In such cases, the mill cannot purchase additional electricity nor can it sell the excess electricity to markets. Additionally, selling the excess heat to markets is exclusive to certain areas with infrastructure for district heating.

The market penetration of the bioproducts play a major role in the profitability of biorefinery investments. In addition to globally quoted market prices, the produced biofuel price includes a component that is dependent on local markets. The local market may potentially consist of a network of mill sites. Multi-site distribution of bioproducts lowers the market risks and expands the expected capacity demand, therefore allowing larger biorefinery investments. This approach of increasing the production capacity is especially important for biofuels that can be used as lime kiln fuel, because scaling the biorefineries after the maximum fuel replacement of a single lime kiln may result in plants below economically feasible capacities. For the surplus production, other markets would have to be found.

3. ECONOMIC BACKGROUND

This chapter introduces the principles behind the economic evaluation of the TEA-tool. First, the financial terminology and indicators used in the final feasibility analysis are introduced. After this, the cash flows related to the biorefinery investment are briefly presented.

The cash flow calculations of the economic model follow a simple pattern that compares the annual cash flows of the biorefinery investment to the reference case. The year 0 of the investment timeline represents the present year, in which the fixed investment costs are assumed to occur. From year 1 to the end of the investment economic life, the implemented biorefineries are assumed to generate constant nominal cash flows. For actual cash flow estimates, the nominal cash flows are discounted to the present day using the inflation-corrected interest rate.

The TEA-tool model does not assess risk nor does it value intangible assets such as the benefit of self-sufficiency or company values. This means that the most profitable investment does not always correspond to the most attractive investment. The motivation for the investment may also originate from other reasons than profitability alone.

3.1 Terminology and economic indicators

The purpose of economic indicators is to represent the economic characteristics of an investment and provide tools for profitability comparison between alternative investment options. Because accessibility and simplicity were defined as desired characteristics of the TEA-tool, the output representing the results had to be limited to simple and understandable economic indicators.

In this tool, net present value, internal rate of return, discounted pay-back period and break-even price are used as key indicators. These indicators were chosen according to wishes given by sales and technology representatives. Additionally, the modified internal rate of return is introduced as a supporting indicator.

None of the presented economic indicators is adequate enough to be used independently to conclude an investment decision. The investment profitability analysis should be conducted using all of the given indicators in conjunction and the results should be interpreted with proper deliberation.

A key idea in feasibility analysis is to understand that a profitable investment is not always feasible. This follows from the fact that the amount of investment funds is limited and therefore the available money should be invested so that it makes the best profit. On the other hand, it is also important to note that the companies behind the investments have different investment strategies and value profit over risk differently.

3.1.1 Discount rate and weighted average cost of capital

The discount rate is a measure for the time value of money. The amount of capital at the present time is more valuable than the nominally equal amount of capital in the future. This results from the assumption that the capital acquired at the present day could be invested to a growing asset until the time of the hypothetical future cash flow. [17, p. 130]

The time value of money indicates that in arbitrage-free markets (no free lunches), nominally equal cash flows occurring at different times cannot be of equal value. The discount rate is used to express the difference between the real values of these cash flows. The formula of expressing the future cash flow in discounted present value will be introduced in chapter 3.1.2.

Concerning this study, it is essential to determine appropriate discount rates for the investments in order to ensure acceptable reliability of the results. The discount rate is a combination of different rate components. The most important components are the market interest rate, the expected rate of return and the risk premium. The market interest rate indicates the cost of raising capital and is relatively easy to define. The expected rate of return reflects the investors' requirements for the returns on their invested capital. The expected rate of return varies between different industries, companies and investments. The risk premium is the price of risk carried by the investment. In other words, a risky asset has higher requirements for returns than a riskless asset.

In addition to these three components, the real discount rate is affected by inflation. The inflation-corrected *real discount rate* r can be expressed as:

$$r = \frac{(1+r_n)}{1+r_i} - 1, \tag{3.1}$$

where r_n denotes the nominal discount rate and r_i the inflation rate.

Because each of the rate components carries uncertainty, defining a unique value for the discount rate is a challenging task. In reality, the discount rate also changes over time. One way to overcome this is to define a fixed discount rate for an investment and to perform a sensitivity analysis concerning the rate fluctuations.

The expected rate of return and the risk premium are difficult to define for investment opportunities that share little to no resemblances to the company's core business. However, if the nature of the investment is close to the company's core competencies, the discount rate can be estimated from the weighted average cost of capital (WACC). The weighted average cost of capital reflects the average discount rate of the company's overall business and can be calculated from formula [17, p. 310]:

$$WACC = \eta_E \frac{E}{E+D} + r_D \frac{D}{E+D} (1 - r_{tax}) , \tag{3.2}$$

where η_E denotes the average growth rate of the company's equity, r_D the debt rate, r_{tax} the corporate tax rate, E the company's equity value and D the company's debt value. The first term of the formula represents the growth requirement from investors' demands and the second term expresses the growth that is required to cover the costs of debt. The weighted average cost of capital should not be blindly used as a discount rate for new investments. It can still be used as a reference value if more detailed estimates of realistic discount rates were unavailable. Depending on the risk premium, the realistic rate could range from 5 to 15 % for the studied biorefinery investments.

3.1.2 Net present value and profitability index

Net present value (NPV) indicates the total value of an investment in units of currency by calculating the sum of expected future cash flows. The time value of money is taken into account by discounting the cash flows by discount rate. The discount rate can be divided into nominal interest rate and inflation rate. In other words, the net present value equals the cumulative discounted cash flow value at the investment life time. The NPV of an investment can be calculated with the following formula [17, p. 339]:

$$NPV = \sum_{n=0}^{N} \frac{C_n}{(1+r)^n} , \tag{3.3}$$

where N is the economic life time of the investment in years, C_n is the total expected cash flow during year n and r is the discount rate. The cash flow discount factor can be expressed in a shorter form as $D_n = 1/(1 + r)^n$. The accuracy of the NPV-formula and its derivative measures are dependent on the accuracy of the future cash flow estimates.

When inflation is taken into account, the net present value of future cash flows can be calculated with one of two equivalent methods:

1. Inflate the nominal cash flows with the inflation rate and discount these inflated cash flows using the plain discount rate.
2. Discount the nominal, non-inflated cash flows directly with the inflation-corrected real rate.

Both of these approaches return the same end result and the choice between these options is a matter of preference. The only difference is that the intermediate state of the nominal cash flows is different. The TEA-tool will use the method number 2.

When the net present value indication is provided, it is often accompanied by the profitability index (PI), which is calculated by dividing the total value of future cash flows (investment + NPV) by the initial investment cost. The profitability index is a relatively self-explanatory indicator and its purpose is to compare the expected relative returns for the bound capital.

With long investment periods it should be taken into account that the discount rate may not (and most likely will not) remain constant during the life of the investment. Another problem with NPV is that against common way of thinking, net cash flows may be valued differently even though they were nominally equal. This is an implication of the capital asset pricing model (CAPM) [17, pp. 240-249] and its assumption on risk-aversive investors. A good example of this is a cash flow with a net present value of zero. A non-existent cash flow has no volatility. This does not necessary apply to all cash flows with a net present value of zero. Considering risk and transaction costs, it is more beneficial to have no cash flows than two equally sized cash flows that negate each other.

The advantage of NPV is that unlike the other indicators, it gives an estimate of the investment's absolute value. It is possible that when comparing alternative investment options, all the relative measures show up against an investment even though the absolute profit would be higher than those of other investments. In the long run, NPV is the most useful single indicator when used without the support of the other indicators.

3.1.3 Internal rate of return

The internal rate of return (IRR) is a relative measure that can be used to compare different investment options with each other. Internal rate of return indicates the implied rate at which the investment is expected to make profit. IRR is calculated by setting the NPV (formula 3.3) equal to 0 and solving the discount rate r from the equation. Because of the high degree polynomials, IRR usually has to be calculated iteratively. [17, p. 341]

A general thumb rule is that the funds of a company should be targeted to investments with the highest IRR. The internal rate of return can also be considered being the threshold of the discount rate under which the investment starts to make negative profit. In other words, if the cost of capital is higher than the IRR of an investment, the funds could be invested into some other asset with a higher payoff.

As a derivative indicator, IRR shares some problems with the underlying NPV formula. Like in NPV, the uncertainty of the implied rate depends on the estimated cash flow accuracy. Additionally the cash flow valuation is made with the assumption that the funds are constantly being reinvested with the same internal rate [17, p. 346].

A more realistic estimate for actual investment returns would be the modified internal rate of return (MIRR) that also takes the reinvestment rate into account. The MIRR is calculated by compounding the positive cash flows occurring during the investment life time with a separate rate, which can typically be assumed to be near or equal to the WACC (equation 3.2) of the company [17, pp. 346-349].

Regardless of its problems, IRR is a useful indicator of making rough comparisons of different investment options. Because numerical values of internal rates are not published in this thesis, simultaneous comparison of IRR and MIRR between the investments would be redundant. Both of these internal rates will still be included in the TEA-tool output.

3.1.4 Discounted pay-back period

Pay-back period indicates the time under which an investment pays back itself. Discounted pay-back period takes the time value of money into account by using discounted cash flows instead of nominal cash flows. The discounted pay-back period can be calculated with the following formula [17, p. 353]:

$$T_p = t + \frac{|NPV_t|}{D_{t+1}C_{t+1}}, \tag{3.4}$$

where *t* represents the full years with negative cumulative discounted cash flow (a), NPV_t is the net present value (€) calculated up to time t, D_{t+1} is the discount factor of period t+1 and C_{t+1} is the net cash flow (€/a) of period *t+1*. The formula can be written in complete form as

$$T_p = t + \frac{\left|\sum_{k=0}^{t} \frac{C_k}{(1+r)^k}\right|}{\frac{C_{t+1}}{(1+r)^{t+1}}}, \tag{3.5}$$

where *r* is the real discount rate.

The discounted pay-back period only indicates the pay-back time. It does not tell if the investment makes profit after that period nor does it provide any information on the profit margins. This implies that a short pay-back time does not necessarily correlate to high overall profits. Despite of this, the pay-back period is usually the most tangible indicator because of its self-explanatory nature.

The pay-back period can be used as a profitability indicator independent from the IRR and NPV, if the required rate of return is included in the discount rate. The benefit of payback period is that it is not affected by the investment life time unlike the net present value and internal rate of return. This is an important factor especially when the expected investment life time is uncertain.

3.1.5 Break-even price

Presenting fixed numerical values for IRR and NPV is problematic due to the sensitivity to bioproduct prices that cannot be reliably defined in undeveloped markets. A more convenient way of indicating feasibility is to calculate the break-even prices for the exportable end products. The break-even price is the product price, at which the investment pays itself back (breaks even) at maturity. In other words, the break-even price is the price, at which the bioproduct has to be sold in order to cover the annual running costs and investment depreciation. This is a simple way of expressing the absolute minimum acceptable value for the end product.

The break-even price of a sold product can either be solved iteratively or calculated analytically. The iterative method should be preferred when the cash flow structure of the investment is complex. The analytical formula can also be constructed without knowing the exact cash flow composition, embedding the individual cash flows implicitly to the formula. Taking the available TEA-tool output into account, an indirect analytical formula for the break-even price (€/MWh) can be expressed as

$$P_{break-even} = \frac{C_{sales,current} - C_{net,current}}{T_{oper}.Q_{sales}}, \tag{3.6}$$

where $C_{sales,current}$ denotes the annual revenue from product sales (€/a) at a fixed reference price, $C_{net,current}$ the annual net cash flow (profit) of the whole investment (€/a) with the same reference price, T_{oper} the annual operating hours of the plant (h/a) and Q_{sales} the product flow to markets (MW). The numerator of the equation represents the amount of sales revenue that is needed to be deducted from the annual cash flow to reach zero profit.

The purpose of expressing the break-even price in this form is that all of the needed variables will be ultimately calculated by the TEA-tool model. The closed formula will be used to calculate additional output in the TEA-tool. The break-even analysis in this study will be executed iteratively.

The break-even price needs to be associated with the expected bioproduct price, after which it can be used as a relative profitability measure. Addressing the profit margin from sales (also called the *markup margin*) is a useful way of analysing the investment feasibility. The sales profit margin does not tell anything about the absolute returns of the investment but it can be used to assess, how deep in the money the product value is compared to the manufacturing value. The break-even prices cannot be used in cross-technology comparisons, unless the compared bioproducts were targeted to same markets with same substitute fuels.

3.1.6 Depreciation

The initial fixed investment cost is usually allocated evenly to the investment's life time. This method is called depreciation. Depreciation is usually related to accounting but it can also correspond to actual loan payments. The advantage of expressing the investment costs through depreciation is that it relates the operating cash flows to the size of the investment.

The annual depreciation represents the nominal annual expense that nets the total value of the initial investment when discounted to year 0. The formula for constant annual depreciation C_{depr} (€) can be directly derived from the NPV formula (equation 3.3) and written as:

$$C_{depr} = \frac{C_{invest}}{\sum_{n=0}^{N} \frac{1}{(1+r)^n}}, \qquad (3.7)$$

where C_{invest} denotes the initial investment (€), r the real discount rate and N the number of economic operating years. The depreciation-corrected annual net cash flow yields zero with an investment life time equal to the discounted payback time.

3.2 Biorefinery investment cash flows

The cash flows related to the biorefinery investments can be divided into three categories: investment expenditure, operating costs and revenues. The investment expenditure covers the initial investment cost occurring at the initial time (year 0) of the investment life time. The operating costs cover all the fixed and variable running costs occurring during the investment life after the initial time. The revenues consist of product sales and savings resulting from the investment.

The investment cost of a biorefinery can be assumed to be a function of the maximum production capacity. In the tool, the correlations between the plant size and investment cost are based on previous offers and cost estimates. More accurate price modelling would not offer much value to the model as the main purpose of the tool is to observe how the input variables (including the investment cost) affect the overall profitability. With this approach the amount of input variables is also reduced. It has to be remembered, however, that the actual investment costs differ a lot depending on the case and contain various price components that are not necessarily functions of the plant size.

The operating costs include general operating and maintenance costs, electricity costs, marginal costs of steam usage and process additive (CO_2, H_2SO_4, NaOH) purchase costs. For pyrolysis and steam explosion, the expenses of mixture biomass purchases is also included. The operating cash flows have to be allocated to represent the actual investments that they result from. In the TEA-tool a convenient way to execute this is to compare a reference scenario to a scenario where one or several biorefineries have been implemented.

The revenue generated by the investment is estimated as the marginal increase in positive cash flows. This means that the profitability indicators calculated by the tool represent the individual investment scenario rather than the whole mill profitability. The same principle applies to operation and maintenance costs, which are addressed only from the amount that they raise the overall expenses.

The price evaluations for the bioproducts are expressed as a single value that is assumed to include all transportation and distribution expenses in addition to the nominal market price. It is important to note that the local operating environment may affect the gate price of imported and exported products and the actual price may differ significantly from the market prices. For this reason, the costs have to be estimated according to the net cash flow generated after the product exports from the mill. Uncertain market conditions can be replicated by decreasing the liquid market price of the products and reducing the availability of the biorefineries.

The value of increased pulp production resulting from LignoBoost implementation will be calculated through the profit margin of pulp sales. This eliminates the need for detailed process cash flow modelling. The method is justified also because the fact that the tool is not supposed to calculate the overall profitability of the mill. The profit margin thinking can be applied to other asset prices as well. The idea is that an internal transaction can be priced after the marginal changes in loss or profit it causes.

3.3 Bioproduct exports and lime kiln fuel savings

When implementing certain biorefineries, a decision has to be made regarding the bioproduct end use. Basically this means, how much biofuel is used locally as a lime kiln replacement fuel instead of exporting the product from the mill site. Assuming that the mill always aims to achieve maximum available profit, the income can be calculated as the maximum of potential lime kiln savings versus the market value of the corresponding biofuel production. The advantage of this assumption is that the bioproduct end use (lime kiln or markets) becomes a function of the market price, thus eliminating the need for separate analysis for these two cases.

The aim is now to define a threshold price, at which the bioproduct sales start to make higher profit than the local lime kiln use. In a market export scenario, the cash flow from sold bioproduct (€/h) can be calculated as

$$C_{sold} = P_{bio} Q_{bio} \ , \tag{3.8}$$

where P_{bio} denotes the bioproduct market price (€/MWh) and Q_{bio} the energy flow of the product (MW). If the price is expressed in relation to a dry product ton (lignin), the formula transforms into

$$C_{sold} = \frac{P_{bio,tDS} H_{bio,wet}}{3.6(1-x_{water})} Q_{bio} \ , \tag{3.9}$$

where $P_{bio,tDS}$ denotes the product price (€/tDS) and $H_{bio,wet}$ the lower heating value of wet product (MJ/kg) and x_{water} the water content of the bioproduct. The multiplier *1/3.6* results from the unit conversions from seconds to hours and from tonnes to kilograms.

In a lime kiln combustion scenario, the fuel-dependent heat demand (MJ/t lime) can be calculated as a weighted average of the specific heat demands for the used fuels. Mathematically this can be expressed as

$$q_{LK,total} = X_{bio}q_{bio} + (1 - X_{bio})q_{default} \,, \tag{3.10}$$

where X_{bio} denotes the net heat rate percentage of the bioproduct, q_{bio} the specific lime kiln heat demand for the biofuel (MJ/t lime) and $q_{default}$ the specific lime kiln heat demand for the default fuel.

The increase in lime kiln heat demand induced by the biofuel use can be written as $K_{bio} = (q_{LK,total} - q_{default})/q_{default}$. After substituting $q_{LK,total}$ in this formula by equation 3.9 the increase in fuel demand can be simplified into

$$K_{bio} = X_{bio}\left(\frac{q_{bio}}{q_{default}} - 1\right). \tag{3.11}$$

The lime kiln fuel savings can be calculated as the net difference in total lime kiln fuel use. In practice, the fossil fuel fraction of the increased gross heat demand is subtracted from the default lime kiln heat load and then multiplied with the fuel price. The formula for fuel savings (€/h) becomes:

$$C_{LK} = (1 - (1 + K_{bio})(1 - X_{bio}))Q_{LK}P_{LK} \,, \tag{3.12}$$

where P_{LK} denotes the price of lime kiln fuel (€/MWh) and Q_{LK} the default heat input (MW) of lime kiln fuel with no biofuel co-firing.

The above formulas can be used to solve the threshold price at which the sold product would start to net higher income than the potential fuel savings. From the profitability point of view, it is assumed that the selection of bioproduct end use is chosen according to the highest profit. When the bioproduct end use is tied to the product market price, the number of required test scenarios is reduced. The biofuel market threshold price P^*_{bio} can be solved by setting the revenue from sold product (equation 3.8) and lime kiln fuel savings (equation 3.12) equal. This gives us the following equation:

$$P^*_{bio}Q_{bio} = (1 - (1 + K_{bio})(1 - X_{bio}))Q_{LK}P_{LK} \,, \tag{3.13}$$

where the considered biofuel fraction of lime kiln fuel can be expressed as $Q_{bio} = (1 + K_{bio})X_{bio}Q_{LK}$. Substituting this and solving the equation with regards to P^*_{bio} (€/MWh) the equation yields

$$P^*_{bio} = P_{LK}\left(\frac{1}{X_{bio}(1+K_{bio})} - \frac{1-X_{bio}}{X_{bio}}\right). \tag{3.14}$$

Alternatively, if the bioproduct market price is expressed as €/tDS (namely lignin), the formula for the market threshold price becomes:

$$P^{*}_{bio,tDS} = \frac{H_{bio,wet}}{3.6(1-x_{water})} P_{LK} \left(\frac{1}{X_{bio}(1+K_{bio})} - \frac{1-X_{bio}}{X_{bio}} \right), \tag{3.15}$$

where $H_{bio,wet}$ is the lower heating value (MJ/kg) of wet biofuel, x_{water} is the water content of the wet biofuel and the default lime kiln fuel price P_{LK} is still expressed as €/MWh.

The market threshold price is used in the feasibility analysis to tie the lignin and pyrolysis oil end uses to the bioproduct price. In this study, the maximum percentage of lime kiln fuel replacement is limited to 50 % for lignin and 30 % for pyrolysis oil. The excess production is exported to markets.

3.4 Process heat pricing

The process heat consumed by the biorefineries can be priced according to the loss of revenue caused by steam usage. Because a modern pulp mill generates more heat than the pulping process consumes, the price for the heat is generally low. For the turbine back pressure steam or hot water condensate, the cost can be negative if there were otherwise no other uses for the excess heat. Steam extracted between the turbine stages can be priced after the lost electricity production that would have taken place if the steam expanded through the end of the turbine.

An ideal expansion of steam from pressure p_1 to p_2 through the turbine would be isentropic. In practice, each turbine has a distinctive isentropic efficiency η_s which is defined as

$$\eta_s = \frac{h_1 - h_2}{h_1 - h_{2s}}, \tag{3.16}$$

where h_1 denotes the steam enthalpy before the expansion, h_{2s} the enthalpy after isentropic expansion to pressure p_2 and h_2 the actual enthalpy after expansion. The isentropic and actual expansion processes are drawn in figure 3.1.

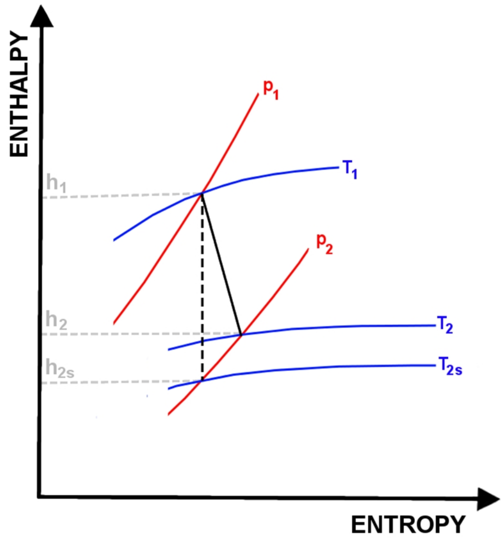


Figure 3.1 *Expansion through a turbine from pressure p_1 to p_2. The ideal isentropic process is represented by the dashed line and the actual process by the solid line.*

The enthalpy difference $h_1 - h_2$ is transformed into kinetic energy in the turbine and further into electricity in a generator coupled with the turbine. The power transferred from the steam can be calculated as

$$Q = \dot{m}(h_1 - h_2) = \dot{m}\eta_s(h_1 - h_{2s}), \tag{3.17}$$

where $\dot{m}$ is the steam mass flow. The generator produces electricity from this enthalpy difference by the amount of $P = \eta_g Q$, where η_g is the generator efficiency.

Because the model focuses on the biorefineries, the pulp mill mass flows will only be modelled in a general level. For steam consumption this means that as an output, the model gives the increase in steam consumption rather than the absolute amount of consumed steam. The turbine balance is also expected to stay fairly constant with small changes in steam consumption. This approach removes the necessity of giving the exact steam parameters as inputs, although more accurate results would be obtained with turbine-specific data.

The steam consumption of a biorefinery is priced according to the theoretical loss in sold electricity that would have occurred if the steam expanded through the turbine. The TEA-tool model is supposed to take the steam consumption into account by defining this price for the consumed steam. The model should also roughly predict the changes in the net steam balance. The steam balance changes related to the increased pulp production are included in the profit margin of pulp. Hot water is considered being free as long as the demand does not exceed the supply.

4. REVIEWED BIOREFINERIES

This chapter aims to give an overview of the biorefinery technologies assessed in the TEA-tool. The process descriptions in the following sub-chapters are intended to give a brief look at the technologies and their relations to the pulping process. For each technology, a general technical description is provided, followed by a summary of constraints limiting the investment size.

This feasibility study is performed for four technologies: LignoBoost, gasification, integrated pyrolysis and steam exploded black pellets. Steam explosion, gasification and pyrolysis utilize a portion of the mill's residue bark that would otherwise be combusted in the utility boiler. LignoBoost, on the other hand, focuses on extracting lignin biofuel from black liquor stream and thus affects directly to the kraft process balance. The choice of the studied technologies is made according to current market conditions and Valmet's marketing intentions. The analysed processes are modelled after the specifications and performance data gathered from Valmet. Alternative process setups and process variations might have completely different profitability dynamics.

Each of the biorefineries is built around the idea of refining biomass into products of higher value and thus increasing the overall cash flows of the mill. The direct operating revenues are generated through bioproduct sales and lime kiln fuel savings. In some cases, implementation of the biorefineries also affects the balance of the underlying mill, therefore generating revenue streams from increased pulp production and potentially increased electricity production.

The operating expenses are caused primarily by process additive purchases, maintenance costs, additional mixture biomass purchases and electricity consumption. Additionally, each of the technologies consume steam, most of which is related to biomass drying steps. In pulp mill integrated biorefineries, the steam consumption plays a relatively small role because of the available low-cost heat. Despite of this, the process heat consumption has to be addressed to ensure that the steam balance does not bottleneck the mill.

4.1 Lignin extraction by LignoBoost

Wood consists mainly of cellulose, hemicellulose, lignin and extractives. In the fibrous structure of wood, lignin works as a binding substance that keeps the cellulose fibres together. Lignin itself is compounded of complex structures of organic polymers [1, p. 6]. The amount of lignin in wood varies usually between 20-30 wt-% [1, p. 2] [18] for the most common softwoods and hardwoods used in kraft pulping. The production of kraft pulp is based on the process of separating wood fibres from each other and dissolving lignin from the wood biomass. Therefore the pulping process generates a side product stream with high lignin density. In kraft process, the lignin content of the cooking liquor is combusted in the recovery boiler. [18]

Instead of combusting in the recovery boiler, lignin can also be extracted from the black liquor stream to produce lignin biofuels with a relatively high energy density. This process is called the LignoBoost. The extracted lignin biofuel can be used locally at the pulp mill or sold to markets. In addition to the produced lignin, the process can also be used to debottleneck the recovery boiler, thus increasing the headroom in pulp production. Due to the relatively high energy density, hydrophobic nature and good grindability, the market potential for lignin pellets are promising. The produced lignin has also uses outside of combustion applications, as it can be used as a raw material for other bioproducts. [18]

4.1.1 Process description

LignoBoost is a patented process in which a fraction of lignin is extracted from the pulp mill black liquor and densified into a high energy content lignin product [19]. The schematic of the process is shown in figure 4.1.

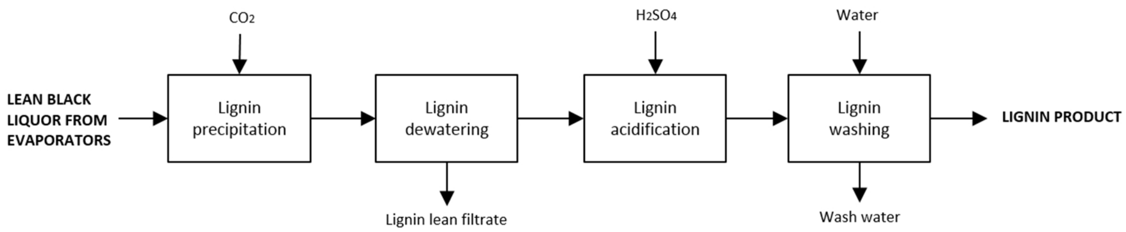


Figure 4.1 *A simplified flow chart of the LignoBoost process.*

First, a stream of black liquor is taken from the evaporation plant to precipitation, where CO_2 is injected in the liquor-water suspension. The CO_2 induces a decrease in pH making the lignin more hydrophobic and eventually precipitating it from the liquid. The lignin precipitate is then filtered from water with a chamber press. After filtering, the lignin cake is dispersed in water and acidified with H_2SO_4 to control the solution pH level. During acidification, a major fraction of the remaining sodium is transferred to the water phase. The idea behind the re-dispersion and acidification is to improve the washing properties of the lignin cake. After acidification, the lignin slurry is once again washed and filtered in another chamber press. Depending on the use, the lignin product can then be either pelletized or pulverized. [19] [20]

Because the added H_2SO_4 would change the sulphur balance of the mill, the added sulphur needs to be extracted after lignin washing. The sulphur is removed from the precipitator dust where the sulphur is bonded into Na_2SO_4. The overall sodium shortage resulting from the removal of Na_2SO_4 needs to be corrected by adding NaOH to the process. [21]

The direct operating costs from the used process chemical are caused by CO_2, H_2SO_4 and NaOH purchases. The CO_2 needed in the lignin precipitation could theoretically be produced from the lime kiln flue gases [19] but this would require complex gas cleaning equipment and is not taken into consideration in this study. The electricity demand of the LignoBoost process is primarily dictated by the vertical pressure plate filters.

4.1.2 Constraints and effects to the mill balance

In kraft pulp mills, the pulp production is usually bottlenecked by the recovery boiler heat load. The extraction of lignin decreases the absolute amount of black liquor flow as well as the average heating value of the black liquor. Thus, the black liquor flow entering the recovery boiler can be increased, directly implying an increase in pulp production. The increase in pulp production has an impact to the whole mill balance through upscaling mass flows. The mass flow changes affect particularly to the evaporator plant, in which the process flows may exceed the design capacity. [19]

Because the lignin removal decreases the black liquor heating value, the liquor flow to the recovery boiler has to be increased in order to maintain the heat load. In addition, while the extraction of lignin lowers the black liquor heating value, the amount of inorganic compounds in black liquor remains unchanged. Therefore the smelt flow from the recovery boiler increases and thus limits the increase in black liquor flow. [18]

The constraints mentioned above generate an optimizing problem where a good balance between increased pulp production and lignin removal percentage has to be found. Both of these design parameters may vary between 5-25 % depending on the case, the maximum value for both variables being around 50 %. Another limiting factor for the process scaling is the amount of effluent water from the second filtration step to the evaporators. Usually an extraction of 25 % of the lignin can be done without causing major problems to the kraft process [20]. [18] [22]

The revenue streams generated by LignoBoost consist of three main cash flows: increased pulp production, savings in lime kiln fuel purchases and lignin sales to markets. The most convenient use for produced lignin at the pulp mill is to use it as a partial replacement fuel for fossil fuels combusted in the lime kiln. Although this requires modifying the burners, the problem of market distribution does not have to be solved. The use of lignin as a lime kiln fuel has been confirmed to be technically viable with gross heat loads of 50-100 % in the kiln [23]. [22]

Although the recovery boiler is usually assumed to be the first bottlenecking component in a kraft pulp mill, capacities of other departments have to be taken into account. The problem with a recovery boiler bottleneck is that the maximum capacity cannot easily be increased without large investments [20].

The market prices of lignin, pulp and lime kiln fuel define the optimum lignin end use and extraction percentage. Theoretically, combusting the produced lignin in the utility boiler could also become a viable option if the increased pulp production dominates the cash flow and the lignin end product value is low. Although unlikely, examining this kind of scenario will also be possible in the TEA-tool.

4.2 Bark gasification

Thermal gasification is a process in which organic solids are turned into gaseous compounds in the presence of under-stoichiometric amounts of oxidizing agent. The product gas can be used as a fuel in various combustion applications. The main combustible components in the product gas are carbon monoxide (CO), hydrogen (H_2) and methane (CH_4). The gasification reactions are endothermic and they occur at the temperature range of 700-1400 °C. The heat required in the process is usually supplied by partial combustion of the feedstock. [24, pp. 124-127]

The choice of oxidizing agent affects the gasification process and the product gas properties significantly. The most common oxidizing agents are air and oxygen. Biogas produced by oxygen gasification has a much higher heating value but the process requires a constant supply of oxygen. [25, pp. 118-119]

Because the production of oxygen is very energy intensive [25, p. 182] and the amount of oxygen consumed is significant, air gasification is usually more commonly used. In this study only air gasification is reviewed.

A typical heating value for air gasified product gas is usually only 3-7 MJ/m^{3}n because of the nitrogen content of approximately 50 vol-%. Compared to the heating value of 7-15 MJ/m^{3}n for oxygen gasified product gas, the low heating value of air gasified gas sets limits to the utilizing options. [2, pp. 568-569]

4.2.1 Process description

The gasifier types vary between applications with different scales, operating environments and process requirements. Considering Valmet’s offering portfolio and the case-related properties of a lime kiln supplying gasifier, only circulating fluidized bed (CFB) gasifier is included in the TEA-tool. In case another type of gasifier is later needed in the comparison, the economic data and mass-energy balances need to be adjusted accordingly.

The most important advantages of the CFB-gasifier are the compatibility with low energy content fuels, easy scalability, a uniform temperature distribution in the reactor and a high moisture content allowed for the input fuel. The power to size ratio is also higher than in bubbling fluidized bed (BFB) reactors. [2, pp. 569-571]

Figure 4.2 shows a flow chart of the gasification process. In a pulp mill environment the reviewed gasification process focuses on producing fuel for the lime kiln from bark residue generated in debarking. Although the gasification reactions are complex and sensitive to operating conditions, the process flow itself is very straight forward.

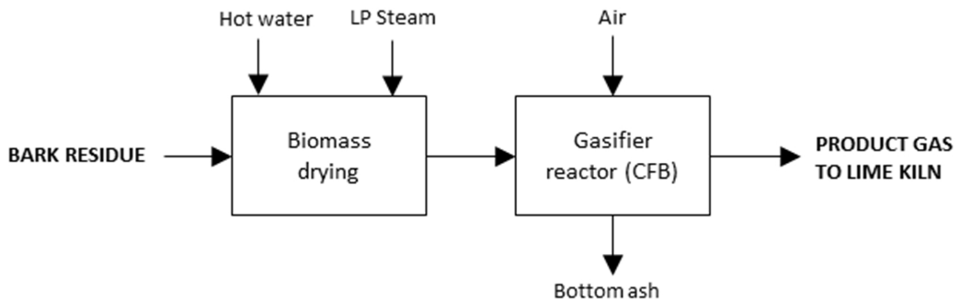


Figure 4.2 *A simplified flow chart of the gasification process.*

The bark flow from woodhandling is dried in pre-dryers operated with heat from the kraft process. After drying the biomass is fed to the gasifier.

The fuel enters the gasifier from the lower part of the reactor into a circulating mass suspension. The air is injected to the circulating flow from the bottom of the reactor. The bed material circulates from the reactor through a cyclone back to the reactor bottom. The gasified product gas is separated from the circulating suspension in the cyclone. The reactor temperature level (750-850 °C) is adjusted by fuel and air flow [7]. The process heat needed in the gasification reactions is supplied by partial combustion of feedstock.

The produced gas is combusted in the lime kiln. The produced gas is injected directly into the lime kiln and no intermediate storage of product gas is required. This allows the gasifier feed to be steered directly by observing the lime kiln energy demand. To lower the ash flow from gasifier to the lime kiln, a second cyclone is sometimes used [4].

In theory, all the bark energy content entering the gasifier reactor is utilized in the lime kiln. This implies that the required amount of dried biomass can be directly calculated from the heat demand and bark heating value [4]. The required biomass input flow obtained this way has to be corrected only by heat losses occurring in the gasification process. In order to maintain a sufficient heat load for the lime kiln, the low heating value of the product gas has to be compensated by high fuel flow.

4.2.2 Constraints and effects to the mill balance

In this study the bark gasifier is only considered being installed as the lime kiln fuel supply. This means that the only function of the gasifier is to transform a portion of residue bark into a fuel suitable for lime kiln use. In addition to bark, the gasifier can operate with other biomass feeds as well. However, bark can be considered as the most convenient gasifier fuel in pulp mill cases.

The bark gasifier causes very little to no disturbances to the pulping process as long as surplus heat is available for biomass drying. Some equipment modifications and increased need of purging may still be needed because of non-process elements ending up to the lime cycle from bark [4].

The process is scaled according to the product gas demand in the kiln. Because a conventional lime kiln could replace up to 100 % of its fuel with gasified biomass [26], the gasifier should be scaled after the maximum lime kiln capacity. Downscaling the plant makes no economic sense, especially because the added capacity does not significantly increase the overall investment costs. Despite of this, the TEA-tool will not limit the gasifier capacity.

4.3 Integrated pyrolysis

When organic biomass is heated in a non-oxidizing environment it transforms into char, volatile organics and other gaseous compounds [2, pp. 192-202] [25, pp. 65-71]. As a chemical process, pyrolysis shares some similarities with gasification, the main differences being the lower operating temperature and the absence of air in the pyrolysis process. The pyrolysis processes can be categorized into fast and slow pyrolysis depending on the retention time [25, p. 71]. The pyrolysis reactor introduced in this study operates in the area of fast pyrolysis utilizing rapid heating and short retention times.

The fast pyrolysis products consist of water, char and organic chemicals in liquid and gaseous phases. The aim of the process is to condense the gaseous pyrolysis substances into liquid bio oil that can be used for combusting or further processing. The uses of pyrolysis oil vary from combustion applications to chemical refining [25, p. 309]. The bio oil produced by the studied pyrolysis process is considered being only suitable for substitute uses for heavy fuel oil. Potentially, pyrolysis oil could also be further refined into transport fuels or other chemicals [27].

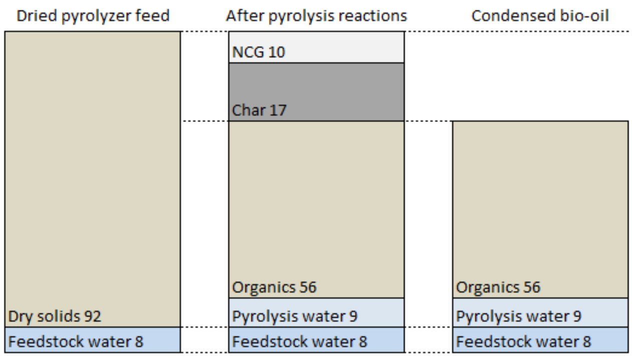
The char and non-condensable gases (NCG) generated in pyrolysis process are used to power the process and generate heat for steam production. In a pulp mill environment, integrated pyrolysis aims to take advantage of the available process heat and biomass residue streams and integrate the pyrolyzer to an existing fluidized bed boiler. The produced pyrolysis oil can be used to replace a fraction of lime kiln fuel or can alternatively be sold to markets.

Some of the most important advantages of pyrolysis oil are better durability and transportation properties than those of a raw biomass residue, relatively low processing costs and the possibility to replace other fossil fuels in combustion applications. The viscosity, moisture, corrosiveness and poor volatility properties of the pyrolysis oil, however, set limitations for the potential uses for different cases. [27]

The pyrolysis reactions occur in the temperature range of 400-650 °C [28, p. 836]. The maximum bio oil yields for fast pyrolysis are reached around the temperature of 450 °C [29] and this is a common temperature used in most fast pyrolysis reactors. For bark residue, the bio oil organics yield can be assumed to span a range of 55 % to 65 % [7] [30, p. 15].

The exact bio oil composition as well as the oil yield depend on process parameters and the feedstock composition. The oil yields are mostly dependent on the feedstock rather than reactor parameters, and for this reason the yields can be assumed to be feed-specific [30, p. 15]. Typical liquid yields will vary between 60-75 wt-% of feedstock dry solids [30, p. 15] [28, p. 841]. A typical water content for pyrolysis oil products is 15-35 % depending on the process parameters and the moisture content of the dried feedstock [28, p. 838] [31] [32, p. 27].

Example values of typical pyrolysis yields have been gathered into figure 4.3. To avoid confusion caused by different naming conventions, it is important to note that in this study, the liquid fraction is divided between organics and water generated in the pyrolysis reactions. Pyrolysis oil yield is therefore the sum of pyrolysis organics and pyrolysis water yields. Alternatively the bio oil yield could be expressed as the gross yield from the biomass to the end products. This is important to remember when gathering yield data from various sources.



Feedstock water content after drying	8 %
Organics yield of dry feed	61 %
Pyrolysis water yield of dry feed	10 %
Non-condensible gases yield of dry feed	11 %
Char yield of dry feed	18 %
Total water content of the bio oil	23 %
Total bio oil yield of dry feed	73 %

***Figure 4.3** Example pyrolysis yields*

When the pyrolysis process is scaled after a certain capacity, the amount of needed biomass input can be derived analytically. First, let us denote the production capacity with $\dot{m}_{product}$. This product flow can be expressed as:

$$\dot{m}_{product} = \dot{m}_{ds}\left(y_{organics} + y_{water} + \frac{x_{feed}}{1-x_{feed}}\right), \tag{4.1}$$

where $\dot{m}_{ds}$ denotes the mass flow of feedstock dry solids, $y_{organics}$ the pyrolysis organics yield, y_{water} the pyrolysis water yield and x_{feed} the moisture content of the feedstock. The required biomass input can be trivially solved from the equation as

$$\dot{m}_{ds} = \frac{\dot{m}_{product}}{(y_{organics}+y_{water}+\frac{x_{feed}}{1-x_{feed}})}. \tag{4.2}$$

Integrated pyrolysis is more sensitive to feedstock water content than gasification because water passes the process to the end product, reducing the heating value of the pyrolysis oil. On the other hand, some water content may be advantageous due to its tendency to lower the oil viscosity [28, p. 838]. In addition to the feedstock moisture, part of the end product water content is generated by the pyrolysis reactions that yield around 10 wt-% of the dry biomass feed [31, pp. 206,209].

4.3.1 Process description

The pyrolysis reactor (pyrolyzer) is integrated to a fluidized bed (BFB or CFB) boiler that works as a heat source for the process. Typical pulp mill sites would usually have an existing bark boiler that serves this purpose. Building a new boiler for the pyrolyzer could theoretically be possible, but in the presence of alternative investment options, this is not considered being economically viable. This being said, the new utility boiler investment scenario will be left with minimal attention in this study. In the TEA-tool, the need for a boiler investment can still be toggled on and off.

A simplified process flow chart of integrated pyrolysis is represented in figure 4.4. The integrated fast pyrolysis process uses sawdust, forest residue or woodchips as its primary feedstock. In a pulp mill environment, it is also attractive to include a fraction of bark residue to the biomass feed. The plant aims to take its feedstock from the biomass residues generated in woodhandling and to transform them into pyrolysis oil product. The uncovered biomass demand has to be imported to the mill site.

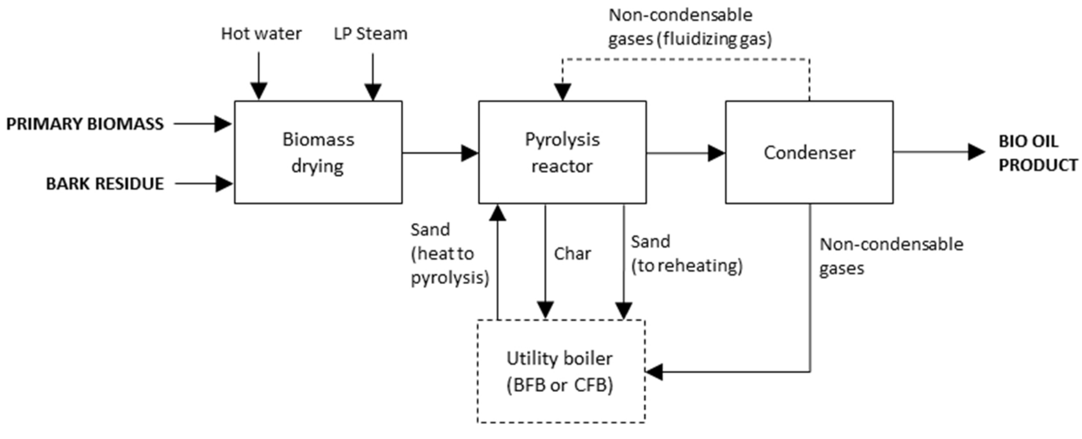


Figure 4.4 *A simplified flow chart of the integrated pyrolysis process.*

The first step of the integrated pyrolysis process is drying, where biomass is dried to a moisture content of 8-10 wt-%. This moisture content can be regarded as a standard for most of the pyrolysis process feeds as it offers a good balance between drying costs and end product quality [11]. The drying heat can be supplied by hot water or low pressure steam. After the dryer the biomass is crushed and stored in a feed silo, from where it is fed to the pyrolysis reactor.

The pyrolysis reactor examined in this study is a so called riser reactor. The structure of the riser reactor closely resembles that of a CFB gasifier. The key difference is that no air input is needed for the pyrolysis reactions. In addition to this, the needed pyrolysis process heat is brought among fluidized bed sand from the boiler, whereas the CFB gasifier would have its own, separate sand circulation. [11]

The process heat input from char and non-condensable gas combustion can be thought of as a partial combustion of feedstock. However, due to the low reaction temperature, the heat is supplied indirectly. Although the energy content of pyrolysis by-products could theoretically cover the whole process heat demand, back-up fuel is usually needed for proper utility boiler operation.

The gaseous pyrolysis products are separated from the circulating stream in a cyclone and condensed into a liquid phase in a separate condenser. The remaining sand-coke mix enters the utility boiler for reheating and combustion. Most of the non-condensable gases from the condenser is transferred back to the pyrolyzer where they act as fluidizing gas. The rest is injected into the utility boiler for extra heat generation thus contributing to the mill steam and electricity production. In a time-independent steady-state scenario, the utility boiler NCG input can be set as equal to the amount of NCG exiting the condenser i.e. no mass accumulates in the process.

The required heat for the pyrolysis process can be expressed through the enthalpy for pyrolysis that is dependent on feedstock properties. The enthalpy of pyrolysis is a parameter that indicates the total amount of heat required for the whole process from ambient temperature feedstock into pyrolysis oil product. A typical value for enthalpy of pyrolysis varies around 1.5 MJ/kg of biomass feed. For integrated pyrolysis this value indicates the gross heat transfer from circulating sand to the pyrolyzer. [33]

4.3.2 Constraints and effects to the mill balance

The char and non-condensable gases injected into the utility boiler contain a significantly higher energy content than the heat required to the pyrolysis process. This means that the sand reheating could theoretically be covered by char and NCG combustion alone. However, the maximum practical amount of pyrolysis heat input into a BFB boiler is 50 % of the total boiler capacity, while a CFB boiler be fully loaded with pyrolysis heat [11]. The TEA-tool will indicate the utility boiler heat load but the user is responsible of evaluating the actual performance constraints. In a pulp mill scenario, the utility boiler (bark boiler) would most likely be a bubbling fluidized bed boiler. Reaching of the 50 % heat load limit of the pyrolysis heat can most conveniently be avoided by maintaining a portion of bark load to the boiler. Adding fuel oil or natural gas input to the boiler is not considered being attractive in this study, although the TEA-tool will be built to allow this kind of implementation as well.

The maximum fraction of bark in the pyrolysis feedstock is limited by the ash content of the gross dry solids input. The allowed limits are defined by the length of the bio oil storing period. This restriction has to be taken into account especially when the bio oil is sold to markets, thus having a long storing period. [11]

For pyrolysis oil products, the lower heating value will vary around 15-17 MJ/kg [28, p. 838] [11]. The low heating value compared to the conventional fuel oils is mainly caused by the high oxygen content that is distinctive for organic biomasses [34, p. 139]. This range will be used as a reference for the tool in cases where more detailed analysis is unavailable. The oil can be assumed to fulfill quality requirements when moisture is below 30 % and solids content is below 0.5 %. [35]

The markets for pyrolysis oil are currently partially incomplete due to competing products and lack of standardization [32]. For this reason, the integrated pyrolysis is primarily considered being installed as a lime kiln fuel supply. However, the option to sell the produced bio-oil to markets works as an alternative motivation for bio oil production in case the use in lime kiln becomes unattractive. From the standpoint of the TEA-tool, this aspect can be taken into account by changing the pyrolysis oil market price and observing how the changes affect the profitability.

4.4 Black pellets by steam explosion

Steam explosion is a process where lignocellulosic biomass is treated with steam in high temperature and pressure, followed by rapid depressurizing. In the process, the internal structure of the biomass is broken down to smaller fragments and the fibres are separated from each other. The output biomass is easily mouldable and hydrophobic, making it an ideal material for pelletizing purposes. [36]

Steam exploded black pellets are more durable, more water repelling and have a much higher energy density than conventional wood pellets. Because of these properties, the pellets can be transported and stored easily. The advantages of black pellets offer an attractive solution for exporting the wood-based biomass further from its point of origin, thus expanding the potential markets of pulp mill wood residues. The black pellet also works as a substitute product for hard coal and conventional wood pellets, easing the market penetration. [36]

4.4.1 Process description

The schematic of a steam explosion process is presented in figure 4.5. In a pulp mill environment, the steam explosion process takes its primary feedstock from the residues generated in woodhandling and debarking. The gross content of the feedstock is a mixture of bark and other wood residues such as sawdust or woodchips. The mixture wood is used to ensure the quality of the pellet product. [12]

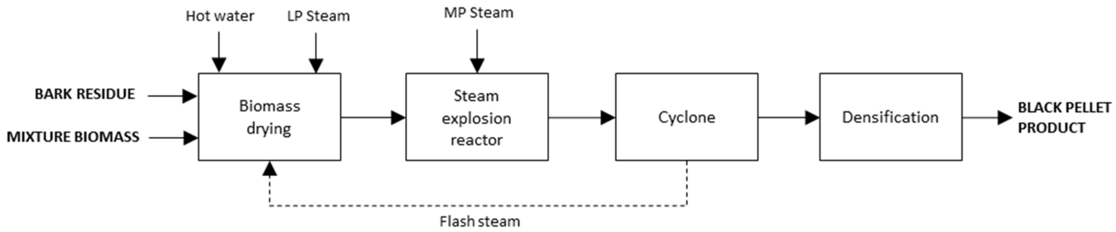


Figure 4.5 *A simplified flow chart of the steam explosion process.*

The process starts with pre-screening, where unwanted materials such as clay and stones are removed from the biomass. After screening, the biomass is pre-dried, hammered and pre-heated. The drying is performed with conventional belt dryers operating with low pressure steam or hot water. [36] [12]

After the dryer, the water content of the biomass is typically around 8 wt-%. The dried and pre-heated biomass is fed to the steam explosion reactor with a plug screw and saturated with medium pressure steam. The steam-saturated wood is then depressurized through a discharge valve. The rapid depressurizing tears the wood structure apart and the fibres are separated in an explosion-like reaction. The biomass is separated from steam in a cyclone and the flash steam is transferred to drying and pre-heating. The biomass is then directed to densification, where it is pelletized into the black pellet product. [36] [12]

The use of flash steam lowers the overall low pressure steam demand for drying compared to the other biorefining processes. The medium pressure steam consumption in the actual explosion process stays fairly constant and can be assumed to be a function of production capacity. [12]

4.4.2 Constraints and effects to the mill balance

The quality of black pellets is highly dependent on the used feedstock and if only residue bark was used, the pellets would not fulfill the market standards because of a low durability. For this reason, 30-50 wt-% of wood is used as a mixture fuel entering the process. If such wood biomass is not available at the mill, the mixture biomass has to be imported. [12]

The heating value of the pellet product can be increased by lowering its moisture content, increasing the process temperature level or using longer retention times. These changes increase the drying energy input and reactor size. In addition to this, the heating value is affected by chemical conversions of the treated biomass. [12]

The pressure level of the turbine medium pressure steam has to be considered when steam explosion is implemented. This is because the pressure level of the steam might not be sufficient for the explosion phase. The specific consumption of medium pressure steam stays constant regardless of the feedstock moisture because the biomass is dried prior to the process [36]. It also has to be noted that the dirty flash vapors cannot be returned to the steam process and therefore new feed water has to be added. Heating of this new feed water increases the overall process heat demand of the steam explosion process.

The dirty residue steam from the process contains volatile organic compounds originating from the biomass feed. This organics fraction can be separated and combusted for partial energy return. The amount of energy transferred this way, however, is insignificant when compared to the energy balance of the whole process. The marginal increase in effluent treatment costs have also been ignored in this study. The condensates from the steam explosion process are mixed with the other condensates of the pulp mill. The increased amount of condensate does not make any major difference to the mill balance. [12]

Given that the main goal of implementing steam explosion is to improve logistical and handling properties of the biofuel, the produced black pellets are always considered being sold to markets. Converting bark residue into black pellets for a local combustion application would provide little to no benefits.

5. IMPLEMENTATION OF THE TEA-TOOL

The main goal of this thesis work was to build a TEA-tool for techno-economic comparison of the four introduced biorefinery technologies. This chapter presents the main designing principles of the tool, explains the logic behind the calculations in a general level and introduces the new graphical user interface.

The balance model contains several intentional simplifications to avoid overcomplicating the calculations. Knowing the assumptions and approximations behind the model implementation is essential to properly understand the results.

5.1 General design

The task of the TEA-tool is to visualize the customer benefit of biorefinery implementation and to define the investment's profitability dynamics under different market conditions. The tool is easy to deploy for basic analyses and with moderate effort it can also be calibrated to represent an existing pulp mill scenario. The technical and economic models behind the tool calculations are simplified to improve accessibility and to enable easy customizability.

The central function of the tool is to manipulate the pulp mill process balance by coupling it with process balance models of the biorefinery integrates. The model constantly analyses the active investment scenario and delivers the investment characteristics to the user in a form of economic indicators and technical constraints.

Figure 5.1 illustrates the user work flow in a situation where default process parameters are used i.e. no model calibration is needed. Essentially, the user generates a set of economic scenarios to be compared with the scenario comparison feature of the tool. The scenario comparison is then used to form an overview of the feasibility dynamics to support further decision making.

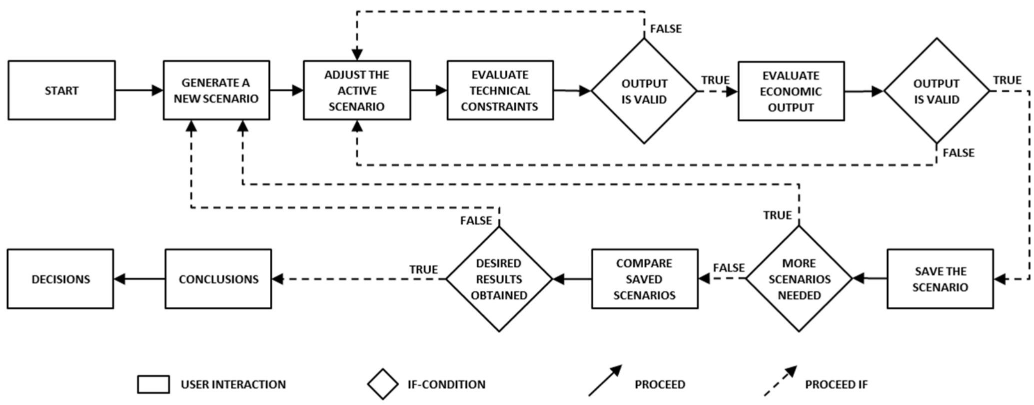


Figure 5.1 *User work flow in a basic analysis.*

The work flow presented in the figure above is intended to be the primary method of end-user interaction. This method is most suitable for preliminary sensitivity analyses and case studies with pre-calibrated mill parameters. Calibrating the model with an actual case mill is considered being advanced user interaction, which requires deeper understanding of the tool logic.

Figure 5.2 displays the overall functioning logic of the TEA-tool, including user interaction and under-the-hood mechanics. The arrows between the logical entities represent the exchange of information. The user input is distinguished from the automated calculations with green lines. The dashed lines denote optional connections that are conditionally active.

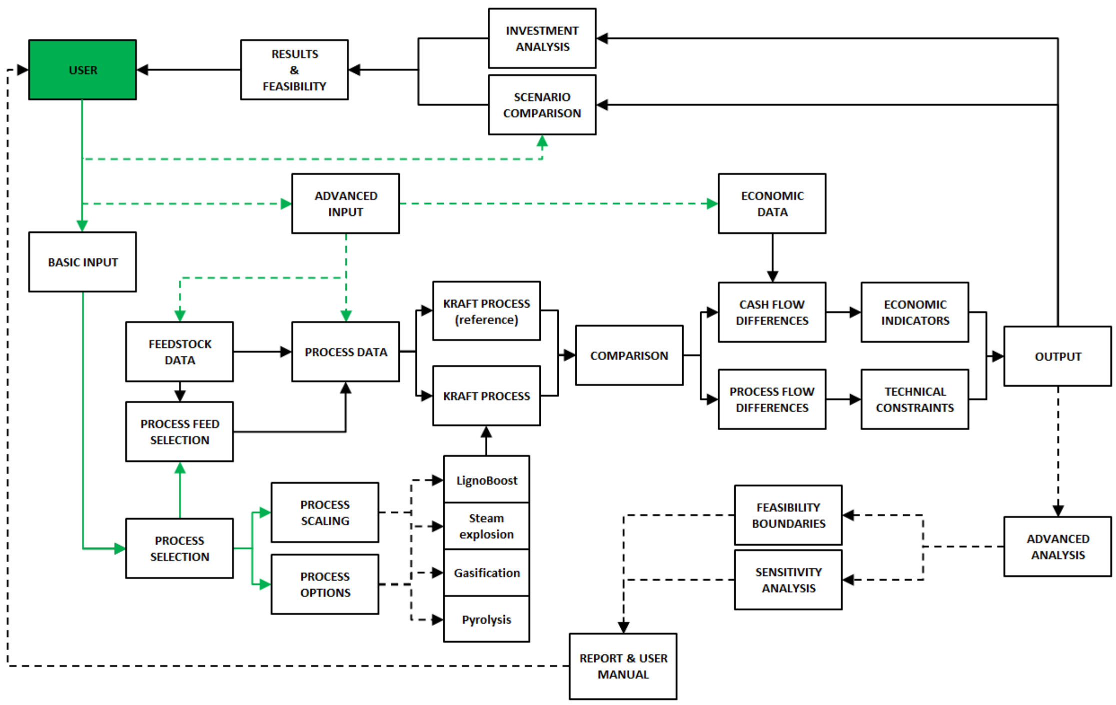


Figure 5.2 *The functioning logic of the TEA-tool. The green lines represent user interaction and black lines represent automated under-the-hood connections. Solid lines denote a flow of information and the dashed lines denote a conditional or optional flow of information.*

The kraft process balance (center of figure 5.2) forms the core of the tool structure, providing a simple process flow model of the pulp mill. The biotechnology flow sheets are coupled with the pulp mill model and can be toggled on and off as preferred. Turning on a bioprocess module immediately affects the overall mill balance. The combination of applied processes is then compared to the balance of the original pulp mill with no integrated biorefineries.

The biorefinery processes are scaled after the requested production capacity, from which the needed feedstock input is calculated. While implementing investment scenarios, the user gets an immediate feedback from the tool in a form of economic and technical constraints. The scenarios can also be saved and compared with each other. The tool does not enforce the user to stay within allowed input value ranges, but most common logical conflicts will trigger warning indications and therefore instruct the user to correct the invalid input values.

The new TEA-tool was intended to be used directly from the MS Excel platform and for this reason a new user interface and means of result visualization had to be designed. The graphical user interface of the TEA-tool is later introduced in chapter 5.3.

During early testing it was discovered that the tool could only handle a couple of continuously calculated scenarios without extending the processing time excessively. This lead to the conclusion that only a few selected dynamic scenarios could be implemented by default. Being independent from the technical limitations, the effect of discount rate was chosen to be included in the dynamic sensitivity analysis of the tool. Advanced sensitivity and feasibility analysis was decided to be included in the written part of the thesis.

5.2 Calculation logic and approximations

The new TEA-tool was intended to expand the analysis of the former LignoBoost TEA-tool into a wider selection of alternative biorefineries. The old tool was used as a reference for the pulp mill process balance. The model was slightly modified to suit the needs of the new modular design. The balances of new technology modules were built using a similar level of detail.

The underlying pulp mill balance was modelled after a specific existing mill. The process was then re-scaled and the yields were slightly modified to conceal the default mill identity. The main purpose of building the tool around a specific reference mill was to ease the model validation. Having a default case mill in the background also enables rough comparisons of the biorefinery investments without the need for a complete list of mill parameters.

Rather than modelling the processes in high detail, the tool calculates the mass and energy flows of the underlying mill with simple correlations. Some of the parameters have been fixed as constants due to insignificant effects to the overall profitability or very little variation in the input values. Several process parameters were linked to the feedstock properties to reduce the amount of required user input to the model.

The feedstock-specific process parameters were included in an integrated feed data table. Examples of feedstock-specific parameters are water content, gross chemical contents, side-product compositions and process yields. In basic tool use, the user responsibility is limited to selecting one of these feeds for each applied bioprocess module. In case the accuracy of the default feed properties is not sufficient, the tool allows dynamic adding of new feeds that can be used straight in the techno-economic evaluations. The feedstock customizability is further discussed in chapter 5.5.

Most of the mill balance data is expressed in relation to the produced air dry ton (ADt) of pulp. This widely used convention allows the mill capacity to be resized without modifying the process parameters. If more accurate analysis is required, the balance of the kraft process can be tweaked to match a specific mill. Calibrating the model to match an actual mill can be performed by following a pre-defined pattern. The TEA-tool model intentionally contains a few strategically defined calibrating parameters that assist in this process. In-depth instructions for calibrating the model are provided in the user manual.

5.2.1 Technical calculation logic and approximations

The mass balance calculations of the TEA-tool process model consist of simple mass and energy balance equations that calculate the input-output correlations with known process yields, material losses and feedstock compositions. The second-hand use of existing correlations causes some inaccuracy to the results. Instead of estimating the level of uncertainty in the process correlations and input parameters, this chapter highlights the most important assumptions of the process-side calculations. Understanding the model approximations is necessary to ensure proper interpretation of the final results.

The amount of needed bark and mixture biomass input to the bioprocesses is calculated backwards from the desired production capacities. If bark demand exceeds the natural supply of the debarking process, additional biomass input for biorefining is needed. The excess bark left unused by the biorefineries will either be sold or combusted in the utility boiler depending on user input. The mill is always assumed to have access to a utility boiler for bark combustion, but replicating the opposite scenario can be made by altering certain input variables. If the process heat demand exceeds the steam supply or if the boiler is not compatible with an integrated pyrolyzer (user input), a new utility boiler investment will be made.

The biomass feed is dried to a user-specified moisture content from the initial moisture content defined in the feed parameters. A pre-defined percentage of the dryer heat is supplied with low pressure steam and the remaining demand is covered with hot water. The dryer heat demand is correlated according to the effective ambient temperature as described in chapter 2.3. The correlated performance curve is visually illustrated in the advanced input sheet and can be adjusted as desired.

The *LignoBoost* module takes a fraction of black liquor from the kraft process, extracts a fraction of the liquor lignin content and returns the slurry back to the kraft process. The amount of black liquor entering the LignoBoost process is determined by the lignin production capacity and lignin separation efficiency. The strong black liquor heating value is modified according to the changes in liquor elementary composition, which in turn is calculated from the difference between black liquor and lignin contents.

The bark flow to the *gasification* process is dependent on the designed lime kiln fuel replacement percentage. The excess bark residue can be used in other biorefinery processes if the capacity allows but due to limited supply of residue bark, this may not be the case if gasifier is already implemented. The required bark flow is directly calculated from the lime kiln heat demand as described in chapter 4.2.1, ignoring the intermediate heating value of the product gas.

The biomass input to the *pyrolysis* process is calculated with known process yields using equation 4.2. The bark fraction of total feed is determined by the user and the remaining demand is covered with the selected primary feedstock. The pyrolysis heat is taken from the utility boiler, which in turn is supplied by the char and non-condensable gas heat from the pyrolysis process. Because the allowed bark fraction of the pyrolysis feed is small (approximately 15 wt-%), the process yields and heating values are assumed to be only dependent on the primary feedstock properties. Changing the amount of mixture bark affects the mass and energy balances of the pulp mill without having an effect to the bio oil properties. Estimating the pyrolysis reactions and end product composition would also require more detailed modelling of the process.

Unlike in pyrolysis, the maximum allowed bark fraction of the *steam explosion* gross feedstock is significant (40-60 wt-%). For this reason, the black pellet heating value is calculated from the elementary composition of the selected feedstock mixture. The amount of biomass dry solids input to steam explosion process is calculated backwards from the given production capacity. The amount of needed bark and mixture biomass is calculated as a weighted average of the respective feedstock fractions. The feed moisture is added to the gross biomass input according to the data given in the feed data table.

The percentage of lime kiln fuel replacement can be set for each biofuel individually. The energy demand in lime kiln left uncovered by the biofuels will be supplied with an auxiliary fuel. Black pellets will always be sold to markets considering the motivation behind the investment decision. Combusting biofuels in the utility boiler (bark boiler) would theoretically be viable only for lignin if the recovery boiler had to be debottlenecked without the possibility of selling lignin to markets. An option for applying this kind of scenario is included to rationalize the problem for the user.

The turbine energy balance is simplified to reduce the number of needed input parameters. The downside of this is that the turbine balance becomes the weakest link of the process model, thus making process optimization impossible in this regard. The electricity production is calculated as a fixed percentage from high pressure steam energy content. With this logic, the steam and electricity balance can easily be calibrated to correspond a case pulp mill without knowing the exact steam parameters. The model assumes that this calibrating parameter stays constant regardless of the changes in medium and low pressure steam consumption. The accuracy of this method is acceptable with small changes in steam balance. Because the turbine is most likely already working near the design capacity in the reference case, this approach is justifiable. To compensate the problems caused by these approximations, the steam consumption of the biorefineries is priced according to the potential decrease in electricity production.

Equations 3.16 and 3.17 can be optionally used for more detailed analysis of the steam parameters. This would increase the model accuracy but at the same time, the amount of needed input parameters would vastly increase. For large operating point deviations the balance should be corrected using Stodola's cone rule [37, pp. 975-977] that estimates the turbine behaviour in off-design conditions. However, optimizing the turbine balance is not the focus of this kind of techno-economic analysis.

5.2.2 Economic calculation logic and approximations

The economic calculations are conducted by comparing a reference pulp mill scenario to a scenario where one or more biorefineries are implemented. The economic indicators (NPV, IRR, PI and discounted pay-back period) describe the characteristics of the biorefinery investment rather than the profitability of the pulping process. An obvious implication of this is that decreasing the reference values of the residue streams increases the profitability and therefore attractiveness of biorefining, even though the net profit of the pulping process was reduced.

The prices of fuels, process feeds, process additives and end products are expressed after principles listed in chapter 3.2. Most importantly, it is necessary to understand the difference between quoted market prices, gate prices and profit margins. Especially the end product prices may vary greatly depending on the scenario. The profit margin from pulp sales is assumed to be a known variable.

The lime kiln fuel savings are calculated directly by comparing the investment scenario to the reference case. The actual savings are calculated as the bioproduct value entering the kiln subtracted by the value of increased fossil fuel demand. The market threshold price is calculated for lignin and bio oil by equation 3.14 and can be used to link the lime kiln use to the bioproduct price.

The annual operating hours of LignoBoost and gasifier are assumed to be equal to the operating hours of the underlying pulp mill. This assumption may not be valid in reality, but it vastly simplifies the process model in scenarios where multiple biorefineries are implemented simultaneously. The operating hours for steam explosion and pyrolysis processes can be given as inputs. The model first calculates the direct operating costs and revenue of the biorefineries according to their availability. The indirect cash flows resulting from changes in the kraft process are calculated by applying downtime compensating cash flows to the maximum availability case. Basically this means that the bark usage, electricity balance and increased lime kiln fuel demand are initially calculated as if the biorefineries were available 100 % of the mill operating hours. After this, these cash flows are corrected by subtracting the difference in the process balances during the biorefinery downtime.

The operating hours denote the average operating hours with the maximum plant capacities. In the first TEA-tool build, the lime kiln fuel production and exported biofuel production share the same operating hours. For example, an availability of 70 % for a pyrolyzer would mean that the bio oil production would be completely halted during 30 % of the annual operating hours of the mill. During this downtime, bio oil would not be produced for markets nor for the lime kiln. In reality it would be more practical to operate the pyrolyzer at a constant base capacity to cover the local fuel needs and steer additional production according to the market situation.

Because the electricity balance is heavily streamlined as described in chapter 5.2.1, the cash flows have to be corrected by adding another cost component to the respective annual cash flows. The electricity production is calculated without taking the increased medium and low pressure steam into account. The error caused by this approximation is corrected by pricing the extracted process steam according to the hypothetical electricity production lo**ss**. The same principle is applied to the internal cost of bark consumption if the bark would otherwise be combusted in the utility boiler. Slight changes in the net electricity balance do not have to be considered as technical constraints if the electricity can be purchased from grid.

Although these approximations cause some inaccuracy to the individual cash flows, the net cash flow should remain fairly accurate. After all, allocating the cash flows is a matter of preference. For example, the marginal cost of bark consumption can be expressed either directly as a raw material expense or indirectly through lost steam production as a part of the electricity balance. Reallocating the cash flows more accurately would not make a difference to the ultimate results. Understanding the logic is still necessary to prevent individual cash flows from being calculated multiple times.

5.3 Visual representation and user experience

One of the focal project goals was to build an easily accessible and intuitive user interface for the TEA-tool model. Although the structure of the tool is complex, a regular user needs to access only a handful of sheets to perform comprehensive feasibility analyses. Whereas the previous descriptions have focused on under-the-hood mechanics of the model, this chapter describes the most important TEA-tool features from the user perspective. The images shown in this chapter are actual screenshots of the TEA-tool with the exception that the confidential economic data and axis scaling are hidden.

The amount of required input variables was reduced to minimum to enable easy deployment of the tool. The main input variables were expressed as self-explanatory terms that any user unfamiliar with the tool could understand. Obviously, being familiar with the applied technologies would still be necessary to use the tool efficiently.

Advanced input variables that could not be left outside of the main input options were offered in a form of drop-down lists that would correspond to certain advanced inputs. These process options consist mainly of process feeds with predefined properties.

By default, the TEA-tool opens into the primary input sheet shown in figure 5.3. This sheet works as the main tool hub, from where the biorefineries can be coupled with the underlying pulp mill model. Performing basic analyses is possible without leaving this tab. The active scenario can be named and saved at any given time for later comparison.

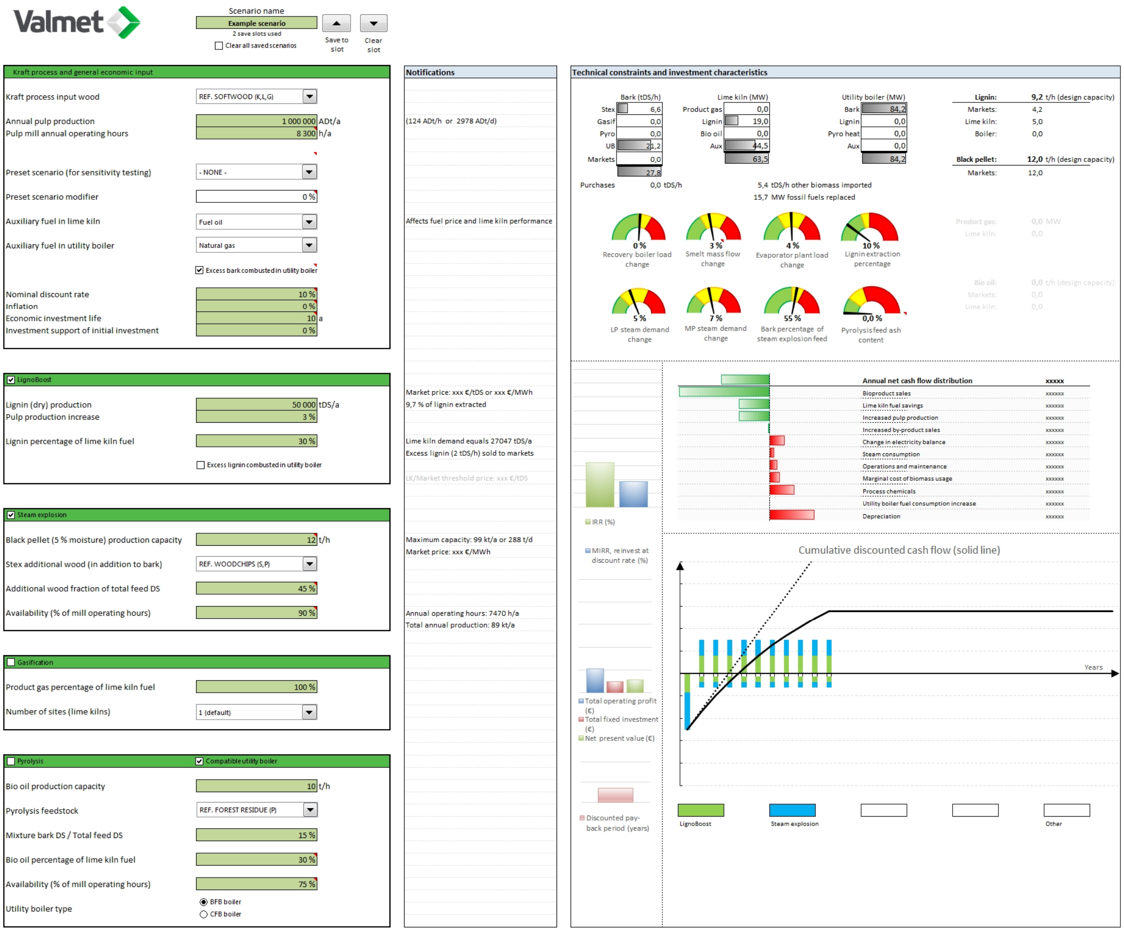


Figure 5.3 *A screenshot of the TEA-tool main input sheet. The numerical economic data and axes of the graphs are hidden from the picture due to confidentiality.*

All the basic user input is included in the left side of the main view. While making changes to these inputs, the technical and economic characteristics of the active scenario is constantly being updated to the analysis module located on the right side of the screen.

The boxes with green labels represent the different input modules that the user has to access in basic tool operation. The first input module is called the *Kraft process and general economic input* module. This input box contains the most important parameters related to the underlying mill, such as raw wood type, production capacity, annual operating hours and lime kiln fuel type. The other four input modules represent the reviewed biorefineries. Close-up screenshots of the general input module and pyrolysis process input module have been shown as examples in figure 5.4.

Figure 5.4 *A screenshot compilation of the main input modules. The example values entered to the above examples do not represent any actual scenario.*

As seen from the example input modules, the basic user input is limited to a minimal number of variables. More advanced data can be entered in the advanced input tab if needed and the feed-specific process parameters are linked to the process feed types. The figure above shows that the fuel and feedstock types can be chosen from drop-down lists that contain a selection of pre-defined feedstock options. New process feeds can be dynamically added as input options on a separate feed data tab.

All the other input variables such as investment cost data, process parameters and process performance correlations are included in the advanced input tab. This tab is not displayed in this report due to confidentiality. The market prices of bioproducts and process chemicals are assumed to be independent from the investment scenario and are therefore also included in the advanced input tab. When required, the product and fuel prices can still be changed from the main sheet using the preset scenario feature. The tool also displays the break-even prices (excluded from the screenshots) of the products for the active investment scenarios. The prices of fuel oil, natural gas and other auxiliary fuels are determined in the advanced inputs. The tool automatically adjusts the fuel prices for economic calculates according to the auxiliary fuel selections.

The user can specify the biofuel fraction of lime kiln heat input for each bioprocess separately. The tool constantly calculates the minimum required production capacity that is needed to cover this demand. This enables quick process scaling in cases where lime kiln use is the primary motivation for biorefining. Additionally, the market threshold price (equation 3.14) is displayed for lignin and bio oil to ease the decision between product exports and lime kiln use.

The biotechnology integrates can be toggled on and off from the respective check boxes located on the top left corner of the process input modules. As long as the technical constraints allow, it is possible to apply multiple simultaneous biorefineries in any desired combination. Prior to proceeding to the economic analysis, the user has to assess the technical constraints displayed in figure 5.5.

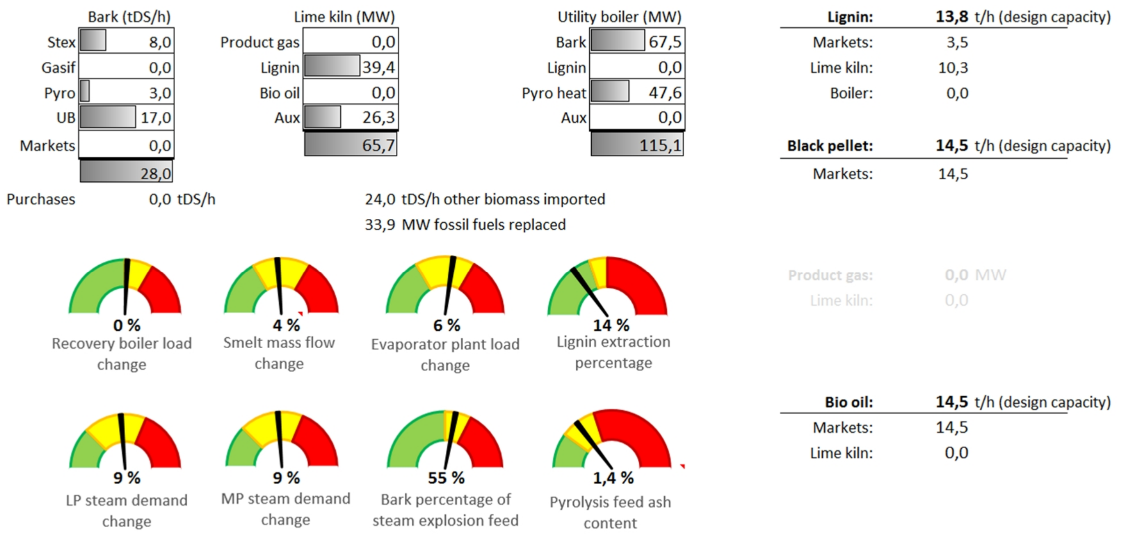


Figure 5.5 *A close-up screenshot of the technical constraint indicators. The numerical data or the constraint gauge band limits do not represent any actual scenario.*

The above technical constraint indicators give a quick overview of the most important technical constraints related to raw material availability and department capacities. The purpose of the gauges is to visualize the available headroom to the technical limitations. Expressing the changes in relative measures aims to reduce the number of needed process parameters. Generally, the yellow bands of the gauges correspond to the acceptable range of change. If any of the needles turn into the red band, the responsibility of estimating the technical limitations regarding the respective indicators changes to the user. The tool also warns about the most common process implementation conflicts with warning labels that appear next to the input options related to the conflict in hand. Examples of such conflicts would be the uncovered demand of lime kiln fuel, shortage of bark residue supply and input values beyond acceptable ranges.

The technical constraint indicators can be used to approve the active investment scenario or to find an optimum plant size by tweaking the process parameters. After the desired parameters for the biorefinery investment are found, the economic characteristic of the investment scenario can be studied in the lower part of the analysis module. The provided economic data includes visualizations from the most important economic indicators, annual cash flow distribution and cumulative cash flow.

As mentioned before, the active investment scenario can be named and saved for later comparison into one of 20 available save slots. Figure 5.6 displays screenshot from the scenario comparison tab.

Figure 5.6 *A screenshot of the TEA-tool scenario comparison sheet, showing a comparison between two saved investment scenarios. The numerical economic data is hidden due to confidentiality.*

In the scenario comparison tab shown in the above picture, the saved scenarios can be compared with each other. Two scenarios can be selected at a time for mutual comparison. The scenario comparison emphasizes the economic analysis of the saved scenarios. The visual representation of the cash flow curves and economic indicators aims to reveal common mistakes made during result interpretation. Examples of these mistakes are given in chapter 6.2.2.

In addition to possibility of saving whole investment scenarios, the input sheet also contains a so called preset scenario feature that allows testing the effects of certain input variable changes without modifying the actual input values. These preset scenarios include the most important factors affecting the profitability. The feature is intended to be used for quick sensitivity analysis. The user can select an input variable from a drop down list and increase or decrease its value by a given percentage. When the quick scenario is toggled off, the scenario resets back to the default state. The effects of the quick scenarios can be included in the saved scenario comparison.

The last user input tab is an optional feed data table that contains feed-specific input parameters. The tool automatically fetches the needed parameters from the feedstock database according to the selections made in the main input sheet. The idea of this methodology is to offer a few ready example feeds, while simultaneously enabling adding of new feedstocks.

5.4 Possible approaches for tool utilization

The TEA-tool can be deployed in a variety of scenario comparisons. The possible approaches for tool utilization can be divided into the following three categories:

1. Feasibility overview and sensitivity analysis of a single technology.
2. Feasibility comparison between different process setups of a single technology.
3. Cross-technology feasibility comparison between alternative biorefineries.

In the first approach, a single technology is toggled on to get an overview of its feasibility dynamics. In practice this means modifying the basic input variables while simultaneously observing their effects on the technical and economic output. The preset scenario feature can also be used to analyse the model's sensitivity to certain advanced inputs, such as bioproduct prices. This kind of basic analyses can be performed without knowing the exact mill parameters.

The second approach can be used to optimize a single biorefinery to a known scenario and to look for boundary conditions, under which the reviewed technology would be profitable. This can be done by generating a number of investment scenarios and saving them for scenario comparison. The scenario comparison feature is then used to compare the scenarios with each other.

The third approach aims to determine, which one of the alternative technologies is the most suitable for a given scenario. In addition to financial profit, the motivation of the biorefinery investment could be related to the technical constraints of the mill. An example of such cross-technology comparison would be the question, whether to invest in multiple gasifiers on several mill sites or to invest in a single pyrolyzer for multi-site bio oil distribution.

Outside of the intended approaches, more advanced analyses can also be performed by exploiting macros or MS Excel's built-in data analysis tools. However, presenting these methods is not relevant from the perspective of this thesis.

5.5 Customizability

The tool is built behind the idea of flexibility. In practice this means that modifying the processes, cash flows, calculation logic, and user interface should be natively supported by the tool design. The tool offers templates for modular tool parts such as correlation modules, cash flows and dynamic constraint gauges. The tool should also be capable of adapting to various special scenarios caused by the differences in operating conditions.

To enable high customization flexibility, each cash flow can be reallocated under different investments by changing a simple index. The annual cash flow function then searches for any cash flows with valid indices. New cash flow sources can be easily added and allocated to any biorefinery.

Due to the case dependent nature of the wood type effects to the kraft process, the wood type options were limited to average softwood and hardwood inputs. This decision was also based on the assumption that the mills would have more accurate process yield data available. The feed data table was constructed to support dynamic adding of new feeds and new feedstock properties.

Figure 5.7 shows the functioning logic of the *feedstock property call function*. In basic tool operation, the user only selects the feedstock from a list of pre-defined options. In advanced tool operation, the user may change the properties of existing feeds or add new entries to the feed database.

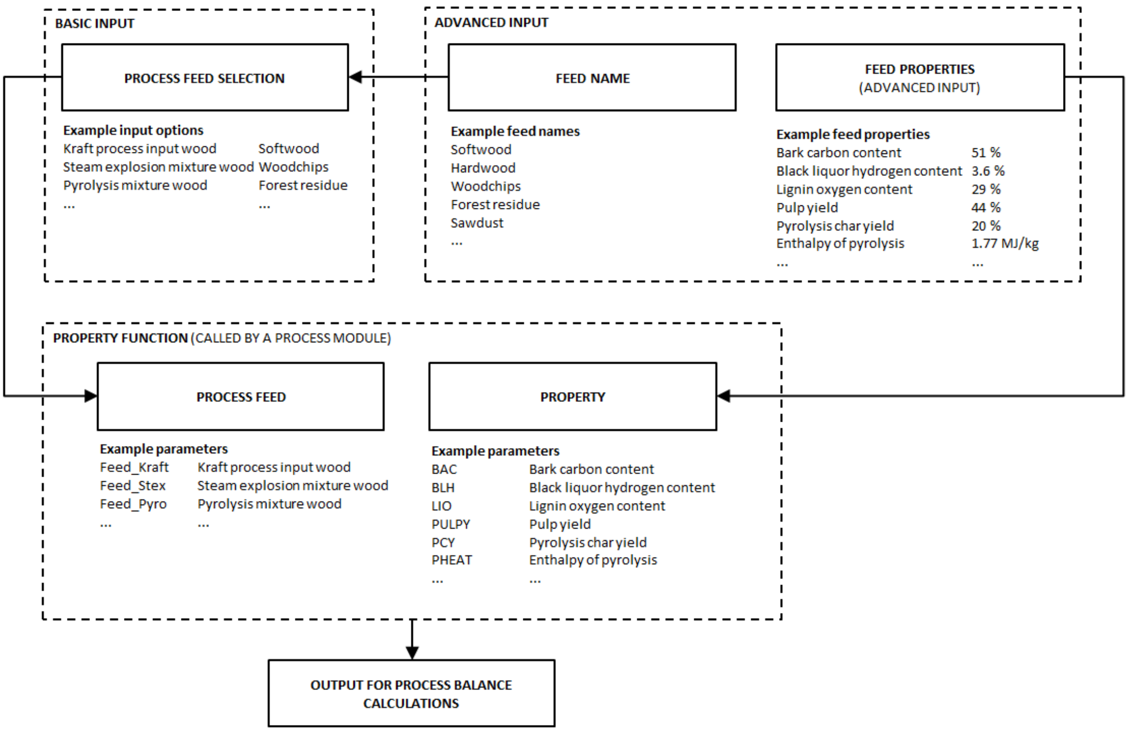


Figure 5.7 *Structure of the feedstock property call function.*

A new feedstock can be added to the database by entering a feedstock name and its properties into a new entry row. This new feedstock entry is immediately available for selection as a basic user input. Any of the specified feeds can be selected for process inputs, after which the feedstock properties can be called by the process modules. For example, the carbon content of the bark residue can be called as *property(Feed_Kraft;BAC)*, in which *Feed_Kraft* denotes the user-selected kraft process input (softwood or hardwood) and *BAC* the bark residue carbon content of the respective feedstock.

The TEA-tool is designed to be easily expandable to new biotechnologies. Adding new technologies requires some effort and proper knowledge of the tool methodology, but can be done without interfering with the existing process modules. The instructions for adding new technologies will be included in the user manual. Changing the reference process (pulp mill) into another process, for example combined heat and power (CHP) plant, is possible but would require major modifications to the tool logic. For simplicity, it would be advisable to build separate branches for the TEA-tool family for each underlying process.

6. FEASIBILITY ANALYSIS

After the building phase, the TEA-tool was used to perform a feasibility comparison between the alternative biorefineries. The purpose of the analysis is to show what the TEA-tool is capable of, while simultaneously defining the boundary conditions of profitable biorefinery investments. Providing a big picture around the investment feasibility also works as a brief introduction for future techno-economic analyses.

The following feasibility study resembles an actual analysis that could be performed with the TEA-tool. By design, an overview of the investment profitability can be studied without calibrating the model to an actual mill. To ensure the reliability of the analysis, the model sensitivity is first tested against changes in each input variable. Based on this preliminary analysis, the economic indicators are then expressed as functions of the most decisive variables.

6.1 Reference values and assumptions

The following feasibility analysis has been conducted with fixed reference values. Therefore generalizing the results for different investment scenarios should be made with caution. The results should still give a good overview of the feasibility dynamics of the reviewed biotechnologies.

The default process parameters of the pulp mill are based on an existing softwood mill. The parameters are slightly modified to ensure good resemblance of typical pulp mills and to protect the identity of the case mill. The reference mill is a softwood mill with a production capacity of 1 000 000 ADt/a and annual operating hours of 8300 h/a. The process parameters such as the black liquor and pulp yields, feedstock composition, recovery boiler balance and specific steam consumptions of different plant departments are obtained from the actual case mill.

In the initial feasibility analysis, the biorefinery processes are assumed being equal to the pulp mill operating hours. This is a reasonable assumption for LignoBoost and gasification processes. For steam explosion and pyrolysis, the effect of operating hours has to be assessed separately. Excess bark is considered being combusted in the bark boiler.

The biorefinery investments reviewed in this study are limited to practical plant sizes that correspond to 10-20 % lignin extraction with a constant recovery boiler load, 80-160 kt/a black pellet production and 60-100 MW bio oil production. In practice, these limits could be exceeded to some extent within the technical limitations. A reasonable estimate is that regardless of the case, the capacities will not exceed 25 % lignin extraction, 200 kt/a black pellet production or 150 MW bio oil production. The gasifier scale is not limited by other than the lime kiln heat demand, which is expected to vary between 50-100 MW.

The price levels of bioproducts and other assets used in the analysis are based on former feasibility evaluations performed by Valmet. The reference prices for the bioproducts are set to 20-50 €/MWh or 120-295 €/tDS for lignin, 33 €/MWh for black pellet and 60 €/MWh for bio oil. The electricity price is fixed to 40 €/MWh and the default lime kiln fuel price is expected to vary between 30-45 €/MWh. With this lime kiln fuel price range, the market threshold prices (equations 3.14 and 3.15) are approximately 160-240 €/tDS for lignin and 27-40 €/MWh for bio oil. It is important to note that the bioproduct markets contain elasticity with regards to these prices. This means that the demand - and furthermore liquidity of the markets - is dependent on the product prices.

The default discount rate has been set to 10 % and inflation to 0 %. Inflation is being ignored at this point because it does not bring any additional information to the results other than changing the real discount rate. The economic life times of the investments are set to 15 years. Changing this parameter would significantly affect to the overall profitability. If the realistic economic life time is uncertain, discounted payback period is suggested to be given the highest priority of the economic indicators.

6.2 Sensitivity of the profitability model

The purpose of the sensitivity analysis is to discover the significance of input variable changes to the TEA-tool profitability model. Rather than obtaining accurate sensitivity data, the main goal of this analysis is to find the variables that will be taken into closer examination in the final feasibility study. Identifying the non-sensitive variables is important as well to reduce the amount of workload in future analyses.

The sensitivity analysis is conducted by observing the output value behaviour when changing one of the input variables at a time. Although this method is not justifiable if the input variables are heavily correlated, this gives a good estimate on the decisive factors affecting the end results. Additionally, a selection of unspecified investment scenarios is taken into closer examination to determine, how the cash flow profile affects the information value of the economic indicators.

6.2.1 Sensitivity to technology-specific variables

The finished TEA-tool was used to analyse the sensitivity of the profitability against different input variable changes. The annual cash flows net present value and internal rate of return were observed, while changing one input variable at a time. The effect of changing these variables were expressed as sensitivity coefficients that would indicate the multiplier, at which the investment's annual net cash flow would change in relation to the change in the respective input. For example, a sensitivity coefficient of 0.5 would denote that a 20 % increase in a given input would reflect 10 % increase (or decrease) in the output.

The initial observation was that the profitability was mostly defined by the bioproduct values, production capacity, initial investment costs and marginal costs of process raw materials (bark, mixture biomass and process chemicals). The sensitivity coefficients for these variables ranged mostly in the range of 0.2-2.5. Table 6.1 shows a summary of sensitivity coefficient ranges for the most important input variables. These ranges are very rough estimates because of the vast number of complex nonlinear interdependencies between the variables and the amount of uncertainty in the operating conditions. Changing multiple variables simultaneously or changing the operating conditions could shift the sensitivity characteristics out of the provided window. This being said, the numerical data should not be used in further analyses due to high inaccuracy. The purpose of the coefficients is to express the approximate relative weights of different parameters to the model sensitivity.

Table 6.1 *Estimated sensitivity coefficient ranges of the most important input parameters.*

Variable	Technology			
	LignoBoost	Gasification	Pyrolysis	Steam explosion
Product price	1.1 - 1.4	1.2 - 1.6	1.2 - 2.2	1.5 - 2.6
Production capacity	0.5 - 1.0	1.3 - 1.4	1.3 - 1.5	1.5 - 1.7
Initial investment costs	0.1 - 0.6	0.1 - 1.0	0.2 - 1.1	0.4 - 1.5
Raw material price	0.3 - 0.5	0.1 - 0.2	0.3 - 0.5	0.3 - 0.6
Annual operating hours	-	-	1.5 - 2.5	1.5 - 2.5

The initial investment costs are automatically correlated to the plant sizes as in the TEA-tool. Plotting the economic indicators against initial investment costs would not provide much value due to the trivial consequences in the profitability figures. The feasibility analysis will be mainly performed with respect to product prices, plant sizes and operating hours. The raw material prices can be estimated rather accurately with known market prices.

In addition to the previously mentioned variables, the sensitivity was also tested against the change of steam consumption, electricity consumption, other operating and maintenance costs, feedstock moisture and dryer heat demand. It turned out that changes in these variables had no significant effects to the overall profitability. Moreover, the absolute values of sensitivity indices varied from 0.01 to 0.1 for these variables. It is important to note that these seemingly insignificant variables could still have an impact on feasibility through technical constraints.

6.2.2 Significance of the cash flow profile

Some of the profitability dynamics can be generalized to the cash flow structure of the investment regardless of the underlying technology. This results from the fact that all the necessary profitability data can be read from the cumulative discounted cash flow curve. The comparison between these anonymous cash flow curves can be made as long as the process-related variables are kept constant and only global variables are changed. The most obvious global variable is the real discount rate, which can affect the attractiveness of an investment drastically. Determining a unique discount rate may also prove to be difficult. Visualizing the impact of this choice helps prioritizing the correct profitability indicators.

The single most descriptive indicator for the shape of a cash flow profile with a given discount rate is the profitability index, which directly expresses the future returns in relation to the initial investment costs. With a fixed discount rate, the cumulative cash flow curves can be divided by their shapes into low profile ($1 < PI < 2$), symmetrical profile ($PI \approx 2$) and high profile ($PI > 2$) investments. The investments in these categories share some distinctive characteristics regardless of their absolute values.

The following analysis presents a number of scenarios that represent the characteristics of different cash flow profiles. To be able to give appropriate amount of attention to the used profitability indicators, it is important to understand how they represent a given cash flow curve. Figure 6.1 shows a reference of how the indicators can be read from the cash flow figures.

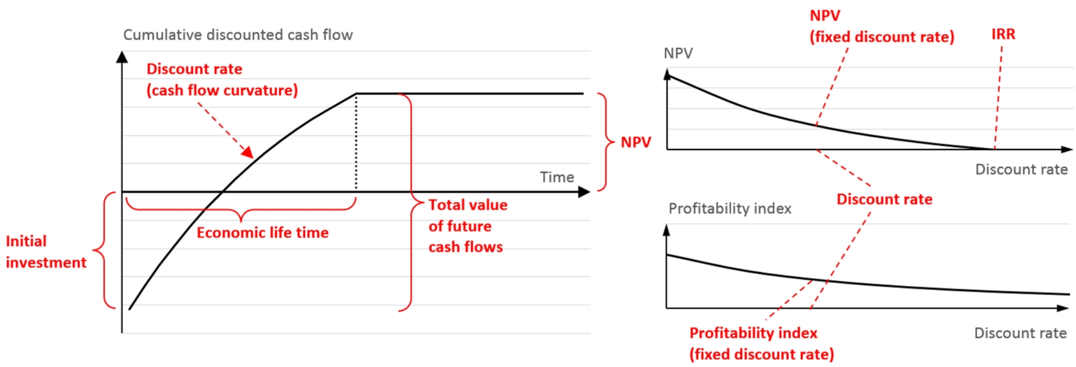


Figure 6.1 *Visual counterparts of the economic indicators displayed in the cash flow curves.*

In the above illustration, the curve on the left represents the cumulative discounted cash flow of the investment. The two smaller sub-figures on the right visualize the NPV and profitability index as a function of the discount rate. The sensitivity of the profitability against discount rate changes was studied by generating a set of example investments with different cash flow structures and analysing them with the scenario comparison feature of the tool.

Even though the internal rate of return is independent from the used discount rate, the significance of the discount rate cannot be totally ignored in the cash flow analysis. The IRR itself represents a high discount rate and may have no information value in low rate scenarios. With small interest rates, the MIRR or profitability index would give a better estimate of the investment’s actual returns compared to the conventional IRR. Using

MIRR instead of IRR would require that the reinvestment rate was identical for both scenarios regardless of the original investment's discount rate. Additionally, the investments should not have an impact on the company's WACC.

With high discount rates, the scenario with a higher internal rate of return will always end up being more profitable. This is a direct consequence of the IRR definition: Internal rate of return is the discount rate, at which the NPV equals zero. In other words, the investment with the higher IRR has a higher tolerance to growing discount rates. With low interest rates, on the other hand, IRR becomes insignificant from the profitability perspective and NPV should be given higher priority. This phenomenon can be seen from figure 6.2 that shows a comparison between two investment scenarios with intersecting NPV curves against the discount rate. The discount rate at which the NPV lines intersect is called the *crossover rate* [17, p. 350].

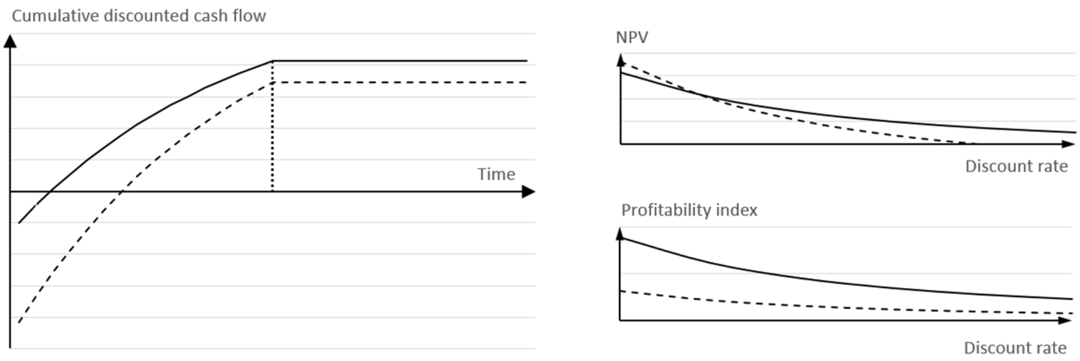


Figure 6.2 *Example comparison between investments with intersecting NPV/discount rate curves.*

In the figure above, both of the investments share the same economic life time and a discount rate of 10 %. The NPV-rate curve on the top-right corner of the figure shows that the investment with the lower IRR becomes more valuable in the lower discount rate range. In other words, lowering the interest rate below the crossover rate would change the profitability order of the investments when only NPV was examined. This kind of behaviour is most likely seen when comparing investments with similar net present values and dissimilar cash flow profiles.

The decision of prioritizing the absolute measures over the relative indicators (or vice versa) has to be made by case. Figure 6.3 displays a comparison between two investment scenarios that are seemingly equal with regards to payback time and the relative measures (IRR and PI). Because the cash flow curves are very dissimilar, the differentiation has to be made between the net present values.

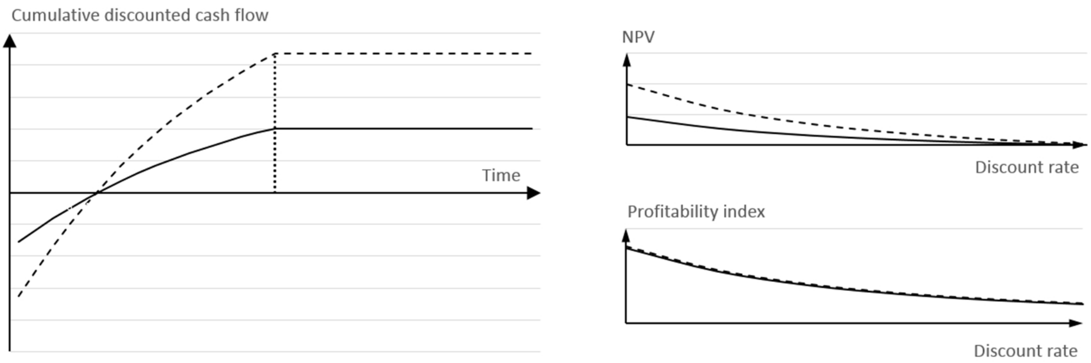


Figure 6.3 *Example comparison between investments with dissimilar NPV curves regardless of the similar IRR, PI and payback period.*

In the above scenario, the information value of the relative indicators is redundant because of similar cash flow profiles. In addition to the cumulative cash flow figure, the similarity of the cash flow profiles can also be seen from the merging profitability index curves. This would be the case regardless of the absolute cash flow sizes and therefore the net present value should be given the highest attention.

One could conclude that the NPV should always be chosen over the relative indicators. This can be easily proven wrong with an example scenario presented in figure 6.4. The scenario with the lower NPV actually dominates the feasibility by the means of IRR, profitability index and payback time.

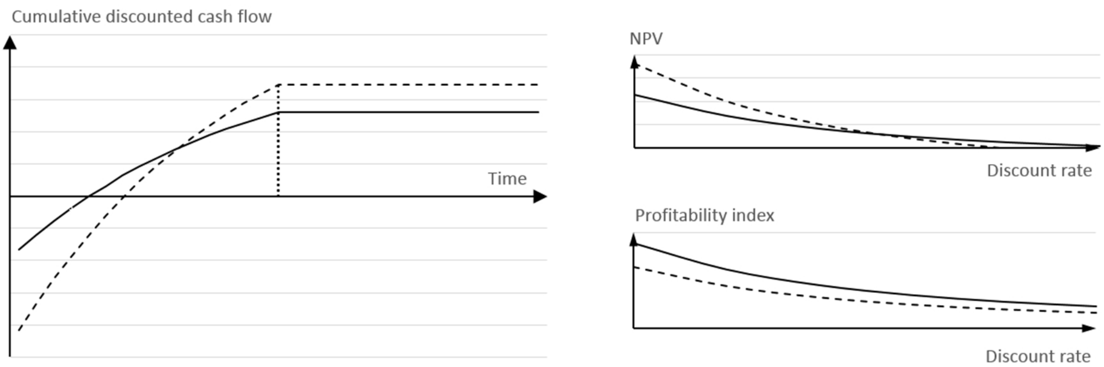


Figure 6.4 *Example comparison between investments, for which the IRR, PI and payback period are more descriptive indicators than the NPV alone.*

Choosing an incorrect investment life time may also distort the NPV and IRR of the investment scenario. This can be avoided by including the expected rate of return into the discount rate and evaluating the profitability using the discounted payback period. Even with this approach, the accuracy of the results is dependent on valid estimates of future cash flows. Relying on payback time alone would also require the chosen discount rate to be valid, containing all the rate components discussed in chapter 3.1.1.

The above examples confirm that the qualitative cash flow profile analysis cannot be bypassed in the profitability comparison. The discount rate sensitivity analysis is also a vital part of any feasibility study. The benefit of this kind of analysis becomes even more obvious when comparing investments with different life times or varying annual cash flows.

If the future cash flows cannot be forecasted with an adequate accuracy, assessing profitability through the cash flow curve is not justified. In this kind of situation, the break-even price of the products could be used as a feasibility measure instead of other economic indicators. The break-even price indicates the lowest acceptable price, at which the products have to be sold to generate profit. Like in payback period analysis, the expected profit margin can be included in the discount rate.

6.3 Feasibility comparison

The goal of the feasibility comparison is to answer to the general question, whether the biorefinery investments are feasible or not. Another goal is to define the boundary conditions for profitable biorefinery implementation with the default tool values. The feasibility comparison is divided into general feasibility evaluation of individual technologies, break-even price analysis of the bioproducts, consideration of biorefinery co-implementation and feasibility comparison under different investment strategies.

The axis scales of the cash flow figures presented in the following analyses are intentionally left out to prevent confidential cost data from being exposed. Knowing the absolute cash flow values is not necessary to understand the analysis. The reviewed scenarios still represent actual investment scenarios generated with the TEA-tool. Whenever multiple plotted curves are displayed next to each other, they are drawn with similar scaling to allow direct comparison between the figures.

6.3.1 Distribution of annual cash flows

Before advancing to the feasibility analysis, the cash flow distributions of the biorefinery investments were studied. Knowing the cash flow distribution eases process optimization and supports the results of the sensitivity analysis.

Table 6.2 shows a compilation of typical cash flow distributions for the biorefinery processes obtained from the TEA-tool output. The distributions have been normalized with regards to the gross revenue. Although the cash flow distribution is not the most convenient aid in profitability comparisons, the table suggests that the relative profits will partly overlap depending on the operating conditions.

Table 6.2 *Typical cash flow distributions of the biotechnologies with default process parameters normalized by the gross revenue (100 units of currency). The sub-components of the cash flows (grey entries) have been left out due to confidentiality.*

	LignoBoost (no production increase)	Steam explosion	Gasification	Pyrolysis
PRODUCT VALUE	60 - 75 (100)	100	100	85 - 88
Product sales				
Lime kiln fuel savings				
INCREASED MILL PRODUCTION	25 - 40 (0)	-	-	12 - 15
Increased pulp sales				
Increased by-produt sales				
Increased electricity production				
GROSS REVENUE	100	100	100	100
DEPRECIATION	11 - 15 (14 - 19)	20 - 31	10 - 17	22 - 39
STEAM AND ELECTRICITY	4 - 5 (10 - 13)	7 - 8	2 - 3	5 - 6
Electricity consumption				
Steam consumption				
RAW MATERIALS, O&M	26 - 29 (34 - 40)	26 - 29	14 - 17	31 - 38
Process chemicals				
Marginal cost of bark				
Marginal cost of other biomass				
Other O&M				
NET PROFIT	79 - 85 (33 - 46)	33 - 47	34 - 73	28 - 50

The table above has to be accompanied with a number of remarks to prevent certain misconceptions. First of all, the net profit values represent relative profit margins and are therefore not directly comparable between technologies with different absolute values. In addition, the net profit is calculated from the gross revenue and should therefore not be confused with the return on investment. The profit margins could vary significantly if the cost data or process parameters were changed from the default values. Especially the depreciation is (by definition) highly sensitive to the investment's economic life time.

The mill production increase for pyrolysis is a theoretical value that results from the assumption that the char and non-condensable gases are used in electricity production. In reality, part of this cash flow would be subtracted by the steam and electricity consumption of the process itself. This indirect notation is used only to differentiate the actual operating costs of the pyrolysis process from the net balance of the pulp mill.

Bark consumption is allocated to the raw material costs. This cash flow could also be included in the steam and electricity balance, but using this method would restrict the excess bark to be combusted in the utility boiler exclusively.

In the initial sensitivity analysis (chapter 6.2.1) it was discovered that the steam and electricity consumption only have a minor impact on the overall profitability with the reference electricity price of 40 €/MWh. Table 6.2 confirms this observation.

6.3.2 Profitability dynamics of the biotechnologies

In this chapter, the distinctive properties of each alternative investment are described. The descriptions focus on net present value and internal rate of return of the investments. Because absolute investment values are highly case-dependent, the NPV and IRR values should be interpreted cautiously.

During early testing of the tool, a few essential observations were made on the biorefinery profitability dynamics. The biorefineries appeared to offer high returns on investment and therefore pay themselves back relatively quickly. In most cases, the payback time varied between 2-5 years. The significance of this is that the biorefineries would generate positive profits with the used 10 % discount rate regardless of the chosen investment life time. Another remark was that the net present values of the investments were highly volatile with regards to the bioproduct prices.

A profitability comparison between the alternative biotechnologies has been presented in figure 6.5. The figure shows the most important economic indicators as a function of liquid market price of the bioproducts, parametrized by the production capacity. The *liquid market price* denotes the price at which the products would be assumed to be sellable regardless of the production volume. In other words, with the liquid price the operating hours would be limited by technical availability rather than product demand. In addition to the four studied technologies, a LignoBoost implementation with no pulp production increase is also displayed. The purpose of this is show the significance of the increased pulp sales in a LignoBoost investment. Expressing the product prices in relation to the energy content eases mutual comparison between the technologies.

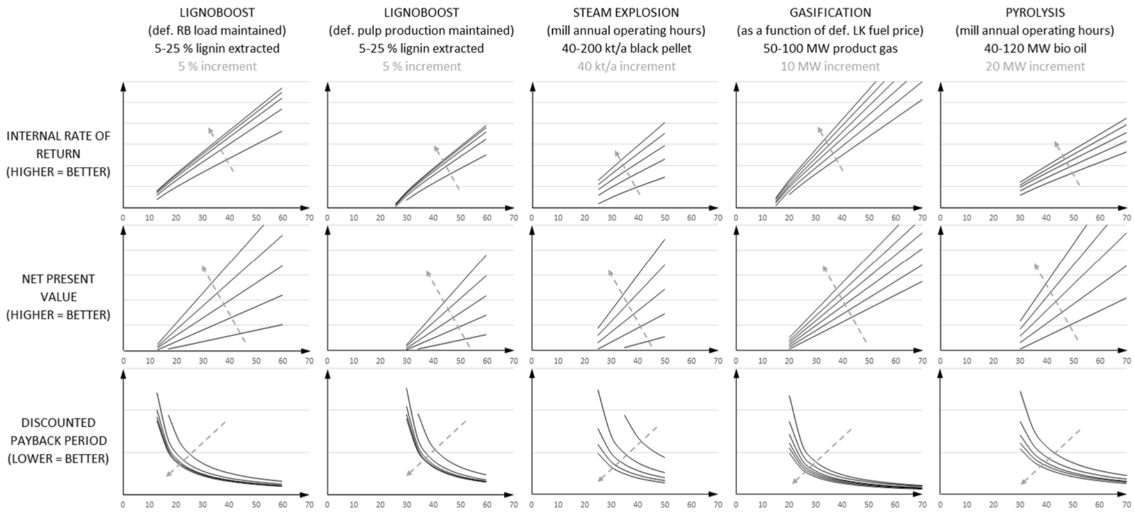


Figure 6.5 *Internal rate of return, net present value and discounted payback period as a function of liquid bioproduct prices (€/MWh) with 100 % biorefinery availability. The gasifier profitability is plotted against the default lime kiln fuel price. The vertical axis values are hidden due to confidentiality issues. The dashed arrows show the direction of capacity increase.*

Rather than examining the whole range of the curves, the profitability indicators should be analysed within the realistic price ranges of the underlying bioproducts. As long as the bioproduct prices are known, the graphs can be used to rank a number of investments to a rough profitability order.

Figure 6.5 shows that IRR and NPV of the investments are linearly dependent on the bioproduct prices in the range of practical production capacities and realistic bioproduct prices. The payback period increases rapidly with declining bioproduct prices after a certain threshold. At the same time, the net present value drops until it reaches zero value at the break-even price. When analysing the net present value, it should always be remembered that the initial investment is not explicitly visible in the figure. Increasing the initial investment cost would push the NPV curve lower and decrease the slope of the IRR curve.

The end product values in figure 6.5 represent the average income from the bioproducts, when sold product and saved lime kiln fuel is taken into account. Below the market threshold price, the low-end tails of the IRR and NPV curves would bend higher depending on the distribution between market sales and utilization in the lime kiln.

Increasing the economic life time would push the NPV-price curves up while simultaneously increasing the slope of the curves. The IRR would remain fairly unchanged with realistic plant sizes and asset prices. The payback time is independent on the economic life time by definition. In cases where the investment life time is uncertain, the payback period provides the highest information value from the three indicators.

Increasing the plant size reduces the payback time to some extent but this benefit starts to converge rapidly with large production capacities. The benefits of large production capacities are the most visible for steam explosion, for which all of the presented indicators show a substantial improvement. LignoBoost and pyrolysis do also benefit from capacity increase through increased NPV, but the IRR starts to converge with large plant sizes. The benefit of maximizing the production capacity is trivial for the gasifier that should always be scaled after the total lime kiln heat demand.

The rapid decline in profitability resulting from decreased product prices occurs around 15-25 €/MWh for lignin, 20-30 €/MWh for black pellet, 20-30 €/MWh for lime kiln fuel (gasification) and 25-35 €/MWh for pyrolysis oil. The expected margin to these prices is the lowest for black pellets, which implies a demand for larger unit sizes. However, with production capacities above 120 kt/a, the rapid lengthening of payback time is not as aggressive as for the other technologies. The margin between the actual prices and break-even prices is further analysed in chapter 6.3.3.

An interesting remark is that for LignoBoost and pyrolysis, the internal rate of return stays within a rather narrow envelope with a given product price regardless of the production capacity. On the other hand, the net present value of the investment can be heavily influenced by increasing production capacity. This means that the feasibility of these investments is mainly defined by production volume.

For LignoBoost, the pulp production increase adds a constant offset to the NPV and IRR curves. This is quite obvious as the revenue from pulp sales is independent from lignin price. The absolute value of the offset is proportional to the production capacity, which implies that decreased revenue from pulp sales brings the net present value curves closer together with a given lignin price.

Increasing the pulp production capacity provides high returns with decent profit margins from pulp sales and therefore debottlenecking can be considered being the primary function of LignoBoost. In practice when other options are available, applying LignoBoost without increasing pulp production capacity is not economically attractive. This applies especially to pulp mills where the production capacity is bottlenecked by several components in addition to the recovery boiler.

If the lignin value dropped deep below the market threshold price, increasing the production capacity would only be feasible with high marginal profits from increased pulp sales. This is a result of the limited lime kiln capacity, which bottlenecks the LignoBoost plant rather quickly in case no other use for the produced lignin was available. For a pulp mill with a production capacity of 1 000 000 ADt/a, replacing 50 % of lime kiln fuel with lignin would correspond a LignoBoost plant with a production capacity of roughly 45 000 tDS/a. This would reflect approximately 9 % lignin extraction from black liquor with no pulp production increase.

Steam exploded black pellets have a great market potential and therefore carry a relatively low market risk compared to the other reviewed bioproducts. The investment also pays itself back rather quickly even with smaller production capacities. The net return on invested capital, however, suffers critically in such cases. The black pellet production capacity has to be scaled up to 120-200 kt/a to compete with the other reviewed biorefineries. A major challenge of steam explosion implementation is the rather high medium pressure steam consumption and strict requirements for the steam parameters. The availability of process steam, however, makes the pulp mill an ideal environment for black pellet production regardless of this limitation. Whenever allowed by the technical constraints, the steam explosion process should be scaled after the largest possible unit size.

From the four alternative technologies, gasification can be considered being the easiest technology to implement. The risk of the investment is also small because of a decent profitability index and a substantially low investment cost compared to the other technologies. Additionally, upscaling the plant does not excessively increase the initial investment cost. The gasifier profitability is mostly defined by the price of default lime kiln fuel.

The drawback of a gasifier investment is the capped maximum capacity at 100 % lime kiln fuel replacement. This also means that the gasifier scale and profitability are tied to the pulp production capacity of the mill. On the other hand, the scalable nature of the process also means that the availability of bark residue is guaranteed. In fact, the percentage of bark residue required to cover the lime kiln heat demand is nearly constant for different mill sizes.

The first look at integrated pyrolysis profitability show up as very positive. The net present value of the investment competes with the best estimates of the other technologies. However, the type of bio oil plant reviewed in this study carries high uncertainty regarding the bio oil markets. The profitability of a pyrolysis plant is heavily dependent on the quoted market price of bio oil. Considering this and the relatively low profitability index, the risk of a pyrolyzer investment is arguably the highest of the four alternatives.

A closer examination also reveals that the limitation of lime kiln heat load sets the pyrolyzer investment in a difficult position: The investment starts to make acceptable profit only after the production capacity of the plant surpasses the maximum amount of lime kiln fuel replacement. The problem with this is that the profitability of a pyrolyzer investment cannot be fully decoupled from the external market situation or the bio oil price fluctuations. Maintaining positive profit would certainly be possible even with lowered bio oil prices, but the feasibility would no more compete with the other technologies. Theoretically the lime kiln bio oil use would exclusively offer acceptable profits when integrated to a large pulp mill with a production capacity of 1 500 000 ADt/a or above. Another solution would be to distribute the produced bio oil to several mill sites. In such cases, the bio oil should be valued after the actual lime kiln fuel savings rather than the market price.

The incompleteness of the bio oil markets is a serious issue because it directly limits the revenue opportunities of the pyrolysis investment. The bioproduct market liquidity is not as restrictive factor for LignoBoost that also profits from increased pulp production. Consecutively, the black pellets work as a straight substitute for other combustible fuels with existing markets and the gasifier directly increases the value of the bark residue stream. Potentially the integrated pyrolysis plant could also offer high returns but it would require better liquidity from bio-oil markets. An attractive solution would be to integrate the bio oil plant into a larger industrial ecosystem with a guaranteed price for the produced bio oil.

The above observations suggest that in an undisclosed market scenario, LignoBoost and gasification would hold the greatest potential of the four alternative technologies. The steam explosion investment competes with these technologies if the black pellet production capacity is large enough. The pyrolysis investment offers potentially high profits but is very susceptible to fluctuations in market conditions. Finding alternative markets beyond combustion applications for lignin and pyrolysis oil would ensure very high returns for both of these investments.

6.3.3 Break-even prices and profit margins of bioproduct sales

The break-even prices of the biofuels were iteratively calculated for the alternative bioproducts using the TEA-tool with an investment life time of 15 years, a discount rate of 10 % and default process parameters. The results of this analysis are shown in figure 6.6 that displays the break-even prices against the plant production capacity. For gasification, the break-even price is expressed as a substitute fuel (default lime kiln fuel) price. Except for the gasifier, the biofuel fraction of lime kiln fuel is set to 0 % to obtain unique price levels for the end products. In the figure, the operating hours are assumed to be equal to the pulp mill operating hours. The sensitivity analysis with regards to availability is presented later in this chapter.

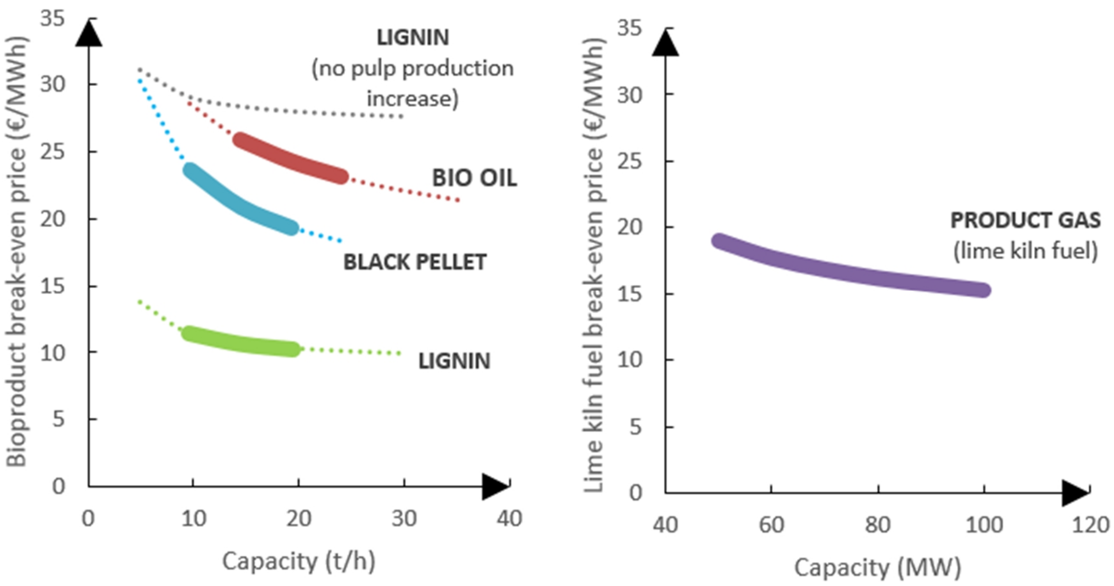


Figure 6.6 *Break-even prices (€/MWh) for bioproducts with different production capacities (10 % discount rate, 15 year investment life time and 100 % availability). The solid lines illustrate the range of practical capacity. For the gasifier, the break-even price is expressed through the default lime kiln fuel price.*

The figure above shows that with practical plant capacities, the break-even prices stay at a nearly constant level. For lignin, black pellets and bio oil, the approximate break-even prices in this range are 10.8 €/MWh, 21.3 €/MWh and 23.2 €/MWh respectively. For gasification, the break-even price of lime kiln fuel is approximately 16.8 €/MWh. In reality, the break-even prices of bio oil and black pellets would be slightly higher if the actual operating hours were taken into account.

The break-even prices represent the minimum acceptable values of the bioproducts regardless of the end use. These minimum values are would not necessarily be defined by bioproduct markets exclusively if other uses were available. The most apparent alternative use would be to combust the biofuels in the lime kiln. For the portion of biofuel used in the lime kiln, the product value has to be defined after the substitute fuel rather than market price. Given that the approximate break-even prices are clearly lower than the market prices of their respective substitute fuels, the bioproducts can be declared feasible with the reference process parameters.

Because of different end use possibilities it is challenging to find a unique market price for lignin. In combustion applications the lignin price can be estimated to vary around that of a wood fuel (20 €/MWh) but the price could rise significantly if the lignin was sold as a raw material for other products. Assuming that the recovery boiler heat load was maintained at the original level by increasing pulp production accordingly, the lignin break-even price would stay well below 15 €/MWh. However, the break-even price of lignin would be heavily dominated by the pulp production capacity. Neglecting the revenue from increased pulp production, the break-even price of lignin rises up to

28.6 €/MWh. Without alternative uses besides combustion applications, applying LignoBoost without pulp production increase would not be feasible.

Figure 6.6 shows that the benefit from plant scale increase is the largest for black pellets, for which the break-even price can be noticeably lowered within the practical capacity range. The dependency on plant scale was also identified as a key factor in chapter 6.3.2. By increasing the production capacity, the relative price margin to the expected black pellet market price of 33 €/MWh is also considerably improved. Upscaling the black pellet production to 24.2 t/h (200 kt/a) lowers the break-even price to 18.3 €/MWh.

Although the break-even price decreases for bio oil also, the relative margin to the nominally higher price cannot be significantly influenced. However, the absolute value of the bio oil production can be increased beyond the value of a black pellet plant by increasing the production volume. For a gasifier, the advantage of increasing the plant scale is obvious.

The sensitivity of bio oil and black pellet break-even prices to the plant availability was tested for different plant scales. The results of this analysis are illustrated in figure 6.7. The figure shows the break-even price as a function of annual operating hours. The reference price levels for bio oil and black pellet (chapter 6.1) are drawn as grey dashed lines. Instead of break-even prices for certain operating hours, the figure can also be used to determine the minimum acceptable operating hours for liquid bioproduct prices, at which the market demand equals the production capacity. Because determining the liquid market prices would be challenging, the formerly mentioned interpretation is more convenient.

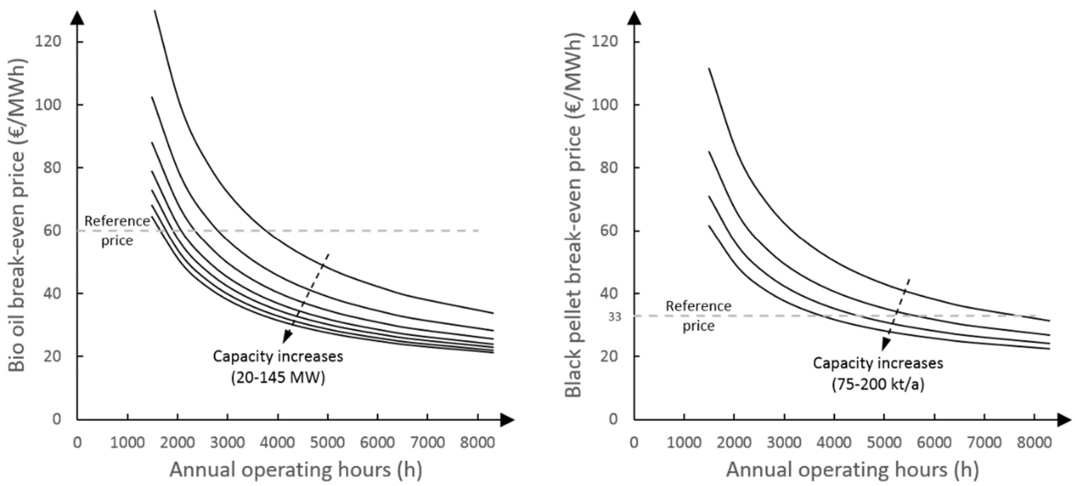


Figure 6.7 *Break-even prices of bio oil (left) and black pellet (right) as a function of annual operating hours parametrized by plant design capacity (constant capacity increment).*

The figure above suggests that with the reference product prices, the lower boundary of the allowed annual operating time would be approximately 1500 h for a pyrolysis plant and 3500 h for a steam explosion plant. These operating hours represent the zero-profit operating times for a 145 MW bio oil plant and a 200 kt/a black pellet plant. Decreasing the plant sizes from these capacities would significantly increase the minimum availability requirement.

The figures show that a black pellet plant has to operate at high operating hours to cover the product manufacturing costs. At the high-end of the operating hours range, added benefit could be achieved mainly by upscaling the plant capacity rather than further improving the plant availability. The reference black pellet price of 33 €/MWh is estimated to be the current liquid price, at which all of the produced pellets can be sold regardless of the production volume [12].

Disregarding the use in lime kiln, the annual operating hours of a bio oil plant can be assumed to be a function of market demand and therefore product price. On the other hand, the actual manufacturing costs of the bio oil are dependent on the annual operating hours. This means that the relation between the operating hours and bio oil production has to be optimized to generate the highest possible profit.

In chapter 6.3.2 it was mentioned that if a pyrolysis plant was designed to answer only to the lime kiln heat demand, the revenues would not compete with the other biotechnologies. Since the net value of the investment is very sensitive to the bio oil price (figure 6.5), the attractiveness of a bio oil plant would be defined by the actual bio oil demand and gate price. With high availability (5000-8000 h/a) the bio oil could be sold with the price of approximately 30-40 €/MWh.

It is important to keep in mind that these break-even price levels have been found with the default discount rate of 10 %, which may not represent the investment's actual growth requirement. Finding the optimum price level and production volume would require more advanced market research.

The absolute break-even prices cannot be directly compared with each other because the expected market prices are not equal. Rough estimates of the investment feasibility can be made by calculating the margin between the actual prices and break-even prices. Depending on the interpretation, this margin can be called as the *markup value* – the difference between the market price and the production cost. Another term would be the *profit margin from sales*. These relative measures should not be confused with the investment's net profit margin that would also take the annual expenses into account.

For LignoBoost, the effect of increased pulp production dominates the net profits. With lignin prices over 30 €/MWh the investment operates at a higher sales profit margin than the other investments. Limiting the increase in pulp production decreases the marginal profit dramatically. After LignoBoost, the gasifier secures the next place in feasibility ranking with a wide range of lime kiln fuel prices. The problem with this conclusion is that because the revenue is generated through fuel savings rather than sales, the gasifier profit margin has to be defined after the cheapest available lime kiln fuel. If the lime kiln has access to low cost fuel, the gasifier profit diminishes rapidly. The amount of replaceable lime kiln fuel also sets an upper limits for the gasifier size.

In favourable market conditions, the bio oil sales could net higher relative profits compared to the gasifier. This is unlikely in cases where the bio oil is sold as a substitute for fuel oil, thus being heavily correlated to the lime kiln fuel price. However, selling the bio oil for other purposes can potentially raise the price above heavy fuel oil price, making the investment highly profitable compared to the other technologies. Even then, the gasifier would provide higher profit if the default lime kiln fuel was priced above 40 €/MWh.

The steam exploded black pellets offer a steady income with a fairly predictable end product price of approximately 33 €/MWh. This price can be expected to stay relatively stable compared to the other bioproducts, since the price is closely tied to the price of the substitute fuels with already existing markets.

6.3.4 Co-implementation of multiple simultaneous biorefineries

Although simultaneous implementation of multiple biorefineries is not considered being the focal point of this study, a brief look at the alternative solutions was taken. The cash flows of the co-implemented biorefineries can be constructed by combining the cash flows of the individual investments. The technical constraints, however, have to be considered by case. Examples of such constraints would be the amount of raw materials, non-utilized potential of lime kiln fuel replacement, department capacities and available steam supply.

Because of practical issues, only one type of biofuel is considered to be combusted in the lime kiln even when multiple biofuel streams are available. This means that the lime kiln will be modified for either lignin, pyrolysis oil or biogas combustion exclusively. The rest of the heat demand is supplied with an auxiliary fuel. Another obvious reason for large replacement fuel fractions is to maximize the operating profit compared to the investment cost. These limitations imply that alternative uses besides lime kiln fuel use are needed when multiple lime kiln fuel providing biorefineries are implemented simultaneously.

The advantage of LignoBoost in this kind of co-implementation is that it is independent from the bark residue stream and therefore does not compete for the same resources. The lignin end product can also be sold to markets without interfering with the other lime kiln biofuels. Implementation of multiple bark-utilizing biorefineries would only be possible if the bark usage could be distributed among the processes in a sensible way. Integrating a pyrolyzer with a BFB bark boiler also limits the amount of available bark. This is because some of the bark has to be fed to the boiler to retain the pyrolysis heat fraction below 50 % of the gross heat load.

6.3.5 Investment strategies and motivation for biorefining

When comparing the feasibility of the biorefinery investment, it is important to define the motivation behind the investment discussion. The most obvious motivation is a long term financial profit. In addition to this, the investment might also be encouraged by a combination of other tangible goals, such as independency from fossil fuels, improved energy self-sufficiency, reduced environmental footprint, acquired market foothold or enhanced company's public image. This chapter lists a few likely investment motivators and evaluates their influence to the investment decision.

From the financial perspective, the attractiveness of the investments can be roughly divided into two categories: relative profit and absolute profit. If the company has multiple mutually comparable investment options available or if the amount of funding is limited, chasing high relative profits can usually be considered being more attractive. This means that the company wants to achieve a certain level of returns for limited available investment funds to satisfy the growth expectations of the investors. Naturally, the expected returns of the investment would have to be in line with the risk management strategy of the company. On the other hand, the company might be willing to increase the absolute profits with the expense of relative returns, especially if no other investment options were available. In the cross-technology comparison, the gasifier was identified as a high IRR investment in chapter 6.3.2 whereas the other biorefineries were categorized as NPV-driven investments.

If the biorefinery investment decision was motivated by independency of fossil fuels (lime kiln fuel replacement), the gasifier would indisputably offer the most convenient solution. Although the profitability of a gasifier investment is sensitive to the substitute fuel price, this kind of investment strategy would most likely emerge in a circumstance of high substitute fuel prices.

The pyrolysis oil plant should not be installed exclusively to lime kiln supply of a single pulp mill. The lime kiln use could exclusively be the motivation for a pyrolysis plant only if the produced bio oil could be distributed to several mill sites. Even then, implementing multiple gasifiers might end up being more profitable due to the relatively low investment costs. This issue was analysed with the TEA-tool for two and three equally sized lime kilns. As anticipated, the benefits of a pyrolyzer became more evident with small amounts of replaced lime kiln fuel: The pyrolyzer could theoretically be chosen over a gasifier for fuel replacements under 30 MW per kiln in a two-site scenario or under 40 MW per kiln in a three-site scenario.

Although lignin could be used to replace large fractions of lime kiln fuel, the decision to invest in a LignoBoost plant would also be highly dependent on the pulp production increase potential. If the LignoBoost was implemented without pulp production increase, the gasifier would generate much higher profits. On the other hand, maintaining a constant recovery boiler load (added pulp production) would value the LignoBoost slightly higher than the gasifier. In such occasion, the decision between LignoBoost and gasifier would have to be made according to the technical constraints. The advantage of a gasifier plant in this kind of comparison would be that it does not affect the pulping process balance.

Another motivation for a biorefinery investment could be to enhance the efficiency of residue bark utilization. This can be the case if the local use of bark serves no actual benefit or if bark disposal causes expenses. From the technologies reviewed in this study, gasification and steam explosion can utilize the largest amounts of residue bark. The gasifier would generally be more attractive in large pulp mills with large residue streams and high capacity lime kilns, where the practical gasifier capacity would not be limited by the lime kiln heat demand. For small capacity pulp mills, the steam explosion investment could end up being more attractive.

Ranking the biorefinery options after highest feasibility is challenging without knowing the exact operating conditions of the underlying pulp mill. The profitability dynamics appeared to be very sensitive to the bioproduct prices, which makes comparing the technologies difficult in a general level. The vast number of input variables makes comparing the technologies a complex task - especially when the bioproduct prices carry high uncertainty. Knowing the local market potential for the bioproducts would ease the task of comparing the technologies with each other. Even with accurate price estimates, the profitability would be hard to express with a single measure as each alternative investment has their distinctive advantages and disadvantages.

7. DISCUSSION AND SUGGESTIONS

This chapter aims to deliver an evaluation on the project's successfulness and to offer subjective recommendations on future TEA-tool development. The purpose is to identify the strengths and weaknesses of the TEA-tool and to openly discuss the reliability of the implemented profitability model.

After reviewing the tool, the general level of project successfulness is analysed by the degree that the project goals were achieved. The project is analysed from the standpoint of tool development and scenario analysis separately. In addition to the project evaluation, actions that are necessary to ramp up the tool use and to take out the full potential from the provided platform are discussed.

7.1 TEA-tool review

Building the tool was an iterative process and the model functionality evolved along the project's life time. The initial setting for the project allowed lots of creative freedom and approaches to choose from. The tool was constantly updated with requested features and the first build of the tool was already expanded beyond the initially planned features. Developing the tool succeeded surprisingly well considering the limited timeframe. Certain aspects of the tool could still have been made more flexible to allow even better customizing potential.

The TEA-tool proved to be a very powerful utility for quick feasibility analyses and cross-technology comparisons. The advantage of the new design was the ability to generate a wide range of alternative investment scenarios with minimal effort. The tool output also works as a great argument to support the benefits of biorefinery investments and the visual representation of the cash flow profiles helps avoiding common pitfalls in the economic evaluations. However, the ambition to include multiple technologies into the same tool caused some inevitable drawbacks.

The gradually implemented expansions induced some unintended complexity to the tool logic. For example, the operating hour limits had to be taken into account by downtime compensation (chapter 5.2.2) rather than direct calculations. Such issues were mostly caused by minor design flaws made in the early stages of tool planning. The initial principle to prioritize wishes from end users over those of administrative users also induced some structural deficiencies to the tool modularity. Expanding the tool further into new modular components would still be relatively simple, but it would require knowledge of certain hardcoded features. In other words, some features are unable to dynamically adjust to changes in the tool logic and would therefore have to be modified manually.

The cash flow model of the tool is well defined and easily approachable. The accuracy of the results is obviously still very sensitive to the reliability of the provided price data. The user's responsibility of entering correct input data is especially important for asset prices, because the TEA-tool does not evaluate the credibility of the implemented market scenarios.

The weakest link of the implemented TEA-tool model is the steam and electricity balance that has been heavily streamlined. The effects to the ultimate profitability figures are minor, but assessing technical constraints is unreliable in this regard. The tool is also lacking the ability to assess environmental impacts and valuing the sustainability of the solutions.

The accuracy of the process model suffers especially if the mill balance shifts far from the reference operating point. However, the accuracy of the technical constraints should be adequate with mass and energy flow changes below 10 %. This limit is most likely not exceeded with practical plant capacities and typical process parameters.

Being aware of the model limitations and calculation logic, the TEA-tool can be deployed in a wide range of special scenarios and market conditions. The reference parameters of the tool could be used for a quick analysis to visualize the significance of different input parameters to the investment profitability. The reference values could also be used in preliminary technology reviews without calibrating the model to an actual mill. A detailed case study would require more reliable and up-to-date input data.

7.2 Reliability of the initial feasibility analysis

The feasibility analysis presented in this thesis was conducted using fixed reference values. To overcome the problem of uncertain input variables, the economic indicators were plotted in relation to the most decisive factors affecting the profitability. This means that the responsibility of reading the profitability curves from valid operating points is given for the reader. With the amount of classified data included in the inputs and outputs, publishing more accurate profitability estimates would not be possible.

The feasibility estimates are the most accurate for steam explosion and gasification investments. This results from the simplicity of the processes and the low amount of uncertain variables. On the contrary, the profitability dynamics of LignoBoost and pyrolysis processes are more complex and only represent the particular scenarios.

The advanced tool input parameters not included in the feasibility comparison are meant for detailed case studies. Examples of these variables would be the process yields, feedstock compositions and process additive consumptions. Altering the advanced variables within reasonable limits would not change the annual cash flows by more than 10 %.

Although the pulp mill model is heavily simplified, it can be accurately calibrated with regards to the addressed process parameters. The mill parameters of the model are chosen to represent a typical pulp mill, and the reference mill could be upscaled or downscaled without breaking the model integrity. The sensitivity analysis also showed that the mill parameters play a minor role in the financial profitability. This means that the primary function of the pulp mill model is to outline the technical boundaries for practical biorefinery sizes.

Performing the study for a pulp mill with a different production capacity would not cause a significant change in the profitability figures, as long as the technical limits were not exceeded. This is because most of the revenue generated by the biotechnologies are related to the biorefinery capacity rather than the mill capacity. For example, a 60 MW gasifier would have the same cash flow curve regardless of the pulp mill size as long as raw material supply and lime kiln heat demand were sufficient.

An exception for the above rule is the LignoBoost investment that also affects the pulp production capacity. The effect of this production increase causes a constant offset to the annual cash flows. The feasibility analysis supposed that the recovery boiler heat load would be kept constant by compensating the reduced black liquor energy content by increasing the overall pulp production accordingly. This may not be the case in an actual investment scenario, in which the recovery boiler could have more headroom for pulp production increase.

7.3 Project successfulness

The project consisted of developing the TEA-tool, utilizing the tool in the initial feasibility evaluation and finally compiling this thesis report. Building the tool should be considered being the main focus of the project, while the written documents are meant to support ramping up the tool deployment. From Valmet’s point of view, the tool implementation can be thought of as the main goal of the project. The project successfulness is evaluated from these aspects separately.

The amount of confidential input data slightly restricted the scope of the general feasibility analysis that did not entirely manage to declare the absolute most profitable biotechnology investments. This was mostly caused by the partially overlapping profitability figures and uncertainty in input variables. The TEA-tool would provide more valuable results in a specific sales scenario with known market conditions. Decisive factors affecting the feasibility were still discovered, which can be considered being the main academic motivation for the initial feasibility study.

Regarding the project goals discussed in chapter 1.2, building the tool can be considered being successful. The general reception of the TEA-tool and its way of representing results was positive. The functionality of the tool was also extended far beyond the initially planned features.

The ability to compare competitiveness of alternative biotechnologies was achieved. The way of presenting the profitability through cash flow curves and basic economic indicators turned out to be well suitable for cross-technology comparisons. The all-in-one approach of the tool allows unbiased comparisons between the alternative technologies, as the level of detail is similar for all the included investment options. In addition to the implemented biorefineries, the tool can later be expanded to other technologies as well.

The accessibility to techno-economic analyses was significantly improved when compared to the previously built calculation templates. Essentially, performing basic analyses with the tool is possible for users with different backgrounds. The tool works for quick feasibility evaluations but is also suitable for comprehensive case studies.

The graphical appearance of the tool would still need some polishing depending on the way the tool is ultimately used. However, the user interface managed to reach the level that was demanded in this stage of tool development.

7.4 Suggestions regarding the future

The next evident step in the TEA-tool lifecycle would be to deploy the tool to support an actual sales project. This would test the tool functionality in practice and add a natural motivation for model calibration. Additionally, this could also reveal potential shortcomings of the profitability model, thus giving insight to the next required tool modifications.

As initially planned, the TEA-tool should be incrementally expanded to cover additional biotechnologies to improve the comprehensiveness of the analysis. In an ideal situation, the biorefineries implemented in the tool would represent the whole offering portfolio of the company. Modifying the under-the-hood mechanics of the tool should be made cautiously to maintain the original level of customizability. Improving the accuracy of the correlations and feedstock properties is still recommended.

To ensure reliability of the TEA-tool model, the investment cost and asset price data should be constantly updated. The imminent consequence of updating the tool is the need for strict version control. Limiting the tool distribution in the first phases of tool deployment is necessary to prevent outdated tool versions from spreading excessively. In its current state, the tool could be used by sales representatives either directly or by extracting the results of the analysis to separate marketing material.

If the tool was later intended to allow customer interaction, building an external user interface would be necessary. A recommended solution would be to build a web browser based interface that could be calibrated and updated remotely. This would improve accessibility, ease version control and eliminate compatibility issues that would be typical for certain standalone platforms.

8. CONCLUSIONS

The global trend of investing in sustainable energy solutions is growing. This creates incentives for developing methods for reducing consumption of non-renewable raw materials. One way of accomplishing this is to invest in biorefining technologies. Because the business for commercial scale biorefining is still relatively undeveloped, it is challenging to assess the feasibility of these solutions. Extending the methods of techno-economic studies benefits the biotechnology suppliers and their customers by improving understanding of the technical limitations and profitability dynamics of the investments.

A kraft pulp mill forms an ideal environment for integrated biorefining due to the availability of residue biomasses and access to low-cost process heat. Although the pulping process utilizes majority of the arriving feedstock with high efficiency, the residue biomasses usually still have lots of unused refining potential. The general idea of integrated biorefining is to improve the efficiency of raw material utilization and to increase the value of the mill's total production portfolio.

The main purpose of this thesis project was to build a techno-economic tool (TEA-tool) that could be used to compare the attractiveness of four alternative biorefinery integrates from the standpoint of a pulp mill customer. The biotechnologies included in the tool were lignin production by LignoBoost, black pellet production by steam explosion, bark gasification and bio oil production by integrated fast pyrolysis. The pulp mill balance model of the new tool was based on an earlier tool for the LignoBoost process.

Developing the TEA-tool succeeded surprisingly well considering the broadness of the subject. The reception of the tool was also very positive. The tool managed to vastly improve the approachability of the techno-economic debate in Valmet by bringing deeper understanding of company's own biotechnology offering. In the future, the flexible tool will allow easy access to a wide range of feasibility analyses without demanding extensive effort from the user.

The finished TEA-tool was used to perform a general feasibility comparison between the investment options. The purpose of the initial feasibility study was to provide a big picture around the biorefinery investments to work as an introduction for early tool deployment. The scope of the analysis was restricted to outlining the profitability boundaries due to confidentiality of certain process parameters and economic data. Despite of this limitation, the analysis managed to capture the core essence of profitability dynamics of the reviewed technologies.

The revenue generated by the studied processes consists of bioproduct sales and savings in local fuel consumption. Additionally, the LignoBoost process can be used to increase the pulp production capacity of the mill. The reviewed technologies were discovered being feasible in the reference market conditions, which boosts confidence in biorefining. The sensitivity analysis also revealed that the profitability of the reviewed technologies would be mostly dependent on bioproduct prices, raw material prices, production capacities and investment costs. These results are not particularly surprising. A more significant outcome of the sensitivity analysis is that with the reference product prices, the profitability would be relatively stable with regards to other input variables. The effects of discount rate and investment's economic life time are obvious.

The attractiveness of a LignoBoost investment was observed to be heavily dependent on the pulp production increase. Implementing the LignoBoost process without pulp production increase could theoretically generate positive profits, but the competitiveness would suffer compared to the alternative investment options. As long as the technical constraints allowed, the LignoBoost process appeared to generate substantial profit. The LignoBoost investment would also benefit from the independency from residue biomass availability.

The gasification process stood out from the compared technologies with high relative returns. This was a result of low investment costs and the ability to add significant value to the mill's bark residue stream. The impact caused by the bark gasifier to the pulp mill balance is minor. The downside of the gasifier investment is that the product gas can only be used locally as a substitute fuel, thus reducing the profitability in cases where other inexpensive fuels are available. The gasifier size is also limited by the lime kiln heat demand.

The integrated pyrolysis was identified being an investment with high potential value. The problem with integrated pyrolysis is that the markets of bio oil are still undeveloped. This reduces the overall attractiveness of the investment. The possibility of distributing bio oil to multiple mill sites would improve the attractiveness of the pyrolyzer investment as the dependency on external markets would be reduced.

The steam explosion investment showed to be competent with the other investment options as long as the production capacity was maximized. The plant size would be constrained by availability of raw materials and medium pressure steam. The main advantages of black pellet production are the predictable cash flows and a high potential in market penetration.

Ranking the biotechnologies into a strict profitability order serves no actual purpose in a rough feasibility comparison. The market demand for the bioproducts is also challenging to forecast without conducting an extensive market research. This causes immense volatility to the possible outcomes in feasibility. On the other hand, majority of the investment scenarios showed to be profitable even with slightly pessimistic estimates on the market demand. The ability to identify this kind of characteristics of the biorefinery investments will definitely offer invaluable support on Valmet's future techno-economic analyses.

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