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Abstract

Conventionally, tall oil soap obtained as a side product from the Finnish pulp mills have been refined to crude tall oil, which is sold to tall oil distillers. In distillation, tall oil pitch is first removed, and the rest of tall oil is fractionated into value added products. Tall oil pitch is mainly used as a fuel and it is often transported back to pulp mill to be used as a lime kiln fuel.

The aim of this thesis was to evaluate the feasibility of depitching tall oil in a kraft pulp mill. The hypothesis was that the energy-intensive depitching process can be performed efficiently by utilizing pulp mill heat generation. The process produces both a higher value tall oil product and renewable fuel for the lime kiln simultaneously.

The pitch separation was modelled and the process energy consumption in model was used to calculate the mill energy balance and production. Economic feasibility of the concept was evaluated through investment and operating costs, which were used to calculate investment payback period. These key values were compared to scenarios with other lime kiln fuel production methods: Heavy fuel oil, bark gas, and both tall oil pitch and bark gas.

Tall oil depitching was concluded to be feasible with short payback period and good applicability to pulp mill. The major technical drawback of the process is that it is not sufficient to provide all of the lime kiln fuel needed. In the future strategies, the depitching could be optimised for phytosterols separation and for the production of different kind of fractions based on market price fluctuation.

Keywords Tall oil pitch, Depitching, Renewable biofuels

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Tiivistelmä

Perinteisesti suomalaisilla sellutehtailla on ollut tapana jalostaa sellunkeiton sivutuotteena syntyvä suopasaippua raakamäntyöljyksi, joka myydään mäntyöljytislaajalle. Mäntyöljytislaamolla raakamäntyöljystä erotetaan ensin piki, ja loppu tislataan korkeamman arvon tuotteiksi. Mäntyöljypiki hyödynnetään lähinnä polttoaineena. Usein se rahdataan takaisin sellutehtaalle meesauunin polttoaineeksi.

Tämän työn tarkoituksena oli tutkia mäntyöljyjen erotuksen kannattavuutta sulfaattisellutehtaalla. Oletuksena oli, että energiaintensiivinen pikierotus voidaan tehdä tehokkaasti sellutehtaan lämmöntuotantoa hyödyntäen. Samalla tuotetaan sekä korkeamman arvon mäntyöljytuotetta, että uusiutuvaa polttoainetta sellutehtaan meesauunille.

Pien erotusprosessi mallinnettiin ja sen lämmönkulutus yhdistettiin koko tehtaan energiataseeseen. Konseptin taloudellista kannattavuutta arvioitiin laskemalla prosessille investointi- ja käyttökustannukset, sekä takaisinmaksuaika. Näitä tunnuslukuja verrattiin kilpaileviin meesauunin polttoaineen tuotantomenetelmiin: raskaaseen polttoöljyyn, kuoren kaasutukseen sekä kaasutuksen ja pienerotuksen yhdistelmään.

Sellutehtaan mäntyöljyjen erotus todettiin soveltamiskelpoiseksi prosessiksi, jolla on lyhyt takaisinmaksuaika. Suurin tekninen haaste prosessissa on se, ettei sillä pysty tuottamaan kaikkea meesauunin tarvitsemää polttoainetta. Tulevaisuudessa pienerotusta voitaisiin optimoida kasvisterolien erotukseen sekä erilaisten jakeiden tuotantoon markkinahintojen vaihtelun mukaan.

Avainsanat Mäntyöljypiki, pien erotus, uusiutuvat polttoaineet

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Espoo, 27th June 2018

Mikko Niemeläinen

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ABBREVIATIONS

ADt	air dry tonne (in 90% dryness)
CaO	calcium oxide
CSS	crude sulphate soap
CTO	crude tall oil
DCTO	depitched crude tall oil
DTO	distilled tall oil
FAME	fatty acid methyl ester
FFE	falling film evaporator
HDS	Hydro Dynamic Separation
HFO	heavy fuel oil
HVO	hydrotreated vegetable oil
RFS	Renewable Fuel Standard
TFE	thin film evaporator
TLO	tall light oil
TOFA	tall oil fatty acids
TOP	tall oil pitch
TOR	tall oil rosin

SYMBOLS

\dot{m}_W	mass flow of evaporated water
ΔH_W	enthalpy change of evaporated water
\dot{m}_B	mass flow of bark into a gasifier
ΔH_B	enthalpy change of bark during drying
Q	heat demand
η	efficiency of heat transfer system

1 Introduction

Several global megatrends, such as increased environmental awareness and resource efficiency are driving the traditional pulp industry towards biorefinery concepts. These concepts have two main principles which are also economically motivated: the elimination of fossil fuels usage during operation and new value added side products for pulp. Most of the interests are directed towards lignin and hemicellulose, but this thesis focuses on the third side stream: the extractives.

The recovery of extractives from pulp production is not a new idea. On the contrary, crude tall oil can be considered as the first remarkable side product of kraft pulp production, which recovery began already in the beginning of the 20th century. There is already an existing refining chain and markets for the products derived from it. The maturity of this segment has decreased the interest towards the development of tall oil processes in pulp mills.

However, the growth of production in pulp mills is offering larger quantities of crude tall oil and further growth to its production can be achieved through process optimisation. Simultaneously, the other principle of biorefineries is still mainly unreached: A majority of lime kilns in the pulp industry is still fuelled with fossil fuels such as natural gas or heavy fuel oil. On the tall oil refining sector, the production of biofuels has become a competing crude tall oil user to traditional distillation, and interests towards phytosterols refining for medical applications are growing. All these factors are opening new possibilities for a more efficient production of tall oil products.

In this thesis, possibilities to increase the efficiency of the tall oil refining chain are investigated from the pulp mill point of view. In the Literature Review, the processes and chemistry of tall oil refining are introduced. The market situation and potential new applications are also briefly viewed. In the Experimental Study, a concept study to further refine crude tall oil in a kraft pulp mill is made and its feasibility is defined through a mill balance comparison between competing scenarios. The process studied in the experimental part is depitching, which creates more valuable crude tall oil product and helps pulp mills to reach carbon neutral production.

LITERATURE REVIEW

2 Tall oil

Tall oil is the group of mostly lipophilic, non-volatile extractives (Figure 1) of wood that are separated from spent sulphate cooking liquors. It is considered as the most important side product of the kraft pulping process in softwood and mixed wood pulp mills (Isotalo, 1996). After refining and fractionation, tall oil offers several base chemicals for the process industry.

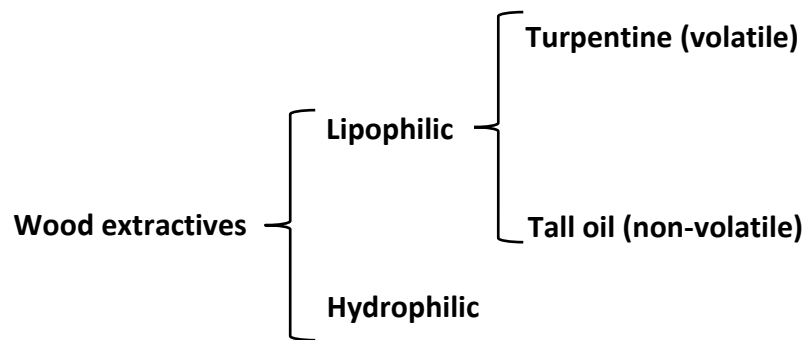


Figure 1. Tall oil as a fraction of wood extractives.

In the Literature Review, the refining of tall oil is first introduced, followed by the characterisation of the end products and their applications. After these, the chemistry of tall oil fractions is further presented. The Literature Review concludes by analysing current trends and future possibilities for tall oil utilisation in more detail.

2.1 Tall oil refining

In this section, the refining and fractionation processes of tall oil are introduced. The thesis focuses on the process chain relevant to the tall oil industry, which begins from soap separation in the evaporation plant and ends in tall oil fractional distillation. This

selection is justified as the earlier wood-handling and cooking processes are optimised for and driven mainly by pulp production, and the further refining of tall oil fractionation products is not selective to only tall oil -based feedstock.

2.1.1 Soap separation from black liquor

During kraft cooking, acidic lipophilic extractives of wood chips are saponified with the sodium hydroxide in the white liquor to form organic sodium salts, as shown in Figure 2. These compounds with hydrophobic and hydrophilic ends tend to form soap micelles when a critical concentration of these particles is achieved, in other words during black liquor evaporation. Both fatty and resin acid salts are needed in order to form micelles (Laxén & Tikka, 2008). When given time to settle, they form a separate phase on the surface of the black liquor, which can be separated with overflow decantation or by skimming with a scraper.

Even though all lipophilic components are not capable of forming soap, they may be incorporated into the micelles, which explains the unsaponifiable content of tall oil (Ödberg et al, 1985). During skimming, black liquor, pulp fibres, lignin and small amount of inorganics, such as calcium compounds, are also retained alongside the soap. Important parameters in soap separation are the fatty acid and rosin acid ratio in soap, ionic strength (residual effective alkali), temperature and the dry matter content of the black liquor, and the turbidity and superficial flow in the soap skimming tanks (Foran, 1983; Ström et al, 1990). The optimal dry matter content of black liquor for soap separation is 28–32 w-%. However, evaporator plants are constructed in such a way that soap is usually skimmed from feed liquor and intermediate black liquor tanks, where the dry matter content is 18–21 and 30–35% (Laxén & Tikka, 2008).

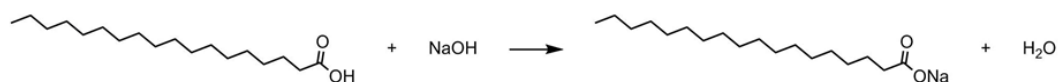


Figure 2. Saponification reaction.

After separation from black liquor, the crude sulphate soap (CSS) is stored in a soap tank. Black liquor tends to further separate from soap during storing, and its removal from the bottom of the tank is beneficial for the quality of the soap (Laxén & Tikka, 2008).

2.1.2 Acidulation of soap

In acidulation, resin and fatty acids are liberated from CSS with the help of acid, which can be sodium bisulphite, sulphuric acid, spent acid from chlorine dioxide production, or a combination of these and carbon dioxide (Johnson & Potier, 1997; Laxén & Tikka, 2008). Acidulation reaction is presented in Figure 3.

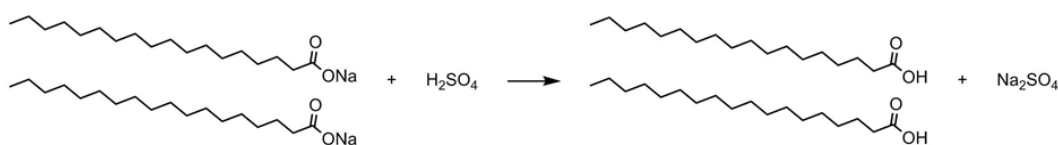


Figure 3. Acidulation of soap.

During acidulation, spent acid, also called brine, lignin and solid gypsum (calcium sulphate formed between acid and calcium compounds) form separate phases from crude tall oil, as shown in Figure 4. Even though phase separation provides an easy way to remove most of the unwanted soap fractions, especially the lignin layer tends to slow down the settling and capture some of the crude tall oil (CTO).

A favourable acidulation reaction temperature is around 98 °C (Foran, 2017). It is beneficial to wash the soap with water before acidulation to lower the residual black liquor concentration and to dissolve impurities from soap. Black liquor and impurities removal reduces acid consumption, improves the quality of the end products and protects the acid content of the soap from esterification reactions catalysed by strong acid (Johansson, 1983). The esterification is further discussed in section 2.1.3.



Figure 4. Phase separation of acidulated tall oil soap, as presented by Foran (2017).

There are several decantation and centrifugation processes used to perform soap acidulation. Traditionally, this acidulation is operated as a batch process, presented in Figure 5. In the most traditional batch systems, hot water is pumped to a reactor followed by sulphuric acid. This mixture is further heated with steam to a final reaction temperature. Crude sulphate soap is added and thoroughly mixed into the solution, and the mix is allowed to react and settle for about 2–3 hours. More acid may be added during reaction to adjust the pH. After settling, the oil phase is separated into a wet oil tank, where it is let to further settle to remove residual brine. Spent acid and solids may be removed from the reactor or left there for following runs, until sequential clean up. In traditional batch systems, lignin precipitates between the oil and spent acid layer, weakening the phase separation. To avoid this, they may be upgraded to semi-batch systems by performing phase separation in a shaker-decanter combination. With this setting, the size of the acidulation reactor may also be significantly reduced (Drew & Propst, 1981).

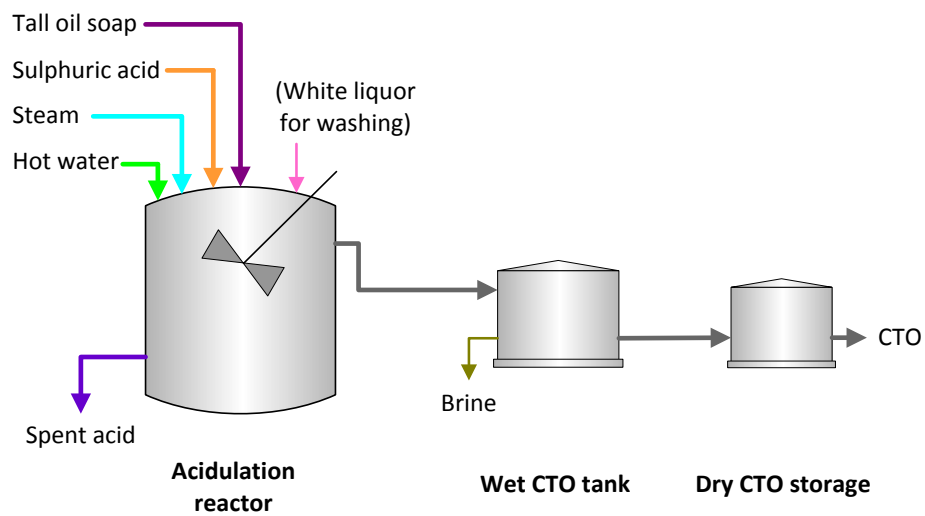


Figure 5. Batch process for soap acidulation, adapted from Drew and Propst (1981).

Hydro Dynamic Separation (HDS), presented in Figure 6, is an example of a continuous decantation process. Soap, acid and water are mixed in-line and fed to a reaction zone of a decanter cone. To avoid the formation of harmful lignin layer, brine is circulated between middle and bottom part of the decanter, and part of the lignin is separated through an inner cone. Colloidal particle removal is further enhanced by lamella sets in the upper section of the decanter. CTO is removed from the decanter through an overflow (Lindqvist, 1983).

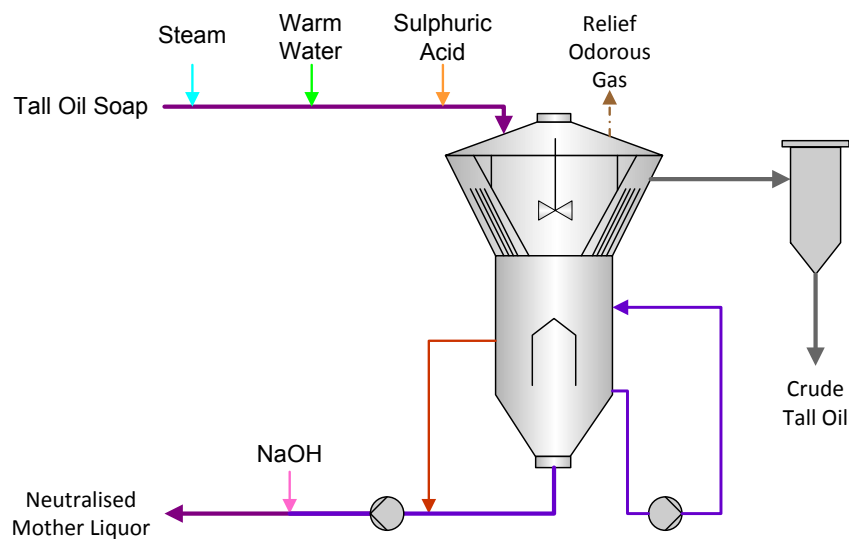


Figure 6. Simplified Hydro Dynamic Separation (HDS) process for soap acidulation, adapted from Laxén & Tikka (2008).

Other option for continuous soap acidulation is to use centrifugation, as shown in Figure 7. Centrifuge rotor exerts a large force to acidulation mixture, enabling fast phase separation and slightly higher yields than other conventional methods (Foran, 2017). However, compared to decantation and batch acidulation, centrifugation equipment needs more mechanical maintenance and is more sensitive to the specific gravity variation of feed fractions. To avoid plugging by solids, lignin can be pre-extracted with a solid bowl centrifuge. The final separation is usually done with a conical plate centrifuge, as shown in Figure 7 (Head Engineering AB, 2011).

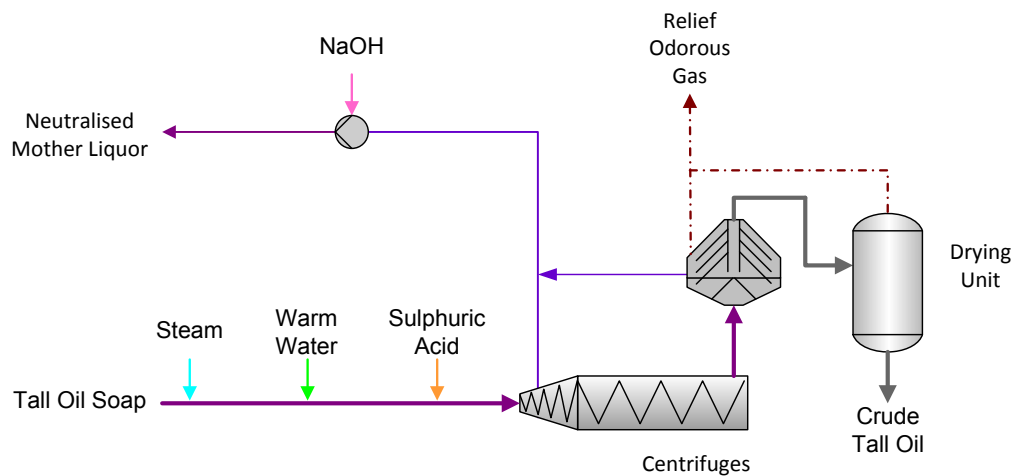


Figure 7. Simplified Pinola TOPP process as an example of centrifugal soap acidulation, adapted from Head Engineering AB (2011).

After acidulation, CTO is vacuum-dried and pumped into a heated storage tank to wait for transportation to a distillation site. During storage, residual solids may sediment onto the bottom of the tank. This sediment is often removed and burned in a recovery boiler on a pulp mill.

2.1.3 Drying and depitching

Before entering distillation process, CTO has been affected by storing and transfer that alter its initial composition. Considering the quality, the most important reaction happening is the esterification of acids with alcohols, forming esters and water, as presented in Figure 8. Esters lower the yield of desired free acids, while water

deteriorates the energy efficiency of distillation. CTO may be additionally dried to increase the stability of vacuum distillation.

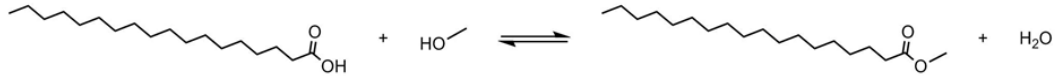


Figure 8. Exemplary esterification reaction between a fatty acid and methanol.

Aim of the depitching is to separate desired acid components from unwanted neutrals through vaporisation. Remaining non-volatile liquid, the tall oil pitch (TOP), is retained from the process. Evaporated components may be fed directly to distillation columns. As all the desired end products are vaporised, process conditions in depitching are rough: Process is operated in vacuum at absolute pressures below 1 kPa and at temperatures up to 270 °C (Huibers, 2000; Norlin, 2012). Lower pressures can be used to lower the operating temperature. Depitching is done with two different systems. The more traditional one is to separate TOP in a stripper column by heating tall oil with direct superheated steam. Steam and volatilised matter is directed onwards to the rosin distillation from the top of the column. Depitching stripper is presented in Figure 11 in section 2.1.4.

Another depitching method closely related to the dry distillation method is to use thin film evaporators. They spread tall oil over a heat exchange surface with rotating blades. The volatile matter is vaporised from the heat exchange surfaces. Thin film evaporator as depitcher is presented in Figure 10 in section 2.1.4. Short residence time is mentioned to be the advantage of the thin film evaporation strategy, whereas good heat transfer is the advantage of the distillation stripper. Drawbacks are the large overall energy consumption and effluent generation of the stripper. The weakness of thin film evaporator is the higher need of mechanical maintenance (Bress, 1982; Freese & Vock, 1982).

2.1.4 Distillation

Distillation is the final tall oil fractionation process, in which depitched tall oil is divided into commercial end products: tall oil rosin (TOR), tall oil fatty acids (TOFA) and distilled tall oil (DTO). In the process, also the most volatile neutrals and fatty acids are separated as a stream of their own, called tall light oil (TLO) or heads. Product streams are presented in Figure 9. Due to high boiling points of the components, distillation is operated in vacuum and at operating temperatures over 200 °C (Norlin, 2012). Regardless of the distillation method, modern distillation system consists of three distillation columns: a rosin column, a heads column and a fatty acid column.

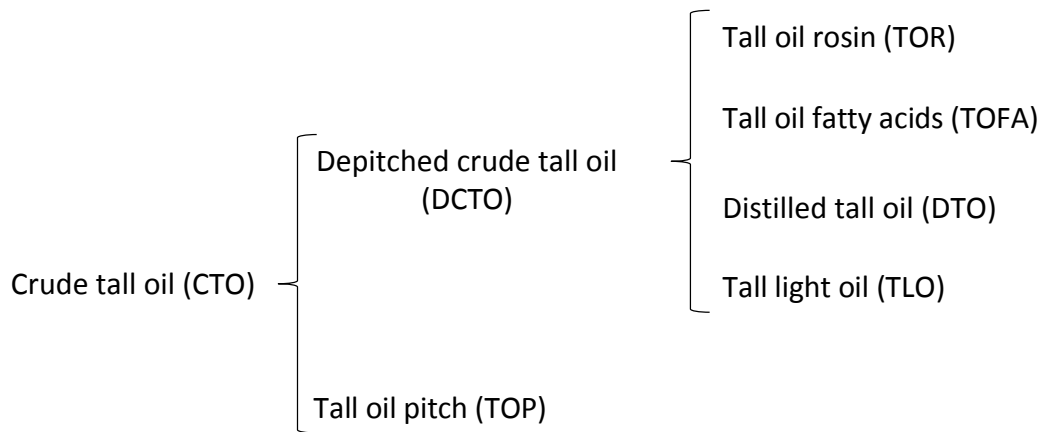


Figure 9. Product fractions of crude tall oil and their abbreviations.

There are two dominant tall oil distillation methods: dry distillation and steam distillation. Dry distillation (presented in Figure 10) is mainly developed and mostly used in Europe, and it is the major method for new plants (Norlin, 2012). According to its name, it does not use entraining media to aid the distillation. Due to this, the process is sensitive to pressure losses inside the columns. Thus, structured packing is a sole option for column internals. Reboilers are either thin film or falling film evaporators. Upsides of the process are low holdup, fast retention time and easier handling of steam condensates (Norlin, 2012).

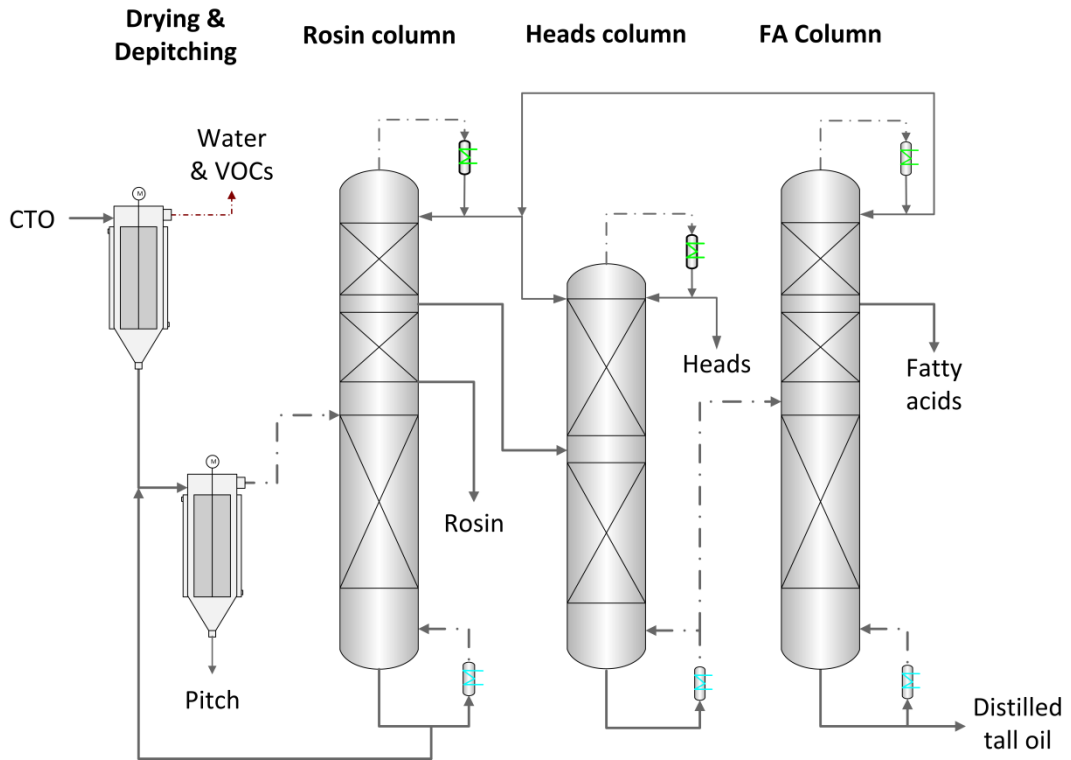


Figure 10. Dry distillation of crude tall oil, edited from Freese & Vock (1982) and Vainiomäki (1995).

Steam distillation, also referred as wet distillation in contrast to dry distillation, has been adapted from petroleum refining and developed for tall oil distillation originally in the United States. Traditional wet distillation uses direct superheated steam heating and bubble-cap tray internals (Norlin, 2012). The steam acts also as an entrainer media to upkeep pressure stability in the column. The use of liquid trays in distillation increases the holdup volume of the columns, and exposes the tall oil to detrimental thermal reactions. In modern wet distillation systems, structured packings can also be used (Bress, 1982). Other issues with wet distillation are large energy demand and the handling of contaminated steam (Hermann & Kuratle, 1983; Norlin, 2012). Wet distillation is presented in Figure 11.

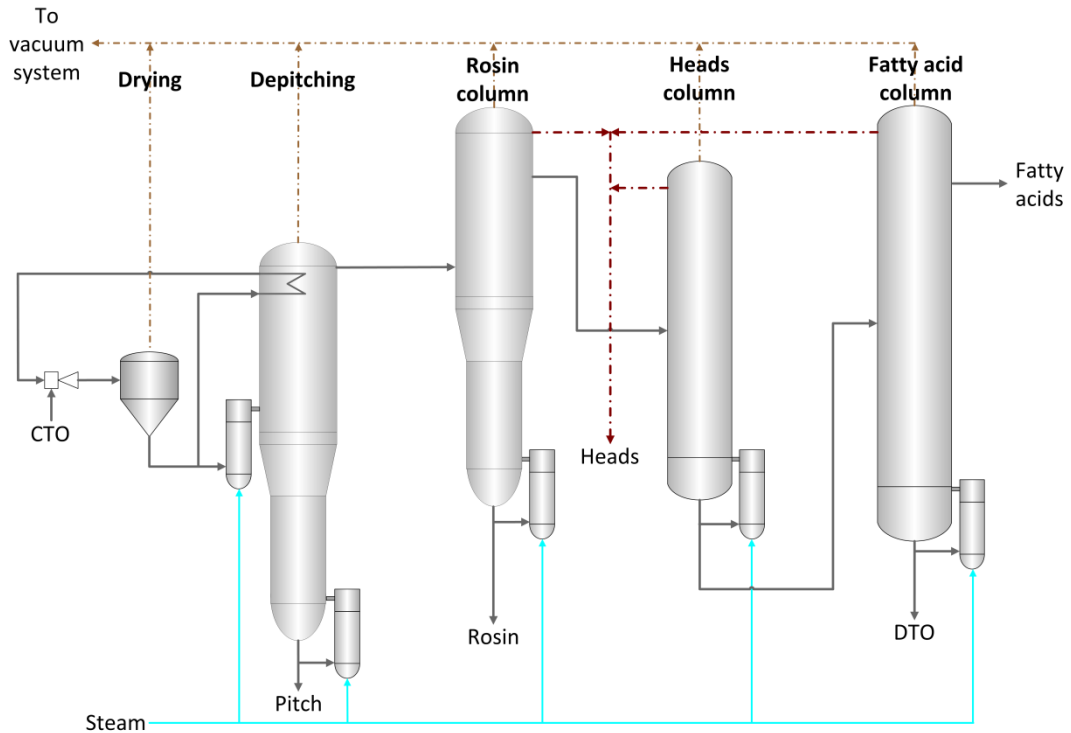


Figure 11. Steam distillation, also called wet distillation of crude tall oil, adapted from (Bress, 1982).

The exact yield of different fractions varies geographically, but roughly 30% of CTO can be fractionated into TOR and TOFA each, around 30% to TOP and TLO, and the rest to DTO. In America, the TOR and TOFA share can be higher (Drew & Propst, 1981).

3 Tall oil end products and markets

This chapter aims to give the reader a general understanding of the CTO markets, and the end use of its fractions: Prices and volumes are given when available. The discussion of future potential of the tall oil industry can be found in Chapter 5.

3.1 Crude tall oil (CTO)

The production of CTO is estimated to be 1.6–2 million tonnes by several sources presented by Peters and Stojcheva (2017). Of this, around 50% is produced in USA and

30% in Scandinavia (Holmbom, 2011; Peters & Stojcheva, 2017). Major part of it is used by tall oil distillers. However, part of it is also burned for energy on pulp mill site, where it can be used to substitute fuel oil. Because of the similar energy content, CTO market price is correlating with the heavy fuel oil (HFO) price, as can be seen in Figure 12. Fractionation through distillation creates significant added value to CTO. During the last years, market prices for distillation products have also increased faster than the price of CTO, as can be noted from the Figure 13.

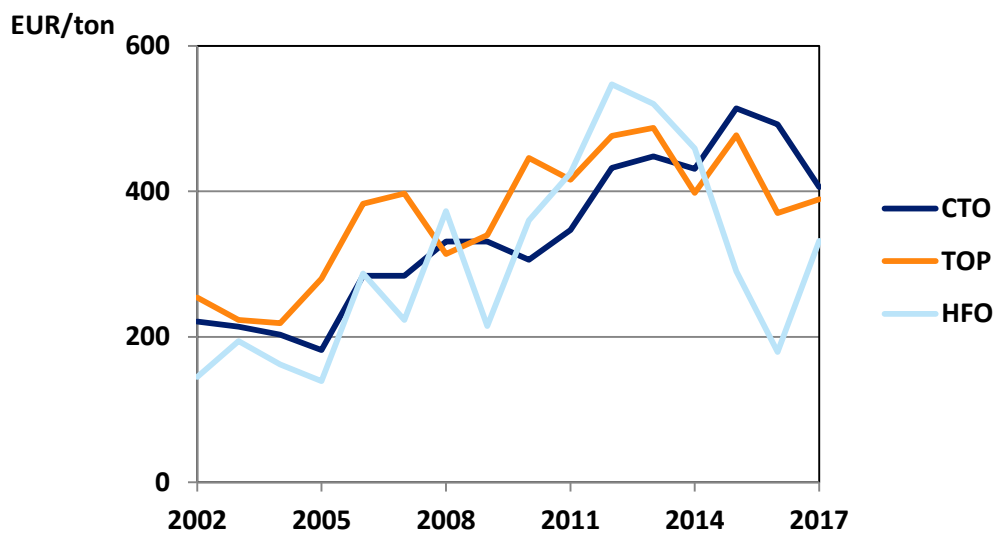


Figure 12. Historical market prices of crude tall oil (CTO), pitch (TOP) and heavy fuel oil (HFO), adapted from Eurostat (2018).

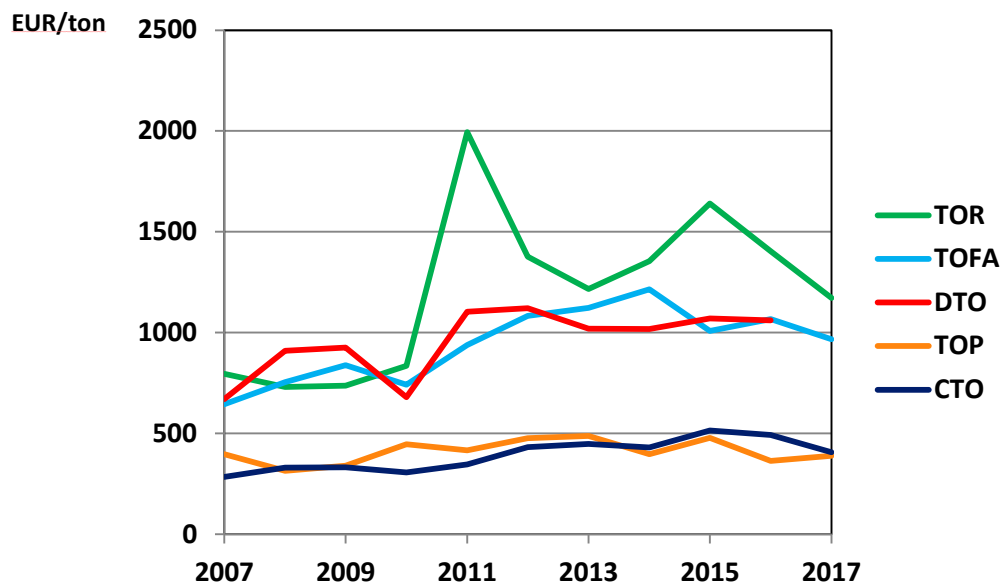


Figure 13. Market prices of tall oil fractions, adapted from Eurostat (2018).

3.2 Tall oil rosin (TOR)

Tall oil rosin is a significant paper size source, but lately the demand of this application has decreased due to cheaper sizing materials and the general downfall of graphical paper production. Nowadays, the most important end uses of tall oil rosin are adhesives and printing inks alongside paper sizing (Poikolainen, 2018). Rosin is also used as a rubber emulsifier (Huibers, 2000; Turner, 2010). One of the most important characteristics of rosin and simultaneously the most important price factor is its colour. The best colour grade rosin has a three times higher market value than the lower grades (Holmbom, 2011).

The global rosin markets are estimated to be around 2.5 million tons, of which 25%, 500 000 tonnes is originating from tall oil (Poikolainen, 2018).

3.3 Tall oil fatty acids (TOFA)

TOFA competes in a larger market of carboxylic acids against other vegetable oils such as soy and rapeseed oils. Favourable aspect for TOFA in this competition is that it is not

competing of farmland against food production. General end use applications for fatty acids are as alkyd resins for paints and as polyamides for adhesives, inks and epoxies (Alalauri, 1995; Huibers, 2000). Rapidly increasing end use is also biofuels, where fatty acids may be utilised as fatty acid methyl esters (FAME), or as a hydrotreated vegetable oil (HVO) to renewable biodiesel. The global fatty acid market size is around 10 million tonnes, of which 5% is from tall oil (Poikolainen, 2018).

3.4 Distilled tall oil (DTO)

DTO is described as “a unique mixture of rosin and fatty acids” (Huibers, 2000), with approximately 40% resin and 60% fatty acid content. DTO can be used in soaps and coating applications, as well as in the glue and paint production. For analytical applications, pimaric and isopimaric acids can be crystallised from DTO (Norlin, 2012; Riistama, et al, 2003; Soltes & Zinkel, 1989).

3.5 Tall light oil (TLO)

TLO consists mainly of neutrals and short-chain fatty acids. It is often considered as a waste fraction alongside pitch (Alalauri, 1995). However, it has interesting solvent properties, and relatively pure palmitic acid could be recovered from it via crystallisation (Huibers, 2000). Still, due to its bad odour it is often incinerated on the distillation site (M. Rintola, personal communication, 16 February 2018; T. Silfverberg, personal communication, 21 March 2018).

3.6 Tall oil pitch (TOP)

TOP has been normally used for energy production and is often sold back to a pulp mill to be used as a lime kiln fuel (M. Rintola, personal communication, 16 February 2018; T. Silfverberg, personal communication, 21 March 2018). This can be seen in the market price of TOP (in Figure 12), which is dependent on HFO price, similarly like CTO.

However, some refined applications for it have been developed. Upgraded TOP can be used in glues and in printing inks, as well as an asphalt emulsifier (Norlin, 2012; Riistama, et al., 2003). In the EU Renewable energy directive, TOP is double accountable for the renewable fuel target (European Parliament, 2015). Despite this, Renewable Fuel Standard (RFS) in USA does not recognise TOP or any other tall oil product as a valid renewable fuel feedstock (United States Environmental Protection Agency, 2017).

4 Tall oil chemistry

In this Chapter, a closer look into the chemical nature of tall oil components is taken to fully understand the behaviour of them during refining. The components are divided into chemically more accurate subcategories, slightly differing from product fractions: Resin acids, Fatty acids and Neutrals. Relation of the compounds and product fractions are given at the end of this Chapter, in Table 2.

Majority of tall oil consists of tri- or higher terpenes and their derivatives, in other words compounds consisting of six or more isoprene units. Lower terpenoids are usually vaporised during the pulp cooking process as turpentine, but some traces of them can also be found in tall oil (Sjöström, 1993). Another significant part of tall oil is fatty acids. These two classes are accompanied by some minor and miscellaneous neutral components.

4.1 Resin acids

Wood resin acids have two major skeletal structures: abietane and pimarane types. In America, also labdane types are found in some wood species: Communic acid from slash pine and lambertianic acid from sugar pine (Ekman & Holmbom, 2000; Sjöström, 1993). These structures alongside related atom numbering are presented in Figure 14 below.

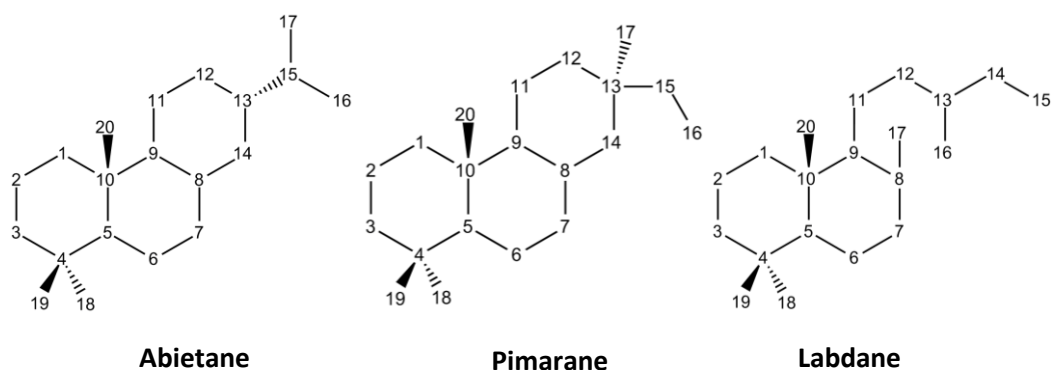


Figure 14. Wood resin acid skeletal structures with atom numbering.

Due to their similarity, it is not necessary to separate different types of resin acids in most applications: They are most often treated just as rosin. However, they differ from each other by their dehydrogenation products: Abietane type produces retene, while pimarane type yield pimarathene (Soltes & Zinkel, 1989). The carboxylic group of resin acids at C18 is sterically hindered, which is why they are more resistant against the esterification reaction than fatty acids (Pensar, 1977).

Resin acids with pimarane basic structure (Figure 15) are tricyclic diterpenes with one double bond inside the rings and one in the ethylene group linked to a C13 carbon. Most usual components of this type are pimaric and isopimaric acids. These differ from each other mostly by the location of the double bond. Also Sandaracopimaric acid (CAS 471-74-9), a stereoisomer of pimaric acid in regard to ethylene bonding to C13, belongs to this group and is often found from tall oil in small concentrations (Holmbom & Avela, 1971). Study by Lawrence (1959) showed that abietane type resin acids concentrate into the rosin fraction during tall oil distillation, indicating that pimarane type resin acids are more vulnerable to polymerisation reactions and are partly removed alongside pitch. They are found to be harmful to aquatic life (Leach & Thakore, 1978; Rogers, et al., 1975).

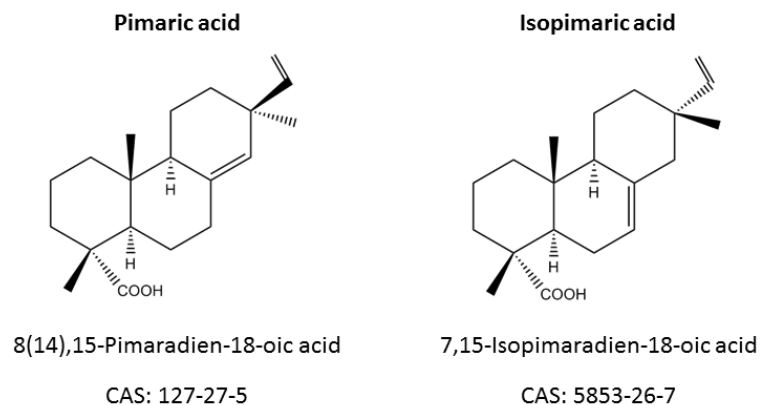


Figure 15. Pimarane type resin acids with their IUPAC names and CAS numbers.

Resin acids with abietane skeleton have two double bonds inside the tricyclic structure and a saturated isopropyl group bound on the C13. Most common component in this group is abietic acid, sometimes also called sylvic acid, that is yellowish powder in room temperature (Nikunen, et al., 1990). Others significant components of this group are levopimaric, palustric, dehydroabietic and neoabietic acids. Dehydroabietic acid is the only component to have an aromatic ring, while neoabietic acid's another double bond is between the isopropyl and the C13 carbon. Due to their conjugated structure, all abietane type resin acids, except dehydroabietic acid, are easily isomerised and oxidised compared to pimarane-structured acids (Sjöström, 1993). Especially levopimaric acid has been noted to isomerise into abietic acid and further into palustric and neoabietic acids in elevated temperatures (Holmbom & Ekman, 1978; Walter, et al., 1989). Abietic, dehydroabietic and palustric acid have been found to be harmful to aquatic life (Leach & Takore, 1976; Leach & Thakore, 1978; Oikari, 1987). These resin acids are presented in Figure 16.

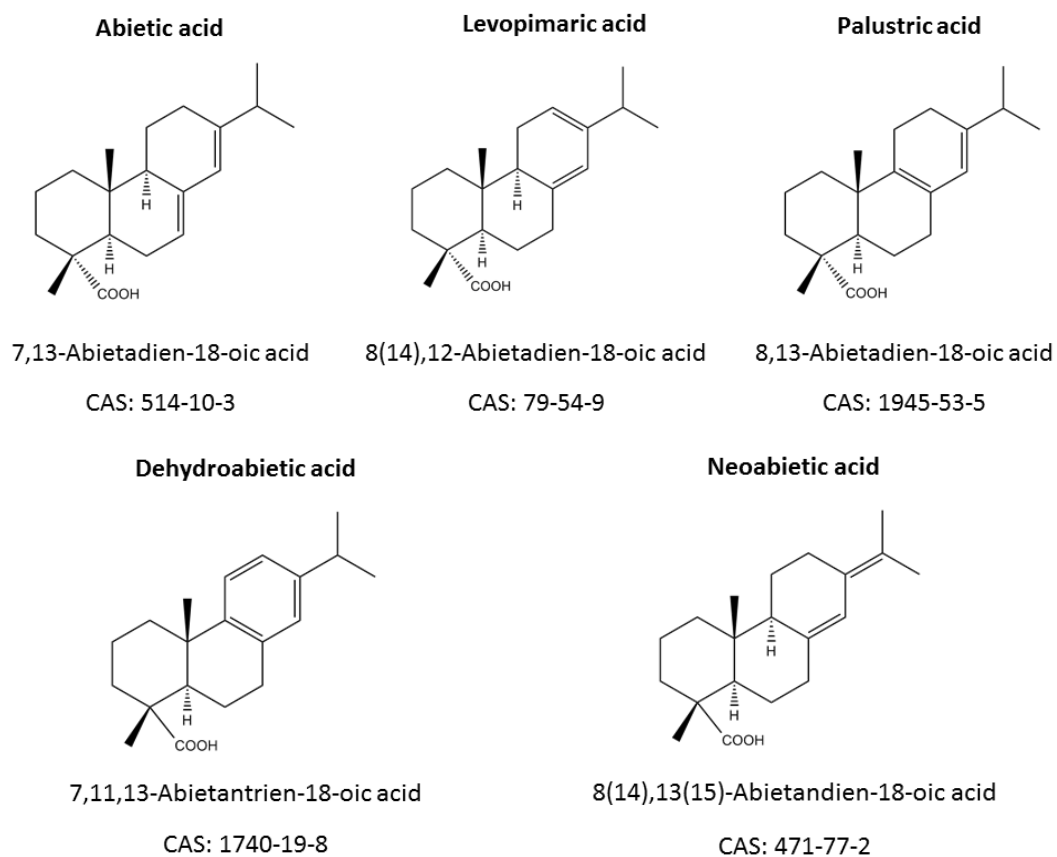


Figure 16. Abietan type resin acids with their IUPAC names and CAS numbers.

Labdane type resin acids (Figure 17) differ significantly from the other two main types as they are bicyclic, usually three double bonds containing compounds. They are usually specific to some American pine species, in which they can be present in high concentrations. Communic, also known as elliotic acid can be found from slash pine (*Pinus elliotii*), whereas lambertianic acid is abundantly available in sugar pine (*Pinus lambertiana*). Structurally also secodehydroabietic acid, aromatic isomerisation product of levopimaric acid formed during kraft pulping, is close to this group (Sjöström, 1993).

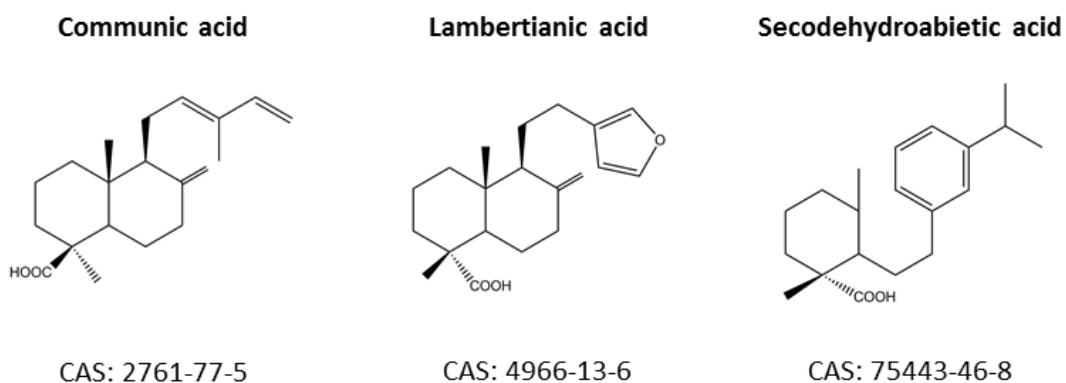


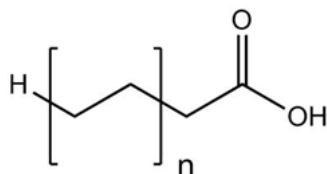
Figure 17. Labdane resin acids and secodehydroabietic acid.

4.2 Fatty acids

Fatty acids present in tall oil originate from fatty compounds, mostly triglycerides of wood that are degraded during pulp cooking. These straight, long-chain acids are present in wood in saturated and unsaturated forms. Their occurrence is strongly dependent of the wood species and geographical location: Triunsaturated fatty acids are more common in Scandinavia region than in America, whereas birch and spruce species are more abundant of saturated forms than pines (Duncan, 1989; Holmbom & Avela, 1971; Holmbom & Ekman, 1978). According to Rydholm (1965), fatty acids are contributing to pitch deposits in pulping and papermaking, which is why their removal along the tall oil is beneficial to the paper quality. Fatty acids are quite easily esterified with alcohols from the unsaponifiable fraction in elevated temperatures. Esterification is an equilibrium reaction: The absence of water, the other esterification end product, speeds up the otherwise relatively slow reaction. It is also noteworthy that free fatty acids are highly corrosive.

Of the saturated fatty acids (Table 1), palmitic acid is the most common. Stearic acid is also universally found from trees, but in lesser concentrations. Other components, such as arachidic and behenic acids, have been found from Finnish birches (*Betula verrucosa* and *Betula pubescens*) (Holmbom & Avela, 1971). Both palmitic and stearic acid in effluents are found to be harmful to fish (Nikunen, et al., 1990). The shortest saturated fatty acids have low boiling points and they partly end up in the rejected light oil fraction during distillation.

Table 1. Most common saturated fatty acids in wood.



n	Formula	Common name	IUPAC	CAS
7	C ₁₆ H ₃₂ O ₂	Palmitic acid	Hexadecanoic acid	57-10-3
8	C ₁₈ H ₃₆ O ₂	Stearic acid	Octadecaonic acid	57-11-4
9	C ₂₀ H ₄₀ O ₂	Arachidic acid	Eicosanoic acid	506-30-9
10	C ₂₂ H ₄₄ O ₂	Behenic acid	Docosanoic acid	112-85-6

Of the unsaturated fatty acids (in Figure 18), oleic and linoleic acids are the two most significant ones. As mentioned before, triunsaturated acids are more uncommon, and during kraft cooking and tall oil refining they are vulnerable to thermally catalysed destructive reactions. The double bonds tend to isomerise into conjugated locations, which may further isomerise from cis- to trans-orientation. This isomeric form exposes them to destructive reactions such as cyclisation. Especially pinolenic acid is known to convert almost completely into cyclopinoleic acid via Diels–Alder reaction in tall oil distillation, causing yield loss (Hase, et al., 1974; Holmbom & Eckerman, 1983). Fatty acids' trans-isomers have also higher melting and boiling points than their cis-isomers, causing handling difficulties (Duncan, 1989).

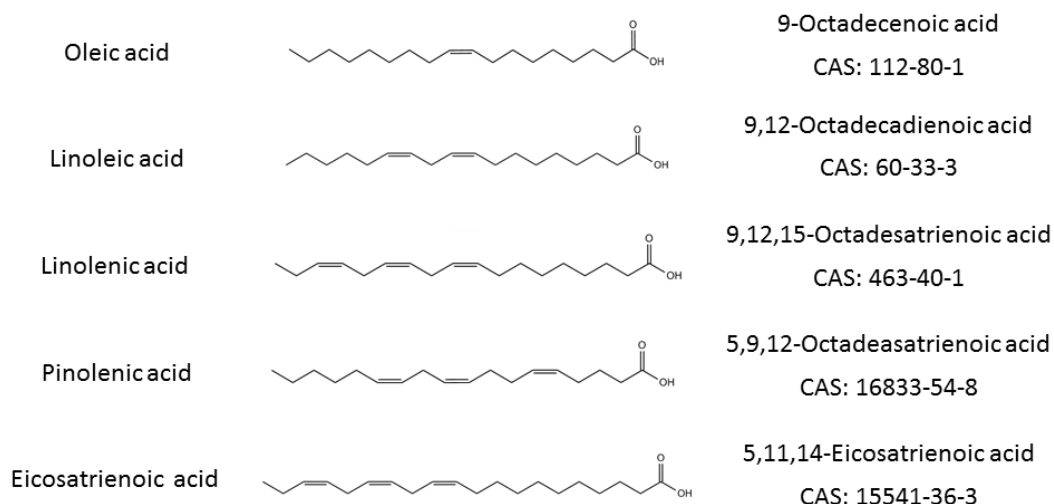


Figure 18. Unsaturated fatty acids present in tall oil.

4.3 Neutrals

Neutrals is a “residual” group of chemical components in tall oil. Terms neutrals, unsaponifiables and pitch are closely related but their differences are presented through classification in Figure 19. The term neutrals refers to tall oil’s non-acidic contents of alcohols, aldehydes, esters, hydrocarbons and phytosterols. The name unsaponifiables refers to chemicals lacking the ability to form soaps, especially during tall oil separation from black liquor. The pitch is the leftover fraction of distillation, consisting of high-boiling neutrals and also some leftover fatty and resin acids. It is noteworthy that besides pitch, a significant amount of neutrals are rectified also in the heads (Holmbom, 1978).

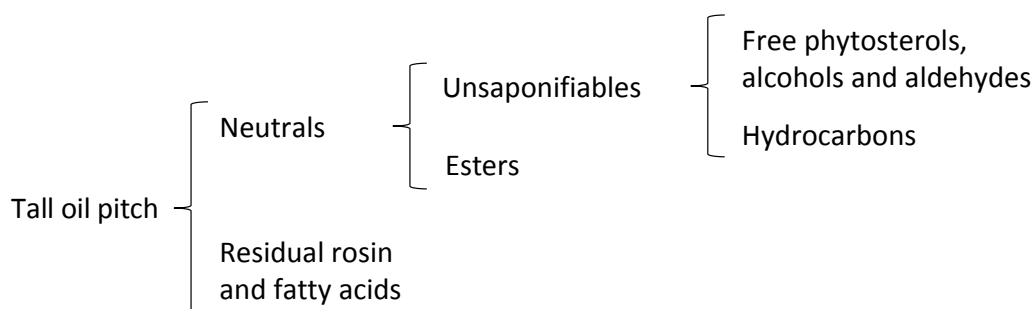


Figure 19. The classification of tall oil pitch components.

Phytosterols (Figure 20) are plant sterols and stanols (saturated sterols). In living wood they are concentrated on the heartwood (Saranpää & Nyberg, 1987), while on tall oil processing they are rectified into the TOP. They are significantly more abundant in tall oil than in other vegetable oils (Holmbom & Avela, 1971; Belitz, et al., 2004). β -Sitosterol, and its saturated version, β -sitostanol are the most commonly acquainted phytosterols (Vikström, et al., 2005; Wong, 2001). Campesterol and the saturated version of it, campestanol are also quite universally found from tall oils in small concentrations. Their biosynthesis intermediates, cycloartenol and lupeol can be found in negligible amounts, mostly from birch, spruce and larch (Vikström, et al, 2005).

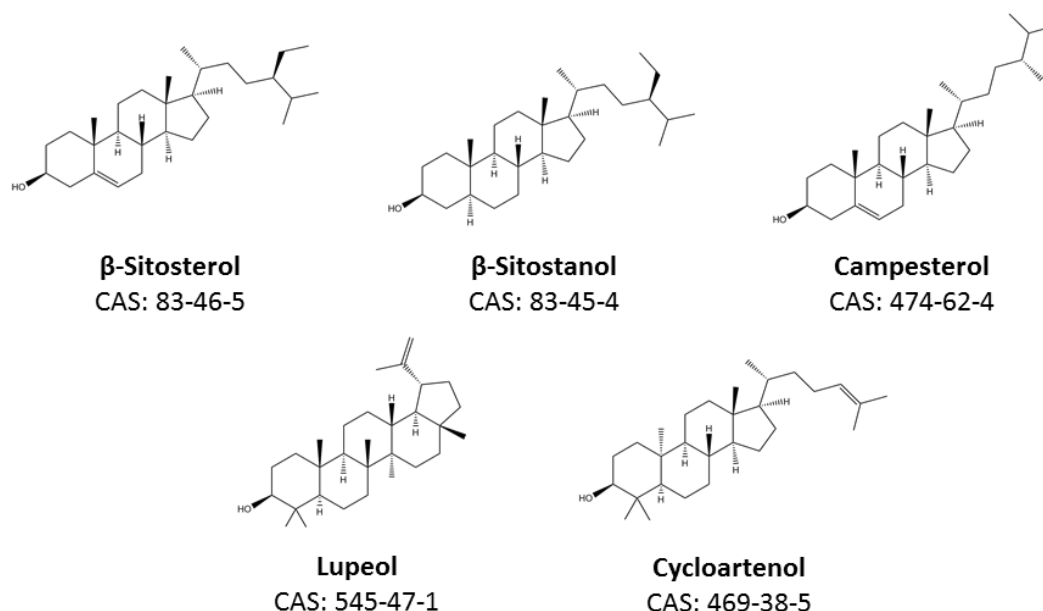


Figure 20. Most common phytosterols.

Neutral components unique to birch may end up in the crude tall oil through mixed wood pulping. In paper production, they are causing significant pitch problems, which is why at some hardwood pulp mills rosin soap is bought to remove them from pulp. Most important birch components are betulinol and betulaprenols. Betulinol is a dihydric alcohol and its removal from pulp is very difficult. At paper machine, it can cause tacky deposits and hinder its runnability. Betulinol can be present in tall oil as a free or as an esterified version, for example as a methyl betulinate. Betulaprenols are straight long-chain alcohols. Also squalene, precursor for triterpenoid biosynthesis, is available in

birch extractives in significant quantities. Commercially, it is most commonly extracted from a shark liver and is often used in cosmetics as an antioxidant and a moisturiser (Lövberg, 1983; Wolosik, et al, 2013). These compounds are presented in Figure 21.

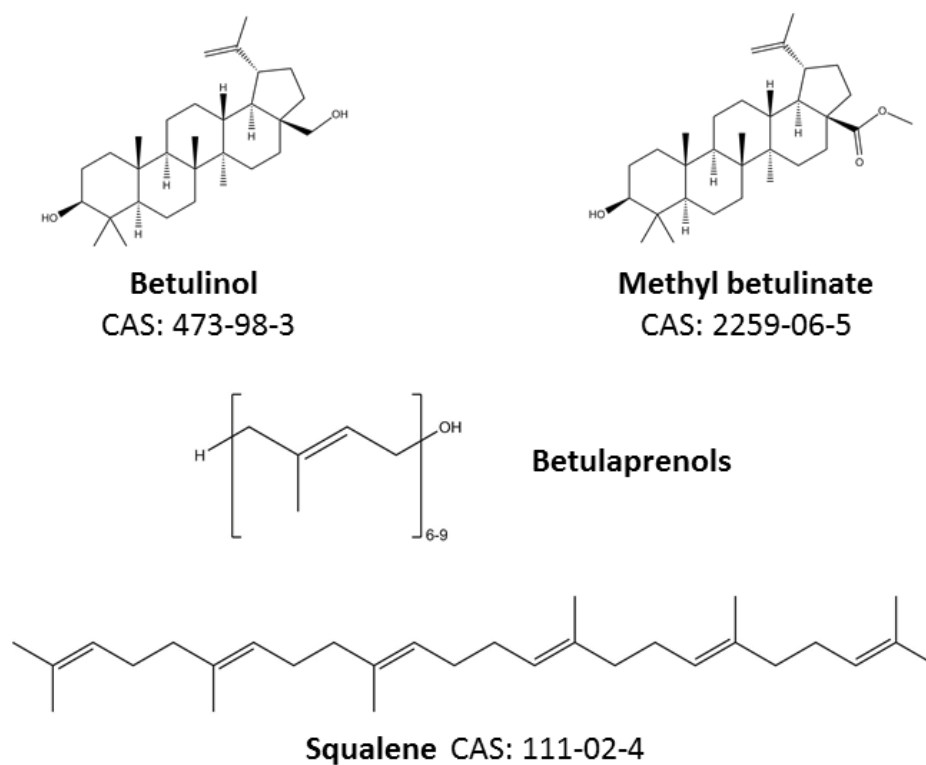
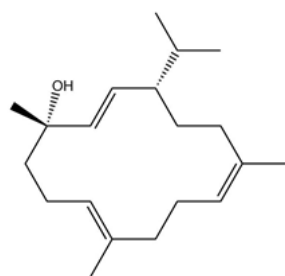
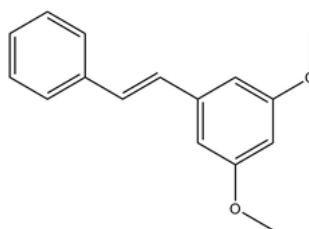


Figure 21. Unsaponifiable components typical to birch.

Also other wood species have some unique, although not as important pitch components as birch. Thunbergol is a diterpenic alcohol specific to spruces (*Picea*). In living wood it is mostly present as a free alcohol in heartwood (Ekman, 1979; Sjöström, 1993). Pinosylvin is a pine-specific dihydric alcohol and a stilbenoid component that is usually present in tall oil in its ester form pinosylvin dimethyl ether. This ether has shown fungicide and pesticide properties, and also its tendency to colourize fatty acid epoxies red has raised interests towards its separation from tall oil (Conner, 1989). The structures of these compounds are presented in Figure 22.



Thunbergol
CAS: 25269-17-4



Pinosylvin dimethyl ether
CAS: 21956-56-9

Figure 22. Some unsaponifiables of specific wood species.

In addition to compounds introduced above, the unsaponifiable fraction contains a variably large amount of alcohols, aldehydes and hydrocarbons with resin and fatty acid skeletal structures (Conner & Rowe, 1975). These are probably remains and side products of the biosynthesis of the acids. The esters of methanol and alcohols mentioned before, and fatty acids are also present. In TOP, some free acids are present, but these can be considered as impurities that separation processes have not been able to selectively remove alongside their product stream (Holmbom & Erä, 1978).

Table 2. Occurrence of tall oil components in distillation products, adapted from Holmbom (1978).

	TOR	TOFA	DTO	TOP	TLO
Resin acids					
Secodehydroabietic	-	-	1-10 w-%		
Pimaric	1-10 w-%	-	1-10 w-%	-	-
Isopimaric	1-10 w-%	-	1-10 w-%	-	
Sandaracopimaric	1-10 w-%		-		
Palustric	1-10 w-%	-	1-10 w-%	-	
Abietic	> 10 w-%	-	1-10 w-%	1-10 w-%	
Neoabietic	1-10 w-%		-	-	
Dehydroabietic	> 10 w-%	-	1-10 w-%	-	
Fatty acids					
Palmitic		-			1-10 w-%
Pinolenic		> 10 w-%	1-10 w-%		1-10 w-%
Linoleic	-	> 10 w-%	1-10 w-%	-	> 10 w-%
Oleic	-	> 10 w-%	1-10 w-%	-	> 10 w-%
Stearic		1-10 w-%	-		-
Eicosatrienoic	-	-	1-10 w-%		
Arachidic	-	-	1-10 w-%		
Behenic	1-10 w-%		-		
- As esters					
Palmitic				-	
Pinolenic				-	
Linoleic	-		-	1-10 w-%	-
Oleic	1-10 w-%		-	1-10 w-%	-
Stearic				-	
Eicosatrienoic	-		-	-	
Arachidic	-		-	-	
Behenic	-			-	
Neutrals					
Pimaradiene					-
Pimaral		-			1-10 w-%
Isopimaral		-			1-10 w-%
Pinosylvin dimethyl ether		-			1-10 w-%
Dehydroabietal		-			-
Abietal		-			1-10 w-%
Squalene	-				
3.5-stigmastadiene	-			-	
β-sitosterol				-	
- As esters					
Dehydroabietic methyl ester		-			-
Abietic methyl ester		-			-
Pimarol				-	
Behenol				-	
Tetracosanol				-	
β-sitostearate				1-10 w-%	

> 10 w-%
1-10 w-%
- < 1 w-%

5 Future outlook and potential end products

It is calculated that current CTO production potential is 2.1–2.6 million tonnes, of which 1.6–2 million tonnes is already retained (Baumassy, 2016; Peters & Stojcheva, 2017; Poikolainen, 2018). Thus, there is a theoretical chance to increase CTO production by around 600 000 tonnes. However, this might not be reached due to losses caused by mixed wood pulping. If the hardwood content in pulp mill feed exceeds 50%, the yield of tall oil is weakened due to the poor soap formation. Depending on heavy fuel oil price fluctuations, it may also be attractive to burn CSS or CTO on site instead of recovery.

Except for direct combustion, interest towards tall oil energy usage as refined biofuels has been increased by the renewable energy targets set by the European Union, and the mitigation principle of the UN Paris Agreement (United Nations, 2015). Also, in the proposal for the new EU renewable energy directive, tall oil is added as valid feedstock for advanced biofuels alongside TOP (European Commission, 2017). For example, there has been recently several master's theses related to biofuel production from tall oil in Aalto University (Lehtinen, 2011; Pohjantähti, 2012; Puustinen, 2015; Rautalin, 2009).

Despite motivations to produce biofuels from CTO, the pine-derived chemicals are maintaining their competitiveness. It is estimated that the profitability of conventional chemical refining equals that of the biofuel production with current policies (Poikolainen, 2018). Pine chemicals markets are growing 4–5% annually, driven mostly by adhesives and coating applications (Baumassy, 2016).

Another interesting way to utilise CTO is the extraction of phytosterols from its neutrals. They are valuable feedstock for the cosmetics and pharmaceutical industry (Lövberg, 1983). Sterol extraction has been commercialised already in the 1970s in the Kaukas chemical mill in Finland via CSR-process developed by Holmbom and Avela (US patent US3965085A, 1976). Later on, the production has been terminated, but new competitive methods for sterol separation are studied, especially using TOP as a feedstock (US patent US8680324 B2, 2014; EP patent WO2017137908, 2017). According to Ukkonen (2016), the sterol market for pharmaceutical applications is growing 8–10% annually.

EXPERIMENTAL STUDY

6 Scope

As can be concluded from the Literature Review, the TOFA and TOR fractions of tall oil are already well utilised and efficiently used by the chemical industry. The greatest improvement potential is for the “leftover” fractions of TOP and TLO. Of these, the pitch fraction is more interesting due to its larger production volume, which is around 20–30% of CTO, roughly double the amount of the TLO fraction. The most value added use for this could be phytosterols separation. However, this growing market is still very limited, and the current depitching conditions may lower the sterol yield (K. Ukkonen, personal communication, 9 April 2018).

From the perspective of pulp mills, the prospects for the tall oil business have been poor in recent years, which can be noted from the CTO market price development in Chapter 3 of the Literature Review. This could be changed by further refining tall oil in a pulp mill by performing depitching already onsite. The process could use the relatively cheap and abundant process energy provided by the recovery and biomass boilers of the pulp mill. Additionally, the extracted pitch can be used as a lime kiln fuel, replacing natural gas or heavy fuel oil. This helps the pulp mill to reach carbon neutral operation. Pitch is already burned in lime kilns after depitching in a distillation site and transportation back to the pulp mill. With onsite depitching, savings in logistics costs could also be made. In the future, when the sterol market has gained sustainable market size, the process could also be modified to produce valuable sterol-rich pitch for the pharmaceutical industry.

The hypothesis in this work is that the tall oil refining chain could be made more efficient by integrating the depitching process into pulp mills. In this Experimental Study, the technical and economic feasibility of this concept is tested by defining the depitching process, calculating its effect to pulp mill mass and energy balances and defining investment cost for the process. These are then used to calculate profitability and the payback period for the investment. These key figures are then compared between TOP separation and competitive scenarios.

7 TOP separation

7.1 Pulp mill

This study calculated the pulp mill mass balance for TOP separation to determine main process flows and department capacities. A greenfield softwood sulphate pulp mill with a capacity of a million air dry tonnes (ADt, 90% dry matter content) of bleached pulp per year was taken as the pulp mill design that was used as the basis for the calculations. The calculations also assumed that the wood feed is 100% pine, and all the wood is supplied to the mill unbarked. The crude sulphate soap obtained from the evaporation plant is acidulated into crude tall oil (CTO). The CTO yield was assumed to be 50 kg/ADt pulp, equalling 49 300 tonnes of CTO annually or 141 tonnes per day. Moreover the assumption was that chlorine dioxide, which is used as a bleaching chemical, is produced on site by reducing sodium chlorate with methanol. Simplified box diagram of the mill model is presented in the Figure 23. The mass balance for the mill model is further presented in Appendix 1.

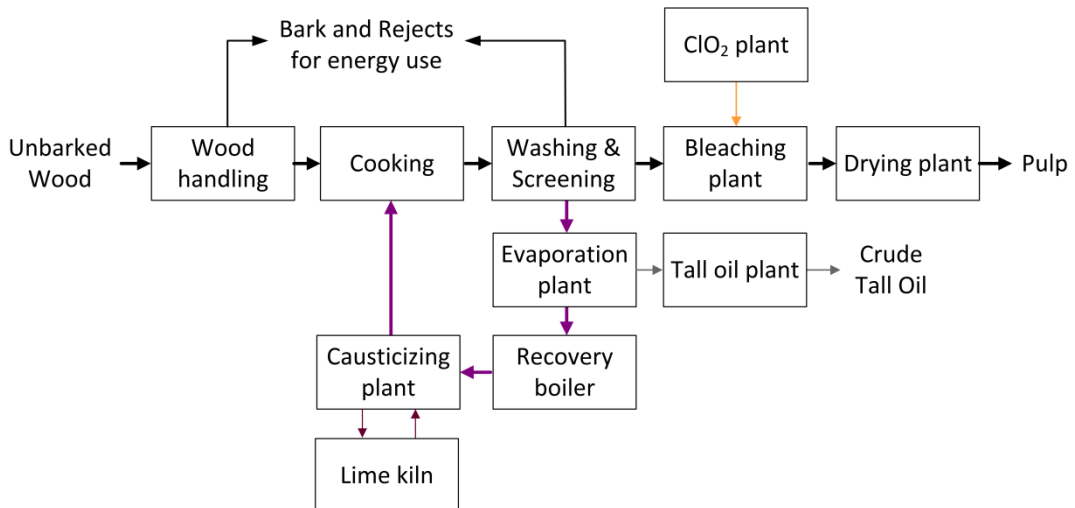


Figure 23. Simplified box diagram of the model mill and its main mass flows.

7.2 Depitching process definition and equipment

In this study, depitching is made similarly to that of a dry distillation plant. This method is technically more applicable to a pulp mill than wet distillation depitching due to its

lower effluent generation and generally better product quality, as mentioned in section 2.1.3 of the Literature Review. CTO feed to the depitching stage is taken from the CTO storage tank after acidulation. The feed is estimated to be at atmospheric pressure and at a temperature of 80 °C. This feed is then pre-heated to 180–200 °C before the actual depitching stage. For the pre-heating, flashed steam condensate from the depitching stage is used: This way energy consumption can be lowered. 180–200 °C is chosen as the temperature range, as in this range the residual water and unwanted volatiles will evaporate. However, it is still a stable temperature regarding degradation reactions of organic components, which tend to rapidly increase at temperatures over 200 °C (Drew & Propst, 1981). To efficiently remove residual vapours from the liquid and to gain a good heat transfer rate, a falling film evaporator (FFE) is used as a pre-heater. To adjust the temperature of tall oil after pre-heating, a recirculation line is added.

Next, from the preheater, heated CTO stream is used directly in the depitching stage, which is executed in a thin film evaporator (TFE) at 267 °C and at an absolute pressure of 1 kPa. In the model, heating happens with a high pressure steam through the heating jacket of the evaporator. In these conditions, suggested by Norlin (2012) and Huibers (2000), most of the free acids in tall oil are evaporated. In the tall oil industry, lower pressure and temperature levels are probably used, but this selected level represents maximum reasonable condition. Liquid TOP leaves the evaporator from the bottom and depitched crude tall oil (DCTO) vapours from the top of the evaporator. TOP is pumped from the TFE vessel to a heated storage tank. To allow better controllability of the depitching process, part of the leaving pitch can be circulated back to the TFE feed line.

DCTO vapours need to be condensed before storage. This is done in a vapour condenser which has a connection to the vacuum system in order to maintain the desired pressure level for the whole process. As there is a risk of resin acids crystallising on the cooling surfaces (Soltes & Zinkel, 1989), a relatively hot cooler media has to be chosen. For the process, a cooling media inlet temperature of 120 °C and a final product temperature of 140 °C are selected. After condensing, the DCTO is stored in a heated tank. Line diagram of the process is presented in Figure 24.

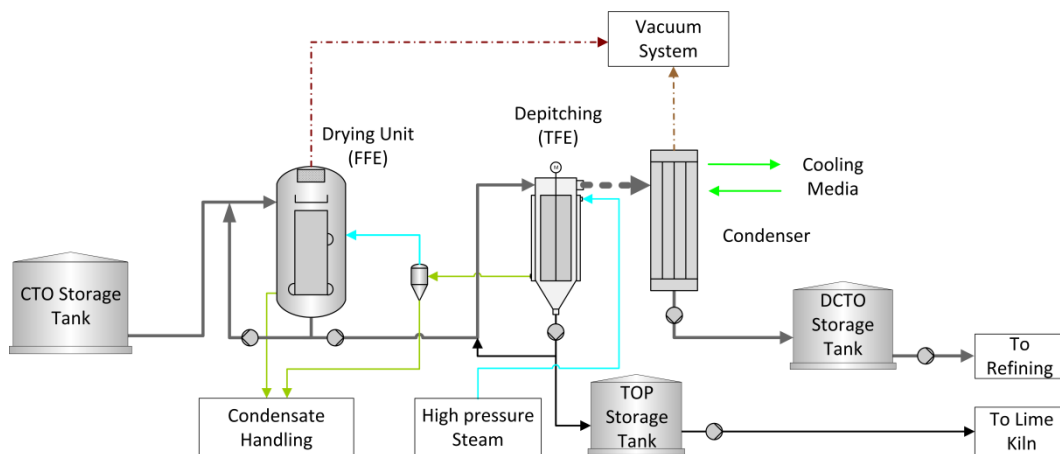


Figure 24. Line diagram for the depitching process.

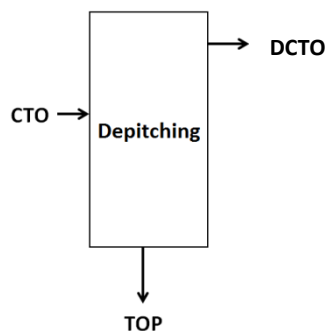
7.3 Process modelling

Bjarne Holmbom extensively defined tall oil components for his doctoral dissertation (Holmbom, 1978), including their distribution between distillation fractions. The CTO measured for this work has a higher pitch and rosin acid content, and a lower fatty acid content than typical Finnish tall oil. However, these results were used as a basis for the process modelling in the present study, as they are more detailed than other measurements reviewed. Component identification and measurements for relevant tall oil fractions can be found in Appendix 2. In the model, the feed is considered to be pure of impurities such as ash or water, as their concentrations and contribution to the process are relatively small.

The measurements show that some degradation and isomerisation reactions occur during fractionation. The TLO and TOP fractions are also quite poorly identified. To balance the model, pseudocomponents representing unknown pitch and heads compounds were introduced. CTO feed in the model was also manipulated to match product streams. Based on their chemical groups and volatility, the measured compounds were divided into ten model compounds and pseudocomponents. The manipulated streams and model compound classification are presented in Appendix 3. Process streams as model components are presented in Table 3.

Table 3. The depitching process streams as model components. Components marked with asterisks are pseudocomponents.

w-% of tall oil fraction	CTO	DCTO	TOP
Abietic acid	23.3	32.5	1.9
Dehydroabietic acid	10.1	14.0	1.2
Pimaric acid	3.8	5.4	0.1
Oleic acid	22.6	32.1	0.3
Palmitic acid	2.9	4.2	0.0
Behenic acid	0.5	0.8	0.0
Lignoceric acid	0.2	0.3	0.0
β -Sitosterol	0.6	0.6	0.7
Pitch*	31.4	4.1	95.7
Heads*	4.4	6.2	0.0



These model compounds were used to simulate the mass and energy flows of the depitching process. Simulation was performed with Aspen Plus v9 software. Even though most of these model components are well-known and commonly used chemicals, empirical data on their thermodynamics are quite scarce. For example, a sufficient amount of experimental liquid vapour pressure data is only available for oleic and palmitic acid in the DIPPR database (The Design Institute for Physical Properties, 2017). As the model components are mostly quite bulky, their thermodynamic behaviour was considered as ideal and unknown vapour pressures were manipulated to give similar process streams as in Holmbom's studies. Other thermodynamic characteristics are estimated with Aspen's component database. For pseudocomponents, a literature source (Drew & Propst, 1981) is used in the estimation of their thermodynamic properties. Pitch separation itself is modelled as a simple flash distillation. More accurate heat transfer models may be obtained by modelling the process as a batch distillation, or as several flash distillations in series (Ilmanen, 2017). However, the accuracy of the simple model is sufficient for this purpose. Mass and energy flows obtained from the model are presented in Figure 25.

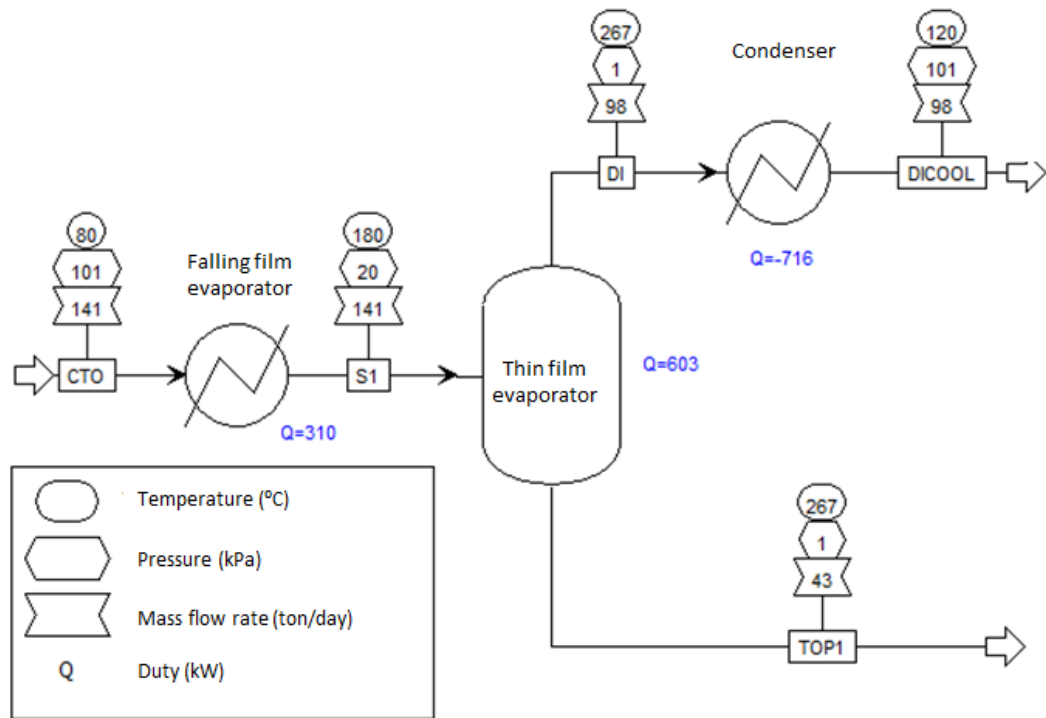


Figure 25. Mass and energy flow results of the model.

7.4 Energy balance

Pulp mill mass balance and depitching process modelling results were used to calculate an energy balance for the mill. In the mill model, heat is produced in a recovery boiler by burning black liquor and in a biomass boiler by burning bark, and wood and pulp reject. Steam generated in these boilers is directed to a turbine plant. In the turbine plant, generators connected to steam turbines produce electricity from the potential difference of high pressure steam. Heat needed in the pulp mill processes is extracted from the turbines as low and medium pressure steam. Pulp mills also produce excess amounts of steam that is used for electricity production with the help of a condensing tail connected to one of the turbines. A simplified diagram of the pulp mill model heat and power generation is shown in Figure 26.

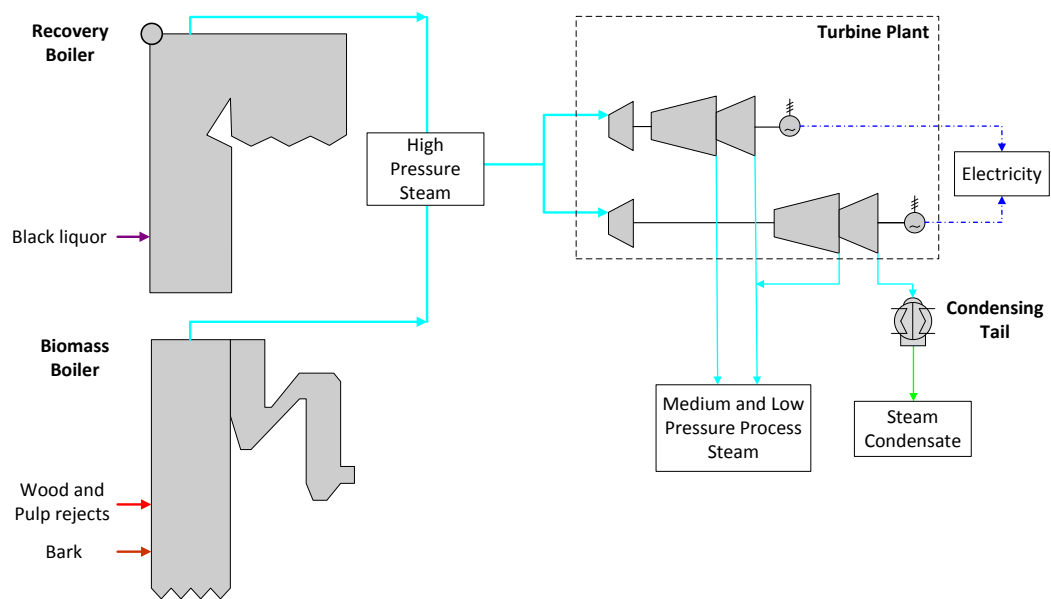


Figure 26. Simplified diagram of pulp mill heat and power generation.

Heat generation for these boilers was calculated from their fuel flows obtained from the mill mass balance. The heat and electricity consumption of the pulp mill and depitching processes was also estimated into the balance. With these values, the steam production and the two steam turbines of the model mill were balanced by using Prosim 5.14 simulation software. During the simulation, the steam turbines were balanced as close to the same electricity production level as possible. The steam network balance shows the electricity production capacity of the mill, which was used to sum up the total energy production and consumption. The energy balance calculation sheet and a steam network diagram can be found in Appendices 4 and 5.

Lime kiln fuel consumption was calculated by using specific heat consumption of 5.5 MJ/kg calcium oxide (CaO), given by Arpalahti et al. (2008). All TOP from the depitching process is burned. Leftover energy is produced by burning heavy fuel oil. Effective heat value for the TOP is estimated to be 38 MJ/kg (Forchem Oy, 2011), and for heavy fuel oil 39.2 MJ/kg (Arpalahti, et al., 2008). Essential production figures for the TOP separation scenario are summed up in Table 4.

Table 4. The key production figures of the TOP separation scenario.

Name	Value	Unit
Pulp production	1 000 000	ADt/a
DCTO production	98	t/d
Heat generation	775	MW
Process heat consumption	379	MW
Electricity generation	219	MW
Process electricity consumption	82	MW
Electricity sales	137	MW
Fuel oil consumption in lime kiln	76	t/d

8 Competitive scenarios

Tall oil depitching in a pulp mill has two main objectives: to produce a refined crude tall oil product and simultaneously renewable fuel for the lime kiln. To study its feasibility, alternative approaches to these objectives have to be compared to it. For the production of a refined crude tall oil product, an alternative choice is to produce crude tall oil in a traditional way, and further refine it in a distillation site. For the lime kiln fuel alternatives, the use of traditional heavy fuel oil was taken as a conventional scenario. Another renewable fuel, wood bark gas was also taken into consideration. Three competitive scenarios were built from these options: A conventional scenario, a bark gasification scenario, and a combined depitching and gasification scenario. All the scenarios are presented in Table 5 below.

Table 5. The tall oil product grades produced and lime kiln fuels used in feasibility study scenarios.

No	Description	Tall oil grade		Lime kiln fuel		
		CTO	DCTO	HFO	TOP	Bark gas
1	Depitching		x	x	x	
2	Conventional	x		x		
3	Bark gasification	x				x
4	Combined depitching and gasification		x		x	x

It is presumed that all the bark generated in the mill is used for energy production, either in a biomass boiler or in a combination of a biomass boiler and a bark gasifier, depending on the scenario. As the calculated mill mass balance (Appendix 1) does not take bark end use or lime kiln fuel consumption into consideration, it is valid for all the scenarios.

8.1 Conventional case

For conventional case, the energy balance was calculated similarly as in the depitching scenario without the additional heat and electricity consumptions of depitching equipment. HFO consumption in the lime kiln was calculated with the same principles and values as in section 7.4, this time without additional fuels. The size of the biomass boiler in this scenario is same as in the depitching one. The energy balance is presented in Appendix 6, and the key production figures in Table 6 in section 8.4 together with other scenarios.

8.2 Bark gasification

Besides tall oil pitch, bark gas is another renewable fuel option for a lime kiln that could be produced onsite from pulp production side streams with a gasifier. In this scenario, the gasifier was dimensioned so that carbon neutral operation can be reached, i.e. no additional fuels are needed in the lime kiln besides bark gas. For the bark gasification mass balance, similar equipment of the Metsä Fibre Joutseno pulp mill was taken into benchmarking. This way, the different efficiency of bark gas fuelled lime kiln compared to a one using liquid fuels can also be taken into consideration. In Joutseno, a combined bark dryer and gasification plant produces 48 MW of thermal power, which fuels up a lime kiln with a design capacity of 600 tons of Calcium oxide (CaO) per day. With a presumed design factor of 90%, this equals an average daily production of 540 tons CaO. Wet bark intake is 22 tons per hour and the dryer evaporation capacity is 12 tons per hour. Thus 54.5% of the original content of bark is evaporated, and the gasifier hourly intake is 10 tons of dry bark (Andritz, 2013).

In the model mass balance, the lime kiln average capacity is 832 tons of CaO per day. The gasification plant capacity in scenario was calculated based on the ratio of lime kiln capacities, calculated in equation 1:

$$\text{Scale ratio} = \frac{\text{Average capacity of the lime kiln in the scenario}}{\text{Average capacity of the lime kiln in Joutseno}} = \frac{540}{832} \approx 0.65 \quad (1)$$

Changes in the efficiency of drying and gasification due to size variation were neglected. Thus, in this scenario the gasification capacity is 74.0 thermal MW and the feed to the gasifier is 15.4 tons per hour. As the bark moisture content in this scenario is assumed to be only 50 w-%, which equals 15.4 tons per hour in the intake of the dryer, only that amount is evaporated. Thus, the bark gasification consumption is 30.8 tons of wet bark per hour. This amount of bark is diverted out of the biomass boiler heat production compared to the previous scenarios, which also decreases the capacity need of the boiler.

Dryer heat demand was estimated to equal the heat needed to warm up the water and bark from an average temperature of 3 °C to 100 °C, and then the heat needed to evaporate the water. The equipment heat transfer efficiency was estimated to be 75%. Heat demand Q is calculated with the Equation 2:

$$Q = \frac{1}{\eta} * \left(\frac{\dot{m}_W}{\Delta H_W} + \frac{\dot{m}_B}{\Delta H_B} \right) \quad (2)$$

where, Q is system heat demand

η is the efficiency of the heat transfer system

\dot{m}_W is the mass flow of the evaporated water

ΔH_W is enthalpy change of the evaporated water

\dot{m}_B is the mass flow of bark into the gasifier

ΔH_B is the enthalpy change of bark during drying

With the equation above, the heat demand was estimated to be 16.2 MW. The electricity demand was evaluated to be 1200 kW. Using these heat and power consumption figures, the energy balance from the conventional scenario was modified for this scenario. The energy balance is presented in Appendix 7, and the key production figures in Table 6 in section 8.4 together with other scenarios.

8.3 Combined depitching and gasification

The last competitive scenario is an addition of a smaller bark gasifier to the depitching case to fully replace the use of HFO in lime kiln. This would be a tall oil refining option in which carbon neutral operation is also reached in the pulp mill. It was presumed that simultaneous burning of TOP and bark gas is possible without any additional losses. Energy balance for this scenario was done by calculating bark consumption figures equalling the remaining HFO usage in the depitching case. Same specific consumption figures were used as in the other bark gasification scenario. Thus, this scenario consists of a gasifier with a thermal capacity of 47.6 MW and a consumption of 19.8 tons of wet bark per hour. This was taken into consideration in the biomass boiler sizing and heat production. For this scenario, an energy balance can be found in Appendix 8 and the key production figures in Table 6 in section 8.4 together with the other scenarios.

8.4 Production figures for the scenarios

The production figures of the scenarios are gathered and presented in Table 6. As the conventional scenario depicts a traditional pulp mill without additional upgrades, it is presented first in the Table. The figures are analysed in section 10.1.

Table 6. Key production figures of each scenario.

Name	Conventional	Depitching	Bark gasification	Depitching + Gasification	Unit
Pulp production	1 000 000	1 000 000	1 000 000	1 000 000	ADt/a
CTO production	141	0	141	0	t/d
DCTO production	0	98	0	98	t/d
Heat generation	775	775	714	736	MW
Process heat consumption	378	379	394	389	MW
Electricity generation	220	219	194	201	MW
Process electricity consumption	82	82	81	81	MW
Electricity sales	138	137	111	120	MW
Fuel oil consumption in lime kiln	118	76	0	0	t/d

9 Economic evaluation

9.1 Investment costs

Investment cost needs to be evaluated to compare economic feasibility. In this study, the investment costs were estimated based on the cost data of Pöyry. As the accuracy of these estimations was rough, a contingency of 20% was added to each of them. As the conventional scenario is featuring a traditional pulp mill, it was used as a base case for investment costs to which other scenarios were compared to.

The investment cost of a depitching plant was estimated for equipment shown in Figure 23. The size of the depitching department is relatively small, so the assumption was made that no investments are needed to common mill site resources or to an individual building. Due to the corrosiveness of the depitching process, special steel alloys such as 904L and 254 SMO were used in the estimation. This increased the investment cost, which was estimated to be 2.2 million euros (M€).

The bark gasification equipment is significantly larger by area than depitching. As the process flows of bark gasification are quite large too, it also needs investments into civil works and into some common mill resources such as instrument air capacity. Due to the significant bark consumption of the gasifier, this scenario needs a significantly smaller biomass boiler than the conventional scenario. The smaller boiler investment cost was

compensated based on steam production to match the lower capacity demand of this scenario. This compensation was excluded from the contingencies basis, as the contingencies related to biomass boiler are included in the conventional scenario. Differences in lime kiln sizing due to gaseous fuel were not taken into consideration in these calculations. The investment cost is estimated to be 17.1 M€.

For the combined depitching and gasification scenario, the investment cost was estimated based on previous scenarios. Price of the bark gasification equipment was scaled from the stand-alone gasification scenario, and summed with the depitching case. Indirect costs are higher than in previous scenarios due to the extent of this investment. In the study, this scenario was also given compensation from the smaller biomass boiler capacity demand. Like in the previous scenario, the compensation was not included in the contingencies basis. The investment cost for this scenario is estimated to be 19.6 M€.

Cost breakdown of the investment calculations can be found in Appendix 9.

9.2 Operational revenue

Operational revenues are based on the difference between production volumes presented in Table 6. Conventional scenario was taken as a base case and the operational revenues of other scenarios were compared against it. Maintenance and staff costs were estimated to be the same for all the scenarios as they can be efficiently integrated to other mill services. The possible taxes and other overheads were also neglected as they are very location and situation specific. In this comparison, price variables were considered for three parameters: tall oil (either crude or depitched) production, electricity sales and fuel oil consumption. Operational revenue dependence of these variables is shown in Figures 25–27. In each of the Figures, the other two variables were set to current market prices shown in Table 7.

Table 7. North Europe market prices for scenario variables.

Commodity	Price	Reference
CTO	410 €/ton	(Poikolainen, 2018)
DCTO	1.3 * CTO	*
Electricity	33 €/MWh	(Nord Pool AS, 2018)
HFO	330 €/ton	(Poikolainen, 2018)

* K. Ukkonen, personal communication, 9 April 2018.

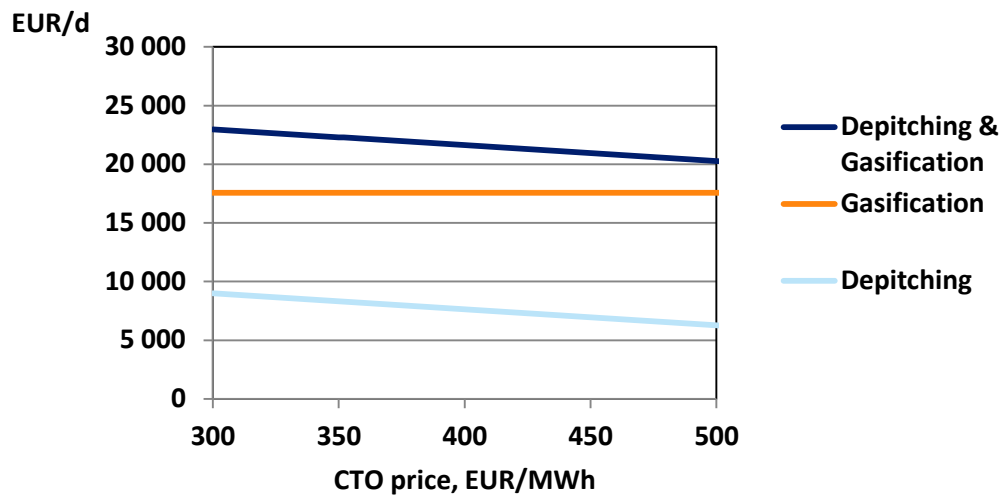


Figure 27. Sensitivity of operational revenue to CTO price.

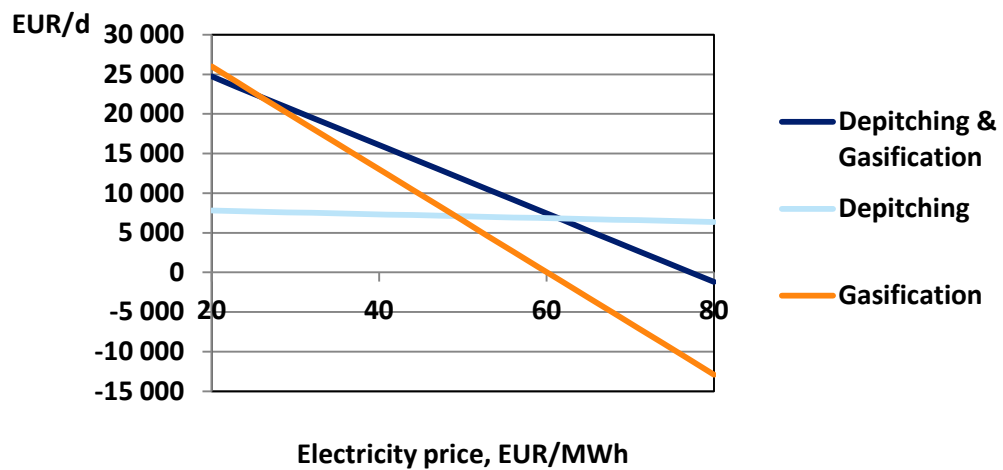


Figure 28. Sensitivity of operational revenue to electricity price.

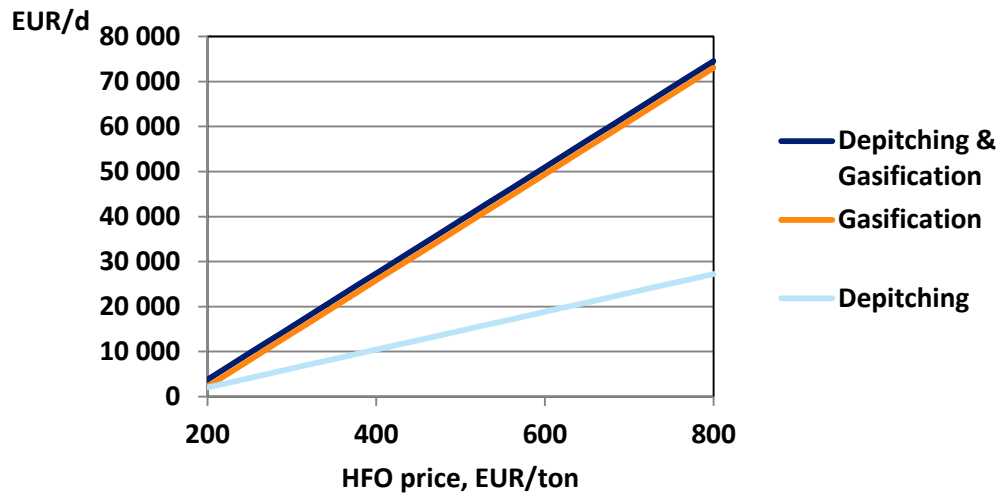


Figure 29. Sensitivity of operational revenue to HFO price.

9.3 Investment payback period

Using these previously calculated operational revenue and investment cost calculations, an investment payback period was calculated by dividing an investment cost with annual operational revenue. In this study, annual operational revenue was obtained by multiplying the daily revenues of scenarios with the operation days of the mill: In scenarios, it is 350 days per year. This calculation used the current market prices of the products. The results are presented in Table 8.

Table 8. Investment costs, annual operational revenues and payback periods for different scenarios.

	Investment cost, €	Annual operational revenue, €	Payback period, years
Depitching	2 221 000	2 622 200	0.84
Gasification	17 105 000	6 144 600	2.78
Combined depitching and gasification	19 618 000	6 687 800	2.93

10 Results

10.1 Technical feasibility

From the two common depitching methods, depitching with a thin film evaporator was evaluated to be more applicable in pulp mills based on the Literature Review. From the production figures in Table 6, it can be noted that tall oil depitching slightly increases process heat consumption and simultaneously lowers electricity generation and sales. Compared to gasification scenario, these changes are minimal, as gasification both needs a significant amount of process heat to bark drying, and on the same time diverts bark away from general process heat production and into lime kiln heating. However, the tall oil pitch is only able to supply 36% of heat energy needed in the lime kiln: During normal operation, additional fuels are needed to produce the rest.

Combined depitching and gasification scenario was introduced to produce depitched tall oil and to reach carbon neutral operation in lime kiln. Compared to just gasification, the drop in pulp mill process heat generation is lower because a smaller amount of bark is diverted from the biomass boiler. Mostly due to this, the combined scenario produces more electricity.

It is interesting to notice that the electricity consumption of the pulp mill decreased in the gasification scenarios, compared to conventional and depitching scenarios. This is mostly due to decreased energy consumption in the cooling water towers, as the use of a condensing tail is lowered. However, this change is small compared to the ones mentioned before.

10.2 Economic feasibility

The figures that are used to analyse the economic feasibility of the scenarios are gathered in Table 8. As can be noticed from the figures, depitching has the lowest investment cost. This is due to the small size of the equipment and overall process area. The investment costs of gasification and combined processes were significantly higher due to the size of the dryer and the gasifier in these processes. The difference in

investment cost between these two is small and in favour of the standalone gasification. However, the difference is only about half of the contingencies given for the scenarios, so it is inconclusive.

In the operational revenue comparison, all the advanced scenarios generated better revenue than the conventional scenario in most of the situations studied. The bark gasification and combined depitching and gasification scenarios produced higher revenues in most of the situations. This is mostly due to their independency from heavy fuel oil. Increased HFO price did not change the profitability of these scenarios; it made the alternative cost higher, in other words the other scenarios less profitable. With electricity price, there is similar but inverse effect: Higher electricity prices lower the profitability of gasification scenarios, as they have lower electricity production capacity.

From the CTO price sensitivity analysis in Figure 27, it can be noticed that high CTO price lowers the profitability of DCTO production scenarios. This is because of the higher price of DCTO is not capable of compensating all the loss in production volumes. Thus, the profitability of these scenarios is relying on savings acquired through TOP utilisation.

Comparison between payback periods reveals that the payback period of depitching is less than 1/3 of the time needed for other scenarios. This is again due to the low investment costs of the equipment. Between the two gasification solutions, there is a difference in favour of standalone bark gasification.

11 Conclusions

This study aimed to evaluate the technical and economic profitability of tall oil depitching in a pulp mill. Technically, tall oil depitching is easy to perform in a pulp mill. The equipment and process area needed are somewhat small, as is its influence on overall pulp mill product streams: The heat and electricity consumptions of the process are negligible. It also has low investment cost and investment payback period. However, it is not able to produce enough pitch to fuel the lime kiln completely.

To reach carbon neutral operation in the lime kiln, a bark gasifier could be added to produce the remaining energy needed. Comparison between bark gasification and

combined depitching and gasification revealed that even though the combined scenario has higher investment cost, the investment payback periods for the scenarios were relatively close to each other. Combined depitching and gasification was not as sensitive to increased electricity price as the standalone gasification, but it was more sensitive to CTO price increase.

Based on the balance comparisons and calculations made in this thesis, it can be concluded that tall oil depitching is a feasible process for the pulp mill, with or without supplementary bark gasification. It helps the mill to reach carbon neutral operation, but it is not a complete solution for this target on its own. Especially it gives more security against electricity price increase when alternative costs are taken into account. In the future, the depitching could be further optimised for different needs, for example to tailor raw materials for biofuels production or distillation, or to phytosterols extraction.

It should be noted that the calculations made in the experimental study contain a vast amount of assumptions and exclusions, most of which are location and case specific. This work offers a comparison between different process scenarios, but the profitability of each scenario should be evaluated case by case.

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Appendix 9: Investment cost breakdown of different scenarios

PULP MILL MAIN DIMENSIONING FIBRE AND CHEMICAL BALANCES INPUT VALUES

**OY DIPPA AB
MILL**

Thesis

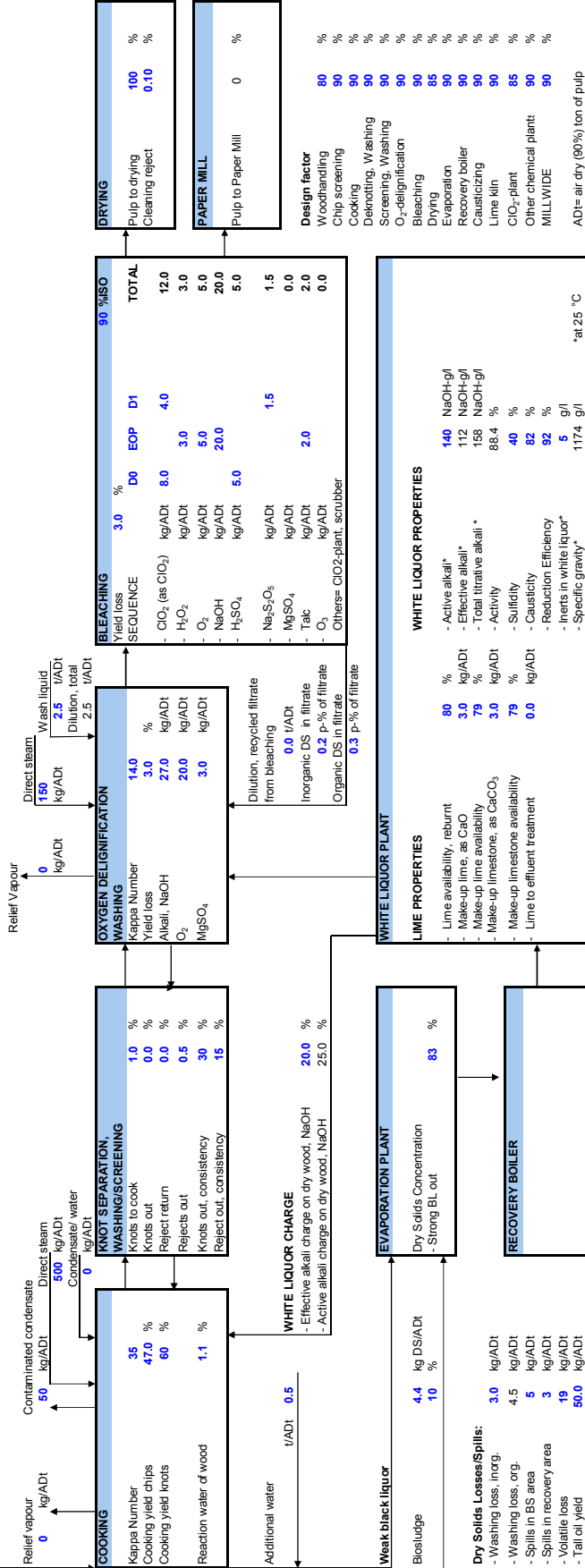
MODEL MILL

PROCESS / WOOD
Sulfate Pine
D0-EOP-D1
ANNUAL PRODUCTION
1 000 000 ADt/a
OPERATION DAYS
350 d/a
Average production
2 857 ADt/d
Design production
3 175 ADt/d

WOOD TO THE MILL
Purchased chips 0 %
Unbarked wood 100 %
Forested/barked wood 0 %
Fresh Roundwood Characteristics
- Wood moisture 50 w-%
- Wood density 410 kgBD/ m³sub
- Bark content, unbarked wood 12 V-%
- Bark density 380 kgBD/ m³s
- Bark content of chips to dig 0.5 w-%

WOOD PROCESSING
Wood losses 0.1 V-%
- Woodyard 1.5 V-%
- Woodhandling 0.2 V-%
- Chip storage 1.0 V-%
TOTAL LOSS 2.8 V-%
Operation days 350 d/a
Operation time wood/chipping 24 h/d
Operation time at chip screening 24 h/d

Storage times
- Roundwood storage 21 d
- Chip storage 5 d
- Hog fuel storage 7 d

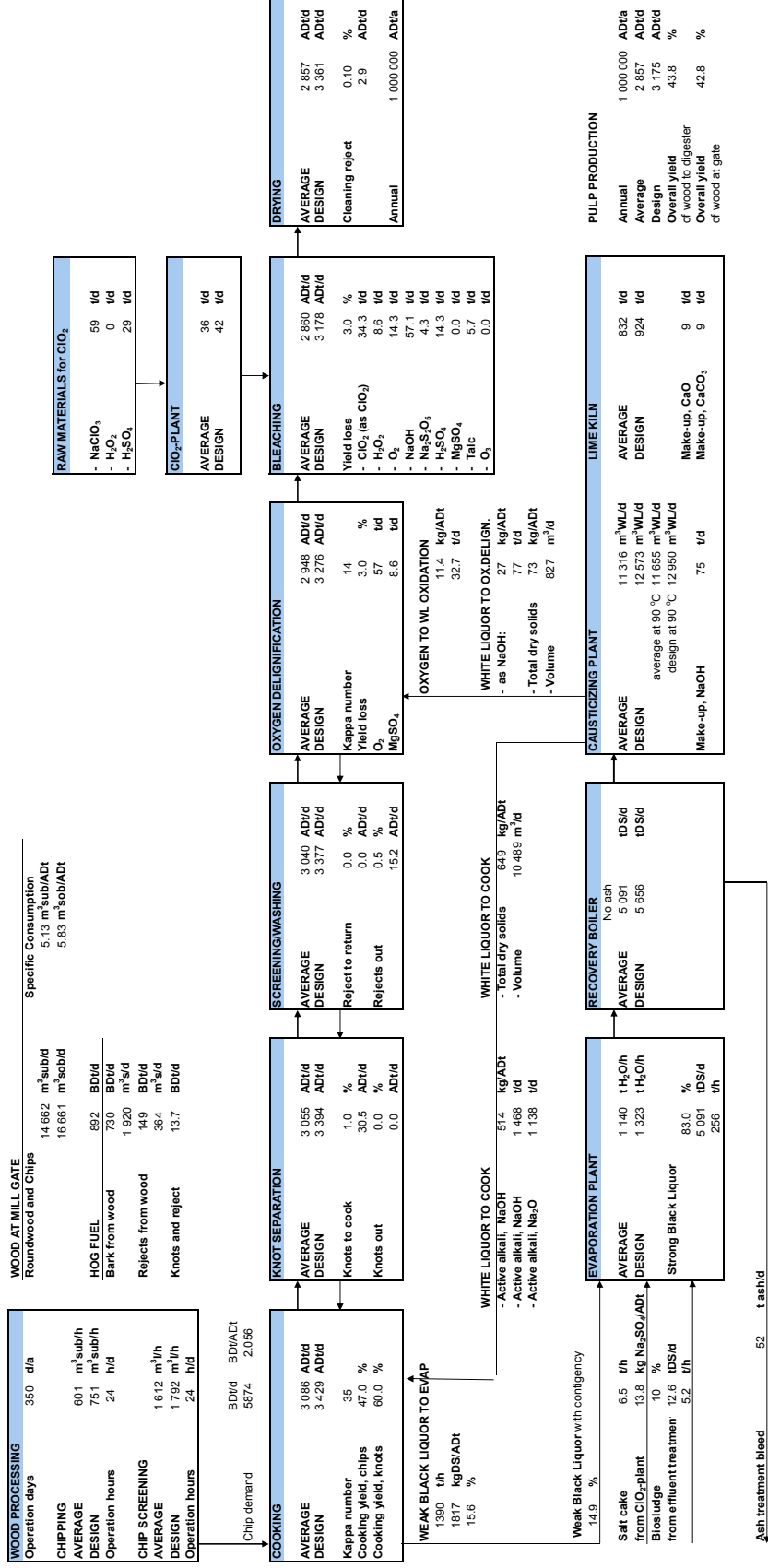


PULP MILL DIMENSIONING FIBRE AND CHEMICAL BALANCES DIMENSIONING

OY DIPPA AB
MILL

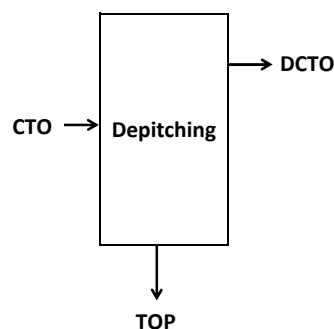
Thesis
MODEL MILL
Sulfate Pine
D0-EOP-D1

PROCESS / WOOD
BLEACHING



Appendix 2: Tall oil chemical constituents

kg/t CTO	CTO	TOP	DCTO
Resin acids	369	8.9	345
Secodehydroabietic	4.5	0	3.7
Pimaric	33.4	0.3	29.6
Isopimaric	26.3	0.4	22.9
Sandaracopimaric	6.8	0.1	6
Palustric	31	0.9	37.2
Abietic	171	3.9	150
Neobietic	37	0.6	16.3
7,9-Abietadienoic acid & other abietadienoic	0	0	2.8
Dehydroabietic & similar	58.8	2.7	76.5
Fatty acids	256	1.1	248
Palmitic	8	0	8.1
17ai (methylhexadecanoic acid)	3.6	0	3.7
Pinolenic	30.5	0.1	29.3
Linoleic	95.9	0.3	92.9
Oleic	67.3	0.3	69
Stearic	2.3	0	2.3
Cyclopinoic	0.3	0	1.3
18:3 ttt (eleostearic acid)	8.4	0	5.8
18:2 tt (linolelaidic acid)	12.3	0.1	10.7
Eicosatrienoic	12.1	0.1	11.6
20:2	2.8	0	2.5
20:1	1.5	0	1.3
Arachidic	2.8	0	2.7
Behenic	5.5	0.1	5
Lignoceric	2.4	0.1	1.7
As esters	39.2	23.2	2
Palmitic	1.3	1	0
17ai (methylhexadecanoic acid)	0.6	0.4	0
Pinolenic	4.2	1.5	0.3
Linoleic	14.4	6	0.8
Oleic	9.8	8.2	0.7
Stearic	0.5	0.4	0
18:3 ttt (eleostearic acid)	1.1	1	0
18:2 tt (linolelaidic acid)	1.8	1.9	0.1
Eicosatrienoic	3.4	0.8	0.1
20:2	0.4	0	0
20:1	0.2	0	0
Arachidic	0.4	0.4	0
Behenic	0.8	1.1	0
Lignoceric	0.3	0.5	0
Unsaponifiables	118	88.8	50.8
Pimaradiene	1	0	1
Hydrocarbon, MW=256-258	0	0	4.1
Pimaral	5.7	0	5.7
Isopimaral	2.1	0	2.1
Pinosylvin dimethyl ether	1.5	0	1.5
Dehydroabietal	1	0	1
Abietal	2.2	0	2.2
Squalene	2	0	1.7
3,5-stigmastadiene	0.3	1.7	3.3
β -sitosterol	4.4	0.4	0.6
As esters	33.7	22.8	0
Dehydroabietic methyl ester	0.2	0	0
Abietic methyl ester	0.2	0	0
Pimarol	5.2	2	0
Docosanol	1.7	1.2	0
Tatracosanol	1.7	1.1	0
β -sitostearate	24.7	18.5	0
Total	718	58.1	618
<i>Pitch, unidentified</i>	235	235	0
Fraction yield, kg/t CTO	1000	293	686
% identified	71.8	19.8	90.1
%	95.3	100	90.1



Appendix 3: Tall oil model

	CTO	TOP	DCTO	% to TOP	% to DCTO	Model component
Resin acids	354	8.9	345	2.5	97.5	
Secodehydroabietic	3.7	0	3.7	0.0	100.0	
7,9-Abietadienoic acid & other abietadienoic	2.8	0	2.8	0.0	100.0	Pimaric acid
Pimaric	29.9	0.3	29.6	1.0	99.0	
Sandaracopimaric	6.1	0.1	6	1.6	98.4	
Isopimaric	23.3	0.4	22.9	1.7	98.3	Abietic acid
Palustric	38.1	0.9	37.2	2.4	97.6	
Abietic	154	3.9	150	2.5	97.5	
Dehydroabietic & similar	79.2	2.7	76.5	3.4	96.6	Neoabietic acid
Neoabietic	16.9	0.6	16.3	3.6	96.4	
Fatty acids	249	1.1	248	0.4	99.6	
Palmitic	8.1	0	8.1	0.0	100.0	
17ai (methylhexadecanoic acid)	3.7	0	3.7	0.0	100.0	
Stearic	2.3	0	2.3	0.0	100.0	Palmitic acid
Cyclopinoleic	1.3	0	1.3	0.0	100.0	
18:3 ttt (eleostearic acid)	5.8	0	5.8	0.0	100.0	
20:2	2.5	0	2.5	0.0	100.0	
20:1	1.3	0	1.3	0.0	100.0	
Arachidic	2.7	0	2.7	0.0	100.0	
Linoleic	93.2	0.3	92.9	0.3	99.7	Oleic acid
Pinolenic	29.4	0.1	29.3	0.3	99.7	
Oleic	69.3	0.3	69	0.4	99.6	
Eicosatrienoic	11.7	0.1	11.6	0.9	99.1	Behenic
18:2 tt (linolelaidic acid)	10.8	0.1	10.7	0.9	99.1	
Behenic	5.1	0.1	5	2.0	98.0	
Lignoceric	1.8	0.1	1.7	5.6	94.4	Lignoceric
As esters	25.2	23.2	2	92.1	7.9	
Pinolenic	1.8	1.5	0.3	83.3	16.7	Pitch
Linoleic	6.8	6	0.8	88.2	11.8	
Eicosatrienoic	0.9	0.8	0.1	88.9	11.1	
Oleic	8.9	8.2	0.7	92.1	7.9	
18:2 tt (linolelaidic acid)	2	1.9	0.1	95.0	5.0	
Palmitic	1	1	0	100.0	0.0	
17ai (methylhexadecanoic acid)	0.4	0.4	0	100.0	0.0	
Stearic	0.4	0.4	0	100.0	0.0	
18:3 ttt (eleostearic acid)	1	1	0	100.0	0.0	
20:2	0	0	0	0.0	0.0	
20:1	0	0	0	0.0	0.0	
Arachidic	0.4	0.4	0	100.0	0.0	
Behenic	1.1	1.1	0	100.0	0.0	
Lignoceric	0.5	0.5	0	100.0	0.0	
Unsaponifiables	140	88.8	50.8	63.6	36.4	
Pimaradiene	1	0	1	0.0	100.0	Heads
Pimaral	5.7	0	5.7	0.0	100.0	
Isopimaral	2.1	0	2.1	0.0	100.0	
Pinosylvin dimethyl ether	1.5	0	1.5	0.0	100.0	
Dehydroabietal	1	0	1	0.0	100.0	
Abietal	2.2	0	2.2	0.0	100.0	
Squalene	1.7	0	1.7	0.0	100.0	
Hydrocarbon, MW=256-258	4.1	0	4.1	0.0	100.0	
3.5-stigmastadiene	5	1.7	3.3	34.0	66.0	Sitosterol
β-sitosterol	1	0.4	0.6	40.0	60.0	
As esters	22.8	22.8	0	100.0	0.0	
Dehydroabietic methyl ester	0	0	0	0.0	0.0	Pitch
Abietic methyl ester	0	0	0	0.0	0.0	
Pimarol	2	2	0	100.0	0.0	
Docosanol	1.2	1.2	0	100.0	0.0	
Tatracosanol	1.1	1.1	0	100.0	0.0	
β-sitostearate	18.5	18.5	0	100.0	0.0	
Total	676	58.1	618	8.6	91.4	
<i>Pitch, unidentified</i>	222	197	25	88.739	11.3	Pitch / Heads
Fraction yield, kg/t CTO	1000	293	686	29.3	68.6	
% identified	67.6	19.8	90.1			
%	89.8	87.1	93.7			

Appendix 4: Mill energy balance for Depitching

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Tall oil Depitching		11.4.2018

ENERGY BALANCES

Design Production	ADt/d	3 175	
Average Production	ADt/d	2 857	
Operational days		350	
Daily Production	ADt/d	1 000 000	

	Steam	Dept. Production t/d	Specific Consumption		Condensate return		Heat Flow MJ/s	Steam Flow kg/s	Cond. Flow kg/s
			GJ/Ut	GJ/ADt	%	Deg.C			
Woodhandling, m3 sub/d	LP	14 662	0.00	0.00	0	0	0.0	0.00	0.0
Cooking	MP1	3 086	1.68	1.81	22	100	60.0	22.08	4.9
Cooking	LP	3 086	0.00	0.00	0	0	0.0	0.00	0.0
O2 delignification	MP1	2 948	0.18	0.19	0	0	6.1	2.19	0.0
O2 delignification	LP	2 948	0.40	0.41	0	0	13.6	4.95	0.0
Bleaching	MP1	2 860	0.06	0.06	0	0	2.0	0.71	0.0
Bleaching	LP	2 860	0.19	0.19	0	0	6.2	2.23	0.0
Drying machine	LP	2 857	2.50	2.50	85	98	82.7	34.29	29.1
ClO2 plant	MP1	36	8.50	0.11	0	0	3.5	1.26	0.0
ClO2 plant	LP	36	25.00	0.32	100	100	10.4	4.45	4.4
Evaporation plant, tH2O/h	MP1	1 140	2.00	0.80	100	100	26.4	11.04	11.0
Evaporation plant, tH2O/h	LP	1 140	8.00	3.19	100	100	105.6	45.09	45.1
Causticizing/Lime kiln	MP1	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Causticizing/Lime kiln	LP	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Recovery boiler, sootblowing	MP2	5 091	0.32	0.57	0	0	18.9	6.05	0.0
Recovery boiler	MP2	5 091	0.00	0.00	100	145	0.0	0.00	0.0
Recovery boiler	MP1	5 091	0.26	0.46	90	145	15.3	6.78	6.1
Recovery boiler	LP	5 091	0.33	0.59	100	145	19.4	9.05	9.0
Biomass boiler, t/h	LP	1 816	0.02	0.01	100	100	0.5	0.21	0.2
Tall oil refining (acidulation)	LP	141	0.63	0.03	0	0	1.0	0.37	0.0
Tall oil refining (depitching)	MP3	98	0.80	0.03	100	100	0.9	0.4	0.4
Pulp Mill Common	MP1	2 857	0.10	0.10	0	100	3.3	1.18	0.0
Pulp Mill Common	LP	2 857	0.10	0.10	0	100	3.3	1.20	0.0
Total (including chemical plant)	MP3			0.03	100	100.00	0.9	0.4	0.4
	MP2			0.57	0	0.00	18.9	6.1	0.0
	MP1			3.53	49	112.48	116.7	45.2	22.0
	LP			7.34	86	103.97	242.7	101.8	87.9
TOTAL				11.47	72	105.65	379.2	153.5	110.3
FEEDWATER	LP							21.7	
LP Heater	LP				100	104		16.2	16.2
Condensing TG					100	44		88.7	88.7

STEAM & WATER PROPERTIES

	bar(abs)	Deg.C
Recovery Boiler	102.00	505
Biofuel Boiler	102.00	505
Steam to Turbine	99.00	505
MP3	99.00	320
MP2 steam	30.00	350
MP1 steam	11.00	195
LP steam	4.50	155
LP heater pressure	1.20	
Condenser pressure	0.090	
FW in tank		143.5
FW for RB boiler		145
FW for Biofuel Boiler		145
Demi water		25
Heated condensate & Demin. Water		102

MAKE-UP WATER DEMAND

	kg/s
Steam from boilers	278.7
Steam desuperheating	1.4
Feedwater from fw tank	280.1
Mill condensate return	126.5
Condenser condensate return	88.7
Feedwater preheat	21.7
Make-up water demand	43.2

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Tall oil Depitching		11.4.2018

HEAT GENERATION

RECOVERY BOILER

Dry solids ratio	tDS/Adt	1.782
Dry solids (virgin)	tDS/d	5 091
Heat value, higher	GJ/tDS	14.0
Efficiency	%	75.8

BIOFUEL BOILER

		Biomass		
		Bark	Wood Rejects	Pulp Rejects
Fuel flow	tDS/d	730.0	149.0	13.7
	t/d	1460	310	46
	GJ/d as fired	12 045	2 383	126
Dry solids content	%ds	50.0	48.0	30.0
Heat value, effective	GJ/tds	19.0	18.7	15.0
	GJ/t	8.3	7.7	2.8
Efficiency	%	87.0	87.0	87.0

HEAT FROM BOILERS

	GJ/Adt	MJ/s	kg/s	t/h
Black Liquor (incl. methanol and CNCG)	18.9	624.9	225.8	812.8
Biofuel	4.4	146.5	53.0	190.6
Total	23.3	771.5	278.7	1003.4

HEAT BALANCE

	GJ/Adt	MJ/s
Heat Generation		
- Black liquor	18.9	624.9
- Biofuel	4.4	146.5
- In feedwater pumps	0.1	3.8
Total	23.4	775.2
Heat Consumption		
- Pulp mill process	11.5	379.2
- BP power generation	3.6	118.2
- Cond. power generation	8.4	277.9
Total	23.4	775.3

POWER BALANCE

POWER CONSUMPTION		DESIGN			BALANCE		Power cons. kW
		Design production t/d	Specif. cons. kWh/Ut	Specif. cons. kWh/ADt	Balance production t/d	Specif. cons. kWh/ADt	
Wood handling	m ³ /sub d	18 328	4.0	23.1	14 662	24	2 841
Cooking	Adt/d	3 429	27.0	29.2	3 086	31	3 684
O2 delignification and bleaching	Adt/d	3 178	88.0	88.1	2 860	93	11 128
Drying machine	Adt/d	3 361	100.0	105.9	2 857	110	13 060
Evaporation Plant	t _{H2O} /h	1 267	74.0	29.5	1 140	32	3 788
Recausticizing plant & Lime kiln	m ³ WL/d	12 573	6.5	25.7	11 316	27	3 252
Recovery Boiler	tDS/d	5 657	46.0	82.0	5 091	87	10 354
Biomass Boiler	t _{steam} /d	5 083	18.0	28.8	4 575	31	3 641
Raw water treatment	m ³ /d	79 365	0.1	1.3	71 429	1.3	158
Raw water intake	l/s	79 365	0.1	1.3	71 429	1	158
Deminerlizing plant	m ³ /d	5 104	0.5	0.8	4 594	1	102
Effluent treatment	m ³ /d	63 492	1.0	20.0	57 143	21	2 526
Cooling water, tower & pumps	m ³ /d	536 851	0.3	42.3	483 166	45	5 341
Turbine plant	Adt/d	3 175	1.0	1.0	2 857	1	126
Compressed air	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Chiller for AC	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Pulp mill common	Adt/d	3 175	12.0	12.0	2 857	13	1 516
Tall oil acidulation	t/d	157	27.3	1.3	141	1	170
Tall oil depitching	t/d	109	9.5	0.3	98	0	42
Oxygen plant	tO2/d	116	560.0	20.4	104.1	22	2 577
ClO2-plant	tClO2/d	42	9170.0	122.3	36	129.8	15 454
PULP MILL TOTAL				651	688		81 900

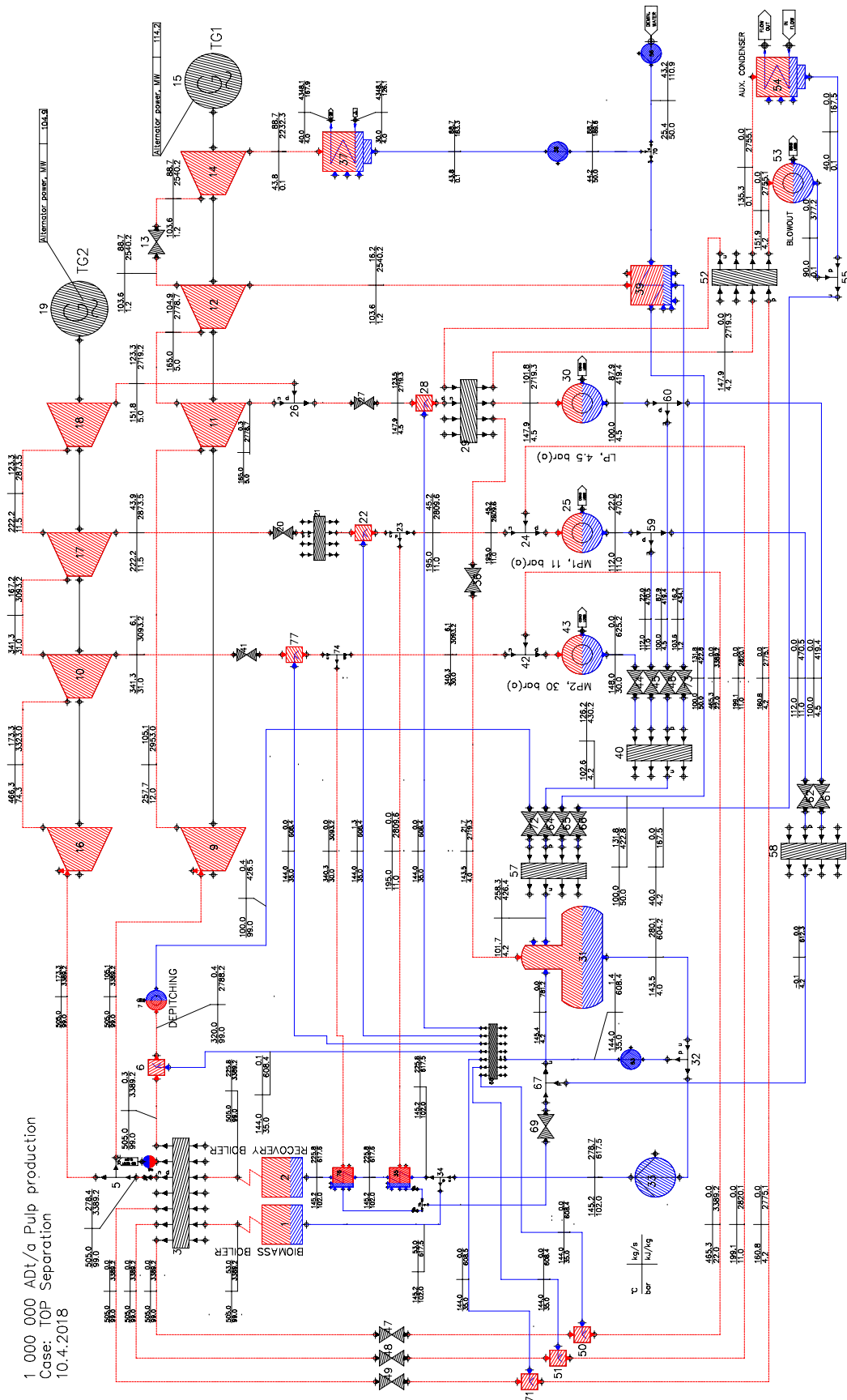
POWER GENERATION

	kWh/ADt	MW
TG1	959	114.2
TG2	881	104.9
Total from turbogenerators	1 840	219.1

POWER BALANCE

	kWh/ADt	MW
Power consumption	688	81.9
Power generation	1 840	219.1
Power sales	1 152	137.2

Appendix 5: Electricity generation and steam distribution for the depitching case



Appendix 6: Mill energy balance for Conventional Case

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Conventional mill Average		3.4.2018

ENERGY BALANCES

Design Production	ADt/d	3 175	
Average Production	ADt/d	2 857	
Operational days		350	
Daily Production	ADt/d	1 000 000	

	Steam	Dept. Production t/d	Specific Consumption		Condensate return		Heat Flow MJ/s	Steam Flow kg/s	Cond. Flow kg/s
			GJ/Ut	GJ/ADt	%	Deg.C			
Woodhandling, m3 sub/d	LP	14 662	0.00	0.00	0	0	0.0	0.00	0.0
Cooking	MP1	3 086	1.68	1.81	22	100	60.0	22.08	4.9
Cooking	LP	3 086	0.00	0.00	0	0	0.0	0.00	0.0
O2 delignification	MP1	2 948	0.18	0.19	0	0	6.1	2.19	0.0
O2 delignification	LP	2 948	0.40	0.41	0	0	13.6	4.95	0.0
Bleaching	MP1	2 860	0.06	0.06	0	0	2.0	0.71	0.0
Bleaching	LP	2 860	0.19	0.19	0	0	6.2	2.23	0.0
Drying machine	LP	2 857	2.50	2.50	85	98	82.7	34.29	29.1
ClO2 plant	MP1	36	8.50	0.11	0	0	3.5	1.26	0.0
ClO2 plant	LP	36	25.00	0.32	100	100	10.4	4.45	4.4
Evaporation plant, tH2O/h	MP1	1 140	2.00	0.80	100	100	26.4	11.04	11.0
Evaporation plant, tH2O/h	LP	1 140	8.00	3.19	100	100	105.6	45.09	45.1
Causticizing/Lime kiln	MP1	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Causticizing/Lime kiln	LP	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Recovery boiler, sootblowing	MP2	5 091	0.32	0.57	0	0	18.9	6.05	0.0
Recovery boiler	MP2	5 091	0.00	0.00	100	145	0.0	0.00	0.0
Recovery boiler	MP1	5 091	0.26	0.46	90	145	15.3	6.78	6.1
Recovery boiler	LP	5 091	0.33	0.59	100	145	19.4	9.05	9.0
Biomass boiler	LP	1 816	0.02	0.01	100	100	0.5	0.21	0.2
Tall oil acidulation	LP	141	0.63	0.03	0	0	1.0	0.37	0.0
Pulp Mill Common	MP1	2 857	0.10	0.10	0	100	3.3	1.18	0.0
Pulp Mill Common	LP	2 857	0.10	0.10	0	100	3.3	1.20	0.0
Total	MP3			0.00	0	0.00	0.0	0.0	0.0
	MP2			0.57	0	0.00	18.9	6.1	0.0
	MP1			3.53	49	112.48	116.7	45.2	22.0
	LP			7.34	86	103.97	242.7	101.8	87.9
TOTAL				11.44	72	105.67	378.2	153.1	109.9
FEEDWATER	LP							21.8	
LP Heater	LP				100	104		16.3	16.3
Condensing TG					100	44		89.3	89.3

STEAM & WATER PROPERTIES		
	bar(abs)	Deg.C
Recovery Boiler	102.00	505
Biofuel Boiler	102.00	505
Steam to Turbine	99.00	505
MP3 steam	99.00	320
MP2 steam	30.00	350
MP1 steam	11.00	195
LP steam	4.50	155
LP heater pressure	1.20	
Condenser pressure	0.090	
FW in tank		143.5
FW for RB boiler		145
FW for Biofuel Boiler		145
Demi water		25
Heated condensate & Demin. Water		102

MAKE-UP WATER DEMAND		kg/s
Steam from boilers		278.7
Steam desuperheating		1.6
Feedwater from fw tank		280.3
Mill condensate return		126.2
Condenser condensate return		89.3
Feedwater preheat		21.8
Make-up water demand		43.0

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Conventional mill Average		3.4.2018

HEAT GENERATION

RECOVERY BOILER

Dry solids ratio	tDS/Adt	1.782
Dry solids (virgin)	tDS/d	5 091
Heat value, higher	GJ/tDS	14.0
Efficiency	%	75.8

BIOFUEL BOILER

		Biomass		
		Biomass 1 Bark	Biomass 2 Wood Rejects	Biomass 3 Pulp Rejects
Fuel flow	tDS/d	730.0	149.0	13.7
	t/d	1460	310	46
	GJ/d as fired	12 045	2 383	126
Dry solids content	%ds	50.0	48.0	30.0
Heat value, effective	GJ/tDS	19.0	18.7	15.0
	GJ/t	8.3	7.7	2.8
Efficiency	%	87.0	87.0	87.0

HEAT FROM BOILERS

	GJ/Adt	MJ/s	kg/s	t/h
Black Liquor (incl. methanol and CNCG)	18.9	624.9	225.8	812.8
Biofuel	4.4	146.5	53.0	190.6
Total	23.3	771.5	278.7	1003.4

HEAT BALANCE

	GJ/Adt	MJ/s
Heat Generation		
- Black liquor	18.9	624.9
- Biofuel	4.4	146.5
- In feedwater pumps	0.1	3.8
Total	23.4	775.2
Heat Consumption		
- Pulp mill process	11.4	378.2
- BP power generation	3.6	117.9
- Cond. power generation	8.4	279.2
Total	23.4	775.3

POWER BALANCE

POWER CONSUMPTION		DESIGN			BALANCE		
		Design production t/d	Specif. cons. kWh/Ut	Specif. cons. kWh/ADt	Balance production t/d	Specif. cons. kWh/ADt	Power cons. kW
Wood handling	m ³ /sub d	18 328	4.0	23.1	14 662	24	2 841
Cooking	Adt/d	3 429	27.0	29.2	3 086	31	3 684
O2 delignification and bleaching	Adt/d	3 178	88.0	88.1	2 860	93	11 128
Drying machine	Adt/d	3 361	100.0	105.9	2 857	110	13 060
Evaporation Plant	t ₂₀ /h	1 267	74.0	29.5	1 140	32	3 788
Recausticizing plant & Lime kiln	m ³ WL/d	12 573	6.5	25.7	11 316	27	3 252
Recovery Boiler	tDS/d	5 657	46.0	82.0	5 091	87	10 354
Biomass Boiler	t/d	2 018	45.3	28.8	1 816	31	3 641
Raw water treatment	m ³ /d	79 365	0.1	1.3	71 429	1.3	158
Raw water intake	l/s	79 365	0.1	1.3	71 429	1	158
Deminerlizing plant	m ³ /d	5 091	0.5	0.8	4 582	1	101
Effluent treatment	m ³ /d	63 492	1.0	20.0	57 143	21	2 526
Cooling water, tower & pumps	m ³ /d	539 193	0.3	42.5	485 274	45	5 364
Turbine plant	Adt/d	3 175	1.0	1.0	2 857	1	126
Compressed air	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Chiller for AC	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Pulp mill common	Adt/d	3 175	12.0	12.0	2 857	13	1 516
Tall oil acidulation	t/d	157	27.3	1.3	141	1	170
Oxygen plant	tO ₂ /d	116	560.0	20.4	104	22	2 577
ClO ₂ -plant	tClO ₂ /d	42	9170.0	122.3	36	129.8	15 454
PULP MILL TOTAL				651	688		81 900

POWER GENERATION

	kWh/Adt	MW
TG1	880	104.8
TG2	965	114.9
Total from turbogenerators	1 845	219.7

POWER BALANCE

	kWh/ADt	MW
Power consumption	688	81.9
Power generation	1 845	219.7
Power sales	1 157	137.8

Appendix 7: Mill energy balance for Bark gasification

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Bark gasification		3.4.2018

ENERGY BALANCES

Design Production	ADt/d	3 175	
Average Production	ADt/d	2 857	
Operational days		350	
Daily Production	ADt/d	1 000 000	

	Steam	Dept. Production t/d	Specific Consumption		Condensate return		Heat Flow MJ/s	Steam Flow kg/s	Cond. Flow kg/s
			GJ/Ut	GJ/ADt	%	Deg.C			
Woodhandling, m3 sub/d	LP	14 662	0.00	0.00	0	0	0.0	0.00	0.0
Cooking	MP1	3 086	1.68	1.81	22	100	60.0	22.08	4.9
Cooking	LP	3 086	0.00	0.00	0	0	0.0	0.00	0.0
O2 delignification	MP1	2 948	0.18	0.19	0	0	6.1	2.19	0.0
O2 delignification	LP	2 948	0.40	0.41	0	0	13.6	4.95	0.0
Bleaching	MP1	2 860	0.06	0.06	0	0	2.0	0.71	0.0
Bleaching	LP	2 860	0.19	0.19	0	0	6.2	2.23	0.0
Drying machine	LP	2 857	2.50	2.50	85	98	82.7	34.29	29.1
ClO2 plant	MP1	36	8.50	0.11	0	0	3.5	1.26	0.0
ClO2 plant	LP	36	25.00	0.32	100	100	10.4	4.45	4.4
Evaporation plant, th2O/h	MP1	1 140	2.00	0.80	100	100	26.4	11.04	11.0
Evaporation plant, th2O/h	LP	1 140	8.00	3.19	100	100	105.6	45.09	45.1
Causticizing/Lime kiln	MP1	11 655	0.00	0.00	0	0	0.0	0.00	0.0
Causticizing/Lime kiln	LP	11 655	0.00	0.00	0	0	0.0	0.00	0.0
Recovery boiler, sootblowing	MP2	5 091	0.32	0.57	0	0	18.9	6.05	0.0
Recovery boiler	MP2	5 091	0.00	0.00	100	145	0.0	0.00	0.0
Recovery boiler	MP1	5 091	0.26	0.46	90	145	15.3	6.78	6.1
Recovery boiler	LP	5 091	0.33	0.59	100	145	19.4	9.05	9.0
Biomass boiler	LP	1 077	0.02	0.01	100	100	0.3	0.12	0.1
Tall oil acidulation	LP	141	0.63	0.03	0	0	1.0	0.37	0.0
Pulp Mill Common	MP1	2 857	0.10	0.10	0	100	3.3	1.18	0.0
Pulp Mill Common	LP	2 857	0.10	0.10	0	100	3.3	1.20	0.0
Bark Drying & Gasification, wet bark	LP	666	2.10	0.49	100	100	16.2	6.90	6.9
Total									
	MP3		0.00	0.00	0	0.00	0.0	0.0	0.0
	MP2		0.57	0.57	0	0.00	18.9	6.1	0.0
	MP1		3.53	49	112.48	116.7	45.2	22.0	22.0
	LP		7.82	87	96.40	258.7	108.6	94.8	94.8
TOTAL				11.92	73	99.43	394.2	159.9	116.8
FEEDWATER	LP							19.9	
LP Heater	LP				100	104		13.5	13.5
Condensing TG					100	44		64.9	64.9

STEAM & WATER PROPERTIES		
	bar(abs)	Deg.C
Recovery Boiler	102.00	505
Biofuel Boiler	102.00	505
Steam to Turbine	99.00	505
MP3 steam	99.00	320
MP2 steam	30.00	350
MP1 steam	11.00	195
LP steam	4.50	155
LP heater pressure	1.20	
Condenser pressure	0.090	
FW in tank		143.5
FW for RB boiler		145
FW for Biofuel Boiler		145
Demi water		25
Heated condensate & Demin. Water		102

MAKE-UP WATER DEMAND		kg/s
Steam from boilers		256.6
Steam desuperheating		1.7
Feedwater from fw tank		258.2
Mill condensate return		130.3
Condenser condensate return		64.9
Feedwater preheat		19.9
Make-up water demand		43.1

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Bark gasification		3.4.2018

HEAT GENERATION

RECOVERY BOILER

Dry solids ratio	tDS/Adt	1.782
Dry solids (virgin)	tDS/d	5 091
Heat value, higher	GJ/tDS	14.0
Efficiency	%	75.8

BIOFUEL BOILER

		Biomass		
		Bark	Wood Rejects	Pulp Rejects
Fuel flow	tDS/d	360.5	149.0	13.7
	t/d	721	310	46
	GJ/d as fired	5 948	2 383	126
Dry solids content	%ds	50.0	48.0	30.0
Heat value, effective	GJ/tds	19.0	18.7	15.0
	GJ/t	8.3	7.7	2.8
Efficiency	%	87.0	87.0	87.0

HEAT FROM BOILERS

	GJ/Adt	MJ/s	kg/s	t/h
Black Liquor (incl. methanol and CNCG)	18.9	624.9	225.8	812.8
Biofuel	2.6	85.2	30.8	110.8
Total	21.5	710.1	256.6	923.6

HEAT BALANCE

	GJ/Adt	MJ/s
Heat Generation		
- Black liquor	18.9	624.9
- Biofuel	2.6	85.2
- In feedwater pumps	0.1	3.5
Total	21.6	713.5
Heat Consumption		
- Pulp mill process	11.9	394.2
- BP power generation	3.5	117.1
- Cond. power generation	6.1	202.2
Total	21.6	713.5

POWER BALANCE

POWER CONSUMPTION		DESIGN			BALANCE		
		Design production t/d	Specif. cons. kWh/Ut	Specif. cons. kWh/ADt	Balance production t/d	Specif. cons. kWh/ADt	Power cons. kW
Wood handling	m ³ /sub d	18 328	4.0	23.1	14 662	24	2 841
Cooking	Adt/d	3 429	27.0	29.2	3 086	31	3 684
O2 delignification and bleaching	Adt/d	3 178	88.0	88.1	2 860	93	11 128
Drying machine	Adt/d	3 361	100.0	105.9	2 857	110	13 060
Evaporation Plant	t ₂₀ /h	1 267	74.0	29.5	1 140	32	3 788
Recausticizing plant & Lime kiln	m ³ WL/d	12 950	6.5	26.5	11 655	28	3 349
Recovery Boiler	tDS/d	5 657	46.0	82.0	5 091	87	10 354
Biomass Boiler	t/d	1 197	44.6	16.8	1 077	18	2 122
Raw water treatment	m ³ /d	79 365	0.1	1.3	71 429	1.3	158
Raw water intake	l/s	79 365	0.1	1.3	71 429	1	158
Deminerlizing plant	m ³ /d	5 098	0.5	0.8	4 589	1	101
Effluent treatment	m ³ /d	63 492	1.0	20.0	57 143	21	2 526
Cooling water, tower & pumps	m ³ /d	443 948	0.3	35.0	399 553	37	4 416
Turbine plant	Adt/d	3 175	1.0	1.0	2 857	1	126
Compressed air	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Chiller for AC	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Pulp mill common	Adt/d	3 175	12.0	12.0	2 857	13	1 516
Tall oil acidulation	t/d	157	27.3	1.3	141	1	170
Oxygen plant	tO2/d	116	560.0	20.4	104.1	22	2 577
ClO2-plant	tClO2/d	42	9170.0	122.3	36	129.8	15 454
Bark drying & Gasification	t/d, wet bark	821	36.6	9.5	739	10.2	1 209
PULP MILL TOTAL				642	678		80 800

POWER GENERATION

	kWh/ADt	MW
TG1	805	95.8
TG2	804	95.7
Total from turbogenerators	1 609	191.5

POWER BALANCE

	kWh/ADt	MW
Power consumption	678	80.8
Power generation	1 609	191.5
Total	930	110.7

Appendix 8: Mill energy balance for combined Depitching and Gasification

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Combined depitching and bark gasification		11.4.2018

ENERGY BALANCES

Design Production	ADt/d	3 175	
Average Production	ADt/d	2 857	
Daily Production	ADt/d	1 000 000	

	Steam	Dept. Production t/d	Specific Consumption		Condensate return		Heat Flow MJ/s	Steam Flow kg/s	Cond. Flow kg/s
			GJ/Ut	GJ/ADt	%	Deg.C			
Woodhandling, m3 sub/d	LP	14 662	0.00	0.00	0	0	0.0	0.00	0.0
Cooking	MP1	3 086	1.68	1.81	22	100	60.0	22.08	4.9
Cooking	LP	3 086	0.00	0.00	0	0	0.0	0.00	0.0
O2 delignification	MP1	2 948	0.18	0.19	0	0	6.1	2.19	0.0
O2 delignification	LP	2 948	0.40	0.41	0	0	13.6	4.95	0.0
Bleaching	MP1	2 860	0.06	0.06	0	0	2.0	0.71	0.0
Bleaching	LP	2 860	0.19	0.19	0	0	6.2	2.23	0.0
Drying machine	LP	2 857	2.50	2.50	85	98	82.7	34.29	29.1
ClO2 plant	MP1	36	8.50	0.11	0	0	3.5	1.26	0.0
ClO2 plant	LP	36	25.00	0.32	100	100	10.4	4.45	4.4
Evaporation plant, th2O/h	MP1	1 140	2.00	0.80	100	100	26.4	11.04	11.0
Evaporation plant, th2O/h	LP	1 140	8.00	3.19	100	100	105.6	45.09	45.1
Causticizing/Lime kiln	MP1	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Causticizing/Lime kiln	LP	11 316	0.00	0.00	0	0	0.0	0.00	0.0
Recovery boiler, sootblowing	MP2	5 091	0.32	0.57	0	0	18.9	6.05	0.0
Recovery boiler	MP2	5 091	0.00	0.00	100	145	0.0	0.00	0.0
Recovery boiler	MP1	5 091	0.26	0.46	90	145	15.3	6.78	6.1
Recovery boiler	LP	5 091	0.33	0.59	100	145	19.4	9.05	9.0
Biomass boiler	LP	1 340	0.02	0.01	100	100	0.4	0.15	0.2
Tall oil acidulation	LP	141	0.63	0.03	0	0	1.0	0.37	0.0
Tall oil refining (depitching)	MP3	98	0.80	0.03	100	100	0.9	0.37	0.4
Pulp Mill Common	MP1	2 857	0.10	0.10	0	100	3.3	1.18	0.0
Pulp Mill Common	LP	2 857	0.10	0.10	0	100	3.3	1.20	0.0
Bark Drying & Gasification, wet bark	LP	476	1.89	0.31	100	100	10.4	4.45	4.4
Total	MP3			0.03	100	100.00	0.9	0.4	0.4
	MP2			0.57	0	0.00	18.9	6.1	0.0
	MP1			3.53	49	112.48	116.7	45.2	22.0
	LP			7.65	87	98.96	253.0	106.2	92.3
TOTAL				11.78	73	101.56	389.4	157.9	114.7
FEEDWATER	LP							20.6	
LP Heater	LP				100	104		14.5	14.5
Condensing TG					100	44		73.0	73.0

STEAM & WATER PROPERTIES

	bar(abs)	Deg.C
Recovery Boiler	102.00	505
Biofuel Boiler	102.00	505
Steam to Turbine	99.00	505
MP3	99.00	320
MP2 steam	30.00	350
MP1 steam	11.00	195
LP steam	4.50	155
LP heater pressure	1.20	
Condenser pressure	0.090	
FW in tank		143.5
FW for RB boiler		145
FW for Biofuel Boiler		145
Demi water		25
Heated condensate & Demin. Water		102

MAKE-UP WATER DEMAND

	kg/s
Steam from boilers	264.4
Steam desuperheating	1.4
Feedwater from fw tank	265.8
Mill condensate return	129.2
Condenser condensate return	73.0
Feedwater preheat	20.6
Make-up water demand	43.1

PROJECT:	Thesis models	1 000 000 ADt/a	Mikko Niemeläinen
ALTERNATIVE:	Case: Combined depitching and bark gasification		11.4.2018

HEAT GENERATION

RECOVERY BOILER

Dry solids ratio	tDS/Adt	1.782
Dry solids (virgin)	tDS/d	5 091
Heat value, higher	GJ/tDS	14.0
Efficiency	%	75.8

BIOFUEL BOILER

		Biomass		
		Biomass 1 Bark	Biomass 2 Wood Rejects	Biomass 3 Pulp Rejects
Fuel flow	tDS/d	492.0	149.0	13.7
	t/d	984	310	46
	GJ/d as fired	8 118	2 383	126
Dry solids content	%ds	50.0	48.0	30.0
Heat value, effective	GJ/tds	19.0	18.7	15.0
	GJ/t	8.3	7.7	2.8
Efficiency	%	87.0	87.0	87.0

HEAT FROM BOILERS

	GJ/Adt	MJ/s	kg/s	t/h
Black Liquor (incl. methanol and CNCG)	18.9	624.9	225.8	812.8
Biofuel	3.2	107.0	38.7	139.2
Total	22.1	731.9	264.4	952.0

HEAT BALANCE

	GJ/Adt	MJ/s
Heat Generation		
- Black liquor	18.9	624.9
- Biofuel	3.2	107.0
- In feedwater pumps	0.1	3.6
Total	22.2	735.5
Heat Consumption		
- Pulp mill process	11.8	389.4
- BP power generation	3.6	117.9
- Cond. power generation	6.9	228.2
Total	22.2	735.5

POWER BALANCE

POWER CONSUMPTION		DESIGN			BALANCE		
		Design production t/d	Specif. cons. kWh/Ut	Specif. cons. kWh/ADt	Balance production t/d	Specif. cons. kWh/ADt	Power cons. kW
Wood handling	m ³ /sub d	18 328	4.0	23.1	14 662	24	2 841
Cooking	Adt/d	3 429	27.0	29.2	3 086	31	3 684
O2 delignification and bleaching	Adt/d	3 178	88.0	88.1	2 860	93	11 128
Drying machine	Adt/d	3 361	100.0	105.9	2 857	110	13 060
Evaporation Plant	t ₂₀ /h	1 267	74.0	29.5	1 140	32	3 788
Recausticizing plant & Lime kiln	m ³ WL/d	12 573	6.5	25.7	11 316	27	3 252
Recovery Boiler	tDS/d	5 657	46.0	82.0	5 091	87	10 354
Biomass Boiler	t/d	1 489	44.9	21.0	1 340	22	2 658
Raw water treatment	m ³ /d	79 365	0.1	1.3	71 429	1.3	158
Raw water intake	l/s	79 365	0.1	1.3	71 429	1	158
Demineralizing plant	m ³ /d	5 093	0.5	0.8	4 584	1	101
Effluent treatment	m ³ /d	63 492	1.0	20.0	57 143	21	2 526
Cooling water, tower & pumps	m ³ /d	478 299	0.3	37.7	430 469	40	4 758
Turbine plant	Adt/d	3 175	1.0	1.0	2 857	1	126
Compressed air	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Chiller for AC	Adt/d	3 175	8.0	8.0	2 857	8	1 011
Pulp mill common	Adt/d	3 175	12.0	12.0	2 857	13	1 516
Tall oil acidulation	t/d	157	27.3	1.3	141	1	170
Tall oil depitching	t/d	104	9.5	0.3	94	0	40
Oxygen plant	tO ₂ /d	116	560.0	20.4	104.1	22	2 577
ClO ₂ -plant	tClO ₂ /d	42	9170.0	122.3	36	129.8	15 454
Bark drying & Gasification	t/d, wet bark	529	36.6	6.1	476	6.5	778
PULP MILL TOTAL				645	682		81 100

POWER GENERATION

	kWh/ADt	MW
TG1	844	100.5
TG2	846	100.7
Total from turbogenerators	1 690	201.2

POWER BALANCE

	kWh/ADt	MW
Power consumption	682	81.1
Power generation	1 690	201.2
Power sales	1 008	120.1

Appendix 9: Investment cost breakdown of different scenarios



Thesis project
OY DIPPA AB

Depitching
Preliminary

Cost Estimate
Summary
- 1000 EUR -

Code	Area	0 Unclassified	1 Civil Works	2 Machinery	3 Piping	4 Electrical Equipment	5 Automation	6 HVAC	7 Spare Parts	0 to 9 Total
10	Depitching		16	1 280	93	6	75		2	1 472
60	Site Common for depitching									
90	Indirect costs	370								370
	- Temporary Facilities and Services									
	- Engineering									
	- Construction Management, Site Supervision									
	- Project Administration and Pre-operational Expenses									
	- Start-up and Training									
	Total	370	16	1 280	93	6	75		2	1 842
	Contingencies approx. 20 %									368
	TOTAL									<u><u>2 211</u></u>

