LAPPEENRANTA UNIVERSITY OF TECHNOLOGY

LUT School of Engineering Science

Master of Science (Technology) in Chemical and Process Engineering

Techno-economic feasibility study of hemicellulose extraction from softwood saw dust

Examiners: Prof. Eeva Jernström
Prof. Tuomas Koiranen

Md Hujjatul Islam

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ABSTRACT

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The main objective of this thesis is to study the feasibility of hemicellulose extraction from sawdust integrated into the wood pellet production process. Extraction process design and simulation was carried out using Aspen Plus simulation tool. The operating parameters for extraction were selected such a way that only high molecular weight hemicelluloses are obtained. In order to accomplish this, extraction time of 20 minutes and temperature 160 ˚C were considered. Solid to liquid ratio was considered 1:7 in order to avoid any problem in the circulation of the sawdust and water mixture. Pinch analysis was done in order to optimize the process for maximum energy recovery. It was found that around 66 MW (in large scale operation) of energy can be recovered and use for heating of reactor feed and drying of sawdust. The conversion and yield of hemicellulose from sawdust were obtained 9.3% and 8.89% respectively. The product was 80% hemicellulose and 20% water. The economic evaluation was done in order to study the feasibility of the extraction process. Small and large scale operation was considered for evaluation. The small scale production capacity was 1.5 ton / h and the large scale capacity was 9.3 ton/h. The return on investment for small scale and large scale operation was 32.8% and 44.9% respectively. The payout time for small scale is 2.67 years and for large scale is 2.02 years. Moreover, the Internal Rate of Return (IRR) was found 32% for small scale and 41% for large scale. It is considered safe to invest in a project that has IRR higher than 25%. Sensitivity analysis shows that variation of product price and production rate have high degrees of risk for investing in the project. Finally, it can be summarized that extraction of hemicellulose by autohydrolysis from sawdust before pelleting is technically and economically feasible except the fluctuation of product price and production rate.

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MD HUJJATUL ISLAM
Lappeenranta, 27th February 2017

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1. Introduction

The increasing interest in environmental issues, the limited amount of fossil raw materials and fuels, climate change, and the increasing world population have inspired the interest of promoting alternative manufacturing chains alongside the petrochemical industry. Combustion of fossil fuels is considered the dominant factor for global warming that releases 7.0 billion tons of carbon in each year into the environment (Zhang, 2008). Therefore, a substantial amount of funds and efforts have been spent in alternative energy resources such as biomass, the wind, geothermal and solar energy for securing the developmental and environmental needs. (Chen, Lawoko, and Heiningen, 2010). High dependency on imported oil, declining fossil reserves and environmental concerns promote the modern bio-based industry for the development of the biorefinery concept that entails the full utilization of every biomass processing streams in the most effective and profitable manner (Testova and Vilonen, 2009). Among lignocellulosic, woody biomass is most abundant in the world with 30% or 4 billion hectares of the world land mass covered by forest which is twice as large as the world's arable (Chen, Lawoko, and Heiningen, 2010). The biorefinery is an auspicious alternative that utilizes renewable biomass as raw material (Ragauskas, 2006). The paper mill uses fiber from wood for paper and board manufacturing. Due to the refining of the pulping process throughout the years, the cellulosic part of the wood is utilized very efficiently, while the rest of the part of the wood is burned to attain energy and heat (Sixta, 2006). For the pulp and paper mill to be a fully developed biorefinery, along with pulping process it is also necessary to develop processes for the non-cellulosic components of the wood. New high-value end-products or large volume bulk chemicals should be manufactured from the other wood components without compromising the main end-product in quality and volume. It is also preferred that the new processes are combined into the current processes that can minimize the investment costs and infrastructures. Apart from the main components of wood, the extractives can also be utilized for producing end products such as turpentine, tall oil, and biodiesel (Krogell, 2015).

Overcoming the difficulty to accurately fractionate biomass into the individual polymers: cellulose, hemicellulose and lignin is the main challenge for conversion of wood into biofuel and valuable chemicals. One of the favorable methods is autohydrolysis where hemicelluloses are dissolved using water or steam as the only solvent. Autohydrolysis is cheap and environmentally friendly process resulting in simpler downstream processes compared to acid and alkaline pre-hydrolysis.

2. Objective

Forest-based industry is going through major changes in worldwide. It is necessary to develop efficient techniques in order to transform wood into high value added products especially utilization of underused wood from various wood-based industry. The main underused parts of woods are harvesting residue, chip screening residue and sawdust. Sawdust is mainly used to produce pellets for energy production. Extraction of one of the main component of wood such as hemicellulose from sawdust before pelleting is of specific interest. In this work, a feasibility study will be carried out focusing on small and medium scale bio-refinery concept for extracting hemicellulose from sawdust. For feasibility study, preliminary process design will be carried out using steady-state simulation tool, such as Aspen Plus, for determining mass and energy balance of the whole process. After that economic analysis will be performed based on the preliminary process design for two different scales.

3: Chemical structure of wood

Trees are the seed-bearing plants which are sub divided into gymnosperms and angiosperms. Coniferous woods or soft woods belong to the first-mentioned category and hardwoods to the second group (Sjöström, 1993). Hardwoods are trees of the deciduous class with broad leaves while the softwoods trees (coniferous) have needle-like or scale-like leaves. Wood consists mainly of cellulose, hemicellulose, lignin and extractives. The chemical composition of wood varies from species to species. The average composition of hardwood and softwood can be seen in Table 1. Typically, polysaccharides represent the leading part of the wood mass.

Table 1: Chemical composition of wood (Sjöström, 1993)

	Hardwood (% of wood)	Softwood (% of wood)
Cellulose	44±6	38±5
Hemicelluloses	24±6	25±4
Glucomannans	2.5±1	17±3
Xylans	23±7	8±3
Other polysaccharides	3±1	6±3
Lignin	24±3	34±7
Extractives	3±2	5±3

Cellulose is the main fraction of wood. Almost 40-50% of the dry substance in maximum wood species is cellulose. Lignin content is around 24-34 % depending on wood species. The non-structural components such as extractives, consist of a vast variety of mostly low-molar mass compounds that can be classified as lipophilic (soluble in nonpolar solvents) and hydrophilic (soluble in polar solvents). Common lipophilic extractives are fatty and resin acids, triglycerides, and stearyl esters and common hydrophilic extractives are sugars, stilbenes, stilbene glucosides, and flavonoids. Their fraction in the wood is only a few percentages, although the amount differs between species. (Sjöström, 1993)

3.1: Hemicellulose

Hemicelluloses are a group of branched polysaccharide polymers built of xylan, glucomannan, glactoglucomannan, arabinogalactan and galactan (Gullichsen, 1999). They were formerly believed to be the intermediates in the biosynthesis of cellulose. But now a days, it is known that the hemicelluloses are the group of heterogeneous polysaccharides that formed through biosynthetic pathways alternative to that of cellulose. Hemicelluloses are comparatively easily hydrolyzed by acids to their main monomeric components consisting of D-glucose, D-mannose, D-galactose, D-xylose, and L-arabinose. Most of the hemicelluloses have a degree of polymerization of around 200. The amount of hemicellulose of the dry weight of wood is usually between 20 to 30%. The chemical composition and structure of hemicellulose in the softwood varies from those in hardwoods (Sjöström, 1993). Table 2 summarizes the main features of hemicelluloses of both hardwood and softwood

Table 2: The major hemicellulose components (Sjöström, 1993)

Hemicellulose type	Occurrence	Amount (% of wood)	Solubility	$\overline{\mathrm{DPn}}$
Galactoglucomannan	Softwood	5-8	Alkali, water	100
(Galacto)glucomannan	Softwood	10-15	Alkaline borate	100
Arabinoglucuronoxylan	Softwood	7-10	Alkali, water	100
Arabinogalactan	Larch wood	5-35	Water	200
Glucuronoxylan	Hardwood	15-30	Alkali, dimethylsulfoxide	200
Glucomannan	Hardwood	2-5	Alkaline borate	200

The most abundant Nordic hardwood species such as birch are mainly composed of xylose and they are often called xylans. The common Nordic softwoods such as spruce and pine have hemicelluloses primarily consist of mannose and glucose and they are called mannans or glucomannans. Hemicelluloses are branched and often contain functional groups such as; carboxyl or acetyl groups other than hydroxyl. Hemicelluloses are mainly amorphous due to the lower molar mass and branching which makes hydrogen bonding between molecules more difficult. Therefore, hemicelluloses are generally easier for water dissolution and more easily hydrolyzed by acid than cellulose (Krogell, 2015).

3.2: Softwood hemicellulose

Galactoglucomannans: Galactoglucomannans are the principal hemicellulose in softwoods which is about 20%. Their main structure is linear or hardly branched chain consist of (1-4)-linked β-D-glucopyranose and β-D-mannopyranose units. (Figure 1)

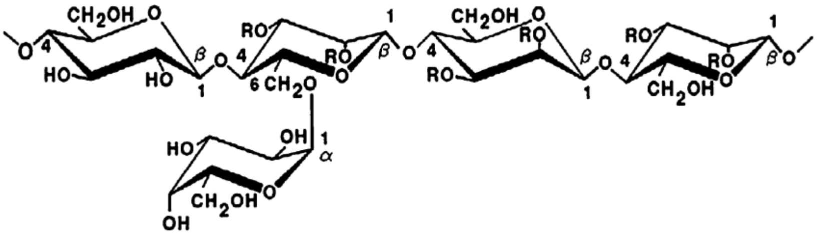


Figure 1. Principal structure of galactoglucomannan (Sjöström, 1993)

A significant structural aspect is that the hydroxyl groups at C-2 and C-3 positions in the chain units are partly replaced by o-acetyl functional groups on the average of one group per 3-4 hexose units. They are easily hydrolyzed by acids and especially so the bond between galactose and the main chain. The acetyl groups are readily broken by alkali than by acid.

Arabinoglucuronoxylan: In addition to Galactoglucomannans, softwood contains an Arabinoglucuronoxylan in the range of 5 -10%. It is composed of (1-4)-linked β-D-xylopyranose units which are moderately substituted at C-2 by 4-methyl-α-D-glucaronic acid groups, on an average of two residues per ten xylose units. (Figure 2)

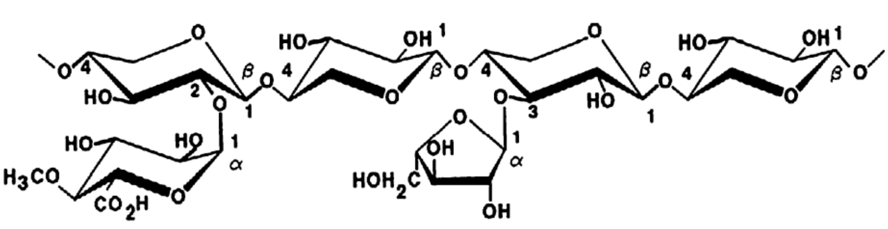


Figure 2. Principal chemical structure of arabinoglucuronoxylans (Sjöström, 1993).

In addition, the framework contains α-L-arabinofuranose units, on the average 1.3 residues per ten xylose units. Due to of their furanosidic structure, the arabinose side chains are easily depolymerized by acids.

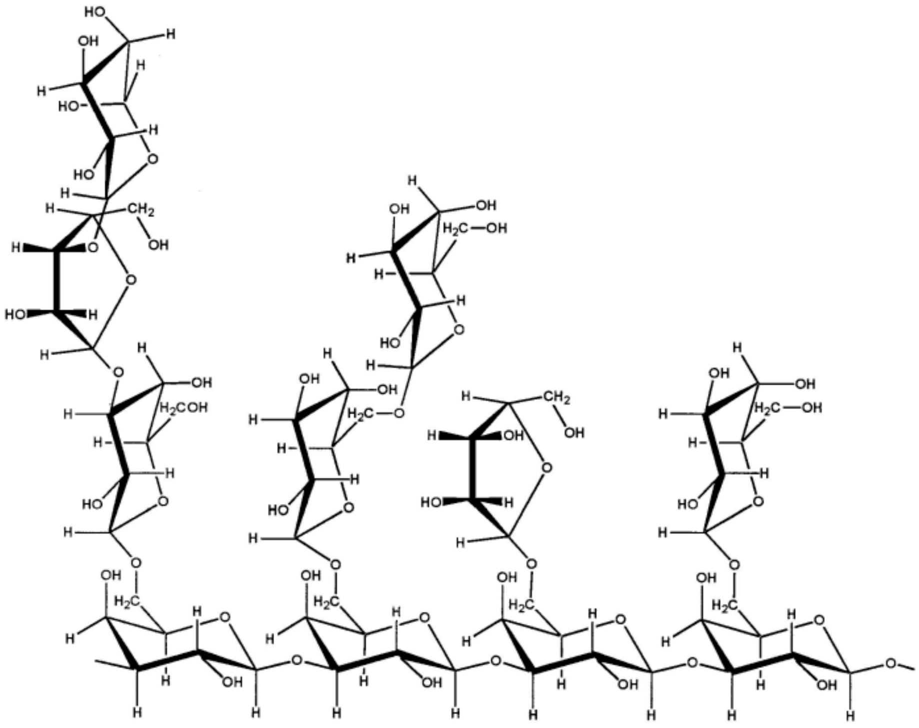


Figure 3: Structural fragment of arabinogalactan macromolecule (Karunaratne, 2012)

Arabinogalactan: The heartwood bears particularly huge amounts of water-soluble arabinogalactan, which is the only inconsequential component in other wood species. Its backbone is built up of (1-3)-linked β-d-galactopyranose units. Nearly every unit carries a branch attached to position 6, largely (1-6)-linked β-D-galactopyranose residues but also L-arabinose (Figure 3).

Other Polysaccharides: Apart of Galactoglucomannans, arabinoglucuronoxylan and arabinogalactan, softwoods contain other polysaccharides, usually present minor quantities. They are mainly starch and pectic substances. The most common units in pectic polysaccharides consist of D-galactosyluronic acid and D-galactose, L-arabinose, and L-rhamnose residue (Sjöström, 1993).

3.3: Hardwood Hemicellulose

Glucuronoxylan: Even if hemicelluloses in various hardwood species varies from each other both quantitatively and qualitatively, the major component is an O-acetyl-4-O-methyleglucurono-β-D-xylan, sometimes called glucuronoxylan. Often the xylose based hemicellulose in both softwoods and hardwoods are named simply xylans (Sjöström, 1993). Depending on the hardwood species, the xylan content differs within the limits of 15-30% of the dry wood. The backbone consists of β-D-xylopyranose units, linked by (1-4)-bonds. Most of the xylose residues contain an O-acetyl group at C-2 or C-3 (about seven acetyl residues per ten xylose units). The xylose units in the xylan chain additionally carry (1-2)-linked 4-O-methyle-α-D-glucuronic acid residues. The xylosidic bonds between the xylose units are easily hydrolyzed by acids whereas the linkages between the uronic acid groups and xylose are very stable (Sjöström, 1993). (Figure 4)

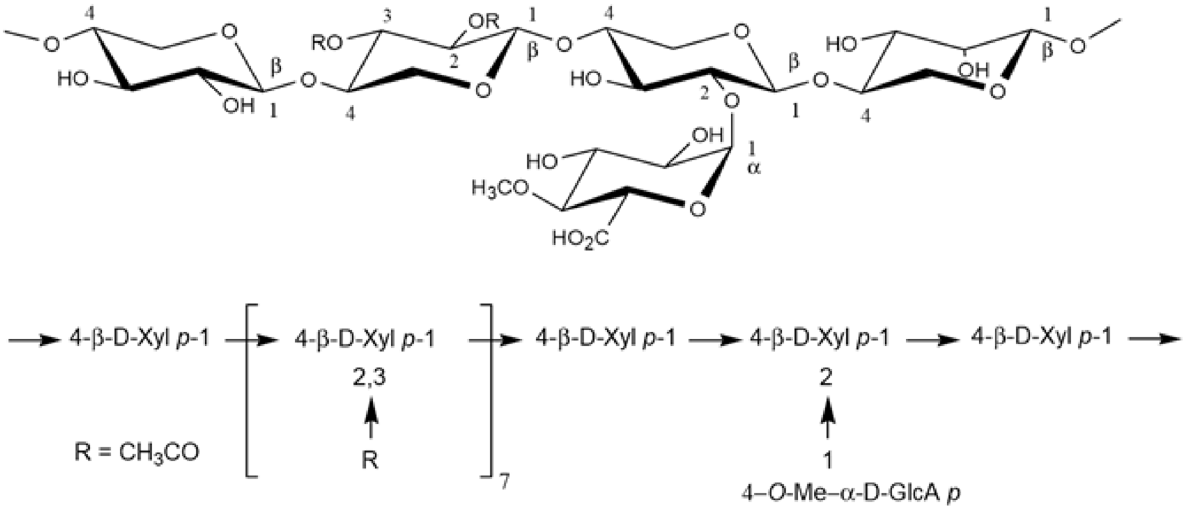
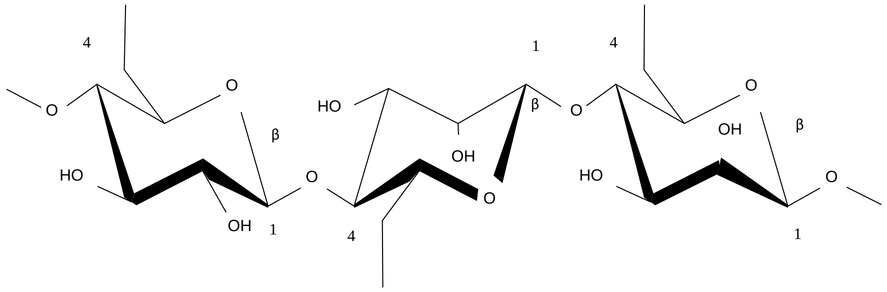


Figure 4. Structural formula of glucuronoxylan (Laine, 2005)

Glucomannan: Apart of xylan, hardwoods also bear 2-5% of glucomannan, which is composed of β-D-glucopyranose and β-D-mannopyranose units connected by (1-4)-bonds. The glucose: mannose ratio varies between 1:2 and 1:1, depending on the type of wood. The mannonsidic bonds between the mannose units are more briskly hydrolyzed by acid than the corresponding glucosidic bonds and glucomannan is easily hydrolyzed in acidic conditions (Sjöström, 1993). (Figure 5)



→ 4-β-D-Glc *p*-1 →4-β-D-Man *p*-1→ 4-β-D-Man *p*-1→

Figure 5: Structural formula of glucomannan (Laine, 2005)

4. Hemicellulose extraction processes

4.1: Acid-catalyzed hydrolysis

Acid-catalyzed hydrolysis has been interpreted as early as the 19th century with commercial applications from the beginning of 20th century. Acid hydrolysis can be divided into two general approaches such as; concentrated acid/low temperature, and dilute acid/high temperature. Biomass pretreatment with concentrated acids is commonly performed using a high acid concentration (over 30%), at ambient to moderate temperatures (<100 ˚C) and atmospheric pressures (Mussatto, 2016). As a catalyst, sulphuric acid (H_2SO_4) is most commonly used although other mineral acids have also been investigated. Concentrated acids eliminate mainly cellulose and hemicellulose producing lignin-rich solid phase. They allow operating at low/medium temperature and pressure that produces small amount degradation products. However, degradation products formation rate is extremely affected by minor variation in temperature. Recovery of acid is a key step for the economic feasibility of concentrated acid pretreatments and the neutralization costs have inhibited the general use of this pretreatment. In addition, equipment corrosion problem is the major disadvantage of this process. (Carvalheiro, Duarte, and Gírio, 2008).

Dilute sulfuric acid processes (0.5-1.5%, temp. 121-160°C) have been most preferred for industrial application because it provides high sugar yields from hemicellulose. H_2SO_4 is usual acid employed although HCl, HNO_3, and H_3PO_4 are also used. Compared to concentrated acid hydrolysis, this pretreatment provides fewer degradation products as well much fewer corrosion problems in reactors and pipes. Carbonic acid acts better than most of the inorganic acids and is especially suitable for hydrolysis under mild pH at high temperatures and pressures. Although, carbonic acid is responsible for an increase in xylan hydrolysis depending on wood (Carvalheiro, Duarte, Gírio, 2008). Around 80- 90% of hemicellulose sugars are usually recovered by dilute acid pretreatment. Besides sugars, the hydrolysates obtained by dilute acid pretreatment may also contain other components such as acetic acid (generated from the acetyl groups linked to the hemicellulose structure); furfural and hydroxymethylfurfural (generated from the degradation of pentose and hexose sugars, respectively) (Mussatto, 2016),(Figure 6). A short summary of acid catalyzed hydrolysis, advantages and disadvantages are illustrated in Table 3.

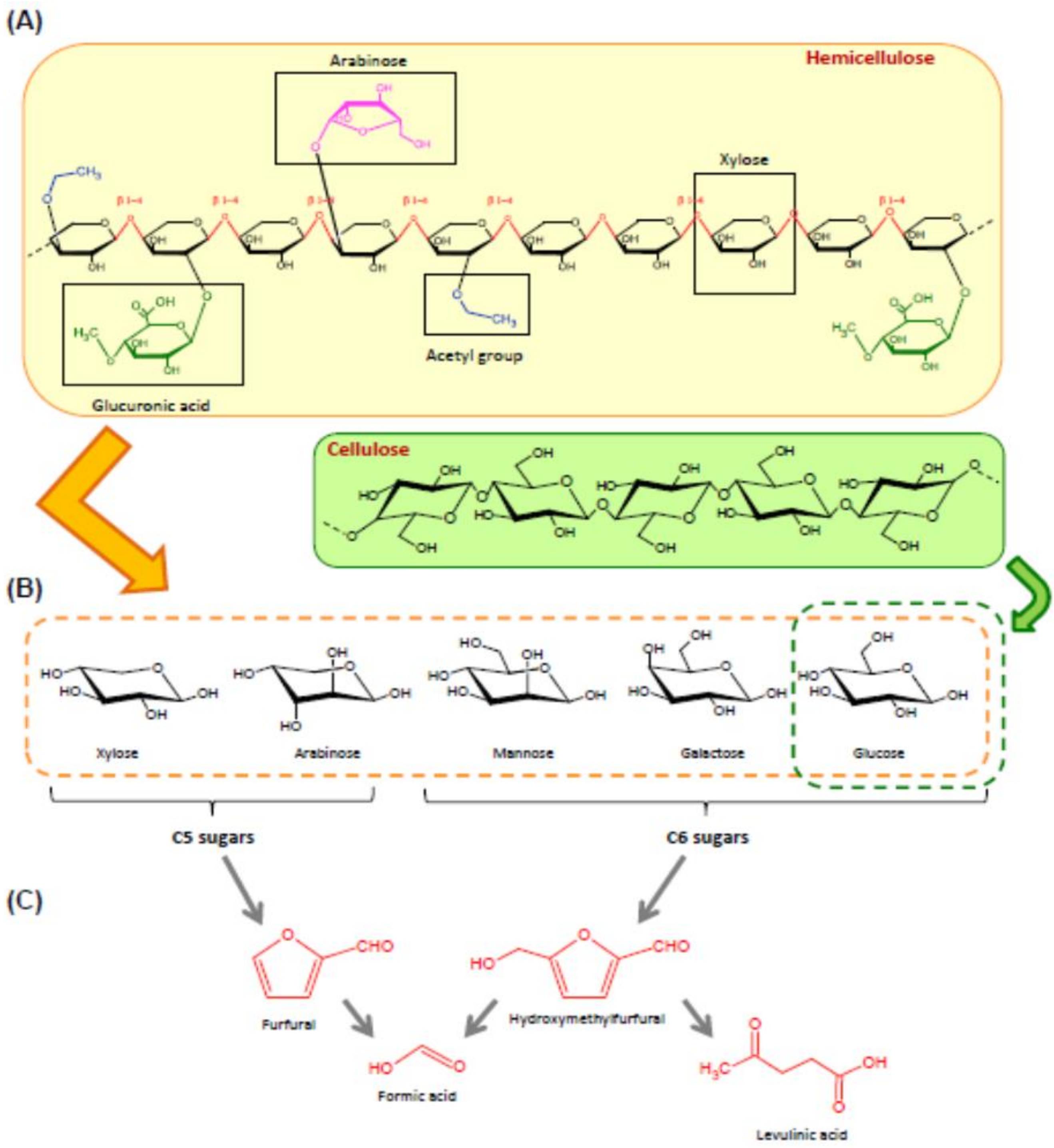


Figure 6: (A) Schematic representation of hemicellulose (example with a xylan backbone) and cellulose structures, (B) sugars that can be released from each structure during pretreatment, and (C), toxic compounds that can be formed by sugar degradation. (Mussatto, 2016).

Table 3: Main effects on lignocellulosic biomass structure; advantages, disadvantages related to acid catalyzed hydrolysis. (Mussatto, 2016, Chapter- 8).

	Dilute Acid Pretreatment	Concentrated acid pretreatment
Main effects on biomass structure	Main effect on hemicellulose; few effects on cellulose and lignin	Main effect on both hemicellulose and cellulose; few effects on lignin
Advantages	• High efficiency to recover sugars from hemicellulose • Increases the material porosity and enhances cellulose accessibility to enzymes • Low consumption of acid	•High efficiency to recover sugars from hemicellulose and cellulose • A subsequent step of enzymatic hydrolysis is usually not required • Requires low temperature and pressure
Disadvantages	• Requires reactors resistant to acids • Requires high temperature and pressure • Necessity of using small particle sizes • Neutralization of the hydrolysate is required before use as fermentation medium. • Formation of toxic compounds	• Requires reactors resistant to acids • High consumption of acid (catalyst recovery needed) • Necessity of using small particle sizes • Neutralization of the hydrolysate is required if the catalyst is not recovered • Formation of toxic compounds(furfural and hydroxymethylfurfural)

4.2: Alkaline Treatments

Alkali pretreatment boosts cellulose digestibility enhancing lignin dissolution. They are more effective for lignin solubilization showing only minor cellulose and hemicellulose solubilization. Sodium and calcium hydroxide (lime) are the most common alkali (conc. 0.05-0.15 g alkali/g biomass). In alkaline hydrolysis, alkali induces swelling prompting to increase in internal surface area. After, a decrease in the degree of polymerization and crystallinity happens with a subsequent separation of structural linkages between lignin and polysaccharides. Then, separation of lignin structure occurs followed by saponification of intermolecular ester bonds that crosslink hemicelluloses and other components. This leads to an increase of porosity with the removal of such crosslinks.

Use of low temperature and pressure is the main advantage of this process. Comparing to NaOH and KOH, lime has reduced cost and less significant safety requirements. It can be recovered from hydrolyzate by reaction with CO_2 which gives carbonate. Carbonate then can be transformed into lime.

4.3: Hydrothermal extraction processes

The liquid hot water (LHW) or auto-hydrolysis is the most widely used treatment process for extraction of hemicellulose from wood. This process has been also used in pulp and paper industry as early as the 1930s as a method for extraction of hemicellulose from wood prior to pulping. The LHW (autohydrolysis) process is catalyzed by hydronium ions and has an identical mechanism to dilute-acid hydrolysis. The catalysts in hydrolysis are hydronium ions generated in situ by water autoionization (autohydrolysis) and acetic acid resulting from acetyl substituents of hemicelluloses. The second phenomena have a much higher contribution to the hydrolysis (Heitz et al., 1986).

The temperature used for hot water extraction varies from 150 ˚C to 240 ˚C which highly dependent on the type of biomass and desired extraction products. Many of the water properties changes in high temperatures such as; the dielectric constant, viscosity and the surface tension decreases (Yang et al 1998). These changes greatly affect the dissolution properties of water. This might be one reason why autohydrolysis is such a prominent method for hemicellulose extraction. The mechanisms for dissolution of hemicelluloses from wood are very difficult to define because of the complex wood matrix and different hemicelluloses in different tree species. Until today they are not fully understood. In order to simplify the extraction process, it can be divided into three different steps (Figure 7).

Figure 7: Simplified layout of the extraction process of hemicelluloses from wood (Krogell, 2015)

The first step is the water that enters into the wood in order to reach to the main location of the hemicelluloses. The next step is the separation or dissolution of the hemicelluloses from the wood matrix into the liquid phase. This is one of the most essential steps. In the third step, dissolved hemicelluloses diffuse out from the wood

4.3.1: Operational condition in autohydrolysis

Hydrothermal treatments mainly comprise of liquid hot water (autohydrolysis) and steam-explosion in the range of 150°C to 230°C. For temperatures below 100°C, there is no hydrolytic effect on the wood whereas above 220°C degradation of some cellulose degradation occurs. The reaction time of hydrothermal treatments diverges from seconds to hours relating to operation temperature. Solids concentration termed as the liquid-to-solid ratio (LSR), may differ between 2 and 100 (w/w). The most common LSR values are around 10. Higher LSR is usually employed with continuous reactors whereas lower LSR is used in the steam explosion.

Influence of particle size is less important for high LSR but for low LSR, an increase of particle size leads to a decrease in hemicellulose solubilization due to an insufficient biomass impregnation. However, the effect of particle size sounds more significant for hydrothermal pretreatments using steam than for those using liquid hot water. For large biomass particles, the joint effect of fast heating and short reaction times may lead to an irregular biomass particle heating. The effect of particle size also depends on the type of biomass. The relationship between temperature and time also strongly influences the composition of both the liquid and solid phases. Although both factors can be grouped in a single parameter, ie, empiric parameters such as the severity factor it is generally accepted that higher temperatures/short reaction times would lead to slightly better results in terms of pentoses yield together with a minimization of inhibitor formation. In order to maximize hemicellulose hydrolysis and minimize degradation reactions, it is important to monitor and control of pH during hydrothermal processing. In fact, even using water only, the pH of hydrolysates can drop significantly. Previously, it was described as a decrease in pH of the hydrolyzates to reach 3.90. This decrease is a function of total weak acids concentration, including degradation products such as formic acid. (Carvalheiro et al., 2016)

4.3.2: Kinetic study of autohydrolysis

In order to better understand and make use of autohydrolysis data, mathematical modeling is an indispensable tool. This is especially relevant to develop technical studies, namely regarding scale-up and economic evaluation. A detailed analysis of the primary steps associated with the dissolution of hemicelluloses is not feasible in practice due to the enormous intricacy of the various chemical reactions. The simultaneous degradations of the extracted compounds convolute the interpretation of the data. The total amount of hemicelluloses in the system is considered to be constant and the loss due to the degradation is compensated to both phases equally. Therefore, the results follow the dissolution of the hemicelluloses primarily present in the wood. The kinetic expression for the dissolution is based on the hypothesis that internal mass transfer resistance is insignificant due to the small particle size and that external mass transfer resistance is overcome by the high turbulence. All the concentrations are calculated as milligrams per liter as the molar masses of the compounds are somewhat unidentified. The reaction rate of an n order chemical reaction in is described as follows;

$$-\frac{d[C]}{dt} = k'[C]^n \qquad (1)$$

Here C = Concentration of the reactant, t = time, k'= rate constant and n = the reaction order. By integrating and rearranging the equation (1) we get;

$$-\ln(\frac{C}{C0}) = k't \qquad (2)$$

It has been proposed that the extraction of hemicelluloses from wood can be considered as a pseudo-first order reaction and thus follow eq. (1). Normally, C_0 and C denotes the concentration of the reactant at the start of the reaction and the reactant concentration at a given time respectively. For hemicellulose extraction, C_0 would mean the original hemicellulose concentration in the wood and C the hemicellulose concentration in the wood during extraction. In this study, C_0 represents the hemicelluloses in the untreated wood whereas C was obtained by subtracting the extracted hemicelluloses from the hemicelluloses in the pristine wood.

Here k´ is given by the modified Arrhenius equation 3.

$$k' = k_0 \exp\left[\frac{-Ea}{R}\left(\frac{1}{T} - \frac{1}{Tref}\right)\right] \qquad (3)$$

Ea stands for the activation energy for hemicellulose dissolution which was determined to be 135Kj/mol, k0 is the rate constant at the reference temperature, R is the universal gas constant, T is the temperature, and Tref is the reference temperature of the experiments in kelvin. The pre-exponential factor (k0) obtained the value 0.0363 (1/min) at the reference temperature (165 °C). (Grénman et al., 2011; Krogell, 2015). Figure 8 represents the changes of hemicellulose concentration from the solid phase to the liquid phase in different temperature based on the above kinetic model. A summary of hemicellulose extraction by autohydrolysis is reported in Table 4.

Table 4: Summary of operation parameters for hemicellulose extraction through autohydrolysis

Raw material type	Particle size, mm	Temperature (◦C)	Pressure (bar)	Time (min)	Wood to water ratio	Conversion %	Types of final product	References
Pinus radiata	0.5-2.8	170	100	40	1:13	59.9	Carbohydrate	Chunlin xu et al(2016)
Pinus radiata	0.5-2.8	170	8	40	1:7	63.9	Carbohydrate	Chunlin xu et al(2016)
Picea Abies	0.5-12.5	170	8	40	1:33	-	Galactogluc-omannan	J. Krogell (2015)
Pinus pinaster	<1	160-240	-	-	1 : 8	95	Monosacch-arides	(González-Muñoz et al., 2011)
Olive tree pruning	<10	170-230	-	10	1 : 5	-	Oligosaccharides	(Cara et al., 2012)
Acacia dealbata	<8	170-240	-	-	1:8	70	Xylooligosaccharides	(Yáñez et al., 2009)

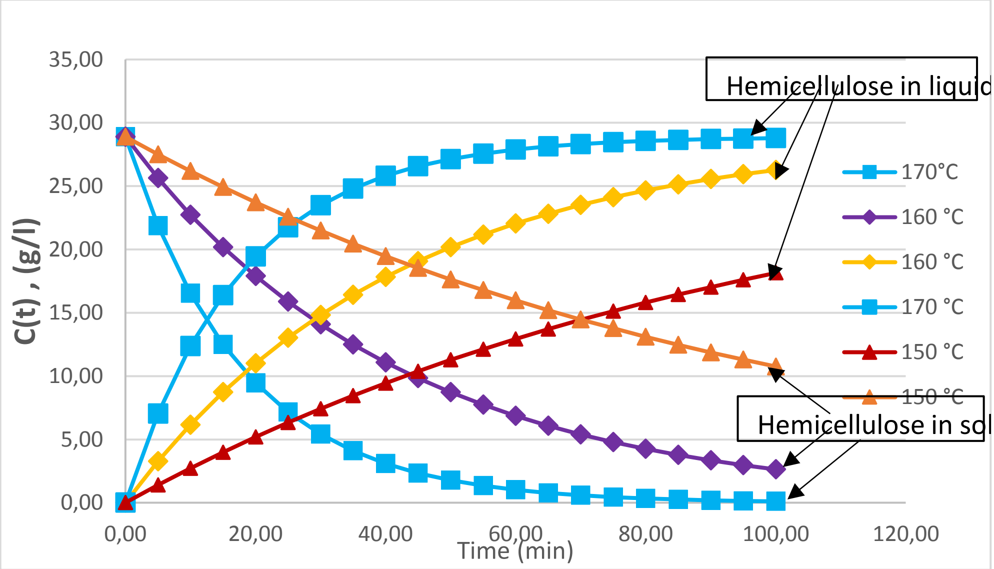


Figure 8: Changes of hemicellulose concentration from the solid phase to the liquid phase in different temperature and time.

4.4. Reactor types and operation mode

Various types of reactors and operation mode are being employed for hemicellulose extraction by autohydrolysis. The most widely used reactor is the autoclave with continuous mixing or re-circulation of the liquor. The hemicellulose yield is affected by reactor operation mode and reactor configuration. The most common is the batch operation, where wood particles and water are heated and held at the desired temperature for a certain residence time before cooling. For non-batch operation, depending on the circulation direction of hot water relative to lignocellulosic biomass, three different process configurations can be employed. These include continuous (concurrent or countercurrent) and semi-continuous (flow-through) reactors. In countercurrent reactors biomass and water flow in opposite directions through the reactor while in flow-through reactors, hot water flows through a stationary bed of biomass, hydrolyzing and dissolving the components and carrying them out of the reactor. Batch reactors usually require less water and energy (both to operate and for downstream processing) whereas continuous reactor systems typically operate at higher temperatures and pressures to achieve a high conversion of biomass within a short extraction time (Mussatto, 2016). A summary of reactor types and operation modes are summarized in Table 5.

Table 5: Summary of reactor types and operation mood

Number	Reactor types	Operating scale / Volume of reactor	Operating mode	References
1	Tubular reactor	Lab/ 0.015 l	Batch	(Ojumu et al., 2003)
2	Rotating bomb digester	Lab / 1.5	Batch	LEE et al., 2009)
3	Tubular reactor	Lab/0.2 l	Batch	Yang and Wyman, 2009)
4	Autoclave	Lab/0.28 l	Batch (Continuous mixed)	Chen, Lawoko, and Heiningen, 2010
5	Autoclave	lab/0.75 l	Batch	Fiserova and Opalena, 2012
6	Autoclave	Pilot/ 50 l	Batch (Recirculation of liquor)	Chunlin xu et al(2016)
7	Jacket-heated Parr reactor	-	Batch	(Lei, Cybulska, and Julson, 2013)
8	Tubular reactor	0.015 L	Batch	(Ojumu et al., 2003)

5. Technoeconomic evaluation methodology

Technoeconomic assessment is very important in the analysis of biorefinery schemes since it is a key tool for providing decision-making information for the new investment steps (Mussatto, 2016). The economic evaluation of an investment project includes the estimation of capital expenditures, annual operating expenditures, revenues, profits and further indicators of economic sustainability. In the following, the methodologies for estimating capital investment and operating costs are briefly discussed along with an evaluation of the economic viability of an investment project. A flow diagram of a typical techno-economic analysis methodology is illustrated in Figure 9.

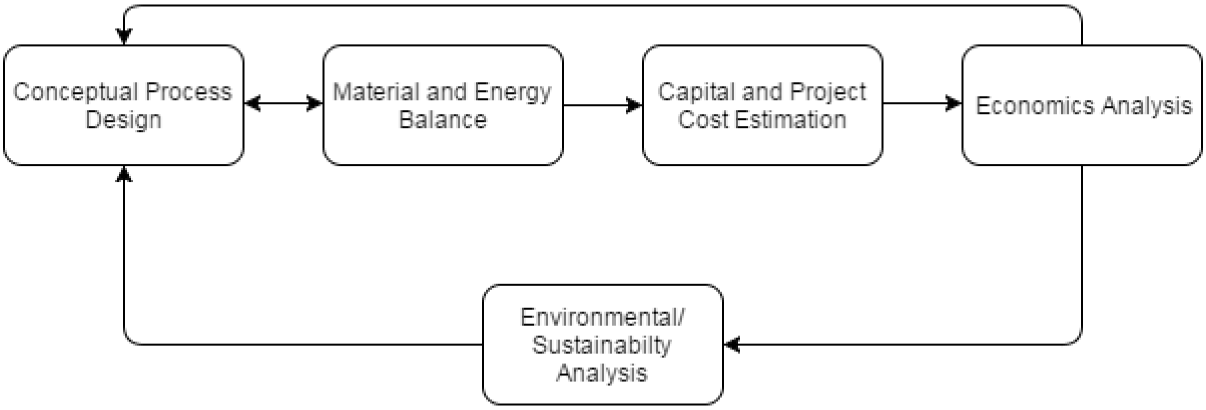


Figure 9: Flow diagram of techno-economic analysis methodologies. (NABC, 2011)

5.1: Capital Cost Estimation

The total investment needed for a project can be roughly divided into the sum of the fixed capital investment and working capital investment. The fixed capital investment is the total cost of the plant ready for start-up. It includes the cost of

1. The inside battery limits (ISBL) investment costs

2. Off-site or OSBL investment

3. Investment for engineering and construction of the plant

4. Contingency costs

ISBL plant costs: The ISBL plant costs comprises the cost of purchasing and installing all the process equipment that consist of the new plant. The direct costs include:

1. All the major process equipment, such as vessels, reactors, columns, furnaces, heat exchangers, coolers, pumps, compressors, motors, fans, turbines, filters, centrifuges, dryers, etc., including field fabrication and testing if necessary.
2. Bulk items, such as piping, valves, wiring, instruments, structures, insulation, paint, lube oils, solvents, catalysts, etc.
3. Civil works such as roads, foundations, piling, buildings, sewers, ditches etc.
4. Installation labor and supervision

Purchased equipment costs can be estimated from vendor data as well as by cost curves. When there is a lack of reliable cost data or estimating software, the correlations given in Appendix 1 (Towler and Sinnott, 2007, Chapter-7) can be used for preliminary estimation. The correlations are according to the equation 4.

$$C_e = a + bS^n \qquad (4)$$

Where Ce = purchased equipment cost

a, b = cost constants

S = size parameter,

n = exponent for that type of equipment

The installation costs are directly related to purchasing equipment cost. Appendix 2 presents the range of installation cost as a percentage of purchased-equipment cost for various types of equipment. George et al. (1988) suggested the piping installed cost for solid processing plant around 16%, solid-fluid processing plant 31% and fluid processing plant 66% of delivered equipment cost.

OSBL investment: The offsite cost or OSBL investment comprises the cost of the additions that must be made to the site infrastructure for adding a new plant or increasing the capacity of existing plant. Offsite costs are typically approximated as a percentage of ISBL costs in the preliminary design stage. It is in the range of 10% to 100% of ISBL costs based on the project scope and infrastructure. For typical chemical projects, offsite costs are usually in the range of 20% to 50% of ISBL cost. An estimation of 40% is usually used as a preliminary estimate if no details of the site are known. (Towler and Sinnott, 2007)

Engineering cost: The engineering cost comprises of the costs of detailed design and other engineering services needed to implement the project. They are mainly detailed design engineering of process equipment, piping and control systems, plant layout, drafting, cost engineering, civil engineering, construction supervision and services. Engineering costs are best estimated based on project scope as they are not directly proportional to project size. A rule of thumb for engineering costs is 30% of ISBL plus OSBL cost for smaller projects and 10% of ISBL plus OSBL cost for larger projects. (Towler and Sinnott, 2007)

Contingency cost: All cost estimates are uncertain and the final installed cost of many items is not known until the installation has been successfully accomplished**.** Therefore**,** Contingency costs are extra costs included into the project budget in order to allow for alteration from the cost estimate. A minimum contingency charge of 10% of ISBL plus OSBL cost should be used on all projects. A higher contingency charges (up to 50%) is needed if the technology is unreliable. (Towler and Sinnott, 2007)

Working capital Investment (WCI): The WCI includes the initial cost of resources as well as money needed for labor and services required to start operating the plant. This is the additional investment that is necessary to start up the plant above the fixed capital investment. Typical WCI includes the cost of:

1. Start-up,

2. Initial catalyst charges,

3. Raw materials and intermediates in the process,

4. Finished product inventories,

5. Funds to cover outstanding accounts from customers.

Common values for the WCI are in the range of 15–20% of the FCI, i.e. about 13–17% of the total investment (Peters and Timmerhaus (1990). Nonetheless, this estimate has been evaluated for conventional chemical plants and for biorefineries the values may differ. For instance, Humbird et al. (2011), proposed an estimate of 5% of the FCI for ethanol production from lignocellulosic biomass.

5.2: Estimation of Manufacturing cost (OPEX)

The costs related to the day-to-day operation of a chemical plant need to be approximated before the economic feasibility of a proposed process can be determined. There are many factors that influence the manufacturing cost of a chemical plant. Total manufacturing costs are the sum of variable manufacturing costs and fixed manufacturing costs.

5.2.1: Variable manufacturing cost

Variable manufacturing costs are proportional to the plant output or production rate. These comprise the cost of

1. Raw material consumed by the process
2. Utilities- fuel burned in process heaters, steam, cooling water, electricity, raw water, instrument air, nitrogen,
3. Consumables- solvents, acids, bases, inert materials, additives, catalysts, adsorbents
4. Effluent disposal
5. Packaging and shipping- drums, bags, tankers etc.

Variable costs are mainly determined by the choice of feedstock, process chemistry, and plant location. By doing a more efficient design and operation of the plant, variable costs can be reduced.

5.2.2: Fixed manufacturing cost

Fixed production costs are costs that are needed regardless of the plant operation rate or output. These costs are not reduced even if the plant reduces its production. Fixed costs include:

1. Operating labor

2. Supervision—typically 25% of operating labor.

3. Direct salary overhead — usually 40% - 60% of operating labor plus supervision. These are non-salary costs such as employee health insurance and other benefits.

4. Maintenance, which includes both materials and labor, and is typically estimated as 3% to 5% of ISBL investment, depending on the expected plant reliability. Plants with more moving equipment or more solids handling usually require higher maintenance.

5. Property taxes and insurance—typically 1% to 2% of ISBL fixed capital.

6. Rent of land (and/or buildings)—typically estimated as1% to 2% of ISBL plus OSBL investment. Most projects consider land is rented rather than purchased, but in some cases, the

land is bought and the cost is added to the fixed capital investment and recovered at the end of the plant life.

7. General plant overhead — charges to cover corporate overhead functions such as human resources, research and development (R&D), information technology, finance, etc.

8. Allocated environmental charges to cover superfund payments —typically 1% of ISBL plus OSBL cost.

9. Running license fees and royalty payments

10. Capital charges—these include interest payments due on any debt or loans used to finance the project, but do not include expected returns on invested equity capital.

11. Sales and marketing costs—in some cases, these are considered part of general plant overhead. They can vary from almost zero for some commodities to millions of dollars a year for branded items such as foods, toiletries, drugs, and cosmetics.

Fixed manufacturing costs are not affected by better design or operation of the plant, other than improvements that allow the plant to be operated safely with a smaller workforce. They are more amenable to control at the corporate level than the plant level. (Towler and Sinnott, 2007)

5.3 Revenues and profitability

The revenues for a project are the income earned from sales of main product and by-products. The production rate of the main product is usually specified in the design basis and is determined by the marketing department based on predictions of overall market growth. The total cost of production is the sum of the fixed and variable production costs. The gross profit is

$$\text{Gross profit} = \text{Main product revenues} - \text{Total cost of production} \qquad (5)$$

5.4 Evaluation of economic sustainability

Often, different process scenarios need to be compared in economic terms. For this purpose, suitable performance figures are needed. A very popular method for evaluating investment projects is the Internal Rate of Return (IRR), which is defined as the discount rate at which the net present value (NPV) of an investment is equal to zero (DeFusco et al., 2007). The higher the IRR, the more encouraging the investment project. Because it suggests that future cash flows could be discounted at a higher discount rate until the NPV equals zero. An IRR of 25% is usually accounted “as the threshold for securing capital investment in new processing technology” in the chemical industry (Brownell, Yu, and Saddler, 1986). This threshold can, therefore, be used as a benchmark which the process under study would have to achieve in order to become attractive for investors.

Although the IRR is very popular, the concept is often not well understood. In particular, it needs to be mentioned that the benchmark of an IRR of 25% is not identical to an annual interest rate of 25%. A simple comparison can be made in order to make the difference clear. Instead of investing the main capital into a biorefinery, it could be saved in a bank and earn annual interest. The interest rate that would provide the same future value at the end of the project lifetime such as 15 years, could be determined. In the case of an IRR of 25%, the equivalent interest rate is around 7%.

6 Conceptual process design and simulation

6.1 Process concept

Autohydrolysis is considered for extraction of hemicellulose from sawdust. It is a green process as no chemicals are used in the process. There are three main phases of the process: extraction of hemicellulose, separation of sawdust and drying, and separation of hemicellulose from extracted liquor. High molecular weight hemicellulose is highly desirable for various applications, such as barriers for packaging, emulsifier, hydrogels and functional fibers (Kisonen et al., 2014). Therefore, the process design is aimed towards obtaining high molecular weight hemicellulose. A block diagram of the whole process is illustrated in Figure. 10

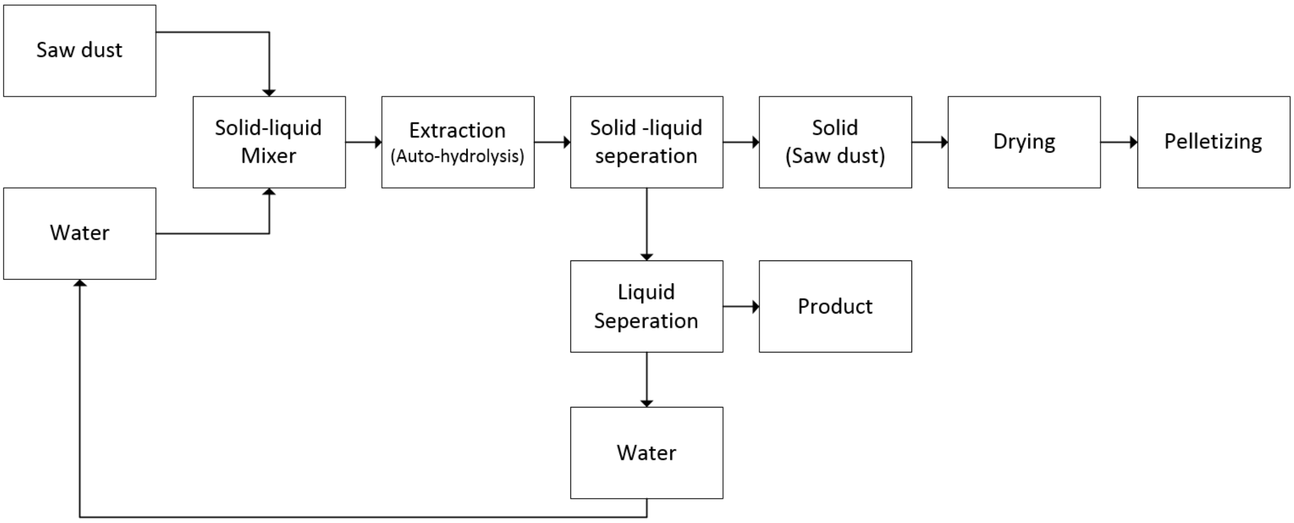


Figure 10: Block diagram for hemicellulose extraction from sawdust

6.2 Selection of operating condition

The main target for selecting appropriate operating condition is to extract high molecular weight hemicellulose. The average molar mass of extracted hemicellulose is suggested to be 20-50 kDa which corresponds to the degree of polymerization of 100-400. The operating conditions such as temperature, extraction time, the particle size of sawdust and pH in the extraction process have certain influences on the molar mass of the extracted hemicellulose (Krogell, 2015). Therefore, the optimal operating conditions are needed to select with the aim to obtain high molecular hemicelluloses with high product yield.

Figure 11 shows the trend of hemicellulose dissolution in different temperature. It can be seen that for a fixed extraction time the product yield of hemicellulose increases substantially with the increase in temperature. For the fixed temperature, the extraction time has a significant influence on the product yield. The Figure shows that the initial extraction, the first 20 min, is critically important for the whole extraction process.

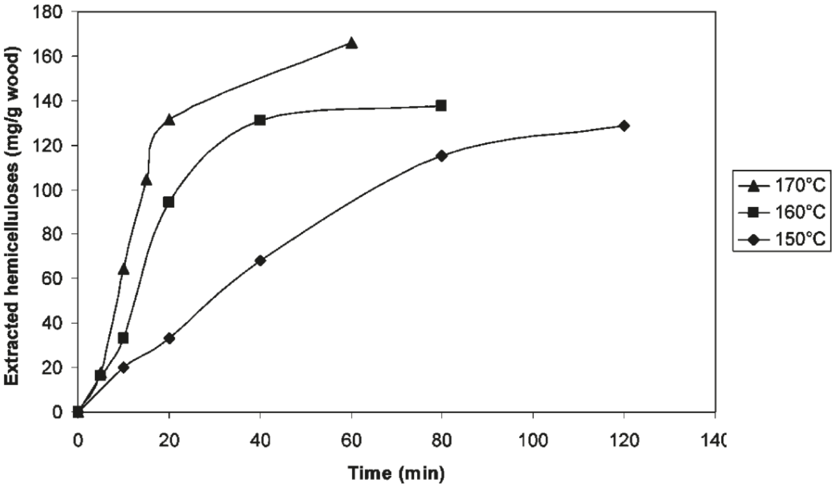


Figure 11: Concentration of extracted hemicellulose in the liquid phase as a function of time (Grénman et al., 2011)

Figure 12 shows the molecular weight of hemicelluloses as a function of extraction time. It can be clearly seen that the lower temperature, e.g. 160 ° C, is a better operating condition for obtaining the high molecular product. Also, the literature study reports that the degradation of hemicelluloses polymers occurs with the longer extraction time, where Figure shows the molecular weight of hemicelluloses decreases with an increase of extraction time. Therefore, the extraction time 20 minutes and temperature 160°C is considered for selecting the operation condition to obtain the polymer product with 20kDa.

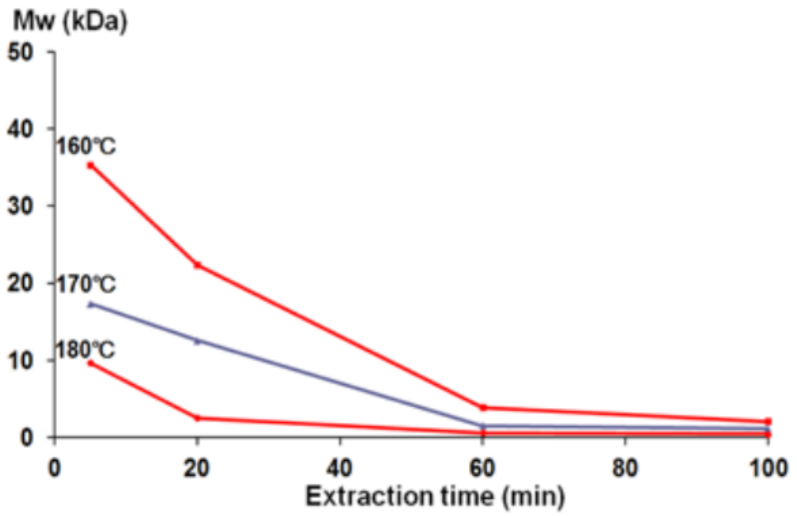


Figure 12: Molecular weight of extracted product vs. time (Song, 2013)

The molar mass of extracted hemicellulose is also influenced by the particle size of sawdust. Figure 13 shows the average molar mass of extracted hemicellulose obtained from extraction of different particle size. At the beginning of the extraction, there is a rapid decrease in molar mass until 20 minutes for all particle sizes. For a fixed extraction time, the smallest particle size (0.5-0.7 mm) has slightly higher molar mass, however, this influence is not severe. On the other hand, the particle size of sawdust has the certain influence on product yield (Figure 14). In the simulation, the particle size 0.5 -2 mm is considered for sawdust.

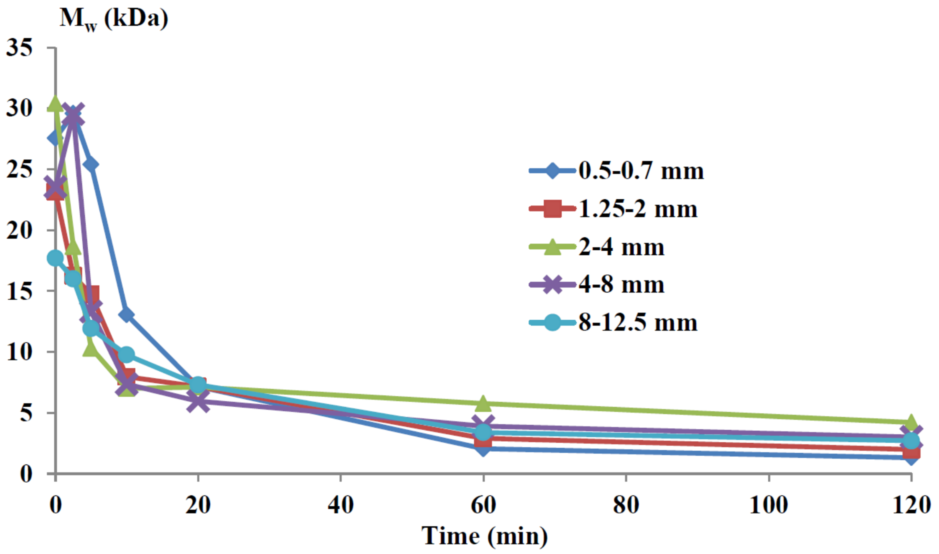


Figure 13: Average molar mass of extracted hemicellulose at 170 ˚C from different wood particle size. (Krogell, 2015)

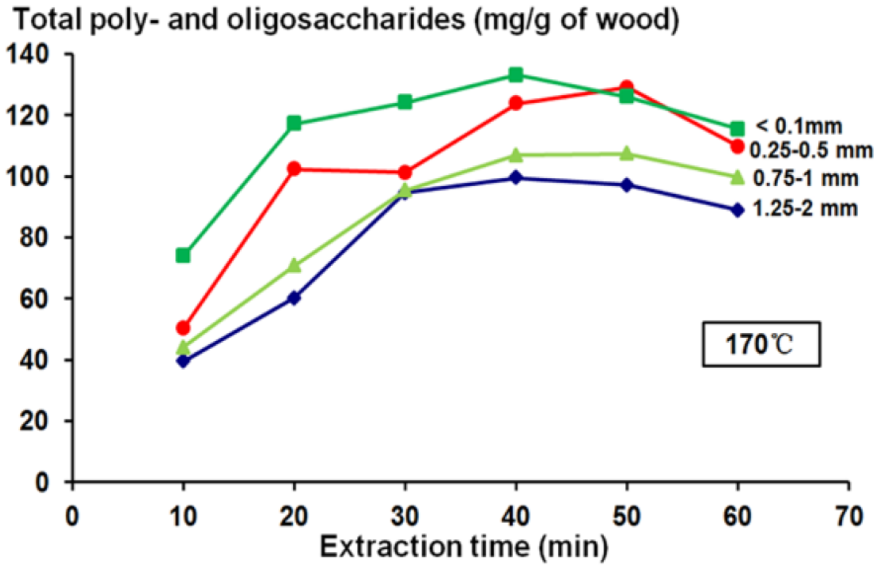


Figure 14: Total carbohydrates as anhydrous sugars extracted from the wood of different particle sizes at 170°C. (Song, 2013)

Other operating conditions are also evaluated in simulation. For example, pH has a small influence on the molar mass of extracted hemicellulose. Therefore, it was not considered in the simulation. The mass ratio of sawdust to water is important in auto-hydrolysis process and also material transportation in a chemical process. In the simulation, Liquid to wood ratio was considered 7:1. Such ratio was considered in order to avoid any problems in the circulation due to smaller particle size of the sawdust according to the practical experimental study of extraction of hemicelluloses (Xu et al., 2016)

6.3 Introduction to Aspen plus

The simulations of hemicellulose extraction from sawdust have been carried out using Aspen Plus. Aspen Plus is a simulation modeling software for chemical process. It was basically developed by the Massachusetts Institute of Technology for the U.S. Department of Energy in order to asses synthetic fuel technologies. This program encompasses a collection of standard unit operation blocks (e.g. pumps, reactors, heat exchangers, flash, dryer, distillation column), that serve as processes taking place in a real chemical plant. The simulation is carried out by defining the specification of unit operations as well as specifying the flow of material and energy streams. Moreover, Aspen Plus also contains a broad area of components database bearing physical properties of a huge amount of pure components. Mathematical routines such as convergence algorithm exist for determining different equations of material and energy balances as well as equilibrium equations. Sequential-modular approach is used to flow sheet convergence. Mass and energy balances for each unit operation blocks are computed sequentially.

Apart from the unit operation blocks, Aspen Plus uses two more mechanisms to simulate chemical processes; FORTRAN blocks and design specifications. For feed-forward control, FORTRAN blocks are used which allow incorporation of user code into a model to control variables. For feedback control, the design specification is employed. This allows the user to set particular design values for any flow sheet variable. Feed stream or block input can be manipulated by the design specification in order to achieve the design value. The flowing three steps are required for constructing a process model in a flow sheet program:

- Flow sheet definition: every inlet streams to the system needs to be specified along with various unit operations and their interrelated streams. All the outlet streams also specified in the flowsheet.
- Chemical components: every chemical component in the process must be defined.
- Operating conditions: all the operating conditions, such as temperature, pressure, heat duties for each unit operation needs to be defined. (Magnusson, 2006)

6.4 Process Simulation

Depending on the calculation pathways selected for basic properties such as liquid, vapor and solid enthalpy and density, the minimal physical properties is necessary by Aspen Plus. In the standard Aspen Plus database, the physical property data for the different components of wood are not available. Therefore, some assumptions were made in order to carry out the simulation. Total five components were taken into consideration for this simulation. They are WATER, SAWDUST, HEMI-S (Hemicellulose in solid state), HEMI-L (Hemicellulose in the liquid state after dissolution) and AIR. Water and air are conventional components and their physical properties are available in Aspen Plus databanks. Wood is composed of cellulose (40%), hemicellulose (30%) lignin (25%) and extractives (5%). Many of the physical and chemical properties of hemicellulose resemble with cellulose. As around 70% of wood is consist of cellulose and hemicellulose, sawdust was considered as cellulose in the solid state from databanks. Hemicellulose in the solid and liquid phase was defined as dextrose where the molar mass is 180.16 and boiling point was set high enough (600 K) in order to avoid vaporization during flash simulation.

Table 6: List of equipment and used symbols in the simulation diagram

Equipment type	Equipment symbol
Heater	H100
Heat exchanger	E101
Heat exchanger	E102
Heat exchanger	E103
Reactor(RCSTR)	R100
Mixing tank	M100
Flash 1	F101
Flash 2	F102
Pump	P100
Solid(sawdust) separator	S100
Solid(sawdust) dryer	D100

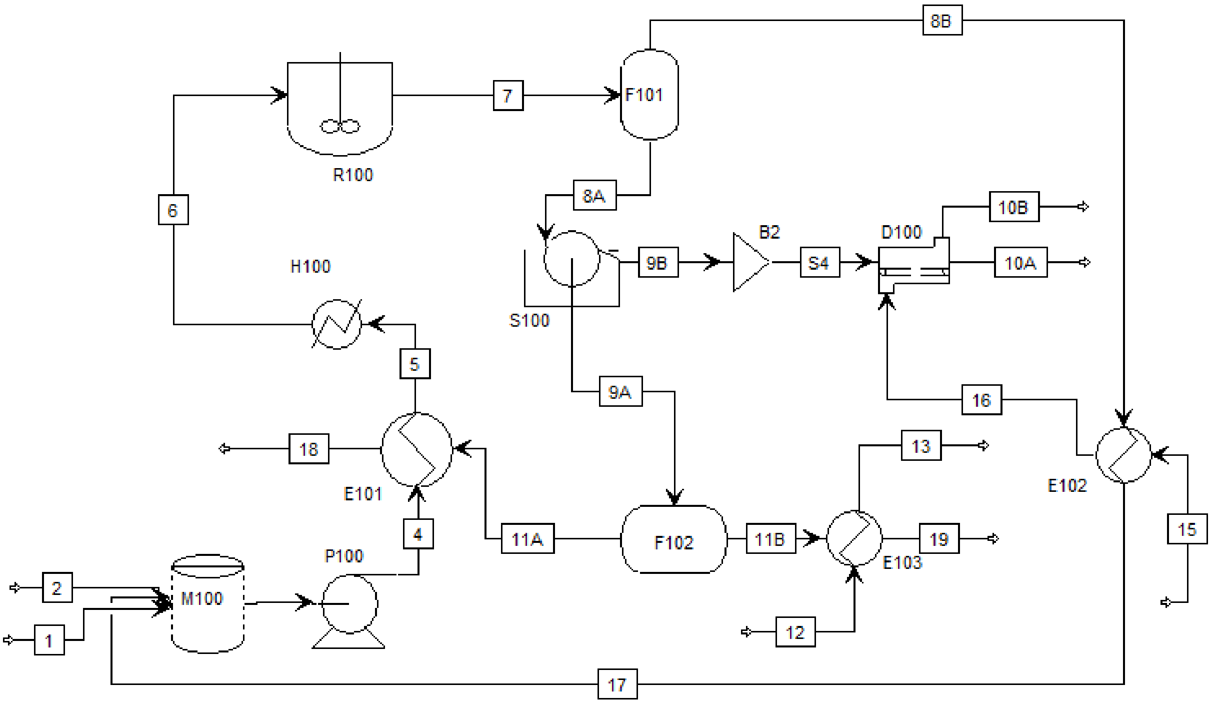


Figure 15: Simulation diagram for hydrolysis of sawdust and product separation

The IDEAL property method is selected as it is a recommended method for solids processing. The simulation diagram is shown in Figure 15 and list of equipment used in the simulation with symbol can be seen in Table 6

6.4. Pinch analysis

Pinch technology provides a methodical approach for optimal energy integration in a process based on thermodynamic principles. The improvements in the process related to this technology are not due to the use of a better unit operation but to the formation of heat integration design. There are many advantages of pinch technology. One of the main advantages of pinch analysis over traditional design method is the ability to define energy target for the process design. The minimal theoretical energy needed for the entire process is the energy target. The fundamental objective of pinch analysis is to match cold and hot process streams with a network of exchangers for minimizing the demands for externally supplied utilities. Pinch technology set up a minimum temperature difference, named as the pinch point which split the overall operating temperature region observed in the process into two temperature regions. After establishing the pinch point, heat from external resources needs to be provided to the process only at temperatures above the pinch. Below the pinch, heat needs to be removed from the process by cooling media. Such procedure will maximize the heat recovery in the process with the formulation of a heat exchanger network based on pinch analysis. (Peters and Timmerhaus, 1991)

Pinch analysis was carried out using Aspen Energy Analyzer Version V8.6 for large scale operation only. Process streams summary are given in Table 7.

Table 7: Stream data for pinch analysis

Stream Number	Stream Type	Stream Description	Temperature interval (°C)		Flowrate kg/h	MCp (kJ/C-s)	Effective Cp (kJ/kg-C)	Enthalpy MW
H1	Hot	Flash F102 effluent	110	98	6.15 x10^5	-	-	326.7
H2	Hot	Flash F101 effluent	101.1	98	4.91 x 10^4	-	-	30.90
H3	Hot	Product stream	110	38	9336	0.3799	2.888	0.54
C1	Cold	Reactor feed	20	160	6.47 x 10^5	0.9718	4.200	105.7
C2	Cold	Drying gas (air)	20	95	7.77 x 10^5	0.2794	1.006	16.28

Composite curves (CC) and grand composite curve (GCC) obtained from the simulation are illustrated in Figure 16 and Figure 17. The overlap between the two curves in the composite curves represents the possibilities for internal heat exchange. The sections of the curve which do not overlap represent the need for external heating and cooling. The temperature at which the curves approach each other with the smallest difference is the pinch temperature. The pinch temperature is 92.6 ˚C which can be seen from the GCC curve in Figure 17. In Figure 16, the composite curves show the overall profiles of heat availability and heat demand in the process over the entire temperature range. The hot composite curve is constructed by calculating total heat content of all the hot streams at the various temperature intervals. The cold composite curve is constructed by calculating total heat demand of all the cold streams at the different temperature intervals.

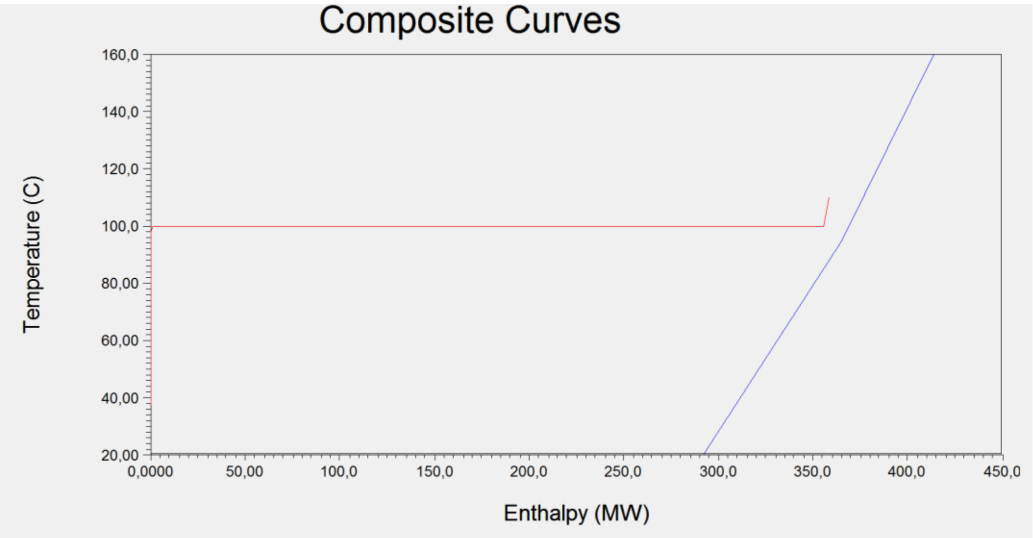


Figure 16: Hot and cold Composite curves where red streams represent hot streams (require cooling) and blue streams represents cold streams (require heating).

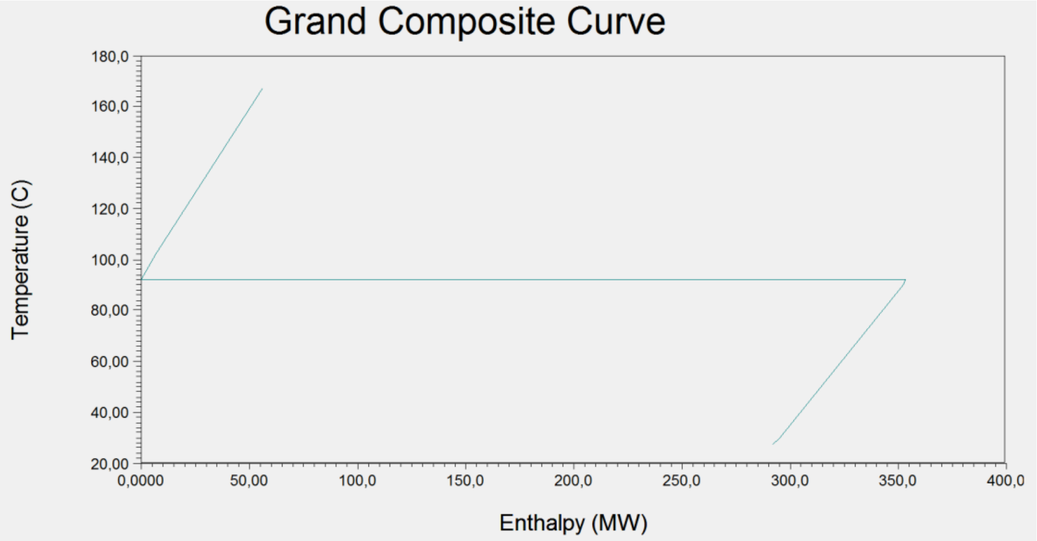


Figure 17: Grand composite curve

Composite curves are constructed by fixing the location of the composite curves with respect to one another with the use of a predefined value of ∆Tmin. ∆Tmin on the composite diagram is where the two curves most closely arrive at each other in temperature when measured in a vertical direction. For heat exchanger networks, the range of 3 to 40°C is considered as the optimal value for ΔTmin. The value is unique for each network and it is necessary to defined before the pinch analysis. If no cooling media are necessary below about 10◦C, the optimal ΔTmin is usually in the range of 10 to 40◦C (Peters and Timmerhaus, 1991). For this simulation, ΔTmin of 15°C is considered.

6.5 Heat integration results

The pinch temperature of the studied system is 92.6 °C, which means that the system has a deficit of heat above and excess of heat below 92.6 °C. The minimum heating and cooling demands are also illustrated by the composite curve in Figure 16. The minimum heating demand is 56.06 MW and cooling demand is 292.6 MW. The potential for internal heat exchange equals to 66 MW. Therefore, 66 MW of energy can be recovered from the hot streams of the process. This energy can be utilized for reactor feed heating as well as drying of sawdust. The cooling demand can be minimized by utilizing the hot stream energy for any other process integrated with pulp and paper mill. Total three heat exchangers were used in order to accomplish this energy recovery and cooling. The detail design specifications of the heat exchangers are given in Appendix 3 for small scale and Appendix 4 for large scale.

6.6. Process description

The process flow diagram is shown in Figure 18 and the stream tables are given in Table 8 for large scale and Table 9 for small scale. Sawdust and water at room temperature (20 °C) are mixed in the mixing tank M100. In the mixing tank, 8% water is recycled water and 92% is feed water. The mass ratio of water to sawdust is 7:1. Pump P100 is used to transport the sawdust and water mixture into the reactor, meanwhile, the pressure is raised from 1 bar to 7 bars. Heat exchanger E101 is used to recover heat from steam coming out from the Flash F102 and the temperature of the mixture increases to 95 °C. Heater H100 is used to raise the temperature of the mixture from 95 ° C to 160° C. In the reactor, R100 extraction is carried out at 160° C and 7 bars. After 20 minutes of extraction time, Flash F101 is used to flash out the steam. The solid-liquid mixture, which contains the extracted hemicellulose and the sawdust, is then fed into rotary vacuum filter S100 to separate the sawdust from the liquid phase. The separated sawdust has a moisture content of 32%. For pelletizing, 8 -12 % moisture content is required in the sawdust. Therefore, dryer D100 is used to dry the sawdust and 10% of final moisture content is obtained in the sawdust.

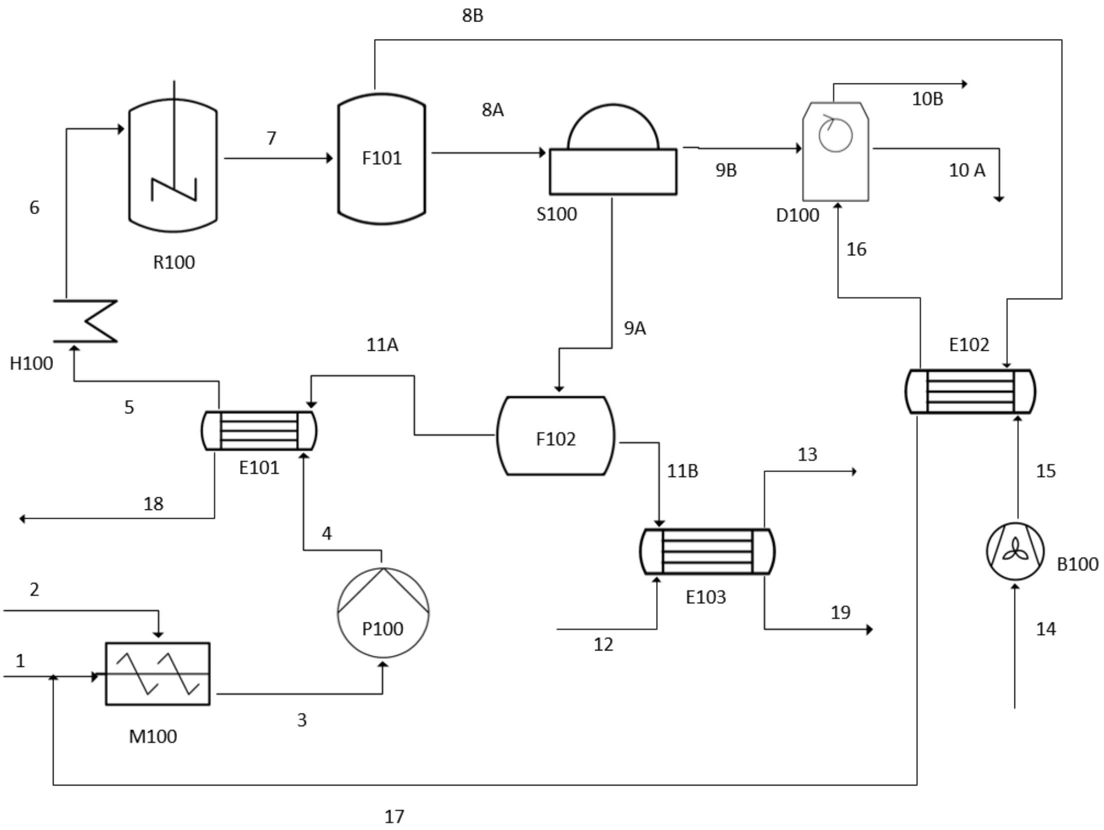


Figure 18: Process flow diagram

Table 8. Stream table (Large scale)

STREAM NUMBER	1	2	3	4	5	6	7	8A	8B	9A	9B
TO	M100	M100	P100	E101	H100	R100	F101	S100	E102	F102	D100
FROM			M100	P100	E101	H100	R100	F101	F101	S100	S100
PHASE	LIQUID	SOLID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID
Mass Flow (ton/hr)											
WATER	541,5	0,0	590,6	590,6	590,6	590,6	590,6	541,5	49,1	516,4	25,1
SAWDUST	0,0	59,9	59,9	59,9	59,9	59,9	59,9	59,9	0,0	0,0	59,9
HEMI-S	0,0	24,5	24,5	24,5	24,5	24,5	16,6	16,6	0,0	0,0	16,6
HEMI-L	0,0	0,0	0,0	0,0	0,0	0,0	7,9	7,9	0,0	7,5	0,4
AIR	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Temperature (˚C)	20,0	20,0	45,4	45,6	84,7	160,0	160,0	99,7	99,7	99,7	99,7
Pressure (bar)	1,0	1,0	1,0	7,0	7,0	7,0	7,0	1,0	1,0	0,5	0,5
Vapor Frac	0,0		0,0	0,0	0,0	0,0	0,0	0,0	1,0	0,0	0,0
Liquid Frac	1,0		1,0	1,0	1,0	1,0	1,0	1,0	0,0	1,0	1,0
Total Flow (ton/hr)	541,5	84,4	675,0	675,0	675,0	675,0	675,0	625,9	49,1	523,9	102,0
Enthalpy (MW)	-2389,5	-0,1	-2588,4	-2588,3	-2559,8	-2499,2	-2532,8	-2351,6	-181,2	-2243,8	-107,9

Table 8: Stream table (Large scale) continued

STREAM NUMBER	10A	10B	11A	11B	12	13	14	15	16	17	18	19
TO			E101	E103	E103		B100	E102	D100	M100		
FROM	D100	D100	F102	F102		E103		B100	E102	E102	E101	E103
PHASE	LIQUID	VAPOR	VAPOR	LIQUID	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	MIXED	MIXED	LIQUID
Mass Flow (ton/hr)												
WATER	7,7	17,5	514,6	1,8	40,0	40,0	0,0	0,0	0,0	49,1	514,6	1,8
SAWDUST	59,9	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
HEMI-S	16,6	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
HEMI-L	0,4	0,0	0,0	7,5	0,0	0,0	0,0	0,0	0,0	0,0	0,0	7,5
AIR	0,0	909,6	0,0	0,0	0,0	0,0	909,6	909,6	909,6	0,0	0,0	0,0
Temperature C	77,6	50,6	109,5	109,5	20,0	32,7	20,0	20,0	89,6	99,6	99,6	37,7
Pressure bar	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Vapor Frac	0,0	1,0	1,0	0,0	0,0	0,0	1,0	1,0	1,0	0,4	0,9	0,0
Liquid Frac	1,0	0,0	0,0	1,0	1,0	1,0	0,0	0,0	0,0	0,6	0,1	1,0
Total Flow ton/hr	84,5	927,1	514,6	9,3	40,0	40,0	909,6	909,6	909,6	49,1	514,6	9,3
Enthalpy MW	-33,0	-58,5	-1895,9	-22,2	-176,5	-175,9	-1,3	-1,3	16,4	-198,8	-1924,3	-22,7

Table 9: Stream Table (Small scale)

STREAM NUMBER	1	2	3	4	5	6	7	8A	8B	9A	9B
TO	M100	M100	P100	E101	H100	R100	F101	S100	E102	F102	D100
FROM			M100	P100	E101	H100	R100	F101	F101	S100	S100
PHASE	LIQUID	SOLID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	LIQUID	VAPOR	LIQUID	LIQUID
Mass Flow ton/hr											
WATER	90,3	0,0	98,4	98,4	98,4	98,4	98,4	90,3	8,2	86,1	4,2
SAWDUST	0,0	10,0	10,0	10,0	10,0	10,0	10,0	10,0	0,0	0,0	10,0
HEMI-S	0,0	4,1	4,1	4,1	4,1	4,1	2,8	2,8	0,0	0,0	2,8
HEMI-L	0,0	0,0	0,0	0,0	0,0	0,0	1,3	1,3	0,0	1,3	0,1
AIR	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
Temperature (˚C)	20,0	20,0	46,6	46,8	94,7	160,0	160,0	99,7	99,7	99,7	99,7
Pressure (bar)	1,0	1,0	1,0	7,0	7,0	7,0	7,0	1,0	1,0	0,5	0,5
Vapor Frac	0,0		0,0	0,0	0,0	0,0	0,0	0,0	1,0	0,0	0,0
Liquid Frac	1,0		1,0	1,0	1,0	1,0	1,0	1,0	0,0	1,0	1,0
Total Flow (ton/hr)	90,3	14,1	112,5	112,5	112,5	112,5	112,5	104,3	8,2	87,3	17,0
Enthalpy (MW)	-398,2	0,0	-431,3	-431,2	-425,4	-416,5	-422,1	-391,9	-30,2	-374,0	-18,0

Table 9: Stream Table (Small scale) continued

STREAM NUMBER	10A	10B	11A	11B	12	13	14	15	16	17	18	19
TO			E101	E103	E103		B100	E102	D100	M100		
FROM	D100	D100	F102	F102		E103		B100	E102	E102	E101	E103
PHASE	LIQUID	VAPOR	VAPOR	LIQUID	LIQUID	LIQUID	VAPOR	VAPOR	VAPOR	MIXED	MIXED	LIQUID
Mass Flow ton/hr												
WATER	1,3	2,9	85,8	0,3	6,5	6,5	0,0	0,0	0,0	8,2	85,8	0,3
SAWDUST	10,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
HEMI-S	2,8	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0	0,0
HEMI-L	0,1	0,0	0,0	1,3	0,0	0,0	0,0	0,0	0,0	0,0	0,0	1,3
AIR	0,0	134,9	0,0	0,0	0,0	0,0	134,9	134,9	134,9	0,0	0,0	0,0
Temperature (˚C)	61,1	52,1	109,5	109,5	20,0	33,0	20,0	20,0	94,7	99,6	99,6	38,0
Pressure (bar)	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0	1,0
Vapor Frac	0,0	1,0	1,0	0,0	0,0	0,0	1,0	1,0	1,0	0,5	0,9	0,0
Liquid Frac	1,0	0,0	0,0	1,0	1,0	1,0	0,0	0,0	0,0	0,5	0,1	1,0
Total Flow (ton/hr)	14,1	137,8	85,8	1,6	6,5	6,5	134,9	134,9	134,9	8,2	85,8	1,6
Enthalpy (MW)	-5,6	-9,8	-316,0	-3,7	-28,7	-28,6	-0,2	-0,2	2,6	-33,0	-321,8	-3,8

For drying, air at room temperature is heated by the recovered heat from steam coming out from flash F101 by using the heat exchanger E102. The liquid phase containing the dissolved hemicellulose is then evaporated using flash F102. The final product, containing 80% hemicellulose and 20% water, is cooled down to 38 ˚C by heat exchanger E103.

6.4 Reactor simulation

Continuous stirred tank reactor (RCSTR) was used to carry out the extraction simulation. Power law kinetic expression was considered using mass concentration as concentration units. The pre-exponential factor, k_0 was converted from 0.0363 (1/min) to 3.3611×10^{-6} (kmol/s/m3)/(kg/m^3) in the simulation. Another important operating parameter in reaction simulation is pressure. Saturated water is used as a solvent in the extraction process. According to the thermos-physical properties of saturated water, the pressure is 6.18 bars for 160˚C saturated water (Incropera et al., 2006). Therefore, 7 bars is considered as operating pressure for reactor simulation. The detail design specification of the reactor is given in Appendix 3 for small scale and Appendix 4 for large scale**.**

6.5 Product separation

Product separation and purification is an essential step in any chemical process. Separation process depends on the type of mixture to be handled. There are two kinds of mixture: heterogeneous and homogeneous. The heterogeneous mixture has two or more phases that are physically separate. The heterogeneous mixtures can consist of two liquids, solid-liquid, liquid-gas and gas-solid. In the homogenous mixture, the constituent phases are identical and they converge into a single uniform phase. The most common homogenous mixtures are liquids as well as gasses and solids (alloys). In this process both homogenous (water and hemicellulose) and heterogeneous (sawdust and water) mixture is present. Therefore, selection and simulation of this two different mixture types are discussed separately.

Sawdust and liquid phase separation

Solid-liquid separation is the most common separation requirement in the process industry. Selection of appropriate separation depends on solids concentration, feed rate as well as the size and nature of the solid particles. The selection of equipment also depends on whether the main aim is to obtain a purified liquid or a solid product, and dryness of the solid required. Characteristics of the performance of the main types of commercially available solid-liquid separation equipment are summarized in appendix 5. It can be seen that there are several types of equipment available for solid-liquid separation. They are filtration, sedimentation, centrifugation, liquid cyclones, screens and ultrafiltration. The particle size of sawdust is in the range of 0.5-2 mm which can be considered as medium size particles. Therefore the appropriate equipment could be filtration, sedimentation and centrifugation (Sorsamäki and Nappa, 2015). Sedimentation is carried out for mainly for pre-concentration of feeds before filtering. Considering the direct cost, separation efficiency and equipment characteristics, vacuum drum filter is chosen and simulated for separating sawdust from the liquor. The detail design specifications are given in Appendix 3 for small scale and Appendix 4 for large scale.

Sawdust drying

After separation of sawdust from the liquor, it needs to be dried in order to proceed with palletization. Drying of sawdust involves a combination of heat transfer and mass transfer. Heat is transferred to the surface of the sawdust to provide the heat of vaporization. The water evaporates and diffuses out from the surface into the bulk of the drying gas or air. Flowing gas is used to ensure that there is always an adequate amount of partial pressure driving force for evaporation.

Convection type dryer is considered for this process. There are several types of equipment for the convective dryer. They are batch tray dryer, conveyor dryer, rotary dryer, fluidized-bed dryer, pneumatic dryer, spray dryer, rotary drum dryer. Considering the continuous operation and relatively low capital and operating cost, the rotary dryer is suitable for drying sawdust. The main cost of the rotary dryer is the heater for the air. But for this process, the rotary dryer was simulated using heat recovered from the steam produced in the Flash unit F101. The steam was used to heat air to 95 ° C which was then fed into the dryer. Therefore, a significant amount of operating cost can be minimized. The detail design specification of the dryer is given in Appendix 3 for small scale and Appendix 4 for large scale.

Separation of hemicellulose from liquid mixture

After separating sawdust from, we have the liquid stream that contains water and dissolved hemicellulose. The mass fraction of water and hemicellulose is 0.985 and 0.015 which means the maximum amount is water. The main product hemicellulose needs to be separated from water. A flash unit which is a single-equilibrium-stage distillation is considered for this stage. In this kind of separation unit, the feed is partially vaporized to give a vapor richer than the feed in the more volatile components. An isothermal flash at atmospheric pressure is simulated. The mass fraction hemicellulose and water in the product stream after flash separation is 0.80 and 0.20 respectively. The design specifications of the two flashes vessel used in the process are given in Appendix 3 for small scale and Appendix 4 for large scale.

6.6 Simulation results and discussion

The simulation was performed in two different scales. One is in small scale and the other is in large scale. The small scale capacity is 1.5 ton/hour whereas the large scale capacity is 9.3 tons/hour. The summary of the simulation results is given in Table 10. Net energy calculation shows the data for energy input for the whole process, however, some certain type of energy consumption, e.g. energy for the mixer and solid-liquid separator, blower were not included in the simulation. The maximum amount of energy is needed for evaporation of the product stream and heating of reactor feed in the form of electric energy. Expensive electric energy consumption can be minimized by fired heating using pellets that will be produced after extracting hemicellulose from the sawdust.

Table 10: Simulation results in two different scale with extraction temperature 160 ˚C and time 20 min.

Operation Scale	Feed Water (ton/h)	Recycled water (ton/h)	Total water (ton/h)	Sawdust (ton/h)	Amount of hemicellulose in feed (ton/h)	Amount of hemicellulose remain after extraction	Product (ton/h)	Conversion %	Yield %	Net energy MW
Small	90	8	98.4	14	4	2.7	1.5	9.32	8.89	63.12
Large	542	49	591	84	25	16.6	9.3	9.32	8.89	386.5

The conversion efficiency is around 9.3 % and yield is 8.89 % for both operating scale. The product has 80% of hemicellulose and 20 % of water. Water can be evaporated subsequently to produce final product hemicellulose. In the real case, the product may contain impurities such as acetic acid, extractives, and low molecular weight hemicellulose. Therefore, membrane filtration would be suitable for separating desired high molecular weight hemicellulose.

The steam produced in the Flash F102 is not recycled in this process. The only heat was recovered in order to heat up the reactor feed. This stream has a vapor fraction of 0.9. In order to recycle the water, it requires a huge amount of cooling water as well as an additional heat exchanger which will increase the plant capital cost. Therefore, it was not considered to recycle this steam but the steam can be utilized in any other process integrated with pulp and paper mill.

7. Economic evaluation

7.1 Estimation of Total Capital Investment

Economic evaluation of both small scale and large scale operation for extraction of hemicellulose from sawdust is reported here. At first capital investment costs are estimated based on equipment purchased costs. The purchased equipment costs are estimated from the cost curves (LOH, Lyons, and White, 2002; Towler and Sinnott, 2007). The equipment purchased cost estimation procedures are explained in Appendix 6. After that, the installation costs are obtained through the percentage of purchased equipment cost as explained in Appendix 7. Piping cost is considered 31% of total purchased equipment cost for solid-fluid processing plant (George et al., 1988). These costs summed up to the inside battery limits or ISBL investment, from which the rest of the capital investment terms (OSBL, engineering, contingency, start-up, and working) are obtained. The list of formulas that are used to calculate all these costs are given in appendix 8. Estimation of total capital investment is reported in Table 11.

Table 11: Estimation of total capital investment

Purchased equipment cost	Small scale		Large scale	
ISBL				
Purchased equipment	Number of unit	Cost (k€)	Number of unit	Cost (k€)
1. Reactor-R100(RCSTR)	1	679	3	3 × 1139
2. Mixing Tank-M100	1	178	1	305
3. Centrifugal fan-B100	1	130	4	4 × 130
4. Heat exchanger E-101	1	62	3	223
5. Heat exchanger E-102	1	43	3	172
6. Heat exchanger E-103	1	27	1	28
7. Pump P-100	1	12	1	25
8. Vacuum drum filter-S100	1	230	3	3 × 279
9. Flash drum -F101	1	39	1	65
10. Flash drum F-102	1	39	1	65
11. Atmospheric tray dryer D-100	1	24	1	42
Total	**1462**		**5700**	
Installation costs	**502**		**1959**	
Piping costs	**453**		**1767**	
Total ISBL costs	**2416**		**9425**	
Total OSBL	**725**		**2828**	
Engineering and supervision	**604**		**2356**	
Contingency	**242**		**943**	
Total fixed capital investment	**3987**		**15552**	
Startup cost	**399**		**1555**	
Total capital investment	**4385**		**17107**	

7.2 Estimation of manufacturing cost

The manufacturing cost is calculated by summing fixed and variable costs. Variable manufacturing costs are determined by the cost of raw materials and utilities. The raw materials for this process are sawdust and water. Sawdust is considered free as the process is considered integrated with pulp and paper mill. The cost of fresh water for the reactor is considered 1.62 €/m^3 ((Lappeenrannan Energiaverkot oy, 2015). Cooling water is considered free but a cost is being estimated for pumping of cooling the water based on the equation 5.

$$P = Q \cdot \Delta P / \eta \qquad (5)$$

Where P is the power in kW, Q is the flow of cooling water in m^3/s, ΔP is the pressure drop in kPa, and η is the pumps efficiency. Both the pressure drop and the efficiency values are estimated as 100 kPa and 80 %, respectively. Total electricity requirement per year for both small and large scale operation is reported in Table 12. Summary of the variable manufacturing costs is given in Table 13.

Table 12: Determination of electricity requirements

	Small scale		Large scale	
Equipment	Power(kW)	MWh/year	Power(kW)	MWh/year
Electricity for pumping cooling water	0.50	4	3.08	24.64
Heater-H100	8850	70800	60600	484800
Flash- F102	54286	434288	325738	2605904
Rotary vacuum filter-S100	2.2	17.6	4	32
Pump-P100	24	192	100	800
Fan-B100	24.27	194	164	1312
Total	**63186,97**	**505495.8**	**386608.72**	**3092870**

Table 13: Variable manufacturing cost

Raw materials	**Units**	**Small scale**				**Large scale**			
		Units/year	€/Unit	k€/year	€/ton	Units/year	€/Unit	k€/year	€/ton
Water	m^3	722925	1,62	1171	94	4336000	1.62	7024	94
Electricity	MWh	505496	50	25275	2030	3092870	50	154643	2071
Total				**26446**	**2125**			**161668**	**2165**

Fixed manufacturing costs are estimated based on labor cost and other factors explained in Appendix 10. It is considered that one operator per shift will be sufficient to operate the plant. Therefore, the number of shift position is 3 for 8 hours each. The monthly wage for each operator is considered € 3000 along with 1.8 % of side cost and benefits. The total amount of labor cost is reported in Table 14 along with the other fixed manufacturing cost items. Total manufacturing cost is reported in Table 15. A fully operational year is considered to be 8000 hours taking maintenance into account.

Table 14: Fixed manufacturing cost

	Small scale		**Large scale**	
	k€/year	**€/ton**	**k€/year**	**€/ton**
Labor costs	195	16	195	3
Supervisions	10	1	10	0
Side costs and benefits	4	0	4	0
Maintenance	48	4	189	3
Insurance	31	3	123	2
Quality control	146	12	878	12
Interest	88	7	342	5
License fees	88	7	342	5
Depreciation	199	16	778	10
Administrative costs	13	1	20	0
Selling and marketing costs	545	44	3291	44
Total fixed cost of production	**1912**	**110**	**6169**	**83**

Table 15: Total manufacturing cost

Cost items	**Small scale**		**Large scale**	
	k€/year	**€/ton**	**k€/year**	**€/ton**
Variable manufacturing cost	26446	2125	161668	**2165**
Fixed manufacturing cost	1367	110	6169	**83**
Total manufacturing cost	**27812**	**2235**	**167837**	**2247**

7.3 Revenues, profitability analysis and discussion

The revenues were estimated from the price of the final product. The final product is bulk hemicellulose with 80% hemicellulose and 20% water. Market survey related to hemicellulose price and market demand were not the scope of this study. Based on the manufacturing cost and expected revenue, a reasonable price for bulk hemicellulose is considered € 2350/ton. Effect of price variation on profitability is discussed in sensitivity analysis section. In Table 16 the estimation of revenues is reported for both small and large scale of production.

Table 16: Revenue estimation

Operating scale	Production (tons/year)	Price (€/ton)	Revenue (k€/year)
Small scale	12448	2350	29253
Large scale	74688	2350	175517

After estimation of the revenues, the annual gross profit and profit per ton of product are calculated and reported in Table 17 and Table 18. From annual gross profit, return on investment and payout time and internal rate of return (IRR) were calculated. For profitability analysis, few assumptions were made such as; 20 years of plant life span, linear depreciation of 5% of fixed capital investment. Cumulative cash flow is calculated and plotted in Figure 19 and Figure 20 for both small and large scale operation. Net Present Values (NPV) are calculated for both small and large scale of operation based on cumulative cash flows and plotted in Figure 21 and Figure 22. The corresponding formulas are reported in Appendix 9.

Table 17: Profitability analysis

	Small scale k€/year	**Large Scale k€/year**
Revenues	29253	175517
Total manufacturing cost	27812	167837
Annual gross profit	1441	7680
Return on Investment (%ROI)	32.8	44.9
Payout time (year)	2.67	2.02
Internal Rate of Return(IRR)	32%	41%

Table 28: Profitability analysis per ton of product

	Small scale **€/ton of product**	**Large Scale** **€/ton of product**
Price	2350	2350
Manufacturing cost	2234	2247
Profit	116	103

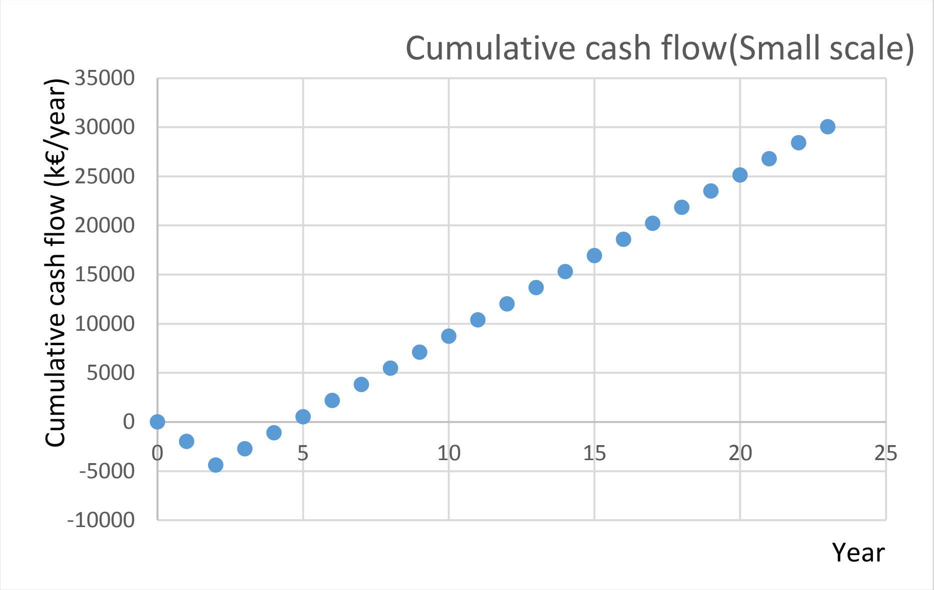


Figure 19: Cumulative cash flow vs operating year (Small scale)

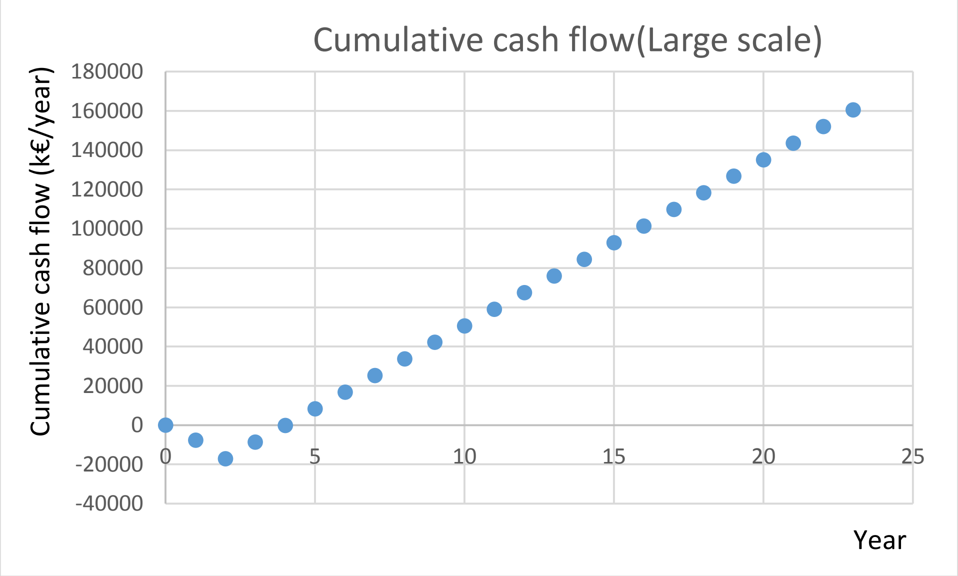


Figure 20: Cumulative cash flow vs. operating year (Large scale)

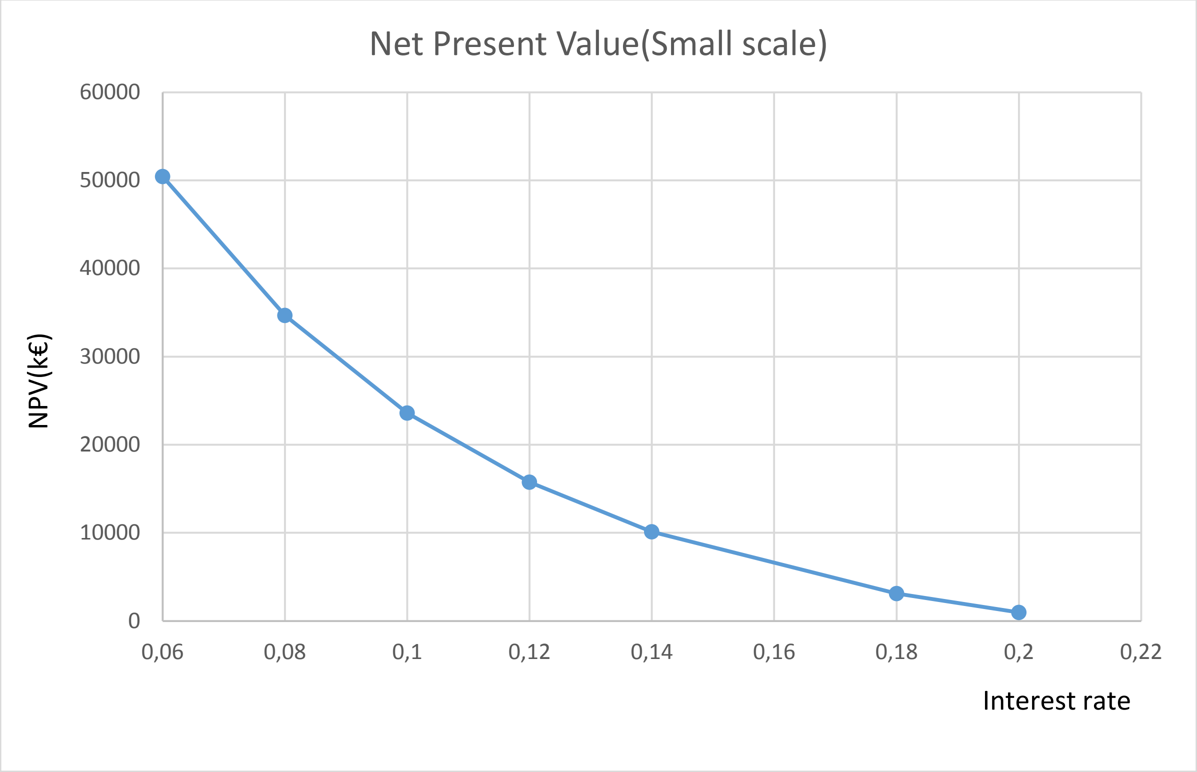


Figure 21: Net Present value (NPV) in different interest rate (Small scale)

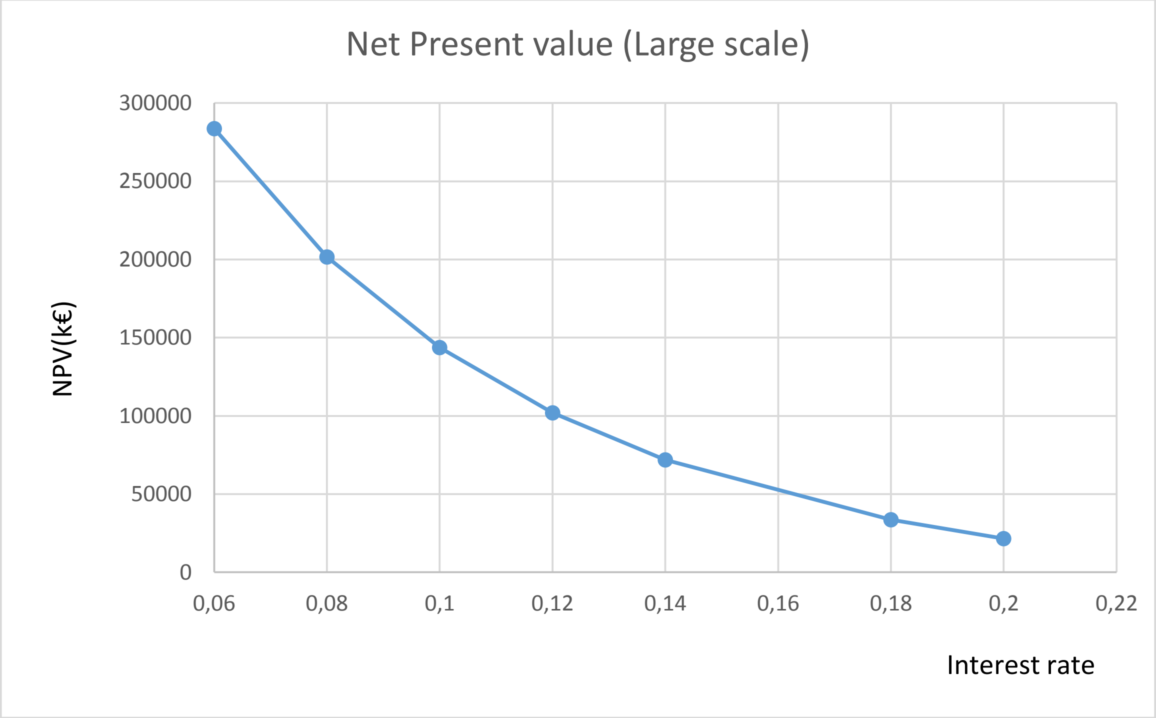


Figure 22: Net present value in different interest rate (Large scale)

Based on the size and scope of the project, the construction may require around from six months to three years to finish. A classical value of 2 years for the time from project inauguration to the start of the plant for production is considered. Over the two-year construction period, there are major capital expenses which can be seen from the cumulative cash flow diagram. This shows the fixed capital investment for purchasing and installing the equipment and auxiliary facilities necessary to start the plant. It is assumed that 50% of the total fixed capital investment was spent in the first year. Construction of the plant is completed at the end of the second year and the production is started. In this stage, the additional investment for working capital required to sail the first few months of operations is presented. This investment is an one-time expense at the start-up of the plant and will be redeemed at the end of the project. After start-up, the plant begins to produce final products for sale. Therefore, the yearly cash flows become positive. This is presented in the positive slope of the cumulative cash flow diagram reported in Figure 19 and Figure 20. The cash flows for the beginning of operation are bigger than those for subsequent years due to the effect of the depreciation allowance.

The NPV is always inferior to the total future value of the project due to the discounting of future cash flows. This is a function of the interest rate defined and the time period considered. NPV were calculated for the period of 23 years with the different interest rate. Net present value is a more useful economic measure than payback period and ROI, as it allows for the time value of money and also for annual variation in expenses and revenues.

Return on Investment is around 32.8% for small scale and 44.9% for large scale. The payout time is 2.7 years for small scale and 2 years for large scale. The IRR was calculated and they are 32% for small scale and 41% for large scale. If the IRR value is higher, it is favorable to invest into the project as future cash flows could be discounted at a higher rate until NPV equals zero. An IRR of 25% is considered the threshold for securing capital investment in a new chemical industry. Therefore, it is safe to invest in both small and large scale of operation for hemicellulose extraction from sawdust.

8. Sensitivity analysis

The main objective of sensitivity analysis is to determine those parameters that have powerful impact on project viability over the expected range of variation of the parameter. The investigated parameters are product price, production rate and fixed capital investment. The variation for product price and production rate were considered ±25% and for fixed capital invest the variation is ±20 %. Varying the production rate by keeping the investment and fixed cost constant investigates the effects of unexpectedly high down time due to maintenance or operation problems. In addition, it investigates also the unexpected difficulties in selling the full volume of product that could be produced.

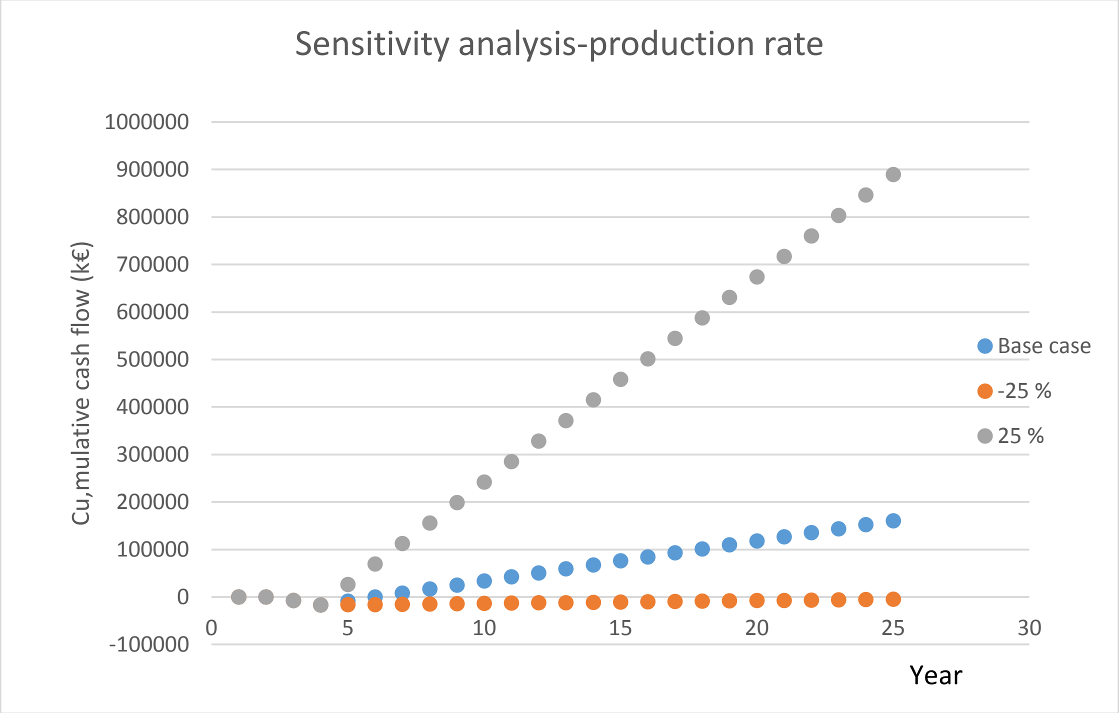


Figure 23: Sensitivity analysis of production rate

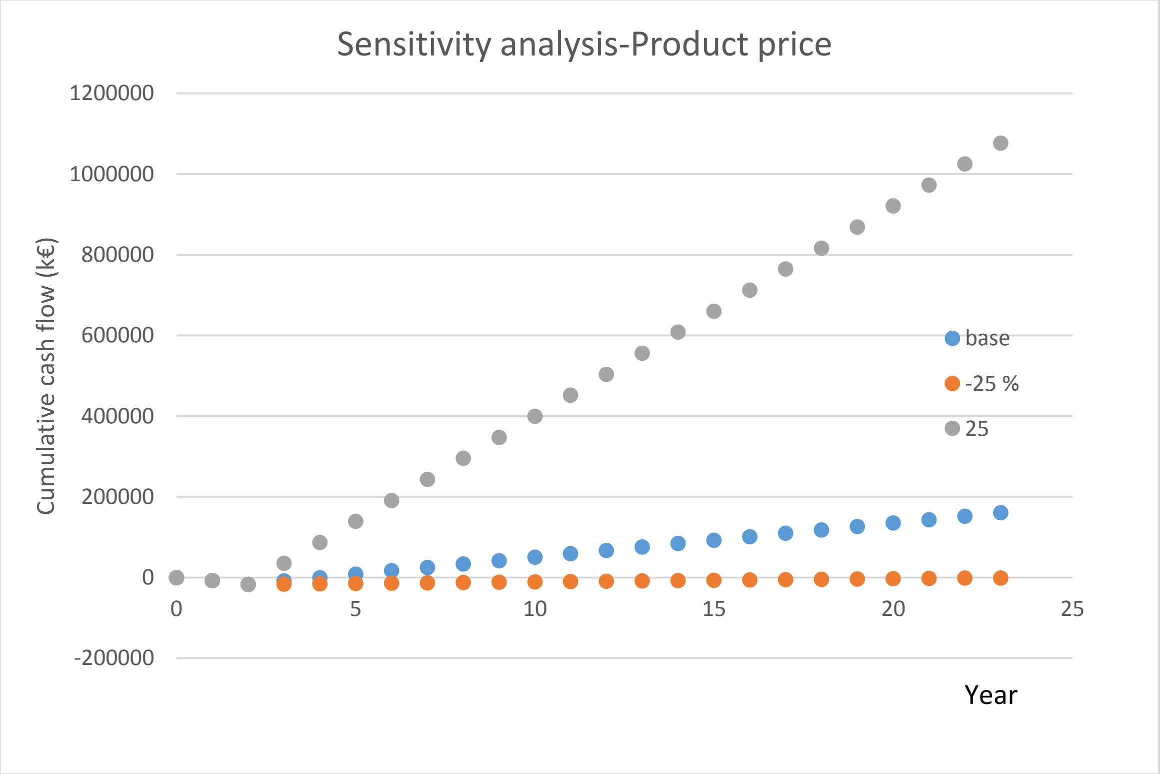


Figure 24: Sensitivity analysis for product price

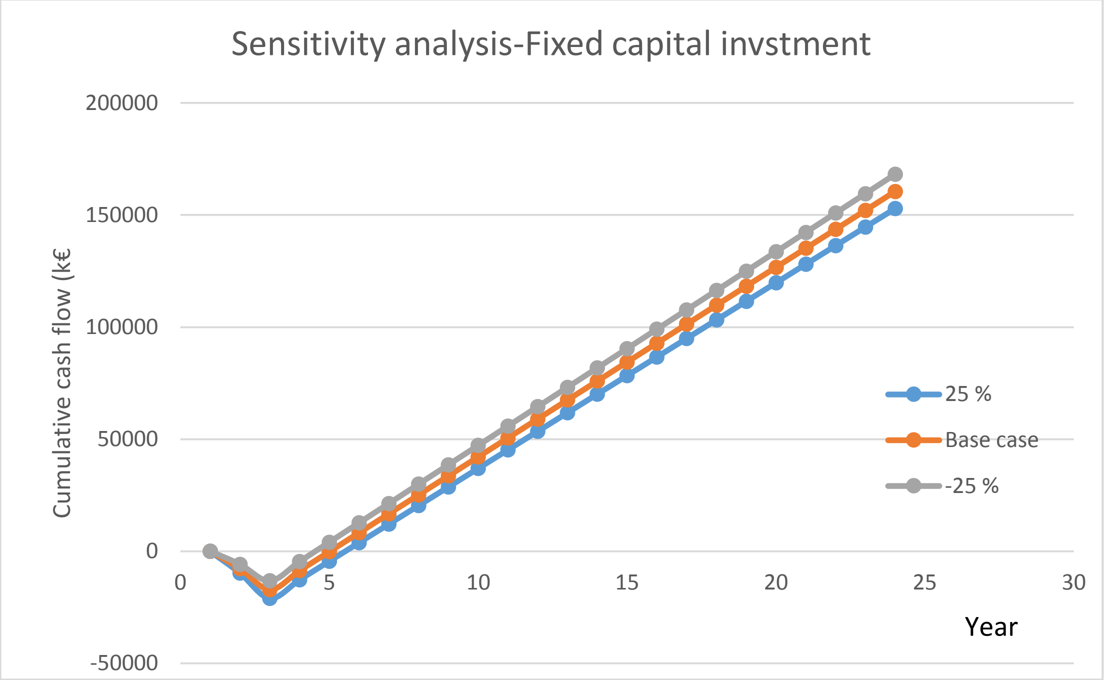


Figure 25: Sensitivity analysis for fixed capital investment

An increase in production rate beyond the design capacity might also be possible if the plant design margins allow some extra capacity or if the yields can be improved by use of better operating parameters. Varying the production rate and product price influences the cash flow significantly (Figure 22, Figure 23). Reduction of product price and production rate for 25% makes the project non-profitable. Variation of ±20% for fixed capital invest and their influence in cash flow is reported in Figure 24. Reduction of 20% fixed capital investment reduces payback period for 1 year. Therefore, it can be summarized that there are high degrees of risk in terms of product price and production rate for making an investment in the project and the project is highly sensitive to any fluctuation of product price and product rate. This is due to the high variable manufacturing cost coming from electricity cost.

9 Conclusion and further study

The main objective of this thesis work was to study the techno-economic feasibility of hemicellulose extraction from sawdust before pelletizing for small and medium scale biorefinery concept. The focus was to extract high molecular weight hemicelluloses (20-50 kDa) by autohydrolysis. Extraction process design and simulation was carried out using Aspen Plus simulation tool. Mass and energy balance for the whole process was estimated through simulation. RCSTR was simulated for extraction. The operating parameters for extraction were selected based on the extraction of high molecular weight hemicellulose. In order to accomplish this, extraction time of 20 minutes and temperature 160 °C were considered. Solid to liquid ratio was considered 1:7 in order to avoid any problem in the circulation of the sawdust and water mixture.

Pinch analysis was carried out in order to optimize the process for maximum energy recovery. It was found that around 66 MW (in large scale operation) of energy can be recovered and use for heating of reactor feed and drying of sawdust. The minimum heating demand needed for the process is 56 MW and cooling demand is 292.6 MW. It means a large amount of cooling water is needed for the process. The cooling demand can be minimized by utilizing the steam for any other process if it is integrated with existing pulp and paper industry. The conversion and yield of hemicellulose from sawdust were obtained 9.3% and 8.89% respectively. The product was 80% hemicellulose and 20% water. The conversion and yield can be increased by using multistage process or by tuning the extraction parameters such as increasing extraction time. But increasing extraction time might reduce the molecular weight of extracted hemicellulose.

The economic evaluation was done in order to study the feasibility of the extraction process. Small and large scale operation was considered for evaluation. The small scale production capacity was 1.5 ton / h and the large scale capacity was 9.3 ton/h. The return on investment for small scale and large scale operation was 32.8% and 44.9% respectively. The payout time for small scale is 2.67 years and for large scale is 2.02 years. Moreover, the Internal Rate of Return (IRR) was found 32% for small scale and 41% for large scale. It is considered safe to invest in a project that has IRR higher than 25%. Sensitivity analysis shows that variation of product price and production rate have high degrees of risk for investing in the project.

The maximum amount of manufacturing cost comes from the electricity cost. Reactor heating and flash separation of the product is responsible for the maximum energy demand. Due to the cost of electricity, the project is very sensitive towards the fluctuation of production rate and product market price. Electricity demand can be minimized by utilizing pellets for reactor heating and flash separation. If it will be possible to minimize the demand for electricity, the profitability of the project will increase significantly and will eliminate all kind of risks. Therefore, further study is needed in order to redesign the process for using pellets as heating sources and further evaluation of the project.

In this study, for the sake of simplicity, it was considered that only hemicellulose dissolution occurs during extraction and liquid phase contains only dissolved hemicellulose. In reality, this is not true. Liquid phase will contain acetic acid as well as some extractives. This was not considered in the modeling of hemicellulose extraction. Further research needs to be done to consider the presence of these compounds in the liquid phase and their subsequent separation.

Finally, it can be summarized that extraction of hemicellulose by autohydrolysis from sawdust before pelleting is technically and economically feasible except the fluctuation of product price and production rate.

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Appendix 1. Purchased equipment cost for common plant equipment (Towler and Sinnot, 2007)

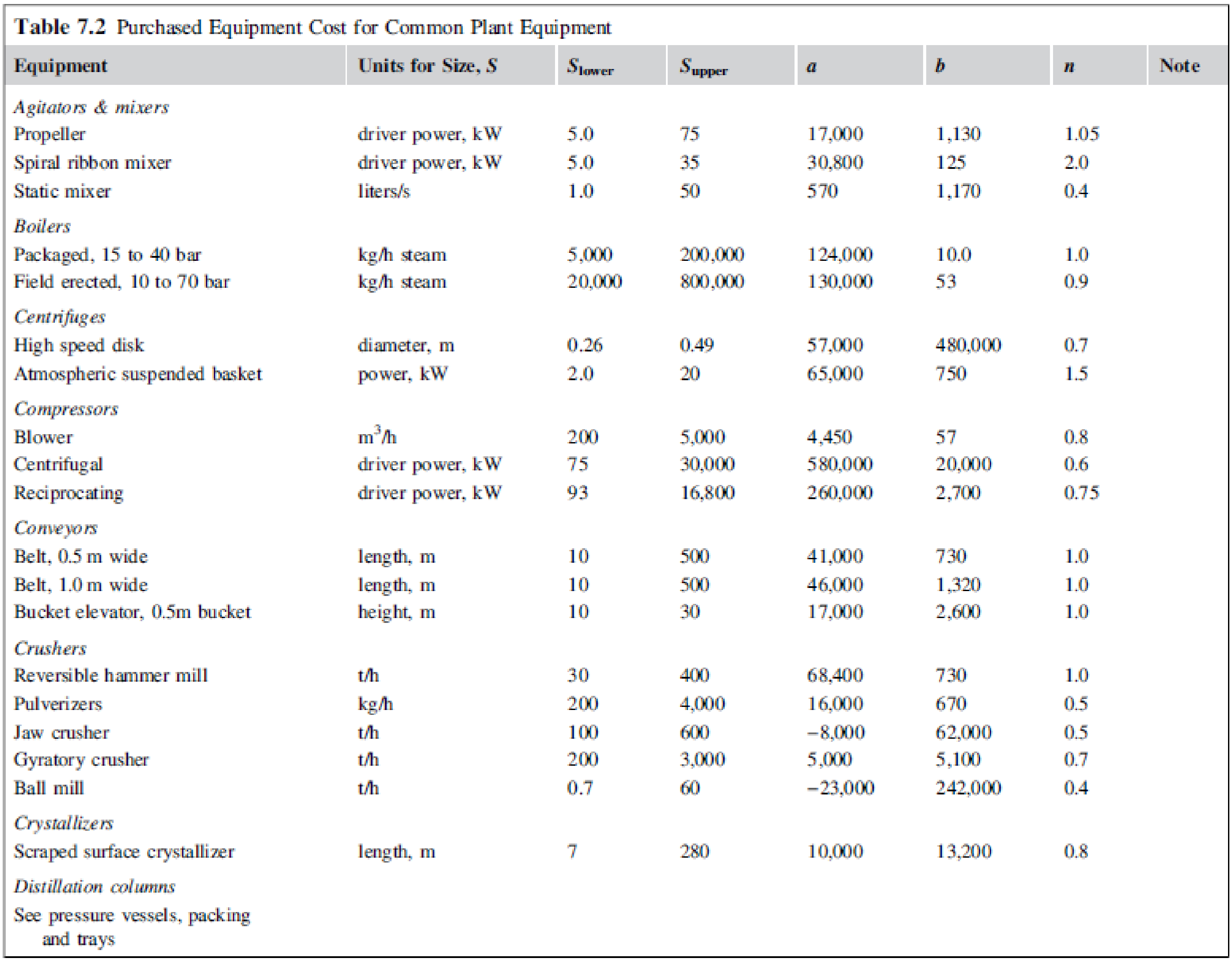


Table 7.2 Purchased Equipment Cost for Common Plant Equipment

Equipment	Units for Size, *S*	S_{lower}	S_{upper}	*a*	*b*	*n*	Note
Agitators & mixers							
Propeller	driver power, kW	5.0	75	17,000	1,130	1.05	
Spiral ribbon mixer	driver power, kW	5.0	35	30,800	125	2.0	
Static mixer	liters/s	1.0	50	570	1,170	0.4	
Boilers							
Packaged, 15 to 40 bar	kg/h steam	5,000	200,000	124,000	10.0	1.0	
Field erected, 10 to 70 bar	kg/h steam	20,000	800,000	130,000	53	0.9	
Centrifuges							
High speed disk	diameter, m	0.26	0.49	57,000	480,000	0.7	
Atmospheric suspended basket	power, kW	2.0	20	65,000	750	1.5	
Compressors							
Blower	m^3/h	200	5,000	4,450	57	0.8	
Centrifugal	driver power, kW	75	30,000	580,000	20,000	0.6	
Reciprocating	driver power, kW	93	16,800	260,000	2,700	0.75	
Conveyors							
Belt, 0.5 m wide	length, m	10	500	41,000	730	1.0	
Belt, 1.0 m wide	length, m	10	500	46,000	1,320	1.0	
Bucket elevator, 0.5m bucket	height, m	10	30	17,000	2,600	1.0	
Crushers							
Reversible hammer mill	t/h	30	400	68,400	730	1.0	
Pulverizers	kg/h	200	4,000	16,000	670	0.5	
Jaw crusher	t/h	100	600	−8,000	62,000	0.5	
Gyratory crusher	t/h	200	3,000	5,000	5,100	0.7	
Ball mill	t/h	0.7	60	−23,000	242,000	0.4	
Crystallizers							
Scraped surface crystallizer	length, m	7	280	10,000	13,200	0.8	
Distillation columns							
See pressure vessels, packing and trays							

Appendix 1 continued

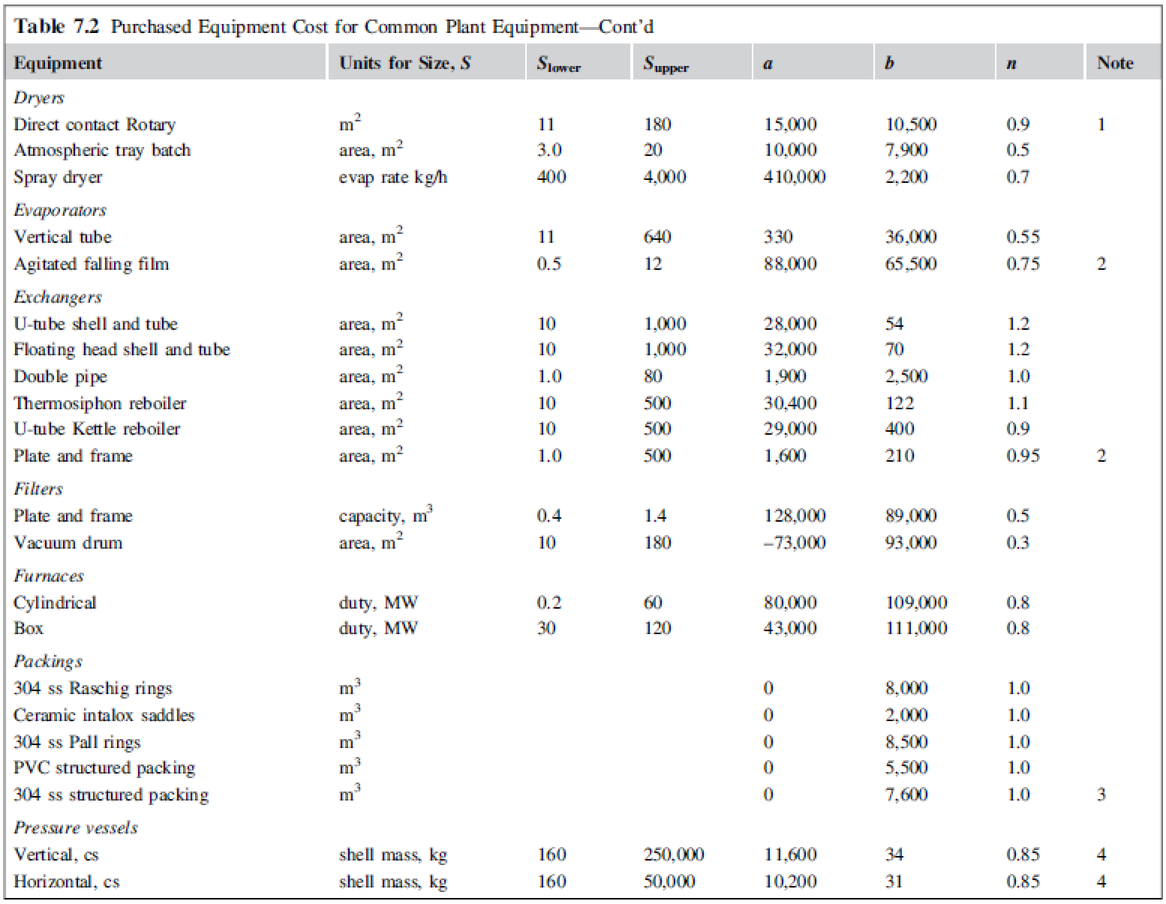


Table 7.2 Purchased Equipment Cost for Common Plant Equipment—Cont'd

Equipment	Units for Size, S	S_{lower}	S_{upper}	a	b	n	Note
Dryers							
Direct contact Rotary	m^2	11	180	15,000	10,500	0.9	1
Atmospheric tray batch	area, m^2	3.0	20	10,000	7,900	0.5	
Spray dryer	evap rate kg/h	400	4,000	410,000	2,200	0.7	
Evaporators							
Vertical tube	area, m^2	11	640	330	36,000	0.55	
Agitated falling film	area, m^2	0.5	12	88,000	65,500	0.75	2
Exchangers							
U-tube shell and tube	area, m^2	10	1,000	28,000	54	1.2	
Floating head shell and tube	area, m^2	10	1,000	32,000	70	1.2	
Double pipe	area, m^2	1.0	80	1,900	2,500	1.0	
Thermosiphon reboiler	area, m^2	10	500	30,400	122	1.1	
U-tube Kettle reboiler	area, m^2	10	500	29,000	400	0.9	
Plate and frame	area, m^2	1.0	500	1,600	210	0.95	2
Filters							
Plate and frame	capacity, m^3	0.4	1.4	128,000	89,000	0.5	
Vacuum drum	area, m^2	10	180	−73,000	93,000	0.3	
Furnaces							
Cylindrical	duty, MW	0.2	60	80,000	109,000	0.8	
Box	duty, MW	30	120	43,000	111,000	0.8	
Packings							
304 ss Raschig rings	m^3			0	8,000	1.0	
Ceramic intalox saddles	m^3			0	2,000	1.0	
304 ss Pall rings	m^3			0	8,500	1.0	
PVC structured packing	m^3			0	5,500	1.0	
304 ss structured packing	m^3			0	7,600	1.0	3
Pressure vessels							
Vertical, cs	shell mass, kg	160	250,000	11,600	34	0.85	4
Horizontal, cs	shell mass, kg	160	50,000	10,200	31	0.85	4

(Continued)

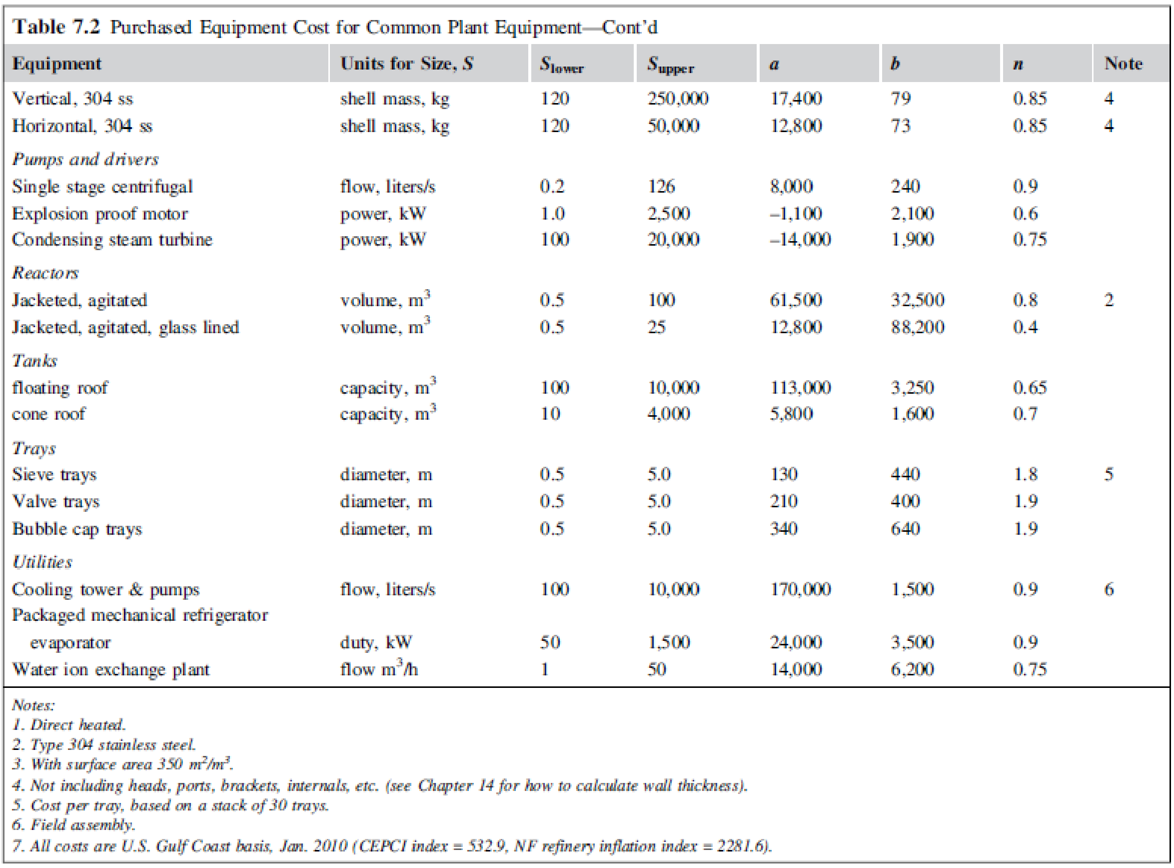


Table 7.2 Purchased Equipment Cost for Common Plant Equipment—Cont'd

Equipment	Units for Size, S	S_{lower}	S_{upper}	a	b	n	Note
Vertical, 304 ss	shell mass, kg	120	250,000	17,400	79	0.85	4
Horizontal, 304 ss	shell mass, kg	120	50,000	12,800	73	0.85	4
Pumps and drivers							
Single stage centrifugal	flow, liters/s	0.2	126	8,000	240	0.9	
Explosion proof motor	power, kW	1.0	2,500	−1,100	2,100	0.6	
Condensing steam turbine	power, kW	100	20,000	−14,000	1,900	0.75	
Reactors							
Jacketed, agitated	volume, m^3	0.5	100	61,500	32,500	0.8	2
Jacketed, agitated, glass lined	volume, m^3	0.5	25	12,800	88,200	0.4	
Tanks							
floating roof	capacity, m^3	100	10,000	113,000	3,250	0.65	
cone roof	capacity, m^3	10	4,000	5,800	1,600	0.7	
Trays							
Sieve trays	diameter, m	0.5	5.0	130	440	1.8	5
Valve trays	diameter, m	0.5	5.0	210	400	1.9	
Bubble cap trays	diameter, m	0.5	5.0	340	640	1.9	
Utilities							
Cooling tower & pumps	flow, liters/s	100	10,000	170,000	1,500	0.9	6
Packaged mechanical refrigerator evaporator	duty, kW	50	1,500	24,000	3,500	0.9	
Water ion exchange plant	flow m^3/h	1	50	14,000	6,200	0.75	

Notes:
1. Direct heated.
2. Type 304 stainless steel.
3. With surface area 350 m^2/m^3.
4. Not including heads, ports, brackets, internals, etc. (see Chapter 14 for how to calculate wall thickness).
5. Cost per tray, based on a stack of 30 trays.
6. Field assembly.
7. All costs are U.S. Gulf Coast basis, Jan. 2010 (CEPCI index = 532.9, NF refinery inflation index = 2281.6).

Appendix 2. Estimation of installation cost for equipment (Peters, Timmerhaus, and West, 2003)

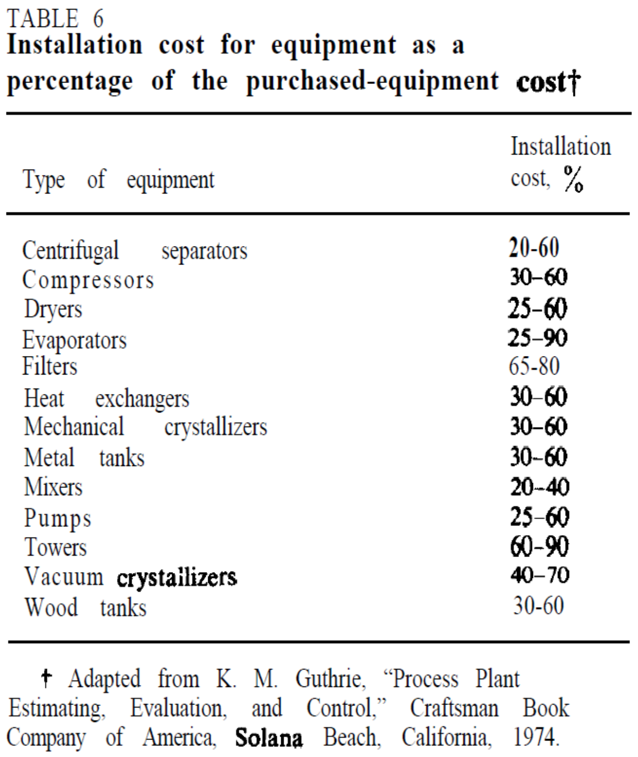


TABLE 6

Installation cost for equipment as a percentage of the purchased-equipment cost†

Type of equipment	Installation cost, %
Centrifugal separators	20-60
Compressors	30-60
Dryers	25-60
Evaporators	25-90
Filters	65-80
Heat exchangers	30-60
Mechanical crystallizers	30-60
Metal tanks	30-60
Mixers	20-40
Pumps	25-60
Towers	60-90
Vacuum crystallizers	40-70
Wood tanks	30-60

† Adapted from K. M. Guthrie, "Process Plant Estimating, Evaluation, and Control," Craftsman Book Company of America, Solana Beach, California, 1974.

Appendix 3: Small scale equipment data sheet

HEAT EXCHANGER –SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : E101 **In series : 1**
NAME : HEAT EXCHANGER **In parallel : None**

General Data

Service : Heat exchanger
Type : Shell and Tube
Position : Horizontal

Capacity	**[kW]**	**:** 5855.74
Heat Exchange Area	**$[m^2]$**	: 230
Overall Heat Transfer Coefficient[W/m2·°C]		: 850
Log. Mean Temperature Diff. (LMTD) [°C]		: 29.95
Passes Tube Side		: 1
Passes Shell Side		: 1
Number of shells in series		: 1
Tube outside diameter	**[m]**	: 0.025
Tube extended length	**[m]**	: 6
Tube pitch	**[m]**	: 0.03
Correction Factor LMTD (min. 0.75)	**[°C]**	: 1
LMTD corrected		: 29.87
Minimum temperature approach	**[˚C]**	: 5

Process conditions

			Shell side	**Tube side**
Medium		**:**	Steam	Mixture of water and sawdust
Mass Stream	**[kg/h]**	**:**	85758.6	98437.5
Temperature IN	**[˚C]**	**:**	109.52	46.8
Temperature OUT	**[˚C]**	**:**	99.64	94.64
Pressure	**[bar]**	**:**	1	7
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 3: Small scale equipment data sheet

HEAT EXCHANGER –SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : E102 **In series :1**
NAME : HEAT EXCHANGER **In parallel : None**

General Data

Service : Heat exchanger
Type : Floating head
Position : Horizontal

Capacity **[kW]** **: 2806.8**
Heat Exchange Area **$[m^2]$** **: 122.19**
Overall Heat Transfer Coefficient **[W/m2·ºC]** **: 850**
Log. Mean Temperature Diff. (LMTD) **[ºC]** **: 27.02**
Passes Tube Side **: 1**
Passes Shell Side **: 1**
Number of shells in series **: 1**
Tube outside diameter **[m]** **:0.025**
Tube extended length **[m]** **:6**
Tube pitch **[m]** **:0.03**
Correction Factor LMTD (min. 0.75) **[ºC]** **: 1**
LMTD corrected **: 27.02**
Minimum temperature approach **[˚C]** **: 5**

Process conditions

			Shell side	Tube side
Medium		:	Cold air	Steam
Mass Stream	**[kg/h]**	:	134936	8183.24
Temperature	**IN [˚C]**	:	20	99.68
Temperature OUT	**[˚C]**	:	94.65	99.64
Pressure	**[bar]**	:	1	1
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 3: Small scale equipment data sheet

HEAT EXCHANGER –SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : E103	**In series : 1**
NAME : HEAT EXCHANGER	**In parallel : None**

General Data

Service : Heat exchanger
Type : Floating head
Position : Horizontal

Capacity	[kW]	: 96.71
Heat Exchange Area	[m^2]	: 2.8
Overall Heat Transfer Coefficient	[W/m2·°C]	: 850
Log. Mean Temperature Diff. (LMTD)	[°C]	: 40.4
Passes Tube Side : 1		
Passes Shell Side : 1		
Number of shells in series : 1		
Tube outside diameter	[m]	: 0.025
Tube extended length	[m]	: 6
Tube pitch	[m]	: 0.03
Correction Factor LMTD (min. 0.75)	[°C]	: 1
LMTD corrected		: 40.2
Minimum temperature approach	[˚C]	: 5

Process conditions

			Shell side	**Tube side**
Medium		**:**	Cold water	Product
Mass Stream	**[kg/h]**	**:**	6500	1556
Temperature IN	**[˚C]**	**:**	20	109.5
Temperature OUT	**[˚C]**	**:**	32.98	37.98
Pressure	**[bar]**	**:**	1	1
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 3: Small scale equipment data sheet

REACTOR- SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : R100		**In series : 1**
NAME	**: Reactor**	**In Parallel : none**
General Data		
Service:		: Reaction
Type		: RCSTR
Position		: Vertical
Tank Design		
Design gauge pressure	[bar]	:7.7
Reactor volume	[m^3]	: 42.47
Liquid Phase volume	[m^3]	: 38.77
Liquid height	[m]	: 3.66
Tank diameter	[m]	: 3.66
Tank height	[m]	: 4
Tank material		: Stainless steel
Impeller design		
Impeller diameter	[m]	: 1.22
Impeller type		: Pitched blade turbine
Number of Impellers		: 1
Process condition		
Temperature	[˚C]	: 160
Pressure	[bar]	: 7
Residence time	[min]	: 20
Mass flow	[kg/h]	: 112500

Appendix 3: Small scale equipment data sheet

MIXING TANK- SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : M100 **In series : 1**

NAME : Mixing tank **In Parallel : none**

General Data

Service: : Mixing of water and sawdust

Type : Mixing tank standard installation

Position : Vertical

Tank Design

Tank volume [m^3] : 42.47

Liquid Phase volume [m^3] : 38.77

Tank diameter [m] : 3.66

Tank height [m] : 4

Liquid height [m] : 3.66

Baffle width [m] : 0.305

Impeller distance from bottom [m] : 1.22

Tank material : SS(1)

Impeller design

Impeller diameter [m] : 1.22

Impeller type : Pitched blade turbine

Number of Impellers : 1

Mixing time [s] : 63.95

Impeller flowrate [m3/s] : 2.31

Impeller speed [rps] : 1.7

Mixing power [kW] : 15.5

Remarks :

(1). SS = Stainless Steel

Appendix 3: Small scale equipment data sheet

FLASH VESSEL – SUMMARY (SMALL SCALE)

EQUIPMENT NR: **NAME :**	**F101** **Flash vessel for water separator**	**F102** **Flash vessel for water separator**
Position	Horizontal	Horizontal
Inlet Pressure [bar] **Outlet Pressure**	7 1	1 1
Inlet Temperature [˚C] **Outlet temperature [˚C]** **Net duty [kW]** **Vapor Fraction**	160 99.68 0 0.08	99.68 109.5 54286.5 0.995
Liquid Volume [m^3] **Diameter [m]** **L or H [m]**	14.81 1.82 5.63	32.29 3.35 3.65
Number **-Series** **Parallel**	 1 -	 1 -
Material of construction (1)	SS	SS
Remarks: (1) SS = Stainless Steel		

Appendix 3: Small scale equipment data sheet

CENTRIFUGAL PUMP – SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : P101 **In series : 1**
NAME : Pump **In parallel : None**

Service : Pressure increase
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid : Mixture of saw dust and water
Temperature [˚C] : 47
Density [kg/m3] : 924
Viscosity [N·s·m−2] : 5×10^{-04}
Volumetric flowrate [m3/h] : 101.18

Power

Capacity [m3/h] : 0.74
Suction pressure [bar] : 1
Discharge pressure [bar] : 7
Fluid power [kW] : 16.86
Brake power [kW] : 24
Theoretical power [kW] : 24
Pump efficiency [%] : 70.02

Construction details	Construction materials(1)
NPSH available [m] : 9.39	**Pump house** :MS
Head developed [m] : 62.88	**Pump rotor** : HT Steel
Design gauge pressure [bar] : 7.7	**Shaft** : HT Steel

Remarks:
(1) : MS = Mild Steel, HT Steel = High Tensile Steel

Appendix 3: Small scale equipment data sheet

FILTER – SPECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : S100 **In series : 1**

Name : Rotary vacuum filter **In parallel : None**

Service : Sawdust-liquid separation

Number : 1

Operating condition & Physical Data

Solid **:** Saw dust

Liquid **:** Extracted liquor

Solid mass flowrate [kg/h] : 12751.6

Average solid density [kg/m3] : 1150.31

Particle size [mm] **:** 0.5-2

Filtration pressure drop [bar] **:** 0.5

Deliquoring pressure drop [bar] : 0.5

Cake thickness [m] : 0.015

Porosity of the cake : 0.45

Mass fraction of solid in filter cake : 0.75

Temperature [˚C] : 99.68

Maximum cake thickness [m] : 1.0088

Design parameters

Filtration angel [deg] : 120,006

Filter area [m^2] : 58

Filter plate diameter [m] : 2.7

Motor power [kW] : 2.2

Drum speed [rpm] **:** 1

Construction material **:** CS (1)

Remarks:

(1) CS = Carbon Steel

Appendix 3: Small scale equipment data sheet

DRYER- SHECIFICATION SHEET (SMALL SCALE)

EQUIPMENT NUMBER : D100 **In series : 1**

Name : Vacuum tray dryer **In parallel : None**

Service : Sawdust drying

Number : 1

Dryer type : Convective

Operating condition & Physical Data

Solid : Saw dust

Drying medium : Hot air

Particle size [mm] : 0.5-2

Heat Duty [kW] : -34.77

Solids residence time [minute] : 20

Calculated heat transfer co-efficient [W/m^2.K] : 2.4

Overall evaporation rate [kg/h] : 2899,79

Solid velocity [m/s] : 0.0015

	Sawdust	**Air**
Inlet flow rate [kg/h]	17002	134936
Inlet temperature [˚C]	99.68	94.65
Outlet temperature [˚C]	61	52
Inlet moisture content [%]	32.85	-
Outlet moisture content [%]	10.11	-

Design parameters

Tray area [m^2] : 3.71

Bed porosity : 0.4

Construction material : CS (1)

Remarks:

(1) CS = Carbon Steel

Appendix 4: Equipment data sheet (Large scale)

HEAT EXCHANGER –SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER :	**E101**	**In series**	**: 1**
NAME	**: HEAT EXCHANGER**	**In parallel**	**: 2**

General Data

Service : Heat exchanger
Type : Shell and Tube
Position : Horizontal

Capacity	**[kW]**	: 28419.8
Heat Exchange Area	**[m^2]**	: 297
Overall Heat Transfer Coefficient	**[W/m2·ºC]**	: 850
Log. Mean Temperature Diff. (LMTD)	**[ºC]**	: 37.58
Passes Tube Side		: 1
Passes Shell Side		: 1
Number of shells in series		: 1
Tube outside diameter	**[m]**	: 0.025
Tube extended length	**[m]**	: 6
Tube pitch	**[m]**	: 0.03
Correction Factor LMTD (min. 0.75)	**[ºC]**	: 1
LMTD corrected		: 37.58
Minimum temperature approach	**[˚C]**	: 15

Process conditions

			Shell side	Tube side
Medium		:	Steam	Mixture of water and sawdust
Mass Stream	**[kg/h]**	:	514552	675000
Temperature IN	**[˚C]**	:	109.52	45.61
Temperature OUT	**[˚C]**	:	99.64	84.65
Pressure	**[bar]**	:	1	7
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 4: Equipment data sheet (Large scale) continued

HEAT EXCHANGER –SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : E102 **In series : 1**
NAME : HEAT EXCHANGER **In parallel : 2**

General Data

Service : Heat exchanger
Type : Floating head
Position : Horizontal

Capacity [kW] : 17648.9
Heat Exchange Area [m^2] : 206
Overall Heat Transfer Coefficient [W/m2·^{o}C] : 850
Log. Mean Temperature Diff. (LMTD) [^{o}C] : 33.616
Passes Tube Side : 1
Passes Shell Side : 1
Number of shells in series : 1
Tube outside diameter [m] :0.025
Tube extended length [m] :6
Tube pitch [m] :0.03
Correction Factor LMTD (min. 0.75) [^{o}C] : 1
LMTD corrected : 33.616
Minimum temperature approach [˚C] : 10

Process conditions

			Shell side	Tube side
Medium		:	Cold air	Steam
Mass Stream	**[kg/h]**	:	909627	49099.5
Temperature	**IN [˚C]**	:	20	99.68
Temperature OUT	**[˚C]**	:	89.64	99.64
Pressure	**[bar]**	:	1	1
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 4: Equipment data sheet (Large scale) continued

HEAT EXCHANGER –SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : E103 **In series : 1**
NAME : HEAT EXCHANGER **In parallel : None**

General Data

Service : Heat exchanger
Type : Floating head
Position : Horizontal

Capacity [kW] : 582,47
Heat Exchange Area [m^2] : 17
Overall Heat Transfer Coefficient [W/m2·°C] : 850
Log. Mean Temperature Diff. (LMTD) [°C] : 40.27
Passes Tube Side : 1
Passes Shell Side : 1
Number of shells in series : 1
Tube outside diameter [m] : 0.025
Tube extended length [m] : 6
Tube pitch [m] : 0.03

Correction Factor LMTD (min. 0.75) [°C] : 1
LMTD corrected : 40.2
Minimum temperature approach [˚C] : 5

Process conditions

			Shell side	**Tube side**
Medium		:	Cold water	Product
Mass Stream	**[kg/h]**	:	40000	9336
Temperature IN	**[˚C]**	:	20	109.5
Temperature OUT	**[˚C]**	:	32.70	37.70
Pressure	**[bar]**	:	1	1
Material			CS(1)	SS(1)

Remarks:

1. SS = Stainless Steel; CS = Carbon Steel

Appendix 4: Equipment data sheet (Large scale) continued

REACTOR- SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : R100 **In series : 1**
NAME : Reactor **In Parallel : 2**

General Data

Service: : Reaction

Type : RCSTR

Position : Vertical

Tank Design

Design gauge pressure [bar] :7.7
Reactor volume [m^3] : 85
Liquid Phase volume [m^3] : 77.7
Liquid height [m] : 4.63
Tank diameter [m] : 4.63
Tank height [m] : 5
Tank material : Stainless steel

Impeller design

Impeller diameter [m] : 1.54
Impeller type : Pitched blade turbine
Number of Impellers : 1

Process condition

Temperature [˚C] : 160
Pressure [bar] : 7
Residence time [min] : 20
Total Mass flow [kg/h] : 675000

Appendix 4: Equipment data sheet (Large scale) continued

MIXING TANK- SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : M100 **In series : 1**

NAME : Mixing tank **In Parallel : none**

General Data

Service: : Mixing of water and sawdust

Type : Mixing tank standard installation

Position : Vertical

Tank Design

Tank volume [m^3] : 255

Liquid Phase volume [m^3] : 232.67

Tank diameter [m] : 6.67

Tank height [m] : 7.29

Liquid height [m] : 6.67

Baffle width [m] : 0.56

Impeller distance from bottom [m] : 2.22

Tank material : SS(1)

Impeller design

Impeller diameter [m] : 2.22

Impeller type : Pitched blade turbine

Number of Impellers : 1

Mixing time [s] : 19.25

Impeller flowrate [m3/s] : 7.68

Impeller speed [rps] : 0.93

Mixing power [kW] : 51.67

Remarks :

(1). SS = Stainless Steel

Appendix 4: Equipment data sheet (Large scale) continued

FLASH VESSEL – SUMMARY (LARGE SCALE)

EQUIPMENT NR: **NAME :**	**F101** **Flash vessel for water separator**	**F102** **Flash vessel for water separator**
Position	Horizontal	Horizontal
Inlet Pressure [bar] **Outlet Pressure**	7 1	1 1
Inlet Temperature [˚C] **Outlet temperature [˚C]** **Net duty [kW]** **Vapor Fraction**	160 99.68 0 0.08	99.68 109.5 325718 0.995
Liquid Volume [m³] **Diameter [m]** **L or H [m]**	66.72 3.048 9.15	187.42 8.07 3.65
Number **-Series** **Parallel**	 1 -	 1 -
Material of construction (1)	SS	SS
Remarks: (2) SS = Stainless Steel		

Appendix 4: Equipment data sheet (Large scale) continued

CENTRIFUGAL PUMP – SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : P101 **In series : 1**
NAME : Pump **In parallel : None**

Service : Pressure increase
Type : Centrifugal
Number : 1

Operating Conditions & Physical Data

Pumped liquid : Mixture of saw dust and water
Temperature [˚C] : 46
Density [kg/m3] : 924
Viscosity [N·s·m−2] : 5×10^{-04}
Volumetric flowrate [m3/h] : 606.38

Power

Suction pressure [bar] : 1
Discharge pressure [bar] : 7
Fluid power [kW] : 101.06
Brake power [kW] : 121.45
Theoretical power [kW] : 121.45
Pump efficiency [%] : 83.21

Construction details	Construction materials(1)
NPSH available [m] : 9.44	**Pump house** :MS
Head developed [m] : 62.81	**Pump rotor** : HT Steel
Design gauge pressure [bar] : 7.7	**Shaft** : HT Steel

Remarks:
(1) : MS = Mild Steel, HT Steel = High Tensile Steel

Appendix 4: Equipment data sheet (Large scale) continued

FILTER – SPECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER : S100 **In series : 1**

Name : Rotary vacuum filter **In parallel : 2**

Service : Sawdust-liquid separation

Number : 1

Operating condition & Physical Data

Solid **:** Saw dust

Liquid **:** Extracted liquor

Solid mass flowrate [kg/h] : 76509.5

Average solid density [kg/m3] : 1150.31

Particle size [mm] **:** 0.5-2

Filtration pressure drop [bar] **:** 0.5

Deliquoring pressure drop [bar] : 0.5

Porosity of the cake : 0.45

Mass fraction of solid in filter cake : 0.75

Temperature [˚C] : 99.68

Maximum cake thickness [m] : 1.0088

Volume flow of filtrate [m^3/h] : 565.88

Design parameters

Filtration angel [deg] : 120,006

Filter area [m^2] : 97

Filter plate diameter [m] : 2.7

Motor power [kW] : 4

Drum speed [rpm] **:** 1

Construction material **:** CS (1)

Remarks:

(1) CS = Carbon Steel

Appendix 4: Equipment data sheet (Large scale) continued

DRYER- SHECIFICATION SHEET (LARGE SCALE)

EQUIPMENT NUMBER	**: D100**	**In series**	**: 1**
Name	**: Vacuum tray dryer**	**In parallel**	**: None**
Service	: Sawdust drying		
Number	: 1		
Dryer type	: Convective		

Operating condition & Physical Data

Solid		: Saw dust
Drying medium		: Hot air
Particle size	**[mm]**	: 0.5-2
Heat Duty	**[kW]**	: 11.20
Solids residence time	**[minute]**	: 20
Calculated heat transfer co-efficient	**[W/m^2.K]**	: 2.7
Overall evaporation rate	**[kg/h]**	: 17487.1
Solid velocity	**[m/s]**	: 0.0041

		Sawdust	**Air**
Inlet flow rate	**[kg/h]**	102013	909627
Inlet temperature	**[˚C]**	99.68	89.64
Outlet temperature	**[˚C]**	77.59	50.63
Inlet moisture content	**[%]**	32.85	-
Outlet moisture content	**[%]**	10	-

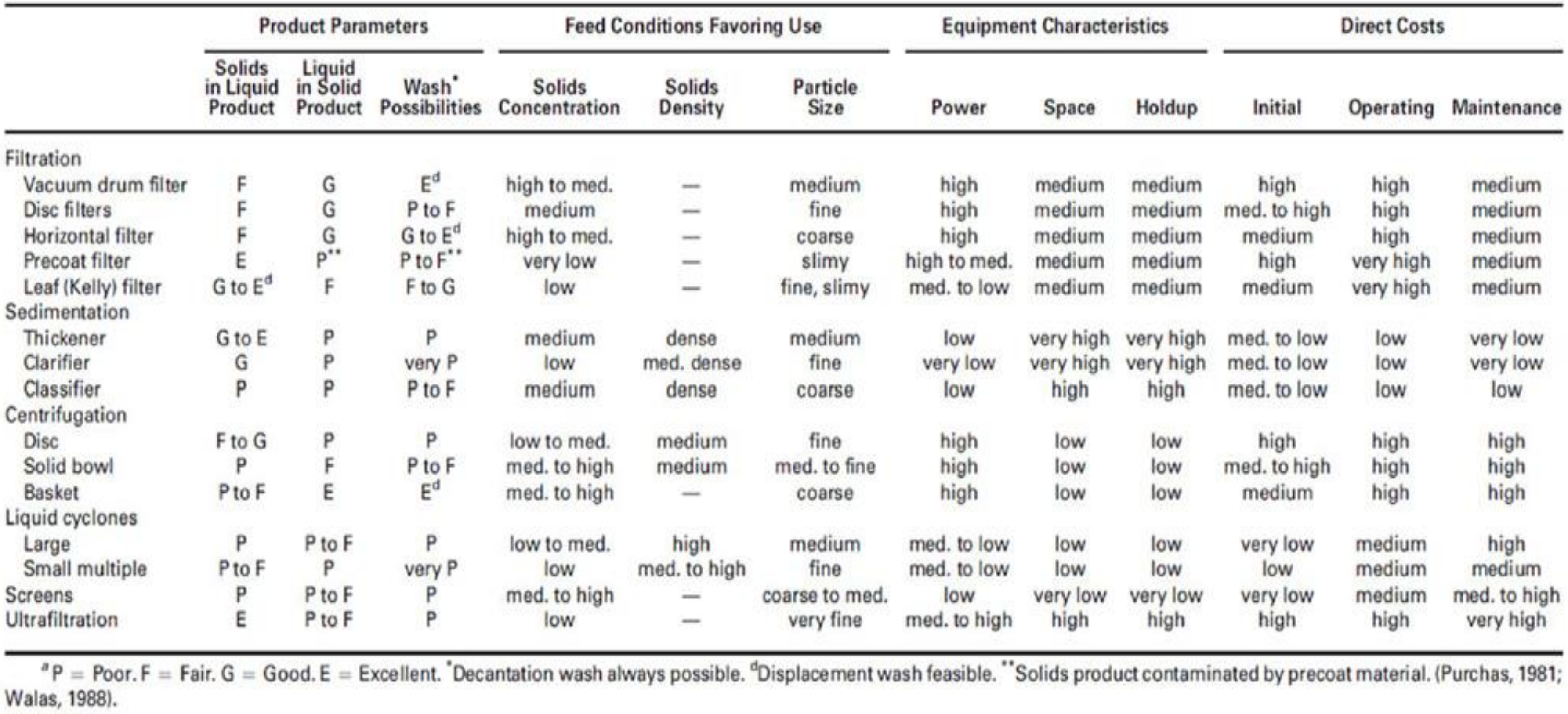
Design parameters

Tray area	[m^2]	: 18.58
Bed porosity		: 0.4
Construction material		: CS (1)

Remasks:

(1) CS = Carbon Steel

Appendix 5: Performance comparison for solid-liquid separation equipment (Couper et al., 2010)



	Product Parameters			Feed Conditions Favoring Use			Equipment Characteristics			Direct Costs		
	Solids in Liquid Product	Liquid in Solid Product	Wash* Possibilities	Solids Concentration	Solids Density	Particle Size	Power	Space	Holdup	Initial	Operating	Maintenance
Filtration												
Vacuum drum filter	F	G	E^d	high to med.	—	medium	high	medium	medium	high	high	medium
Disc filters	F	G	P to F	medium	—	fine	high	medium	medium	med. to high	high	medium
Horizontal filter	F	G	G to E^d	high to med.	—	coarse	high	medium	medium	medium	high	medium
Precoat filter	E	P^{**}	P to F^{**}	very low	—	slimy	high to med.	medium	medium	high	very high	medium
Leaf (Kelly) filter	G to E^d	F	F to G	low	—	fine, slimy	med. to low	medium	medium	medium	very high	medium
Sedimentation												
Thickener	G to E	P	P	medium	dense	medium	low	very high	very high	med. to low	low	very low
Clarifier	G	P	very P	low	med. dense	fine	very low	very high	very high	med. to low	low	very low
Classifier	P	P	P to F	medium	dense	coarse	low	high	high	med. to low	low	low
Centrifugation												
Disc	F to G	P	P	low to med.	medium	fine	high	low	low	high	high	high
Solid bowl	P	F	P to F	med. to high	medium	med. to fine	high	low	low	med. to high	high	high
Basket	P to F	E	E^d	med. to high	—	coarse	high	low	low	medium	high	high
Liquid cyclones												
Large	P	P to F	P	low to med.	high	medium	med. to low	low	low	very low	medium	high
Small multiple	P to F	P	very P	low	med. to high	fine	med. to low	low	low	low	medium	medium
Screens	P	P to F	P	med. to high	—	coarse to med.	low	very low	very low	very low	medium	med. to high
Ultrafiltration	E	P to F	P	low	—	very fine	med. to high	high	high	high	high	very high

^{a}P = Poor. F = Fair. G = Good. E = Excellent. *Decantation wash always possible. ^{d}Displacement wash feasible. **Solids product contaminated by precoat material. (Purchas, 1981; Walas, 1988).

Appendix 6: Equipment cost estimation

The purchased equipment costs for most the equipment used in this process design are based on the following equation.

$$C_e = a + bS^n$$

Where Ce = purchased equipment cost

a, b = cost constant

S = size parameter

n = exponent for that type of equipment (Towler and Sinnott, 2007, Table 7.2)

Estimation of purchased equipment cost

Equipment	Units for size, S	a1*	a2**	b1	b2	S1	S2	n1	n2	Small Scale cost, $	Large scale cost, $	Small scale cost k€	Large scale cost k€
R100	m3	61500	61500	32500	32500	42,47	85	0,8	0,8	713653	1197605	678	1138
M100 Propeller	m3 kW	113000 17000	113000 17000	3250 1130	3250 1130	42,47 15.5	255 51,67	0,65 1,05	0,65 1,05	150165 37088	232161 88118	143 35	221 84
E-101	area, m2	28000	28000	54	54	230	850	1,2	1,2	64853	204887	62	195
E-102	area, m2	28000	28000	54	54	122,19	617,63	1,2	1,2	45252	148578	43	141
E-103	area, m2	28000	28000	54	54	2,8	17	1,2	1,2	28186	29618	27	28
P-100	kW	8000	8000	240	240	24	121,45	0,9	0,9	12192	26037	12	25
S100	area, m2	-73000	-73000	93000	93000	58	97	0,3	0,3	241421	293872	229	279
D-100	area, m2	10000	10000	7900	7900	3,71	18,58	0,5	0,5	25216	44053	24	42

***1 = Small scale, ** 2 = Large scale**

The purchased cost for Flash F101, F102 and blower B100 was calculated from the cost curve**.** The summary for the cost estimation is given below.

Appendix 8: Equipment cost estimation continued

Estimation of purchased equipment cost

Equipment	**Small scale**	**Large scale**

	Capacity	$	Capacity	$
F101	Gallons	30000	Gallons	50000
F102	Gallons	30000	Gallons	50000
B100	459514 Gallons/minute	100000	4 × 774416 Gallons/minute	4 × 100000

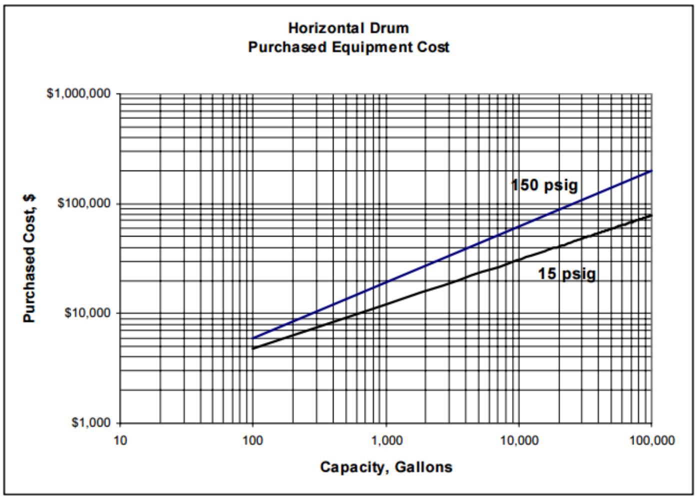


Figure: Cost curve for horizontal drum (LOH, Lyons, and White, 1998).

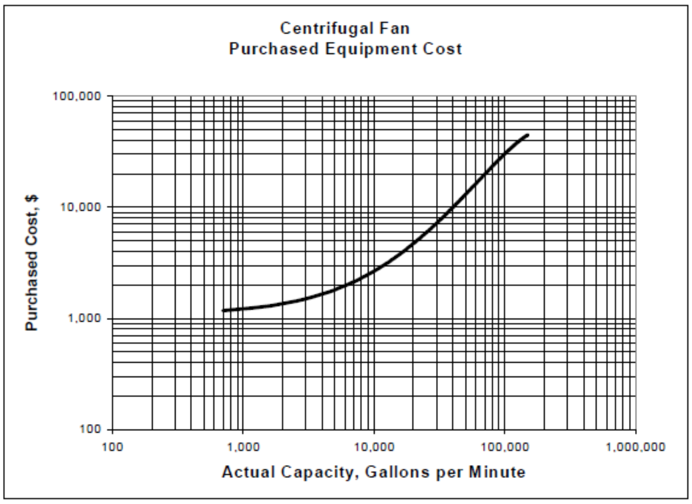


Figure: Cost curve for centrifugal fan (LOH, Lyons, and White, 1998).

Table: Small scale equipment cost indexing

Purchased equipment	Price (€)	Year	Index year	Index 2016	Price (€)	Price (k€)
1. Reactor-R100(RCSTR)	663697	2010	533	545	678643	679
2. Mixing Tank-M100	174145	2010	533	545	178067	178
3. Centrifugal fan-B100	93000	1998	390	545	130104	130
4. Heat exchanger E-101	60313	2010	533	545	61671	62
5. Heat exchanger E-102	42084	2010	533	545	43032	43
6. Heat exchanger E-103	26213	2010	533	545	26803	27
7. Pump P-100	11339	2010	533	545	11594	12
8. Vacuum drum filter-S100	224522	2010	533	545	229577	230
9. Flash drum -F101	27900	1998	390	545	39031	39
10. Flash drum F-102	27900	1998	390	545	39031	39
11. Rotary drum dryer D-100	23451	2010	533	545	23979	24
						1462

Table: Large scale equipment cost indexing

Purchased equipment	Price (€)	Year	Index year	Index 2016	Price (€)	Price (k€)
1. Reactor-R100(RCSTR)	3341318	2010	533	545	3416559	3417
2. Mixing Tank-M100	297859	2010	533	545	304567	305
3. Centrifugal fan-B100	372000	1998	390	545	520418	520
4. Heat exchanger E-101	190545	2010	533	545	194836	195
5. Heat exchanger E-102	138178	2010	533	545	141289	141
6. Heat exchanger E-103	27545	2010	533	545	28165	28
7. Pump P-100	24214	2010	533	545	24760	25
8. Vacuum drum filter-S100	819903	2010	533	545	838366	838
9. Flash drum -F101	46500	1998	390	545	65052	65
10. Flash drum F-102	46500	1998	390	545	65052	65
11. Rotary drum dryer D-100	40969	2010	533	545	41892	42
						5641

Appendix 7: Installation cost estimation

Installation costs for equipment are calculated based on Appendix 2. Details of the installation cost are summarized below.

Table: Installation cost estimation

Equipment	Percentage %	Small scale k€	Large scale k€
R100	30	204	1025
M100	25	45	76
B100	25	33	130
E101	30	19	58
E102	30	13	42
E103	30	8	8
P100	25	3	6
S100	65	149	545
F101	30	12	20
F102	30	12	20
D100	25	6	10
Total		**502**	**1941**

Appendix 8: Estimation of fixed capital investment

ISBL cost:

$$\mathrm{ISBL} = \text{purchase of equipment } + \text{installation of equipment} + \text{piping cost}$$

OSBL cost:

$$\mathrm{OSBL} = 0{,}3 \cdot \mathrm{ISBL}$$

Engineering and supervision cost:

$$\text{Engineering and supervision} = 0{,}25 \cdot \mathrm{ISBL}$$

Contingency cost:

$$\text{Contingency} = 0{,}1 \cdot \mathrm{ISBL}$$

Total fixed capital investment:

$$\mathrm{TFCI} = \mathrm{ISBL} + \mathrm{OSBL} + \text{Engineering and supervision} + \text{Contingency}$$

Start-up cost:

$$\text{Start up} = 0{,}1 \cdot \text{TFCI}$$

Total capital investment:

$$\text{Total Capital Investment} = \text{TFCI} + \text{Start up}$$

(Towler and Sinnott, 2007)

Appendix 9: Profitability analysis

1. Annual gross profit:

$$\text{Annual gross profit} = \text{Revenue} - \text{Total cost of production}$$

2. Return on investment:

$$\%\ \text{ROI} = \frac{\text{Annual profit}}{\text{Total capital investment}} \cdot 100$$

3. Payout time:

$$\text{Payout time (yr)} = \frac{\text{Total capital investment}}{\text{Annual gross profit} + \text{depreciation}}$$

4. Net Present Value (NPV):

$$\text{NPV} = \sum_{n=1}^{n=t} \frac{\text{CFn}}{(1+\text{i})\text{n}}$$

Where CF_n = Cash flow in year n

t= project life in years

i= interest rate

Appendix 10: Fixed manufacturing costs

The fixed operation costs are mainly calculated as percentages of different factors and the cost of labor. Manufacturing costs are calculated as the addition of variable and fixed manufacturing costs.

Table: Estimation of fixed manufacturing cost (Towler and Sinnott, 2012)

	1 operators/shift position	
Labor cost	3	64800 €/year each
Supervisions	5 % of operating labor	
Side costs and benefits	1,8 % of operating labor and supervisions	
Maintenance	3 % of ISBL investment	
Insurance	1 % of ISBL and OSBL investment	
Quality control	0,5 % of sales revenues	
Interest	2 % of total capital investment	
License fees	2 % of total capital investment	
Depreciation	5 % of fixed capital investment	
Administrative costs	5 % of labor, supervision, and maintenance	
Selling and marketing costs	2 % of total product costs	
Research and development costs	2 % of total product costs	