CHALMERS
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Soda Based Production of Dissolving Pulp from Wheat Straw

Process Simulation and Techno-Economic Assessment

Master’s thesis in Innovative and Sustainable Chemical Engineering

LINNEA NILSSON

DEPARTMENT OF CHEMISTRY AND CHEMICAL ENGINEERING

CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2023
www.chalmers.se

Master's thesis 2023

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Supervisors: Maria Hernández Leal, Benjamin Storm & Tomas Rydberg, IVL
Examiner: Diana Bernin, Department of Chemistry and Chemical Engineering, Chalmers

Master's Thesis 2023
Department of Chemistry and Chemical Engineering
Chalmers University of Technology
SE-412 96 Gothenburg

Typeset in LaTeX
Gothenburg, Sweden 2023

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Abstract

Due to the large environmental impact of today's textile industry, both on a global and local scale, change to the industry is needed. One way to reduce the impact of the textile industry is to replace the fibres used in textile production. One alternative resource for textile production is dissolving pulp which can be produced from lignocellulosic materials. Through a techno-economic assessment this thesis evaluates the production of dissolving pulp from wheat straw through a soda cooking process on an industrial scale. The evaluation is made based on a previously performed lab trial scaled up and simulated in Aspen Plus. In the simulation the pulping, bleaching, drying, evaporation and boiler section are included as well as a heat integration of the process.

Through cost estimations the break-even price of the dissolving pulp was determined for two scenarios, one where the drying of the pulp was included and one where it was assumed that the dissolving pulp would be utilised directly after pulping and thus not requiring drying. The break-even price for the scenario with drying is 7194 SEK/dton dissolving pulp and 5736 SEK/dton dissolving pulp for the scenario without drying. This shows that the process has potential to be competitive on the market if further developed and improved upon.

Keywords: textile fibres, dissolving pulp, soda pulping, wheat straw, process simulation, scale-up

Acknowledgements

The work in this thesis has been carried out as part of the project "CircuCel" (Cirkulär cellulosa till framställning av textilfiber), project number 2021-02093 funded by the Swedish innovation agency VINNOVA. The work is also linked to and co-sponsored by the project "Calimero" (Industry CAse studies anaLysis to IMprove EnviROnmental performance and sustainability of bio-based industrial processes), project number 101060546 funded by The European Union.

Writing this thesis has not been an easy task and it would not have been completed without the help, input and inspiration from everyone involved in the project.

Thanks to Maria Hernández Leal and Benjamin Storm for the helpful insight in the usage of Aspen Plus and consultation about the simulation and economic assessment.

Thanks to Manuel Mammen Parayil for great discussion about assumptions and setup for the simulation.

Thanks to Joanna Wojtasz-Mucha for providing much needed information and explanation on the lab process.

Thanks to Thomas Rydberg for coordinating this thesis with many other interesting projects.

Linnea Nilsson, Gothenburg, May 2023

Contents

List of Acronyms

Below is the list of acronyms that have been used throughout this thesis listed in alphabetical order:

Ara	Arabinose
ASL	Acid Soluble Lignin
CEPCI	Chemical Engineering Plant Cost Index
DF	Dilution Factor
DP	Dissolving Pulp
DR	Displacement Ratio
Gal	Galactose
Glu	Glucose
HEN	Heat Exchanger Network
HTC	Heat Transfer Coefficient
HX	Heat Exchanger
Klason	Klason lignin
Man	Mannose
MMCF	Man Made Cellulosic Fiber
MP	Medium Pressure
l/s	liquid to solid
Xyl	Xylose

1
Introduction

1.1 Background

Today's textile industry is growing in an unsustainable manner. Production of textiles comes with a significant impact on the environment, human health and society. If the textile industry continues to grow and operate in the same way as today it will not be possible to keep global warming below the 1.5°C goal set in the Paris Agreement [1].

The majority of textiles produced is made from synthetic fibers, cotton and man made cellulosic fibers (MMCF), synthetic fibers make up 64% of the produced fibers, cotton stands for 22% of production and MMCF for 6% [1]. Synthetic fibers are mainly made from non-renewable resources with energy intensive processes which primarily is powered by fossil fuels. Cotton on the other hand is a renewable resource but the cultivation uses large amounts of water, fertiliser and pesticides. In the production of textiles the use of different chemicals is great. Chemicals are used to get the desired properties of the textile, such as colour, flame protection and water repellence, these chemicals are often hazardous and can have huge environmental and health impacts [2].

There are also many ethical issues in the textile industry, even though the industry can bring income to societies the working conditions are very poor with long work hours, low wages, child labour [2] and sexual harassment [3]. The safety of the employees is likewise very poor, factories have little to no safety equipment or routines and often operate in unsafe buildings. The use and pollution of local waterways affects the local communities ability to use the water for other necessities such as drinking and bathing [2].

It is clear that the textile industry must change to reach global sustainability goals, both in volume but also with the methods used in production. One alternative is the production of MMCFs from dissolving pulp, a product made from cellulosic materials. The Swedish company TreeToTextile uses wood based dissolving pulp to produce textile in a process that reduces the energy demand by 30%, the water usage by 80% and the use of chemicals by 80% compared to conventional viscose production [4]. This report focuses on the production of dissolving pulp from wheat straw as an alternative in the textile industry. Wheat straw is a byproduct from the agricultural industry and by developing a process for production of dissolving pulp its economic value may increase whilst also lowering the impact of the textile industry.

1.2 Aim

The aim of this master thesis is to develop a process simulation of the production of dissolving pulp from wheat straw through a soda cooking process on an industrial scale. The process will be evaluated through a techno-economic point of view. The projects also aims to investigate the use of heat integration in the process.

1.3 Scope

The process simulation will be based on lab trials performed at Chalmers University of Technology not included in this thesis. The data from the lab trials will be used to scale up the process in the process simulation program Aspen Plus. Included in the process simulation will be the pulping, bleaching, drying, evaporation and boiler section of the process. The full chemical recovery cycle will not be simulated but will be considered in the evaluation of the process. By using Aspen Energy Analyser the potential for heat integration in the process will be investigated.

This thesis will only evaluate the process through a techno-economic point of view. No evaluation of the environmental impact of the process will be performed. Several raw materials can be used for producing dissolving pulp but only the use of wheat straw will be studied. The thesis will also be limited to only studying the production of dissolving pulp and not any further usage or refining of dissolving pulp.

2
Theory

2.1 Lignocellulosic Materials

Plant cell walls consist of cellulose, hemicelluloses, lignin and extractives. The distribution of these components vary between different plants and the different parts of the plant [5].

Cellulose is a polymer made up of glucose connected by 1→4 β-glycosidic bonds. The cellulose polymer is linear, crystalline, unbranched and has a high degree of polymerisation, typically between 10 000 and 15 000. Hemicellulose is a polymer with a lower degree of polymerisation than cellulose, often 150 to 200. The hemicelluloses are composed of different monosaccharides and the polymers are amorphous and branched. The most common monosaccharides in hemicelluloses are xylopyranose, mannopyranose, arabinofuranose and galactopyranose. The different monosaccharides present in lignocellulosic materials can be seen in figure 2.1. The structure and composition of hemicellulose varies between plant species. Lignin is a complex structure consisting of a mixture of aromatic and aliphatic parts. It is a 3-dimensional network connected mainly by ether and carbon-carbon bonds [5][6].

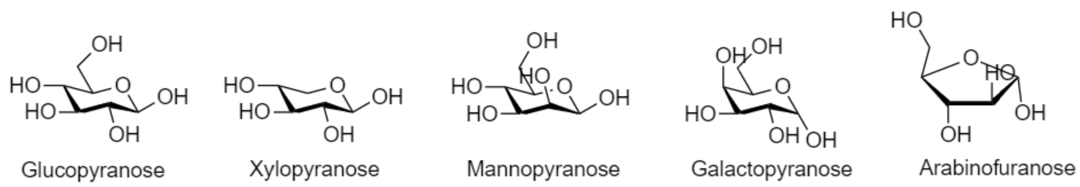

Figure 2.1: Monosaccharides in lignocellulosic materials

In the cell wall the cellulose forms a skeletal matrix that is surrounded and held together by the hemicellulose and lignin. The major part of the cell wall is made up of cellulose, hemicellulose and lignin but there are also other minor components present. Those that are extractable by various neutral solvents are called extractives and they can vary greatly between species. The major categories of extractives are resins and polyphenols [5][6].

2.1.1 Wheat Straw

Wheat straw is a byproduct from wheat farming that is used today as animal feed, fertiliser, fuel and to a small extent as a base material in pulp production [7]. In Europe at least one third of the wheat straw is required to remain on the field in order to ensure that a sufficient amount of nutrients is kept in the field, around 20% of wheat straw is needed for bedding and animal feed. The rest of the wheat straw can be utilised for energy and other production areas. There is potential to increase the use of wheat straw for energy and other production areas as there is more straw available than what is being used [8]. The amount of straw acquired from harvesting wheat varies between reports, some papers claim that the straw/grain ratio is 0.5-0.66 [9] whilst others approximate it to 1.3-1.4 [7]. The difference in the approximated straw/grain ratios could be due to the average straw length varying between different species of wheat that is grown in different areas [9].

2.2 Pulping

Pulping is performed to separate the components of lignocellulosic materials and to liberate the cellulose fibers. There are four categories of pulping: chemical, semi-chemical, chemi-mechanical and mechanical pulping. The more chemicals, both in number and volume, used in a pulping process the more components will be degraded and solubilised thus resulting in a lower yield [10].

The focus on this report is on soda cooking, a chemical pulping process, and will therefore be explained further.

2.2.1 Soda Cooking

In soda cooking, an aqueous solution of sodium hydroxide is used as cooking liquor to degrade the lignin in the biomass thus delignifying the material. The process also breaks down polysaccharides present in the biomass. The cooking can be performed either batchwise or continuously and is typically operated at temperatures between 150 and 170°C. As it is not desired for the cooking liquor to vaporise the cooking takes place in pressurised vessels called digesters. After cooking, pulp is recovered which needs to be washed and bleached to reach desired properties which depend on the produced pulp. Black liquor which consists of spent cooking liquor and dissolved and delignified biomass can be sent through a chemical recovery process which recovers the cooking liquor whilst also providing energy for the pulping plant [10][11].

Soda cooking is considered to be a more environmentally friendly pulping technology compared to standard chemical pulping processes such as kraft pulping or sulphite pulping. This is because soda cooking uses fewer chemicals and does not produce toxic gases such as hydrogen sulphide or sulphur dioxide. The soda cooking also consumes less chemicals making it more suitable for raw materials with a lower lignin content and is in general used for non wood materials [12]. The use of less chemicals also leads to the process being more economical [13].

2.2.2 Washing

Washing of pulp can take place in many different types of equipment all with the purpose of removing black liquor still present in the pulp. This is done by displacing the black liquor with wash liquor. The efficiency of the displacement is dependent on many factors of the pulp such as pulp porosity, fibre size distribution and sorption on the fibre surface. The amount of wash liquor required is dependent on the type of equipment used as well as the number of wash stages but should be minimised to increase the energy recovery potential from the spent liquor. This is because less water in the black liquor exiting the washer results in less steam being needed to evaporate the water in the evaporation section which will be explained further in 2.5.1 [11].

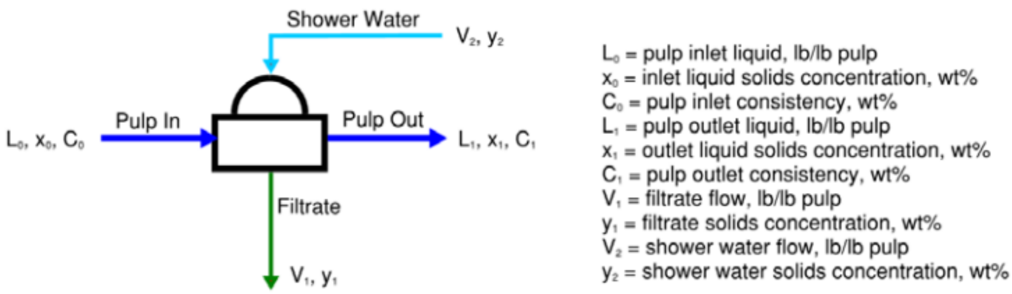

Figure 2.2: Diagram of a washing step [14]

The efficiency of a washer is commonly defined by the Norden efficiency factor. The Norden efficiency factor normalized to a 10% discharge consistency (E_{10}) is defined as follows [14]

$$Dilution\ Factor\ (DF) = V_2 - L_1 \tag{2.1}$$

$$E_{10} = \frac{ln\left[\frac{L_0(x_1 - y_2)}{L_1(x_0 - y_1)}\right]}{ln(1 + DF/9)} \tag{2.2}$$

Where all elements are explained in fig 2.2.

Another indicator of the washing efficiency is given by the displacement ratio [14]

$$Displacement\ Ratio\ (DR) = \frac{x_0 - x_1}{x_0 - y_2} \tag{2.3}$$

2.3 Bleaching

During cooking the pulp does not reach complete delignification due to the need to avoid depolymerising the cellulose in the pulp. Therefore bleaching is required to reach a clean, white pulp with high cellulose content. The bleaching of pulp is carried out in several steps using different bleaching agents to delignify and clean the pulp. In the first stages the bleaching chemical demand is high as the pulp contains a high amount of lignin and other impurities. In the later stages the selectivity of the bleaching chemicals is of importance as the low amounts of lignin need to be dissolved whilst avoiding depolymerisation of cellulose [11].

2.4 Drying

The final step in the pulping process is the drying section. The water in the pulp is removed to reduce the cost of transportation and prevent biological degradation of the pulp [15]. Drying of pulp consists of three stages: drainage, mechanical compression and thermal operation. The drainage is performed using gravitation, pulsation or vacuum and it increases the pulp consistency to 15-25%. In the mechanical compression the water is pressed from the pulp resulting in a pulp consistency of 33-55%. The more water that can be removed during the first steps the better as the final step of evaporating the remaining water consumes energy. All moisture isn't removed during the drying, around 5-9% remains in the pulp [16].

2.5 Black Liquor Treatment

The chemical recovery of black liquor consists of four parts: black liquor evaporation, condensate treatment, black liquor burning and chemical recovery. These processes generates steam, recovers cooking chemicals and provides the pulping process with wash water [11].

2.5.1 Evaporation Plant

The black liquor from the pulping plant consists mainly of water which needs to be removed to increase the heating value of the black liquor before it is burned. This is done in an evaporator plant which typically decreases the water content of the black liquor to 20-25%. To make the evaporator plant more effective several evaporators are used in series where the vapour from one evaporator unit is used to heat the next unit. The evaporator units can be configured in different ways but in a modern pulping plant the evaporator plant usually consists of 6-7 units in a counter-current setup. The condensate from the evaporators is mostly water but it also consists of organic compounds with low molecular weight such as methanol and ethanol. If the condensate is to be reused as wash water or in the preparation of white liquor these compounds must be removed. The organic compounds have a lower boiling point than water and can therefore be effectively separated using a stripper column [11].

2.5.2 Recovery Boiler

The concentrated black liquor is combusted with air as an oxidant in a recovery boiler. The combustion of biomass mainly forms CO_2 and H_2O but also other gases such as H_2, CO and NO_x [17]. Under most circumstances the combustion efficiency is at least 99%, i.e 99% of the black liquor solids entering the boiler is completely combusted into CO_2 and H_2O [18]. Apart from the gas, salt smelt is also produced in the recovery boiler which consists mainly of sodium salts [11][18].

The liquor is transported to the boiler via liquor guns which sprays the liquor into the boiler where it forms drops. Air is fed to the boiler through air ports located at different heights in the boiler [11][18]. To hinder leakage of combustion gases there is in general a weak under-pressure in the boiler [19], which is typically between 0,25 and 1 kPA [18].

Lining the walls of the boiler are heat exchanger tubes were water at its boiling point is evaporated. The evaporated water is then fed to a superheater where the hot flue gas from the boiler superheats the steam. The flue gas thereafter pass through an economizer where the water that is fed to the evaporator tubes is brought to its boiling point. The flue gas is then brought through an electrostatic precipitator to recover ash before being released into the atmosphere. To maximise the energy recovery from the flue gas the exit temperature of the gas should be as low as possible. It is however important to keep the flue gas above a temperature of 150℃ to avoid condensation of the flue gas which could cause corrosion in electrostatic precipitation unit. The steam produced in a recovery boiler is typically between 60 to 75 bars and 350 to 400℃ [11].

2.5.3 Chemical Recovery

The recovery of white liquor starts by dissolving the salt smelt from the boiler in a weak white wash, this forms green liquor which is then filtered to separate out solid materials that are present in a small amount in the green liquor. Next the green liquor is combined with burned lime mud in a causticising plant which consists of a slaker and several caustication vessels. In the causticising plant the white liquor and lime mud is formed. The liquor is then separated from the mud and is ready to be used in the pulping plant. The lime mud is washed to remove the remaining white liquor present, this wash water is the weak white wash which is used to dissolve the smelt. After washing the lime mud is burned in a lime kiln, the kiln requires fuel to operate and apart from burned lime mud, carbon dioxide is also formed. The process of the chemical recovery system as well as the reactions that take place can be seen in figure 2.3 [10][11]. The efficiency of the chemical recovery is around 97% and is therefore economically favourable [20].

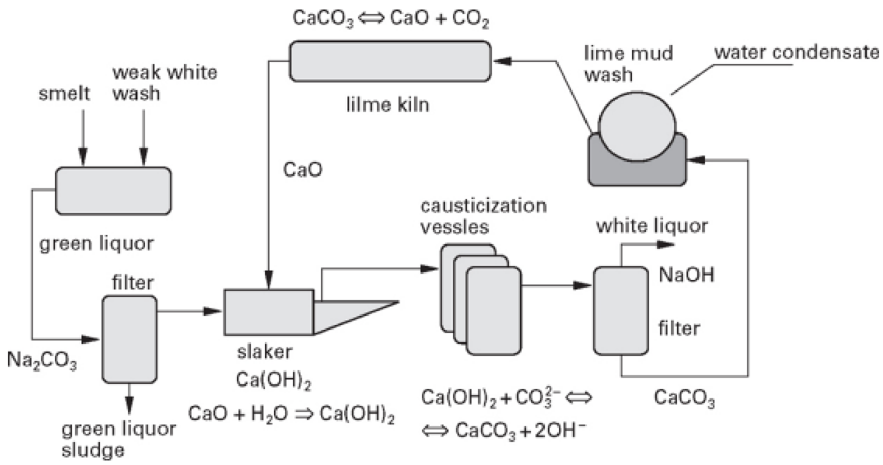

Figure 2.3: Diagram of the chemical recovery process [11]

2.6 Dissolving Pulp

For some cellulose derivatives such as textiles, esters and ethers, pulp with high cellulose content, high brightness and high cellulose reactivity is needed. Pulp with a cellulose content over 90% is called dissolving pulp (DP) and can be used to produce bio-based speciality chemicals [13][21][22]. To achieve the high cellulose content of DP a pre-hydrolysis step is needed before the cooking process to partially depolymerise the hemicellulose in the biomass [10]. With the growing interest to replace fossil-based polymers and fibers the demand for DP is increasing [12][13].

2.7 Economics

The capital cost for process equipment can be estimated in many different ways. In this report one method presented by Towler et al. is used which is based on cost parameters [23]. The capital cost for a process unit is given by

$$Cost = a + b \cdot S^c \tag{2.4}$$

where a, b and c are tabulated parameters and S a sizing parameter. The type of sizing parameter is different for different types of equipment.

For a heat exchanger the sizing parameter is given by the area of the heat exchanger. The area can be estimated using

$$A = \frac{Q}{U \cdot \Delta T_{lm}} \tag{2.5}$$

where A is the area in m^2, Q the heat duty in W, U the overall heat transfer coefficient in W/m^2K and ΔT_{ln} the logarithmic mean temperature difference in K given by

$$\Delta T_{lm} = \frac{(T_1 - t_2) - (T_2 - t_1)}{ln\frac{T_1 - t_2}{T_2 - t_1}} \tag{2.6}$$

where T_1 is the inlet temperature of the hot side, T_2 the outlet temperature of the hot side, t_1 the inlet temperature of the cold side and t_2 the outlet temperature of the cold side, all in K.

Costing data for process equipment is often not up to date, to factor in changes in pricing cost indexes are used. The cost in year x is given by

$$C_x = C_y \cdot \frac{I_x}{I_y} \tag{2.7}$$

where C_y is the cost in year y, I_x is the cost index in year x and I_y is the cost index in year y.

3
Methods

3.1 Lab process

Prior to the simulation of the soda-cooking process lab trials were performed at Chalmers University of Technology. These labs are not part of this thesis but the simulation performed is based on the method and data from the lab process.

The production of DP from wheat straw on lab scale started with a pre-hydrolysis step. The wheat straw was treated with a solution of 0.15w% HCl at a 15:1 liquid to solid (l/s) ratio, the temperature was then raised to 160℃. The pre-hydrolysis was performed for 75 min. After the pre-hydrolysis the wheat straw was washed until a neutral a pH-value had been reached. Next the cooking of the wheat straw was performed. This was done by adding 4w% NaOH with a l/s ratio of 15:1, the cooking was done for 120 min at 170℃. After the soda-cooking the pulp was washed until pH was neutral and then defibrated at 30 000 revolutions.

After cooking bleaching was performed in 3 steps, EDTA-bleaching and two peroxide-bleaching steps. In the EDTA-bleaching EDTA and H_2SO_4 was added with 0.5w% on dry pulp and 0.3w% on dry pulp respectively. The l/s ratio was 20:1 and the bleaching was performed at 60℃ for 60 min. The two peroxide-bleaching steps were identical. 5w% on dry pulp H_2O_2 and 0.1M NaOH was added with a 20:1 l/s ratio, the bleaching was done at 90℃ for 120 min. Between each bleaching step the pulp was washed until a neutral pH-value had been reached. After all the bleaching steps had been performed the pulp was defibrated at 30 000 revolutions.

3.2 Simulation in Aspen Plus

Based on the lab process the soda-cooking was modelled and simulated in Aspen Plus version 12.1. To adjust from lab scale to an industrial scale changes were made to the model, no washing was done between the pre-hydrolysis and cooking step and no defribration was simulated. The reactants were scaled linearly but the amount of water in the cooking and bleaching chemicals was reduced by 20% which can be done according to Piccinno et. al. [24]. A simplified flowchart of the simulation can be seen in figure 3.1 The lab data used for the simulation can be found in table A.1.

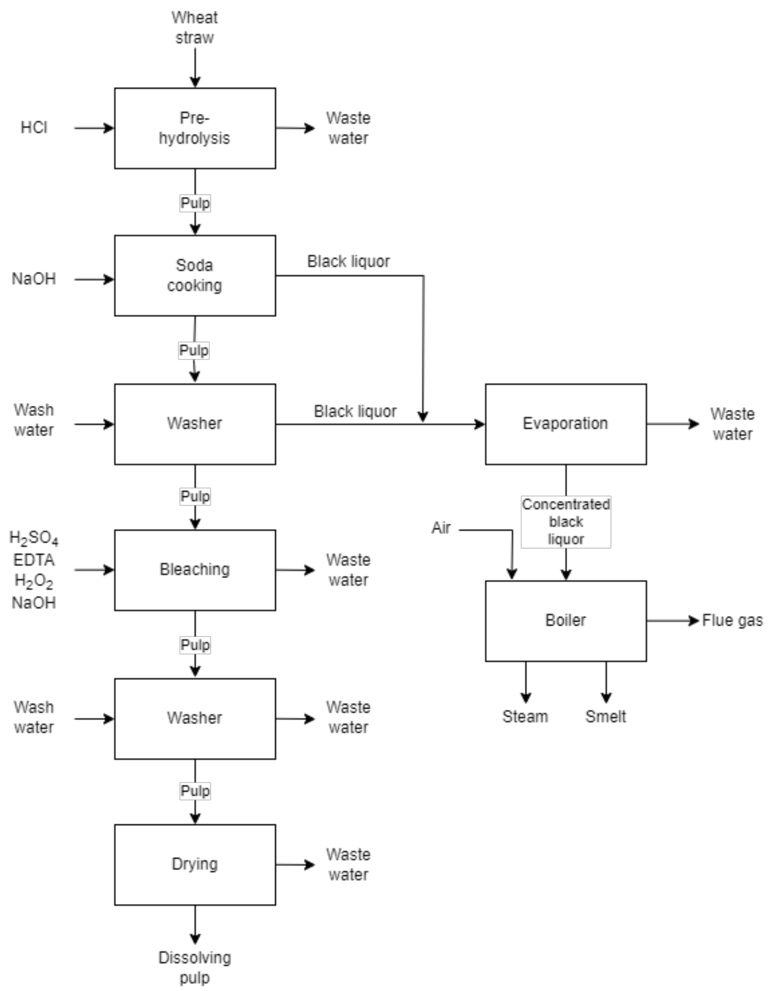

Figure 3.1: Flowchart of simulation

The wheat straw was defined by specifying the components of the biomass. Cellulose, D-galactose, D-xylose and L-arabinose were added from the Polymer databank, klason lignin and Acid Soluble Lignin (ASL) were defined as vanillin and the ash and other components were defined as silicone dioxide. All the components of the biomass were set to solids therefore the solids heat of formation at 25℃ (DHSFRM), the molar volume parameter of solid (VSPOLY) and the solid heat capacity coefficients (CPSPO1) had to be manually defined. These values were defined according to table 3.1. The data was based on a previous project simulating soda cooking [25]. All other components necessary for the process was added through databases in Aspen Plus.

Table 3.1: User defined parameters in Aspen Plus

Component	DHSFRM [cal/mol]	VSPOLY [m^3/kmol]	CPSPO1 [kJ/kmol-K]
Cellulose	-233 200	0.249	C_1 = -1.5328; C_2 = 0.67527
Galactose	-233 200	0.249	C_1 = -1.5328; C_2 = 0.67527
Xylose	-182 100	0.203	C_1 = -1.2489; C_2 = 0.55022
Arabinose	-182 100	0.203	C_1 = -1.2489; C_2 = 0.55022
Lignin	-108 248		C_1 = -0.472915; C_2 = 0.79267

To estimate physico-chemical data for the simulation the NRTL model was used as it is suitable for systems with polar substances and different phases. As the process streams contains both liquids and solids, without known particle size distribution, the MIXCISLD stream class was used for the simulation. For the streams in the simulation the biomass components were defined as CI Solids and the liquid components were defined as Mixed components. All inlet streams to the process were set to a temperature of 25℃ and a pressure of 1 bar. In all heat exchangers the minimum temperature difference was set to 10℃. The amount of wheat straw used was assumed to be 100 000 dton/year which can be covered by the production of wheat in the Götaland regions during 2021 [26].

3.2.1 Pulping

The incoming wheat straw with a moisture content of 11.4 w% was mixed with pre-hydrolysis liquor to a l/s ratio of 15:1. The pressure was raised to 7 bar and the temperature to 160℃. The pressure was chosen to keep the stream in liquid phase. The pre-hydrolysis was simulated through a stoiciometric reactor were the biomass components went from solid to dissolved in fractions according to the lab data. After the pre-hydrolysis the waste liquor was separated from the pulp. This was done through a components separator together with a design specification setting the water content of the pulp out of the separator to 80% as that is in the range for drainage of pulp [16]. It was assumed that the dissolved biomass and pre-hydrolysis chemicals were perfectly mixed, therefore a calculator was used to set the split fraction of dissolved components entering the waste stream to be equal to the split fraction of water entering the waste stream. All of the solid biomass components were set to enter the pulp stream.

The pulp was then mixed with the cooking liquor to a l/s ratio of 15:1. The pressure was raised to 8 bar and the temperature to 170℃. The cooking was simulated in the same way as the pre-hydrolysis step with a stoichiometric reactor followed by a components separator which separated the pulp in the same way as in the prehydrolysis step.

3.2.2 Washing

The washing of the pulp was simulated with a separator. The efficiency of the washing was calculated according to equations 2.1, 2.3 and 2.2. It was calculated as a two stage pressure diffuser washer. Data used for the calculations can be found in table A.2. The water content of the pulp was assumed to be constant through the washing system and the amount of wash water fed to the washer was assumed to be 1.1 times the amount of water present in the pulp, it was pumped to a pressure of 9 bar. All liquid removed in the washing step was sent to the recovery boiler.

3.2.3 Bleaching

The bleaching was simulated in a single step as no lab data was available of the composition of the pulp in between the different bleaching steps. The bleaching chemicals were specified according to the lab process and the bleaching was simulated in a stoiciometric reactor followed by a components separator specified like the seperators in the pre-hydrolysis and cooking step. The bleaching was performed at 80°C and 1 bar.

After the bleaching a washer was added in the same way as in the washing step but was calculated as a single-stage pressure diffuser and the wash water was pumped to a pressure of 7 bar.

3.2.4 Drying

The dryer was simulated as two separators, one for the mechanical drying and one for the thermal drying. In the mechanical dryer the moisture content of the pulp was set to 50% and the moisture content of the pulp after the thermal dryer was set to 6%. The moisture content was set with design specifications.

3.2.5 Evaporation

All streams going to the recovery boiler were combined to one stream with a mixer which was then sent to the evaporator section of the simulation. The evaporator was setup as a co-current evaporator with 5 evaporators. The evaporators were simulated as flash tanks combined with heaters that condensed the vapour from the evaporators. The heat from the condensation was then used in the next tank to evaporate the water. This was done by connecting an energy stream from the heater to the evaporator. The first evaporator was set to operate at 160°C, then each consecutive evaporator was 10°C lower to have a suitable temperature difference between the condensers and evaporators. With a design specification the vapour fraction in the first evaporator was varied so that the excess energy from the final evaporator would be zero.

3.2.6 Boiler

The boiler was modelled as a stoichiometric reactor were the biomass components reacted with air into carbon dioxide and water. The conversion of the biomass was set to 99%, and the reactor was set to operate at 0.999 bar. The heat duty of the reactor was set to 0 so that all the heat from the reactions was converted into heat in the flue gas. The flow of air was adjusted with a design specification so that the amount of oxygen leaving the boiler would be 2% to ensure that there was excess oxygen in the process.

To simulate the heat recovery from the boiler the hot flue gas was sent through three heat exchangers, an evaporator, a superheater and an economizer. All heat exchangers were set up as shortcut heat exchangers with a countercurrent flow. On the cold side of the economizer, water that was set to a pressure of 60 bar was fed and brought up to its saturation temperature. The water was then led to the evaporator were all the water was vapourised. In the superheater the steam was heated to 350℃. In the economizer, the amount of water was varied with a sensitivity analysis to find the massflow of water needed to reach a stack temperature of 220℃. This temperature was chosen to give some flexibility to the boiler so that in case of lower production the temperature of the flue gas would stay above 150℃.

3.2.7 Chemical Recovery

The chemical recovery was not simulated. Instead the amount of recovered NaOH from the process was calculated with a calculator block. 97% of the NaOH present in the recovery stream was assumed to be recovered.

3.3 Heat Integration

Two scenarios were considered when evaluating the heat integration of the process, one were both the black liquor from the cooking and washing steps were sent to the black liquor treatment, and one were only the black liquor from the cooking step was sent to the black liquor treatment. The heat integration was performed in Aspen Energy Analyzer. Data of the streams that needed to be heated and streams that had potential of cooling were manually added from the simulation. The heat transfer coefficient (HTC) for each stream was chosen through the HTC database in Aspen Energy Analyzer. For the different scenarios different designs were tested and evaluated. The minimum temperature difference used in the heat exchangers was 10℃. Based on the sizing and load data from Aspen Energy Analyzer the cost of the designs could be calculated and compared. The energy requirement for the evaporator plant was not considered in the heat integration but it was assumed that it would be completely covered by the steam produced in the recovery boiler. The steam not needed in the evaporator was considered to be able to combined with steam purchased by the plant.

The steam and electricity required for the dryer section was estimated based on data for energy consumption of drying tissue paper and can be found in table A.3. As the amount of black liquor recovered differed between the scenarios the cost of make-up NaOH was considered when comparing the scenarios. The heat exchangers from the most cost efficient design were added to the simulation in Aspen Plus and was used in the cost evaluation of the process. The yearly cost for the heat exchangers was calculated by assuming a interest rate of 10% and a lifespan of 10 years.

3.4 Cost Evaluation

For the cost evaluation two scenarios were studied, one were the pulp needed to be dried to a moisture content of 6% and one were it was assumed that the pulp would be further processed directly after the pulping and therefore only mechanical drying was necessary.

3.4.1 Equipment cost

Based on the data from the simulation the equipment could be sized and the investment cost estimated using the cost data in table A.4.

The digesters, washers and bleachers were calculated as vertical pressure vessels made of stainless steel. Stainless steel was used as it is corrosion resistant [23]. The thickness of the vessels was estimated based on the diameter according to table A.5 The cost for the filters for the drainage of the vessels was calculated as sieve trays.

The dryer cost was calculated as a horizontal stainless steel pressure vessel with a diameter of 3.5 m which is a common size for a Yankee dryer typically used in drying of tissue-grade pulp [16] [27].

The cost of the evaporators was calculated as shell and tube heat exchangers, the area was calculated according to equation 2.5 with an overall heat transfer coefficent estimated to 1500 W/m^2K according to data in Chemical Engineering Design [23].

For the boiler data, a field erected boiler was used to estimate the cost. The size of the evaporator, superheater and economizer could be taken directly from the simulation and was used to estimate the cost. For the evaporator and economizer shell and tube heat exchanger was used and for the superheater a double pipe heat exchanger was used due to the small area of the superheater.

The cost of the pumps was calculated as single stage centrifugal pumps. Pumps were considered to be placed on each inlet stream to all of the unit operations of the process. As industry standard two pumps were assumed to be needed as a fail-safe measure.

3.4.2 Capital cost

As the equipment cost only takes in consideration the investment cost of the equipment for the plant capital cost factors were used to estimate the extra cost needed for erecting the plant. The factors used can be found in table A.6. The total cost for the equipment was multiplied by the fixed capital cost factor to get a more accurate estimate of the total capital cost. As the data for the equipment cost was based on prices in 2010 data from the Chemical Engineering Plant Cost Index (CEPCI) was used to calculate the investment cost of the plant in 2022 [28].

3.4.3 Operating cost

The variable operating costs were calculated using data from the simulation. The prices for the raw materials and utilities was compiled and estimated from different sources and can be found in table A.7

The fixed operating costs was calculated according to table 3.2. It was assumed that 20 employees were needed with a wage of 35 000 SEK/month and a factor of 1.45 for allowances and overhead.

Table 3.2: Cost estimation of fixed operating cost [17]

No	Specification	Cost Estimation
1	Maintenance	5% of capital cost
2	Personnel	Calculated separately
3	Laboratory costs	20% of (2)
4	Supervision	20% of (2)
5	Plant overheads	50% of (2)
6	Capital charges	10% of capital cost
7	Insurance	1% of capital cost
8	Taxes	2% of capital cost
9	Royalties	1% of capital cost
10	General OH and R&D	20% of $\sum_1^9 No_n$

3.4.4 Evaluation

To evaluate the economics of the plant a break-even price was calculated for the dissolving pulp. The break-even price was determined by assuming full production during the first year, an interest rate of 10% and a lifetime for the plant of 10 years. The operating time for the plant was assumed to be 8400 h/year. Based on these assumptions an economic calculation was performed over the lifespan of the plan where the total cumulative cost for the plant was calculated. The total cumulative cost was then divided by the total production of dissolving pulp to obtain the break-even price.

4
Results

4.1 Process Simulation

The input of 112 900 ton/year of wheat straw with a moisture content of 11.4% generated 45 900 ton/year of dissolving pulp with a moisture content of 6%, this is equivalent to 43 500 dton/year of dissolving pulp, dton refers to the weight completely without moisture. The chemical requirements for the process is presented in table 4.1.

Table 4.1: Chemical requirements of the final process

Chemical	Flow [kg/h]
HCl	175
Total NaOH	4074
Make-up NaOH	196
H_2SO_4	16
EDTA	26
H_2O_2	457

Most of the NaOH from the cooking process was recovered but some make-up was required as presented in table 4.1. This both due to the efficiency of the chemical recovery and that the NaOH used in the bleaching is not recovered. The total input of the chemical streams can be found in table A.10.

The created process model is presented in several steps below.

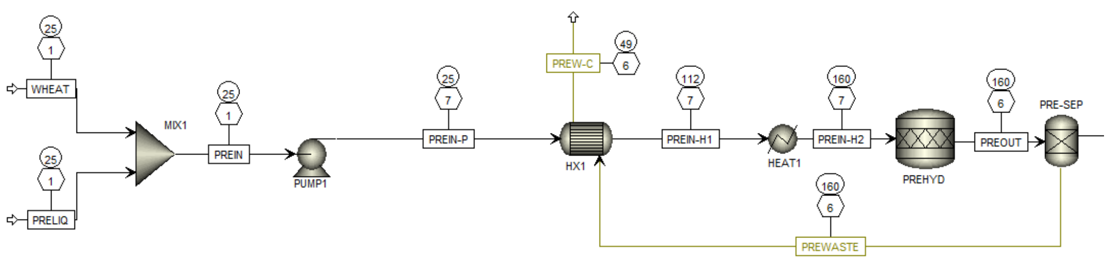

Figure 4.1: The pre-hydrolysis section of the simulation. Name of flows in rectangles, temperatures [°C] in circles and pressures [bar] in hexagons.

The simulation starts with the pre-hydrolysis step which is seen in figure 4.1. The input to the pre-hydrolysis was the wheat straw and pre-hydrolysis liquor. The heating demand could be partially covered by the waste stream from the pre-hydrolysis step.

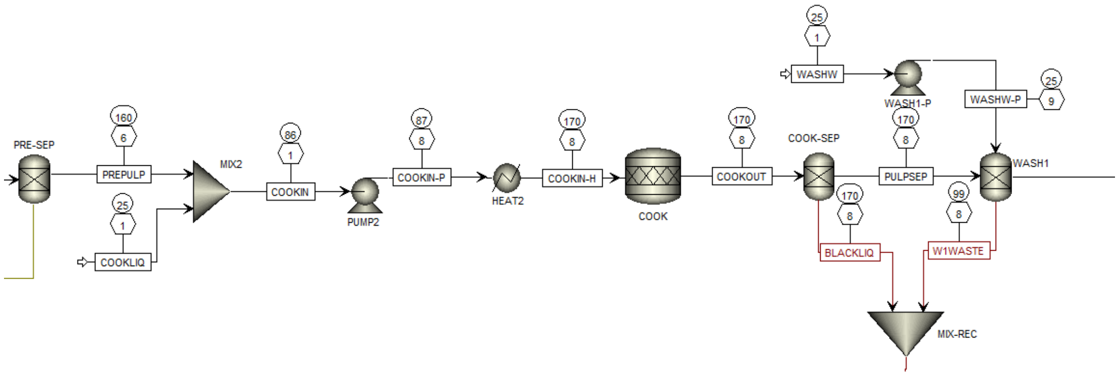

Figure 4.2: The cooking and washing sections of the simulation. Name of flows in rectangles, temperatures [°C] in circles and pressures [bar] in hexagons.

Following the pre-hydrolysis was the cooking and washing presented in figure 4.2. The inputs to the reactor was the pulp from the pre-hydrolysis and white liquor. The pulp from the cooking step was then washed. It was calculated that 97.93% of the dissolved components was washed from the pulp and the required wash water was 35 444 kg/h. The black liquor from the cooking and washing step was mixed and sent to the evaporation section.

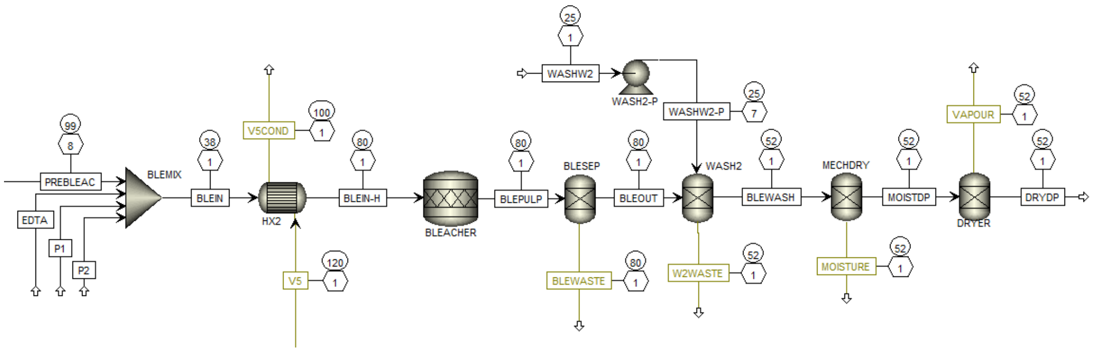

Figure 4.3: The bleaching, final washing and drying sections of the simulation. Name of flows in rectangles, temperatures [°C] in circles and pressures [bar] in hexagons.

The bleaching step, final washing step and drying can be seen in figure 4.3. The bleaching was simulated as a single unit with four inputs; washed pulp and three bleaching liquor streams. The heating demand for the bleaching was covered by steam from the evaporator plant.

The second washing stage was calculated to be able to remove 87.5% of the dissolved components and required a wash water flow of 23 355 kg/h.

The washed pulp was then dried, the steam requirement for drying pulp to a moisture content of 6% was estimated to be 5259 kW and the electricity requirement was estimated to be 5229 kW.

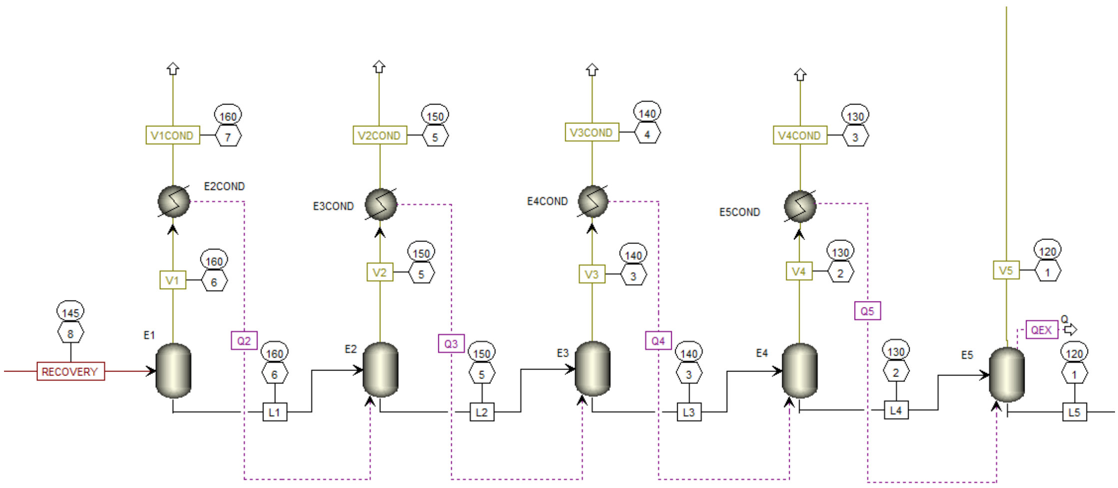

Figure 4.4: The evaporation section of the simulation. Name of flows in rectangles, temperatures [°C] in circles and pressures [bar] in hexagons.

The 5-step co-current evaporation section can be seen in figure 4.4. The entering stream was black liquor from the cooking and washing step. The steam required for the evaporation section was 11 460 kW. The evaporation section reduced the water content of the black liquor from 91.7% to 17.5%.

It was tested to simulate the evaporator train in a counter-current setup but the iterations required to solve the simulation were not possible for more than three evaporators and the requirements for the concentrated black liquor could not be met with three evaporators.

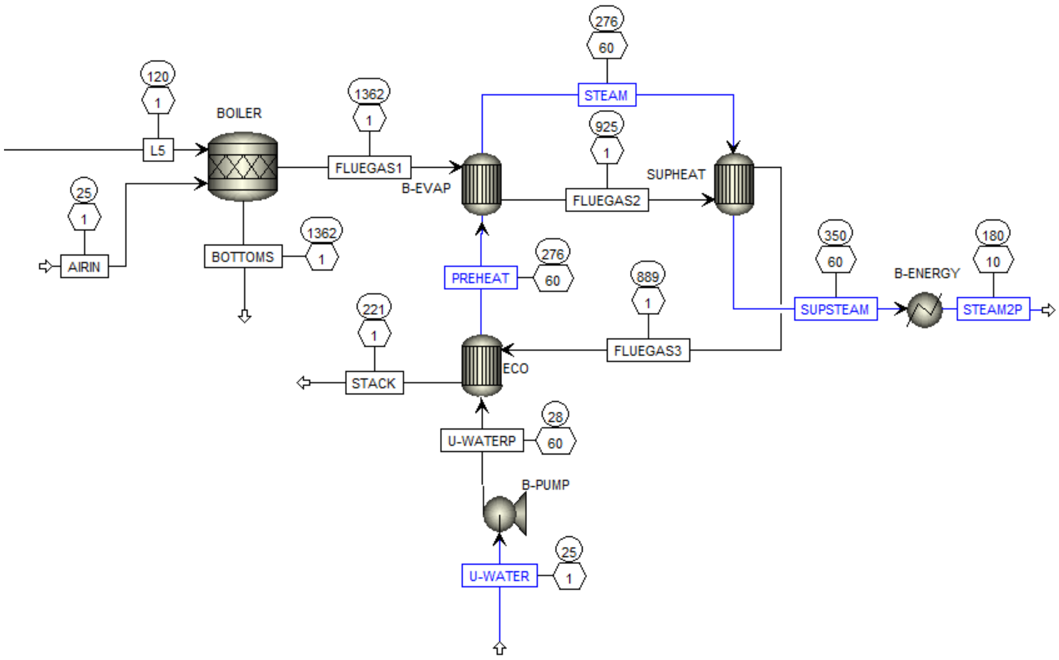

Figure 4.5: The boiler section of the simulation. Name of flows in rectangles, temperatures [°C] in circles and pressures [bar] in hexagons.

The concentrated black liquor entered the boiler which can be seen in figure 4.5. The amount of air required in the boiler was 28 000 kg/h. The boiler produced 21 750 kg/h of steam, the energy from the produced steam was 14 195 kW. The amount of NaOH entering the boiler was 3998 kg/h, from this the amount of recovered NaOH was calculated to be 3878 kg/h

The energy requirements for the process without heat integration is presented in table 4.2.

Table 4.2: Energy requirement for process without heat integration

Process	Type	Demand [kW]
Pre-hydrolysis	MP-steam	24 830
Cooking	MP-steam	10 357
Bleaching	MP-steam	10 038
Drying	Steam	5259
Drying	Electricity	5229
Pumps	Electricity	82
Evaporation	MP-steam	11 460
Boiler	Steam production	-14 195
Total steam		47 744
Total electricity		5309

As can be seen in table 4.2 the pre-hydrolysis step had the highest steam demand in the process. The electricity requirements were low for the pumps but high for the drying of the pulp.

4.2 Heat Integration

Of the two scenarios studied in the heat integration, black liquor from cooking and washing and black liquor only from cooking, it was quickly discovered that the second scenario had a much higher cost. This was due to the lower amount of NaOH that could be recovered even though the amount of steam available for the pulping process was higher. Therefore further work was only performed on the first scenario.

The most cost efficient design found for the heat integration is presented in figure 4.6. This design required 4 heat exchangers to cover the heating demand of the process: 2 process-to-process exchangers and 2 heaters.

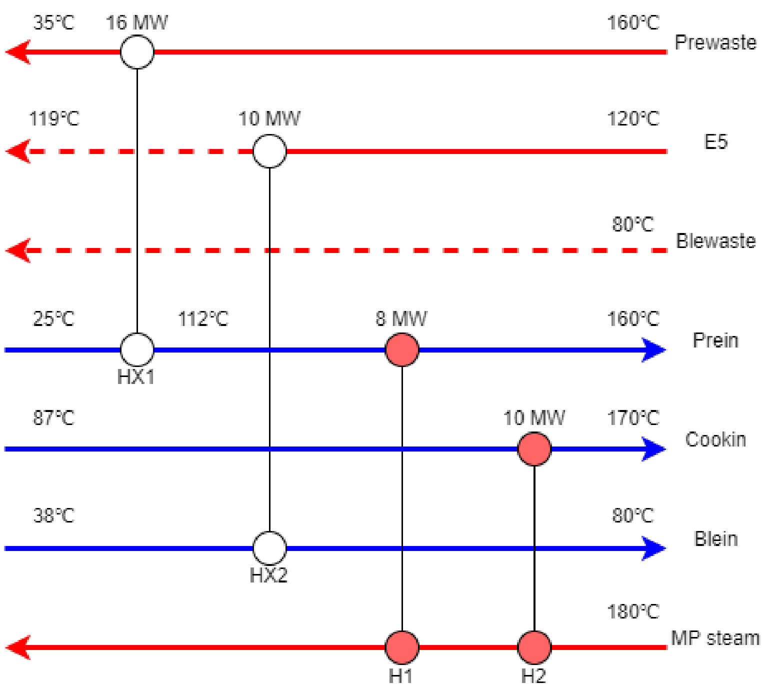

Figure 4.6: The final design of the HEN

As can be seen in figure 4.6 the heating of the pre-hydrolysis stream (Prein) was partially covered by the waste stream from the pre-hydrolysis (Prewaste). The rest of the heating demand was met by MP-steam. The input to the cooking section (Cookin) was heated entirely by steam and the bleaching section (Blein) was completely covered by steam generated from the last evaporator in the evaporation section (E5). As can be seen in figure 4.6 the heat from the waste stream of the bleaching step (Blewaste) was not utilised in the heat integration and there was some energy left in the steam from the evaporator section. The waste streams were assumed to not require cooling as the heat requirements for wastewater treatment is unknown. Details of the heat exchangers, steam demand and cost is presented in table 4.3. The total steam demand for the process including the drying section was 21 674 kW.

Table 4.3: Demand, area and cost of heat integration design

Streams	Duty [kW]	HX area [m^2]	Cost 2022 [SEK]	Cost 2022 [SEK/year]
PREIN-PREWASTE	16 036	648	2 325 819	3 931 020
BLEIN-E5	10 038	96	612 263	1 034 826
PREIN	8793	128	691 233	1 168 298
COOKIN	10 357	157	765 588	1 293 971
Variable costs	Demand			Cost 2022 [SEK/year]
Steam	16 415 kW			13 953 731
NaOH	149 kg/h			7 605 020
Total cost/year				28 986 442

4.3 Cost Evaluation

The fixed capital cost factor was calculated to be 4.32, this was used to estimate the total fixed capital cost of the plant. The capital cost of the plant including the drying section is presented in table 4.4.

Table 4.4: Capital cost of plant including the drying section

Unit	Cost 2022 [SEK]
Heat Exchangers	4 394 817
Digesters	12 620 161
Bleachers	11 923 571
Washers	8 974 234
Dryer	3 558 950
Evaporators	13 813 219
Boiler	9 349 497
Total	64 634 448
Fixed Capital Cost	279 220 816

The capital cost for the dryer was relatively small and the fixed capital cost for the plant excluding the dryer was 272 708 691 SEK.

The variable operating cost of the plant including the drying section is presented in table 4.5.

Table 4.5: Variable operating cost of plant including the drying section

	Cost 2022 [SEK/year]
Wheat straw	73 114 528
Process water	4 939 116
HCl	2 249 100
NaOH	10 095 160
H_2SO_4	787 141
EDTA	8 583 806
H_2O_2	17 523 994
Electricity	61 471 045
Steam	1 820 573
Total	180 584 463

The difference between in the operating costs between the plant with and without the dryer lies in the electricity and steam consumption. The variable operating cost for the plant without a dryer section was 120 057 742 SEK/year.

The fixed operating cost of the plant including the drying section is presented in table 4.6.

Table 4.6: Fixed operating cost of plant including the drying section

	Cost 2022 [SEK/year]
Maintenance	13 961 041
Labour	12 180 000
Laboratory costs	2 436 000
Supervision	2 436 000
Plant overheads	2 436 000
Capital charges	27 922 082
Insurance	2 792 208
Taxes	5 584 416
General OH and R&D	14 507 991
Total	87 047 946

The fixed operating cost for the plant without a dryer section was 85 563 182 SEK/year. The total operating cost for the two scenarios was 267 632 409 SEK/year for the plant with a dryer section and 205 620 924 SEK/year for the plant without.

The comparison between the two scenarios and the break-even price is presented in table 4.7. The entire economic calculation can be seen in Appendix A.7.

Table 4.7: Economic evaluation of the process plants

	With drying	**Without drying**
Fixed capital cost [SEK]	279 220 816	272 708 691
Annual operating cost [SEK/year]	267 632 409	205 620 924
Break-even price [SEK/dton DP]	7194	5736

5
Discussion

The model developed in Aspen Plus together with the economic evaluation provides a base for development of a soda based pulping plant. In further development of the process there are several aspects that need to be considered.

5.1 Simulation

In a full scale plant there are various features required that are not part of the model which only includes the pulping, bleaching, drying, evaporation and boiler section of the plant. In a real life scenario more steps are required. These include pre-treatment of straw with storage, drying and cutting, wastewater treatment, flue gas cleaning and the complete chemical recycling process. All these processes will result in a higher capital and operating cost and therefore a higher break-even price.

Furthermore the simulation in Aspen Plus was done as a continuous process and the processing time for each unit was not considered when determining the equipment needed. In the laboratory scale the different steps had varying processing times. The timing of the process must therefore be considered if the process is developed further and intermediary storage may become necessary between some units.

5.2 Evaporation and Boiler Section

The steam demand in the evaporators is greatly affected by the water content of the black liquor. In the simulation the water content of the black liquor was 91.7% which is higher than in a typical kraft pulping process where it is around 80-85% [11]. Reducing the water content in the black liquor would lead to steam savings. This can be done by either reducing the amount of wash water used or reducing the amount of water in the white liquor. In the model the amount of wash water was only 1.1 times larger than the water content in the pulp. Theoretically in displacement washing a 1:1 ratio between the wash water and pulp water is needed to clean the pulp. A 1.1:1 ratio is therefore quite low and should probably not be reduced further to ensure that a clean pulp is obtained. The effectiveness of the washing also affects how much biomass is present in the black liquor which in turn affects the energy produced in the boiler. Lowering the amount of water in the white liquor is therefore a more reasonable approach to reduce the steam demand of the evaporators. In the simulation the amount of water was reduced by 20% compared to the lab process. The amount of liquor used was determined based on the dry content of the pulp. There is a large amount of water present in the pulp after the pre-hydrolysis step that the white liquor dissolves in. If the liquor amount had been based on pH-values it may have been possible to see if the amount of water required could have been further reduced. Further studies of how a reduction of water would affect the pulping process is needed to see if high quality dissolving pulp can be produced with a reduced amount of water. A reduction of water in the processing liquids would also lead to a reduced heating demand.

Due to constraints in Aspen Plus the evaporation section was simulated as a co-current evaporation train and 5 units were used. It is custom to use 6-7 evaporator units in a counter-current setup in the pulping industry as this is more effective. Therefore the energy requirement for the evaporation section of the plant may be lowered if a counter-current setup were to be implemented with more units thus lowering the steam demand of the plant. A more extensive analysis of the evaporation section is therefore an important step in improving the developed process. There is also the need to account for the boiling-point rise of the black liquor during the evaporation. Taking the boiling-point rise into account will result in a less effective evaporation section but is important when further developing the process to get a more accurate picture of the process.

When performing the heat integration 2 scenarios were considered, one where all the black liquor from the process was sent to the recovery boiler and one where only the black liquor from the cooking step was sent to the recovery boiler. The results shows that utilising all of the black liquor from the process is more economical due to the amount of NaOH that can be recovered. The evaluation of the heat integration did not take the cost of the evaporation and boiler section of the plant in consideration. If black liquor from just the cooking step had been utilised the size of the evaporation and boiler section could have been reduced thus lowering the capital and operational cost. However the black liquor not utilised needs to be cleaned meaning the waste water treatment of the plant would need to be larger thus increasing capital and operational costs. By recovering more NaOH the first scenario is also more beneficial from an environmental point of view.

Another aspect to consider is the integration of the black liquor recovery with an already existing pulping plant. The evaporation and boiler section of the plant stand for more than a third of the total capital cost of the plant as seen in table 4.4 which would not be needed if the plant was integrated with an already existing plant resulting in large savings. This would also reduce the steam demand of the plant. Integrating the recovery process requires more factors to be examined. First, there needs to be an existing plant that can handle the additional black liquor stream and have enough land to erect an additional pulping plant. The majority of the existing pulp plants in Sweden are kraft mills [29], integrating the black liquor from the soda cooking process would thus result in the produced white liquor being unsuitable for the soda cooking as it contains sodium sulphide. Not being able to recover the sodium hydroxide will increase the operating costs of the process. Another aspect to consider is how a black liquor integration will affect the heat integration of the plant. In the suggested design, steam from the final evaporator is used to heat the bleaching stream. This source of heat will consequently need to be replaced. In the heat integration analysis, designs that didn't utilise the steam from the evaporators were investigated and found to be suitable for the process. The steam network will also need to be reconsidered in the case of integrating the black liquor recovery with an existing plant. No steam will be produced at the soda plant but excess steam might be available from the neighbouring plant.

5.3 General Economics

In the economic evaluation pumps were only placed at the inlets to the different units, it is likely that more pumps are required in an actual process plant. For example in the heat exchanger network (HEN) the steam and hot fluids need to be transported around the plant. Furthermore the electricity requirements of the pumps could only be evaluated for the pumps increasing the pressure in the process, as seen in table 4.2 this demand is relatively low. Due to the high electricity prices today an increase in energy demand will have a large effect on the variable operating cost.

The high electricity price is also the main reason for the difference in break-even price between the two scenarios including and excluding drying. A drying section is not necessary in the process if the pulp is to be used directly after pulping. It is therefore more profitable and energy efficient to integrate the pulping plant directly to a process further refining the dissolving pulp. As the electricity price was high during 2022 the price difference between the two scenarios is large, however prices today (May 2023) are lower than the price used in the economic evaluation of the project [30]. Therefore the difference between the two scenarios may be lower if the electricity price does not rise again.

As can be seen in table 4.2 the steam demand for the pre-hydrolysis is relatively high. The main reason for the high demand is due to the large change in temperature, the inlet stream needs to be raised from 25℃ to 160℃. However the use of the waste stream from the pre-hydrolysis to pre-heat the inlet stream led to a large reduction of the steam demand. The temperature of the inlet streams affect the steam demand of the plant, it may therefore be worthwhile to investigate if some inlet streams may have higher temperatures in a real plant.

As the process of producing dissolving pulp from wheat straw is new and no conventional plant exists today the lifespan of the plant was assumed to be only 10 years. The majority of European mills in operation are 10-40 years and can be expected to operate for up to 60 years [31]. A longer lifespan of the plant would result in a lower break-even price. Another assumption affecting the break-even price is that the plant could go into full production during the first year of construction. In reality a plant would take a couple of years to construct and operate at a lower production in the begin of its operation making it less profitable than if full scale production was in place.

The estimation of the capital cost of the process is based on data from 2010. Even though the CEPCI was used to adjust the prices to 2022's level the data used for the price estimation is outdated. Due to changes in for instance technology, construction practices and labour efficiency a newly constructed plant may have other sizing and cost estimation data. However new data is not readily available and was therefore not used. The method used in this project is a quick method that gives a rough estimation of the cost which should be taken in consideration in its evaluation.

5.4 Future work

As mentioned in the discussion there are several aspects of this thesis that can be further worked upon. Firstly the additional processing steps required for a complete plant need to be investigated and how these will affect the capital and operational cost. The continuity of the process must also be considered, how a batch process would be setup or if it would be possible to operate the process continuously.

There are several aspects of the evaporation section of the plant that can potentially increase the efficiency and should be investigated. This includes the water content of the black liquor, the setup of the evaporation train both in consideration to number of units and configuration, and if it would be advantageous to integrate the recovery process with a neighbouring plant.

The performed cost evaluation was done as a quick and basic estimate. To get a more accurate cost of the process a more detailed evaluation is required. The start-up time and lifespan of the plant needs to be considered, the use of newer data for cost estimation is required and the affect on the operating cost when prices fluctuate should be investigated.

6 Conclusion

The techno-economic assessment performed in this thesis provides a base for development of a soda-based production of dissolving pulp from wheat straw. The market price of wood based dissolving pulp is around 10 000 SEK/ton [25]. There is therefore a possibility for a profitable production but improvements can be made to increase its competitiveness in the market. The developed process has a break-even price of 7194 SEK/dton DP, a steam demand of 21 674 kW and an electricity demand of 5309 kW. In the scenario where the drying was excluded the break-even price was 5736 SEK/dton DP, the steam demand was 16 415 kW and the electricity demand was 82 kW. Compared to the base case with a steam demand of 47 744 kW the heat integration reduced the steam demand of the process. Due to the simplifications made in the project further work is needed to get a more accurate overview of the process and economics.

The production of dissolving pulp from wheat straw shows promise of being a feasible alternative to production of wood-based dissolving pulp but further development is needed.

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A

Appendix 1

A.1 Lab Data

The lab data used to construct the model in Aspen Plus is collected from a thesis by Sjöstedt [32] and can be seen in table A.1

Table A.1: Composition of wheat stalks/pulp in w% after treatment stages

Treatment stage	Glu	Xyl	Gal	Ara	Man	Klason	ASL	Ash	$Others^1$
No treatment	41.56	12.08	0.8	2.15	0	23.86	2.55	6.8	10.2
$Pre-hydrolysis^2$	56.44	9.62	0.15	0.42	0	26.26	1.46	2.67	2.97
Cooking	94.8	2.45	0	0.08	0	1.3	0	0.6	0.77
Bleaching	95.4	2.38	0	0.08	0.79[3]	0.74	0	0.19	0.42

1: Adjusted to reach composition of 100%
2: Based on pre-hydrolysis for wheat straw
3: Assumed to be 0 in simulation

A.2 Washer Data

Table A.2: Data used for calculating setup of washers [14]

Washer type	Feed Consistency [%]	Discharge Consistency [%]	DR	E_{10}
Pressure Diffuser	8-12	8-12	0.875	5

A.3 Drying Data

Table A.3: Data used for calculating energy consumption of dryer [16]

Paper grade	Steam [kWh/ton H_2O]	Electricity [kWh/ton H_2O]
Tissue	1056	1050

A.4 Capital Cost Data

Table A.4: Data used for estimating equipment cost in USD 2010 [23]

Equipment	Unit for size, S	a	b	c
Boiler				
Field Erected	kg steam/h	130 000	53	0.9
Exchangers				
U-tube shell and tube	area, m^2	28 000	54	1.2
Double pipe	area, m^2	1 900	2 500	1.0
Pressure vessels				
Vertical	shell mass, kg	17 400	79	0.85
Horizontal	shell mass, kg	12 800	73	0.85
Pumps				
Single stage centrifugal	flow, liter/s	8000	240	0.9
Trays				
Sieve trays	diameter, m	130	440	1.8

Table A.5: Minimum wall thickness for pressure vessels [23]

Vessel diameter [m]	Minimum thickness [mm]
1	5
1 to 2	7
2 to 2.5	9
2.5 to 3	10
3 to 3.5	12

Table A.6: Factors for estimation of fixed capital costs [23]

No	Type	Factor
1	Equipment erection	0.5
2	Piping	0.6
3	Instrumentation and control	0.3
4	Electrical	0.2
5	Civil	0.3
6	Buildings	0.2
7	Lagging and paint	0.1
8	Design and engineering	0.25
9	Contingency	0.1
$f_{tot} = (1 + \sum_1^7 f_n)(1 + \sum_8^9 f_n)$	**Total**	4.32

A.5 Operating Cost Data

Table A.7: Prices used to calculate operating cost

Input	Cost	Unit	Source
Wheat straw	0.73	SEK/kg	[17]
Process water	0.01	SEK/kg	Estimated from [23]
HCl	1.53	SEK/kg	[33]
NaOH	6.13	SEK/kg	[33]
H_2SO_4	5.99	SEK/kg	[33]
EDTA	39.16	SEK/kg	[33]
H_2O_2	4.57	SEK/kg	[33]
Electricity	1378	SEK/MWh	Cost 2022 in area SE3 [34]
Steam	0.01	SEK/kWh	Estimated from [23]

A.6 Stream Data

Table A.8: Specifications for input streams

Stream/component	Flow [kg/h]
WHEAT	13 438
Water	1533
Biomass	11 905
PRELIQ	142 889
Water	142 714
HCl	175
COOKLIQ	58 743
Water	54 716
NaOH	4027
EDTA	83 513
Water	83 471
H_2SO_4	15.7
EDTA	26.1
P1	41 808
Water	41 527
NaOH	20.9
H_2O_2	261
P2	41 832
Water	41 610
NaOH	26.1
H_2O_2	195.5

A.7 Economic Calculations

Table A.9: The economic calculation for the process with drying

Year	2023	2024	2025	2026	2027
Cap. cost [SEK]	27 922 082	27 922 082	27 922 082	27 922 082	27 922 082
Rent [SEK]	27 922 082	25 129 873	22 337 665	19 545 457	16 753 249
Op. cost [SEK]	267 632 409	267 632 409	267 632 409	267 632 409	267 632 409
Net cost [SEK]	323 476 572	320 684 364	317 892 156	315 099 948	312 307 740
Cum.cost [SEK]	323 476 572	644 160 936	962 053 092	1 277 153 040	1 589 460 780
Production [kg/y]	43 219 779	43 219 779	43 219 779	43 219 779	43 219 779

Year	2028	2029	2030	2031	2032
Cap. cost [SEK]	27 922 082	27 922 082	27 922 082	27 922 082	27 922 082
Rent [SEK]	13 961 041	11 168 833	8 376 624	5 584 416	2 792 208
Op. cost [SEK]	267 632 409	267 632 409	267 632 409	267 632 409	267 632 409
Net cost [SEK]	309 515 531	306 723 323	303 931 115	301 138 907	298 346 699
Cum.cost [SEK]	1 898 976 311	2 205 699 635	2 509 630 750	2 810 769 657	3 109 116 356
Production [kg/y]	43 219 779	43 219 779	43 219 779	43 219 779	43 219 779

Table A.10: The economic calculation for the process without drying

Year	2023	2024	2025	2026	2027
Cap. cost [SEK]	27 270 869	27 270 869	27 270 869	27 270 869	27 270 869
Rent [SEK]	27 270 869	24 543 782	21 816 695	19 089 608	16 362 521
Op. cost [SEK]	205 620 924	205 620 924	205 620 924	205 620 924	205 620 924
Net cost [SEK]	260 162 662	257 435 575	254 708 488	251 981 401	249 254 314
Cum.cost [SEK]	260 162 662	517 598 237	772 306 725	1 024 288 126	1 273 542 440
Production [kg/y]	43 219 779	43 219 779	43 219 779	43 219 779	43 219 779
Year	2028	2029	2030	2031	2032
Cap. cost [SEK]	27 270 869	27 270 869	27 270 869	27 270 869	27 270 869
Rent [SEK]	13 635 435	10 908 348	8 181 261	5 454 174	2 727 087
Op. cost [SEK]	205 620 924	205 620 924	205 620 924	205 620 924	205 620 924
Net cost [SEK]	246 527 227	243 800 140	241 073 054	238 345 967	235 618 880
Cum.cost [SEK]	1 520 069 668	1 763 869 808	2 004 942 862	2 243 288 828	2 478 907 708
Production [kg/y]	43 219 779	43 219 779	43 219 779	43 219 779	43 219 779

DEPARTMENT OF CHEMISTRY AND CHEMICAL ENGINEERING
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden
www.chalmers.se