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Energy Analysis of Hemicellulose Extraction at a Softwood Kraft Pulp Mill

Case Study of Södra Cell Värö

Master of Science Thesis

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CHALMERS UNIVERSITY OF TECHNOLOGY
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The thesis was performed in collaboration with Södra Cell Värö and Södra Innovation.

Cover: Arial photo of the Värö Pulp Mill, showing biological cleaning plant (front) paper room (left), recovery boiler (back) and the wood yard (right). Source: Södra Cell, Värö.

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Göteborg, Sweden, 2013

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Abstract

The Swedish pulp and paper industry is struggling with lower revenues and high raw material cost. The implementation of biorefinery concepts is not only interesting from a diminishing fossil feedstock perspective but also as a way of creating new revenue streams for the pulp and paper industry. Production of dissolving pulp would be beneficial both due to the higher market price of the pulp but also since the hemicellulose extracted could be upgraded to value-added by-products.

The conversion from Kraft to dissolving pulp production has been evaluated in a case study of Södra Cell Värö pulp mill. An energy analysis has been performed with pinch technology and the potentials for steam savings and electricity production have been quantified for different levels of process integration.

It has been shown that the softwood Kraft pulp mill at Värö can be converted to a dissolving pulp mill by introducing extraction of hemicelluloses prior to the digestion step. The proposed dissolving pulp plant would not only be self sufficient energy wise but in fact be able to produce both more electricity and excess steam than today.

The most rigorous process integration proposed enable 71 MW electricity production as well as a surplus of 38 MW LP steam. A less rigorous option with lower investment cost can maximize the electricity production of the current back pressure turbine to 62 MW (from the current capacity at 51 MW), while leaving 28 MW of LP steam available for use elsewhere. The future saw mill and pellet factory, which is planned to be built close by, will require 20 MW of LP steam which has to be subtracted when assessing the steam surplus. Furthermore there might be an additional demand of LP steam for upgrading the extracted hemicellulose.

Keywords: biorefinery, pinch analysis, energy integration, process integration, water hydrolysis, hemicellulose, softwood, Kraft pulp mill

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Jon Bood & Linus Nilsson

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1. Background

Pulp and paper production has been a cornerstone in Swedish industry for many years. During the latest years however, pulp and paper demand has decreased and large new plants are being built, especially in developing countries. These factors lead to intense competitiveness within the pulp and paper industry. It has therefore become interesting to convert existing pulp mills into biorefineries with the possibility of producing alternative products, e.g. dissolving pulp, within the traditional lines of the mill. Another incentive for the biorefinery concept is the increased oil prices, making it profitable to extract side streams for further refinement into biomass-based speciality chemicals or materials, competing with or replacing the crude-oil based ones. (Lundberg, et al., 2012b) (Hämäläinen, et al., 2011) (Marinova, et al., 2010) (Mateos-Espejel, et al., 2010) (Shi, et al., 2011)

This thesis has been proposed in cooperation between the Department of Energy and Environment, Division of Heat and Power, at Chalmers University of Technology (Chalmers) and Södra Cell Värö (Värö), with the aim of investigating the possibilities for hemicellulose extraction within the existing plant at Värö. Hemicellulose extraction is used in order to be able to produce clean specialty cellulose, referred to as dissolving pulp, which then can be processed into viscose fibre.

1.1. Gap in research

The research within the area of hemicellulose extraction and dissolving pulp production is currently a hot topic. However, there seems to be a focus on converting hardwood mills to dissolving pulp mills, and mostly on the chemical and process wise difficulties of doing so. Not many articles or reports highlight the need for proper process integration to obtain a feasible energy situation for the dissolving pulp mill. This is of great importance since the required extraction step operates with large amounts of water at high temperatures (~175°C).

A few articles have been written on the subject, amongst others by Valeria Lundberg who focused at computer models of different pulp mills developed by the research institute Innventia AB, which were based on the model mills originally developed within the Future Resource-Adapted pulp Mill (FRAM) research programme.

This thesis is a theoretical case study of converting the actual mill at Värö from a softwood Kraft pulp mill to softwood dissolving pulp mill with focus on proper energy and process integration of the mill.

1.2. Aim

The aim of the thesis is to investigate the possibilities for extraction of hemicellulose and dissolving pulp production by adding an extraction step to the conventional fibre line at the Värö pulp mill. More specific, the thesis will investigate the potentials for energy integration in such a plant design and how to make it feasible from an energy perspective. To achieve this, both the existing and the new energy balance for the mill will be investigated. The possibilities for heat integration will be studied using pinch technology.

1.3. Specification of research topics

The specific parts to be investigated during the thesis are:

- Study the existing energy balance for Värö.
- Study the different alternatives to extract hemicellulose at a softwood Kraft pulp mill.
- Study the changes introduced to the energy balance by incorporation of hemicellulose extraction.
- Identify the potential energy efficiency measures using Pinch technology.

1.3.1. Limitations

The thesis will only cover investigations at Värö, and the results might therefore not be valid for other pulp mills.

No alternatives for the use or upgrading of the extracted hemicellulose will be investigated.

The environmental impact of the changes introduced when converting a Kraft pulp mill to a dissolving pulp mill will not be analysed or quantified.

The chemistry behind the hemicellulose extraction will not be investigated or optimized. Recommendations given are based on literature findings alone.

1.4. Method

The thesis will involve an energy study of the existing Kraft pulp mill at Värö, to get a good understanding of the situation at the mill today. Suitable process alternatives for hemicellulose extraction will be investigated and a new energy analysis will be performed with the new process operation unit incorporated.

The analysis will be performed using Pinch Technology (see *2.4. Pinch Technology*), highlighting possible alternatives for energy integration of the dissolving pulp mill and suggest some retrofits that could save energy. Many of these retrofits will also be valid for the current Kraft pulp mill.

2. Theory

The theory obtained from a literature study was the foundation for the future understanding and proceedings of the work. The literature review was conducted in order to gain relevant knowledge of the actual process at Värö, as well as different process alternatives to extract hemicellulose and how to integrate new process units into industrial plants.

2.1. Södra Cell Värö

Värö is a Kraft pulp mill also known as a sulphate pulp mill where the delignification of wood material is performed by mostly NaOH and Na_2S as cooking chemicals. The pulp mill at Värö, Figure 2.1 below, was the first paper pulp factory in the world to be transformed to totally chlorine free (TCF) bleaching of the pulp. Värö has a capacity of producing 425 000 ADt pulp annually in five different pulp categories, each suitable for different paper purposes. (Södra Cell Värö, 2010)

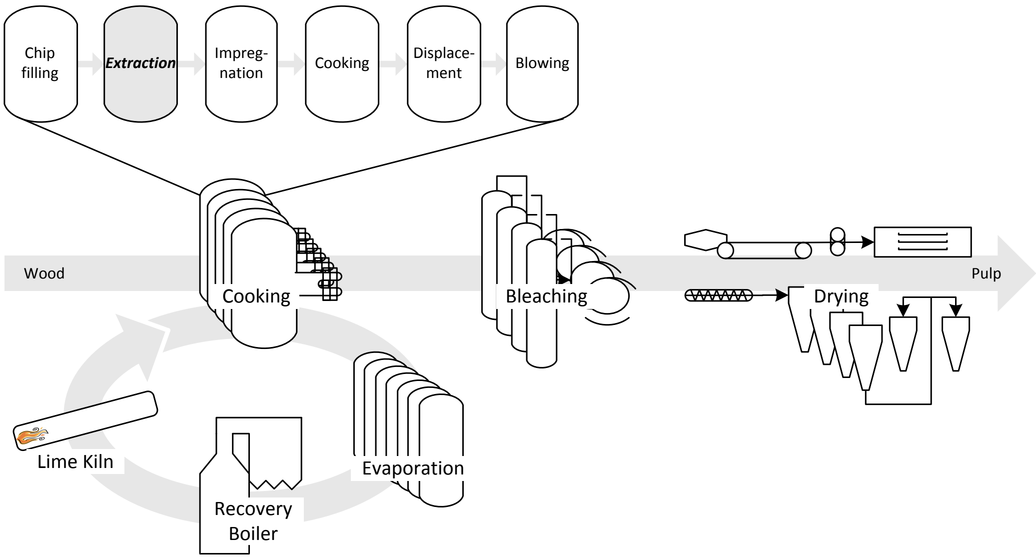


Figure 2.1. Overview of the mill at Södra Cell, Värö

The Värö pulp mill processes wood chips of uniform size distribution from spruce (70 %) and pine (30 %). The wood chips are cooked in digesters where they are subjected to high temperature and pressure as well as the cooking chemicals (called white liquor). This result in one liquid stream called black liquor as well as a dispersion of the solid wood fibres. The black liquor contains most of the lignin, some residues of degraded hemicellulose, cellulose and extractives as well as the used cooking chemicals. This liquor is then evaporated to produce thickened black liquor with high heat value, suitable for combustion in the plant's recovery boiler. In the recovery boiler the organic material is combusted for heat recovery, while the inorganics are recovered as a smelt taken out at the bottom of the combustion unit. The cooking chemicals are regenerated through a series of processes, going from green liquor after the smelt, to white liquor ready for use. (Lindsten, 2004) (Fritzon, 2002)

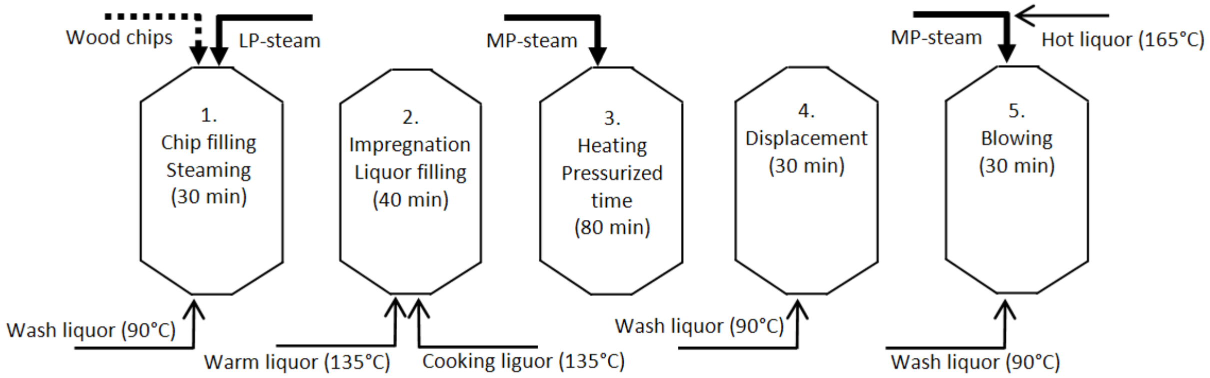


Figure 2.2. Schematic of the cooking cycle divided into five parts. The first four parts takes place in the same vessel while the last part, the blowing, takes place in a separate blowing vessel.

The Värö mill uses batch cooking in two digester lines with five digesters in each line. In Figure 2.2, the whole cooking cycle is displayed in five steps. The first four of these steps take place in the same vessel, the digester, while the final step, the blowing, takes place both in the digester and in a separate blowing vessel. In step one in Figure 2.2, the chips are transported from a storage bin to the digester where steam and wash liquor is used to displace all the air in the wood chips. In the second step of the cooking, the wood chips are first impregnated with warm liquor. The digester is then filled with the cooking liquor displacing all the black liquor which is sent to the evaporators. In step three the actual cooking takes place. The chips are heated up to the cooking temperature which is maintained through the whole cooking process. After the cooking, the pulp is washed with wash liquor; displacing all the liquor used during the cooking, step four in the figure above. The heat in the displaced liquor is recovered and then it is used for impregnation of the wood. Degassing valves are used to remove gases that are formed during the whole cooking process. In the last step, step five, the pulp is flashed in a blow tank, releasing steam at atmospheric pressure, before further treatment. (Fritzon, 2002)

In the evaporators, black liquor with a dry content of 16-17% is concentrated to a dry content of about 75%. This is done by heating the black liquor with steam in several effects. Steam condenses in the first effect evaporating water from the black liquor thus producing new steam at a lower temperature which is then used in the second effect. As the black liquor gets more concentrated the boiling point is elevated. This means that the pressure has to be decreased with each effect to compensate for the new saturation temperature at the following effect. The boiling point elevation and temperature difference across the evaporators are limiting factors, which mean that by using low pressure steam the number of effects is limited to between five and seven. According to Gustaf Collin[1], the plant at Värö rebuilt their evaporation system in 2009 and is now using seven effects. (Lindsten, 2004)

The plant at Värö has, during the latest years, undergone several improvement investments. The evaporation unit and the causticizing have a potential capacity increase of 86% and the back pressure turbine has a maximum capacity of producing 62 MW of electricity. On top of

[1] Gustaf Collin (Energy Coordinator, Södra Cell Värö) 25 Sep, 2012.

that the recovery boiler can quite easily be rebuilt to incorporate a 63% increase in capacity[2]. The potential capacity increase is listed in Table 2.1.

Table 2.1. Potential capacity increase for the different process parts at Värö.

Process equipment	Present use	Maximum capacity potential	Potential capacity increase
Evaporation	52 MW (LP steam)	96 MW	86 %
Causticizing	-*	-*	86 %
Recovery boiler	279 MW (HP steam)	455 MW	63 %
Turbine	51 MW	62 MW	22 %

* The present capacity use is unknown in the causticizing plant but the potential increase percentage is known.

At Värö there are predominantly three different water temperature classifications. These are defined as follows: Cold Water (CW) at the same temperature, 2°C, as the Viskan River, Warm Water (WW) at 47°C and lastly Hot Water (HW) at 65°C. Steam levels at the plant are defined as follows: low pressure (LP) steam at 4.5 bar, medium pressure (MP) steam at 11 bar and high pressure (HP) steam at 86 bar.

2.2. Hemicellulose Extraction

Hemicellulose is a group of polysaccharides that build up some of the organic material in wood, usually between 20-35% of the dry content depending on wood specie. (Henriksson, et al., 2005)

There are some variations in the type of hemicellulose found in different species. Coarsely classified between hardwood and softwood; hardwoods can be said to have a high content of xylans while softwoods have a high content of glucomannans. This means that the methods for hemicellulose extraction have to differ slightly depending on the wood species being processed at the plant under study. (Schild, et al., 2010)

For extraction of xylans from the hardwood specie Eucalyptus globules, alkaline or near-neutral extraction is proposed as good alternatives since the xylans are extracted in their oligomeric form and can be further processed for use. However, for softwood hemicelluloses like glucomannans, alkaline or near-neutral extraction cause rapid degradation of the polysaccharides due to the peeling reaction and are thus not suitable extraction methods. (Schild, et al., 2010)

One of the key issues is to extract the hemicellulose without degradation of the polymers, so that it can be used to produce high value by-products (e.g. films, plastics, chemicals or other materials) (Schild, et al., 2010). For softwood, extraction is preferably performed with water hydrolysis where most of the dissolved hemicellulose will be found in its oligomeric or polymeric form in the hydrolysate (the stream containing the extracted hemicellulose)

[2] Gustaf Collin (Energy Coordinator, Södra Cell Värö) 25 Sep, 2012.

(Mateos-Espejel, et al., 2012). On the other hand if the hydrolysate is to be burned in the recovery boiler, then steam hydrolysis would be preferable (Mateos-Espejel, et al., 2012). According to the models supplied by Innventia (Innventia AB, 2013) the water hydrolysis step takes place with 4:1 water to wood ratio at a temperature of 175°C.

Due to formation of acetic acid the water hydrolysis is somewhat acidic, with a pH between 3 and 4, resulting in a need for a neutralisation step before the actual cooking procedure (Mateos-Espejel, et al., 2012). The drawback of this method is that there might be a formation of sticky precipitates from dissolved lignin, which can be somewhat problematic for large-scale industrial processes. It is described in literature that this problem can be reduced slightly by lowering the extraction temperature and prolonging the extraction time or alternatively adding a catalyst to the mixture (Schild, et al., 2010). Another way of reducing the lignin precipitates is to keep the pressure elevated during the whole hydrolysis step and not to release it when draining the hydrolysate from the wood chips (Mateos-Espejel, et al., 2012).

Another way of extracting hemicellulose is by using steam hydrolysis followed by a neutralization step. This method limits the sticky precipitates but degrades the carbohydrates of the hemicelluloses making them hard to refine. The practice has thus been to send that stream, together with the dissolved lignin in the black liquor, for combustion in the recovery boiler. This however would increase the load of the recovery boiler, which according to Leschinsky et al. might limit the amount of raw material that can be processed. (Leschinsky, et al., 2009) Ideally, the aim would be to use the hemicellulose-containing hydrolysate stream for purification and upgrade it to value-added by-products, making a transition into a forest based biorefinery.

The extraction of hemicellulose would imply changes in both the steam and electricity balance of the mill due to changes in the pulp production capacity and/or altered material composition of the black liquor, moreover due to energy demand for the extraction step. Pulp and paper mills often are limited in their production and expansion by the capacity of the recovery boiler (Lundberg, et al., 2012a). This however seems not to be the case at Värö, since the recovery boiler can be rebuilt to facilitate extra capacity.

2.3. Dissolving pulp

Dissolving pulp is a pure cellulose product, with a maximum hemicellulose content of 3% and uniform polymer size distribution (Mateos-Espejel, et al., 2012). To be able to produce this in a Kraft pulp mill the best method is to remove the hemicelluloses in a hydrolysis step and then do extra bleaching to remove the last fractions of lignin in the pulp (Gellerstedt, 2005).

By removing the hemicellulose from the wood chips, the yield of pulp in the mill is decreased from 43% in a conventional Kraft pulp mill to 33% with a water hydrolysis step, producing dissolving pulp (Delin, et al., 2005) (Innventia AB, 2013). Depending on the bottleneck of the pulp mill under study, different options have to be considered in order to maximize the dissolving pulp production to be able to turn a profit from the plant conversion.

Dissolving pulp was traditionally produced in sulphite pulp mills. This pulping method was favoured since the cooking chemicals dissolve both lignin and hemicellulose, resulting in a relatively pure stream of cellulose even prior to additional cleaning. However, in a sulphate mill, like Södra Cell Värö, the cooking chemicals mostly delignify the wood material, resulting in a pulp with a high amount of the hemicellulose left. (Köpcke, et al., 2010)

To be able to redesign a pulp mill from producing Kraft pulp to dissolving pulp, it is (for reasons mentioned above) required to extract the hemicellulose prior to the digestion process. The hemicellulose extraction affects the energy balance of the whole mill as the composition of the material streams is changed and a lot of hot water and energy is needed for the water hydrolysis.

Dissolving pulp (or dissolved cellulose) can be used as raw material for different products. Some of the applications are as starting material for Viscose and LyoCell as well as for production of cellulose derivatives e.g. Carboxymethyl Cellulose (CMC) which can be used as a thickening and dispersion agent in food and cosmetic products. (Gellerstedt, 2005)

During the latest years the world wide demand for dissolving pulp has been growing by 10% annually and many factories are being built or converted to meet this new demand. The long term price is according to Macdonald expected to level out at approximately 1200 US$ per tonne, as compared to 819 US$[3] per tonne for Kraft pulp. However the price estimation by Macdonald seems a bit overoptimistic and according to Södra a more updated estimation predicts the price to around 1000 US$ per tonne[4]. This means that to be able to profit from a dissolving pulp production conversion the key is to maintain or increase the pulp production rate of the plant and to be able to produce high-value products from the extracted hemicellulose. (Macdonald, 2011)

2.4. Pinch Technology

An energy balance for such a large system as Värö is extremely complex and therefore a special methodology, called pinch technology, is used for the analysis (Harvey, 2011). Pinch technology can be used to determine:

- How much heat must minimally be added to the process?
- How much heat must minimally be cooled externally?
- How much heat can be recovered internally?
- How should the heat exchanger network be designed to maximize the heat recovery?

A Pinch analysis starts by gathering stream data and classifying each stream either as a cold or hot stream. A *cold stream* is one that needs to be heated and a *hot stream* is one that needs to be cooled. The minimum temperature difference (ΔT_{min}) in each heat exchanger (HX) is determined by the state and composition of the streams exchanging heat. A lower value gives the ability to utilize more heat while a higher value gives smaller heat exchangers due to the increased driving force for heat transfer. The trade-off thereby lies within the cost of energy versus cost of extra heat exchanger area.

[3] FOEX Indexes Ltd. (www.foex.fi) for Northern Bleached Softwood Kraft Pulp. 29 Jan, 2013.

[4] Maria Edberg (Senior Project Manager, Södra Innovation) 18 Feb, 2013.

All the thermal streams are presented in a table with initial and target temperature as well as energy demand or excess. The streams are then graphically displayed as a complete system in a composite curve (CC) or a grand composite curve (GCC). From these curves it is fairly easy to extract information regarding how much energy that can be heat exchanged internally and how high the external heating and cooling demands are. The curve also shows the pinch temperature (pinch) which is of great importance when the heat exchanger network (HEN) is to be designed.

Designing a HEN with a lot of different streams can be difficult and it gets even harder when an existing HEN is to be improved, i.e. retrofitted. Three important rules in pinch technology describe how it should be designed to reach maximum energy recovery (MER). Breaking any of these rules, bulleted below, is what is referred to as a “Pinch violation”.

- Do not transfer heat through the pinch, i.e., from a hot stream above the pinch to a cold stream below the pinch.
- Do not use external cooling above the pinch.
- Do not use external heating below the pinch.

The HEN is designed separately for the part above and the part below the pinch. Above the pinch the main goal is to cool all the hot streams down to pinch temperature without using any cold utility (external cooler etc.). Below the pinch the goal is to heat all the cold streams to pinch temperature without using any hot utility (external heaters etc.). The easiest way to do this is to match each hot stream above the pinch with a cold stream that cools the hot stream completely. The same thing should be done below the pinch in order to achieve a MER network design.

It can however be very difficult to create a MER when retrofitting an existing HEN and it would not be economically feasible to do so. Some streams cannot be heat exchanged economically since the physical distance between them is too large. This is often dealt with by dividing the big factory system into smaller parts where heat exchanging between the streams is possible. In order to find a reasonable retrofit from an economical point of view, different rules of thumb can be used:

- Retain as many existing heat exchangers at original positions as possible.
- Install as few new heat exchangers as possible.
- Install no more new pipe than necessary.

The process integration potential of two separate systems can be evaluated by doing a so called foreground/background analysis. The process to be heat integrated, the foreground system, is fitted into the existing GCC of the main process, the background system. By doing this the total energy demand of the fully integrated, i.e. maximum integration possible between the systems, process can be visualized.

3. Methodology

During the start of the project the main focus was to get a better understanding of the process adopted at Värö to better be able to evaluate the potential to extract hemicellulose. Three other master theses have been performed at Värö earlier; 2002 the digester was modelled by Anna Fritzon and 2004 the evaporators were modelled by Henrik Lindsten. Although several years have passed and much has changed at the plant, these two reports still gave a basic understanding of how the plant is designed. Another thesis by Emma Sundin, with focus on the Air drier, was conducted at Värö in 2011.

The plant at Värö was investigated in its present design to get an understanding of the current energy balance at the plant. Later, the energy balance for the new design proposition including the unit operation for hemicellulose extraction was investigated. A pinch analysis was done for the whole plant to identify options for retrofitting and the possibility of integrating the hemicelluloses extraction step with the rest of the plant.

3.1. Data gathering at Värö

The stream data in the theses works dated 2002 and 2004 was used to obtain basic understanding during the beginning of the effort to identify hot and cold streams at Värö. No stream data was however kept from previous theses works since several streams where removed or added as the plant has been modified during the last ten years. All data for the streams was collected for the time period between the 14th and 18th of January 2012 when the production was stable and the outdoor weather conditions was deemed representable for a year round production.

Stream data was collected using the internal control system with the help of the contact people at Värö to make sure that the values were representative. The plant was also studied from detailed construction drawings of the process, to make sure that no streams were missed and that all the measured data was collected from the correct places and sensors at the plant.

It was not possible to find data for all the streams in the information system and some temperatures and flows have therefore been approximated. In some cases it was not possible to find data for the above stated time period and in those cases a more recent time period, between the 4^{th} and 8^{th} November 2012, has been chosen. All the streams and measurements that in some way were affected by this are listed and explained in detail in *Appendix 2*.

The bark boiler is usually only run for approximately three months a year and has not been accounted for in the energy analysis since it is not a part of the normal production.

In the near future the saw mill capacity will increase at Värö which means that the heat demand for this part of the process will increase. Connected to the saw mill, a pellet factory will be built which would further increase the heat demand. These future increases in heat demand have, however, not been taken into account when making the GCC. Instead the possibility to deliver excess heat was investigated and evaluated for addressing the future demand.

3.2. Hemicellulose extraction

Water hydrolysis was chosen for the hemicellulose extraction due to the versatility of the process. It can be used either if the hemicellulose is to be upgraded or simply evaporated and burned in the recovery boiler for additional electricity production.

There are two possible process alternatives for hemicellulose extraction by water hydrolysis; continuous or batch extraction. The best option in this case is probably batch extraction since the rest of the digestion is taking place in a modified super-batch process. The idea is to do the extraction as a preceding step in the different digester vessels already in place. This would probably make the transition to dissolving pulp production easier to implement. Extraction of hemicellulose can, according to Mateos-Espejel et al, be performed in a displacement batch process which is very similar to the one at Värö. In the same article it is argued that the residence time in each digester would increase from 200 minutes for Kraft cooking to 269 minutes with water hydrolysis included, which will put limits on how much raw material that can be processed. The actual cooking cycle at Värö takes 210 minutes but because it is hard to know how the extraction time affects the cooking time, 200 minutes has been used as proposed in literature. (Mateos-Espejel, et al., 2012)

3.2.1. Increased digester capacity

The residence time increase in each digester is assumed from the work of Mateos-Espejel and is set to a total of 269 minutes for dissolving pulp production with water hydrolysis as compared to 200 minutes for regular Kraft pulp production (Mateos-Espejel, et al., 2012). This leads to an increase in residence time of about 35%.

The increase of wood intake, to compensate for the inferior yield in the dissolving pulp production case, is calculated from the FRAM models to an approximate of 31%. This leads to the conclusion that the total digester capacity needs to increase with 77%, resulting in a total demand of 18 digesters instead of the ten that is used today. By adding one extra digester to each line as well as adding one additional line with six digesters, this would be accomplished.

Provided that the digesters operate at full capacity, the mean time between each digester sequence start was 40 minutes during the studied time period. By adding digesters when converting to dissolving pulp production as discussed above, this will change to 45 minutes. This fact is important since each line of digesters is connected by the same liquor handling pump system and if this time is shortened or increased this might lead to problems with the displacement sequences. If modifying the two existing lines is impossible, another alternative could be to add all the extra capacity needed in a third line.

3.3. Streams in the dissolving pulp plant

The dissolving pulp plant is assumed to look the same as the current plant at Värö with the exception that there is a hemicellulose extraction facility before the digestion plant.

From the computer model of a conventional Kraft pulp mill and a model for a dissolving pulp mill, a comparison has been made on how the conversion affects streams within the production line. The streams from the Kraft plant at Värö have been scaled up in accordance with the comparison, listed in *Appendix 5*.

Because of the fact that the yield of dissolving pulp production is lower than conventional Kraft pulp production, the raw material intake will have to increase if the production rate is to be kept constant. If constant pulp production is to be maintained in the dissolving pulp plant, it results in 31% increase of wood intake. This increases the load on the digesters that not only have to incorporate the increased amount of wood but also an increased residence time due to the hemicellulose extraction step that has to be carried out before the actual cooking. The capacity in wood handling, debarking and chipping would also have to be increased.

After the digestion step, the amount of pulp to the rest of the fibre line remains unchanged with a constant pulp production rate. This means that the streams in the screening, bleaching and drying facilities are assumed unchanged, leaving neither any unused capacity nor any need to upgrade these parts of the plant. However it is possible that some of the bleaching chemicals need to be changed or increased due to the higher purity demands i.e. higher lignin removal, but this will most likely not affect any thermal streams in the energy analysis of the plant. (Innventia AB, 2013)

The digesters has a demand of MP steam, for heating of the pulp and liquors, which have been calculated from the total volume of the digester and the displacement procedures for warm, hot and white liquor, respectively. The temperature after the final stage in the cooking sequence was higher in the dissolving pulp case and thus the MP steam demand in each individual digester decreased. On the other hand, when taking into account the additional digester capacity, the total MP steam demand increased by 44%.

The additional white liquor needed, according to the models, leads to an increased load in the causticizing plant by approximately 45%.

The evaporators have a capacity high enough to handle the increase of thin black liquor to evaporation, which is assumed to arrive there at unchanged dry content compared to the Kraft production plant. There is even unused capacity in the evaporation system, which opens possibilities for even greater production and, if no better alternative is found, even to evaporate the hydrolysate stream for combustion in the recovery boiler. However the latest alternative might lead to an extensive capacity increase in the recovery boiler which is not possible and would then limit the whole production capacity of the plant. Neither would it lead to any further electricity production within the existing turbine, since its capacity limit is already met in the proposed dissolving pulp plant.

3.4. Methods and models

3.4.1. Pro-Pi (Pinch analysis software)

The energy balances and Pinch analysis was carried out in Microsoft Excel, using the add-in ProPi which was developed by Chalmers Industriteknik AB. The GCC and other important features of the energy optimization were modelled in the same software and provided the necessary information for a HEN design and some retrofit ideas.

3.4.2. Models used

The models used for approximating the changes in process streams when a Kraft pulp mill is converted into a dissolving pulp mill have been obtained as stream flowsheets from computer models developed by Innventia. (Delin, et al., 2005) (Innventia AB, 2013)

3.4.3. Värö mill schematics

The energy analysis of the present Kraft pulp mill at Värö was mapped and evaluated using extensive flowsheet documentation of all the streams, valves and indicators at the plant. This, was then compared to the streams available in the local information and control program used at the site, first to confirm streams and get a better understanding of how and when the different stream were used in the process and secondly, to extract data for the time period chosen.

4. Results

This section starts by presenting some key data for both the existing Kraft pulp mill at Värö as well as the proposed dissolving pulp plant. The results of the energy analyses are presented both in form of GCCs and lists of the occurring pinch violations. Some retrofit suggestions are presented as well as a few different cases in which hemicellulose extraction could be implemented. Lastly the energy production potentials and possible uses for excess energy are presented.

All the calculations and comparisons made for the Kraft pulp mill and its conversion to a dissolving pulp mill are performed with the assumption that the pulp production can be kept constant despite the inferior pulp yield, by increasing the wood input to the mill.

4.1. Key data

In Table 4.1 below, the steam and electricity production is presented. The normal operating mode which was investigated includes running the Recovery Boiler (RB) at full load during a somewhat mild winter period while not running the Bark Boiler (BB) at all. All the additional steam that theoretically could be produced in the BB is therefore not included in the analysis, but could be produced if necessary to compensate for seasonal and production variations.

Table 4.1. Key data for the original Värö pulp mill and the theoretical dissolving pulp mill

Key data	Kraft pulp plant	Dissolving pulp plant
Pulp production	1 227 ADt/d	1 227 ADt/d
Raw material intake	2 824 ADt/d	3 712 ADt/d
Yield	43%*	33%
LP steam production, RB	161 MW	202 MW
MP steam production, RB	50 MW	85 MW
Electricity production	51 MW	62 MW
District heating, Varberg	29 MW (10 MW by LP steam)	29 MW (0 MW by LP steam)
HP Steam production capacity, Bark Boiler	30 MW	30 MW
Water consumption	17 ton/ADt	26 ton/ADt

* This is the pulp yield of the Kraft mill model and not the actual Värö mill. The actual yield at Värö is approximately 47-48%, however due to the fixed model data this figure was not possible to use for the comparison between the mill types.

The electricity production will increase when converting to dissolving pulp production. This is due to the increased flow of black liquor and therefore also steam production in the recovery boiler. The turbine efficiency at Värö, 88%, was calculated from the current electricity production. Depending on how the transition to dissolving pulp production is performed, the steam and electricity balance will change. This is further evaluated in section *4.7.3. Steam and electricity production*.

4.2. Grand composite curves

The GCCs are an excellent tool to visualize the total energy balance of a system. It shows the theoretically minimum energy consumption making it easy to calculate the potential for energy savings.

The GCCs are calculated with individual ΔT_{min}, which can be found in *Appendix 3*, depending on the heat transfer properties of individual streams.

4.2.1. Kraft pulp plant

The GCC of Södra Cell Värö is presented below in Figure 4.1. The minimum hot and cold utility for the process are 157 MW and 60 MW respectively. The pinch temperature of the process is 101°C.

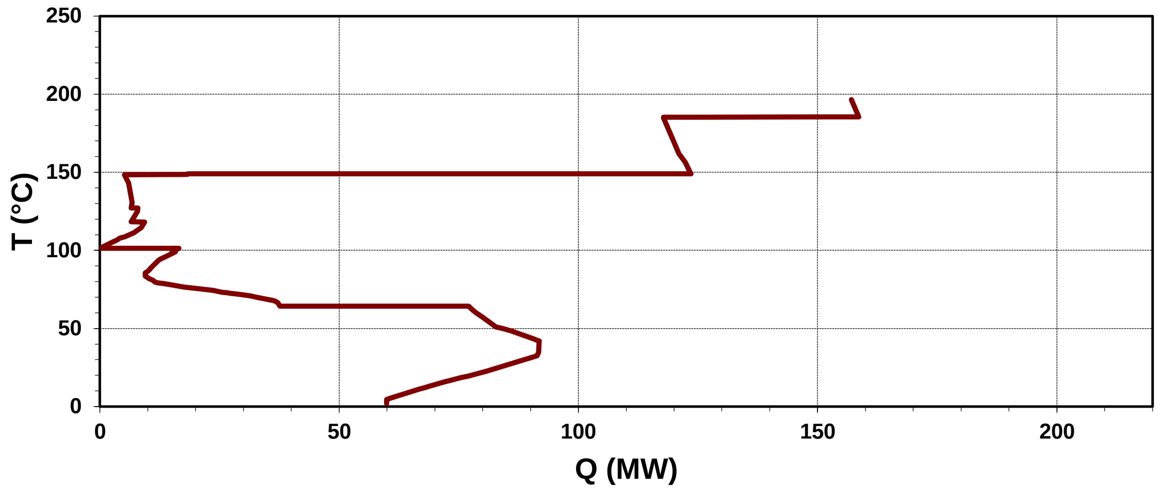


Figure 4.1. GCC of the Kraft pulp mill at Värö.

The total steam production at the Kraft pulp plant is currently 211 MW (161 MW LP and 50 MW MP), see Table 4.1 above. The total steam consumption during the same period was 192 MW. The gap between production rate and consumption of steam is known at Värö and some deviations were expected. The deviation during normal conditions is usually around 5%, but was slightly higher during the studied time period, see *5.4. Uncertainties and fluctuations*.

To identify the total amount of pinch violations, the steam demand of the actual plant is compared to the minimum hot utility demand in the GCC. In the plant studied, the calculated steam consumption does not add up to the steam production at the plant. This means that there are either unknown steam consumers or losses of steam which cannot be quantified in this study. Due to the uncertainties regarding the actual steam demand of the studied plant, a minimum and maximum total pinch violation value can be approximated. To obtain values comparable to the GCC, the effect in the dumping condenser (4.8 MW) and the steam consumption for district heating (DH) to Varberg (9.6 MW) must be subtracted from the steam production and demand of the plant (since these streams are not included in the GCC). This results in a total steam production of 197 MW and a demand of 178 MW. This difference is referred to as unknown steam consumers in Table 4.6 and Table 4.7, and is discussed in *5.4.*

Uncertainties and fluctuations. The comparison is made both between the steam consumption and demand against the minimum hot utility demand in the GCC, resulting in an interval of Pinch violations for Värö between 20.5 MW and 40 MW, which corresponds to approximately 10% of the total steam production. The stream data sheet for Värö pulp mill which is the foundation for the GCC can be found in *Appendix 4*.

4.2.2. Dissolving pulp plant

Below, in Figure 4.2, the GCC for the dissolving pulp plant is presented. The minimum hot utility is 211 MW which is an increase of 34% compared to the Kraft case. This compared with the total steam production in the recovery boiler results in 76 MW of pinch violations which limits the complete process integration potential to the same amount. The minimum cold utility demand is 58 MW which is actually a slight decrease when compared to the Kraft case, probably due to the fact that the amount of wood and flow rate of CW needing heat has increased more in proportion to the rest of the plant. The pinch temperature is still 101.3°C.

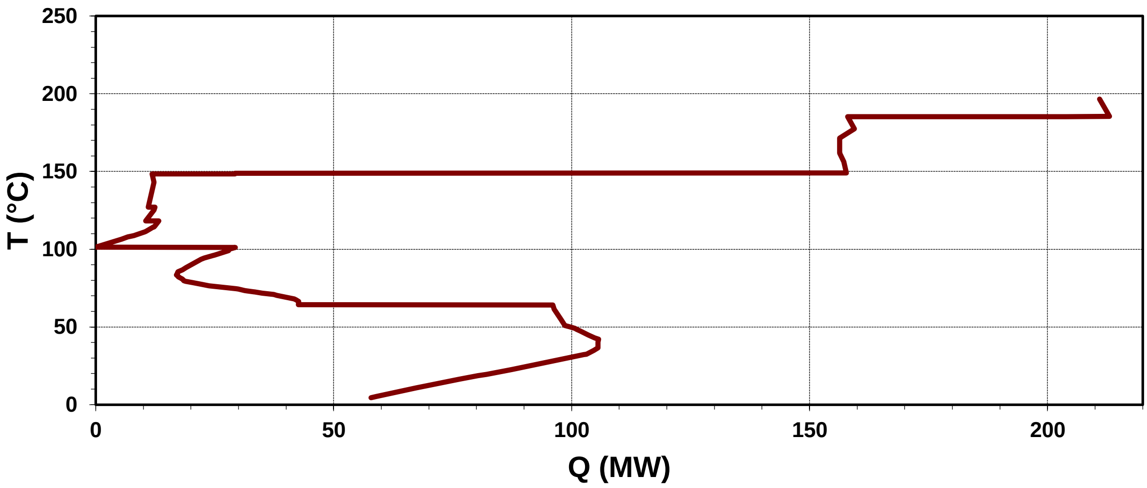


Figure 4.2. GCC of the theoretical dissolving pulp mill.

The GCC of the dissolving pulp plant is very much alike the one for the Kraft pulp plant, which is no surprise since only three new streams have been introduced. The major difference is that some of the streams have been scaled up, resulting in for instance more excess heat in the blowing steam condenser and the surface condenser, the two horizontal lines at 103°C and 66°C.

4.3. Pinch violations

Below, in Table 4.2, is a list of pinch violations, occurring both in the Kraft and dissolving production cases. It includes the size of the stream and with how much it violates the pinch rules for both cases. In the dissolving case some streams are scaled up according to scaling factors in *Appendix 5*.

Table 4.2. Identified pinch violations of the Kraft and dissolving pulp mill

Pinch violation	Kraft [MW]	Dissolving [MW]	Comments
Heat exchanging of cooking chemicals	0.7	1.2	Heat is transferred through the pinch from hot liquor to white liquor.
Production of HW with LP steam	3.1	3.1	HW produced in digestion and causticizing by LP steam. Heating below the pinch.
Heat exchanging between liquor condensate at stripper	2.2	3.0	Heat is transferred through the pinch from Clean condensate to Mixed and Dirty condensate.
Heating of air to Paper room with MP steam	0.5	0.5	Heating of air below the pinch.
Heating of HW with flash steam	2.8	2.8	Flash steam above the pinch is used to heat water below. Water was previously used in the HW box at the wire.
Heating of air to cyclone driers with LP & MP steam[5]	> 6.8[4]	> 6.8[4]	Heating of air below the pinch.
Heating of office facilities with LP steam	0.2	0.2	Heating below the pinch.
Pre-heating of feed water with flue gases	6.7	9.0	Hot flue gasses above the pinch heating feed water below the pinch.
Heating of air to bark drier with LP steam	3.1	4.1	Heating of air below the pinch.
Total pinch violation	24.5	29.0	

The pinch violations that were found at Värö are described in more detail in *Appendix 6*.

4.3.1. Pinch violations of the cyclone drier

The pinch violation in the cyclone drier is very uncertain due to lack of measurement data for the air flow rates.

To calculate the pinch violation, it was assumed that all the LP and MP steam was supplied to the high temperature HX, heating cold air from 4°C to 148°C. It was also assumed that the low temperature HX only used hot air from the air drier as heating media. From these assumptions a pinch violation of the air heated up to the pinch temperature was calculated. If the assumptions are false, the pinch violation will increase; hence it is a minimum pinch violation, resulting in 6.8 MW.

4.4. Retrofit suggestions

The retrofits suggested for the Hot and Warm Water System (HWWS) are based on producing HW from WW instead of CW as is being done today. None of them are in fact pinch violations but the heat can be used more efficiently thus eliminating other pinch violations in which steam is used. The additional WW needed for the suggested retrofits can be produced by heating CW with a large effluent stream from the bleaching plant, which is today cooled by cooling towers.

If some of the pinch violations are corrected as in the following retrofit proposal, 13 MW steam could be saved in the existing Kraft pulp plant. This means an additional 12.5 MW of LP steam that can be used for other purposes, while an additional 0.5 MW of MP steam could be run through the turbine, slightly increasing the electricity production.

[5] See further comments for this minimum pinch violation in *Appendix 6 – Heating of air to cyclone driers*.

Table 4.3 shows a list of retrofit suggestions for the Kraft pulp mill at Värö. For more details regarding the retrofit proposals, see *Appendix 7*.

Table 4.3. Retrofit suggestions for the Kraft pulp mill.

Pinch violation	Effect [MW]	Retrofit suggestion	Solved effect [MW]	Comments
Heating of HW at digestion plant	2.2	HWWS revamp	2.2	Heating WW instead of CW to produce HW in the Acc0 cooler.
Pre-heating of feed water with flue gases	6.7	HWWS revamp	6.4	Green liquor cooler heats WW instead of CW which makes heat from the Mist condenser free for use here.
Heating of HW at causticizing plant	0.9	HWWS revamp	0.9	Heating WW instead of CW to produce 82°C water from the washing.
Heating of office facilities with LP steam	0.2	HWWS revamp	0.2	Heating WW instead of CW to produce 82°C water from the washing.
Heating of HW with flash steam	2.8	Air drier	2.8	Use the flash steam to heat air into the air drier. Replaces LP or MP steam.
Heating of air to paper room with MP steam	0.5	Air drier	0.5	Replace steam by heat exchanging with outgoing air.
Pinch violations	13.3	Total solved	13.0	

In the Kraft case approximately 13 MW of the 25 MW of pinch violations could be solved quite easily. This equals to 53% of the pinch violations solved.

In Table 4.4, a list of retrofit suggestions for the dissolving pulp plant is summarized. Since the dissolving pulp plant is fictive, the easily corrected pinch violations are the same as in the existing Kraft pulp plant, with the difference that the size of some streams are bound to change. Despite the change in size the retrofit suggestion is very similar to the existing Kraft pulp plant. In the dissolving pulp plant, 52% of the pinch violations were corrected and the steam savings adds up to 15 MW of steam ready for use elsewhere.

Table 4.4. Retrofit suggestions for the Dissolving pulp mill

Pinch violation	Effect [MW]	Retrofit suggestion	Solved effect [MW]	Comments
Heating of HW by LP steam at digestion plant	2.2	HWWS revamp	2.2	Heating WW instead of CW to produce HW in the Acc0 cooler to remove steam demand.
Pre-heating of feed water with flue gases	9.0	HWWS revamp	8.5	Green liquor cooler heats WW instead of CW which makes heat from the Mist condenser free for use here.
Heating of HW by LP steam at causticizing plant	0.9	HWWS revamp	0.9	Heating WW instead of CW to produce 82°C water from the washing.
Heating of office facilities with LP steam	0.2	HWWS revamp	0.2	Heating WW instead of CW to produce 82°C water from the washing.
Heating of HW with flash steam	2.8	Air drier	2.8	Use the flash steam to heat air into the air drier. Replaces LP or MP steam.
Heating of air to paper room with MP steam	0.5	Air drier	0.5	Replace steam by heat exchanging with outgoing air.
Pinch violations	15.6	Total solved	15.1	

4.5. Hemicellulose extraction

By performing a background/foreground analysis, see Figure 4.3, the process integration potential of the hemicellulose extraction is 26 MW. This is a measure of how much energy that maximally can be recovered by heat integration of the extraction step with the rest of the plant. The minimum hot utility needed for the extraction would in such a case be 17 MW.

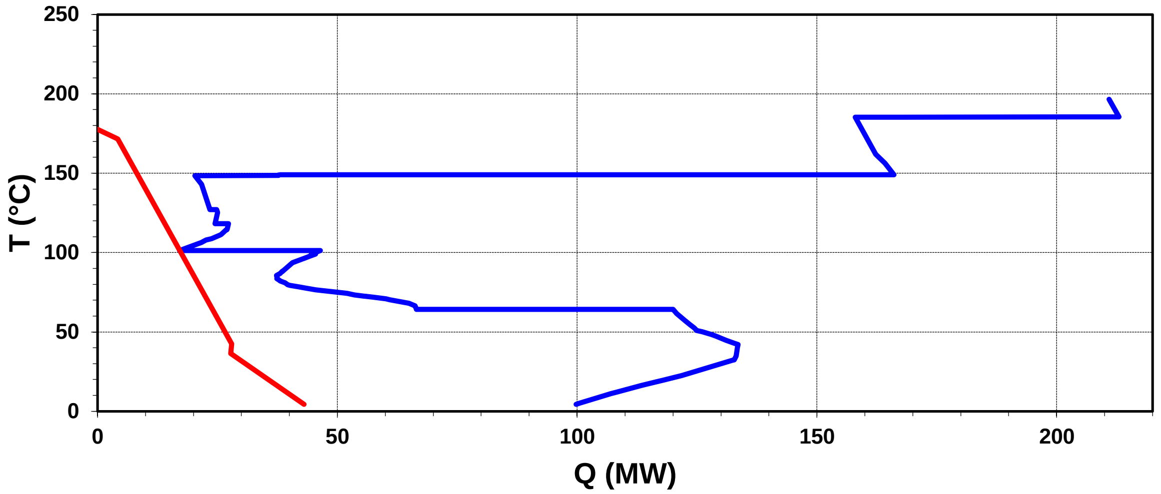


Figure 4.3. Background/foreground curve showing total process integration potential of hemicellulose extraction.

Assuming that all the proposed retrofits are implemented when converting to dissolving pulp production, four cases for integration of the hemicellulose extraction unit with the rest of the plant are considered.

The hydrolysate stream is assumed to be cooled to from 170°C to 40°C, which however depends on the desired use of the hemicellulose. If it is to be sent to evaporation it should probably have the same temperature as the black liquor stream of >70°C. For fermentation, biogas production, biological waste water treatment or any other biological processing a temperature around 40°C would be preferable. The target temperature of the hydrolysate stream cooling controls which retrofit and process integration idea that would be most beneficial.

The four cases to be investigated are presented below:

Case 1: The extraction plant is not heat integrated at all.

Case 2: The extraction plant is internally integrated (i.e. hydrolysate is heat exchanged with ingoing water).

Case 3: The extraction plant is integrated with the rest of the plant.

Case 4: The extraction plant is rigorously integrated with the rest of the plant.

The steam demands of the four investigated cases are presented in Table 4.5 below. Each case is discussed and presented further in the following subsections.

Table 4.5. Heating demand of extraction water

Location	Dissolving, Case 1 [MW]	Dissolving, Case 2 [MW]	Dissolving, Case 3 [MW]	Dissolving, Case 4 [MW]
Extraction water heating by MP steam	20.6	20.6	20.6	20.6
Extraction water heating by LP steam	92.1	22.4	9.9	5.0
Extraction water heating by heat recovery measures	0	69.6	82.2	87.0
Total heating demand	112.7	112.7	112.7	112.7

4.5.1. Case 1 – no heat integration

In this case the only energy saving measures is from the retrofit suggestions presented earlier. All the ingoing extraction water is heated with steam and all outgoing hydrolysate is cooled with cooling water being discarded. Case 1 is obviously not a relevant case in practice, since it is clearly a very inefficient way to do the heating and cooling in this unit. However, in this thesis it is used to show the importance of proper heat integration of the hemicellulose extraction unit. The total energy demand of Case 1 is calculated to 113 MW, consisting of 21 MW of MP steam and 92 MW of LP steam.

4.5.2. Case 2 – internal integration within the extraction unit

By installing a heat exchanger that is heating the ingoing extraction water with the outgoing hydrolysate, a lot of heat is recovered internally and a significant amount of steam can be saved. The MP steam remains constant to reach the high temperature of the extraction water but the heat integration saves 70 MW of LP steam in one heat exchanger. To be able to do this heat integration in a batch extraction, an accumulator or storage tank of some sort is needed since the ingoing water is heated with outgoing hydrolysate.

4.5.3. Case 3 – extraction unit is integrated with the whole plant

The internal heat recovery to the heat extraction water can be made even more efficient by process integrating with the rest of the mill. It demands more investment and more piping but the energy savings compared to Case 2 might still make it worth the investment costs.

From the comparison between the Kraft and dissolving pulp mill models it can be seen that the WW demand of the two are equal. In the surface condenser the capacity increases by 36% which results in an increased WW production of 14 MW. Due to temperature limits for heat exchanging with the extraction water stream, 13 MW can be used for heating.

By using this heat to preheat the extraction water, a temperature of 28°C can be reached. After mixing this water with the wood, the temperature of the wood/liquid mix becomes approximately 32°C. By heat exchanging with the outgoing hydrolysate stream, cooling it from 175°C to 40°C, the temperature of the extraction water and wood mix increases to 131°C, leaving an additional 31 MW to be heated by LP and MP steam. Heating is done as far as possible with LP steam, reaching a temperature of the extraction water and wood of 146°C.

From that temperature MP steam is used to reach the final temperature of 175°C. The heating needed becomes; 10 MW LP steam and 21 MW MP steam.

This means that an additional 13 MW of LP steam can be saved compared to Case 2. An economic analysis would have to be made in order to decide which of these two options that is considered most beneficial.

4.5.4. Case 4 – improved process integration

An effort was made to further improve the performance of the process integration, by a little more complex heat integration with the rest of the mill.

The effluent stream from the bleaching plant (V1) is, at Värö, currently used for production of DH to Varberg and the residual heat is needed to perform the suggested retrofits in *4.4. Retrofit suggestions*. This stream has a substantially high flow rate and is at suitable temperature to preheat the extraction water. In order to be able to increase the temperature of the extraction water as much as possible before having to use steam, the heat supplied to DH by V1 has to be decreased. This is done by using heat from the blowing steam condenser, resulting in an increase of the V1 temperature from 54°C to 58°C.

The V1 stream is split to match the F*Cp value of the extraction water and wood mixture. The extraction water is heated with warm water and mixed with the wood chips, reaching 32°C as in Case 3. The wood chip/water mix is heated to a temperature of 52°C with the V1 stream, transferring totally 14 MW heat. The hydrolysate stream can then heat the mixture to a temperature of 138°C which results in a final LP steam demand of 5 MW.

The part of the V1 stream that was split off, together with remaining heat in the hydrolysate stream can be used to produce the 12 MW of WW needed to make the proposed retrofits possible.

This case of course demands more heat exchangers and more piping as well as being more complex to implement. However the final LP steam savings add up to 5 MW compared to Case 3 and up to 17 MW compared to not doing any process integration with the mill at all in Case 2.

4.6. Steam balance

From the models, it was shown that the flow rate of black liquor into the recovery boiler increased by 36% but on the other hand, the heat value of the black liquor (calculated as amount of steam produced in MW per kg of black liquor to the boiler) decreased slightly. With these scaling factors included, the HP steam production would be increased by 33%, from 279 MW to 371 MW.

In Table 4.6, an extensive list of the MP steam users in the plant is presented. Of all the MP steam users, approximately 15% is unknown. Some of this can possibly be accounted to letdown valves or disturbances in the measuring equipment. The unidentified steam demand in the existing Kraft pulp mill is included in the MP steam users for the dissolving pulp plant, to eliminate the risk of subtracting any real steam losses that might have been left out from the data collection.

Table 4.6. MP steam balance for the original Kraft pulp mill (Värö) and the different cases.

Location	Kraft w/o retrofit [MW]	Dissolving, Case 1 [MW]	Dissolving, Case 2 [MW]	Dissolving, Case 3 [MW]	Dissolving, Case 4 [MW]
Extraction water heating	-	20.6	20.6	20.6	20.6
Digester heating and wood chip transportation/digester draining	20.2	29.1	29.1	29.1	29.1
Bleaching step 2	1.5	1.5	1.5	1.5	1.5
Bleaching step 4	3.1	3.1	3.1	3.1	3.1
Heating of paper room facilities	0.5	-	-	-	-
Cyclone drier (air heating)	1.2	1.2	1.2	1.2	1.2
Feed water pre-heating	4.3	5.8	5.8	5.8	5.8
Recovery boiler	11.8	15.7	15.7	15.7	15.7
MP steam demand by known consumers	42.5	77.4	77.4	77.4	77.4
Unknown steam consumers	7.5	7.5	7.5	7.5	7.5
Total MP steam production	50.0	84.8	84.8	84.8	84.8

Since the heating of extraction water to its top temperature (i.e. 175°C) is done with the same amount of MP steam in all four cases and that the same retrofits are considered, the MP steam demand will remain constant. Note that the total MP steam production is set to match the steam demand of the plant for the four cases.

In Table 4.7, an extensive list of the LP steam users in the plant is presented. The unknown steam users account for 7% of the total steam demand which could be accounted to losses or perhaps deviating measuring equipment. The amount of LP steam produced is dependent on the MP steam required, which means that it will be constant through the four cases. However there will be a difference in the excess of LP steam depending on the level of steam savings from heat integration.

Table 4.7. LP steam balance for the original Kraft pulp mill (Värö) and the different cases.

LP steam consumer	Kraft w/o retrofit [MW]	Dissolving, Case 1 [MW]	Dissolving, Case 2 [MW]	Dissolving, Case 3 [MW]	Dissolving, Case 4 [MW]
Total LP steam production	161.0	202.2	202.2	202.2	202.2
Extraction water heating	-	92.1	22.4	9.9	5.0
Packing of wood chips	3.5	4.6	4.6	4.6	4.6
HW production	3.1	-	-	-	-
Bleaching plant	9.7	9.7	9.7	9.7	9.7
Evaporation	51.8	70.3	70.3	70.3	70.3
Stripper column	3.0	4.1	4.1	4.1	4.1
MeOH column	0.8	1.1	1.1	1.1	1.1
Air drier	21.7	18.9	18.9	18.9	18.9
Cyclone drier	9.8	9.8	9.8	9.8	9.8
Feed water tank	13.1	9.0	9.0	9.0	9.0
Recovery boiler	6.4	8.5	8.5	8.5	8.5
Chemical plant	1.1	1.1	1.1	1.1	1.1
Offices facilities	0.2	-	-	-	-
Saw mill	7.8	7.8	7.8	7.8	7.8
Bark drier	3.1	3.1	3.1	3.1	3.1
District heating to Varberg	9.6	-	-	-	-
Dumping condenser	4.8	-	-	-	-
LP steam demand by known consumers	149.5	240.0	170.4	157.9	153.0
Unknown steam consumers	11.5	11.5	11.5	11.5	11.5
Difference	0	-49.3	20.3	32.9	37.7

The effects of the retrofits are included for the four cases, but not for the Kraft case, which is the original setup at Värö today. In reality, the heating demand for feed water pre-heating would increase when converting to dissolving pulp production but since the retrofits suggested decreases the steam demand in the feed water tank, the resulting load decreases as well, as can be seen in Table 4.7. Note that the dumping condenser duty is removed in the four proposed dissolving pulp production cases. This is due to the fact that dumping of steam is unwanted; however the dumping condenser must remain in place to take care of possible steam surplus and variations in the factory. The LP steam demand for district heating to Varberg is not included since the possibilities of DH production was investigated in a foreground/background analysis, presented in *4.7.1. District heating production to Varberg*.

4.6.1. Case 1 – no heat integration

By not performing any heat integration measures it would be very hard to implement a feasible transition to a dissolving pulp mill. The lack of LP steam at 49 MW would need to be added from other types of fuel at the mill, which are not present in these amounts today.

4.6.2. Case 2 – internal integration within the extraction unit

By doing this simple heat integration measures, the mill is now self sufficient in energy. There is 20 MW of excess LP steam available at the plant which can be used to handle variations in heat demand or to supply energy to other heat sinks at the mill.

4.6.3. Case 3 – extraction unit is integrated with the whole plant

By integrating the hemicelluloses extraction process with the rest of the plant, a surplus of up to 33 MW of LP steam can be obtained. This is done by fairly easy heat integration measures, presented earlier, i.e. by using the additional WW produced in the surface condenser to preheat the fresh extraction water.

4.6.4. Case 4 – improved process integration

In this case the process integration is, as mentioned, more complex and would thus demand a higher investment and more heat exchangers. However, all the retrofits could still be implemented resulting in a LP steam surplus of 38 MW.

4.7. Possible uses for surplus energy at the plant

In this section the possible uses for the excess energy in the dissolving pulp plant will be presented.

4.7.1. District heating production to Varberg

From the GCC it can be seen that approximately 50 MW of district heating could be produced by utilizing excess process heat more efficiently, as presented by the green line in the GCC below in Figure 4.4. It would however be very difficult to produce all these 50 MW of DH since it would require a MER-network which would be too complex to implement in reality. The aim is therefore to produce the same amount of DH that is produced today, the red line in Figure 4.4, and to show that the part of DH that in the existing plant at Värö is heated by LP steam now can be heated by other streams with surplus energy.

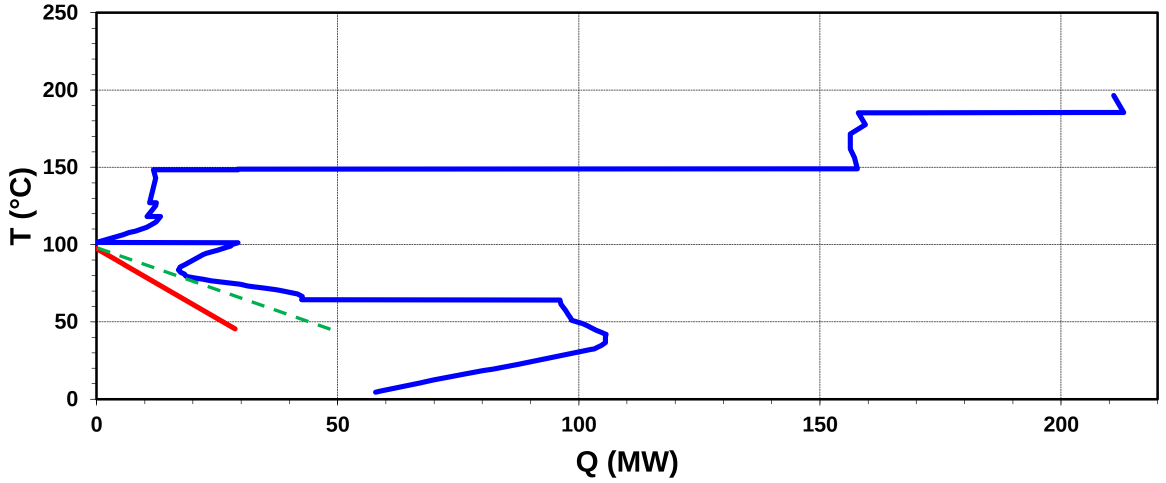


Figure 4.4. GCC of the dissolving pulp mill with current D.H. production (red) and potential D.H. production capacity (green dashed).

The effect in the blowing steam condenser, used for HW production in the Kraft pulp mill, increases by 77% which means an extra 13 MW of heat at 103°C is available. According to the models, the HW demand decreases by 28% when converting to dissolving pulp production. Since the continuous plant in the model is not completely applicable to the batch process at Värö, it is assumed that the HW demand of the new plant remains constant. This assumption leads to approximately 13 MW heat at 103°C in the blowing steam condenser that could be used to heat water for DH to Varberg. This will effectively replace all the 10 MW LP steam used to produce DH today. In Case 4, all the 13 MW is used to replace both the LP steam and some part of the V1 stream.

4.7.2. Saw mill and Pellet factory

Södra Timber recently bought a new saw mill, to replace the old one located at the Värö pulp mill premises. This will induce an increased heating demand for drying of the sawn goods. The total heating demand which consists of LP steam is assumed to increase by 130% compared to the actual heating demand today. This will result in an increase in demand of LP steam by 10 MW.

Furthermore Södra plans to build a new pellet factory on the area of the Värö pulp mill. The production capacity of the pellet factory is assumed to be 9 ton pellets/hour to be dried and the heat consumption is calculated to 1.1 MWh of heat to dry 1 ton of pellets. This adds up to roughly 10 MW of LP steam demand to the new pellet factory.

By implementing the proposed retrofits and performing proper heat integration of the extraction unit, as presented in Case 3, both these new steam demands could be facilitated within the new energy balance of the pulp mill without having to burn any additional fuels. There would even be an additional margin of 13 MW of LP steam that might be needed to handle variations in heating demand or to facilitate a hemicelluloses upgrading plant. Regarding Case 4 this would result in 18 MW of excess LP steam.

4.7.3. Steam and electricity production

There are several ways to combine electricity production and steam surplus at the proposed dissolving pulp plant. Tradeoffs between them make it hard to present just one without doing a proper economical analysis of the alternatives. Below, three different ways of producing electricity at the plant are proposed:

(A): Improving back pressure turbine capacity.
(B): No turbine investment.
(C): Improving back pressure turbine capacity and investing in additional condensing turbine.

Table 4.8. Alternatives for production of steam and electricity.

	A		B		C	
	Steam [MW]	El [MW]	Steam [MW]	El [MW]	Steam [MW]	El [MW]
Dissolving Case 1 w/o Retrofit	-64	71	-55	62	N/A	N/A
Dissolving Case 1	-49	“	-40	“	N/A	N/A
Dissolving Case 2 w/o Retrofit	5	“	15	“	0	72
Dissolving Case 2	20	“	30	“	“	76
Dissolving Case 3 w/o Retrofit	18	“	28	“	“	75
Dissolving Case 3	33	“	43	“	“	79
Dissolving Case 4 w/o Retrofit	23	“	32	“	“	77
Dissolving Case 4	38	“	48	“	“	80
Total process integration of hemicellulose extraction	46	“	56	“	“	83
Complete process integration of whole plant	76	“	86	“	“	93

Performing total process integration of the hemicellulose extraction, as mentioned in *4.5. Hemicellulose extraction* would result in a steam surplus of 46 MW. On the last row in Table 4.8, the complete process integration of the mill and the extraction step is presented. This corresponds to the total pinch violations of the dissolving pulp plant, see Figure 4.2. The mill at Värö is however already built and it might not be possible to achieve these levels of process integration since heat at required temperature might be used elsewhere.

By calculating the turbine electricity production from the HP steam flow rate, let down through the turbine to MP and LP steam, the result is greater than what the current turbine is capable of. The MP steam flow rate is determined by the MP steam demand at the plant, which makes it more or less constant through the different cases presented. In the sections below the different options for steam and electricity production will discussed more thoroughly.

4.7.3.1. (A) – Improving back pressure turbine capacity

Since it has been shown that more electricity potentially could be produced than the maximum effect of the existing turbine, one obvious option would be to invest in additional back pressure turbine capacity. Options could be either to invest in an extra small turbine or to upgrade to one larger turbine.

4.7.3.2. (B) – No turbine investment

If the turbine investment is deemed non profitable, another option could be to let down the excess HP to produce additional LP steam. Upon letdown from HP to LP steam, the resulting LP steam will be overheated. To compensate for this, water is added to the LP steam to reach saturation. By calculating with constant HP production and maximum electricity production of 62 MW, the resulting LP steam production increases with 9.8 MW.

This would increase LP steam production capacity that might be needed if upgrade of hemicelluloses is desired.

4.7.3.3. (C) – Improving back pressure turbine capacity and investing in additional condensing turbine

Further investments on electricity production could be made to maximize the potential electricity production. This could be done by investing in a condensing turbine, and thus to condensate the excess LP steam to generate more electricity.

By doing this 17 MW additional electricity could be produced, included the 9 MW from a bigger back pressure turbine. If however the expanded saw mill and the pellet factory are included as steam consumers then the additional electricity produced by a condensing turbine would merely be 12 MW including 9 MW from the back pressure turbine. For calculations of the condensing turbine effect, see *Appendix 8*.

4.7.3.4. Hemicellulose upgrading

The steam demand for the contingent use or upgrade of hemicellulose needs to be further evaluated. It will however, most likely, demand some steam and it is therefore of interest to have a surplus of LP steam at the dissolving pulp plant. How much on the other hand is unknown and hard to speculate around.

5. Discussion

Through the thesis work it has become evident that a Kraft pulp mill can in fact be converted into a dissolving pulp mill with maintained self sufficiency on heat and energy. It has been shown that after the conversion to dissolving pulp production at Värö, the production rate can be maintained if the capacity of the digestion plant is increased.

5.1. The choice of hemicelluloses extraction option

The different ways to extract hemicelluloses prior to cooking all have different advantages and disadvantages. The two alternatives that were considered for Värö in this thesis were both based on water. The water hydrolysis method was ultimately chosen, since it is favourable in one key aspect compared to the steam hydrolysis. The hemicelluloses extracted with hot water is separated from the pulp stream and can be used for other purposes, like upgrading to high value products, in contrast to extracting with steam where the hemicelluloses ends up in the thin black liquor. Since the use of the hemicelluloses was not studied in this thesis it seemed as the best idea to leave every option open.

To integrate the hemicellulose extraction into the super-batch process adopted at Värö today is also good for limiting lignin precipitation. According to Mateos-Espejel the precipitation is reduced by keeping the pressure elevated during the whole extraction step and not to release it when draining the hydrolysate from the wood chips.

5.2. Validity of the proposed process integration cases

The four cases studied for the conversion to a dissolving pulp are not equally relevant. Case 1 cannot be considered as a real case since it is introduced only to show that the hemicelluloses extraction step accounts for a large part of the total energy balance at the mill. It also serves to show that the extraction step has to be properly integrated, either internally or with the whole mill, in order to make the conversion to dissolving pulp production feasible.

The difference between Case 2 and Case 3 show great potential for energy savings when the extraction step is assessed together with the rest of the mill. This leads to a higher investment cost in form of one additional HX but savings in running costs since the demand of LP steam is decreased by up to 13 MW.

The most rigorous process integration is performed in Case 4 where an additional 5 MW of LP steam is saved when compared to Case 3. The investment cost is probably much higher due to the more complex integration with the rest of the mill. However it may still be profitable when comparing to the potential usages of the excess steam, depending on the energy cost.

For all the integrated cases (Case 2,3 & 4), there is a need for some accumulator tank to be able to heat exchange the extraction water against the hydrolysate in the proposed way.

5.3. Pinch violations and implementability of retrofits

There are some uncertainties regarding the pinch violations at Värö. Insufficient data in different parts of the mill e.g. bark drier and cyclone drier, combined with losses in the steam network and dumping of streams to cesspools causes the total amount of pinch violations to remain uncertain. It is therefore also uncertain whether most of the pinch violations have been found or if something has been overlooked. If overlooking is the case, it does not affect the suggested retrofits in this thesis but merely shows that there are more to be done in order to get Värö as energy efficient as possible.

The retrofits suggested have not been investigated from any implementability or construction stand point, but only to solve as much of the pinch violations as possible. Some of them might therefore be unfavourable to implement. The office facilities heating is one example of where the cost for additional piping might make the retrofit unfavourable, since it only replaces 200 kW LP steam. Another retrofit that might not be favourable is to replace the LP steam supplied directly to hot and warm water tanks since it couples the system, making it harder to control the temperature.

However, most of the retrofit suggestions are deemed feasible and also easy to implement. Depending on how much LP steam that is deemed necessary the retrofits might or might not be so favourable. Even if they are quite easy to realize there is always a cost for additional piping and HX area, which has to be put in perspective to the use of the steam saved. One reason for undertaking the retrofits even if the LP steam is not needed at the moment could be that it is easier to fix them when the plant is being converted to dissolving pulp production.

5.4. Uncertainties and fluctuations

The time period chosen for data extraction was like mentioned earlier a period of mild winter deemed as representative for average weather condition without any peaks in temperature. The weather conditions of the period seem representative for both mild winter but also extend to a fair part of spring and autumn as well. On top of the prevailing weather conditions the pulp plant at Värö was running at even production rate without any major irregularities or disruptions in the production.

However some values were not valid during the time period of choice. The different reasons and measures taken to overcome these are listed in *Appendix 2*. The time period chosen for the more recent time period in November 2012 is deemed quite equal in terms of production rate even if the period had to be chosen to avoid an unexpected stop in the production. Temperature-wise this alternative time period was a bit warmer than the one originally studied.

There are also some data that was either calculated from mixing points or obtained from internal energy documentation from 2011. All these are of course approximations but comparing with true measurements which, according to personnel at Värö, in unlucky times can have a precision of ±20%[6], they are considered fully reasonable for this type of overall energy study.

In the data regarding the existing Värö plant there are uncertainties around the Stripper and Methanol column with regards to their respective condensers. The steam demand for reboiling is however known. The temperatures in the condensers are most likely below the pinch temperature. The left out cooling demand would neither affect the pinch temperature nor the hot utility demand even if it would be included in the GCC.

Data is also missing for the cyclone drier which makes the pinch violations and actual heating demand for the air in the GCC uncertain. However the approximations made, were from best available data found. If, for instance, data was available for the outgoing streams from the cyclone drier they could have been used in a retrofit replacing LP steam to heat the ingoing air. Thereby a lot of the additional pinch violations would have been solved.

The calculated electricity production was performed using the real turbine efficiency (calculated from the current electricity production at Värö) at the current operating conditions at 50 MW. If the flow rate of steam through the turbine increased to produce the maximum effect of 62 MW, the efficiency might be changed and thus our calculations might not be fully valid.

As mentioned previously, in *4.2.1. Kraft pulp plant*, the steam consumption does not add up to the steam production at Värö. The difference is known to exist by staff at Värö but the underlying reasons have not been fully investigated. During the studied time period the losses were higher than usual, at approximately 10%, as opposed to the 5% deviation that is quite normal. According to staff at Värö, the mismatch could be due to steam losses to the surroundings or merely disturbances in measurements devices. Despite the unknown steam losses, all the real steam consumers have been included in the study.

5.5. Extra capacity

The Kraft pulp mill at Värö is probably quite unique due to the fact that a lot of investments have been made in the latest years and therefore has unused capacity in large parts of the plant. Some of these parts are otherwise common bottlenecks, limiting the possible pulp production rate upon conversion to dissolving pulp. Like mentioned earlier, maintaining or increasing the pulp production rate is one of the key elements to successfully convert to a dissolving pulp mill.

The digestion part is most likely the bottleneck of the production today, and will be even more so if converted to a dissolving pulp mill, where both an increase in wood intake and residence time (hydrolysis included) will be necessary. Therefore it seems unavoidable to invest in an upgraded digestion plant. It may be in the way of adding a separate hemicelluloses extraction step prior to the batch digestion lines, but could also be done by carrying out the extraction within the existing vessels and increase the number of digesters.

[6] Andreas Martinsson (Process engineer, Södra Cell Värö) 16 Nov, 2012.

Adding a separate step would probably be necessary if any other extraction method than water hydrolysis is to be used[7]. For instance the digester material would not stand for acidic hydrolysis. After the hemicellulose has been extracted, the actual digestion of the wood takes less time which might make it possible to maintain production rate by adding only a few digesters. The extraction system however would have to incorporate storage tanks to be compatible with the current batch digestion lines. Like mentioned earlier it would probably be easier to expand the digestion plant and perform the water hydrolysis within the existing digesters. This would probably simplify the rebuilding and transition into a dissolving pulp mill, as the extraction will be a part of the cooking process which the staff and engineers are already familiar with.

The pulp mill at Värö has a higher yield, approximately 47-48%, than the model Kraft mill, 43%. The lower model yield was used in the comparison since it was the one that could be compared to the dissolving pulp model. The difference in yield might lead to an underestimated increase in raw material intake and therefore also the need of additional digester capacity. On the other hand it is reasonable to assume that the percentage change in yield between the two mill types is the same in the models as in a proposed conversion.

The RB can, as mentioned in *2.1. Södra Cell Värö*, be upgraded by moving a wall. Representatives as Värö have stated that this would be an easy option to increase the capacity by, at most, 62 % which easily facilitates the upgrade to a dissolving pulp plant with maintained pulp production rate. However it is not clear how much an expansion like this would cost and how beneficial it would be compared to other methods, e.g. lignin separation to reduce the load on the RB. This option is however not investigated in this study, since particularly at Värö it might be possible to keep the production rate constant without that much effort as compared to a regular Kraft pulp mill.

The heating value of lack liquor to the RB is almost constant between a Kraft and a dissolving pulp plant, according to Innventia. Since the hemicelluloses is extracted (or at least a part of it) the heating value of the organic part of the black liquor increases since the fraction of lignin is greater. However this effect is counteracted by the increased amount of inorganic substances reduced by endothermic reactions, which effectively lowers the amount of steam produced in the RB. According to data from Innventia, the steam produced per mass of black liquor at a dissolving plant was calculated to 98% of that at a Kraft pulp plant.

5.6. Potential use of surplus energy

In the suggested conversion to a dissolving pulp plant, there will be an excess in LP steam, which is free for use as best suited. There are numerous ways of using the steam, some more feasible than others.

First of all, DH to Varberg has not been considered in the original GCC nor has it in the steam excess calculations. This means that if DH is to be heated like today, with LP steam, the excess would be reduced by 9.6 MW. However it has been shown that the LP steam used for this purposed, effectively can be replaced by steam from the blowing steam condenser, which

[7] Lennart Sundquist (Process engineer, Södra Cell Värö) 11 Dec, 2012.

almost doubles in size when converting to a dissolving pulp mill. One option to facilitate this change could be to use the existing HX for HW production and install a new one to facilitate the additional condensing demand as well as the high temperature heating of DH. An investment like this would be needed anyway since the current condenser is too small to handle the increased load.

The most logical in this case would be to look upon options for steam consumers that connect to the actual plant. The Södra Timber has already bought a new saw mill which has been placed on the premises of Värö pulp mill and the drying of the sawn goods will be carried out by air heated with LP steam. A new pellet factory is also planned in connection to the saw mill, which will compress saw dust to firing pellets to be sold as biofuel. These two options are of course first in hand to consider since they are practically self written as new steam consumers in the nearby future.

With the Pellet factory and Saw mill taken into account there is however approximately an additional 13 and 18 MW of LP steam in Case 3 and 4 respectively. This could be used to facilitate a hemicellulose upgrade facility, where the hemicelluloses could for instance be used to produce biofuels or speciality chemicals with increased value.

The increased capacity of the RB leads to such an increased flow of HP steam that some of it has to be let down outside the back pressure turbine. Instead of this exergy waste, options to invest in additional effect in the turbine, either a large new one or a small complementary on the side, to maximize the electricity production could be evaluated.

To produce even more electricity, investing in a condensing turbine to extract the most possible energy for electricity production could be considered. The prices are however very high for new turbines which might make this suggestion unattractive, especially considering the fluctuating and seasonally varying steam excess in a Swedish pulp plant.

In this thesis the options for use of the hemicellulose have not been examined. It is however of great interest to investigate what to do with the hydrolysate stream and thereby also whether any steam is needed for the upgrade facility. If for instance it would be fermented to ethanol some type of separation equipment would be needed.

The hydrolysate stream is also very dilute and would most likely need to be concentrated in some way. Some articles show that some type of membrane separation could be feasible but many also suggest some kind of evaporation and then even more steam would be needed.

Regardless of what the hemicellulose should be used for it will most likely need some of the excess steam that could be produced at the mill, making Case 3 & 4 with retrofits included the most beneficial. The steam needed for upgrading of hemicelluloses contradict some of the proposals for use of the excess LP steam e.g. installing of condensing turbine or HP steam export.

It is also a question whether the retrofits are worth to realize if there is no obvious use for the excess steam in Case 3 & 4. For instance only 2 of the 15 MW of retrofits would be needed to break even and be self sufficient in energy if Case 3 is implemented. In Case 4 none of the retrofits would be needed and the plant would still produce some excess steam. However most of the retrofits proposed are quite easy and the excess steam could be used for hemicellulose upgrade or to produce additional electricity in a condensing turbine as discussed above.

5.7. Future work

The data and information around the paper room, and specifically around the cyclone drier, is insufficient for any deeper energy analysis. The value of minimum pinch violations in the cyclone drier implies that there is great potential for improvements and increased heat integration. It would be recommended to assess what type of data is lacking and what type of measurement devices that could be used to collect it. This would help to investigate how the equipment could be heat integrated more efficiently to reduce the steam demand for drying operations.

The use and/or upgrade of hemicellulose need to be assessed, not only to perform a qualitative economic evaluation of the mill but also to decide on which level of process integration that would be most beneficial. When the hemicellulose upgrading plant is designed, a deeper economic evaluation can point out the best compromise between investment cost and revenues from value-added products.

The technical difficulties in converting to dissolving pulp production, e.g. expanding the digester and recovery boiler capacity, potential modification of bleaching process and implementing the hemicellulose extraction into the cooking sequence. Furthermore the actual change in process streams at Värö would need to be assessed to make sure that the data from the continuous models is applicable to the actual pulp mill.

6. Conclusions

The current energy balance at Värö was studied and some possibilities for improvement were identified and some retrofit proposals were made.

Pinch violations at the Värö Kraft pulp mill add up to at least 25 MW which means that there are great possibilities for energy savings. Process integration retrofits have been suggested, which if implemented would save 13 MW steam. In the proposed dissolving pulp plant these values are 29 MW and 15 MW respectively.

According to literature, it is possible to convert a softwood Kraft pulp mill into a dissolving pulp mill in a number of ways. At Värö the most beneficial/practical method would be extraction of the hemicellulose prior to the cooking, by water hydrolysis. Hemicellulose extraction can take place within the existing displacement batch digesters and thus be implemented in a manner limiting new types of process equipment and lignin precipitation.

By increasing the energy efficiency of the mill and the hemicellulose extraction unit, it is possible to design a process which is self sufficient in terms of steam, i.e. without having to rely on any other fuel sources.

Complete process integration of the dissolving pulp mill would result in a maximum of 76 MW excess steam. The most rigorous integration proposed in this study (Case 4) show that it is possible to obtain 38 MW excess LP steam at the plant, given that the back pressure turbine capacity is increased to produce 71 MW electricity. Of the excess steam, 20 MW will be used for a saw mill and pellet plant while the additional 18 MW could potentially be used for hemicellulose upgrade. Further studies can show the amount of excess steam needed and then the steam saving efforts can be adjusted to fit those needs.

A proposition with lower investment cost would be to keep the existing turbine, producing 62 MW of electricity. This combined with a less rigorous integration (Case 3 w/o retrofits) would result in 28 MW of excess LP steam, leaving 8 MW available for use elsewhere. This might however limit the possibility for upgrading the hemicellulose.

When the need for steam has been assessed, the remaining trade-off lies between producing less steam by not implementing process integration measures and increased electricity production by additional turbine investments. The different cases presented in this thesis show in detail the trade off that can be made between performing more rigorous process integration and saving less steam with fewer changes to the process.

References

Delin, L., et al. 2005. *FRAM 09: Bleached market kraft pulpmill.* Stockholm : STFI, 2005.

Fritzon, A. 2002. *Simulation of Pinch Analysis of Värö Mill.* Gothenburg : Department of Heat and Power Technology, Chalmers University of Technolgy, 2002.

Gellerstedt, G. 2005. Ch. 55 - Cellulose products and chemicals from wood. [ed.] M. Ek. *Ljungberg textbook.* Stockholm : Fiber and Polymer Technology, 2005.

Harvey, S. 2011. *Industrial Energy Systems, Course compendium - Autumn 2011.* Gothenburg : Chalmers University of Technology, Division of Heat and Power Technology, 2011.

Henriksson, G., Brännvall, E. and Lennholm, H. 2005. Ch. 2 - The Trees. [ed.] M. Ek. *Ljungberg textbook.* Stockholm : Fiber and polymer Technology, 2005.

Hämäläinen, S., Näyhä, A. and Pesonen, H.L. 2011. Forest biorefineries, A business opportunity for the Finnish forest cluster. *Journal of Cleaner Production.* 2011, pp. 1884-1891.

Innventia AB. 2013. *Personal Communication.* Göteborg : s.n., 2013.

Köpcke, V., et al. 2010. Optimization of treatment sequences for the production of dissolving pulp from birch kraft pulp. *Nordic Pulp and Paper Research Journal.* 1, 2010, Vol. 25, pp. 31-38.

Leschinsky, M., Sixta, H. and Patt, R. 2009. Detailed mass balances of the autohydrolysis of eucalyptus globulus at 170. *Bioresources.* 2009, Vol. 4, 2, pp. 687-703.

Lindsten, H. 2004. *Möjligheter till besparingar på Södra Cell Värö med inriktning på indunstningen.* Göteborg : Chalmers Tekniska Högskola, 2004.

Lundberg, V., et al. 2012a. Energy Analysis for Conversion of a Kraft Pulp Mill into a Dissolving Pulp Mill. *Chemical Engineering Transactions.* 2012a, Vol. 26.

Lundberg, V., et al. 2012b. Process integration of near-neutral hemicellulose extraction in a Scandinavian kraft pulp mill - Consequenses for the steam and Na/S balances. *Applied Thermal Engineering.* 2012b, Vol. 43, pp. 42-50.

Macdonald, C. 2011. *Dissolving Pulp Draws Renewed Interest.* Tembec, Que : Pulp & Paper Canada, 2011. 112.

Marinova, M., Mateos-Espejel, E. and Paris, J. 2010. From Kraft mill to forest biorefinery: An energy and water perspective. II. Case study. *Cellulose Chemistry and Technology.* 2010, Vol. 44, 1-3.

Mateos-Espejel, E., et al. 2010. Simulation of a Kraft Pulp Mill for the Integration of Biorefinery Technologies and Energy Analysis. *Pulp & Paper Canada.* 2010, Vol. 111, 3.

Mateos-Espejel, E., Radiotis, T. and Jemaa, N. 2012. *Implications of Converting a Kraft Pulp Mill to a Dissolving Pulp Operation with a Hemicellulose Extraction Stage.* Ponte-Claire, Québec : FPInnovations, 2012.

Schild, G., Sixta, H. and Testova, L. 2010. Multifunctional alkaline pulping, delignification and hemicellulose extraction. *Cellulose Chemistry and Technology.* 2010, Vol. 44, pp. 35-45.

Shi, H., et al. 2011. A combined acidification/PEO flocculation process to improve the lignin removal from the pre-hydrolysis liquor of kraft-based dissolving pulp production process. *Bioresource Technology.* 2011, Vol. 102, pp. 5177-5182.

Sundin, E. 2011. *Energy optimization of pulp drying, Södra Cell Värö.* Stockholm : Royal Institute of Technology, 2011.

Södra Cell Värö. 2010. Södra Cell Värö. *Södra Cell AB.* [Online] 2010. [Cited: 11 09 2012.] http://www.sodra.com/PageFiles/10517/sodrac_varo_sv.pdf.

Appendix 1

Nomenclature

Värö	Södra Cell Värö pulp mill
FRAM	Future Resource-Adapted pulp Mill
ADt	Air Dry ton (90% dryness)
CW	Cold Water (2°C)
WW	Warm Water (47°C)
HW	Hot Water (65°C)
LP	Low Pressure steam (4.5 bar)
MP	Medium Pressure steam (11 bar)
HP	High Pressure steam (86 bar)
ΔT_{min}	Minimum temperature difference
HX	Heat exchanger
CC	Composite Curve
GCC	Grand Composite Curve
HEN	Heat Exchanger Network
MER	Maximum Energy Recovery
RB	Recovery Boiler
BB	Bark Boiler
DH	District Heating to Varberg
HWWS	Hot and Warm Water System

Appendix 2

Approximations and assumptions for thermal streams at Värö

Some of the measurement points at Värö were not installed at the time period under study and has therefore been investigated during other time periods. Other streams did not have any measurement devices installed and where that was the case, either values has been calculated from mixing points or approximated from nearby tanks etc.

The streams that did not have sufficient measurement were:

- Thin liquor cooling between Liquor tank 1 and Black liquor tank 2.
  - Data has been gathered between: 4-8 Nov 2012.
- Cooling of Backwater between Liquor tank 2 and AWP1.
  - Data has been gathered between: 4-8 Nov 2012.
- Feed water preheating with RB-flue gases.
  - Data has been gathered between: 4-8 Nov 2012.
- The temperature in Filtrate tank, Diffuser.
  - Approximated as the temperature in the Pressure diffuser.
- Some of the WW and HW did not have any temperature measurement.
  - Where this occurred, their temperature was set to 47°C and 65°C respectively.
- Preheating of MAVA: The temperature after the heat exchanging was lower than before the heat exchanger for the cold stream, during the investigated time period.
  - The temperature difference of the cold stream was set to 1°C. This means that it is possible to calculate but the effect on the heat exchanger is minimal.
- The return temperature of the office district heating fluid is missing.
  - Approximated from the assumption that the temperature difference is the same as in other "facility heating"-systems at the plant. This however only accounts for approximately 200 kW of heat which is negligible.
- The pulp drying machine had very limited data for the air stream.
  - Data for the mass flow of dry air and the absolute moisture content was obtained from a previous master thesis that was performed in 2011. The same moisture content of the drying air was used for calculations of the Cyclone drier. (Sundin, 2011)
- The total flow of air through the Cyclone drier was given as a qualified guess from the engineers at Södra Cell Värö. The heating demand of the air did however not match the effect of the steam delivered to the heat exchangers.
  - The measured flow of steam is deemed more accurate and is therefore used in the stream data sheet and the GCC. The minimum pinch violation is calculated from the enthalpies for heating of the humid air.
- The water flow rate which was used to calculate the effect of the turpentine condenser was unknown.
  - It was obtained from temporary measurement devices, documented in a (water modelling) corporate report from 2010.

- The flow rate of the water side of the Blowing steam condenser was unknown.
  - Obtained from the same 2010 corporate report, where it had been approximated from the rpm and capacity of the pump.
- The flow rate of water for cooling of Ack0 was also unknown.
  - Obtained from the same 2010 corporate report.
- Cooling of Ack0: Temperature in the tank is unknown.
  - The temperature was calculated from a mixing point near the outlet of the tank.

Appendix 3

Individual minimum temperature difference used

The temperature difference presented in the table below is for that side in the heat exchanger. For instance a water/water heat exchanger would have 2.5+2.5 in minimum temperature difference.

Stream	Temperature difference [°C]
Water	2.5
Other liquid	3.5
Steam	0.5
Flash steam	2
Gases	8

Appendix 4

Stream data table for existing plant at Värö

Stream	Type	T_{start} [°C]	T_{target} [°C]	Q [kW]	Process part	Comments
Blowing steam condenser	Hot	103.3	103.2	16585	Digestion	Condensation of blowing steam, condenses at 1.25 bar(a). (77-421-042)
Turpentine condenser	Hot	90.1	50.5	1120	Digestion	Condenses gas from gas trap. (77-421-040)
Heating of white liquor	Cold	90.8	108.1	1788	Digestion	Heat exchanged with below. (77-411-027 & 77-411-028)
Cooling of hot liquor	Hot	165.4	146.5	1788	Digestion	Heat exchanged with above. (77-411-027 & 77-411-028)
Cooling of Acc. 0	Hot	80.0	74.3	6041	Digestion	Cooling of Accumulation tank 0. (77-411-026)
HW production	Cold	70.6	80.0	2235	Digestion	HW cistern heated with LP steam. (77-421-037)
Steam demand	Cold	148.4	148.5	3470	Digestion	LP steam for packing the wood chips.
Steam demand	Cold	184.8	184.9	20200	Digestion	MP steam for heating digesters and blowing.
Filtrate Tank Diffuser to COP1	Hot	87.0	82.0	1068	Wash	Cooling of filtrate liquor before wash step in COP-press. (77-431-005)
Back water (Liquor Tank 2 to AWP1)	Hot	87.0	78.0	4506	Wash	Cooling to of Back water before AWP1. (77-431-004)
Thin liquor cooling to Blow Tank	Hot	67.9	38.3	4090	Wash	Thin liquor cooling from Liquor tank 1. (77-431-060)
Steam demand Oxygen bleaching	Cold	184.8	184.9	0	Oxygen bleaching	Only MP steam due to pressure. (Rarely used)
Heating of water to Filter 4	Cold	1.9	46.5	14337	Bleaching	Heating of chemically purified water in to Filter 4. (77-431-060, 77-421-033, 77-421-041 & 77-451-003)
Gas cooling after step 4	Hot	99.7	90.1	293	Bleaching	Condensation from Blowing steam fall tube: S4. (77-451-066)
Heating of KLR to Filter 2 & 3	Cold	81.2	83.3	293	Bleaching	Condensing liquor steam. (77-451-066)
Cooling of BB2 to AWP white wash	Hot	82.7	77.8	4613	Bleaching	(77-451-060)
Cooling of BB4	Hot	83.2	71.5	9216	Bleaching	Back water 4 to Blowing fall tube. (77-451-003 & 77-451-455)
HW to Diluter Screw Feeder	Cold	1.9	65.0	17730	Bleaching	Dilutes the pulp in the Screw feeder.
Steam demand Step 2	Cold	184.8	184.9	1498	Bleaching	MP steam due to pressure (10bar).
Steam demand Step 4	Cold	184.8	184.9	3052	Bleaching	MP steam due to pressure (10bar).
Steam demand	Cold	148.4	148.5	9748	Bleaching	Total LP steam demand for the Bleaching Plant.
Surface condenser	Hot	66.3	66.2	39395	Evaporation	Two real condensers after last evaporation effect.
Steam demand for evaporation	Cold	148.4	148.5	51793	Evaporation	LP steam demand for evaporation.

Primary condenser (MeOH column)	-	-	-	-	Stripper	(77-761-006)
Second. condenser (MeOH column)	-	-	-	-	Stripper	(77-761-007)
Cooling of KLR	Hot	117.8	84.4	11774	Stripper	Cooling of ”Clean liquor condensate”.
Heating of KLB	Cold	66.8	107.7	6288	Stripper	Heating of ”Mixed liquor condensate”.
Heating of KLS	Cold	68.3	102.9	5486	Stripper	Heating of ”Dirty liquor condensate”.
Trim condenser	-	-	105.1	-	Stripper	Partial condensation of Stripper gases. (85-712-251)
Steam demand Stripper	Cold	148.4	148.5	3007	Stripper	LP steam for reboiling in Stripper column.
Steam demand MeOH column	Cold	148.4	148.5	805	Stripper	LP steam for reboiling in MeOH column.

Heating of Paper Room Facilities	Cold	3.0	34.9	499	Paper room	Heated by MP steam at 180°C.
Heating of Paper Room Facilities	Cold	3.0	36.8	2074	Paper room	Heating of Paper Room Facilities with outgoing drier air.
Steam demand, Wire Steam box	Cold	148.4	148.5	2876	Paper room	Steam demand, pressurized to 1.4 bar. Mixture of LP steam and Fresh steam condensate.
Flash steam condenser	Hot	129.1	129.0	1395	Paper room	Condensation of Flash steam from condensate cistern: 83-521-151, to pre-heat air to Air drier.
Air Cooling from Air drier, step 1	Hot	104.0	59.0	3599	Paper room	Cooling outlet air from Air Drier. Steps are adapted to cooling and condensing energy in Moliere diagram.
Air Cooling from Air drier, step 2	Hot	59.0	50.0	8967	Paper room	See step 1.
Air Cooling from Air drier, step 3	-	50.0	*Soft*	-	Paper room	See step 1. *Not used today but useful for retrofit. Cooling to 20 degrees could be a soft target.*
Heating of air to Air Drier	Cold	36.8	122.7	9882	Paper room	Heated with outgoing air, flash steam (above) and MP steam.
Steam demand, Air drier	Cold	148.4	148.5	21690	Paper room	LP steam demand in the heating radiators inside the Air drier.
Flash steam condenser	Hot	120.2	120.1	2808	Paper room	Heating HW to BSB-cistern (below).
HW to BSB-tank	Cold	52.0	70.0	2808	Paper room	Heated with the flash steam above.
WW demand in tank	Cold	2.7	52.1	7730	Paper room	Heated with outgoing air from Air drier in scrubbers.
WW to Back Water Tank	Cold	1.9	45.0	3419	Paper room	WW flow to BSB-cistern.
Air cooling, Wet stage, Cyclone drier	-	52.6	*Soft*	-	Paper room	Cooling of wet air out from Cyclone drier. *Not handled today, useful for retrofit.*
Cooling of air from drier cyclone 1	-	84.8	*Soft*	-	Paper room	See above.
Cooling of air from drier cyclone 2	-	95.6	*Soft*	-	Paper room	See above.
Heating of air to Cyclone drier	Cold	4.3	148.3	10946	Paper room	Total hot air demand for Cyclone drier.

Cooling of Green liquor	Hot	97.1	87.6	2322	Causticizing	Cooling and expansion of Green liquor. (77-771-001)
HW heating	Cold	148.4	148.5	865	Causticizing	HW heated with LP steam.
HW to lime washer	Cold	1.9	70.0	940	Causticizing	HW demand for the Lime washer.
Mist condenser	Hot	84.1	68.4	6353	Causticizing	Condensing mist from the smelt dissolver tank.

Cooling of V1	Hot	76.8	53.6	19146	District heating	Cooled to produce district heating to Varberg.
District heating demand Saw mill	Cold	105.4	111.9	4128	District heating	District heating to the saw mill, timber drier.
Steam demand Saw mill	Cold	148.4	148.5	3676	District heating	Steam demand for special timber drying heat exchangers.
District heating to Tomato farm	Cold	37.7	50.0	703	District heating	Heated with HW.
R&D facilities heating	Cold	49.2	54.7	42	District heating	Heated with HW, produced from the mist condenser.
Office facilities heating	Cold	65.0	75.0	200	District heating	Heated with LP steam.

Steam demand Feed water tank	Cold	148.4	148.5	13127	Recovery Boiler	LP steam demand to de-aerate the feed water tank plus heating of the feed.
Heating of VKT to Feed water	Cold	16.1	17.0	134	Recovery Boiler	Heating of VKT to cover losses in feed water.
Flue gas cooling	Hot	204.5	133.2	9884	Recovery Boiler	Flue gases are cooled to pre-heat Feed water (below).
Feed pre-heating	Cold	83.0	106.2	9884	Recovery Boiler	Heated with the hot flue gases (above).
Steam demand, Feed pre-heating	Cold	184.8	184.9	4335	Recovery Boiler	Feed pre-heating with MP steam before economizer 1.
Steam demand Recovery Boiler	Cold	148.4	148.5	6352	Recovery Boiler	LP steam for air pre-heating etc.
Steam demand Recovery Boiler	Cold	184.8	184.9	11767	Recovery Boiler	MP steam for air pre-heating, oil pre-heating and start-up oil burner.

Cooling of BSB to bio cleaning	Hot	65.0	36.0	3795	Miscellaneous	Performed with Cooling towers.
Cooling of V1 to bio cleaning	Hot	53.6	36.0	14427	Miscellaneous	Performed with Cooling towers (after district heating to Varberg).
VKT production	Cold	1.9	20.0	3368	Miscellaneous	Heating of water to produce VKT from VKK, first heated by HW and then topped with LP steam).
Turbine cooling	Hot	40.2	18.8	581	Miscellaneous	Cooling demand of lubricant oil and turbine generator.
Leakage steam cooling	Hot	99.6	99.5	718	Miscellaneous	Condensation of steam leakage from the turbine.
Heating demand, hot air to Bark drier	Cold	8.3	88.3	7319	Miscellaneous	Heating of Air to dry Bark. Heated with HW and LP steam.
Steam demand, Chemical plant	Cold	148.4	148.5	1106	Miscellaneous	LP steam demand for Chemical Plant.

Appendix 5

Stream data table for dissolving pulp plant

The table below is for constant pulp production and shows the additional streams needed for hemicellulose extraction.

Stream	Type	T_{start} [°C]	T_{target} [°C]	Q [kW]	Process part	Comments
Wood chips & humidity in wood	Cold	39.9	175.0	29976	Hydrolysis	Heating of wood chips, wood water in wood and packing steam condensate. (see table below for steam)
Heating of extraction water	Cold	1.9	175.0	82732	Hydrolysis	Heating of fresh water to hydrolysis.
Cooling of hydrolysate	Hot	175.0	40.0	69617	Hydrolysis	Cooling of the outgoing hydrolysate containing hemicellulose.

The table below is for constant pulp production and depicts relative load increase in percent, compared to Kraft pulp production, as well as the actual load of the thermal streams in the dissolving pulp plant. The temperatures are the same as in the Kraft production case in *Appendix 4*.

Stream	Type	Increase [%]	Q [kW]	Process part	Comments
Blowing steam condenser	Hot	77	29422	Digestion	Condensation of blowing steam, condenses at 1.25 bar(a). (77-421-042)
Turpentine condenser	Hot	0	1120	Digestion	Condenses gas from gas trap. (77-421-040)
Heating of white liquor	Cold	61	2887	Digestion	Heat exchanged with below. (77-411-027 & 77-411-028)
Cooling of hot liquor	Hot	61	2887	Digestion	Heat exchanged with above. (77-411-027 & 77-411-028)
Cooling of Acc. 0	Hot	6	6433	Digestion	Cooling of Accumulation tank 0. (77-411-026)
HW production	Cold	0	2235	Digestion	HW cistern heated with LP steam. (77-421-037)
Steam demand	Cold	31	4560	Digestion	LP steam for packing the wood chips.
Steam demand	Cold	44	29051	Digestion	MP steam for heating digesters and blowing.
Filtrate Tank Diffuser to COP1	Hot	0	1068	Wash	Cooling of filtrate liquor before wash step in COP-press. (77-431-005)
Back water (Liquor Tank 2 to AWP1)	Hot	0	4506	Wash	Cooling to of Back water before AWP1. (77-431-004)
Thin liquor cooling to Blow Tank	Hot	0	4090	Wash	Thin liquor cooling from Liquor tank 1. (77-431-060)
Steam demand Oxygen bleaching	Cold	0	0	Oxygen bleaching	Only MP steam due to pressure. (Rarely used)
Heating of water to Filter 4	Cold	0	14337	Bleaching	Heating of chemically purified water in to Filter 4. (77-431-060, 77-421-033, 77-421-041 & 77-451-003)
Gas cooling after step 4	Hot	0	293	Bleaching	Condensation from Blowing steam fall tube: S4. (77-451-066)
Heating of KLR to Filter 2 & 3	Cold	0	293	Bleaching	Condensing liquor steam. (77-451-066)

Cooling of BB2 to AWP white wash	Hot	0	4613	Bleaching	(77-451-060)
Cooling of BB4	Hot	0	9216	Bleaching	Back water 4 to Blowing fall tube. (77-451-003 & 77-451-455)
HW to Diluter Screw Feeder	Cold	0	17730	Bleaching	Dilutes the pulp in the Screw feeder.
Steam demand Step 2	Cold	0	1498	Bleaching	MP steam due to pressure (10bar).
Steam demand Step 4	Cold	0	3052	Bleaching	MP steam due to pressure (10bar).
Steam demand	Cold	0	9748	Bleaching	Total LP steam demand for the Bleaching Plant.
Surface condenser	Hot	36	53505	Evaporation	Two real condensers after last evaporation effect.
Steam demand for evaporation	Cold	36	70343	Evaporation	LP steam demand for evaporation.
Primary condenser (MeOH column)	-	-	-	Stripper	(77-761-006)
Second. condenser (MeOH column)	-	-	-	Stripper	(77-761-007)
Cooling of KLR	Hot	36	15992	Stripper	Cooling of ”Clean liquor condensate”.
Heating of KLB	Cold	36	8540	Stripper	Heating of ”Mixed liquor condensate”.
Heating of KLS	Cold	36	7451	Stripper	Heating of ”Dirty liquor condensate”.
Trim condenser	-	-	-	Stripper	Partial condensation of Stripper gases. (85-712-251)
Steam demand Stripper	Cold	36	4083	Stripper	LP steam for reboiling in Stripper column.
Steam demand MeOH column	Cold	36	1093	Stripper	LP steam for reboiling in MeOH column.
Heating of Paper Room Facilities	Cold	0	499	Paper room	Heated by MP steam at 180°C.
Heating of Paper Room Facilities	Cold	0	2074	Paper room	Heating of Paper Room Facilities with outgoing drier air.
Steam demand, Wire Steam box	Cold	0	2876	Paper room	Steam demand, pressurized to 1.4 bar. Mixture of LP steam and Fresh steam condensate.
Flash steam condenser	Hot	0	1395	Paper room	Condensation of Flash steam from condensate cistern: 83-521-151, to pre-heat air to Air drier.
Air Cooling from Air drier, step 1	Hot	0	3599	Paper room	Cooling outlet air from Air Drier. Steps are adapted to cooling and condensing energy in Moliere diagram.
Air Cooling from Air drier, step 2	Hot	0	8967	Paper room	See step 1.
Air Cooling from Air drier, step 3	-	-	-	Paper room	See step 1. *Not used today but useful for retrofit. Cooling to 20 degrees could be a soft target.*
Heating of air to Air Drier	Cold	0	9882	Paper room	Heated with outgoing air, flash steam (above) and MP steam.
Steam demand, Air drier	Cold	0	21691	Paper room	LP steam demand in the heating radiators inside the Air drier.
Flash steam condenser	Hot	0	2808	Paper room	Heating HW to BSB-cistern (below).
HW to BSB-tank	Cold	0	2808	Paper room	Heated with the flash steam above.
WW demand in tank	Cold	0	7730	Paper room	Heated with outgoing air from Air drier in scrubbers.
WW to Back Water Tank	Cold	0	3419	Paper room	WW flow to BSB-cistern.
Air cooling, Wet stage, Cyclone drier	-	-	-	Paper room	Cooling of wet air out from Cyclone drier. *Not handled today, useful for retrofit.*
Cooling of air from drier cyclone 1	-	-	-	Paper room	See above.

Cooling of air from drier cyclone 2	-	-	-	Paper room	See above.
Heating of air to Cyclone drier	Cold	0	10946	Paper room	Total hot air demand for Cyclone drier.

Cooling of Green liquor	Hot	43	3329	Causticizing	Cooling and expansion of Green liquor. (77-771-001)
HW heating	Cold	0	865	Causticizing	HW heated with LP steam.
HW to lime washer	Cold	43	1347	Causticizing	HW demand for the Lime washer.
Mist condenser	Hot	43	9078	Causticizing	Condensing mist from the smelt dissolver tank.

Cooling of V1	Hot	0	19146	District heating	Cooled to produce district heating to Varberg.
District heating demand Saw mill	Cold	0	4128	District heating	District heating to the saw mill, timber drier.
Steam demand Saw mill	Cold	0	3676	District heating	Steam demand for special timber drying heat exchangers.
District heating to Tomato farm	Cold	0	703	District heating	Heated with HW.
R&D facilities heating	Cold	0	42	District heating	Heated with HW, produced from the mist condenser.
Office facilities heating	Cold	0	200	District heating	Heated with LP steam.

Steam demand Feed water tank	Cold	33	17486	Recovery Boiler	LP steam demand to de-aerate the feed water tank plus heating of the feed.
Heating of VKT to Feed water	Cold	33	179	Recovery Boiler	Heating of VKT to cover losses in feed water.
Flue gas cooling	Hot	33	13166	Recovery Boiler	Flue gases are cooled to pre-heat Feed water (below).
Feed pre-heating	Cold	33	13166	Recovery Boiler	Heated with the hot flue gases (above).
Steam demand, Feed pre-heating	Cold	33	5775	Recovery Boiler	Feed pre-heating with MP steam before economizer 1.
Steam demand Recovery Boiler	Cold	33	8461	Recovery Boiler	LP steam for air pre-heating etc.
Steam demand Recovery Boiler	Cold	33	15674	Recovery Boiler	MP steam for air pre-heating, oil pre-heating and start-up oil burner.

Cooling of BSB to bio cleaning	Hot	0	3795	Miscellaneous	Performed with Cooling towers.
Cooling of V1 to bio cleaning	Hot	0	14427	Miscellaneous	Performed with Cooling towers (after district heating to Varberg).
VKT production	Cold	0	3368	Miscellaneous	Heating of water to produce VKT from VKK, first heated by HW and then topped with LP steam).
Turbine cooling	Hot	33	774	Miscellaneous	Cooling demand of lubricant oil and turbine generator.
Leakage steam cooling	Hot	33	956	Miscellaneous	Condensation of steam leakage from the turbine.
Heating demand, hot air to Bark drier	Cold	31	9619	Miscellaneous	Heating of Air to dry Bark. Heated with HW and LP steam.
Steam demand, Chemical plant	Cold	0	1106	Miscellaneous	LP steam demand for Chemical Plant.

Appendix 6

Description of pinch violations

Heat exchanging (L1 & L2) in the digestion

Heat is exchanged through the pinch from hot liquor at 165°C to white liquor at 91°C. This is not easily corrected and is from a technical point of view an easy way of recovering heat in the batch digestion process. Totally 727 kW is transferred through the pinch.

Heat exchanging between liquor condensates at the stripper

Cleaned liquor condensate out from the stripper is used to preheat the dirty and the mixed liquor condensates before entering the stripper. It is hard to do anything about this pinch violation in an easy way since the streams on the cold side and on the hot side of the heat exchanger are so closely connected and dependent on each other. Totally 2 238 kW of heat is transferred through the pinch.

Heating air with Medium Pressure Steam in the Paper room

The air preheating for both the paper drying machine and the premises where it is located is heated using medium pressure steam at a temperature of 185°C. This is a major pinch violation, temperature wise, especially for the heating of outside air at temperatures around 2°C for the time period studied. On top of this there is a large air flow out of the paper drying machine that is currently only cooled to around 50°C, before being let out to the atmosphere. The pinch violation is 500 kW for the heating of outside air and below 100 kW for heating of ingoing air to the drier for the studied period.

The reason for using medium pressure steam was simply that the low pressure steam pipe to the paper facilities today is production limiting. A revamped piping system however, would allow energy savings and increased electricity production at the mill.

Flash steam for heating hot water

Currently there is flash steam heating HW that is basically being sent directly to cooling and biological treatment of sewage. This is due to a recent rebuild trial, where the HW heating of the pulp on the wire is no longer used. If the trial is deemed successful this water or the steam at 120°C, could be used somewhere else, which would save approximately 2 800 kW. One proposal is that the flash steam could instead be used as heating medium in the final air heater in the dry section of the Air drier, which today is heated with MP steam.

Heating of air to Cyclone driers

Cool outside air is heated through two different air inlet batteries with slightly different heat sources. The first one is preheated through a glycol circulation (for freeze protection) and the glycol itself heated with LP steam. The air is then heated with LP steam followed by MP steam. In the second air battery the glycol circulation is replaced by an air-air heat exchanger using hot outlet air from the air drier as heat source. The individual intermediate temperatures are unknown as well as the flow of air through each battery. The only known data are temperatures before and after as well as the total air flow.

Today there is some confusion of how much energy that is supplied to the cyclone driers. On one hand there is a steam demand of 10 946 kW of LP and MP steam, while on the other the heating demand of the air, 22 Nm3/s from 4-55°C and 18 Nm3/s from 4-160°C, is calculated from Moliere tables for humid air to merely 4000 kW. The flow rates of the cyclone drier are guesses based on fan size, which might be one reason that the values do not match.

The steam measurements were deemed more accurate and from them a minimum pinch violation was calculated. The assumption was made that all the steam was supplied to the high temperature HX and that the low temp HX only used hot air from the air drier. From this assumption a pinch violation of the air heated up to the pinch temperature was calculated. If the assumption that steam is only used in the one HX is false, the pinch violation will increase; hence it is a minimum pinch violation.

Production of hot water with LP steam

In two heat exchangers, one connected to the HW cistern in the digestion plant and one producing HW in the causticizing, LP steam is used to heat water up to 65°C. Together these two violations account for 3 100 kW heat that could be supplied internally.

Office facilities heating with LP steam

The office facilities heating system (uses water at 75°C) is today heated with LP steam. This pinch violation accounts for 200 kW.

Steam demand for district heating to Varberg

This district heating stream is not included in the GCC since the potential for district heating will be studied instead of the current district heating. Today, 9 500 kW out of 28 700 kW used for district heating to Varberg is produced with LP steam which is a pinch violation. It might however be hard to solve since the temperature needed is rather high, 95°C, due to the long transport distance.

Flue gas cooling to pre-heat feed water to the feed water tank

To compensate for losses in steam condensate some cold deionised water is added to the system. Steam condensate is mixed with cold deionised water before being pre-heated by the flue gases from the recovery boiler. Heating condensate below the pinch temperature, with hot flue gases, is a pinch violation of 6 592 kW. Removing this would increase the temperature of the feed water entering the feed water tank and consequently reduce the LP steam demand for heating that today is approximately 13 100 kW.

Production of deionised cold water

Heated from 2°C to 20°C by firstly HW and then topped off by LP steam. The heating with LP steam is a pinch violation but no measurement is available to determine the ratio between the HW heating and the LP steam heating. The LP steam is therefore assumed to be used only to regulate the exact temperature and the pinch violation is considered to be irresolvable.

Heating of air to bark drier

Outside air is heated to 88°C with HW followed by LP steam. The heating with steam is a pinch violation but heating with HW is not. Since no intermediate temperatures are measured at the plant it is assumed that HW is used for heating as much as possible, to 54.5°C. This would result in a residual pinch violation, where air is heated with LP steam, of 3 093 kW.

Appendix 7

Retrofit idea for the Air drier

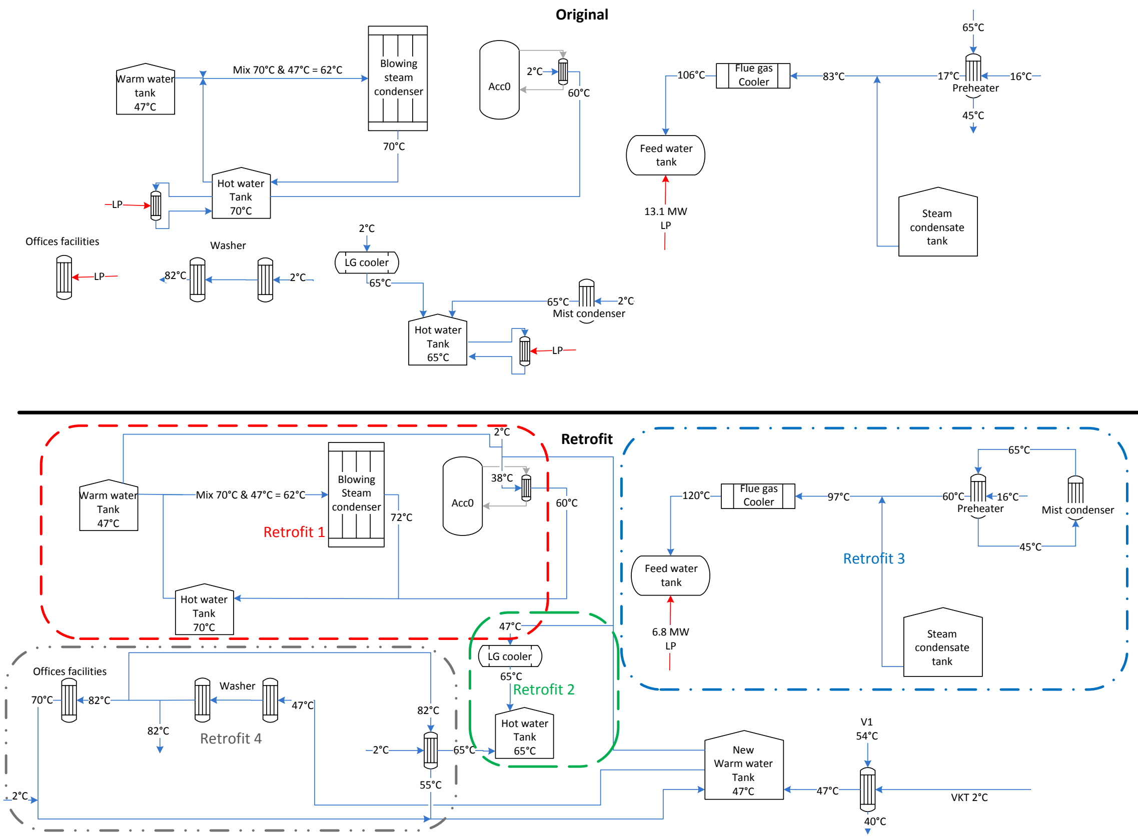
The Air drier outlet is vented towards the atmosphere at approximately 50°C. In the last heat exchangers, outside air to the paper room facilities is heated with the outgoing air, but the heat exchangers are not fully utilized today. The outgoing air's temperature is considered a soft target and could possibly be lowered a fair amount of degrees, thus being able heat a greater flow of air into the facilities. This might be able to replace the MP steam that is used there today.

The proposal is to preheat as much as possible of the ingoing air to the premises with the outgoing airstream from the paper drying machine and, if necessary, to use LP steam to supply the extra steam needed.

Furthermore it is suggested that the third air heater of the ingoing air to the drying machine is heated with low pressure steam instead of medium pressure steam to increase the possibility of producing electricity in the turbine.

Retrofit idea for Kraft pulp plant HWWS

Below is a figure of the proposed HWWS retrofits, described in following subsections.



Cooling of Accumulator tank 0 (Retrofit 1)

Cooling of Accumulator Tank 0 (Acc0) is today carried out by heating CW from 2°C to 60°C. The temperature in Acc0 is 80°C, and it therefore has a potential to produce HW from WW instead of the case today.

Currently, in the cooking plant, there is an additional need of HW which is heated by 2234 kW LP steam. By revamping the cooling system of Acc0 it might be possible to replace this steam demand with internal heat exchanging. One way of doing so, which depends on how the flow rate through the Acc0 heat exchanger can be modified, is presented below.

The flow of fresh CW through Acc0 is 25 kg/s in the original case, which implies that this amount of water needs to be added to the system.

It is preferable to heat up a certain amount of CW to WW to match the heat previously supplied by the LP steam. This means that the flow of CW needs to be lowered to 11.87 kg/s. To match the water addition to the system (25 kg/s) we add 13.12 kg/s of CW before the Acc0 heat exchanger. The flow of 25 kg/s at elevated temperature is not enough to facilitate the cooling demand of Acc0, thus it needs to be increased. To keep the outlet HW temperature at 60 degrees as in the original case, additional WW has to be sent through the heat exchanger, resulting in a total flow of 65.74 kg/s. Since extra water now is sent, from the WW tank, through the Acc0 heat exchanger, it is necessary to compensate for this in the blowing steam condenser in order to maintain constant levels in the HW and WW tanks. Therefore the new flow of WW through the blowing steam condenser is 105.26 kg/s, instead of today's 146 kg/s, and the HW flow is increased to maintain constant total flow rate through the blowing steam condenser. This results in a higher outlet temperature of the condenser thus replacing the steam demand of the HW tank.

Cooling of Green liquor (Retrofit 2)

Green liquor is cooled from 97°C to produce HW from CW at a rate of 8.8 kg/s resulting in a heat load of 2 322 kW. If the cooling of this hot stream instead would take place with WW, producing HW, the heat could be used more efficiently and almost facilitate the total HW demand in the causticizing plant. The HW production in the causticizing plant is 36.23kg/s today and it is possible to produce 30.86 kg/s of this in the green liquor cooler alone (if starting from WW). To heat the remaining 1 413 kW water from 2 to 65°C heat must be recovered from another process part, see point 4. This is done to facilitate a pinch violation correction of 6 350 kW in the Feed water pre-heating system by using excess heat from the mist condenser, originally used to produce HW in the causticizing plant.

Condensing of steam from smelt dissolver tank (Retrofit 3)

In the Mist Condenser there is a heat load of 6 353 kW, heating CW to HW while the condensing media is cooled from 84 to 68°C. This is not considered a pinch violation but the heat can be used more efficiently.

The HW produced here originally, can be produced partially in the green liquor cooler as discussed above, while the heat load of the mist condenser could potentially be used to produce HW for Feed water preheating.

By heating WW instead of CW it is possible to produce 76.0 kg/s of HW at 65°C. This water is used to heat deionised cold water to make up for condensate losses in the steam system. By using the available heat from the mist condenser it is possible to preheat the cold deionised water to 60°C, instead of 16°C which is done today. By doing this it would increase the temperature of the condensate stream before entering the flue gas cooler by approximately 15°C, almost eliminating the previous pinch violation and as a consequence reducing the steam demand in the feed water tank to 6 776 kW instead of 13 127 kW.

The HW is cooled from 65 to 45°C and re-circulated to the mist condenser, assuming no losses in the heat exchangers. By doing so, no extra CW needs to be heated up for this heat exchanger loop. The remaining 6 776 kW of LP steam used in the Feed water tank should still be sufficient for air removal in the tank.

Cooling of washing water (Retrofit 4)

Two heat exchangers in series are used to produce water at 82°C, starting with CW entering the first, with hot side at 87 to 82°C, and the second heat exchanger starting at approximately 30°C, with 87 to 78°C on the hot side. Doing this heat exchanging operation by heating WW it might be able to facilitate the 1413 kW of HW needed in the causticizing plant.

The flow rate of CW through the serial heat exchangers is calculated to 21.47 kg/s from the measurements on one of the heat exchangers. It is desirable to run these on WW instead of CW with the same outlet temperature of 82°C that is used today. By doing so, the flow of water would increase to 38.10 kg/s. This means that there is an additional 16.63 kg/s of water at 82°C (after subtraction of the 21.47 kg/s originally produced) to use for heating of HW elsewhere.

The offices facilities heating system has a heat demand of 200 kW to heat the internal heating system from 65 to 75°C. This is currently done by LP steam and replacing it with water being cooled from 82 to 70°C would demand a flow rate of 3.99 kg/s, which is readily available.

The need of HW heating in the causticizing plant is 1413 kW and using the remaining flow of water at 82°C it would result in a temperature difference of 26.7°C. The residual stream of 12.64 kg/s water at 55.3°C could then be mixed with the returning stream from the offices heating and diluted with CW to produce WW. This would result in 21.00 kg/s WW that can be used in the new WW supply needed.

All these retrofit suggestions demands WW that is currently not produced at the mill but can be produced quite easily by using the excess heat in the "V1" stream currently being cooled by cooling towers before biological cleaning.

Water production

The retrofit demands (11.87 + 30.86 + 38.10 - 21.00) 59.83 kg/s of WW which results in a heating effect of 11 254 kW to heat the 2°C CW to 47°C. This can be done by only using the excess heat in V1, which contains 14 427 kW heat when cooled from 53.5°C to 36°C before the biological cleaning unit.

Redesigning the secondary water system like this would save 9653 kW LP steam and 11 254 kW cooling in the cooling towers. Consequently resulting in savings both in electricity used, in the cooling towers, and decreasing the amount of LP steam needed at the plant. The reason that these do not match is that it seems like some of the measurements from the washing heat exchangers in step 4 is in contradiction with each other. The ingoing water is stated to be "cold mechanically treated water" (normally 2°C) but according to calculations from other measurements it seems to be around 20°C.

Retrofit idea for dissolving pulp plant HWWS

The retrofit idea for the dissolving pulp plant is basically the same as for the Kraft pulp plant, thus the retrofits can refer to the schematic picture in the previous section.

Cooling of Accumulator tank 0 (Retrofit 1)

The effect of the blowing steam condenser increases with 71% to 29.4 MW, according to the model data for the black liquor flash. The additional steam produced can be used to pre-heat extraction water in another heat exchanger.

As in the Kraft case retrofit, by heating WW instead of CW, the production of HW could increase to remove the LP steam demand for HW production in the digestion plant.

Cooling of Green liquor (Retrofit 2)

The cooling demand of the green liquor has increased by 43% to 3329 kW. The water demand in the causticizing plant has increased by 46% according to the models, resulting in a HW demand of 52.9 kg/s. By retrofitting as in the Kraft plant (heating WW instead of CW) 44.2 kg/s HW can be produced in the green liquor cooler. This means that an additional 8.7 kg/s of HW need to be produced elsewhere.

Condensing of steam from smelt dissolver tank (Retrofit 3)

The mist condenser capacity increases with the smelt increase by 43% to 9078 kW. At the same time the deionised water for heating before being mixed with steam condensate only increases by 33% along with the load of the recovery boiler. This means that we get some extra heat from the mist condenser that cannot be used to heat the deionised water which is fixed at 60°C, not to get too high temperature in to the Feed water tank. This is simply solved by sending all the HW produced through the HX and then to the new WW tank at slightly elevated temperature. This means that 618 kW of additional heat is delivered to the WW tank.

Bu using this heat to preheat the deionised water before mixing with the fresh steam condensate, 8460 kW of the pinch violation of 8971 kW is resolved in the flue gas cooler.

Cooling of washing water (Retrofit 4)

The water consumption for washing is assumed to remain unchanged as the amount of pulp in the fibre line is assumed constant through the washing.

This means that we still have the same amount of water at 82°C as in the Kraft case, to be used elsewhere in the plant. This means that we still have 16.63 kg/s water at 82°C to use for other heating purposes.

The office facilities still needs 200 KW of heat which can be taken care of by, like in the Kraft retrofit design, using 3.99 kg/s of 82°C water for this purpose. The return flow is sent straight to the WW tank at 70°C, with an excess of heat at 383 kW.

The remaining flow of 12.64 kg/s is used to heat HW in the causticizing plant. The need of 2278 kW heat results in a returning stream at 39°C. This is sent to the WW tank with a deficit of 428 kW heat. In total we still send some excess heat to the WW tank by including the water after the mist exchanger so this should not be a problem.

Water production

To facilitate this retrofit we need an additional (44.2+38.1-16.63) 65.67 kg/s of WW produced, which results in 12.4 MW of heating that can be done from the V1 that currently is being cooled in cooling towers. By doing so the residual V1 is cooled in this HX from 53.6°C to 38.5°C.

Appendix 8

Calculations for condensing turbine

The calculations are performed as follows.

P_1 = 4.5 bar, T_1 = 148.5°C for saturated LP steam.

Isentropic expansion to P_2 = 0.056 bar.

The enthalpies are obtained from a Mollier diagram in: Mörstedt & Hellsten, Data & Diagram, pp 58. Liber, Malmö.

H_1 = 2745 kJ/kg

$H_{2,is}$ = 2105 kJ/kg

From calculations of the isentropic efficiency at 0.88 this results in:

$H_2 = H_1 - \eta_{is} (H_1 - H_{2,is})$ = 2182 kJ/kg

The mass flow of LP steam is calculated from the enthalpies of LP steam and condensate which adds up to 15.35 kg/s LP steam. The mechanical and electrical efficiency is 0.97.

$P\ [kW] = \dot{m} (H_1 - H_2)\ \eta_{el\&mech}$ = 8385 kW

If the saw mill and the pellets factory are included, the mass flow of excess LP steam is instead 5.92 kg/s and the effect produced in a condensing turbine would be 3233 kW electricity.

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