New environmental objectives will lead to changes in future pulp and paper industry. Water loop closures, new processes to remove Non-Process Elements (NPE) and the demand for reduced fossil fuel consumption are possible results of these new objectives. A first step toward reducing the fuel consumption in a plant is to energy integrate existing processes. Then, by designing the secondary heat system differently, medium temperature (70-100°C) heat can be made available and used in the plant to replace live steam through process modification, thus further decreasing fuel consumption. With a higher temperature where heat can be made available, a commensurate amount of live steam can be replaced. The amount of heat that can be made available varies between plants; two model mills producing market pulp (2000 ADMT/day) are investigated here to help quantify the expected amounts. Two different secondary heat systems are shown for each model mill, one reference design and one novel design where excess heat is made available. The heat made available in these model mills is used in the evaporation plant to replace live steam. The investment cost for the novel secondary heat system making heat available and for
using that heat in the evaporation plant compared to the investment cost for the reference system is between M$5-10 for the two model mills. The total heat demand for the plant is lowered with up to 17% when using the heat made available in the evaporation plant to replace live steam. That, together with the extra investment cost, results in a profit of M$0.6-1.5/year.

**Introduction**

The pulp and paper industry is facing new challenges and the conditions under which it is working are changing. New environmental objectives, partly due to new scientific and technological advances and partly due to the global environmental situation, will render reduced consumption of fossil fuels necessary in order to minimize contributions to global warming. These new environmental objectives will probably lead to more closed water loops, new processes to remove non-process elements (NPE) and a demand for decreased fuel consumption. Because of these expected future changes there are new incentives to change processes. This paper concentrates on process modifications that facilitate integration between different processes in the mill, thus decreasing the heat demand. There is excess heat that can be made available in the plant at medium temperature (70-100°C), and this can be used in the plant to replace live steam, which decreases the heat demand.

The decreased heat demand means that oil, natural gas, and/or biomass fuel can be saved. If the plant is already energy efficient before process modifications are performed, then the saved fuel is biomass; this can be used elsewhere to replace oil and/or produce electricity. In order to evaluate these process modifications pulp mill engineers need to know the
quantity and quality of the excess heat and the cost to make it available. To illustrate this, a state of the art model market pulp mill and a future model mill with a new type of dryer have been studied here.

The two model mills have been evaluated from a system perspective how using excess heat affects the overall heat balance. This paper starts by describing the different model mills, process by process. Then these processes are energy integrated and the benefits of integration are shown as well as the amount and temperature of the heat that can be made available. In order to make the heat available for use, the secondary heat system needs to be designed differently from normal design practices in use today. A novel design is then presented herein and compared to a reference secondary heat system; where this excess heat is used is also shown. Finally, in a cost analysis the total investment cost for the novel system making the excess heat available and the extra cost involved in using this excess heat is compared with the investment cost for the reference secondary system, followed by conclusions.

The work presented in this paper is part of the Swedish National Program “The Eco Cyclic Pulp Mill”, financed by MISTRA, the Swedish Foundation for Strategic Environmental Research and the Swedish Energy Administration [1]. The vision of this program is an eco-cyclic kraft pulp mill producing high quality products, using as much as possible of the energy and biomass potential in the raw material entering the mill. There is a sub-project called Energy Potential, which comprises this work, whose aim is to identify efficient energy systems in the minimum impact mill that are economically and technically attractive.
**Aim and objectives**

The overall aim of this project is to evaluate the use of excess heat to reduce the total heat demand and evaluate the consequences for the model mills. A three-step process does this:

- The excess heat available after process integration is identified and the amount and temperature of the medium temperature heat that can be made available is quantified.
- It is then evaluated how to make this excess heat available and how it is used to reduce the live steam demand.
- Then, an overall economic evaluation is performed.

**Process Integration**

For the process integration study Pinch Analysis [2] has been used. This is a well-established tool used for improving energy efficiency mainly in the petrochemical and chemical industry. Its use in the pulp and paper industry has been less frequent, but recently the interest for this tool has increased [3-5]. In this paper Pro Pi [6], an Excel based program, has been used as a Pinch Analysis Tool.

Earlier studies typically show the potential for process integration in the pulp and paper industry mainly by representing the energy system with Grand Composite Curves (GCC’s). These GCC’s show the lowest possible energy demand using economically viable temperature differences between heat-exchanged streams. All streams are included to show the true minimum heat demand. In this study, however, certain restrictions have been applied: not all streams are included due to the present design of the individual processes.
As a result the GCC’s do not represent all the true streams in the system but rather the defined equipment. The resulting GCC’s show a certain heating and cooling demand. Part of this cooling demand can be regarded as usable excess heat, if changes in the equipment are possible. The excess heat can then replace live steam and reduce the total heat demand. This paper shows how part of the cooling demand below the pinch temperature is made into usable excess heat by changing certain processes. For example, heat below the pinch temperature can be used in the evaporation plant if a redesign of the evaporation plant is allowed. The cooling demand below the pinch temperature above 70°C, is therefore referred to as usable excess heat.

Process integration in the pulp and paper industry is usually restricted as described above and several studies take this into consideration, but there have been few discussions about quantifying the amount of heat below the pinch temperature that can be made available for use, thereby lowering the live steam demand [7]. This amount of heat available is different under various circumstances; looking at the GCC’s for different defined models suggests some general conclusions. This paper does not address the means for achieving process integration in these cases, but rather assumes that process integration has been accomplished elsewhere; calculations are performed, however, based on the heat which remains after process integration.

In a pulp plant, heat below the pinch temperature is used for production of warm and hot water. As a result, in a modern kraft market pulp mill heat at a medium temperature (between 70-100°C) is rejected as waste heat, since the plant does not have the need for all the warm and hot water produced. This is why the design of the secondary heat system is
very important when making heat available below the pinch temperature. Designs of the secondary heat system are in general too large, since all heat below the pinch is used to produce a surplus of warm water, whereas a novel design only produces the warm and hot water needed in the plant, leaving the remaining heat for other uses.

**The model mills**

The base model that has been studied is the ”state of the art”, new green-field pulp mill as defined in the national Swedish research program “The Eco Cyclic Pulp Mill”[1]. It is a model mill consisting of the best techniques built and run in the Nordic pulp and paper industry today. It produces 2000 bleached ADMT/day. It has a total heat demand of 10.40 GJ/ADMT and there is already a surplus of biomass fuel that is used to produce condensing power. The individual processes are described below.

Generally, pulp dryers leave low temperature, low quality heat as humid air. In order to recover the heat in a dryer and assist energy integration, it would be ideal if the excess heat were of higher temperature as well as quality. For the second model studied here we have therefore exchanged the traditional airborne pulp dryer for a superheated steam dryer, that leaves excess heat as steam at a 100 °C. This dryer is currently under development for paper drying and here we assume that the quality of the pulp can be preserved in such a dryer. We chose this dryer in order to see how the overall energy system could be affected if such a dryer would be developed. Except for the dryer the second model is the same as the base case. The dryer substitution does not affect the total heat demand before process integration (10.40 GJ/ADMT).
Cooking plant

The requirement of the digester is to achieve a pulp yield of 46% with a kappa number of 22, represented here by a two-vessel vapor/liquid phase digester with black liquor impregnation, resembling ITC cooking according to Kværner (Figure 1). It consists of a chip bin, a pre-steaming vessel, a pre-impregnation vessel, the digester vessel and three flash chambers. Certain conditions are static here, for example the steam from flash one and two is used in the chip bin and the pre-steaming vessel and the black liquor to the pre-impregnation vessel from the first flash chamber is heat exchanged with white liquor to the digester. The black liquor leaving the cooking plant is not considered a hot stream since it has the temperature that is needed in the evaporation plant. With this layout the streams, representing the heating and cooling demand in the cooking plant are given in Table 1. The resulting composite curves (Figure 2) show that no further integration is possible within the cooking plant. They also show that there is a large amount of medium temperature heat available from the relief vapors condenser, the wash liquor and the steam from the third flash chamber.
Figure 1: The cooking plant
### Table 1: Stream representation for the cooking plant

<table>
<thead>
<tr>
<th>Streams</th>
<th>Type</th>
<th>$T_{\text{start}}$ (°C)</th>
<th>$T_{\text{target}}$ (°C)</th>
<th>Q (GJ/ADMT)</th>
<th>$\Delta T$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Heating of the C8-circulation</td>
<td>Cold</td>
<td>134</td>
<td>160</td>
<td>0.68</td>
<td>5</td>
</tr>
<tr>
<td>Heating of the ITC-circulation</td>
<td>Cold</td>
<td>134</td>
<td>160</td>
<td>0.10</td>
<td>5</td>
</tr>
<tr>
<td>Heating of white liquor to the transfer-circulation</td>
<td>Cold</td>
<td>124</td>
<td>160</td>
<td>0.24</td>
<td>5</td>
</tr>
<tr>
<td>Live steam</td>
<td>Cold</td>
<td>187</td>
<td>188</td>
<td>0.64</td>
<td>0</td>
</tr>
<tr>
<td>Condensing of the relief vapors</td>
<td>Hot</td>
<td>100</td>
<td>99</td>
<td>0.84</td>
<td>5</td>
</tr>
<tr>
<td>Wash liquor to the bottom of the digester</td>
<td>Hot</td>
<td>93</td>
<td>85</td>
<td>0.37</td>
<td>5</td>
</tr>
<tr>
<td>Flash steam from the third flash chamber</td>
<td>Hot</td>
<td>85</td>
<td>85</td>
<td>0.89</td>
<td>5</td>
</tr>
</tbody>
</table>

![Composite Curves for the digester](image)

**Figure 2:** Composite Curves for the digester
Oxygen delignification

The oxygen delignification is a two-step delignification without a washing stage in between. The kappa number after oxygen delignification is 9. The heat consumption in the oxygen delignification is 0.28 GJ/ADMT, which is satisfied by medium pressure steam. There is a 90°C limitation for the MC-pumps and therefore cooling is required before pumping. The washing equipment’s efficiency is taken into account in the oxygen delignification plant as well as the bleach plant. The streams in the oxygen delignification are presented in Table 2 and the Composite Curves in Figure 3. As can be seen there is no room for heat exchanging, but the heat from the cooling can be used in other parts of the plant.
Table 2: Stream representation for the oxygen delignification

<table>
<thead>
<tr>
<th>Streams</th>
<th>Type</th>
<th>$T_{\text{start}}$ (°C)</th>
<th>$T_{\text{target}}$ (°C)</th>
<th>Q (GJ/ADMT)</th>
<th>ΔT</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam to O$_2$ delignification stage</td>
<td>Cold</td>
<td>187.96</td>
<td>188</td>
<td>0.30</td>
<td>0</td>
</tr>
<tr>
<td>Cooling before dilution 1</td>
<td>Hot</td>
<td>97.662</td>
<td>90</td>
<td>0.08</td>
<td>5</td>
</tr>
<tr>
<td>Cooling before the wash filter</td>
<td>Hot</td>
<td>97.662</td>
<td>90</td>
<td>0.29</td>
<td>5</td>
</tr>
</tbody>
</table>

![Composite Curves for the oxygen delignification](image)

Figure 3: Composite Curves for the oxygen delignification
Bleach plant

The bleaching sequence is Q OP D Q PO were Q is a chelation stage, OP is a peroxide enhanced oxygen stage, D is a chlorine dioxide stage and PO is a oxygen enhanced peroxide stage. The heating in the bleach plant is achieved with medium pressure steam to stage OP and PO. The D stage requires a lower temperature than the rest of the bleach plant. This is temperature is attained without the need for cooling, but there is a cooling need caused by the 90°C limitation of the MC-pumps. The effluent from the bleach plant is 9.8 m$^3$/ADMT. The total heat demand for the bleach plant is 0.72 GJ/ADMT. The stream representation for the bleach plant can be seen in Table 3. The Composite Curves for the bleach plant is shown in Figure 4. There is no room for heat exchanging in the bleach plant but there is a small amount of heat available for use elsewhere.
Table 3: Stream representation for the bleach plant

<table>
<thead>
<tr>
<th>Streams</th>
<th>Type</th>
<th>T_{start}(°C)</th>
<th>T_{target}(°C)</th>
<th>Q(GJ/ADMT)</th>
<th>ΔT</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam to OP stage</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.37</td>
<td>0</td>
</tr>
<tr>
<td>Steam to PO stage</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.34</td>
<td>0</td>
</tr>
<tr>
<td>Cooling of OP filtrate</td>
<td>Hot</td>
<td>96</td>
<td>90</td>
<td>0.09</td>
<td>5</td>
</tr>
</tbody>
</table>

Figure 4: Composite Curves for the bleach plant

Dryer

The dryer is a conventional airborne pulp dryer. The air intake is heat exchanged with the exhaust air and is therefore excluded in the stream data, but the remaining heat in the exhaust air is included. The total heat demand for the dryer is 2.05 GJ/ADMT. Excluding the intake/exhaust heat exchange, there is no room for heat exchanging in the dryer as can be seen in the Composite Curves for the dryer (Figure 5). The stream representation for the dryer is given in Table 4.
Figure 5: Composite Curves for the airborne dryer

Table 4: Stream representation for the airborne dryer

<table>
<thead>
<tr>
<th>Stream</th>
<th>Type</th>
<th>$T_{\text{start}}$(°C)</th>
<th>$T_{\text{target}}$(°C)</th>
<th>Q(GJ/ADMT)</th>
<th>$\Delta T$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Evaporation and superheating</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>1.94</td>
<td>10</td>
</tr>
<tr>
<td>Heating of paper sheet</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.04</td>
<td>10</td>
</tr>
<tr>
<td>Heating of remaining water</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.01</td>
<td>10</td>
</tr>
<tr>
<td>Heating of drying air</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.05</td>
<td>10</td>
</tr>
<tr>
<td>Heating of leakage air</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.01</td>
<td>10</td>
</tr>
<tr>
<td>Exiting air after heat exchanging, below dew point</td>
<td>Hot</td>
<td>59</td>
<td>50</td>
<td>0.86</td>
<td>5</td>
</tr>
<tr>
<td>Exiting air below dew point 2</td>
<td>Hot</td>
<td>50</td>
<td>40</td>
<td>0.57</td>
<td>5</td>
</tr>
<tr>
<td>Exiting air below dew point 3</td>
<td>Hot</td>
<td>40</td>
<td>30</td>
<td>0.36</td>
<td>5</td>
</tr>
</tbody>
</table>
Recovery system

The parts of the recovery system, which are significant for the energy analysis, are the evaporation plant, the effluent treatment and the smelt-dissolving tank. The other parts of the recovery have little influence on the live steam demand and are therefore excluded in this analysis. The combined stream data can be seen in Table 5.

Table 5: Stream representation for the recovery system

<table>
<thead>
<tr>
<th>Streams</th>
<th>Type</th>
<th>$T_{\text{start}}(°C)$</th>
<th>$T_{\text{target}}(°C)$</th>
<th>Q(GJ/ADMT)</th>
<th>$\Delta T$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam for evaporation and steam stripper</td>
<td>Cold</td>
<td>154</td>
<td>155</td>
<td>3.70</td>
<td>0</td>
</tr>
<tr>
<td>Steam from evaporation to condenser</td>
<td>Hot</td>
<td>55</td>
<td>54</td>
<td>3.77</td>
<td>0</td>
</tr>
<tr>
<td>Steam from smelt-dissolving tank</td>
<td>Hot</td>
<td>100</td>
<td>85</td>
<td>0.45</td>
<td>5</td>
</tr>
<tr>
<td>Waste water</td>
<td>Hot</td>
<td>78</td>
<td>45</td>
<td>1.61</td>
<td>5</td>
</tr>
</tbody>
</table>

Evaporation and effluent treatment

The evaporation plant consists of six-effects, counter current (Figure 6). The amount of black liquor evaporated is 10.32 MT/ADMT with a dry solid content of 18.4%. The exiting dry solid content is 80%. Because of the high dry solid content to the recovery boiler 25% of all the steam used in the evaporation plant is medium pressure steam. Foul condensates are stripped in a steam stripper that is integrated in the evaporation plant. The total heat demand for the evaporation plant including steam stripping is 3.70 GJ/ADMT. With this layout there are no streams that can be heat exchanged in the evaporation plant, but there is a large amount of heat available from the surface condenser at 55°C.
**Figure 6: Evaporation plant**

**New type of dryer**

In this model a superheated steam dryer replaces the airborne dryer. This is an impingement dryer under development today for paper drying [8, 9] and here assumed to be used for pulp drying as well, causing no damage to the pulp. Compared to a dryer that dries paper the temperature has been lowered considerably, to 150°C, so that the pulp quality can be preserved. This new technique produces a large amount of excess heat at 100°C and has the same heat demand as a conventional dryer. If such a dryer is installed there is a large possibility to use this excess heat at 100°C for other processes, possibly after heat pumping, thereby lowering the total heat demand for the mill. In this model the excess heat is assumed to be used in the evaporation plant without heat pumping. The stream representation can be seen in Table 6 and the Composite Curves in Figure 7.
Table 6: Stream representation for the new dryer

<table>
<thead>
<tr>
<th>Stream</th>
<th>Type</th>
<th>$T_{\text{start}}$</th>
<th>$T_{\text{target}}$</th>
<th>Q</th>
<th>$\Delta T$</th>
</tr>
</thead>
<tbody>
<tr>
<td>Superheated steam for the dryer</td>
<td>Cold</td>
<td>100</td>
<td>150</td>
<td>2.05</td>
<td>5</td>
</tr>
<tr>
<td>Steam from the dryer</td>
<td>Hot</td>
<td>101</td>
<td>100</td>
<td>2.17</td>
<td>5</td>
</tr>
</tbody>
</table>

Figure 7: Composite Curves for the new dryer
**Process integration of the model mills**

As shown above in the Composite Curves for the individual processes, the possible heat recovery in each specific process, without making process changes, has reached its limit; now the different processes need to be integrated with each other. The potential for integration can be seen in a Grand Composite Curve (GCC) as well as the cooling demand under the pinch temperature that can be made available as usable excess heat.

**The base case**

For the base case model (Figure 8), there are two cases. First, the case where the flue gases from the recovery boiler and limekiln are included, usually the heat from flue gases is obtained in a scrubber at a relatively low temperature where there is already a large surplus of excess heat. Second, since it is difficult and/or expensive to recover the heat at higher temperatures than attained in a scrubber, there is also a case where the flue gases are not included. The total heat demand for the two cases, when integrated, is 9.0 GJ/ADMT when flue gases are included, and 9.3 GJ/ADMT for the case that excludes flue gases. The biggest difference between the two cases is found below the pinch temperature. The excess heat available is of course greater if the hot flue gases are included, but because of the problems and costs to recover the heat in the flue gases at a high temperature, the analysis will concentrate on the model where the flue gases are not included. The usable excess heat available below the pinch temperature for the base case is 2.2 GJ/ADMT at a minimum temperature of 80°C.
Figure 8: Grand Composite Curve for the base case model with and without flue gases

Model with new type of dryer

For the model with the new type of dryer, (Figure 9) the total heat demand is the same as the base case 9.3 GJ/ADMT. The flue gases are not included in the model with the new dryer for the same reasons as for the base case. The biggest difference between the base case and the model with a new dryer can be seen below the pinch temperature. There is a larger amount of usable excess heat above 80°C for the model with a new dryer, 4.4 GJ/ADMT compared to 2.2 GJ/ADMT for the base case. The potential to use this heat below the pinch temperature in other processes through process modification, is then of course greater for the model with the new dryer.
Secondary heat system

The secondary heat system is very important when it comes to making usable heat available. Most of the available heat in a plant can be found as warm water because the secondary heat system uses the excess heat in different parts of the mill to make warm and hot water. Here a novel secondary heat system has been designed where some of the excess heat at medium temperature is not used; excess heat at lower temperatures is used to heat warm water instead, and only the amount of warm and hot water that is needed is produced. The criterion when choosing streams has been to leave as much as possible of the medium temperature heat for alternative use.

Figure 9: Grand Composite Curve for the model with the new dryer
Figure 10: GCC for the reference system including the reference evaporation plant

Figure 11: GCC for the novel system including the non-conventional evaporation plant
Usually when making changes of this kind, an iterative process must be performed. Here no iterative process is conducted because the effects of the changes on the system can be predicted and taken into account before the changes are made. The changes that are made and the effects they have on the system are illustrated in Figure 10 and Figure 11. First, the temperature in the surface condenser is lowered to 40°C as can be seen in Figure 11, indicating that the temperature of the heat from the surface condenser used in the novel secondary heat system is 35°C [7]. Secondly, the excess heat that is used in the evaporation plant does not replace steam at a one-to-one ratio, i.e., 1 GJ of excess heat does not replace 1 GJ of live steam. When comparing Figure 10 to Figure 11 it is visible that the live steam demand for the evaporation plant has been lowered. Figure 10 and Figure 11 also show that the excess heat added to the evaporation plant does not replace live steam at a one-to-one ratio. More excess heat is added in Figure 11 than live steam reduced compared to Figure 10. Using excess heat in the evaporation plant lowers the total heat demand and the cooling demand is lowered by a commensurate amount; this can also be seen when comparing Figure 10 to Figure 11. For example, 2.2 GJ/ADMT is added to the evaporation plant below the pinch temperature and 1.5 GJ/ADMT less is added to the evaporation plant above the pinch temperature. The net increase in the surface condenser is 0.7 GJ/ADMT; the total cooling demand below the pinch temperature has however decreased by 1.5 GJ/ADMT as well as the live steam demand by the same amount. The stream data for the plant is shown in Table 7, where ΔT is an economically reasonable minimal temperature difference for the individual heat exchangers.
Table 7: Stream data for the plant

<table>
<thead>
<tr>
<th>Streams</th>
<th>Process</th>
<th>Type</th>
<th>$T_{start}^{°C}$</th>
<th>$T_{target}^{°C}$</th>
<th>Q (GJ/ADMT)</th>
<th>ΔT</th>
</tr>
</thead>
<tbody>
<tr>
<td>Steam for evaporation and steam stripper</td>
<td>Evaporation</td>
<td>Cold</td>
<td>154</td>
<td>155</td>
<td>3.70</td>
<td>0</td>
</tr>
<tr>
<td>Steam from evaporation to condenser</td>
<td>Evaporation</td>
<td>Hot</td>
<td>55</td>
<td>54</td>
<td>3.77</td>
<td>0</td>
</tr>
<tr>
<td>Evaporation and superheating</td>
<td>Dryer</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>1.94</td>
<td>10</td>
</tr>
<tr>
<td>Heating of paper sheet</td>
<td>Dryer</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.04</td>
<td>10</td>
</tr>
<tr>
<td>Heating of remaining water</td>
<td>Dryer</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.01</td>
<td>10</td>
</tr>
<tr>
<td>Heating of drying air</td>
<td>Dryer</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.05</td>
<td>10</td>
</tr>
<tr>
<td>Heating of leakage air</td>
<td>Dryer</td>
<td>Cold</td>
<td>129</td>
<td>130</td>
<td>0.01</td>
<td>10</td>
</tr>
<tr>
<td>Exiting air after heat exchanging, below dew point</td>
<td>Dryer</td>
<td>Hot</td>
<td>59</td>
<td>50</td>
<td>0.86</td>
<td>5</td>
</tr>
<tr>
<td>Exiting air below dew point 2</td>
<td>Dryer</td>
<td>Hot</td>
<td>50</td>
<td>40</td>
<td>0.57</td>
<td>5</td>
</tr>
<tr>
<td>Exiting air below dew point 3</td>
<td>Dryer</td>
<td>Hot</td>
<td>40</td>
<td>30</td>
<td>0.36</td>
<td>5</td>
</tr>
<tr>
<td>Drying with superheated steam</td>
<td>New dryer</td>
<td>Cold</td>
<td>100</td>
<td>150</td>
<td>2.05</td>
<td>5</td>
</tr>
<tr>
<td>Steam from superheated dryer</td>
<td>New dryer</td>
<td>Hot</td>
<td>101</td>
<td>100</td>
<td>2.17</td>
<td>5</td>
</tr>
<tr>
<td>Steam to O₂ delignification stage</td>
<td>O₂ stage</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.30</td>
<td>0</td>
</tr>
<tr>
<td>Cooling before dilution 1</td>
<td>O₂ stage</td>
<td>Hot</td>
<td>98</td>
<td>90</td>
<td>0.08</td>
<td>5</td>
</tr>
<tr>
<td>Cooling before the wash filter</td>
<td>O₂ stage</td>
<td>Hot</td>
<td>98</td>
<td>90</td>
<td>0.29</td>
<td>5</td>
</tr>
<tr>
<td>Steam to OP stage</td>
<td>Bleach plant</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.37</td>
<td>0</td>
</tr>
<tr>
<td>Steam to PO stage</td>
<td>Bleach plant</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.34</td>
<td>0</td>
</tr>
<tr>
<td>Cooling of OP filtrate</td>
<td>Bleach plant</td>
<td>Hot</td>
<td>96</td>
<td>90</td>
<td>0.09</td>
<td>5</td>
</tr>
<tr>
<td>Heating of the C8-circulation</td>
<td>Digester</td>
<td>Cold</td>
<td>134</td>
<td>160</td>
<td>0.68</td>
<td>5</td>
</tr>
<tr>
<td>Heating of the ITC-circulation</td>
<td>Digester</td>
<td>Cold</td>
<td>134</td>
<td>160</td>
<td>0.10</td>
<td>5</td>
</tr>
<tr>
<td>Heating of white liquor to the transfer-circulation</td>
<td>Digester</td>
<td>Cold</td>
<td>124</td>
<td>160</td>
<td>0.24</td>
<td>5</td>
</tr>
<tr>
<td>Live steam</td>
<td>Digester</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.64</td>
<td>0</td>
</tr>
<tr>
<td>Condensing of the relief vapors</td>
<td>Digester</td>
<td>Hot</td>
<td>100</td>
<td>100</td>
<td>0.84</td>
<td>5</td>
</tr>
<tr>
<td>Wash liquor to the bottom of the digester</td>
<td>Digester</td>
<td>Hot</td>
<td>93</td>
<td>85</td>
<td>0.37</td>
<td>5</td>
</tr>
<tr>
<td>Flash steam from the third flash chamber</td>
<td>Digester</td>
<td>Hot</td>
<td>85</td>
<td>85</td>
<td>0.89</td>
<td>5</td>
</tr>
<tr>
<td>Bleach chemical preparation MP steam</td>
<td>Chemicals</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.03</td>
<td>0</td>
</tr>
<tr>
<td>Bleach chemical preparation LP steam</td>
<td>Chemicals</td>
<td>Cold</td>
<td>144</td>
<td>144</td>
<td>0.08</td>
<td>0</td>
</tr>
<tr>
<td>Makeup boiler feed water</td>
<td>Water</td>
<td>Cold</td>
<td>10</td>
<td>95</td>
<td>0.47</td>
<td>5</td>
</tr>
<tr>
<td>Makeup boiler feed water, steam</td>
<td>Water</td>
<td>Cold</td>
<td>148</td>
<td>148</td>
<td>0.25</td>
<td>0</td>
</tr>
<tr>
<td>Hot water</td>
<td>Water</td>
<td>Cold</td>
<td>50</td>
<td>90</td>
<td>1.23</td>
<td>5</td>
</tr>
<tr>
<td>Warm water</td>
<td>Water</td>
<td>Cold</td>
<td>10</td>
<td>50</td>
<td>1.39</td>
<td>5</td>
</tr>
<tr>
<td>Steam from smelt-dissolving tank</td>
<td>Recovery</td>
<td>Hot</td>
<td>100</td>
<td>85</td>
<td>0.45</td>
<td>5</td>
</tr>
<tr>
<td>Waste water</td>
<td>Waste</td>
<td>Hot</td>
<td>78</td>
<td>45</td>
<td>1.61</td>
<td>5</td>
</tr>
<tr>
<td>Waste water, new dryer</td>
<td>Waste</td>
<td>Hot</td>
<td>79</td>
<td>45</td>
<td>1.45</td>
<td>5</td>
</tr>
<tr>
<td>Loss</td>
<td>Loss</td>
<td>Cold</td>
<td>188</td>
<td>188</td>
<td>0.14</td>
<td>0</td>
</tr>
<tr>
<td>Loss</td>
<td>Loss</td>
<td>Cold</td>
<td>144</td>
<td>144</td>
<td>0.22</td>
<td>0</td>
</tr>
</tbody>
</table>

The reference heat-exchanging network has been designed to cool all the heat below the pinch temperature while producing warm and hot water as can be seen in Figure 12. As can be seen in this picture the pinch temperature is 95 °C and all the excess heat is below this temperature. The streams for producing warm and hot water is large, therefore the streams have been split in order to be effectively heat exchanged. There is a limitation on the intake of raw water so some of this warm and hot water needs to be cooled, for example in a cooling tower, and re-circulated in order to cool all the heat below the pinch. The maximum temperature on the wastewater is 45°C due to the effluent treatment plant; so all the heat below the pinch needs to be cooled to 45°C before release to the effluent treatment plant.
<table>
<thead>
<tr>
<th>Hot water</th>
<th>Hot water</th>
<th>Hot water</th>
<th>Hot water</th>
</tr>
</thead>
<tbody>
<tr>
<td>50</td>
<td>50</td>
<td>95</td>
<td>95</td>
</tr>
<tr>
<td>50</td>
<td>50</td>
<td>92</td>
<td>92</td>
</tr>
</tbody>
</table>

**Figure 12: Reference heat exchanger network with the temperatures shown in Celsius**

In the novel network (Figure 13), only the necessary warm and hot water is produced, which means that not all the heat below the pinch temperature is cooled and the flow through the heat exchanger network is smaller. Instead, the medium temperature heat is left for use in the evaporation plant, finally ending up in the surface condenser at a lower temperature than before. Only the necessary warm and hot water is produced, and considered together with the lower temperature in the surface condenser, the need for external cooling of the wastewater before release to the effluent treatment plant in the novel system can be thus minimized and the cooling need of the circulated water in the system has decreased.
The base case

For the base case the number of heat exchanger units are seven for the reference network and eight for the novel network including the heat exchanger units necessary in order to distribute the excess heat (distribution system). The medium temperature excess heat after heat exchanging both in the reference system and the novel system can be seen in Figure 14. This figure is not a part of a GCC, so the temperatures shown are the actual temperatures. The figure shows that the medium temperature heat available differs remarkably between the two secondary heat systems. The novel system leaves usable heat above 80°C of 2.2 GJ/ADMT.

Figure 13: Novel heat exchanger network with the temperatures shown in Celcius
Figure 14: Excess heat for base case left by the novel and reference network

**Model with new type of dryer**

For the model with a new dryer the number of heat exchanger units are seven for the reference network and eight for the novel network including the heat exchanger units in the distribution system for the excess heat. The networks look the same as for the base case (Figure 12, Figure 13). The excess heat after heat exchanging both the reference and the novel methods can be seen in Figure 15. One of the benefits from the new dryer is that the water from the dryer is in the form of steam at 100°C for use in the plant. This will of course appear below the pinch temperature, but is not included in the networks as a surplus that needs to be cooled. It is considered separately and would need a separate steam reformer in order to produce usable steam for use in the mill. There is a large difference between the reference and novel network for the new dryer as well. The novel system
leaves usable excess heat above 80°C of 4.4 GJ/ADMT and no need for external cooling of the effluent to the effluent treatment plant or cooling of surplus warm or hot water.

Figure 15: Excess heat for the model with new dryer left by the novel and reference network

The driving forces in the novel heat exchanger networks are smaller and should for this reason result in a larger area requirement compared to the reference network. This is not the case, however, because in the reference network the streams that have to be heat exchanged are much larger. The area requirements for the four networks can be seen in Table 8 as well as the number of units. The number of units also includes the number of heat exchangers in the distribution system (see “Economy”). The area requirement for the distribution system is not included but can be seen in Table 11.
Table 8: Area requirements for different networks for a plant producing 2000 ADMT/day

<table>
<thead>
<tr>
<th></th>
<th>Number of heat exchanger units</th>
<th>Area requirements (m²)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reference network, base case</td>
<td>7</td>
<td>8400</td>
</tr>
<tr>
<td>Novel network, base case</td>
<td>8</td>
<td>4900</td>
</tr>
<tr>
<td>Reference network, new dryer</td>
<td>7</td>
<td>8300</td>
</tr>
<tr>
<td>Novel network, new dryer</td>
<td>8</td>
<td>5100</td>
</tr>
</tbody>
</table>

**Evaporation designs for use of excess heat**

In this article the excess heat will be used in the evaporation plant. There are other processes where this excess heat could be used, for example in the bleach plant or upgraded by a heat pump and used where low pressure live steam is used today.

When using excess heat in the evaporation plant the excess heat is introduced at one or more intermediate temperature level and cascaded through the effects below this temperature. This method is described more thoroughly by Algehed et al. [7, 10]. Depending on the temperature of the excess heat and the design of the evaporation plant, the excess heat can be used in one or more effects.

**The base case**

The evaporation plant for the base case is a seven-effect evaporation train, as described in "The model mills" above, but with a lower surface condenser temperature of 40°C and one level where excess heat is supplied. This is the most economical design for these conditions according to recent studies [7, 10]. For the base case the area requirements for the distribution system has been calculated as well as the area in the steam reformer. 2.2 GJ/ADMT at 70°C excess heat is introduced in the fifth effect and cascaded through two effects. This results in a saving of 1.5 GJ/ADMT of live steam.
**Model with new type of dryer**

The heat made available by designing the secondary heat system in a novel fashion is distributed and transformed in the same way as for the base case, i.e., a steam reformer at the evaporation plant. The steam from the new dryer also goes through a separate steam reformer. Since there is steam at two different temperature levels heat will be introduced in two different effects in the evaporation plant. Because of the viscosity of black liquor there is a temperature limitation for the different levels of dry solid content. With the large amount of excess heat most of the evaporation happens in the lower effects, resulting in a lower temperature and higher dry solid content in the higher effects. Because of this, the overall temperature difference is not enough when using a six-effect evaporation plant (the temperatures are too low at a high dry solid content) but it is sufficient when using a five-effect evaporation plant. This results in a lower area requirement for this model compared to the base case, (Table 12) even though the amount of excess heat is doubled, but less energy is saved compared to what could have been saved in a six-effect evaporation plant. 2.2 GJ/ADMT of excess heat is introduced in the fourth effect at 95°C and 2.2 GJ/ADMT is introduced in the fifth effect at 70°C. This results in a savings of only 1.8 GJ/ADMT of live steam. Solving the viscosity problem would considerably increase the amount of saved live steam. If there had been effluents to pre-evaporate the excess heat could also have been used more efficiently and more live steam could have been saved.
Economy

It is important in the planning of a process not only to know if it is technically feasible but also to know the cost for implementing it. It is very difficult to say exactly how much it is going to cost to build the secondary heat system and use the excess heat in the plant. Here, approximate investment costs for the new system where excess heat is made available and used in the evaporation plant compared to the investment costs for a reference secondary system is shown. They are not retrofit situations but rather green field plants. Included in the investment cost for the new system are:

- The novel secondary heat system
- The collection of excess heat
- Transformation of the excess heat to usable heat
- The evaporation plant

This is compared to the investment cost for the reference secondary heat system and the reference evaporation plant.

For the model with the new dryer the cost for transforming the steam from the dryer to usable steam is not taken into account as well as the cost for the new dryer, since it is still under development. There is a need for a cooling tower in the reference system as well as the novel system; for the novel system the cooling tower load will be lower. The cost for a cooling tower is site specific, depending on where the plant is located; it has therefore been excluded in this analysis. The cost should be larger for the reference system than the novel
system, not only because it has a larger cooling tower but also because the running cost for the novel system would be smaller. The considerations for investment costs in this analysis are budget costs only, which include heat exchange area cost of $400/m².

<table>
<thead>
<tr>
<th>Table 9: Conditions for the economic evaluation</th>
</tr>
</thead>
<tbody>
<tr>
<td>Annuity factor</td>
</tr>
<tr>
<td>Cost for electricity</td>
</tr>
<tr>
<td>Boiler efficiency</td>
</tr>
<tr>
<td>Power to heat ratio</td>
</tr>
<tr>
<td>Value of biomass</td>
</tr>
</tbody>
</table>

When building the new system the plant uses less live steam than the reference system, but when reducing the steam demand the electricity production is reduced as well. This has to be compensated with the cost for buying the same amount of electricity that has been reduced. The conditions for those calculations are presented in Table 9. An average value for buying electricity in the Swedish pulp and paper industry today is $20/MWh were a higher value is more representative of the conditions in the North American pulp and paper industry. In this plant the saved fuel is biomass because it is already energy efficient and the value for biomass is set to $7/MWh, which by some is considered low. The biomass consists of both bark and lignin, in order to be able to sell lignin there has to be a process available for precipitation of lignin. Today there is ongoing research in this field, also in the Eco-cyclic Pulp Mill research program. If the biomass had been processed the value when selling it would of course be greater. Since this is a green field plant an annuity factor of 0.1 is applied even for the energy investments, but 0.2 is usually used for such investments.
Secondary heat system

The investment cost for the novel and reference secondary heat system for the two models can be seen in Table 10.

Table 10: Investment cost for the different secondary heat systems

<table>
<thead>
<tr>
<th>Number of heat exchanger units</th>
<th>Area requirements (m²)</th>
<th>Area cost (M$)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Reference network, base case</td>
<td>7</td>
<td>8400</td>
</tr>
<tr>
<td>Novel network, base case</td>
<td>8</td>
<td>4900</td>
</tr>
<tr>
<td>Reference network, new dryer</td>
<td>7</td>
<td>8300</td>
</tr>
<tr>
<td>Novel network, new dryer</td>
<td>8</td>
<td>5100</td>
</tr>
</tbody>
</table>

Distribution system

The distribution system includes the heat exchangers to transfer the excess heat to hot water and the steam reformer. There is an optimization between the investment cost for the distribution system and the extra investment cost in the evaporation plant. Shown here (Table 11) is the minimum investment cost of the distribution system if designed together with the evaporation plant.
Table 11: Investment costs for the distribution system

<table>
<thead>
<tr>
<th></th>
<th>Number of heat exchanger units</th>
<th>Area requirements (m²)</th>
<th>Area cost (M$)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Base case</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Distribution system</td>
<td>3</td>
<td>4400</td>
<td>1.9</td>
</tr>
<tr>
<td>Steam reformer</td>
<td>1</td>
<td>5200</td>
<td>1.8</td>
</tr>
<tr>
<td><strong>Model with new dryer</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Distribution system</td>
<td>3</td>
<td>4500</td>
<td>1.9</td>
</tr>
<tr>
<td>Steam reformer</td>
<td>1</td>
<td>5300</td>
<td>1.8</td>
</tr>
</tbody>
</table>

Evaporation plant

The cost for using this excess heat in the evaporation plant is shown (Table 12) and compared to the cost for a conventional evaporation plant where no excess heat is used [7, 10].

Table 12: Investment cost for the evaporation plant

<table>
<thead>
<tr>
<th></th>
<th>Live steam demand for evaporation (GJ/ADMT)</th>
<th>Area requirements (m²)</th>
<th>Investment cost (M$)</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Base case</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>No excess heat</td>
<td>3.7 GJ/ADMT</td>
<td>42000</td>
<td>25.2</td>
</tr>
<tr>
<td>Excess heat</td>
<td>2.2 GJ/ADMT</td>
<td>60000</td>
<td>32.8</td>
</tr>
<tr>
<td><strong>Model with new dryer</strong></td>
<td></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Model with new dryer</td>
<td>1.9 GJ/ADMT</td>
<td>49000</td>
<td>28.2</td>
</tr>
</tbody>
</table>

For the model with the new dryer there is a much larger amount of excess heat available and the excess heat is available at two temperature levels. Because of the temperature sensitivity of the viscosity of the black liquor, only a five-effect evaporation can be used; this results in a lower area requirement compared to the base case. This is discussed further in the section “Evaporation design for use of excess heat” above. Had there been effluents to evaporate as well the excess heat could have been used more efficiently and the reduction of live steam demand would have been greater.
Summary of the Results

The reduction of live steam demand for the evaporation plant as well as the total live steam reduction can be seen in Table 13. This does not include the reduction in total live steam demand that could be achieved by implementing the results from a process integration study.

Table 13: Reduction of live steam demand for the two model mills

<table>
<thead>
<tr>
<th>Live steam demand for evaporation (GJ/ADMT)</th>
<th>Reduction of total live steam demand</th>
</tr>
</thead>
<tbody>
<tr>
<td>Base case</td>
<td></td>
</tr>
<tr>
<td>No excess heat used</td>
<td>3.7</td>
</tr>
<tr>
<td>Excess heat used</td>
<td>2.2</td>
</tr>
<tr>
<td>Model with new dryer</td>
<td>1.9</td>
</tr>
<tr>
<td></td>
<td></td>
</tr>
</tbody>
</table>
| With these reductions and the economic conditions discussed above, there is an economic gain in designing the secondary heat system differently from today and using the excess heat made available in the plant. The results can be seen in Table 14, where the numbers in parenthesis is referring to the case where the cost for electricity is $30/MWh.

Table 14: Results

<table>
<thead>
<tr>
<th></th>
<th>Base case</th>
<th>Model with new dryer</th>
</tr>
</thead>
<tbody>
<tr>
<td>Profit from the sold biomass</td>
<td>3.0 M$/year</td>
<td>3.6 M$/year</td>
</tr>
<tr>
<td>Extra annual investment cost for the new system</td>
<td>1.0 M$/year</td>
<td>0.5 M$/year</td>
</tr>
<tr>
<td>Cost for buying electricity</td>
<td>1.5 (2.2) M$/year</td>
<td>1.8 (2.7) M$/year</td>
</tr>
<tr>
<td>Net profit</td>
<td>0.5 (-0.2) M$/year</td>
<td>1.3 (0.4) M$/year</td>
</tr>
<tr>
<td>Payback period</td>
<td>6.7 (13.5) years</td>
<td>—</td>
</tr>
</tbody>
</table>

This results in a payback period of 6.7 years for the base case with the lower electricity price and 13.5 year when a higher electricity price is used. If choosing a higher price for the biomass sold, $9/MWh instead of $7/MWh the payback period is reduced to 4.3 (6.3) years. The payback period refers to the cost for the new design compared to the
reference design. In a retrofit situation the saved fuel would probably be oil, the savings
would of course be larger and the payback period would be reduced. The profit would be
larger if the investment and running cost for the cooling towers is taken into account since
that cost is probably larger for the reference system.

Since the cost for the new dryer as well as the cost for making its steam into usable steam is
not included here a payback period cannot be calculated.

The investment cost for these new systems is larger than for the reference system but with
reduced use of live steam, biomass can in these cases be sold and that would result in a
yearly net profit. If the steam saved could have been used to replace oil, which would be
the case in many retrofit situations, the yearly net profit would have been larger.

**Conclusions**

There is a large potential to reduce the live steam demand by designing the secondary heat
system differently and using the excess heat in the evaporation plant.

- The total heat demand is reduced by:
  
  14% for the base case
  
  17% for the model with the new type of dryer

- For the model with the new type of dryer the excess heat available cannot be used to its
full potential, thus the incentive to install such a dryer is minimized.

**References**


