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## FOSSIL FUEL REPLACEMENT IN THE PULP MILLS



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## ABSTRACT

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Many kraft pulp mills have a desire to increase their production capacity. In such cases the recovery boiler is often one of the bottlenecks. Recently two new approaches have become available that promise increased pulp mill capacity and creation of completely fossil-fuel free mill concept. These are lignin removal from black liquor (LignoBoost) and biomass gasification for lime kiln

To study the effect of lignin removal to black liquor an example pulp mill was chosen. The aim was to look at the effects of lignin removal on actual mill conditions. Therefore data was gathered on the mill and the calculations done were based on the actual operation of the mill as far as known.

The removal of lignin decreases the organic content of black liquor, but the inorganic portion remains essentially unchanged. The heating value of the black liquor decreases with increased lignin removal.

The new process for extracting lignin from black liquor should not affect the BPR of black liquor very much. Removal of high molecular mass lignin affects only marginally the effective average molecular weight of the non water black liquor portion. The new process for extracting lignin from black liquor may be an opportunity for decreasing the viscosity of black liquor as lignin removal removes high molecular mass components which are a significant source of black liquor viscosity.

For the boiler used in this study we note

- At the same steam generation rate, smelt and black liquor flows increase.
- At lignin removal rate of roughly 20 % the boiler superheating limit is reached. Typically at lower loads auxiliary fuel needs to be added to make full superheating
- Lower furnace will start behaving problematically (TRS, SO<sub>2</sub>, reduction) at about 30 % lignin removal rate
- Minimum load that the boiler can run corresponds to about 50 % lignin removal

The main financial benefits are from the lowered operating costs and from debottlenecking the mill. Lowered operating costs are almost solely based on price of natural gas. If recovery boiler is the bottleneck of the mill, then lignin removal also allows further increase of pulp production and brings in quite a lot of new revenue. It seems that taking out lignin to reduce the recovery boiler load does not bring in significant extra revenue because the current price for the sold lignin is rather low

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## ABBREVIATIONS

ADt	air dry ton
BB	boiler bank
BDt	bone dry ton
BFB	bubbling fluidized bed
BPR	boiling point rise
BSK	bleached softwood kraft
CFB	circulating fluidized bed
DS	dry solids
DNCG	diluted non-condensable gas
ESP	electrostatics precipitator
FG	flue gas
HHRR	hearth heat release rate
HHV	higher heating value
HMML	high molecular mass lignin
HSL	hearth solids loading
HTP	heat treatment process
LF	lower furnace
NPE	non-process elements
SH	superheating
Sub	solid under bark
TRS	total reduced sulphur
WL	white liquor

## 1 INTRODUCTION

Many kraft pulp mills have a desire to increase their production capacity. In such cases the recovery boiler is often one of the bottlenecks. Some recovery boilers can be upgraded to handle increased load (Williamson, Santyr 1988; Oreder 1992), but this approach is often expensive and some of the older recovery boilers at the mills are already upgraded to the limit. An alternative approach that is becoming commercially available is to use a black-liquor gasifier as a booster (Landälv, 2007). This technology has experienced implementation problems (Brown et al., 2007), but it can sometimes be cheaper than upgrading the recovery boiler (Berglin, Andersson 2001). To date, gasifiers are not used widely.

Recently two new approaches have become available that promise increased pulp mill capacity and creation of completely fossil-fuel free mill concept. These are

- lignin removal from black liquor (LignoBoost) (Axelsson et al., 2006, Tomani et al, 2005a, Tomani et al, 2005b)
- biomass gasification for lime kiln

As there are no large scale mills currently operating which utilize these technologies it became important to study the application of these technologies.

## 2 NOVEL METHODS TO INCREASE BIOENERGY USAGE

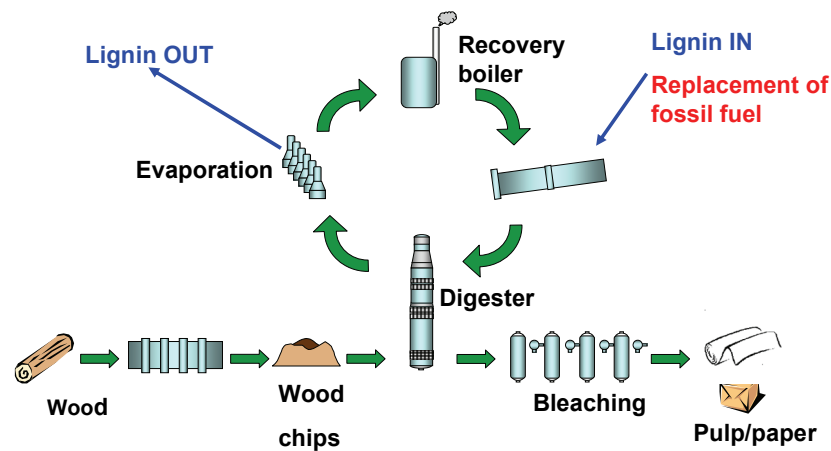
### 2.1 Lignin removal

Major wood components are cellulose 40-50 %, hemicellulose 23-32 % and lignin 15-30 %. About half of the original wood is converted to kraft pulp and the rest of the organics in wood are led through evaporation plant to energy production in recovery boiler.

The separation of lignin is an option that is considered by the pulp mills for several reasons. Firstly, the heat transfer capacity of the recovery boiler is often a bottleneck that limits pulp production. Removing part of the lignin from the black liquor decreases the heat load on the recovery boiler and more pulp can be produced. The separated lignin could be used to replace e.g. fuel oil or natural gas in the lime kilns or be combusted in a power boiler if the energy is required. Secondly the modern pulp mills have energy surplus and this energy surplus can be exported to other users in the form of biofuel. Thirdly separated lignin can be used as a raw material in chemicals (Öhman et al., 2007a).

In this study the focus is to debottleneck the capacity of recovery boiler by lignin removal from the black liquor and use as a fuel in lime kiln, figure 1.

## Lignin from Kraft Black Liquor



**Figure 1.** Lignin removal from black liquor (Axegård et al. 2007).

There are two ways to produce lignin:

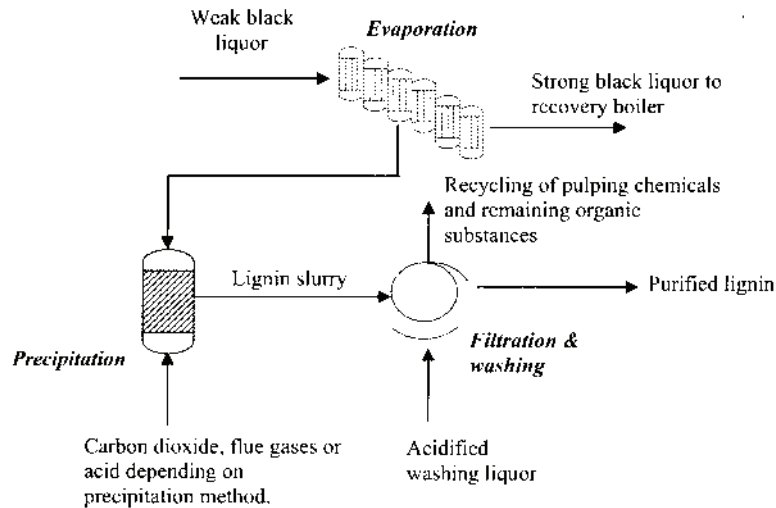
- Membrane filtration after digester
- Precipitation and dewatering from black liquor from evaporation stage

The latter is so called LignoBoost method. The principal of manufacturing of lignin with this method is illustrated in the figure 2. The main processes of lignin removal are:

- Precipitation
- Separation
- Washing

Extracting of lignin is based on precipitation by filtration. Black liquor is led from evaporation in 30-40 dry solid content to the precipitation vessel. The success of precipitation process is dependent on pH and temperature. The hydrogen source can be either sulphuric acid or CO<sub>2</sub>. Sulphuric acid is easy to handle in the mixing stage but the sulphur balance in the mill can be affected. By using CO<sub>2</sub> avoids this balance problem but the mixing stage is more complicated. The lignin precipitate is filtered and then washed to purify the product. The remainder of the black liquor is returned to the chemical recovery system in the pulp mill.

The traditional one stage process of lignin removal has problems with plugging of the filter cake. This resulted in low flow of wash liquor through the cake and high sodium contents in the final lignin. These problems were caused by changes in lignin solubility, caused by excessive pH and ionic strength gradients in the cake during the washing process (Öhman et al. 2007c). Also large lignin losses and high water content in lignin are mentioned (Axegård).



**Figure 2.** Traditional one stage process of lignin removal (Gellerstedt 2007).

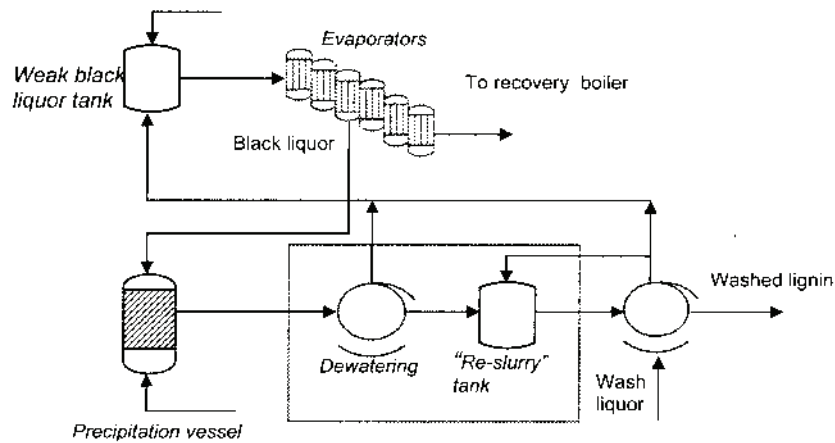
An effective washing is necessary after filtration for three main reasons: (Öhman, 2007b).

- Sodium is enriched in the lignin precipitate and has to be washed out and returned to the mill to avoid disturbing the sodium/sulphur balance and an excessive demand of make-up chemicals
- If the lignin is to be used as a biofuel, the sodium content has to be low to prevent corrosion and other problems associated with low melting ash
- The amount of wash water used should be minimized, since the recycling of spent wash water to the mill recovery system leads to increased energy and capacity demands on the evaporations.

Because of encountered problems a modified method for one stage lignin removal process was developed, figure 3.

If the problems encountered with lignin solubility are caused by large gradients in pH and ionic strength, it would be desirable to even out these profiles during washing. In the improved process lignin is precipitated by acidification and filtration like in one stage process. But instead of washing the lignin directly after filtration, the filter cake is re-dispersed once again. The new slurry is filtered and finally washed using displacement washing. The filter cake is re-dispersed in liquor where the pH are controlled to approximately that of the final wash liquor, the gradient during the washing stage will be small (Öhman, 2007c).





**Figure 3.** A modified method for lignin removal from black liquor (Öhman 2007c)

Typical properties of lignin produced in pilot scale trials in the Bäckhammar kraft mill in Sweden are given in table 1. The analysis is close to softwood lignin analysis from Blunk and Jenkins, 2000. Effective heat value (dry) is higher than other wood derived fuels such as bark, chips, saw dust etc. Ash content of lignin is lower compared to other wood derived fuels. Typical biomass ash content varies between 0,4 – 6,0 (Alakangas 2000). As the ash and sodium content are low in the lignin, it is possible to use lignin as fuels in the lime kiln. However if lignin washing is compromised, then potassium, sulphur and sodium might pose a problem for burning.

**Table 1.** Typical properties of lignin produced in the Bäckhammar mill trials 2004 (Lignoboost, 2008)

<b>Parameter</b>	
Dry solid content, %	70,0
Ash content, % on dry weight	0,2
Effective heat value (dry), MJ/kg	25,4
Effective heat value (30 % moisture), MJ/kg	17,1
<b>Element, w-%</b>	
Carbon	64,0
Oxygen	26,4
Hydrogen	5,7
Nitrogen	0,1
Chloride	0,005
Sodium	0,03

The LignoBoost demonstration plant was constructed to the Swedish pulp mill Bäckhammars bruk in Kristinehamn and the lignin production started up on December 2006. The annual capacity of the mill is 4 000 tonnes/year and the lignin is used in co-firing with coal. Lime kiln tests have also been done.

## 2.2 Biomass gasification

The main reason to gasify wood is the possibility to replace the fossil fuel with carbon dioxide free biofuel in energy production and to increase the power to heat ratio of combined power and heat production. Several types of gasifiers suitable for wood gasification in energy production have been developed. There are three main types of biomass gasifiers (Maniatis 1998):

- Fixed bed (Graham and Barynin 2003)
- Bubbling fluidized bed (BFB) (Zevenhoven-Onderwater et al. 2001)
- Circulating fluidized bed (CFB) (Li et al. 2004)

Fixed bed and circulating fluidized bed gasifier technology has been applied in pulp and paper industry producing lime kiln fuel from biomass. Most practical experience exists from using circulating fluidized bed gasification.

Circulating fluidized bed gasification process may be pressured or work in atmospheric pressure. Circulating fluidized bed gasification is suitable for over 60 MW. Figure 4 shows an atmospheric CFB gasifier. It has a gasifier reactor that is a vertical refractory lined steel cylinder. The fuel is fed to the lower part of the gasifier at the level where the upward flowing gas stream does not contain free oxygen. When entering the reactor the biofuel particles start to dry in the hot gas flows at temperatures of 850 °C - 950 °C, and the pyrolysis also occurs. During the pyrolysing the volatile liberate from the fuel and form the combustible gas. The fixed carbon remains in solid form as char. In the reactor solid suspension includes sand which transports the thermal energy from the bottom of reactor to the upper section balancing the temperature differences and ensuring the stable conditions to the drying and pyrolysis.

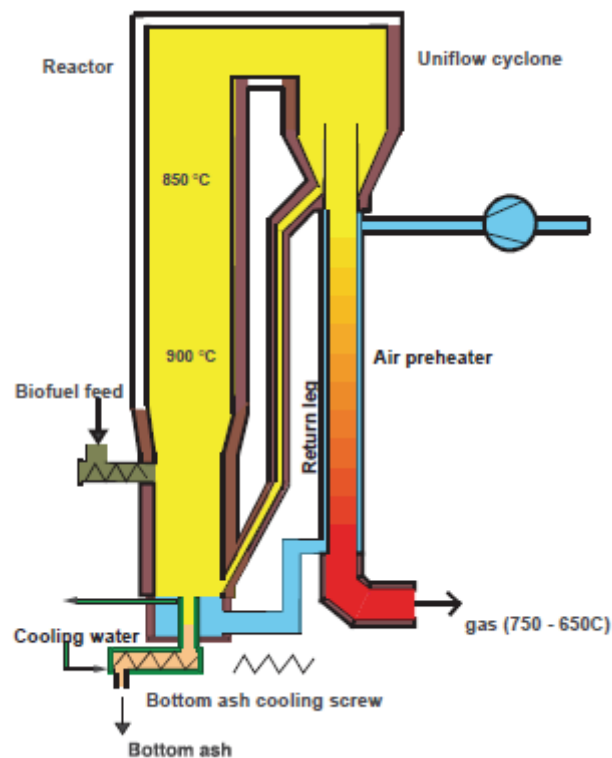
From the reactor, the gases flow to the cyclone, which separate the circulating material from the gas. From the cyclone, the product gas is led directly to the burners via an air preheater and ducts. The separated solids which are primarily sand and carbon fall through the return leg to the bottom part of the gasifier. Gasifying air is preheated indirectly with producer gas and is blown into the bottom of the reactor. There, the carbon residue is burned generating sufficient heat to maintain the gasifier at the desired temperature.

Typical gas composition of wood gas is in table 2 (Huhtinen, Hotta 2000). The flammable components in product gas are hydrogen, carbon monoxide, methane and tars. The product gas include carbon dioxide, steam, char coal, ash and in air gasification nitrogen, too. The heating value of product gas in air gasification is 3-7 MJ/m<sup>3</sup>. If pure oxygen is used in gasification the dilution of product gas components with air nitrogen is avoided and the heating value is 7-15 MJ/m<sup>3</sup>. Because oxygen production needs a lot of electricity, the technology has not been applied.

**Table 2.** Typical gas composition on dry basis for a CFB wood gasifier (Huhtinen, Hotta, 2000)

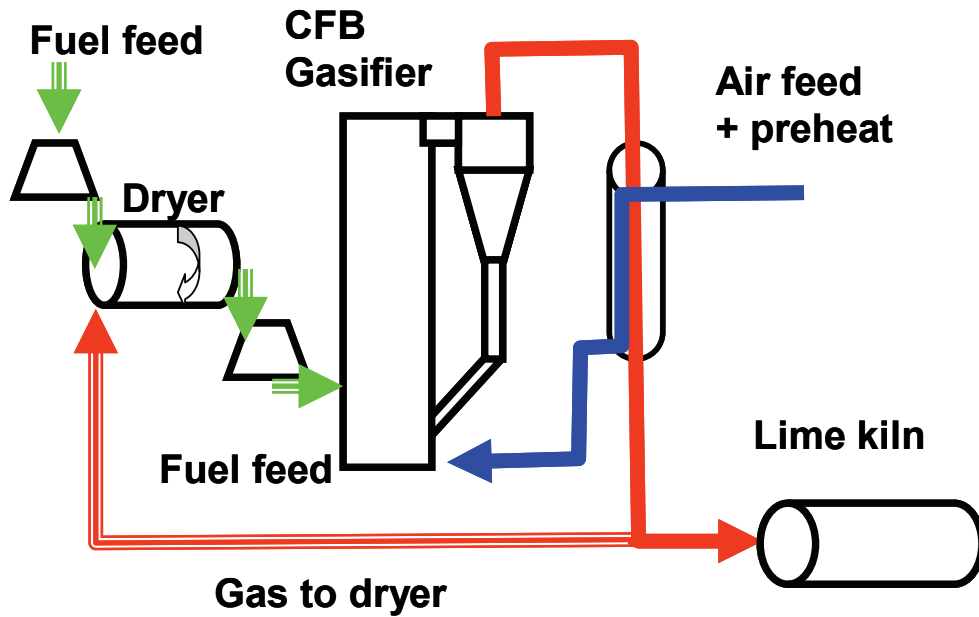
Component	Amount, vol.% (dry basis)
CO	21-22
CO <sub>2</sub>	10-11
CH <sub>4</sub>	5-6
H <sub>2</sub>	15-16
N <sub>2</sub>	46-47

If the fuel contains high amounts of chlorine and alkali metals or aluminium, there will be corrosion and fouling problems in boilers and the gas can not be burned directly. In the hot gas cleaning method the product gas is first cooled by preheating of gasification air and high-pressure boiler feed water. The cooled gas is cleaned in bag filters. Calcium hydroxide is injected to the gas before the bag filters for binding of HCl. The cleaned product gas is led to gas burners.



**Figure 5.** CFB gasifier (Huhtinen, Hotta 2000).

In the lime kiln application, bark or wood requires drying before gasification to produce gas that has sufficiently high heating value to achieve the desired flame temperature in the lime kiln. Fuel drying decreases the amount of flue gas, too. The drying can be done with flue gases taken after the lime kiln or a bark or recovery boiler, figure 6 (Isaksson 2007).



**Figure 6.** CFB gasifier connected to a lime kiln (Isaksson 2007).

At the eighties during the oil crises five circulating fluidized bed gasifiers were installed to produce fuel for lime kilns in Finland, Sweden and Portugal. Of these the biomass gasifier of Metso Power Company has been in function since 1987 in Södra Cell Värö pulp mill in Sweden, Figure 7. The fuel is dried in a drum type dryer before feeding to the gasifier (Isaksson 2007).



**Figure 7.** CFB gasifier connected to a lime kiln in Värö

### 3 EFFECT OF LIGNIN REMOVAL TO BLACK LIQUOR

To study the effect of lignin removal to black liquor an example pulp mill was chosen from eastern Finland. The aim was to look at the effects of lignin removal on actual mill conditions. Therefore data was gathered on the mill and the calculations done were based on the actual operation of the mill as far as known.

#### 3.1 Pulp mill 600 000 ADt BSK/a

The evaluation is based on the Nordic BSK mill. The mill produces mainly pine (softwood) pulp. The mill has an modern large recovery boiler which fires high dry solids. The virgin black liquor dry solids from the evaporator is typically about 78 %. The pulping kappa before oxygen delignification is 30 and after oxygen delignification 14.

#### 3.2 Effect of lignin removal to black liquor properties

Effect of lignin recovery was studied by examining how lignin removal would affect the black liquor composition of target Nordic BSK mill. The black liquor composition was based on typical distribution of organics in wood, Table 3

**Table 3.** Organics in wood

		<b>Pine</b> softwood	<b>Spruce</b> softwood	<b>Birch</b> hardwood	<b>Euca</b> hardwood
Cellulose	%	39	41	40	45
Hemicellulose	%	30	30	37	25
Lignin	%	27	27	20	27
Extractives	%	4	2	3	3
Summary		100	100	100	100

Based on the mill balances at 600 000 ADt/d operation with softwood, distribution of organics in the wood and the black liquor composition method based on Grace, the black liquor composition was calculated. The assumptions used to model mill operation with sample calculation are shown in Appendix I

The reference mill uses mainly pine for the black liquor. The black liquor as fired elemental composition was converted from as fired liquor to virgin liquor using ash properties prediction (Vakkilainen, 2000). The virgin liquor composition was then compared to elemental analysis based on the mill balances and calculations. The calculated properties correspond to measured properties at mill, Table 4.

The lignin removal from black liquor was simulated using Grace's theory of black liquor heating value formation (Adams et al., 1997 p.79) and the MILLFLOW pulp mill balance program. The black liquor composition at various lignin removal rates can be seen in table 5 and Appendix Ic.

**Table 4.** The black liquor elemental composition

		<b>Balance</b>	<b>Virgin</b>	<b>As fired</b>	<b>Mill</b>
HHV	MJ/kgds	14.20	14.16	12.95	12.90
C	mass-%, dry	34.76	34.70	31.85	31.80
H	mass-%, dry	3.48	3.50	3.20	3.20
N	mass-%, dry	0.10	0.08	0.07	0.07
S	mass-%, dry	5.05	5.40	6.56	6.60
Na	mass-%, dry	20.44	20.00	20.74	20.80
K	mass-%, dry	2.20	2.20	2.62	2.60
Cl	mass-%, dry	0.12	0.15	0.21	0.20
Inorganics	mass-%, dry	27.8	27.8	30.1	30.2
S/(Na <sub>2</sub> +K <sub>2</sub> )	mol-%	33.29	36.4	42.2	42.4
Cl/(Na+K)	mol-%	0.34	0.5	0.61	0.58
K/(Na+K)	mol-%	5.95	6.1	6.90	6.85
Dry solids	%	78	78.0	79.5	80.0

It can be seen that the removal of lignin decreases correspondingly the organic content of black liquor, but the inorganic portion remains essentially unchanged. In the simulations done it was assumed that reduction (char bed processes) remain unchanged. As the heating value of the black liquor decreases with increased lignin removal, then it is probable that at some point the char bed temperature starts to be affected. Then decreasing reduction would change the inorganic portion of black liquor.

**Table 5.** The black liquor composition at various lignin removal rates

Lignin removal		<b>0 %</b>	<b>5 %</b>	<b>10 %</b>	<b>15 %</b>	<b>20 %</b>	<b>25 %</b>	<b>30 %</b>
Heating value of BLDS	kJ/kg	14148	13974	13795	13610	13419	13222	13018
Heat in to recovery	MW	524	510	495	481	467	452	438
Massflow to recovery boiler	tDS/d	3439	3384	3330	3275	3221	3166	3112
HHRR	kW/m <sup>2</sup>	3376	3282	3187	3093	2999	2905	2811
Black liquor DS generation	kg/ADt	1807	1779	1752	1724	1696	1668	1641
Lignin in black liquor	kg/BDt	616	585	554	523	493	462	431
Total organic to BL	kg/BDt	1358	1327	1296	1265	1234	1204	1173
Total inorganic to BL	kg/BDt	650	650	650	650	650	650	650
<b>Lignin removal</b>		<b>35 %</b>	<b>40 %</b>	<b>50 %</b>	<b>60 %</b>	<b>70 %</b>	<b>80 %</b>	<b>90 %</b>
Heating value of BLDS	kJ/kg	12807	12589	12128	11633	11100	10523	9897
Heat in to recovery	MW	423	409	380	352	323	294	265
Massflow to recovery boiler	tDS/d	3057	3003	2894	2785	2676	2567	2458
HHRR	kW/m <sup>2</sup>	2717	2623	2435	2248	2061	1874	1688
Black liquor DS generation	kg/ADt	1613	1585	1530	1475	1419	1364	1308
Lignin in black liquor	kg/BDt	400	369	308	246	185	123	62
Total organic to BL	kg/BDt	1142	1111	1050	988	927	865	803
Total inorganic to BL	kg/BDt	650	650	650	650	650	650	650

The hearth heat release rate (HHRR) decreases faster than the massflow to the recovery boiler. So of the typical two indices (HSL and HHRR) using the latter for evaluating the furnace behavior is recommended.

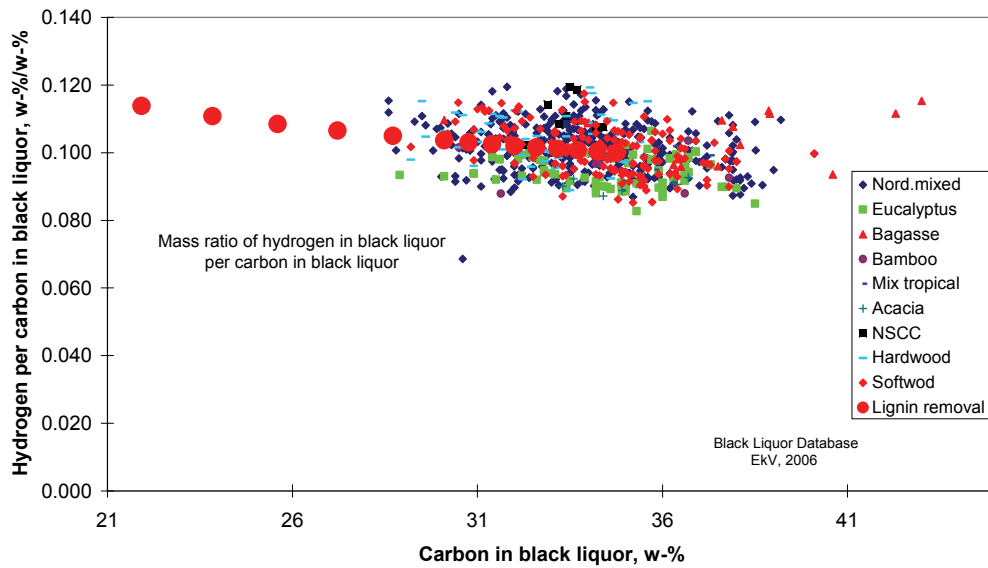
### 3.2.1 Effect of lignin removal to black liquor composition

Lignin removal was simulated using MILLFLOW pulp mill balance program. The lignin removal affects the black liquor properties as organics are removed from the liquor. The analysis presented is for liquor from the fibre line. So no additional flows e.g. from chemical manufacture have been taken into account. Results of the calculations are shown in table 6 and Appendix Id.

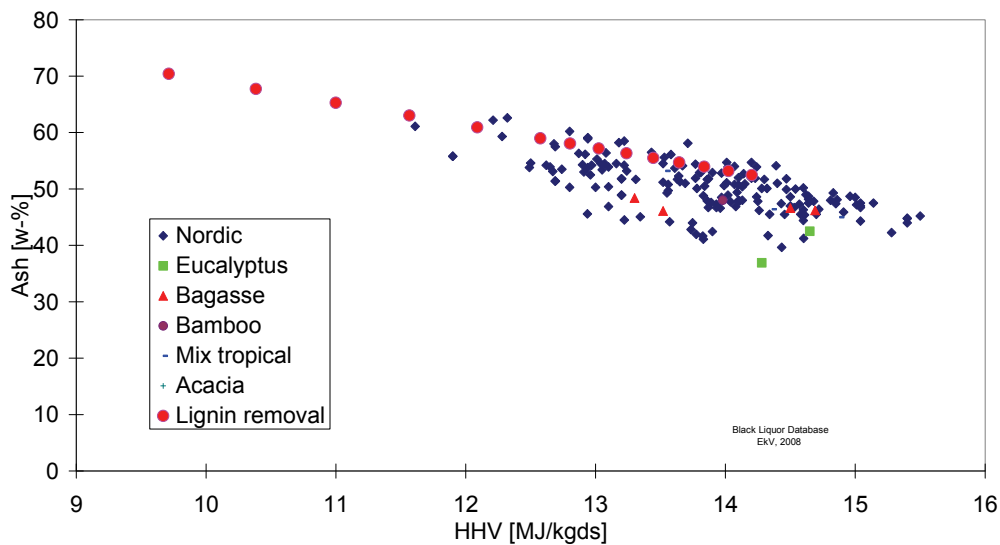
**Table 6.** The black liquor elemental composition at various lignin removal rates

Lignin removal		0 %	5 %	10 %	15 %	20 %	25 %	30 %
C	%	34.65	34.08	33.49	32.88	32.26	31.62	30.95
H	%	3.46	3.42	3.37	3.33	3.28	3.23	3.17
N	%	0.10	0.10	0.10	0.10	0.10	0.10	0.10
S	%	4.09	4.15	4.21	4.27	4.34	4.41	4.47
Na	%	21.40	21.71	22.03	22.36	22.70	23.05	23.41
K	%	2.20	2.23	2.27	2.30	2.33	2.37	2.41
Cl	%	0.15	0.15	0.15	0.16	0.16	0.16	0.16
O by diff	%	33.95	34.16	34.38	34.60	34.83	35.07	35.32
Inorganics		26.36	26.68	27.02	27.37	27.72	28.09	28.47
S/(Na <sub>2</sub> +K <sub>2</sub> )	mol-%	25.8	25.8	25.8	25.8	25.8	25.8	25.8
Cl/(Na+K)	mol-%	0.4	0.4	0.4	0.4	0.4	0.4	0.4
K/(Na+K)	mol-%	5.7	5.7	5.7	5.7	5.7	5.7	5.7
Dry solids	%	78.0	78.0	78.0	78.0	78.0	78.0	78.0
Lignin removal		35 %	40 %	50 %	60 %	70 %	80 %	90 %
C	%	30.27	29.56	28.07	26.48	24.78	22.95	20.98
H	%	3.12	3.06	2.95	2.82	2.69	2.54	2.39
N	%	0.10	0.10	0.10	0.10	0.11	0.11	0.11
S	%	4.55	4.62	4.77	4.94	5.12	5.31	5.51
Na	%	23.78	24.17	24.98	25.84	26.77	27.76	28.84
K	%	2.45	2.49	2.57	2.66	2.75	2.86	2.97
Cl	%	0.17	0.17	0.18	0.18	0.19	0.19	0.20
O by diff	%	35.57	35.83	36.38	36.97	37.60	38.28	39.00
Inorganics		28.86	29.27	30.12	31.03	32.00	33.05	34.18
S/(Na <sub>2</sub> +K <sub>2</sub> )	mol-%	25.8	25.8	25.8	25.8	25.8	25.9	25.9
Cl/(Na+K)	mol-%	0.4	0.4	0.4	0.4	0.4	0.4	0.4
K/(Na+K)	mol-%	5.7	5.7	5.7	5.7	5.7	5.7	5.7
Dry solids	%	78.0	78.0	78.0	78.0	78.0	78.0	78.0

The ratio of carbon to hydrogen remains fairly constant, figure 8. Therefore the combustion of organics will not change too much even if lignin is removed from the black liquor. The inorganic portion of black liquor increases quite a bit. As seen in figure 9 the ash analysis of lignin removed liquor would give very high numbers. Ash is mostly oxidized, therefore the smelt flow per mass unit of black liquor will not increase as much.



**Figure 8.** Hydrogen to carbon ratio at various lignin removal rates compared to general trend in black liquors

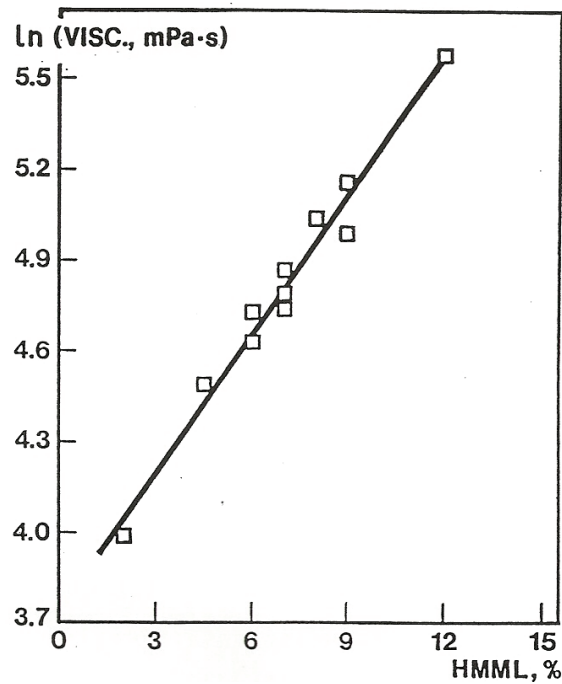


**Figure 9.** Ash at various lignin removal rates compared to general trend in black liquors

### 3.2.2 Effect of lignin removal to black liquor viscosity and BPR

The viscosity of the black liquor is determined by its composition. The two main groups are lignin and polysaccharides (Söderhjelm, 1986). The polysaccharides like xylan are dissolved during the cook as long chain molecules which increase the original viscosity level of black liquor. The relationship is clearly shown when we compare the behavior of black liquors originating from different mills, figure 10.



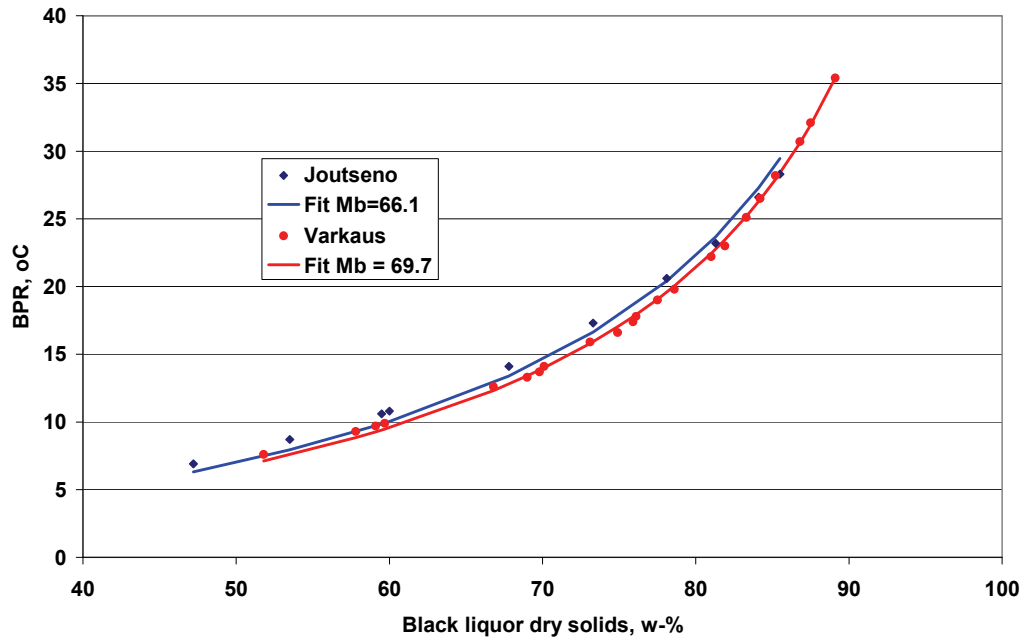


**Figure 10.** Effect of lignin content to black liquor viscosity.

In practice this means that the new process for extracting lignin from black liquor may be an opportunity for decreasing the viscosity of black liquor. Effect of lignin removal to black liquor viscosity has been studied by Moosavifar et al. 2006. They concluded that there was some decrease in viscosity.

Black liquor boiling point raise has been studied by Järvinen and Kankkunen 2000. They concluded that a simple two component approach could give very satisfactory results in modeling BPR for mill liquors, figure 11.

In practice this means that the new process for extracting lignin from black liquor should not affect the BPR of black liquor very much. Removal of high molecular mass lignin affects only marginally the effective average molecular weight of the non water black liquor portion. Effect of lignin removal to black liquor BPR has been studied by Moosavifar et al. 2006. They concluded that there was little effect on boiling point rise.



**Figure 11.** Fits of BPR to two mill liquors using approach from Järvinen and Kankkunen 2000.

### 3.3 Effect of lignin removal to evaporator operation

Removal of lignin changes black liquor properties. As discussed previously the main black liquor parameters that affect the evaporator performance do not change significantly.

#### 3.3.1 Effect of BPR to evaporator operation

Removal of lignin does somewhat lower the boiling point rise of black liquor. This reduction is not significant. BPR increases with dry solids concentration. Therefore the effect of boiling point rise could decrease the evaporation especially in high concentration effects.

It is expected that the final concentration and the evaporation capacity are not affected because of effects of lignin removal on boiling point rise.

#### 3.3.2 Effect of viscosity to evaporator operation

Removal of lignin decreases black liquor HMML-lignin concentration. This should decrease the black liquor viscosity. Viscosity increases with increased dry solids content. For evaporators this poses a flow limit at some dry solids content. The increase of viscosity is very steep after a certain point (Holmlund and Parviainen 2000). An increase in temperature will lower the viscosity. The practical limit for handling the liquor is the pumping limit of 300 – 500 mPa·s. The viscosity at every evaporator effect must always be below this level, and is usually much lower in the low dry solids effects of an evaporation plant.

A heat treatment process can also reduce the viscosity. When black liquor temperature is above 120 – 130 °C and it contains residual alkali, the cooking reactions continue.

These reactions break the long organic molecules and polysaccharides into shorter forms and results in an irreversible reduction of the liquor viscosity.

It is expected that the final concentration and the evaporation capacity are not affected because of effects of lignin removal on black liquor viscosity.

### **3.3.3 Effect of pH to evaporator operation**

To remove the lignin black liquor pH must be lowered. This causes the lignin to separate. With lignin other undesirable components e.g. silica are precipitated. These must be separated from lignin by washing.

If separated residue from lignin removal is evaporated, then the pH must be increased by addition of alkali. Alternatively green liquor could be used. Not much trial evaporation has been done of the residue, but it is expected that this can be handled in standard evaporation without significant modification. The evaporation load increases as the washing media, which can be partly secondary condensate must be re-evaporated

### **3.4 Effect of lignin removal to recovery boiler operation**

Removal of lignin changes black liquor properties and reduces the flow of organics to recovery boiler. This means that recovery boiler load decreases. Decreasing load leads to decreasing superheating. If recovery boiler load decrease is extensive then lower furnace temperatures are affected resulting in unsatisfactory operation.

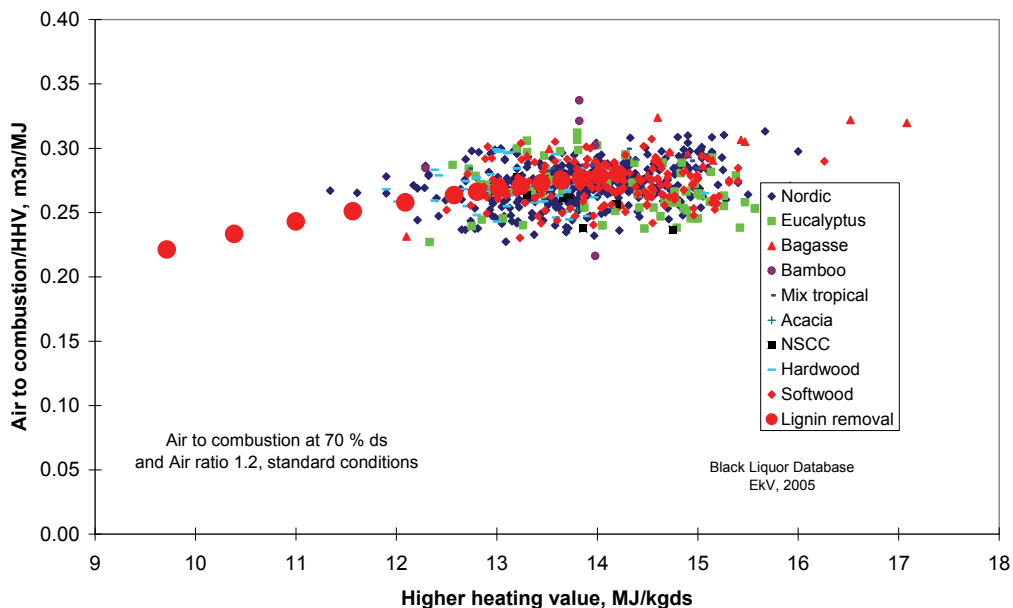
#### **3.4.1 Effect of lignin removal to recovery boiler flows**

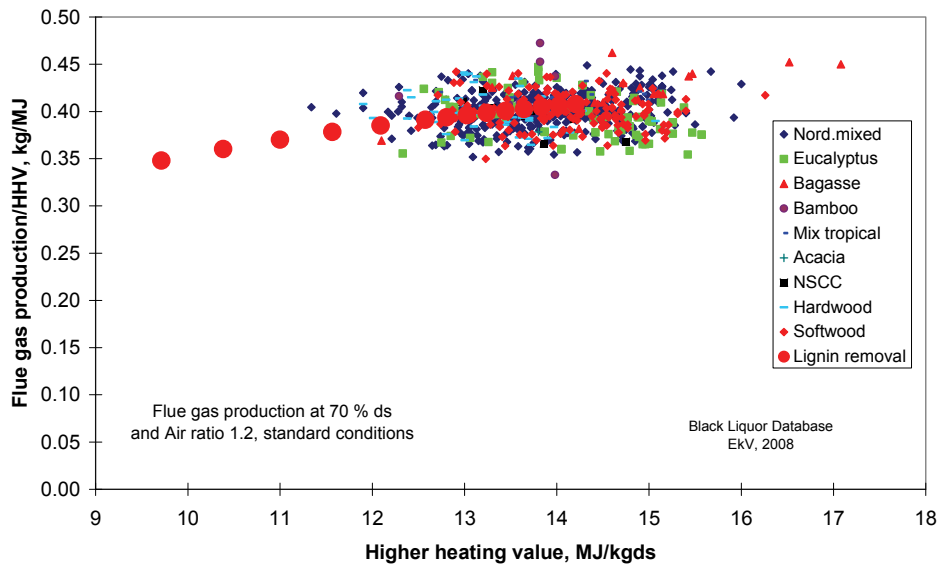
The virgin black liquor elemental compositions were used to calculate recovery boiler flows. For each black liquor composition the corresponding air flows and flue gas flows were calculated based on mass balances. Additionally lower furnace temperatures, ash formation rates and the recycle ash rates were calculated based on ash properties prediction (Vakkilainen 2000). The results are summarized at table 7 and Appendix Ie.

**Table 7.** The furnace behaviour at various lignin removal rates

Lignin removal		0 %	5 %	10 %	15 %	20 %	25 %	30 %
NetHeat	kW/kgds	8959	8772	8580	8382	8178	7968	7751
Air at 1.164	m <sup>3</sup> n/kgds	3.585	3.507	3.427	3.345	3.260	3.172	3.082
FG at 1.164	m <sup>3</sup> n/kgds	4.317	4.231	4.143	4.052	3.958	3.861	3.761
FurnH/FG	kW/m <sup>3</sup> n	2660	2665	2669	2675	2680	2686	2693
Fg	m <sup>3</sup> n/s	170.0	164.1	158.2	152.3	146.4	140.6	134.7
Air	m <sup>3</sup> n/s	121.3	116.8	112.4	108.0	103.6	99.2	94.8
LFtemp	°C	1136	1142	1142	1142	1142	1142	1142
BBash	g/m <sup>3</sup> n,dry	3.7	3.7	3.7	3.7	3.7	3.7	3.7
ESPash	g/m <sup>3</sup> n,dry	22.8	22.8	22.8	22.8	22.8	22.8	22.8
Recycle ash	kg/kgds	0.08	0.08	0.08	0.07	0.07	0.07	0.07
Lignin removal		35 %	40 %	50 %	60 %	70 %	80 %	90 %
NetHeat	kW/kgds	7527	7296	6810	6291	5735	5137	4492
Air at 1.164	m <sup>3</sup> n/kgds	2.988	2.892	2.690	2.474	2.242	1.993	1.724
FG at 1.164	m <sup>3</sup> n/kgds	3.658	3.552	3.329	3.090	2.834	2.560	2.263
FurnH/FG	kW/m <sup>3</sup> n	2700	2708	2726	2748	2775	2811	2860
Fg	m <sup>3</sup> n/s	128.8	123.0	111.3	99.7	88.1	76.5	64.9
Air	m <sup>3</sup> n/s	90.4	86.0	77.3	68.5	59.8	51.2	42.5
LFtemp	°C	1142	1136	1114	1092	1069	1047	1025
BBash	g/m <sup>3</sup> n,dry	3.7	3.7	3.7	3.7	3.7	3.7	3.7
ESPash	g/m <sup>3</sup> n,dry	22.8	22.6	20.7	18.9	16.8	14.9	12.0
Recycle ash	kg/kgds	0.06	0.06	0.05	0.04	0.03	0.02	0.02

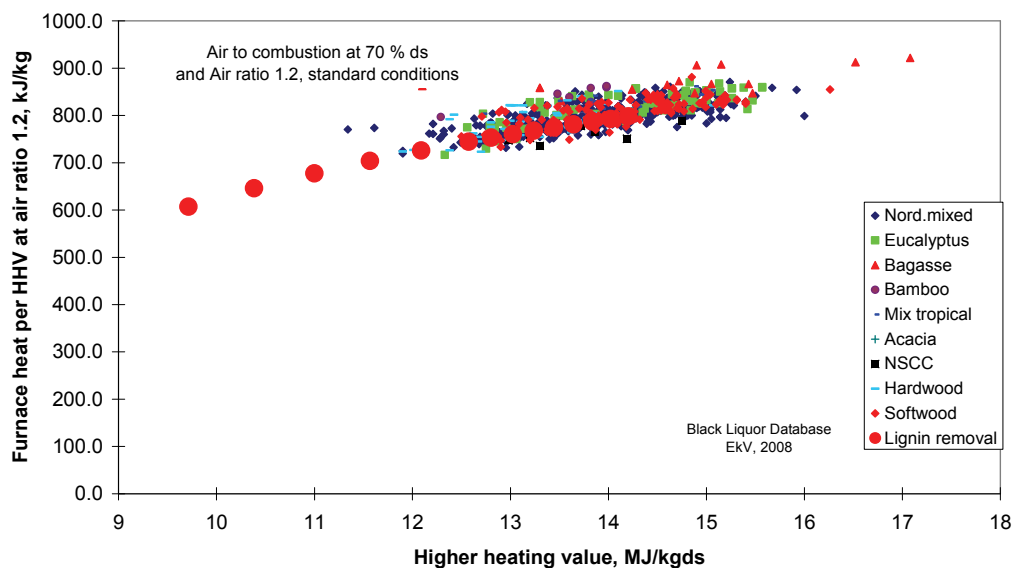
With increased lignin removal the air requirement at air ratio of 1.164 per mass unit of black liquor decreases. Similarly decreases the air requirement per unit of heat liberated, figure 12. Even though the air requirement decreases it still remains in the area which is typical for black liquors in general.

**Figure 12.** Air requirement at various lignin removal rates compared to general trend in black liquors



**Figure 13.** Flue gas at various lignin removal rates compared to general trend in black liquors

With increased lignin removal the flue gas generation at air ratio of 1.164 per mass unit of black liquor decreases. Similarly decreases the flue gas generation per unit of heat liberated, figure 13. Even though the flue gas generation decreases it still remains in the area which is typical for black liquors in general.



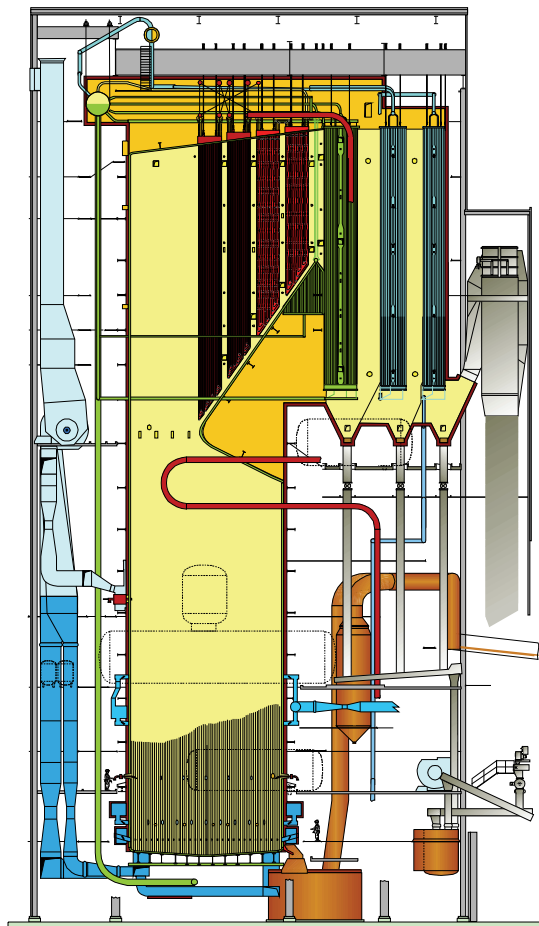
**Figure 14.** Furnace heat at various lignin removal rates compared to general trend in black liquors

Removal of lignin reduces the heat available at furnace, figure 14. The portion of heat in black liquor that is needed for reduction increases. Therefore the net heat available in furnace per unit mass of black liquor decreases even more. It can be seen that at high lignin removal rates the heat available at furnace is less than is typical for black liquors. The difference to the general trend is not very high.

Predictions for lower furnace temperature and ash generation indicate that at lignin removal rate of about 50 %, furnace behavior starts deviating significantly from those conditions that normally occur in recovery boilers.

The predicted lower furnace temperature and ash generation rates are correlations from mill data. As the higher heating value of lignin removed black liquor falls below 11 MJ/kg<sub>sd</sub>, the mill operational data does not exist anymore. So therefore the predictions should be looked more as trends not absolute values.

It can be concluded that lignin removal does not significantly affect the burning properties of black liquor. Even though for the same heat requirement more black liquor needs to be burned, the air and flue gas flows are close to the original values. Current recovery boilers are therefore suitable for lignin removed liquor combustion without significant retrofits.



**Figure 15.** Side view of the studied recovery boiler

### 3.4.2 Effect of lignin removal to recovery boiler operation

The virgin black liquor elemental compositions were used to calculate recovery boiler operation. For each black liquor composition the corresponding operation was calculated using boiler calculation based on Vakkilainen 1993.

This fairly modern recovery boiler, figure 15, has the following features;

- One drum boiler with 3-part superheater and water screen (optional)
- Steam design data 9.2 MPa / 490 °C
- Design black liquor dry solids 80 % with pressurized heavy liquor storage tank
- Liquor temperature control with flash tank, indirect liquor heaters for backup
- DNCG burning in the boiler
- Low emissions of TRS, SO<sub>2</sub> and particulates
- Flue gas cleaning with ESP (no scrubbers)
- Lower furnace cross section of 166.8 m<sup>2</sup>

The calculated recovery boiler operation data is summarized at table 8 and Appendix If. As lignin is removed the organic content in the black liquor decreases. This is seen as decreasing massflow to the recovery boiler. The hearth heat release rate decreases faster than the massflow as the removed lignin has higher heating value than the black liquor average. Therefore the recovery boiler steam generation also drops faster than the firing rate indicates.

**Table 8.** The recovery boiler operation at various lignin removal rates

Lignin removal		0 %	5 %	10 %	15 %	20 %	25 %	30 %
Massflow (virgin)	tDS/24h	3157	3105	3052	3000	2948	2895	2843
HHR	kW/m <sup>2</sup>	3112	3021	2931	2840	2750	2660	2569
Steam	kg/s	128.3	124.0	119.6	115.2	111.3	107.3	103.5
Nose	°C	974.5	955.7	936.9	918.2	899.4	880.4	861.8
SH out	°C	584.2	573.8	563.1	552.6	541.4	529.7	518.6
BB out	°C	446.2	439.0	431.7	424.5	417.2	409.6	402.4
Eco out	°C	168.9	166.7	164.6	162.5	160.3	158.1	155.9
Steam out	°C	490.0	490.0	490.0	490.0	486.7	480.8	474.9
DeSH	°C	22.7	16.1	9.6	3.1	-3.2	-9.2	-15.1
dP tot	bar	14.7	14.0	13.3	12.7	12.0	11.3	10.6
Lignin removal		35 %	40 %	50 %	60 %	70 %	80 %	90 %
Massflow (virgin)	tDS/24h	2791	2738	2633	2529	2424	2319	2214
HHR	kW/m <sup>2</sup>	2479	2389	2209	2029	1850	1671	1492
Steam	kg/s	99.5	95.4	87.2	78.8	70.1	61.2	52.0
Nose	°C	843.2	824.3	786.9	749.6	712.3	675.2	638.2
SH out	°C	507.1	495.2	473.0	450.9	429.2	407.9	387.7
BB out	°C	395.1	387.5	373.6	360.0	346.9	334.5	323.3
Eco out	°C	153.7	151.5	147.3	143.2	139.2	135.4	131.8
Steam out	°C	468.9	463.0	451.3	440.5	430.2	420.3	411.3
DeSH	°C	-21.1	-27.0	-38.7	-49.5	-59.8	-69.7	-78.7
dP tot	bar	10.0	9.3	8.2	7.1	6.1	5.2	4.4

With decreasing firing rates the flue gas temperatures at various surfaces decrease correspondingly. At about 50 % lignin removal rate the flue gas temperature at the nose and the hearth heat release rate get lower than what is normally considered the recovery

boiler minimum load, so predictions with very high lignin removal rates have mostly academic interest.

The same steam loads were run using the no lignin removal black liquor composition. This recovery boiler part load operation data is summarized at table 9 and Appendix Ig.

**Table 9.** The recovery boiler operation at various steaming rates corresponding to respective lignin removal rates

<b>Boiler load</b>		100 %	97 %	93 %	90 %	86 %	82 %	79 %
Massflow (virgin)	tDS/24h	3157	3049	2939	2830	2717	2602	2494
HHRR	kW/m <sup>2</sup>	3112	3005	2896	2789	2678	2564	2458
Steam	kg/s	128.3	124.0	119.6	115.2	111.3	107.3	103.5
Nose	°C	974.5	956.0	937.4	918.9	899.8	880.4	862.1
SH out	°C	584.2	573.7	563.1	552.6	540.3	527.9	516.2
BB out	°C	446.2	438.9	431.7	424.4	416.5	408.5	401.0
Eco out	°C	168.9	166.7	164.5	162.3	159.9	157.4	155.2
Steam out	°C	490.0	490.0	490.0	490.0	484.3	477.2	470.6
DeSH	°C	22.7	14.9	8.4	1.3	-5.7	-12.8	-19.4
dP tot	bar	14.7	13.9	13.2	12.4	11.6	10.9	10.1
<b>Boiler load</b>		76 %	72 %	65 %	58 %	51 %	44 %	38 %
Massflow (virgin)	tDS/24h	2384	2272	2053	1837	1618	1402	1186
HHRR	kW/m <sup>2</sup>	2350	2239	2023	1810	1595	1382	1169
Steam	kg/s	99.5	95.4	87.2	78.8	70.1	61.2	52.0
Nose	°C	843.4	824.2	787.3	750.7	713.5	677.0	640.5
SH out	°C	504.4	492.3	469.2	446.4	423.7	401.6	379.5
BB out	°C	393.4	385.8	371.3	357.5	343.8	331.2	318.6
Eco out	°C	152.8	150.5	146.0	141.9	137.5	133.4	129.3
Steam out	°C	464.0	457.3	444.7	432.5	420.3	408.7	397.1
DeSH	°C	-26.0	-32.7	-45.3	-57.5	-69.7	-81.3	-92.9
dP tot	bar	9.4	8.7	7.4	6.1	5.0	4.0	3.0

Because recovery boiler heat transfer depends mostly on flue gas flow the calculated thermal behaviour at similar steaming rates is fairly close. The lignin removed black liquor is able to keep superheating closer to original than the corresponding partial load operation. Therefore it should be easier to run lignin removed black liquor than less liquor at the same steaming rate.

The heart heat release rate at base case (0 % lignin removal) is above 3100 kW/m<sup>2</sup>. This means that the boiler is at base case very close to its maximum capacity. It should be noted that steaming rates below 70 % of nominal are only run occasionally. So any prediction that has steaming rate of 70 kg/s is of academic interest only.



When comparing the two cases (same steam rate with lignin removal and with load reduction) we can summarize

- Lignin removal increases the dry solids rate. As organics are reduced but inorganics remain, then for the same steam load more lignin removed solids needs to be fired
- Flue gas side temperatures; furnace nose, superheater out, boiler bank out, eco out remain roughly the same
- Lignin removal retains superheating a bit better than does corresponding load reduction.

For the specific boiler in question we note

- At lignin removal rate of roughly 20 % the boiler superheating limit is reached. Typically at lower loads auxiliary fuel needs to be added to make full superheating
- Lower furnace will start behaving problematically (TRS, SO<sub>2</sub>, reduction) at about 30 % lignin removal rate
- Minimum load that the boiler can run corresponds to about 50 % lignin removal

It should be noted that all of these values can be somewhat altered with investing to higher dry solids and/or to a new air system. Increasing the production of the mill and correspondingly increasing the recovery boiler production will improve the situation and allow higher lignin removal rate.

### 3.5 Effect of lignin removal to lime kiln operation

Lime kiln load does not change if part of the organic load is removed from black liquor. So lignin removal does not affect the lime kiln per se. An interesting option is to make lime kiln fossil fuel independent. Then either the lime kiln is fired by gasified biomass (biogas) or with removed lignin.

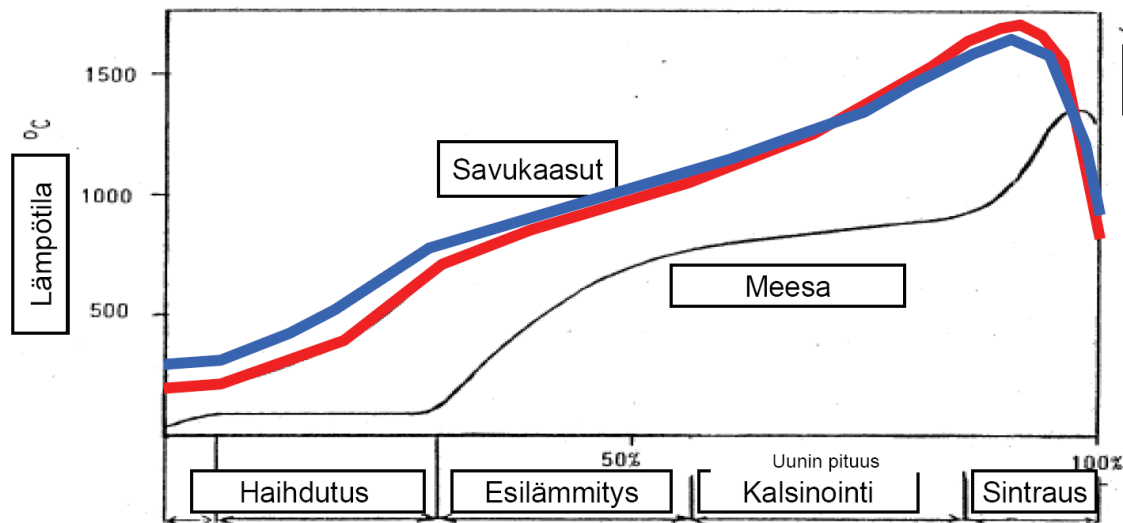
#### 3.5.1 Burning lignin in lime kiln

Lime kilns use mainly natural gas and heavy fuel oil to dry, heat and calcine the lime mud. The example mill uses natural gas in the lime kiln. Lime kiln operation depends on the fuel used. Typical lime kiln fuels are compared to oil and natural gas in table 10. The values for gasified bark (biogas) are from Isaksson (2007).

**Table 10.** Properties of lime kiln fuels

		Oil	Natural gas	Biogas	Lignin
Moisture	%	~0.1	0	10	30
Effective heating value	MJ/kg	40	50	16.7	17.1
Adiabatic combustion temp.	°C	2210	2050	1870	1980
Air requirement	m <sup>3</sup> n/kgpa	12.2	14.9	5.2	5.3
Flue gas production	m <sup>3</sup> n/kgpa	13.4	15.7	5.9	6.0
Energy consumption	GJ/t	5.92	6.17	6.46	6.34
Flue gases	m <sup>3</sup> n/t	3250	3390	3500	3430
CO <sub>2</sub> - production	kg CO <sub>2</sub> /t	466	340	~0	~0

The heating values for lime kiln biofuels are lower than for natural gas or heavy fuel oil. This in itself does not mean much. Lower adiabatic temperatures and larger flue gas flows mean that the maximum temperature in the lime kiln is lower.



**Figure 16.** Lime kiln flue gas temperature profiles (Isaksson 2007). Red is oil firing, blue is biogas.

Figure 16 shows kiln temperature profiles for two cases oil (red) and biogas (blue). Firing biogas produces lower firing end temperatures. Because the capacity of the lime kiln is determined by the heat exchange from flue gas to solids, same capacity requires higher firing rates for biogas and results in higher drying end temperatures.

Lignin as a lime kiln fuel seems to be between biogas and natural gas. So the kiln capacity is lowered by 5 to 10 %. Because the flue gas losses increase (back end temperature increases) the fuel consumption increases 3 % (as heat).

### 3.5.2 Ringing in a lime kiln burning lignin

Lime kiln main operational problem is formation of hard deposits called rings that have to be removed manually. Lime kiln rings are connected to variability of fuel heating value or mass flow, free sodium and free sulfur in fuel.

If the washing of lignin is done to degree done in e.g. LignoBoost project the free sodium and sulfur should not be a problem.

In trials to be conducted the stability of lignin heating value needs to be tested, but it can be estimated that this is not a major problem.

## 4 MILL MAIN BALANCES

Mill main balances were calculated to find out how lignin removal affects the kraft mill balances especially the recovery boiler and the evaporator.

The main balances were calculated using MILLFLOW excel spreadsheet as basis for balances.

### 4.1 Studied cases

The studied cases are concepts for non fossil fuel kraft pulp mill. The biofuels production alternatives are washed lignin and biomass (wood/bark) gasification. The biomass needs to be suitable for usage (Na-content, NPE).

The cases studied are

1. 600 000 ADt/d mill with conventional operation (Base case)
2. 600 000 ADt/d mill which uses (wood/bark) gasification for lime kiln fuel (LKgasif case)
3. 600 000 ADt/d mill which uses extracted lignin for lime kiln fuel (LKlignin case)
4. Maximum lignin extraction with recovery boiler (evaporator) as a bottleneck (no auxiliary fuel used); case 4a 600 000 ADt/a (LigMax600 case), case 4b 704 000 ADt/a (LigMax704 case)
5. Maximum lignin extraction with recovery boiler (evaporator) as a bottleneck (gasified wood/bark as auxiliary fuel); 600 000 ADt/a (LigAux600 case)
6. For comparison a high lignin removal case with auxiliary firing (lower furnace organic rate similar to base case but with auxiliary firing) 930 000 ADt/a (LigAux930 case)

### 4.2 Effect of lignin removal to pulp mill main balances

For each case a full pulp mill balance was calculated using MILLFLOW spreadsheet. The results are summed up in the Table 11.

As can be seen the fiberline operation remains essentially unchanged if the mill production is kept constant. For higher pulp production rates the fiberline production increases proportionally.

For recovery operation the causticizing and the lime remain the same for constant pulp production. For increased production rates they increase.

**Table 11.** Pulp mill main balances for the studied cases

		Base/ LKgasif	LKlignin	LigMax 600	LigMax 704	LigAux 600	LigAux 930
<b>Pulp</b>	t/a	600000	600000	600000	704000	600000	930000
<b>Lignin removal rate</b>	%	0	14	20	20	50	50
	t/d	0	146.7	190.5	209.6	524.0	812.1
<b>Fiberline</b>							
Woodroom	M <sup>3</sup> n/sub	6697	6697	6697	7858	6697	10380
Digester	BDt/d	1668	1668	1668	1957	1668	2586
Oxygendelignification	BDt/d	1594	1594	1594	1870	1594	2471
Screening	BDt/d	1668	1668	1668	1957	1668	2586
Pulpwashing	BDt/d	1668	1668	1668	1957	1668	2586
Bleaching	BDt/d	1543	1543	1543	1810	1543	2391
Pulpdryer	BDt/d	1543	1543	1543	1810	1543	2391
<b>Recovery</b>							
Evaporation	t H <sub>2</sub> O/d	16218	17349	16999	19903	15250	23504
Recovery boiler	tDS/d	3543	3401	3340	3845	3035	4472
Causticizing	m <sup>3</sup> WL/d	6683	6683	6683	7842	6683	10359
Lime kiln	t lime/d	548	548	548	643	548	849
<b>Water and effluent</b>							
Water treatment	m <sup>3</sup> /d	46286	46286	46286	54309	46286	71743
Effluent treatment	m <sup>3</sup> /d	39429	39429	39429	46263	39429	61115
<b>Bleaching chemicals</b>							
Chlorine Dioxide	tClO <sub>2</sub> /d	17	17	17	20	17	26
Peroxide	tH <sub>2</sub> O <sub>2</sub> /d	5	5	5	6	5	8
Oxygen	tO <sub>2</sub> /d	35	35	35	41	35	54
Caustic	tNaOH/d	17	17	17	17	17	17
Sulphuric acid	tH <sub>2</sub> SO <sub>4</sub> /d	15	15	15	15	15	15
<b>Lignin chemicals</b>							
Carbon dioxide	tCO <sub>2</sub> /d	0	44.0	57.2	62.9	243.6	157.2
Caustic	tNaOH/d	0	10.3	13.3	14.7	56.8	36.7
Sulphuric acid	tH <sub>2</sub> SO <sub>4</sub> /d	0	13.2	17.1	18.9	73.1	47.2

The fiberline production rates for cases LigMax704 and LigAux930 are very high and fall clearly outside normal retrofit areas (+5 ... 10 %). These high production cases show what is the effect of lignin removal.

At 20 % lignin removal one can dimension the recovery boiler 1/7 smaller than without lignin removal. For mill of 700 000 ADt/a this means investment savings of 20 M€.

Case LigAux930 corresponds to the size of mills one is currently thinking of building in Europe. The recovery boiler size with 50 % lignin removal would correspond roughly to 75 % of boiler size without lignin removal. This would mean investment cost decrease of 42 M€.

In both cases above the actual lignin removal equipment would increase the cost for the mill. However it can be assumed that the decrease in recovery boiler cost would cover the increase in lignin removal equipment cost.

## 5 MILL ENERGY BALANCES

The main balances were calculated using MILLFLOW excel spreadsheet as basis for balances. The mill condensing tail flow was set to balance the steam balance. The example pulp mill has condensing tail in its turbine. Most of the older pulp mills do not have a condensing tail, but run only backpressure operation. It should be noted that in the example mill there is no bark boiler. Recovery boiler is the sole steam producer in the mill. The bark is sold to outside of the mill.

The individual department operation in different cases is seen in table 10. The summary of energy balance for each case is seen in Table 12. For each case a more detailed steam and energy balance is listed in appendix II.

**Table 12.** Pulp mill steam and electricity balances for the studied cases

		Base	LKgasif	LKlignin	LigMax 600	LigMax 704	LigAux 600	LigAux 930
Lignin removal rate	%	0	0	14	20	20	50	50
Steam usage								
Pulp mill steam usage total	kg/s	94.2	94.2	96.2	96.1	111.1	90.5	138.1
Steam production								
Recovery Boiler	kg/s	128.3	128.3	116.1	111.1	128.0	103.6	154.0
Condensing tail	kg/s	-31.0	-31.0	-17.4	-13.6	-14.2	-11.3	-12.9
Spraying	kg/s	6.0	6.0	6.0	6.0	6.8	6.0	8.7
To feedwater preheating	kg/s	-7.6	-7.6	-8.5	-7.4	-9.5	-7.8	-11.6
Electricity usage								
Pulp mill total usage	MW	53.9	56.1	53.7	53.9	59.4	53.4	70.1
Electricity production								
Condensing turbine	MW	28.5	28.5	16.0	12.5	13.7	10.4	11.9
Back pressure turbine	MW	64.9	64.9	65.9	65.1	78.3	61.6	94.1
Electricity production total	MW	93.4	93.4	81.9	77.6	91.9	72.0	106.0
Electricity for sale	MW	39.5	37.3	28.2	23.7	32.5	18.6	35.9

In the most drastic lignin removal i.e. auxiliary firing cases (LigAux600 and LigAux930) the recovery boiler was topped with 20 % of heat in recovery boiler furnace produced with auxiliary fuel firing. It has been assumed that the extra fuel could be generated by gasifying biomass. In all lignin removal cases (except Base and LKgasif) the lime kiln has been fired with lignin.

In all cases the pulp mill produces more electricity than it consumes. This is because the mill is modern and thus energy efficient i.e. it has steam surplus. The second reason for high electricity generation is high steam parameters. The third reason is extra electricity generation done with condensing tail in the steam turbine.

Pulp mill steam usage depends mainly on the production rate. It has been assumed that no primary steam is needed to dry bark/lignin. Instead secondary heat can be used for this purpose. In lignin removal cases the steam usage increases mainly because of the extra load in the evaporator. For high pulp production cases the extra steam usage is mainly because of the increased capacity.

As can be seen the condensing tail capacity decreases with increasing lignin removal. This decreases the electricity available for sale. The difference between 20 % lignin removal case and base case is 15.8 MW.

## 6 ECONOMICS OF INCREASES BIOMASS USAGE

Economics of lignin removal were studied by calculating the marginal operating costs for the studied cases. In the simple economic balance the following changes were taken into account;

- Revenue from additional pulping
- Revenue from decreased lime kiln fuel usage
- Revenue from lignin sales vs
- Lost power production
- Lost bark sales
- Increased lime mud dumping (make-up costs)

**Table 13.** Data for pulp mill base economics for the studied cases

		Base	LKgasif	LKlignin	LigMax 600	LigMax 704	LigAux6 00	LigAux 930
<b>Lignin removal rate</b>	%	0	0	14	20	20	50	50
Pulp	t/a	600000	600000	600000	600000	704000	600000	930000
Operation days	d/a	350	350	350	350	350	350	350
Lignin removal rate	BDt/d	0	0	146.7	190.5	209.6	524	812.1
Lignin for sale	t/a	0	0	0	21900	31450	188650	332700
Lime kiln	t lime/d	548	548	548	548	643	548	849
Lime kiln fuel	MWh/a	316680	316680	316680	316680	371579	316680	490623
Lime kiln CO <sub>2</sub>	t/a	62748	0	0	0	-10876	0	-34512
Bark production	BDt/d	564	564	564	564	661	564	873
Bark for sale	BDt/a	197400	116900	197400	197400	231420	197400	305690
Lime dumping	kg/ADt	20	28	21	21	21	21	21
Extra fuel	MW	0	0	0	0	0	59	88

The assumed costs for different cost components are

- Pulp (marginal price for additional tons) 200 €/ADt
- Chlorine dioxide 400 €/t
- Peroxide 80 €/t
- Oxygen 40 €/t
- Carbon dioxide 60 €/t
- Caustic 240 €/t
- Sulphuric acid 70 €/t
- Electricity 40 €/MWh
- Bark 8 €/MWh
- CO<sub>2</sub> credits 30 €/t
- Natural gas 35 €/MWh
- Gasified biomass 12 €/MWh
- Lignin price 57 €/t (~12 €/MWh)
- Quicklime 30 €/t

It has been assumed that lignin fetches higher price sold than bark. This is based on the fact that lignin is drier and has higher heating value than bark. An indication of the

magnitude of cost change can be seen in Table 14. There the costs associated with each cost item have been compared to the cost at base case.

**Table 14.** Cost difference data for pulp mill base economics for the studied cases (base case is the case for comparison)

		Base	LKgasif	LKlignin	LigMax600	LigMax704	LigAux600	LigAux930
<b>Lignin removal</b>	%	0	0	14	20	20	50	50
Pulp	t/a	60000	600000	600000	600000	704000	600000	930000
Lime kiln fuel	k€/a	0	11100	11100	11100	13024	11100	17205
Electricity	k€/a	0	-800	-3800	-5300	-4705	-7100	-8515
CO <sub>2</sub> credits	k€/a	0	1900	1900	1900	2200	1900	2900
Bark sales	k€/a	0	-2900	0	0	1200	0	3900
Extra fuel	k€/a	0	0	0	0	0	-5900	-8900
Pulp sales	k€/a	0	0	0	0	20800	0	66000
Lignin sales	k€/a	0	0	0	1200	1800	10800	19000
Chemicals	k€/a	0	-200	-1600	-2000	-2800	-8400	-7300
Total	k€/a	0	9100	7600	6900	31519	2400	84290

It seems that for operating costs all cases provide profitability compared to base case. Increased production is responsible for more than half of the additional revenue in cases where additional pulp is produced.

With costs used the price from bark really is not significant in the mill economics. Neither is the cost associated with decrease in electricity sales. Savings in lime kiln fuel are significant.

Creating a fossil fuel free pulp mill by taking out lignin to replace lime kiln fuel seems a very profitable and attractive idea. The main financial benefits are from the lowered operating costs and from debottlenecking the mill. Lowered operating costs are almost solely based on price of natural gas, which has been climbing up and which is not expected to decrease in the near future. If recovery boiler is the bottleneck of the mill, then this method also allows further increase of pulp production.

It seems that taking out lignin to reduce the recovery boiler load is not itself very attractive. Case LigMax600 does not bring in significant extra revenue. This is because the price for the sold lignin is rather low. This mirrors the conclusions of Olsson et al. 2006. If pulp production capacity can be increased, then this brings in quite a lot of new revenue. This is seen in LigMax704 case.

The 930 000 ADt pulp mill, with lignin removal is very attractive especially in Russia, where biomass fuels are available. The profitability of this kind of operation is heavily dependent on whether one can sell the lignin that has been taken out.

## 7 CONCLUSIONS

The removal of lignin decreases the organic content of black liquor, but the inorganic portion remains essentially unchanged. The heating value of the black liquor decreases with increased lignin removal.

The new process for extracting lignin from black liquor should not affect the BPR of black liquor very much. Removal of high molecular mass lignin affects only marginally the effective average molecular weight of the non water black liquor portion. The new process for extracting lignin from black liquor may be an opportunity for decreasing the viscosity of black liquor as lignin removal removes high molecular mass components which are a significant source of black liquor viscosity.

When comparing recovery boiler operation at the same steam rate with lignin removed black liquor and with load reduction we can summarize

- Lignin removal increases the apparent dry solids firing rate. As organics are reduced but inorganics remain, then for the same steam load more lignin removed solids needs to be fired
- Flue gas side temperatures; furnace nose, superheater out, boiler bank out, eco out remain roughly the same
- Lignin removal retains superheating a bit better than does corresponding load reduction.

For the specific boiler in question we note

- At the same steam generation rate, smelt and black liquor flows increase.
- At lignin removal rate of roughly 20 % the boiler superheating limit is reached. Typically at lower loads auxiliary fuel needs to be added to make full superheating
- Lower furnace will start behaving problematically (TRS, SO<sub>2</sub>, reduction) at about 30 % lignin removal rate
- Minimum load that the boiler can run corresponds to about 50 % lignin removal

It should be noted that all of these values can be somewhat altered with investing to higher dry solids and/or to a new air system.

Creating a fossil fuel free pulp mill by taking out lignin to replace lime kiln fuel seems a very profitable and attractive idea. The main financial benefits are from the lowered operating costs and from debottlenecking the mill. Lowered operating costs are almost solely based on price of natural gas, which has been climbing up and which is not expected to decrease in the near future. If recovery boiler is the bottleneck of the mill, then this method also allows further increase of pulp production.

It seems that taking out lignin to reduce the recovery boiler load is not itself very attractive. Case LigMax600 does not bring in significant extra revenue. This is because the price for the sold lignin is rather low. If pulp production capacity can be increased, then this brings in quite a lot of new revenue. This is seen in LigMax704 case.

The 930 000 ADt pulp mill, with lignin removal is very attractive especially in Russia, where biomass fuels are available. The profitability of this kind of operation is heavily dependent on whether one can sell the lignin that has been taken out.



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**APPENDICES**

- I Effect of lignin removal
- Ia Mill operational criteria
- Ib Prediction of dust and ash
- Ic Effect of lignin removal to Black liquor
- Id Effect of lignin removal to Black liquor elemental analysis
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- II Mill main and energy balances
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**Appendix Ia, Mill operational criteria**

<b>Wood species in cooking</b>	Pine	
Air dry unbleached cellulose	1776	t/24 h
Chips to digester	3627	BDt/d
Wood moisture	45.0	%
Yield in Oxygen delignification	99	%
Yield from wood to dry unbleached cellulose	46	%
Pulp kappa before oxygen delignification	30	
Pulp kappa after oxygen delignification	14	
Washing efficiency unbleached pulp	99	%
Washing efficiency bleached pulp	99	%
Soap recovery efficiency	70	%
Water created by neutralization of acids in wood	100	kg/ADt
Sulfur in NCG	2	kg/ADt
<b>White liquor</b>		
Active alkali, Na <sub>2</sub> O-% per wood	17.5	%
White liquor sulfidity	38.7	%
White liquor dry solids	15	%
Causticity	81	%
Reduction in white liquor	95	%
Black liquor dry solids	78	%
<b>Oxidized White liquor</b>		
Oxidized White liquor used in Oxygen delignification	37	kg/ADt
Methanol formation in Oxygen delignification	0.2	kg/ADt

## Appendix Ib, Prediction of dust and ash

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### Enrichment

### Nordic BSK

#### Virgin liquor analysis

	Design value	Estimate from HHV
HHV MJ/kgds	14.16	14.2
C mass-%, dry	34.70	34.7
H mass-%, dry	3.50	3.5
N mass-%, dry	0.08	0.1
S mass-%, dry	5.40	5.4
Na mass-%, dry	20.00	18.6
K mass-%, dry	2.20	2.2
Cl mass-%, dry	0.15	0.2
B mass-%, dry	0.00	0.0
Inorganics mass-%, dry	27.8	26.3
S/(Na2+K2) mol-%	36.4	38.9
S/Na2 mol-%	38.7	41.6
Cl/(Na+K) mol-%	0.46	0.49
K/(Na+K) mol-%	6.07	6.50
B/Na mol-%	0.00	10.81
Dry solids %	78	78
FurnHeat kW/kgds	12606	12638
NetHeat kW/kgds	10016	10070
Air at 1.0 m3n/kgds	3.06	2.98
Air at 1.0 kg/kgds	3.956	3.857
Air/HHV m3n/MJ	0.216	0.211
FG at 1.2 m3n/kgds	4.51	4.35
m3n(dry)/kgds	3.60	3.82
FG/HHV m3n/MJ	0.318	0.307
FG/NHHV m3n/MJ	0.450	0.432
FurnH/FG kW/m3n	2798	2905
SO2 ppm	0	<-predicted
HCl ppm	0	<-predicted
	164.9	
LFtemp oC	1128	<-predicted
Furnace lo kW/m2	3300.0	
NCG-load gS/kgds	4.0	
BBash g/m3n, dry	3.7	<-predicted
kg/kgds	0.013	
ESPash g/m3n, dry	22.2	<-predicted
kg/kgds	0.080	
Dry solids % - as fired	79.5	

### As fired analysis

HHV	MJ/kgds	12.95
C	m-%	31.9
H	m-%	3.2
N	m-%	0.07
S	m-%	6.6
Na	m-%	20.7
K	m-%	2.6
Cl	m-%	0.2
B	m-%	0.0
O	m-%	34.7
Cl/(Na+K)	mol-%	0.61
K/(Na+K)	mol-%	6.90
S/(Na2+K2)	mol-%	42.19
E Cl		1.35
e K		1.14
T(Firstm)		598.3
T15		810.8
T70		841.6
Sintering index		300.00
SH Front deposit	(=carryo.)	
Na	m-%	34.28
K	m-%	3.69
Cl	m-%	0.12
CO3	m-%	24.58
SO4	m-%	37.22
S	m-%	0.10
BO2	m-%	0.00
(Na2+K2)SO4	m-%	55.79
(Na2+K2)CO3	m-%	44.21
(Na2+K2)Cl	m-%	0.20
(Na2+K2)S	m-%	0.25
NaBO2	m-%	0.00
Cl/(Na+K)	mol-%	0.21
K/(Na+K)	mol-%	5.95
S/(Na2+K2)	mol-%	49.30
CO3/(Na2+K2)	mol-%	51.67
B/Na	mol-%	0.00
E Cl		0.34
e K		0.86
T(Firstm)		525.4
T15		789.0
T70		799.6
Sintering index		1518
Initial density kg/m3		1516
Density at 12 kg/m3		2090
Strength at 1:MPa		30.7
pH		12.36

### ESP ash (=fume)

Na	m-%	28.01
K	m-%	7.45
Cl	m-%	0.96
CO3	m-%	4.32
SO4	m-%	59.16
S	m-%	0.10
BO2	m-%	0.00
(Na2+K2)SO4	m-%	90.16
(Na2+K2)CO3	m-%	7.94
(Na+K)Cl	m-%	1.65
(Na2+K2)S	m-%	0.26
NaBO2	m-%	0.00
Cl/(Na+K)	mol-%	1.93
K/(Na+K)	mol-%	13.52
S/(Na2+K2)	mol-%	87.92
CO3/(Na2+K2)	mol-%	10.21
B/Na	mol-%	0.00
E Cl		3.14
e K		1.96
T(Firstm)		523.5
T15		690.9
T70		785.2
Sintering index		1720
Initial density kg/m3		311
Density at 12 kg/m3		410
Strength at 1:MPa		0.2
pH		11.01
BB ash	(=mixture)	
Na	m-%	32.77
K	m-%	4.86
Cl	m-%	0.30
CO3	m-%	21.09
SO4	m-%	40.88
S	m-%	0.10
BO2	m-%	0.00
(Na2+K2)SO4	m-%	61.55
(Na2+K2)CO3	m-%	38.16
(Na2+K2)Cl	m-%	0.50
(Na2+K2)S	m-%	0.25
NaBO2	m-%	0.00
Cl/(Na+K)	mol-%	0.54
K/(Na+K)	mol-%	8.02
S/(Na2+K2)	mol-%	55.34
CO3/(Na2+K2)	mol-%	45.35
B/Na	mol-%	0.00
E Cl		0.88
e K		1.16
T(Firstm)		515.0
T15		755.9
T70		782.9
Sintering index		2606
Initial density kg/m3		275
Density at 12 kg/m3		2090
Strength at 1:MPa		30.7
pH		12.24







## Appendix Ie, Effect of lignin removal to furnace behavior

<b>Lignin removal</b>		<b>0 %</b>	<b>5 %</b>	<b>10 %</b>	<b>15 %</b>	<b>20 %</b>	<b>25 %</b>	<b>30 %</b>
<b>FurnHeat</b>	<b>kW/kgds</b>	11484	11274	11059	10836	10608	10372	10128
<b>NetHeat</b>	<b>kW/kgds</b>	8959	8772	8580	8382	8178	7968	7751
<b>Air at 1.164</b>	<b>m3n/kgds</b>	3.585	3.507	3.427	3.345	3.260	3.172	3.082
<b>Air/HHV</b>	<b>m3n/MJ</b>	0.218	0.216	0.214	0.211	0.209	0.207	0.204
<b>FG at 1.x</b>	<b>m3n/kgds</b>	4.317	4.231	4.143	4.052	3.958	3.861	3.761
<b>FG/HHV</b>	<b>m3n/MJ</b>	0.305	0.303	0.301	0.298	0.296	0.293	0.290
<b>FurnH/FG</b>	<b>kW/m3n</b>	2660	2665	2669	2675	2680	2686	2693
<b>Fg</b>	<b>m3n/s</b>	170.0	164.1	158.2	152.3	146.4	140.6	134.7
<b>Air</b>	<b>m3n/s</b>	121.3	116.8	112.4	108.0	103.6	99.2	94.8
<b>LFtemp</b>	<b>oC</b>	1136	1142	1142	1142	1142	1142	1142
<b>BBash</b>	<b>g/m3n,dry</b>	3.7	3.7	3.7	3.7	3.7	3.7	3.7
<b>ESPash</b>	<b>g/m3n,dry</b>	22.8	22.8	22.8	22.8	22.8	22.8	22.8
<b>Recycle ash</b>	<b>kg/kgds</b>	0.08	0.08	0.08	0.07	0.07	0.07	0.07
		<b>35 %</b>	<b>40 %</b>	<b>50 %</b>	<b>60 %</b>	<b>70 %</b>	<b>80 %</b>	<b>90 %</b>
<b>FurnHeat</b>	<b>kW/kgds</b>	9877	9618	9073	8491	7867	7196	6473
<b>NetHeat</b>	<b>kW/kgds</b>	7527	7296	6810	6291	5735	5137	4492
<b>Air at 1.164</b>	<b>m3n/kgds</b>	2.988	2.892	2.690	2.474	2.242	1.993	1.724
<b>Air/HHV</b>	<b>m3n/MJ</b>	0.201	0.198	0.192	0.184	0.175	0.164	0.151
<b>FG at 1.x</b>	<b>m3n/kgds</b>	3.658	3.552	3.329	3.090	2.834	2.560	2.263
<b>FG/HHV</b>	<b>m3n/MJ</b>	0.287	0.283	0.276	0.267	0.257	0.245	0.231
<b>FurnH/FG</b>	<b>kW/m3n</b>	2700	2708	2726	2748	2775	2811	2860
<b>Fg</b>	<b>m3n/s</b>	128.8	123.0	111.3	99.7	88.1	76.5	64.9
<b>Air</b>	<b>m3n/s</b>	90.4	86.0	77.3	68.5	59.8	51.2	42.5
<b>LFtemp</b>	<b>oC</b>	1142	1136	1114	1092	1069	1047	1025
<b>BBash</b>	<b>g/m3n,dry</b>	3.7	3.7	3.7	3.7	3.7	3.7	3.7
<b>ESPash</b>	<b>g/m3n,dry</b>	22.8	22.6	20.7	18.9	16.8	14.9	12.0
<b>Recycle ash</b>	<b>kg/kgds</b>	0.06	0.06	0.05	0.04	0.04	0.03	0.02

**Appendix If, The recovery boiler operation at various lignin removal rates**

<b>Lignin removal</b>		<b>0 %</b>	<b>5 %</b>	<b>10 %</b>	<b>15 %</b>	<b>20 %</b>	<b>25 %</b>	<b>30 %</b>
<b>Massflow (virgin)</b>	<b>tDS/24h</b>	3157	3105	3052	3000	2948	2895	2843
<b>HHR</b>	<b>kW/m<sup>2</sup></b>	3112	3021	2931	2840	2750	2660	2569
<b>Steam</b>	<b>kg/s</b>	128.3	124.0	119.6	115.2	111.3	107.3	103.5
<b>Nose</b>	<b>°C</b>	974.5	955.7	936.9	918.2	899.4	880.4	861.8
<b>SH out</b>	<b>°C</b>	584.2	573.8	563.1	552.6	541.4	529.7	518.6
<b>BB out</b>	<b>°C</b>	446.2	439.0	431.7	424.5	417.2	409.6	402.4
<b>Eco out</b>	<b>°C</b>	168.9	166.7	164.6	162.5	160.3	158.1	155.9
<b>Steam out</b>	<b>°C</b>	490.0	490.0	490.0	490.0	486.7	480.8	474.9
<b>DeSH</b>	<b>°C</b>	22.7	16.1	9.6	3.1	-3.2	-9.2	-15.1
<b>dP tot</b>	<b>bar</b>	14.7	14.0	13.3	12.7	12.0	11.3	10.6
<b>Lignin removal</b>		<b>35 %</b>	<b>40 %</b>	<b>50 %</b>	<b>60 %</b>	<b>70 %</b>	<b>80 %</b>	<b>90 %</b>
<b>Massflow (virgin)</b>	<b>tDS/24h</b>	2791	2738	2633	2529	2424	2319	2214
<b>HHR</b>	<b>kW/m<sup>2</sup></b>	2479	2389	2209	2029	1850	1671	1492
<b>Steam</b>	<b>kg/s</b>	99.5	95.4	87.2	78.8	70.1	61.2	52.0
<b>Nose</b>	<b>°C</b>	843.2	824.3	786.9	749.6	712.3	675.2	638.2
<b>SH out</b>	<b>°C</b>	507.1	495.2	473.0	450.9	429.2	407.9	387.7
<b>BB out</b>	<b>°C</b>	395.1	387.5	373.6	360.0	346.9	334.5	323.3
<b>Eco out</b>	<b>°C</b>	153.7	151.5	147.3	143.2	139.2	135.4	131.8
<b>Steam out</b>	<b>°C</b>	468.9	463.0	451.3	440.5	430.2	420.3	411.3
<b>DeSH</b>	<b>°C</b>	-21.1	-27.0	-38.7	-49.5	-59.8	-69.7	-78.7
<b>dP tot</b>	<b>bar</b>	10.0	9.3	8.2	7.1	6.1	5.2	4.4

**Appendix Ig,** The recovery boiler operation at various steaming rates corresponding to respective lignin removal rates

<b>Boiler load</b>		<b>100 %</b>	<b>97 %</b>	<b>93 %</b>	<b>90 %</b>	<b>86 %</b>	<b>82 %</b>	<b>79 %</b>
<b>Massflow (virgin)</b>	<b>tDS/24h</b>	3157	3049	2939	2830	2717	2602	2494
<b>HHRR</b>	<b>kW/m<sup>2</sup></b>	3112	3005	2896	2789	2678	2564	2458
<b>Steam</b>	<b>kg/s</b>	128.3	124.0	119.6	115.2	111.3	107.3	103.5
<b>Nose</b>	<b>°C</b>	974.5	956.0	937.4	918.9	899.8	880.4	862.1
<b>SH out</b>	<b>°C</b>	584.2	573.7	563.1	552.6	540.3	527.9	516.2
<b>BB out</b>	<b>°C</b>	446.2	438.9	431.7	424.4	416.5	408.5	401.0
<b>Eco out</b>	<b>°C</b>	168.9	166.7	164.5	162.3	159.9	157.4	155.2
<b>Steam out</b>	<b>°C</b>	490.0	490.0	490.0	490.0	484.3	477.2	470.6
<b>DeSH</b>	<b>°C</b>	22.7	14.9	8.4	1.3	-5.7	-12.8	-19.4
<b>dP tot</b>	<b>bar</b>	14.7	13.9	13.2	12.4	11.6	10.9	10.1
<b>Boiler load</b>		<b>76 %</b>	<b>72 %</b>	<b>65 %</b>	<b>58 %</b>	<b>51 %</b>	<b>44 %</b>	<b>38 %</b>
<b>Massflow (virgin)</b>	<b>tDS/24h</b>	2384	2272	2053	1837	1618	1402	1186
<b>HHRR</b>	<b>kW/m<sup>2</sup></b>	2350	2239	2023	1810	1595	1382	1169
<b>Steam</b>	<b>kg/s</b>	99.5	95.4	87.2	78.8	70.1	61.2	52.0
<b>Nose</b>	<b>°C</b>	843.4	824.2	787.3	750.7	713.5	677.0	640.5
<b>SH out</b>	<b>°C</b>	504.4	492.3	469.2	446.4	423.7	401.6	379.5
<b>BB out</b>	<b>°C</b>	393.4	385.8	371.3	357.5	343.8	331.2	318.6
<b>Eco out</b>	<b>°C</b>	152.8	150.5	146.0	141.9	137.5	133.4	129.3
<b>Steam out</b>	<b>°C</b>	464.0	457.3	444.7	432.5	420.3	408.7	397.1
<b>DeSH</b>	<b>°C</b>	-26.0	-32.7	-45.3	-57.5	-69.7	-81.3	-92.9
<b>dP tot</b>	<b>bar</b>	9.4	8.7	7.4	6.1	5.0	4.0	3.0

## Appendix IIa, Steam and electricity balances for case Base

### Mill steam usage and production

Project Name	BSMax_Base	<b>Pulp production</b>	
Client name	Study	Pine 1	600000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	6697	100	391	0.0	0.0	0.0	3.2
Cooking & O2	BDt/d	1668	1980	1927	0.0	0.0	8.8	5.7
Bleaching	BDt/d	1543	580	522	0.0	0.0	1.5	2.6
Drying	BDt/d	1543	2378	2140	0.0	0.0	0.0	16.8
Evaporation	t H2O/d	16218	460	4352	0.0	0.0	6.1	29.6
Recovery Boiler	tDS/d	3543	720	1488	5.9	0.0	3.0	1.7
Causticization	m3WL/d	6683	10	39	0.0	0.0	0.0	0.3
Lime reburning	t lime/d	548	120	38	0.0	0.0	0.0	0.3
Raw water	m3/d	46286	1	27	0.0	0.0	0.0	0.2
Effluent treatment	m3/d	39429	0	0	0.0	0.0	0.0	0.2
NCG	m3/h	32858	10	192				1.5
Auxiliary condenser	kg/s	1714	10	10				0.1
Auxiliary departments	BDt/d	1714	530	530	0.0	0.0	0.0	4.4
<b>Pulping steam usage total</b>				<b>11655</b>	<b>5.9</b>	<b>0.0</b>	<b>19.3</b>	<b>66.7</b>
ClO2+O2 Plant	tClO2/d	31	15500	280	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>11936</b>				<b>94.2</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	1.00	2.86	0
Recovery Boiler	kgDS/s	41.00	3.13	8.96	18513
Condensing tail	kg/s				-31.0
Spraying	kg/s				6.0
To feedwater preheatin	kg/s				-9.2
<b>Steam to consumers</b>					<b>94.2</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	6697	9000	10	12	3.27
Cooking	BDt/d	1668	1530	38	36	2.53
O2 delignification	BDt/d	1594	1530	28	27	1.82
Screening	BDt/d	1668	1530	33	32	2.20
Washing	BDt/d	1668	1530	21	20	1.40
Bleaching	BDt/d	1543	2070	67	78	5.04
Drying	BDt/d	1543	2070	150	176	11.29
Evaporation	t H2O/d	16218	23328	3.9	4.8	3.24
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	6683	8400	5.1	5.8	1.60
Lime reburning	t lime/d	548	575	44	45	1.03
Raw water	m3/d	46286	70000	0.3	0.4	0.73
Effluent treatment	m3/d	39429	60000	1	1.3	2.07
NCG	m3/h	32858	54760	0.5	1	0.91
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	1714	2000	36	39	2.79
<b>Pulp production total</b>					<b>786</b>	<b>50.52</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>838</b>	<b>53.87</b>

### Electricity production

Condensing turbine	kg/s	31	60	23500	22073	28.51
Back pressure turbine	kg/s	97	140	16000	16012	64.94
<b>Electricity production total</b>					<b>1454</b>	<b>93.45</b>

## Appendix IIb, Steam and electricity balances for cases LKgasif

### Mill steam usage and production

Project Name	BSMax_Gasif	<b>Pulp production</b>	
Client name	Study	Pine 1	600000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	6697	100	391	0.0	0.0	0.0	3.2
Cooking & O2	BDt/d	1668	1980	1927	0.0	0.0	8.8	5.7
Bleaching	BDt/d	1543	580	522	0.0	0.0	1.5	2.6
Drying	BDt/d	1543	2378	2140	0.0	0.0	0.0	16.8
Evaporation	t H2O/d	16218	460	4352	0.0	0.0	6.1	29.6
Recovery Boiler	tDS/d	3543	720	1488	5.9	0.0	3.0	1.7
Causticization	m3WL/d	6683	10	39	0.0	0.0	0.0	0.3
Lime reburning	t lime/d	548	120	38	0.0	0.0	0.0	0.3
Raw water	m3/d	46286	1	27	0.0	0.0	0.0	0.2
Effluent treatment	m3/d	39429	0	0	0.0	0.0	0.0	0.2
NCG	m3/h	32858	10	192				1.5
Auxiliary condenser	kg/s	1714	10	10				0.1
Auxiliary departments	BDt/d	1714	530	530	0.0	0.0	0.0	4.4
<b>Pulping steam usage total</b>				<b>11655</b>	<b>5.9</b>	<b>0.0</b>	<b>19.3</b>	<b>66.7</b>
ClO2+O2 Plant	tClO2/d	31	15500	280	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>11936</b>				<b>94.2</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	1.00	2.86	0
Recovery Boiler	kgDS/s	41.00	3.13	8.96	18513
Condensing tail	kg/s				-31.0
Spraying	kg/s				6.0
To feedwater preheatin	kg/s				-9.2
<b>Steam to consumers</b>					<b>94.2</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	6697	9000	10	12	3.27
Cooking	BDt/d	1668	1530	38	36	2.53
O2 delignification	BDt/d	1594	1530	28	27	1.82
Screening	BDt/d	1668	1530	33	32	2.20
Washing	BDt/d	1668	1530	21	20	1.40
Bleaching	BDt/d	1543	2070	67	78	5.04
Drying	BDt/d	1543	2070	150	176	11.29
Evaporation	t H2O/d	16218	23328	3.9	4.8	3.24
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	6683	8400	5.1	5.8	1.60
Lime reburning	t lime/d	548	575	44	45	1.03
Raw water	m3/d	46286	60000	0.3	0.3	0.66
Effluent treatment	m3/d	39429	50000	1	1.1	1.86
NCG	m3/h	32858	54760	0.5	1	0.91
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	1714	2000	36	39	2.79
<b>Pulp production total</b>					<b>782</b>	<b>50.25</b>
Gasifier	t/d	190	200	304	312	2.47
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>872</b>	<b>56.07</b>

### Electricity production

Condensing turbine	kg/s	31	60	24500	22073	28.51
Back pressure turbine	kg/s	97	140	16700	16012	64.94
<b>Electricity production total</b>					<b>1454</b>	<b>93.45</b>

## Appendix IIc, Steam and electricity balances for case LKlignin

### Mill steam usage and production

Project Name	BSMax_LKlignin	<b>Pulp production</b>	
Client name	Study	Pine 1	600000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	6697	100	391	0.0	0.0	0.0	3.2
Cooking & O2	BDt/d	1668	1980	1927	0.0	0.0	8.8	5.7
Bleaching	BDt/d	1543	580	522	0.0	0.0	1.5	2.6
Drying	BDt/d	1543	2378	2140	0.0	0.0	0.0	16.8
Evaporation	t H2O/d	17349	460	4655	0.0	0.0	6.5	31.7
Recovery Boiler	tDS/d	3401	720	1428	5.7	0.0	2.8	1.6
Causticization	m3WL/d	6683	10	39	0.0	0.0	0.0	0.3
Lime reburning	t lime/d	548	120	38	0.0	0.0	0.0	0.3
Raw water	m3/d	46286	1	27	0.0	0.0	0.0	0.2
Effluent treatment	m3/d	39429	0	0	0.0	0.0	0.0	0.2
NCG	m3/h	32858	10	192				1.5
Auxiliary condenser	kg/s	1714	10	10				0.1
Auxiliary departments	BDt/d	1714	530	530	0.0	0.0	0.0	4.4
<b>Pulping steam usage total</b>				<b>11899</b>	<b>5.7</b>	<b>0.0</b>	<b>19.6</b>	<b>68.6</b>
ClO2+O2 Plant	tClO2/d	31	15500	280	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>12179</b>				<b>96.2</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	1.00	2.86	0
Recovery Boiler	kgDS/s	39.36	2.95	8.44	16748
Condensing tail	kg/s				-17.4
Spraying	kg/s				6.0
To feedwater preheatin	kg/s				-8.5
<b>Steam to consumers</b>					<b>96.2</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	6697	9000	10	12	3.27
Cooking	BDt/d	1668	1530	38	36	2.53
O2 delignification	BDt/d	1594	1530	28	27	1.82
Screening	BDt/d	1668	1530	33	32	2.20
Washing	BDt/d	1668	1530	21	20	1.40
Bleaching	BDt/d	1543	2070	67	78	5.04
Drying	BDt/d	1543	2070	150	176	11.29
Evaporation	t H2O/d	17349	23328	3.9	4.6	3.33
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	6683	8400	5.1	5.8	1.60
Lime reburning	t lime/d	548	575	44	45	1.03
Raw water	m3/d	46286	60000	0.3	0.3	0.66
Effluent treatment	m3/d	39429	50000	1	1.1	1.86
NCG	m3/h	32858	54760	0.5	1	0.91
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	1714	2000	36	39	2.79
<b>Pulp production total</b>					<b>783</b>	<b>50.35</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>835</b>	<b>53.70</b>
<b>Electricity production</b>						
Condensing turbine	kg/s	17.4	60	24500	22073	16.00
Back pressure turbine	kg/s	99	140	16700	16012	65.86
<b>Electricity production total</b>					<b>1273</b>	<b>81.86</b>

## Appendix II.d, Steam and electricity balances for case LigMax600

### Mill steam usage and production

Project Name	BSMax_LigMax	<b>Pulp production</b>	
Client name	Study	Pine 1	600000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	6697	100	391	0.0	0.0	0.0	3.2
Cooking & O2	BDt/d	1668	1980	1927	0.0	0.0	8.8	5.7
Bleaching	BDt/d	1543	580	522	0.0	0.0	1.5	2.6
Drying	BDt/d	1543	2378	2140	0.0	0.0	0.0	17.6
Evaporation	t H2O/d	16999	460	4561	0.0	0.0	6.4	31.0
Recovery Boiler	tDS/d	3340	720	1403	5.6	0.0	2.8	1.6
Causticization	m3WL/d	6683	10	39	0.0	0.0	0.0	0.3
Lime reburning	t lime/d	548	120	38	0.0	0.0	0.0	0.3
Raw water	m3/d	46286	1	27	0.0	0.0	0.0	0.2
Effluent treatment	m3/d	39429	0	0	0.0	0.0	0.0	0.2
NCG	m3/h	32858	10	192				1.5
Auxiliary condenser	kg/s	1714	10	10				0.1
Auxiliary departments	BDt/d	1714	530	530	0.0	0.0	0.0	4.4
<b>Pulping steam usage total</b>				<b>11779</b>	<b>5.6</b>	<b>0.0</b>	<b>19.4</b>	<b>68.8</b>
ClO2+O2 Plant	tClO2/d	31	15500	280	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>12060</b>				<b>96.1</b>

### STEAM PRODUCTION

	Unit	Balance	Steam kg/s/unit	Heat production		Steam kg/s
				MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	1.00	2.86	0	0.0
Recovery Boiler	kgDS/s	38.65	2.88	8.23	16030	111.1
Condensing tail	kg/s					-13.6
Spraying	kg/s					6.0
To feedwater preheatin	kg/s					-7.4
<b>Steam to consumers</b>						<b>96.1</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	6697	9000	10	12	3.27
Cooking	BDt/d	1668	1530	38	36	2.53
O2 delignification	BDt/d	1594	1530	28	27	1.82
Screening	BDt/d	1668	1530	33	32	2.20
Washing	BDt/d	1668	1530	21	20	1.40
Bleaching	BDt/d	1543	2070	67	78	5.04
Drying	BDt/d	1543	2070	150	176	11.29
Evaporation	t H2O/d	16999	23328	3.9	4.7	3.30
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	6683	8400	5.1	5.8	1.60
Lime reburning	t lime/d	548	575	44	45	1.03
Raw water	m3/d	46286	60000	0.3	0.3	0.66
Effluent treatment	m3/d	39429	50000	1	1.1	1.86
NCG	m3/h	32858	54760	0.5	1	0.91
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	1714	2000	36	39	2.79
<b>Pulp production total</b>					<b>783</b>	<b>50.32</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>835</b>	<b>53.67</b>

### Electricity production

Condensing turbine	kg/s	13.6	60	24500	22073	12.51
Back pressure turbine	kg/s	98	140	16700	16012	65.07
<b>Electricity production total</b>					<b>1207</b>	<b>77.58</b>

## Appendix IIe, Steam and electricity balances for case LigMax704

### Mill steam usage and production

Project Name	BSMax_LigMax	<b>Pulp production</b>	
Client name	Study	Pine 1	704000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	7858	100	391	0.0	0.0	0.0	3.8
Cooking & O2	BDt/d	1957	1980	1927	0.0	0.0	10.3	6.7
Bleaching	BDt/d	1810	580	522	0.0	0.0	1.7	3.1
Drying	BDt/d	1810	2378	2140	0.0	0.0	0.0	19.7
Evaporation	t H2O/d	19903	460	4552	0.0	0.0	7.5	36.4
Recovery Boiler	tDS/d	3845	720	1376	6.4	0.0	3.2	1.8
Causticization	m3WL/d	7842	10	39	0.0	0.0	0.0	0.4
Lime reburning	t lime/d	643	120	38	0.0	0.0	0.0	0.4
Raw water	m3/d	54309	1	27	0.0	0.0	0.0	0.3
Effluent treatment	m3/d	46263	0	0	0.0	0.0	0.0	0.3
NCG	m3/h	38553	10	192				1.8
Auxiliary condenser	kg/s	2011	10	10				0.1
Auxiliary departments	BDt/d	2011	530	530	0.0	0.0	0.0	5.1
<b>Pulping steam usage total</b>				<b>11744</b>	<b>6.4</b>	<b>0.0</b>	<b>22.7</b>	<b>79.6</b>
ClO2+O2 Plant	tClO2/d	31	15500	239	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>11983</b>				<b>111.1</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	2.86	0	0.0
Recovery Boiler	kgDS/s	44.51	8.23	15731	128.0
Condensing tail	kg/s				-14.2
Spraying	kg/s				6.8
To feedwater preheatin	kg/s				-9.5
<b>Steam to consumers</b>					<b>111.1</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	7858	9000	10	11	3.51
Cooking	BDt/d	1957	2070	38	39	3.19
O2 delignification	BDt/d	1870	2070	28	29	2.30
Screening	BDt/d	1957	2070	33	34	2.77
Washing	BDt/d	1957	2070	21	22	1.76
Bleaching	BDt/d	1810	2070	67	72	5.42
Drying	BDt/d	1810	2070	150	161	12.13
Evaporation	t H2O/d	19903	23328	3.9	4.3	3.54
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	7842	8400	5.1	5.3	1.73
Lime reburning	t lime/d	643	660	44	45	1.19
Raw water	m3/d	54309	70000	0.3	0.3	0.78
Effluent treatment	m3/d	46263	60000	1	1.1	2.21
NCG	m3/h	38553	64252	0.5	1	1.07
Cooling towers	MW	242	350	108	132	1.33
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	2011	2500	36	40	3.38
<b>Pulp production total</b>					<b>743</b>	<b>56.07</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>788</b>	<b>59.42</b>

### Electricity production

Condensing turbine	kg/s	14.2	60	24500	23073	13.65
Back pressure turbine	kg/s	114	140	16700	16512	78.27
<b>Electricity production total</b>					<b>1219</b>	<b>91.92</b>



## Appendix IIf, Steam and electricity balances for case LigAux600

### Mill steam usage and production

Project Name	BSMax_LigAux	<b>Pulp production</b>	
Client name	Study	Pine 1	600000 ADt/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADt/a
		Eucalyptus 3	0 ADt/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	6697	100	391	0.0	0.0	0.0	3.2
Cooking & O2	BDt/d	1668	1980	1927	0.0	0.0	8.8	5.7
Bleaching	BDt/d	1543	580	522	0.0	0.0	1.5	2.6
Drying	BDt/d	1543	2378	2140	0.0	0.0	0.0	16.8
Evaporation	t H2O/d	15250	460	4092	0.0	0.0	5.7	27.9
Recovery Boiler	tDS/d	3035	720	1275	5.1	0.0	2.5	1.5
Causticization	m3WL/d	6683	10	39	0.0	0.0	0.0	0.3
Lime reburning	t lime/d	548	120	38	0.0	0.0	0.0	0.3
Raw water	m3/d	46286	1	27	0.0	0.0	0.0	0.2
Effluent treatment	m3/d	39429	0	0	0.0	0.0	0.0	0.2
NCG	m3/h	32858	10	192				1.5
Auxiliary condenser	kg/s	1714	10	10				4.4
Auxiliary departments	BDt/d	1714	530	530	0.0	0.0	0.0	0.1
<b>Pulping steam usage total</b>				<b>11182</b>	<b>5.1</b>	<b>0.0</b>	<b>18.5</b>	<b>64.6</b>
ClO2+O2 Plant	tClO2/d	31	15500	280	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>11463</b>				<b>90.5</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	2.86	0	0.0
Recovery Boiler	kgDS/s	35.13	2.95	14940	103.6
Condensing tail	kg/s				-11.3
Spraying	kg/s				6.0
To feedwater preheatin	kg/s				-7.8
<b>Steam to consumers</b>					<b>90.5</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	6697	9000	10	12	3.27
Cooking	BDt/d	1668	1530	38	36	2.53
O2 delignification	BDt/d	1594	1530	28	27	1.82
Screening	BDt/d	1668	1530	33	32	2.20
Washing	BDt/d	1668	1530	21	20	1.40
Bleaching	BDt/d	1543	1800	67	73	4.67
Drying	BDt/d	1543	2070	150	176	11.29
Evaporation	t H2O/d	15250	23328	3.9	5.0	3.16
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	6683	8400	5.1	5.8	1.60
Lime reburning	t lime/d	548	575	44	45	1.03
Raw water	m3/d	46286	70000	0.3	0.4	0.73
Effluent treatment	m3/d	39429	60000	1	1.3	2.07
NCG	m3/h	32858	54760	0.5	1	0.91
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	1714	2000	36	39	2.79
<b>Pulp production total</b>					<b>779</b>	<b>50.07</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>831</b>	<b>53.42</b>

### Electricity production

Condensing turbine	kg/s	11.3	60	24500	22073	10.39
Back pressure turbine	kg/s	92	140	16700	16012	61.56
<b>Electricity production total</b>					<b>1119</b>	<b>71.96</b>

## Appendix IIg, Steam and electricity balances for case LigAux930

### Mill steam usage and production

Project Name	BSMax_LigAux	<b>Pulp production</b>	
Client name	Study	Pine 1	930000 ADT/a
Location	Scandinavia BSK	Eucalyptus 2	0 ADT/a
		Eucalyptus 3	0 ADT/a

### STEAM CONSUMPTION

Department	Unit	Balance	Heat		Steam			
			MJ/unit	MJ/ADt	MP3 kg/s	MP2 kg/s	MP kg/s	LP kg/s
Woodhandling	m3n/sub	10380	100	391	0.0	0.0	0.0	5.0
Cooking & O2	BDt/d	2586	1980	1927	0.0	0.0	13.7	8.8
Bleaching	BDt/d	2391	580	522	0.0	0.0	2.3	4.0
Drying	BDt/d	2391	2378	2140	0.0	0.0	0.0	26.0
Evaporation	t H2O/d	23504	460	4069	0.0	0.0	8.8	42.9
Recovery Boiler	tDS/d	4472	720	1212	7.5	0.0	3.7	2.1
Causticization	m3WL/d	10359	10	39	0.0	0.0	0.0	0.5
Lime reburning	t lime/d	849	120	38	0.0	0.0	0.0	0.5
Raw water	m3/d	71743	1	27	0.0	0.0	0.0	0.3
Effluent treatment	m3/d	61115	0	0	0.0	0.0	0.0	0.3
NCG	m3/h	50929	10	192				2.3
Auxiliary condenser	kg/s	2657	10	10				0.1
Auxiliary departments	BDt/d	2657	530	530	0.0	0.0	0.0	6.8
<b>Pulping steam usage total</b>				<b>11096</b>	<b>7.5</b>	<b>0.0</b>	<b>28.5</b>	<b>99.8</b>
ClO2+O2 Plant	tClO2/d	31	15500	181	0.0	0.0	0.2	2.1
Power Boiler	kg/s	0	13000	0	0.0	0.0	0.0	0.0
					<b>0.0</b>	<b>0.0</b>	<b>0.2</b>	<b>2.1</b>
<b>Pulp mill steam usage total</b>				<b>11277</b>				<b>138.1</b>

### STEAM PRODUCTION

	Balance	Steam kg/s/unit	Heat production		Steam kg/s
			MJ/unit	MJ/Adt/d	
Power Boiler	kg/s	0.00	1.00	2.86	0
Recovery Boiler	kgDS/s	51.76	2.97	8.51	14329
Condensing tail	kg/s				-12.9
Spraying	kg/s				8.7
To feedwater preheatin	kg/s				-11.6
<b>Steam to consumers</b>					<b>138.1</b>

### Department usage

Department	unit	Balance units/d	Design units/d	Design		Electricity MW
				kWh/unit	Actual kWh/unit	
Woodhandling	m3n/sub	10380	11000	10	10	4.45
Cooking	BDt/d	2586	2700	38	39	4.18
O2 delignification	BDt/d	2471	2700	28	29	3.02
Screening	BDt/d	2586	2700	33	34	3.63
Washing	BDt/d	2586	2700	21	21	2.31
Bleaching	BDt/d	2391	2430	67	68	6.73
Drying	BDt/d	2391	2430	150	151	15.07
Evaporation	t H2O/d	23504	23328	3.9	3.9	3.83
Recovery Boiler	tDS/d	3967	4450	46	51	8.51
Causticization	m3WL/d	10359	11000	5.1	5.3	2.27
Lime reburning	t lime/d	849	900	44	45	1.60
Raw water	m3/d	71743	70000	0.3	0.3	0.89
Effluent treatment	m3/d	61115	60000	1	1.0	2.52
NCG	m3/h	50929	84878	0.5	1	1.41
Cooling towers	MW	24	350	108	842	0.84
Compressed air	m3/h	700	800	40	43	1.25
Miscellaneous	BDt/d	2657	3000	36	38	4.24
<b>Pulp production total</b>					<b>670</b>	<b>66.77</b>
ClO2 production	tClO2/d	31	40	120	137	0.18
Oxygen+peroxide	tO2/d	77	85	940	989	3.17
<b>Pulp mill total usage</b>					<b>704</b>	<b>70.12</b>
<b>Electricity production</b>						
Condensing turbine	kg/s	12.9	60	24500	22073	11.86
Back pressure turbine	kg/s	141	140	16700	16012	94.12
<b>Electricity production total</b>					<b>1064</b>	<b>105.98</b>



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