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Techno-economic analysis of energy efficiency measures in a pulp mill converted to an ethanol production plant

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KEYWORDS: Kraft pulp mill, Ethanol production, Pinch analysis, Thermal process integration, Steam surplus, Lignin separation, Power generation

SUMMARY: A conceptual ethanol production plant, based on conversion of a kraft pulp mill, has been studied. The process uses softwood as raw material, alkaline pre-treatment combined with delignification, and biochemical conversion of sugars to ethanol (i.e. hydrolysis and fermentation). The plant has been analysed by pinch methods in order to find steam-saving possibilities. It is shown in the study that a large amount of steam surplus can be found if energy efficiency measures are implemented. In order to study the possible effect on the profitability of the plant when introducing steam-saving measures, the process has been analysed from a techno-economic point of view. It is shown that implementing energy efficiency measures could have a substantial effect on profitability if the by-product (in this case lignin biofuel or power) is high-valued. It is also shown that lignin as by-product might be more profitable than power, mainly because the demand for CO₂ in lignin extraction might be supplied by CO₂ produced in fermentation of sugars to ethanol. If investments are made to convert a pulp mill to ethanol production, energy efficiency measures should be included in the discussion since they might play an important role in minimising ethanol production cost.

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The transport sector is responsible for a large part of global greenhouse gas emissions today. In contrast to most other sectors, GHG emissions from transport (mainly CO₂) increased substantially in the European Union from 1990 to 2001 (Fuglestvedt et al. 2007). In order to come to terms with emissions in a society where demand for transportation will continue to grow rapidly, it is essential to utilise sustainable and efficient sources and technologies. Near-zero-emission technologies, such as electric and hybrid electric engines, hydrogen fuel cells, biodiesel and ethanol, need to take over the role of petroleum-based fuels in the future transport system, in order to reach sustainability.

In the long term, electric engines and hydrogen fuel cells have the potential of being much more efficient than combustion engines. Liquid biofuels, for example biodiesel (RME, DME) and ethanol, have the short-term advantage of being more easily implemented into the existing infrastructure. Both biodiesel and ethanol can be mixed with today's fuels, thus facilitating a rapid reduction of greenhouse gas emissions from the transport sector.

Ethanol is the primary biofuel used in the transport sector today, and is expected to remain dominant in the near future (IEA 2006). It can be produced in many different ways, of which not all are easily shown to be environmentally beneficial in comparison with fossil fuels. The major part of all ethanol biofuel sold today is produced by way of hydrolysis/fermentation of sugar or starch. Feedstocks used are for example wheat, corn, and sugar cane.

In order to improve environmental performance of ethanol production, while at the same time vastly expanding the global production, lignocellulosic raw material should be used. Lignocellulosic biomass accounts for about 50% of all global biomass (Claassen et al. 1999). In lignocellulosic ethanol production the main by-product is lignin, which due to its relatively high energy content is interesting to use in production of electricity, heat, or other biofuels.

The most important parameters for economics of lignocellulosic ethanol production are feedstock costs, investment costs, and reaching high ethanol yield and concentration in the fermentation step of the process. Since ethanol production from lignocellulosic feedstock, even with optimal conversion performance, has relatively low theoretical yield (for softwood around 410-455 litres ethanol/metric ton dry raw material), it is also of high importance to make use of by-products. This implies that the energy efficiency of the plant should be as high as possible, in order to keep a high efficiency in the production process (Galbe et al. 2002).

There are many similarities between lignocellulosic ethanol plants and pulp mills, since both use wood-based feedstock and include several similar unit operations. Since increased competition in pulp production can be expected in the future, many pulp mills could have decreasing profit margins. Within the forest industry, research on how to invest in new markets for non-profitable pulp mills, in order to save them from closing down, has mainly been focusing on using compounds in the black liquor to produce new high-value products. New alternatives such as black liquor gasification followed by biofuel production, as well as biofuel production from hemi-cellulose and lignin extraction for pellets production, have been introduced and studied in recent years (Van Heiningen 2006, Frederick et al. 2007).

Research by STFI-Packforsk in Sweden has suggested a new possibility in order to save unprofitable pulp mills, namely conversion into biorefineries with ethanol as main product line, and lignin/electricity as by-products. This ethanol process would use conventional hydrolysis/fermentation, but alkaline pre-treatment and

chemical recovery according to pulp mill process design. Possible positive aspects of this, compared to other ethanol processes, could be lower investment cost, sulphur-free process, simplified distillation, and less inhibition in hydrolysis/ fermentation (von Schenck et al. 2007).

The main part of the study presented here is a techno-economic analysis of the ethanol process suggested by von Schenck et al. (2007), evaluating the potential for improving profitability when introducing energy efficiency measures. The results are presented in relation to a base case, which is the process without implemented energy efficiency measures. This is assumed to be the best way of displaying the consequences of steam-saving measures in the process, due to the difficulties inherent in generating reliable economic data in the not yet commercialised lignocellulosic ethanol process.

Process description

FRAM type mill

The pulp mill is based on a model mill, designed within the Swedish national research programme “Future Resource-Adapted Mill” (Delin et al. 2004). The simulation model has been developed together with industry in order to produce a model of a typical Scandinavian kraft pulp mill. The model is assumed to be in steady state, as it is mainly designed for use in modification studies.

The capacity of the mill is 1000 ADt/day of bleached market pulp, with a pulp yield of about 50%.

Some other key aspects of the type mill are:

- There is no condensing turbine.
- The backpressure turbine is too small to accommodate the HP steam produced in the recovery boiler, and part of the steam is let down through expansion valves.
- The digester uses a conventional 2-flash system, and chip bin pre-steaming is not used.
- The evaporation plant has a steam economy corresponding to 5.5 effects, and reaches a DS of 72%.
- The foul condensate stripper is not thermally integrated.
- The lime kiln is fired with fuel oil.

Further information can be found in Delin et al. (2004).

The ethanol plant

The ethanol production plant presented and used here has its basis in the FRAM model mill, where changes have been made to the design of the fibre line, in order to produce ethanol instead of bleached market pulp. It is assumed that the process can be made sulphur-free, and that the chemical recovery cycle of the converted pulp mill will be usable without any major modifications, except if introducing a lignin separation unit. Presented in Fig 1 is a block diagram of the conceptual process.

The capacity of the ethanol production process is approximately 1800 tonnes dry wood/day, compared to

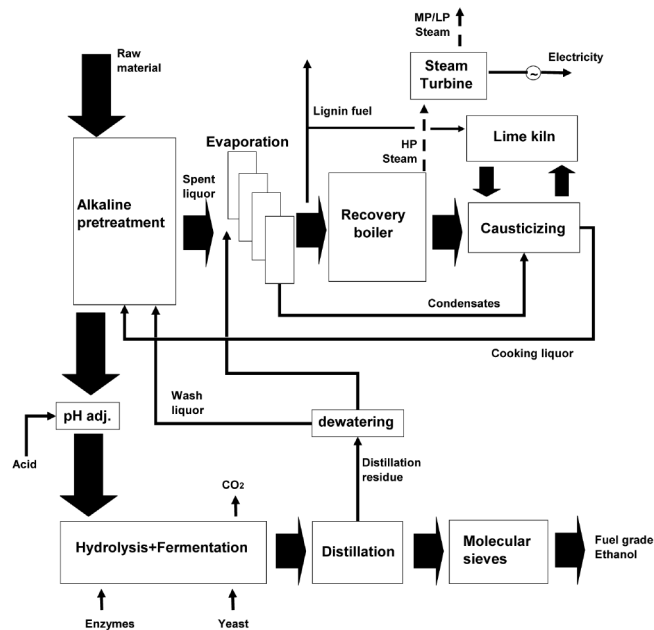


Fig 1. Block diagram of the conceptual ethanol production process.

the pulp mill which has a capacity of 2065 tonnes dry wood/day. This is because the sodium hydroxide produced from sulphide hydrolysis in the kraft mill process will not be available in the ethanol process. This is assumed to decrease the production rate of the digester at a given alkali charge (von Schenck et al. 2007). The ethanol plant produces about 340 m³ of ethanol/day, which corresponds to approximately 55% of theoretical ethanol yield, if only hexoses are fermented. The yield of ethanol compared to raw material needed is approximately 15%, and thus 1 tonne of dry wood introduced to the plant will generate 150 kg of ethanol. The composition of the raw material used in the analysis is presented in Table 1.

The main units in the ethanol production process are described below, as are the data used for the analysis of the conceptual ethanol production process.

a) Pre-treatment unit

The pre-treatment of incoming raw material is assumed to be taking place in the same units as in a pulp mill. Sodium hydroxide is used to delignify the wood chips entering the digester. Data used in the model for pre-treatment are shown in Table 2, and are assumed to be similar to the model mill pre-treatment data.

Table 1. Composition of softwood entering the ethanol plant (von Schenck et al. 2007).

Cellulose, C6 hexoses	%	40,0
Hemicellulose		
- C6 hexoses	%	14,0
- C5 pentoses	%	8,0
- Other	%	5,0
Lignin	%	26,0
Extractives	%	3,0
Other	%	4,0
Total	%	100,0

Table 2. Data for pre-treatment (von Schenck et al. 2007).

Pre-treatment data		
NaOH charge	% on wood	19
Kappa number		30
Cellulose yield	% on cellulose	80
Hemicellulose yield	% on hemi	40
Total carbohydrate yield	% on wood	42,8
Dry substance in wood to PT		50
Dry substance after PT and wash	%	10

b) Hydrolysis/fermentation

The hydrolysis is assumed to be enzymatic, and separated from fermentation in order to utilise a higher temperature in hydrolysis (65°C), as compared to fermentation (40°C). It is assumed that only hexoses are fermented in the process. Estimations of yields in hydrolysis and fermentation are taken from Aden et al. (2002), and experimental studies by STFI Packforsk (Olm et al. 2007). Table 3 below present data for hydrolysis and fermentation used in the calculations.

Table 3. Data for hydrolysis and fermentation (von Schenck et al. 2007).

Hydrolysis / Fermentation data		
Enzyme dosage	kg/t (C5+C6)	20
Hydrolysis yield	% on carbohydrates	85
Fermentation yield C6	% on C6	48
Fermentation yield C5	% on C5	0

c) Distillation

The distillation units are assumed to be similar to the design presented in Aden et al. (2002), using two columns in series. The first is a beer column, where most of the CO₂, water, and other residues are removed, and the second is a rectifying column where ethanol is separated from water. The columns are initially assumed to be driven by LP and MP steam. Steam demand has been estimated by comparing with ethanol processes in the literature. In the conceptual process feed ethanol concentration is 5.4%, and after distillation it is assumed to have reached near azeotropic concentration. The ethanol is then dried in a molecular sieve up to over 99% purity before leaving the process.

d) Chemical recovery cycle

Evaporation, recovery boiler, causticizing and lime kiln are assumed to be similar to the units in the pulp mill. It is estimated that the causticizing plant and lime kiln will be the limiting units in the ethanol plant. This is because of the assumption that the process is sulphur-free. Thus, as mentioned before, there will be no production of sodium hydroxide from sulphide hydrolysis, which will lead to the consequence that the production of sodium hydroxide will be the limiting step in the process.

e) Steam/electricity production

Compared to the pulp mill, the biorefinery is estimated to have considerably lower steam consumption. This indicates that excess steam will be produced. If no investments are made in the ethanol process, excess HP steam will be expanded to LP steam and vented to atmosphere.

Heat integration

The possibilities for heat integration in the ethanol process have been discussed briefly by von Schenck et al. (2007). The heat integration part of this study is done with reference to the discussions made there, as well as discussions made in previous studies produced at the department, concerning heat integration in pulp mills (Axelsson et al. 2006). The stream data used for evaluating possibilities are included in Appendix 1. Included in the stream data are the hot utility demands of LP, MP, and HP steam, respectively. The steam demand of the process, as given in the appendix, is approximately 149 MW.

In Fig 2 the hot utility targets for varying global ΔT_{min} are plotted. Studying this figure it can be seen that the steam demand corresponds to a global ΔT_{min} of approximately 28 K, and that there exist large possibilities of decreasing this demand if the global minimum temperature difference for heat exchange in the process is reduced.

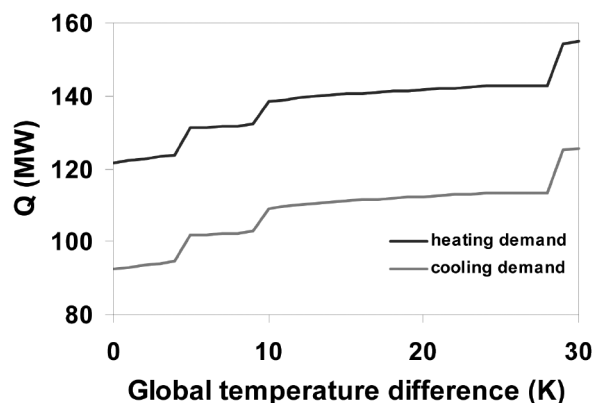


Fig 2. Demand curves for the ethanol process.

In reality, reducing global ΔT_{min} in the process implies solving pinch violations. If reducing the global ΔT_{min} to 4 K, the violations according to Table 4 can be found in the process as suggested by von Schenck et al. (2007).

All but one of the violations found in the process are related to utilisation of LP and MP steam below the pinch temperature, thus increasing the amount of steam needed above the pinch, in the heat deficit region. But the last violation, utilising flash steam from the spent liquor across the pinch, is the most important reason for having a high utility demand.

As can be seen in Fig 2, there are three steps in the curve where a small decrease in global ΔT_{min} will give a larger decrease in hot utility demand. The step at 28 K is

Table 4. Pinch violations in the studied process.

Pinch violation	Stream	Theoretical saving [MW]
Add Q below pinch	Building heating (LP)	1,1
Add Q below pinch	Wood Yard (LP)	1,4
Add Q below pinch	Wood chips (LP)	2,5
Add Q below pinch	W. liquor – imp. (MP)	1,0
Add Q below pinch	W. liquor – digest (MP)	4,1
Q across pinch	Spent liquor Flash steam 1 *	14,8
Total pinch violations		25

* Spent liquor flash steam should be used for Wood chips and White liquor above pinch, and in dist reboiler. Otherwise it will be used across the pinch

related to the possibility of using flash steam at 128°C when pre-steaming wood chips at 100°C. This is assumed to be done in the ethanol process, thus implying a global ΔT_{\min} of 28 K. The step at 9 K is related to using flash steam at 109°C for pre-steaming wood chips at 100°C. This would in turn generate the possibility of using more flash steam at 128°C above the pinch temperature, thus reducing the hot utility demand. Finally, at 4 K global ΔT_{\min} , the flash steam at 128°C could be used in the distillation reboiler at 124°C, which would give further decrease in hot utility demand, due to possibilities of using almost all flash steam in the process at above pinch temperature.

Resulting from the discussion above, the individual ΔT_{\min} -approach for each stream has been estimated. The values of these are included in **Appendix 1**. Values for streams also existing in the model pulp mill have been set in previous studies by Axelsson et al. (2006). The spent liquor flash steam does have a minimum approach temperature of 2 K in these studies, which is in accordance with the discussion above. The added streams due to conversion to ethanol production have not been studied previously, and thus ΔT_{\min} -values need to be estimated here. For simplicity, all ethanol streams are set to have ΔT_{\min} -values of 5 K in this study, except for the reboiler of the low temperature distillation column. ΔT_{\min} for this stream is set to 2 K, in order to be able to integrate according to the previous discussion. This set value might be too low in practice, but since the distillation in this process is only conceptual, and since in reality it should be possible to operate the column at a lower pressure in order to utilise the flash steam in the reboiler, it should be acceptable.

Looking at the Grand Composite Curve of the process at the minimum temperature approaches assumed in this study (Fig 3), the existence of excess heat available at a high temperature (i.e. above 80°C) is seen. The two main sources for this are the distillation column condenser and the foul condensate stripper condenser.

As seen in Fig 3, the amount of excess heat at a high temperature is approximately 23 MW. There exist several alternative ways of making use of this excess heat. For example, internal heat pumping in the distillation units could be implemented (between the condenser in the high temperature column and the reboiler in the low temperature column), or the heat could be used in a district heating network if one is available near the site. In this study, however, it is assumed that the excess heat is used in order to decrease the steam demand of the evaporation train.

Two different ways of investing in energy efficiency measures have been analysed in this study, namely:

- *Conventional measures*: investing in solving pinch violations, as well as upgrading evaporation from 5.5 to 7 effects steam economy.
- *Process integration*: investing in the conventional measures, as well as process integration of the evaporation plant.

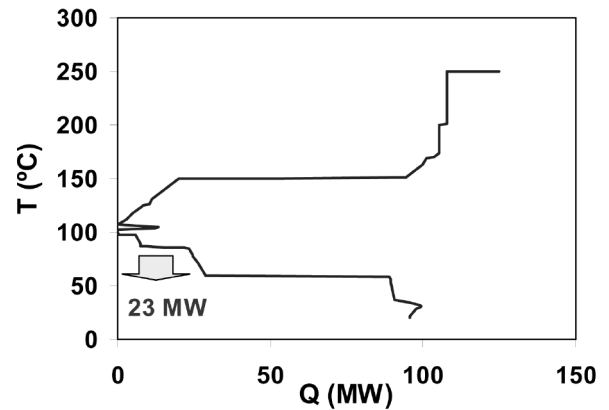


Fig 3. GCC for ethanol process at low global ΔT_{\min} .

The theoretically possible steam savings that could be made while implementing the two alternative measures above are shown in Table 5. It is assumed in this study that when upgrading the evaporation plant, the condensate stripper will be integrated with the evaporation train, thus decreasing the amount of excess heat by approximately 5.7 MW. It is also assumed that a further 2 MW of steam can be saved when implementing conventional measures. This is because some of the excess heat available could be used for preheating make-up boiler water from 75 to 95°C, which is as close to the pinch temperature of 101°C as is assumed possible. Making further use of excess heat when applying conventional energy efficiency measures is not included in this study.

Process-integrated evaporation

Using excess heat for integration of the evaporation plant implies that several changes in the process are needed. Since the excess heat should be used in as many evaporator effects as possible, it is assumed that the surface condenser temperature is decreased and that new effects are added at the low temperature side of the evaporation train. It is also assumed that the total evaporation capacity is constant in the study. This means that increasing the steam economy from 5.5 to 7 effects, and integrating the plant, will decrease the live steam demand accordingly. In this study it is assumed that 4 of the 7 effects can be used as sinks for excess heat. Thus, process integration will reduce steam demand by about 10 MW. Fig 4 visualises the concept of integrating evaporation used in this study.

Table 5. Potential steam savings in the ethanol process.

	Savings process integrated evaporation [MW]	Savings conventional evaporation [MW]
7 effects	12.9	12.9
integrated stripper	7.3	7.3
increased dry solids	6.9	6.9
Excess heat ^a	10.0	2.0
Sum	37.1	29.1

^a It is assumed that the redesigned evaporation will have integrated stripper, thus excess heat will decrease. Using the remaining excess heat for integrated evaporation will theoretically lead to 10 MW of steam savings. If conventional methods are used, approximately 2 MW of excess heat can be used to preheat boiler water

Techno-economic analysis

It has been shown that process integration and other energy efficiency measures can reduce steam demand in the conceptual ethanol production plant. The feasibility of implementing these measures is, however, not discussed. The main part of this study has been the production of a techno-economic analysis of the ethanol process, in order to study the consequences of investing in lower energy demand.

Due to the fact that the lignocellulosic ethanol production process is still under development, it is difficult to obtain reliable economic data for the process. Therefore this analysis will focus on how the yearly earnings of the process vary relative to a base case. The base case will be the conceptual ethanol plant without any investments in energy efficiency measures or by-product export.

Two alternative ways of turning steam savings into increased earnings have been studied, namely extraction of lignin for sale as biofuel, or increasing power production by investing in new turbine capacity. Thus six different alternatives have been evaluated, as shown in Table 6.

Investment costs for the new 7-effect evaporation plant with integrated stripper are estimated from Axelsson et al. (2006). The specific investment cost for solving pinch violations is set to 0.3 €/W in this study. Economic data used for the analysis are shown in Table 7.

Lignin separation

The lignin separation is assumed to be done using the LignoBoost concept, as described in Olsson et al. (2006). The unit is connected to the evaporator plant, where part or all of the spent liquor flow is diverted into the lignin separation unit, and CO₂ is added in order to lower the pH. A pH of about 9 will, according to studies on lignin extraction of kraft black liquor (Olsson et al. 2006), precipitate up to 70% of the lignin in the diverted stream. Since wash water, which is recycled to the evaporator, is used in the lignin separation unit, the steam demand and heat exchange area of the evaporator will increase. Data used in the techno-economic analysis for calculations of lignin extraction are presented in Table 8.

Implementing energy efficiency measures will decrease steam demand in the ethanol production process. This in turn leads to increased possibilities of extracting lignin from the spent liquor. There is, however, an upper limit on the amount of lignin which can be extracted. Since lignin has a relatively high heating value in the spent liquor, lowering the lignin content of the liquor flowing into the recovery boiler will also lower the heating value of the total liquor stream. This will affect recovery boiler operation, since the adiabatic flame temperature might decrease. In kraft pulp mill operation it has been estimated that approximately 0.42 tonnes lignin/ADt pulp can be extracted without compromising recovery boiler operation, when reaching a black liquor DS of 80% out from evaporation (Olsson et al. 2006). For the FRAM type mill, this can be recalculated as approximately 0.21 tonnes lignin/tonne dry wood introduced to the plant. The maximum amount of lignin extracted at 80% DS out from eva-

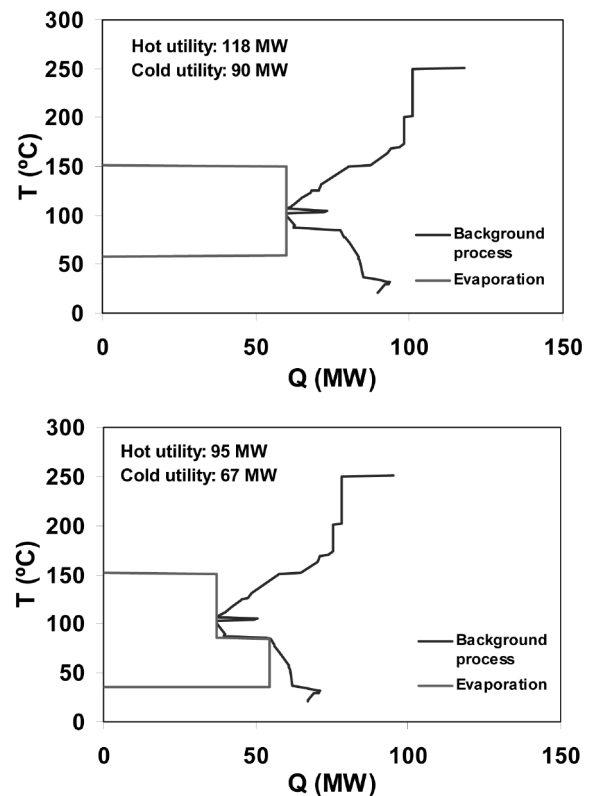


Fig 4. BG/FG-curves of evaporation and background process (assuming integrated four condensate stripper):

- Conventional evaporation with 5.5-effect steam economy (above).
- Process-integrated evaporation with 7-effect steam economy (below).

Table 6. Cases studied in the techno-economic analysis.

Alternative cases studied in analysis	Power	Lignin
No efficiency measures	Case 1A	Case 1B
HX improved, new conventional evaporation	Case 2A	Case 2B
HX improved, new integrated evaporation	Case 3A	Case 3B

Table 7. Economic variables used in the analysis.

Annuity factor	0.1	Strategic investment
Lignin selling price	5 - 30	[€/MWh]
Electricity price	30 - 80	[€/MWh]
New evaporation plant (7 effects, integrated stripper)		
- Case 2 (Conventional)	9.1	[M€] (Olsson et al, 2006)
- Case 3 (Process-integrated)	9.5	[M€] (Olsson et al, 2006)
Pinch violations*	7.5 (8.1*)	[M€]

* If excess heat is used for heating make-up boiler water (strategic investment)

Table 8. Parameters/variables used in economic analysis of lignin separation unit.

wash water in lignin separation	2 [t/t lignin]	
Increased steam demand of evap	8 - 16 [% of evap]	
Power demand of lignin extraction	82 [kWh/t lignin]	(Von Schenck et al, 2007)
Chemicals		
- Acid	80 [kg/t lignin]	(Von Schenck et al, 2007)
- CO ₂	-	On-site production assumed (fermentation)
Acid cost	38 [€/ton]	(Von Schenck et al, 2007)
Investments needed in evap (to increase HX-area)	6,7 - 13,1 [M€]	Estimated using simulation tool OptiVap (Olsson et al, 2006)
Cost for lignin separation plant	5.95*LR ^{0.6} [M€]	LR = lignin extraction rate [kg/s] (Olsson et al, 2006)

poration can thus be estimated as approximately 75% of total lignin. Lower amounts of dry substance in the liquor from evaporation will decrease the heating value, and thus also decrease possibilities to extract lignin.

In this study the evaporation plant initially has a 5.5-effect steam economy, with a 72% DS of liquor out from evaporation. Upgrading evaporation to 7-effect steam economy will result in 80% DS in the strong liquor.

In the process without steam-saving measures (*Case 1 in Table 6*), approximately 38% of total lignin can be extracted. Since approximately 75% lignin could be extracted at 80% DS without compromising boiler operation, it is assumed in this study that 38% at 72% DS will be possible. The assumption made is that having twice the amount of lignin in the strong liquor will compensate for a decrease in heating value caused by an 8% decrease in DS.

For the upgraded evaporation (*Cases 2 and 3 in Table 6*) the maximum amount of lignin can be extracted. This value is set to 70% of total lignin in the process. This is within the limits stated above regarding recovery boiler operation.

Since the ethanol production process is assumed to be sulphur-free, the comparison with kraft pulp mill recovery boiler operation will be conservative, as no endothermic sulphur-reduction reactions are assumed to take place.

Increased power production

The increased amount of unutilised LP steam in the process, after energy efficiency measures are introduced, could also be used for power production in a turbine. Hence the calculations on increasing power production include investment in a condensing turbine, as well as investing in a new back-pressure turbine, sized according to amount of available HP and MP steam in the process. In *Table 9* the data used for calculations in the techno-economic analysis of increased power production are presented.

Results

The results of the calculations done in the techno-economic analysis are shown in *Table 10*. Steam savings are presented as relative values compared to a base case, which is the ethanol production plant prior to making any investments. Lignin extraction is presented in absolute values, while power production is presented both as an absolute value and relative to the base case (Δ power). Investment costs are calculated as described in the preceding text. The amount of cooling needed is calculated from the steam data and steam savings in each case. The cooling cost decreases by up to 90 k€/year when implementing efficiency measures in the process. The main reasons for this are solving pinch violations and improving the steam economy of evaporation. In *Case 1A* the cooling is increased somewhat due to the investment in a condensing turbine. The relative increase in operating cost in the table below is due to the need for acid in the lignin separation unit.

As can be seen in the above table, maximum capacity of lignin extraction is reached when applying energy efficiency measures. In fact, there is still a steam surplus remaining after extraction of the lignin. Thus it is assumed in this study that further investments are made in order to make use of this LP steam in a condensing turbine. Another alternative could be to only decrease steam demand enough for the LP steam to sufficiently cover the demand of the process. The PBP of investments would decrease somewhat if this alternative was chosen, but is still low (below 4 years) for both alternatives within the limits of this study. Since the process would have decreased power production when not investing in a condensing turbine, and since the annual earnings would be lower, the alternative with maximum possible steam savings, and investment in condensing turbine, has been chosen in this study.

A few interesting results can be seen in *Table 10*. For example, it is shown that the 8 MW difference in steam savings between conventional and process-integration measures will give a small increase in power production. The increase is 12 GWh/a, or approximately 1.5 MW. This is because the saved steam is of low pressure, and thus is only used in the condensing turbine. It can also be seen that the difference in investment cost between process-integration and conventional measures is small. This is because the cost for process-integrating evaporation is assumed to be lower per MW saved than if conventional measures are applied (Axelsson et al. 2007).

Lignin separation

As shown in *Table 8*, investing in lignin separation will have several economic consequences for the ethanol

Table 9. Data used for turbines in the analysis.

Back-pressure turbine		
- isentropic efficiency	0.85	(Olsson et al, 2006)
- investment cost	$1.09 * P^{0.6}$	P = generated power [MW] (Olsson et al 2006)
Condensing turbine		
- isentropic efficiency	0.81	(Olsson et al, 2006)
- investment cost	$1.96 * P^{0.6}$	P = generated power [MW] (Olsson et al 2006)

Table 10. Lignin extraction and power production for base case ethanol plant (no investments made) and the 6 different cases studied.

		Initial EtOH-plant	Turbines			Lignin Extraction		
			1A	2A	3A	1B	2B	3B
Steam saving	GJ/tdw in	-	-	3	3	-	3	3
	MW	-	-	54	62	-	54	62
Power production	GWh/a	190	312	404	415	187	196	207
Sold power		92	214	305	317	84	89	100
Δ power	GWh/a	-	122	213	225	-3	5	17
Lignin	GWh/a	-	-	-	-	364	652	652
Cooling	MW	118	134	108	114	118	80	86
Investments	M€	-	15	37	38	16	48	49
Pinch violations		-	-	8	7	-	8	7
Evaporation plant		-	-	9	10	7	22	23
Condensing turbine		-	-	6	11	11	-	5
Back-pressure turbine		-	-	9	9	9	-	-
Lignin Extraction Plant		-	-	-	-	9	13	13
Δ operating costs	k€/a	-	-	-	-	182	327	327

process. The steam demand will increase by 8-16%, since the wash water from the lignin extraction plant is recycled to the evaporator train. This also implies that the flowrate into the evaporators will increase, thus creating a need for larger heat-exchange area. Finally, acid is used to prevent dissolution of, and sodium binding in, the lignin product, thereby increasing the operating costs for the process. The resulting effect on economics for the three different cases with lignin separation can be seen in *Table 10*, and the variation with regard to lignin price can be seen in *Figs 5* and *6*.

As is expected from the discussion in the preceding section, lignin extraction is not sensitive to electricity price. This can be concluded from the values in *Table 10*, since the amount of power sold does not change very much compared to the base case. As can be seen by comparing *Figs 5* and *6*, *Case 3* is somewhat better if the electricity price is high, whilst the other two cases are somewhat less profitable. The difference is small, however.

An interesting observation in the figures is that the slopes of the lines for *Case 2* and *3* are steeper than for *Case 1*. This is because in these cases the maximum amount of lignin can be extracted. Thus the annual earnings will increase more rapidly with increasing lignin selling price. The investment needed in these two cases is high, as can be seen in *Table 10*, which is indicated by the higher PBP for *Cases 2* and *3*, as compared to *Case 1*. Since the annual earnings are higher for the cases where energy efficiency measures have been implemented, and since the PBP is below 4 years at a lignin selling price of around 17 €/MWh, it could be argued that efficiency measures are of interest if the price of lignin is high.

Increased power production

As can be seen in *Table 10*, investing in energy efficiency measures, as well as new turbine capacity, can lead to substantial increases in power production. Between 122 and 225 GWh/a of electricity can be sold to the grid, in the three cases studied. The investment costs are related to purchase of new turbines, and are lower than investments needed for lignin separation. As can be seen in *Fig 6*, *Cases 2* and *3* give higher annual earnings than *Case 1* continuously within the limits of this study. The PBP is lower for the case without energy efficiency measures, however. At a power price of about 50 €/MWh, the PBP is below 4 years for all cases. If the assumption is made that the price of power in the future will be higher than 50 €/MWh, investing in energy efficiency measures and increased power production should be advantageous.

Lignin extraction versus increased power production

Fig 8 shows which lignin selling price is needed at a specific electricity price, in order for increased power production and lignin extraction to have equal annual earnings. The three cases studied in this paper are fairly similar, although the slope of *Case 1* differs from those of *Cases 2* and *3*.

In previous studies (Ådahl 2004) the future price ratio between electricity and biofuel has been estimated, and assumed to be fairly constant, at a value of 2.9. In Olsson

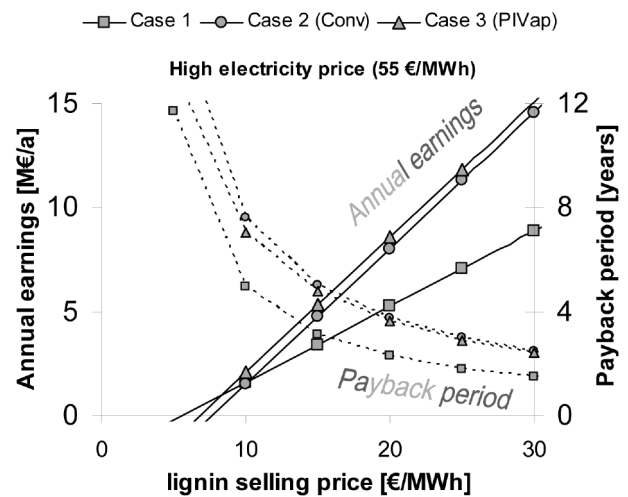


Fig 5. Annual earnings and payback period for investment versus lignin selling price (high electricity price (55 €/MWh)).

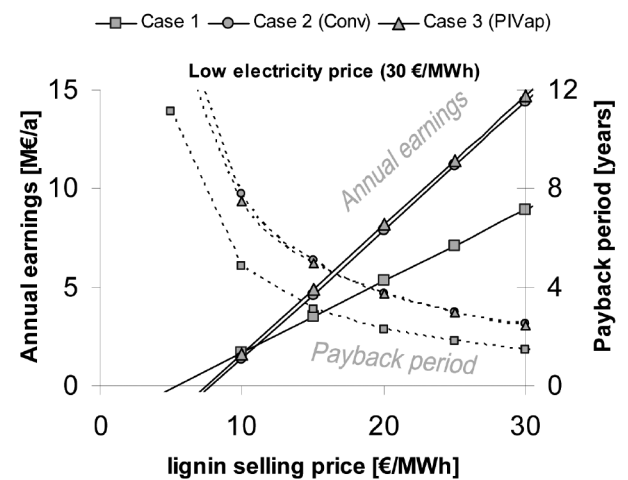


Fig 6. Annual earnings and payback period for investment versus lignin selling price (low electricity price (30 €/MWh)).

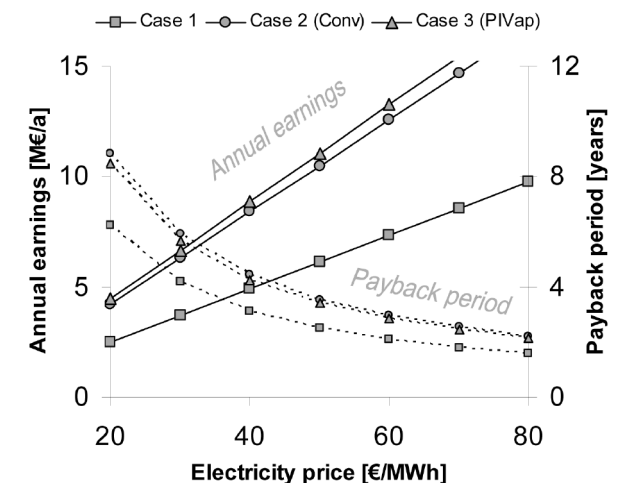


Fig 7. Increased power production.

et al. (2007) this value has been recalculated to be used for comparing electricity and lignin in a pulp mill, and here it is estimated that a value of 2.3-2.5 could be expected in the future.

The values for the electricity/lignin price ratio calculated in this study, and depicted as the slope of the lines in

Fig 8, are between 2.9 and 3.1. Thus if the real ratio is lower than 2.9, lignin would give higher annual earnings than electricity. This indicates that if a ratio of 2.3-2.5 was realised, lignin separation would be the more interesting choice.

Comparison with pulp mill studies

Previous studies of typical kraft pulp mills have shown large possibilities of steam savings in the process. Possibilities of using this steam surplus for increased power or lignin production have also been investigated in a techno-economic analysis.

When converting a pulp mill into ethanol production, important process parameters are changed in several ways. Two energy integration implications of conversion are that the steam demand of the process is decreased, and that the heat balance of the process is changed. This will in turn affect the possibilities for producing by-products, and for investing in energy efficiency measures.

Comparing ethanol production with kraft pulp mill studies presented by Olsson et al. (2006), it can be seen in Fig 9 that investing in lignin extraction shows more promise in the biorefinery. The investment costs are higher for the biorefinery case, since the cost for evaporator area is set at a higher value today than in the study by Olsson et al. (2006). If the investment costs were set at the same value for the pulp mill and biorefinery, the difference would be larger.

CO₂ is a large contributor to the operating cost of running a lignin separation plant. In an ethanol production plant, a lot of CO₂ will be produced in the fermentation reaction. Thus it should be possible to integrate these two units in order to decrease the lignin separation unit operating costs. In Fig 10 the effect of introducing a CO₂ price in an economic analysis of lignin separation in the biorefinery is shown. The low and high prices for CO₂ (9 and 24 €/tonne lignin extracted) are taken from Olsson et al. (2006). It can be seen that the possibility of internal production of CO₂ has a large effect on the profitability of lignin separation.

Effect on ethanol production cost

When discussing ethanol production it is interesting to study how investments affect the production cost of ethanol. Thus, instead of showing the results of the study in annual earnings for each case, the results could be expressed as the potential decrease in ethanol production cost. As mentioned previously in the article the yield of ethanol in the process is approximately 340 m³ of ethanol per day, which thus leads to a yearly production of approximately 112,000 m³ ethanol. Dividing the annual earnings by the production of ethanol will give an estimate of how much the ethanol cost will decrease in € per cubic meter ethanol. As can be seen in Fig 10, the change in ethanol cost varies depending on what measures are taken, but investing in energy measures could have a large effect on the production cost. The lignin selling price and electricity price on the axes in the figure are set so that an electricity price on the second y-axis is 2.5 times the corresponding lignin selling price on the prima-

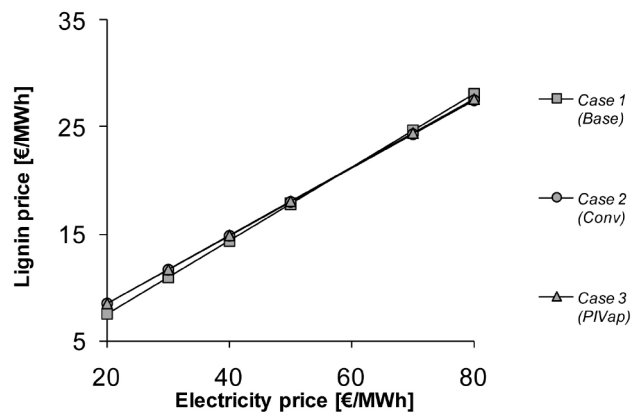


Fig 8. Lignin price where annual earnings for lignin extraction and increased power production are equal, at a given electricity price.

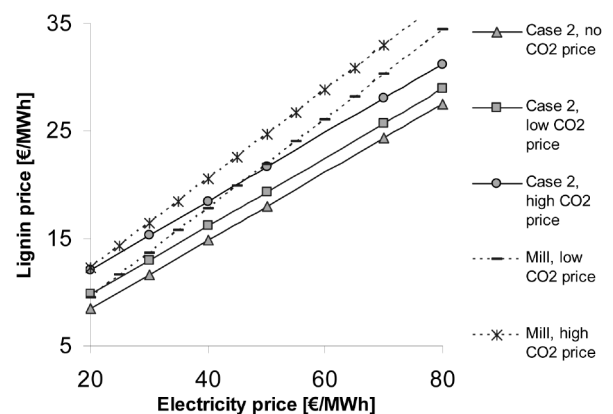


Fig 9. Effects of CO₂ cost on lignin price where annual earnings are equal for increased power production and lignin extraction, at a given electricity price.

ry y-axis, as depicted by the gridlines in the figure. For example an electricity price of 40 €/MWh corresponds to a lignin price of 16 €/MWh. Thus the different lines for lignin extraction and increased power production can be easily compared using the figure, assuming that the ratio of 2.5 is valid in future scenarios. As seen in the figure, this also implies that lignin extraction will have a larger effect on ethanol production cost, unless the prices are very low.

An assumed ethanol production cost, as estimated by von Schenck et al. (2007), could be between 620 and 715 Euro per cubic meter, depending on what assumptions are made in the analysis. According to Fig 10 increasing the by-product vending in all three cases would lead to a decrease in production cost. For Case 1, where no energy efficiency measures are made, investing in turbines or lignin extraction could give 74-80 €/m³ lower production cost within the limits of this study. This in turn would mean up to 13% lower production cost. If energy efficiency measures are implemented, the production cost could decrease by up to 150 €/m³, which in turn could lead to over 20% lower production cost.

If we expect a power price of about 60 €/MWh in the future, and if we assume that the ratio estimated by Olsson et al. (2006) is correct and approximately 2.5, giving an estimated lignin selling price of 24 €/MWh, the corresponding decrease in ethanol production cost would be between 50 and 80 €/m³ ethanol when increasing

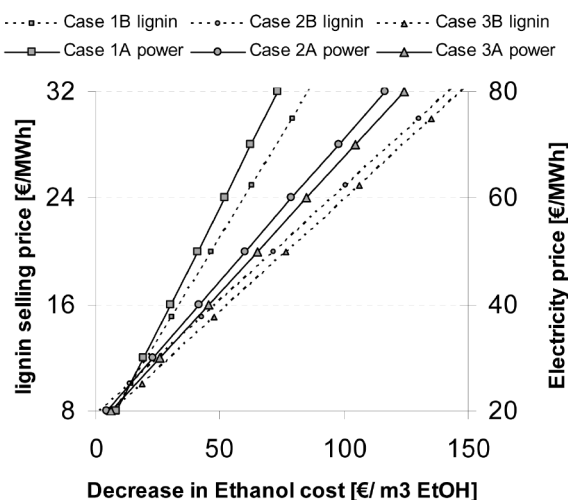


Fig 10. Effect on ethanol production cost if investing in case 1-3.

power production, and between 60 and 100 €/m³ ethanol for lignin extraction. Thus it is estimated here that lignin separation would have a bigger impact on the ethanol production price, and that the decrease could be substantial (up to about 15% in the case described here).

Conclusions

This paper introduces a process for lignocellulosic ethanol production, based on a converted kraft pulp mill. It also analyses possibilities of implementing energy efficiency measures in ethanol production. The possibilities of increasing energy efficiency, when implementing both conventional and process integration measures, are shown to be large. Approximately 54–62 MW of steam could be saved, resulting in a decrease in steam demand of about 36–42%. This can be compared to results gained if only pinch violations had been solved. The theoretical savings potential would then be approximately 25 MW. Thus the implementation of further energy measures will increase the theoretical steam-saving potential in the process from 17% to 42%, i.e. by a factor 2.5.

No distinction between the two alternative measures studied in this analysis can easily be made, but it can be stated that applying energy efficiency measures will generally give a decrease in ethanol production cost, especially if the by-product price is high. Process integration measures give slightly better results in the study, whilst applying conventional measures does give further possibilities due to the existence of unutilised excess heat, which have not been included in this work.

It is shown in the techno-economic analysis that using steam surplus for production of either lignin biofuel or power could have a large effect on the ethanol production price (decreasing it by up to 15%). This indicates the importance of energy efficiency studies in ethanol production.

It is also shown in the study that lignin extraction might be more interesting than increasing power production. One important reason for this is that CO₂ used in lignin separation can be produced on-site in the fermentation unit of the ethanol production plant.

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Nomenclature

GHG	Greenhouse gases
RME	Rapeseed Methyl Esters
DME	Di-methyl Esters
ADT	Air-dry tonne (90% dry substance)
DS	Dry substance
FRAM	"Future resource adapted mill", national swedish research programme
PT	Pre-treatment
LP	Low pressure steam, 4.5 bar (a)
MP	Medium pressure steam, 11 bar (a)
HP	High pressure steam, 60 bar (a)
C6	hexose-monosaccharide
C5	pentose-monosaccharide
ΔT_{\min}	Minimum approach temperature for heat exchange in the process (in pinch analysis)
GCC	Grand Composite Curves in pinch analysis (as described in Kemp 2007)
CC	Composite Curves in pinch analysis (as described in Kemp 2007)
BG	Background curve in pinch analysis (as described in Kemp 2007)
FG	Foreground curve in pinch analysis (as described in Kemp 2007)
HX	Heat exchanger
PBP	Payback period
a	annually
tdw	tonne dry wood
ww	warm water
t	tonne

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Appendix 1. Stream data for the ethanol process

Below the stream data used in the pinch analysis of the ethanol process is shown. The initial heat demand (prior to implementation of energy efficiency measures) is also presented, based on data from von Schenck et al. (2007).

Cold streams	Temp. [°C]		Load [MW]	$\Delta T/2$ [°C]
	Start	Target		
ww 85	18	85	3.1	2.5
Make up boiler w.	18	75	6.5	2.5
Building heating (LP)	21	22	1.1	8
Wood Yard (LP)	18	30	1.4	4
Wood chips (Flash steam)	99	100	11.2	3.5
Wood chips (LP)	100	120	7.0	3.5
LP to stripper	150	151	7.3	0.5
LP to evap	150	151	59.8	0.5
LP to rest	150	151	6.7	0.5
Dist1 preheater	37	100	13.5	5
Dist1 reboiler (LP)	123	124	16.8	2
Dist2 reboiler (MP)	164	165	2.1	2
W.liquor-digest (MP)	85	165	15	3.5
W.liquor-imp.(MP)	85	105	1.0	3.5
Digester circ. (MP)	165	170	2.2	3.5
Hi-Heat (MP)	128	160	8.9	3.5
MP to rest	200	201	2.6	0.5
HP steam	449	450	16.9	0.5

Hot streams	Temp [°C]		Load [MW]	$\Delta T/2$ [°C]
	Start	Target		
General cooling	40	35	9.0	3.5
Surface condenser	61	60	60	2
Excess condensate	60	39	0.6	3.5
Stripper condenser	100	100	5.1	2
Stripper sec.cond.	90	90	0.6	4
Steam smelt dissolver	80	60	3.7	2
Spent liquor fl ash steam 1	128	127	14.8	2
Spent liquor fl ash steam 2	109	107	14.3	2
Condensate	107	75	0.8	3.5
Black liquor	105	93	5.6	3.5
Digester bottom	90	89	0.7	3.5
Fermentation	90	37	10.3	5
Dist1 condenser	60	59	0.2	5
Dist2 condenser	92	90	14.6	5
Condensate from Dist1	123	47	13.5	5

Total steam demand	
LP	100 MW
MP	31.8 MW
HP	16.9 MW
Total	149 MW