

TECHNICAL ECONOMIC EVALUATION OF A HARDWOOD BIOREFINERY USING THE "NEAR-NEUTRAL" HEMICELLULOSE PRE-EXTRACTION PROCESS

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ABSTRACT

Forest products companies may increase revenue by producing biofuels and chemicals in addition to wood, pulp and paper products in a so called Integrated Forest Biorefinery (IFBR). One form of such a biorefinery is where the hemicelluloses, which normally end up in the black liquor of a hardwood Kraft mill, are extracted prior to pulping and used for the production of ethanol and acetic acid. The extracted liquor undergoes hydrolysis, separation, fermentation and distillation for the production of acetic acid and ethanol. A computer model was developed for this process using WinGEMS and ASPEN Plus software. The capital and operating cost for the process were estimated by standard methods. The amounts of ethanol and acetic acid produced are relatively small compared to current corn to ethanol and commercial acetic acid plants. This situation results because of the small amount of hemicelluloses extracted which is dictated by pulp quality and pulp yield considerations. The rate of return on investment varies between 7.1% and 13.0% depending upon the size of the pulp mill (750 to 1500 tonne per day) for the case where the utilities and waste treatment facilities are sufficiently large to handle the additional requirements for the process and a suitable vessel is available for upgrading to handle the extraction process. If the utilities and waste treatment system have to be upgraded, then the rate of return on investment decreases to 1.1% to 6.5% depending upon the size of the pulp mill (750 to 1500 tonne per day). In making the capital cost estimates 20% of the cost of a new installed digester was included as the cost of upgrading the extraction vessel and tying it in to the existing pulping process. Installing a new extraction vessel causes the discounted cash flow rates of return to be negative however because of the added capital cost.

INTRODUCTION

Background

Interest in renewable resources as a source of energy and chemicals has increased because of expected worldwide shortages of easy-to-access oil and gas [1], and growing concerns regarding the accumulation of greenhouse gases in the atmosphere. The concept of an Integrated Forest Products Biorefinery (IFBR) is being advanced by a number of investigators, who envision converting cellulose, hemicellulose and lignin from woody biomass, dedicated annual crops and municipal waste into bioenergy and basic chemicals [2, 3, 4, 5, 6, 7, 8, 9, 10]. The IFBR concept evaluation in this study seeks to integrate pulp production and a number of by-products derived from hemicellulose. In this process a portion of the hemicelluloses is extracted from wood prior to pulping and converted into acetic acid and ethanol while using the extracted wood chips to produce market Kraft pulp.

The main purpose of this work here is to present the results of a technical and economic analysis of a "Near-Neutral" hemicellulose pre-extraction process integrated within an existing hardwood Kraft mill. In this analysis an attempt was made to determine the economy feasibility of building a commercial biorefinery that is co-located at an existing Kraft pulp mill and is fully integrated with the mill in terms of mass and energy.

The existing Kraft pulp mill was considered as the base case. For the IFBR version the pulp production was maintained constant and the hemicellulose extraction process was added to the fiber line. The feedstock to the biorefinery was assumed to be mixed US Southern hardwoods (35% Gum (both sweet and black), 35% Southern Red Oak, 15% Red Maple, 12% Poplar and Sycamore and 3% Southern Magnolia). The feed stock is extracted in a separate impregnation vessel prior to the continuous digester for pulp production. In the analysis it is assumed that an existing continuous digester could be converted into the required extraction vessel.

The conceptual biorefinery process is shown in Figure 1. The pre-extraction and modified cooking process is carried out using green liquor (mostly $\text{Na}_2\text{CO}_3 + \text{Na}_2\text{S}$) and white liquor (mostly $\text{NaOH} + \text{Na}_2\text{S}$) respectively. The charge of green liquor used in the pre-extraction is about 3% (as Na_2O on dry wood) counted as total titratable alkali (TTA), while the charge of effective alkali (EA) in the modified kraft cook is reduced by 3% (as Na_2O on original dry wood) compared to the base case of conventional kraft cooking to

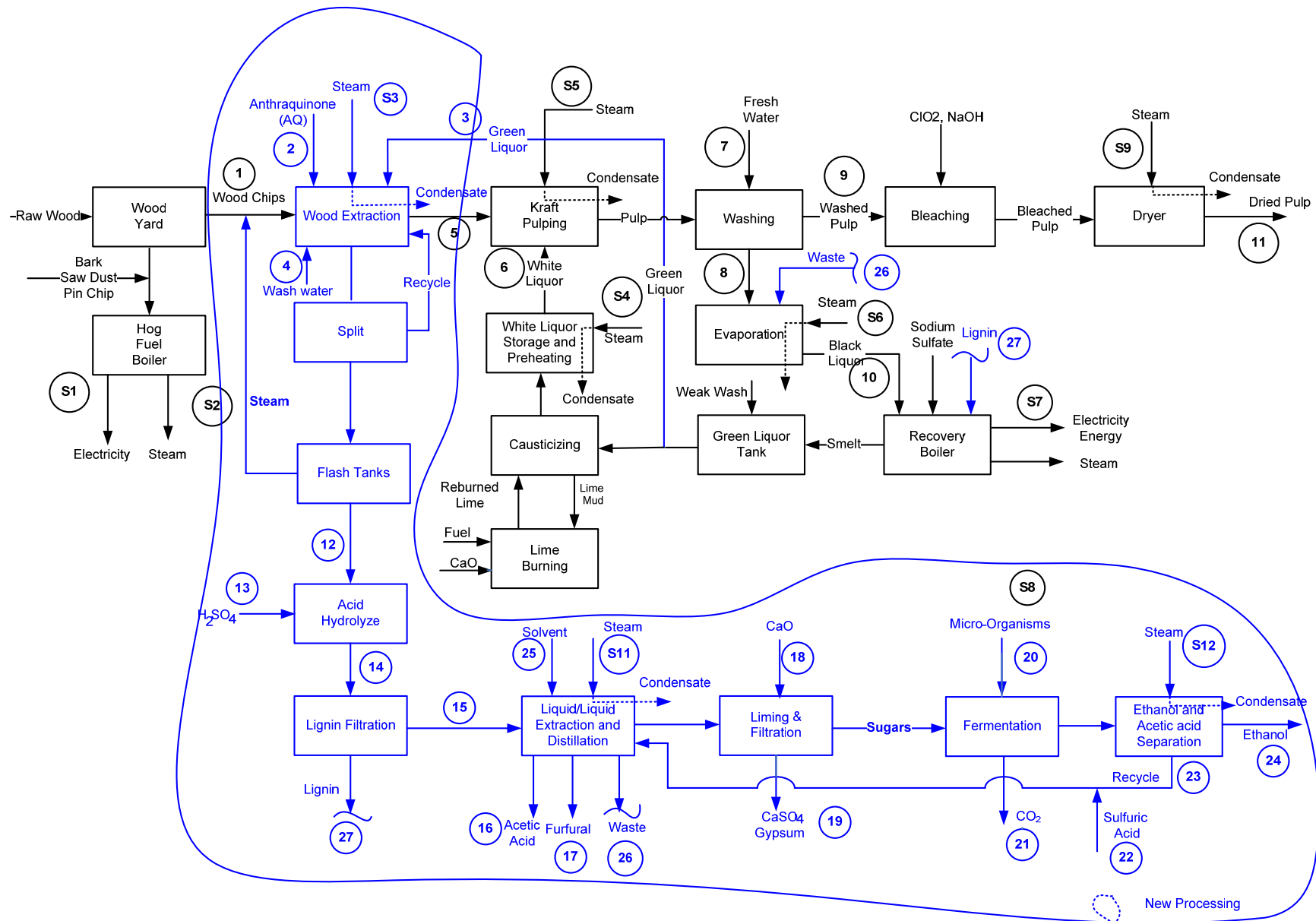


Figure 1. Modified Kraft Pulp Mill for the Production of Acetic Acid and Ethanol

the same bleachable grade hardwood pulp. A charge of 0.05% (on wood) of anthraquinone is included in the green liquor. The processing of the pre-extracted wood is shown in black in Figure 1 represents the conventional Kraft pulp mill. The unit processes required for hemicellulose extraction and conversion to ethanol and acetic acid are marked in blue. They include wood extraction for hemicellulose removal, flashing of the extract to produce preheating steam, recycling a portion of the extract back to the extraction vessel for the purpose of raising the solids content of the extract, acid hydrolysis using sulfuric acid for conversion of the oligomeric carbohydrates into mono sugars and cleavage of lignin-carbohydrate covalent bonds, filtration to remove precipitated lignin, liquid-liquid extraction followed by distillation to remove acetic acid and furfural from the sugar solution, liming to raise the pH to that required for fermentation, fermentation of five and six-carbon sugars and glucuronic acid to ethanol and finally distillation and upgrading the product to pure ethanol (99+%).

Process Advantages and Disadvantages of the “Near Neutral” Extraction Process

There are several important advantages associated with the “near- neutral” hemicellulose extraction process. Firstly, the quantity and quality of the pulp is unchanged compared to conventional Kraft cooking as shown in Figures 2 and 3 respectively. Figure 2 shows that pre-extraction with 3% green liquor (+0.05% AQ) at 160°C for 110 minutes followed by modified kraft cooking at 12% EA (30% sulfidity) gives the same yield within experimental error (blue triangular symbols) based on original wood as regular Kraft cooking at 15% EA and 30% sulfidity (red star symbols). Figure 3 shows that the tear-tensile strength of the pre-extracted Kraft pulp is as good as or slightly better than that of the Kraft pulp produced using conventional methods.

A second advantage of the “near-neutral” extraction process is that the recovery cycle is off-loaded because the amount of organics in black liquor is reduced and less white liquor is needed for pulping. This change would allow a mill to significantly increase its pulp production rate if the recovery cycle is the bottleneck. Environmental advantages of the extraction process are that the methanol content of the black liquor is reduced by about 40%, while its TRS content is also diminished because of the lower white liquor charge and shorter pulping time during the modified Kraft cook. Finally, a new feed stock stream of sugars is produced which may be used for the production of biofuels and renewable chemicals.

Disadvantages of the “Near Neutral” extraction process compared to autohydrolysis-based water extraction is that less sugars are removed and that the extract contains inorganic salts originating from the green liquor. These two aspects have a negative aspect on the economic production of biofuels such as ethanol from such an extract via biochemical processes. Also, the process involves several changes in pH which cause the operating cost to increase.

PROCESS DESIGN

Wood Composition

The main constituents present in wood are cellulose (42.6%), hemicellulose (29.7%) and lignin (27.4%). Additionally there is about 0.3% inorganic (ash) and about 2 to 3% extractives which are low molecular weight compounds consisting of fats (acids and alcohols), waxes and sterols [11] in hardwoods.

Plant Sizes and Assumptions

The basis for the design study is summarized by Mao [12]. Three sizes of pulp mills were considered; a small plant producing 750 tonne per day of market Kraft pulp, a medium plant of 1,000 tonne per day and a large pulp mill of 1,500 tonne per day. The base case is the Kraft pulp mill.

Pulping Conditions

The Kraft pulping conditions for the base case was 15% EA, 170 °C temperature to give an H-factor of 1500 hours. For the “Near-Neutral” hemicellulose pre-extraction process the total H-factor was maintained the same; that is an H-factor of 700-hours was used in the extraction vessel and 800 hours in the cooking digester. The corresponding pulping conditions after pre-extraction are 12% EA at 170 °C.

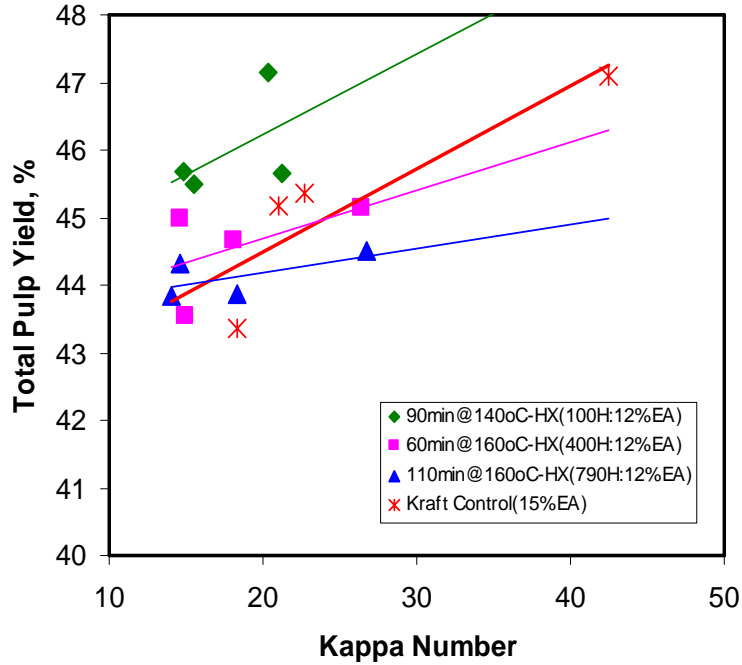


Figure 2 Yield vs. Kappa number

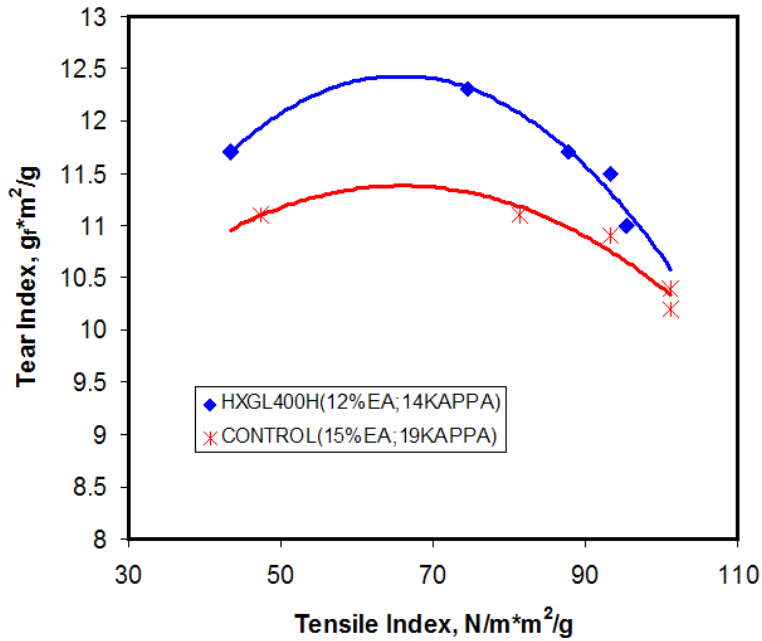


Figure 3 Tear Index vs. Tensile Index

Extraction Liquor

Green liquor was selected because it is produced in the recovery cycle of the Kraft mill, lignin precipitation is minimized in the final near-neutral pH extract, and the pulp yield (based on original wood) and physical properties of the pulp are maintained. The use of pure water as extraction liquid results in a low pH final extract (pH down to 3.5) and a lower total pulp yield as seen in Figure 4. The base case is a Kraft cook at 160°C, 15% EA, 30% sulfidity, L/W of 4.5 L/kg and an H-factor of 1500 hrs to give a yield of 46.6±0.3 %.

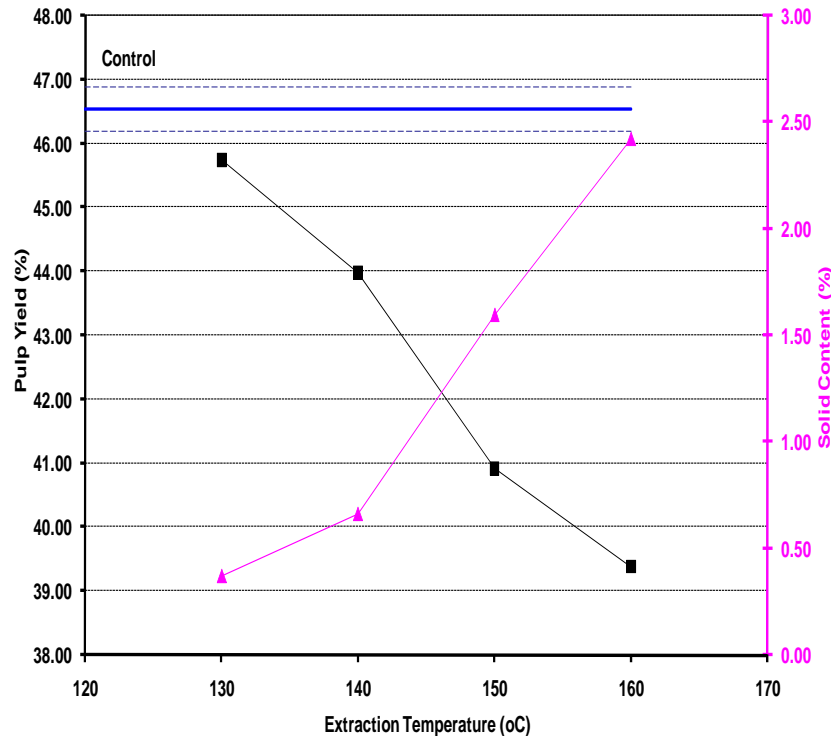


Figure 4 Kraft Pulp Yield and Solid Content versus Water Extraction Temperature

The total pulp yield after water autohydrolysis pre-extraction is also based on the original oven dry wood. The extraction time at the different temperatures was 90 minutes, except at 160°C which was 60 minutes. The kappa numbers of the pre-extracted Kraft pulps varied from 16.1 to 13.1. The total pulp yield was corrected to 16 kappa using a correction of 0.25%/kappa unit. Also shown in Figure 4 is the solid content of the extract. It can be seen that increasing extraction yield is correlated with a significant decrease in total yield of pre-extracted-Kraft pulp.

The composition of the “near- neutral” extract used in the design analysis is given in Table 1 and is based on extraction experiments performed by van Heiningen [13]. The main organic components in the extract are xylan, acetate groups, glucuronic acids groups and lignin. During the extraction process acetyl groups, which are side chains on the xylan hemicelluloses, are cleaved to give dissolved sodium acetate which will be converted to acetic acid in the hydrolysis stage of the plant.

Process Simulation

The simulation model for the modified Kraft pulp mill first was developed primarily using the WinGEMS computer code [14] which was designed for solids/liquid systems occurring in the pulp and paper industry. Compound blocks, or subroutines designed to simulate complicated unit operations were used in this model. In performing the energy balance calculations associated with distillation operations and also in designing the liquid-liquid extraction process modeling was done using ASPEN and by scaling data provided by Wooley et al. [8].

“Near-Neutral” Extraction Process

Wood chips are pre-steamed in a steaming vessel for 15 minutes and then heated to the extraction temperature of 160 °C in 60 minutes and then extracted for 110 minutes at 160 °C. Under these conditions approximately 10% of the mass of the wood is extracted (Table 1). The extraction liquid containing hemicellulose and lignin is separated from the chips, and the chips are conveyed into the pulping digester and cooked at 170 °C to a total H-factor of 1500 hrs including the pretreatment H-factor.

Lignin Removal and Acid Hydrolysis

After extraction, the extract liquor at high temperature (160 °C) and elevated pressure (110 psig) is sent to a flash tank to recovery steam and lower the temperature (130 °C) and pressure (20 psig). The dissolved solids content of the extract leaving the extraction vessel is approximately 8.5 %. The extract may then be further concentrated using a multiple effect evaporator system, but this was considered undesirable since it would concentrate the residual salts and adversely affect fermentation. The result of not-concentrating the extract means that higher volumes of liquor must be processed.

The hydrolysis of the component carbohydrates is assumed to follow the kinetics specified by Garrote and co-workers [15] for hydrolysis of extract from Eucalyptus using sulfuric acid. The concentration range of sulfuric acid studied was 0.5 to 4 weight % corresponding to a pH of 1.0 to 0.1 respectively. In our study we adopted a pH of 1.0 at 126°C for the hydrolysis. This required addition of sulfuric acid to make a 2.84% solution due to consumption of sulfuric acid by carbonate and protonation of acetate and dissociated uronic acids.

Table 1
Composition of Extraction for
1,000 Tonne per day Pulp Mill Production

Composition of Extract	Value
Flow rate	3,171 Tonne/day
Suspension solids content	0%
Dissolved organics	5.61%
Arabinan	0.06%
Galactan	0.17%
Mannan	0.11%
Glucan	0.22%
Xylan	1.889%
Acetyl group	1.11%
4-O-MGA	1.33%
Lignin	0.728%
Dissolved inorganics	2.84%
H ⁺	0%
Na ⁺	1.11%
SO ₄ ²⁻	0.70%
HCO ₃ ⁻	1.03%
Water	2,903 tonne/day

A hydrolysis time of 1.3 hours was needed for complete hydrolysis of the xylo-oligomers. At these conditions it was assumed that all the extracted lignin is precipitated and can be separated from the liquor by filtration.

Liquid-Liquid Extraction

After lignin separation a liquid- liquid extraction process is used to extract acetic acid and furfural from the mixture liquor. A sketch of the process , taken from the ASPEN+ simulation of the process, is illustrated in Figure 5. Two alternatives were considered.

Alternative 1. Acetic Acid Recycle. Alternative 1 is illustrated in Figures 1 and 5 where the acetic acid in the bottoms product from the crude ethanol distillation column is recycled back to the liquid-liquid extraction section of the plant (stream 23 in Figure 1). The acetic acid in this stream is a fermentation product of the extracted uronic acids (will be discussed later). In this case two liquid-liquid extraction columns are necessary to remove acetic acid. Also, additional sulfuric acid is required to lower the pH of the recycle stream to match the pH in the liquid-liquid extraction process. This then necessitates additional lime and gypsum removal.

Alternative 2. No Acetic Acid Recycle. An alternative process configuration was investigated in which the bottoms product from the ethanol and acetic acid separation section of the plant was treated as a waste stream and sent back to the recovery process. This alternative avoids putting in a second extraction column but reduces the acetic acid product by approximately 30% with concomitant loss of revenue. This alternative was rejected based on process economics.

In the configuration selected (Alternative 1), sugar feed from the hydrolysis reactor (stream 6 in Figure 5) is sent to the first extraction column (F) where it is extracted with an appropriate solvent. In the simulation reported here ethyl acetate was used as the solvent to extract acetic acid and furfural from the sugar/water/salt mixture. The raffinate from the extraction column F (stream 18), is composed of an aqueous mixture of sugars, which goes to the liming process. In the near-neutral extraction process, acetic acid is formed from the fermentation of uronic acids. Acetic acid ends up in the bottoms product of the ethanol upgrading column and is sent back to the liquid-liquid extraction process as stream 23 in Figure 1. Consequently in the liquid-liquid extraction process illustrated in Figure 5 a second liquid-liquid extraction column (Extractor E) is required to process this recycle stream which is illustrated as stream 8 in Figure 5 (acid-mix). Extract from both Extractors F and E is mixed and then sent to a solvent recovery column (G) where the solvent is recovered and the bottoms product is sent to a flash tank (H). The vapor fraction from the flash tank (13) is sent to an acid column (I) where acetic acid and furfural are separated. The bottoms product from the flash tank is mixed with the raffinate from extractor (E) and discharged as waste. The aqueous waste stream (stream 17 in Figure 5) contains predominately sodium sulfate with traces of acetic acid. The waste stream is blended back into the weak black liquor stream going to the evaporators (stream 26 in Figure 1). Based on this technology the recovery efficiency for acetic acid and furfural was estimated to be about 90%. The extracted acetic acid and furfural can be further purified by distillation (not shown) and are sold as final products

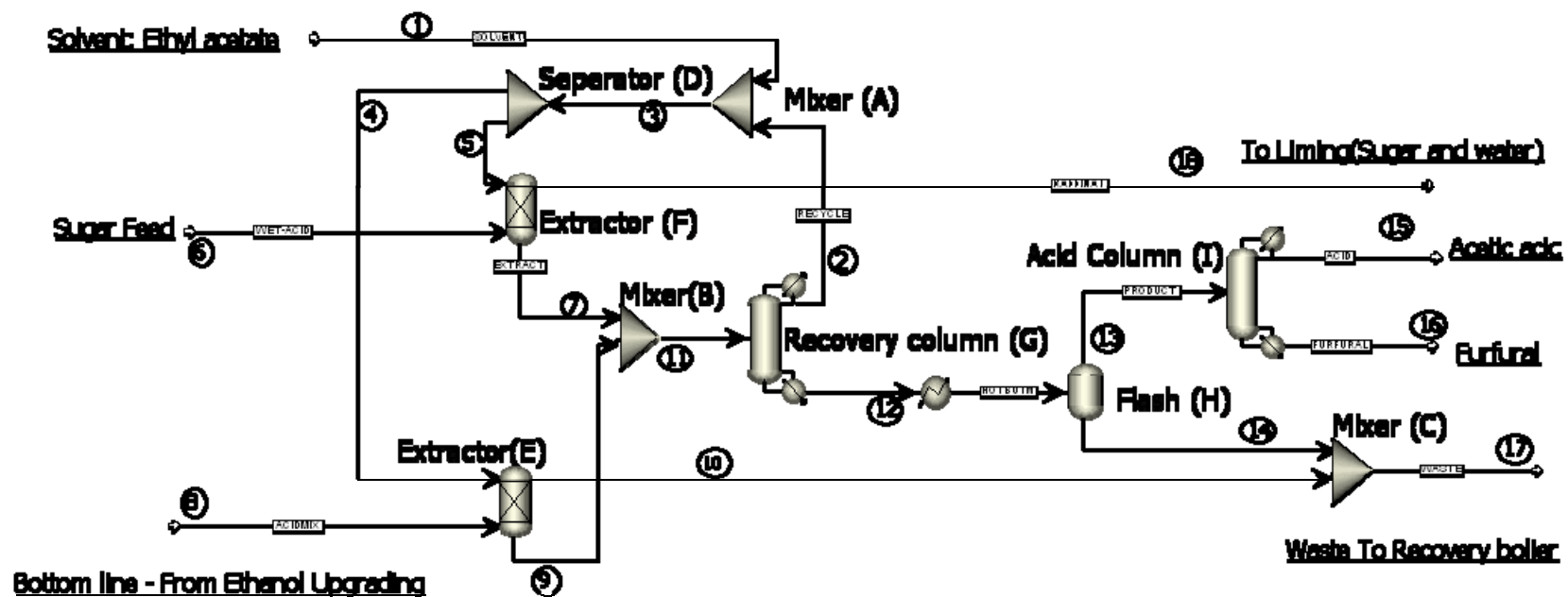


Figure 5. Liquid-Liquid Extraction Process for Separation of Acetic Acid and Furfural from Mixed Sugar Feed

Liming and Gypsum Removal

After liquid- liquid extraction and acetic acid and furfural recovery, calcium oxide (CaO) is used to “lime” the extract liquor. The main function of liming is to adjust the pH to near neutral condition for ethanol fermentation. The lime also acts as a bactericide. The calcium precipitates the sulfate ions as gypsum, $\text{CaSO}_4 \cdot 2\text{H}_2\text{O}$, to a concentration level which can be tolerated by the fermentation culture. Gypsum is removed by filtration.

Ethanol and Acetic Acid Production

In the fermentation step [8, 16, 17], it was assumed that C5 and C6 organics can be fermented simultaneously by using E.Coli (KO11). Also, glucuronic acid was assumed to be fermented into acetic acid and ethanol by using E. Coli B as described by Lawford and Rousseau [18]. The micro-organisms for conversion of the component sugars to ethanol would be produced on site. The efficiency of converting sugars and glucuronic acid into ethanol and acetic acid was set at 90% of the theoretical yield. Ethanol was processed after fermentation by pre-distillation to 50% purity and then further distillation to 95% which is the azeotrope concentration. The ethanol is then further concentrated to 99.9% by using molecular sieve technology [19]. Bottoms product containing acetic acid and sodium salts from the distillation columns was recycled back to the liquid-liquid extraction process.

Savings in the Lime Kiln

In the “near neutral” extraction process less white liquor is required in the cooking step. This will result in a corresponding decrease in the amount of calcium carbonate that needs to be removed in the white liquor clarifier and decomposed to lime in the kiln. This reduction in flow of CaCO_3 has a significant effect on the amount of energy required to operate the lime kiln. This is a distinct advantage of using green liquor rather than sodium hydroxide or white liquor in the pre-extraction step. The extraction step uses approximately 3% green liquor, calculated on dry wood basis and expressed at the total titratable alkali again expressed on a Na_2O basis. It allows a reduction in the effective alkali charge of white liquor in the modified Kraft cook of 3%; 15% EA in conventional cooking based upon dry wood basis and 12% EA in the modified cooking step. As a result, the reduction in CaCO_3 that must be processed in the lime cycle will be reduced by 80% based on the use of 30% Sulfidity in the white liquor or cooking liquor.

RESULTS

Production Rate

For a 1,000 tonne per day pulp mill the production rate for ethanol would be approximately 39.5 tonne/day (4.62 million gallons per year of 100% ethanol) and 31.6 tonne per day of acetic acid (3.97 million gallons per year of 100% acetic acid). In addition about 0.92 tonne per day of furfural are produced (320 tonne per year) and incinerated. Figure 6 illustrates the production rates for ethanol, acetic acid and furfural as a function of the production rate of the Kraft mill. Even for the 1,500 tonne per day Kraft mill, the amount of ethanol produced is relatively small when compared to a modern corn to ethanol plant which typically produces 50-100 million gallons per year [20]. Similarly, for comparison purposes, a modern acetic acid plant would have a production rate of 500,000 tonne per year, which is about 30 times larger than the capacity of the 1,000 tonne per day pulp mill [21]. However, although these wood derived chemical production rates are relatively small, the production of ethanol and acetic acid represents significant sources of revenue for the pulp mill.

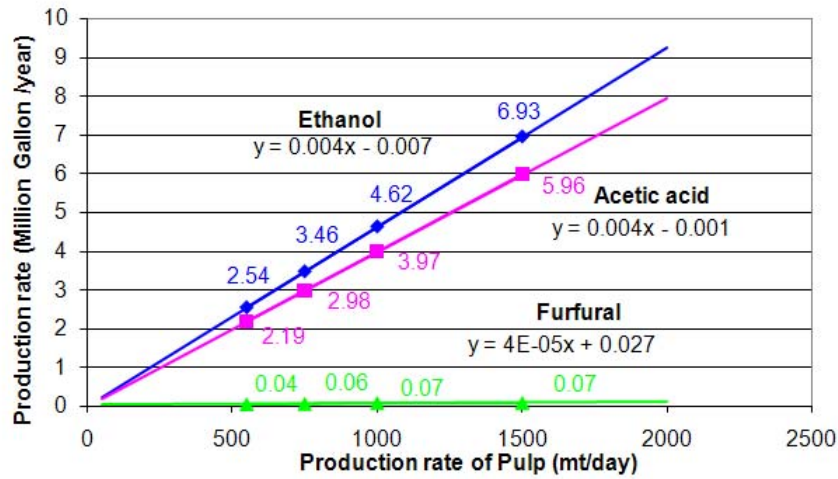


Figure 6 Products rate vs. plant size

Energy Consumption

The addition of the hemicellulose extraction process to a Kraft pulp mill reduces the energy which is obtained from the residual pulping liquor. No attempt was made in the present analysis to optimize the energy use by performing a pinch analysis [22]. The net energy output is illustrated in Figure 7 in terms of millions of BTU per hour as equivalent steam and electrical energy as a function of the pulp mill size. A comparison is made to the base Kraft mill case where no hemicelluloses are pre-extracted. The summary of the energy requirement for a 1,000 tonne per day pulp mill is illustrated in Tables 2 and 3. On average, the modified Kraft mill would produce approximately 35% less steam than the conventional Kraft mill because 10% of the wood mass is extracted and additional energy is required for the pre-evaporation and distillation operations [12]. Also, energy is lost during the extraction process because the extraction liquor is removed from the digester and flashed.

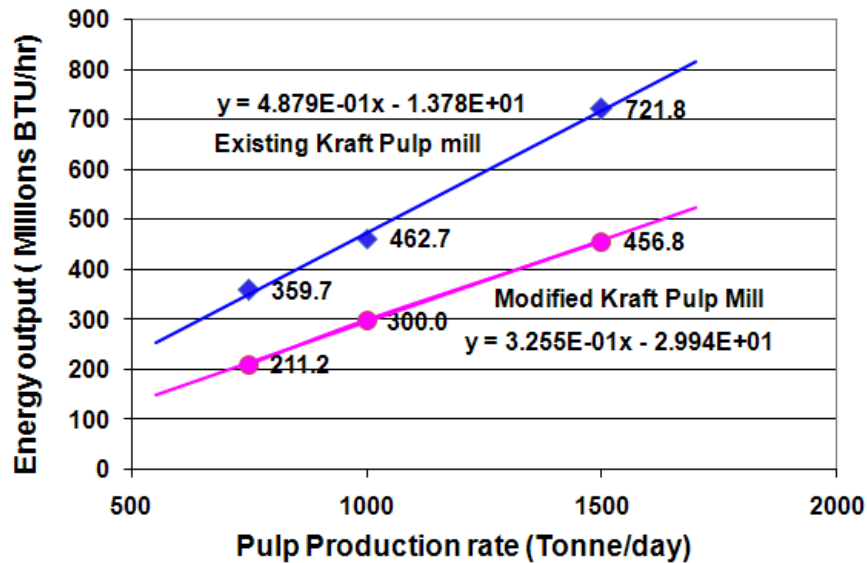


Figure 7 Energy output vs. plant size

Effect on Operation of the Lime Kiln

In the present situation the amount of Na_2CO_3 which needs to be processed in the lime cycle will be reduced by 20% for 30% sulfidity white liquor. This is equivalent to a reduction of 2.48% (as Na_2O on dry wood). The oil consumption per tonne of dry lime mud (CaCO_3) is 0.085 tonne of oil based upon the work of Gullichsen and Fogelholm [23], resulting in savings of about 9 tonne of fuel oil per day for a 1000 tonne per day pulp mill assuming a pulp yield of 47% and a causticizing efficiency of 80%. At a fuel oil price of about \$2.2/gallon the oil saving are approximately \$2 million dollars per year.

The savings in oil in the lime kiln of approximately \$2 million dollars per year helps offset the loss of steam which is used to manufacture the new products. The value of the lost steam, assuming a steam cost of \$7.00 per million BTU is \$9.6 million dollars per year.

ECONOMIC ANALYSIS

The economic analysis was performed following standard methods described by Peters and Timmerhaus [24] and Turton et al. [22]. In the analysis reported here it was assumed that an extraction vessel of some type was available at the site. For example in an existing Kraft pulp mill with two fiber lines in which one fiber line has been shut down, there may be an idle digester that could be converted into an extraction vessel and the plant operated as a single fiber line. A more likely situation in a single fiber-line pulp mill is that an impregnation vessel is available that could be converted into an extraction vessel since the wood must be impregnated with the pulping liquor in either case. In making the capital cost analysis, only 20% of the cost of a new continuous digester was applied to the project for modification of the impregnation vessel and/or connecting two continuous digesters in series. The capital cost estimates were obtained by scaling data provided by Wooley [8] and correcting for inflation and changes in unit price using the CEPCI index. In making the capital cost estimates a 15% contingency factor was applied to the estimated capital cost from the scaled data of Wooley [8]. Similarly, some mills would have sufficient waste treatment and utilities to accommodate the new process while other mills would have to expand their waste treatment and utility systems to account for the additional processing. Consequently two sub-cases were considered; one in which the utilities did not need to be upgraded and the case where they did.

**Table 2
Energy Balance for Modified 1000 Tonne per Day Kraft Pulp Mill**

Energy Balance Relative to Base Case				
Stream Number and Description		Existing Kraft Pulp Mill Case	Modified Kraft Pulp Mill	Units
Steam Required (Enthalpy)	S3 Extraction	0.0	84.6	MMBTU/hr
	S4 WL Preheater	20.9	15.6	MMBTU/hr
	S5 Upper Heater	84.6	41.5	MMBTU/hr
	S5 Lower Heater	41.8	64.7	MMBTU/hr
	S5 Washer Heater	29.6	41.1	MMBTU/hr
	S6 BL Evaporator	224.3	209.0	MMBTU/hr
	S9 Steam Dryer	68.0	68.2	MMBTU/hr
	S10 Extract Evaporator	0.0	0.0	MMBTU/hr
	S11 HAC Separation	0.0	11.8	MMBTU/hr
	S12 EtOH Distillation	0.0	12.7	MMBTU/hr
Total Steam Production (Enthalpy)	Hogful Boiler	129.4	129.1	MMBTU/hr
	Recovery Boiler	802.5	720.1	MMBTU/hr

Table 3
Summary of Energy Balance for Modified 1000 Tonne per Day Pulp Mill
Relative to Base Case

Stream Number and Description	Existing Kraft Pulp Mill Case	Modified Kraft Pulp Mill	Units
Steam Required in Extraction	0.00	84.6	MMBTU/hr
Steam Required in Pulping	469.2	440.1	MMBTU/hr
Steam Required in Ethanol Process	0.0	24.5	MMBTU/hr
Oil Savings in Lime Kiln	0.0	(15.3)	MMBTU/hr
Net Energy Output	As steam	462.7	MMBTU/hr
	As Electricity	27.1	MW

Note: Output of electricity was calculated by assuming 20% of the energy content in available biomass is converted to electrical energy.

Basis for Analysis

The capital cost was obtained from vender quotations on important equipment and by scaling cost data presented by Mitchell [9]. The basis for the economic analysis is summarized by Mao [12].

The plant was assumed to be constructed over a two year period. The selling price of ethanol was taken to be \$2 dollars per gallon while the selling price of acetic acid was assumed to be \$4 dollars per gallon. No additional wood was used in the process since experimental data showed that the overall pulp yield remains the same with the present “near-neutral” pre-extraction process. It was also assumed that the investment was completely borrowed and repaid over the life of the project which was taken to be 10 years. The capital recovery factor was assumed to be 10% based on a 10% interest rate; capital recovery factor of 16.3%. The viability of the project was determined by estimated the discounted cash flow rate of return and payback period as described by Turton [22].

Capital Investment

In making the capital cost estimate, it was assumed in one version of the analysis that the waste treatment and utilities were insufficient at the mill site to support the new process and had to be upgraded. The relationship between size and Total Project Investment is shown in Equation 1 for the case where utilities and waste treatment had to be upgraded to account for the additional processing of hemicellulose extract. These additional charges accounted for approximately 20% of the capital investment. Total project cost for the three plant sizes considered is summarized in Table 4. It is clear from the equation that the capital cost increases as a logarithmic function with the size of pulp mill. In this power law the scaling exponent is equal to 0.699 for pulp mill production rates between 750 and 1,500 tonne per day. This scaling exponent is slightly larger than the normal 6-tenths rule. Total project costs were estimated to be between 32.5 and 52.8 million dollars depending upon the size of the pulp mill. The capital cost could be lowered if the utilities and waste treatment system were of sufficient size to handle the added processing.

$$\text{Total Project Cost (\$)} = 317,219 \times (\text{Plant size, Tonne/Day})^{0.699} \quad (1)$$

Table 4
Estimate Capital Cost for the Near Neutral Hemicellulose Extraction Process Assuming Extraction Vessel Available and Upgraded Utilities

Pulp Mill Size (Tonne/Day)	Total Project Cost (Millions \$)
750	32.5
1,000	39.8
1,500	52.8

Operating Cost

The operating cost was estimated as a function of the size of the pulp mill and is shown in Equations (2). The operation cost for the various plant sizes were estimated by Mao [11] and fit to a linear equation.

$$\text{Operation Cost (\$/year)} = 17,470 (\text{Plant Size, Tonne/D}) + 1,315,000 \quad (2)$$

The unit production costs were estimated from the operating cost by assigning a portion of the operating cost to the production of both ethanol and acetic acid according to the amount of product produced on a mass basis. Using this simplified method of proportioning operating costs, approximately 43% of the operating cost was attributed to the production of ethanol and 57% to acetic acid. Table 5 illustrates the unit production cost for ethanol and acetic acid. Production costs for ethanol varied between \$1.94 for the large pulp mill (1,500 tonne per day) and \$1.85 per gallon for the small pulp mill (750 tonne per day). Similarly for the acetic acid the unit production cost varied between about \$2.46 and \$2.58 per gallon depending upon plant size.

Table 5
Ethanol Unit Production Cost

Pulp Mill Size (Tonne/Day)	Ethanol (Dollar/Gallon)	Acetic acid (Dollar/Gallon)
750	1.94	2.58
1000	1.90	2.52
1500	1.85	2.46

Discounted Cash Flow Analysis

An example of the discounted cash flow diagram is illustrated in Figure 8 for the 1000 tonne per day pulp mill for the case where the utilities and wastewater treatment system had to be upgraded. For this case the rate of return on investment was is 3.4%. Table 6 summarizes the discounted rate of return on investment for the different plant sizes. As the pulp mill size increases the discounted cash flow rate of return increase because additional product is produced and also because the operating cost per unit of product is lower.

Table 6
After Tax Discounted Cash Flow Rate of Return on Investment
Assuming Capital Investment Required for Upgrade of Waste Treatment System

Plant size (Pulp Production Rate)	Rate of Return on Investment
750 tonne/day	1.1 %
1000 tonne/day	3.4%
1500 tonne/day	6.5%

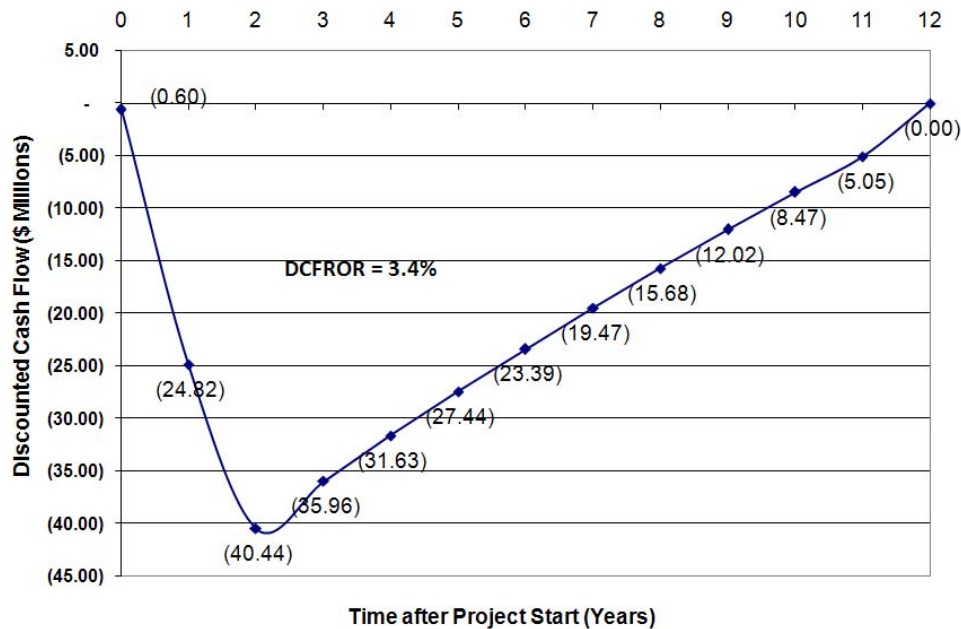


Figure 8. Discounted Cash Flow Diagram with Extraction Vessel Available for 1000 tonne per Day Case

Extraction Vessel AND Adequate Utilities and Waste Treatment Facilities Available

The case was considered where the utilities and waste treatment at the site were sufficient to handle the new process. Hence the capital cost would be reduced accordingly. Here again, it was assumed that the extraction vessel was available for hemicellulose extraction but needed to be modified to accommodate the extraction process and required approximately 20% of the cost of a new vessel. The capital cost, unit production cost and discounted cash flow rate of return for this scenario are summarized in Table 7. Approximately 20% less total capital investment is required; that is the total capital cost was estimated to be approximately \$25.7 million for the 750 tonne per day small pulp mill and \$42.2 million for the 1,500 tonne per day large plant. For this case, the unit production costs are lowered to approximately \$1.76 to \$1.80 per gallon of ethanol and to \$1.80 to \$2.38 per gallon for acetic acid depending upon the plant size.

Table 7
Economy Analysis for Case of Extraction Vessel Available and No Waste Treatment and Utility Systems Upgrade

Plant Size (Tonne/day)	Capital Cost (Million Dollar)	Unit Production cost (\$/Gallon)		DCFROR
		Ethanol	Acetic acid	
750	25.7	1.80	2.38	7.1%
1000	31.6	1.78	2.06	10.0%
1500	42.2	1.76	1.80	13.0%

CONCLUSIONS

The hardwood biorefinery considered here uses green liquor with the addition of AQ for wood extraction prior to modified Kraft cooking to preserve both pulp yield and quality. During extraction about 10% of the wood goes into solution. The extract contains mostly hemicellulose derived organic compounds and has a near-neutral pH. The extracted chips require a 3% lower EA charge (as Na₂O on original oven dry wood) of white liquor and a lower H factor in the subsequent kraft cook as compared to a regular kraft cook.

In the present analysis, the extraction and pulping were assumed to be conducted using continuous processes. The assumption inherent in the analysis reported here is that an existing impregnation vessel can be converted into an extraction vessel. The rate of return on investment varies between 1.1% and 13.0% depending upon the plant size and whether the onsite utilities and wastewater water are of sufficient capacity to handle the additional processing from the hemicellulose extraction process. Installing a new extraction vessel causes the discounted cash flow rates of return to be negative however because of the added capital cost. There are significant advantages to the new process, namely the pulp yield and quality are not degraded, the recovery boiler and the lime kiln are off loaded by about 20%, and two new by-products from the pulping process are generated. The off loading of the recovery cycle allows a potential increase in production rate if this section of the mill is forms the bottleneck for the production rate.

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