

Lignin recovery from kraft black liquor: preliminary process design

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ABSTRACT *A process design and a capital cost estimate were prepared for a lignin recovery process. The process includes CO₂ precipitation, lignin separation, acid washing, and product drying. The process equipment was sized to recover 9.5% of the lignin originally present in the black liquor for a typical kraft mill producing an annual average of 700 a.d. tons/day. The economic evaluation was based on the sale of 21% of the recovered lignin as a specialty chemical, with the remainder used as a fuel for the lime kiln. A computerized mass balance was developed to examine the impact of lignin recovery on the mill chemical balance. The proposed process would have a total installed capital cost of about \$19 million (Canadian) and a pay-back time of four years.*

KEYWORDS

Black liquor
Carbon dioxide
Kraft pulping
Lignin
Lime kilns
Process simulation
Recovery

Lignin from chemical pulping processes is a renewable source of many valuable chemicals currently derived from petroleum. Unlike the recovery of lignosulfonate, the recovery of kraft lignin has not been practiced widely. Not only is kraft lignin valued as a specialty chemical, but it could also replace some or all of the fossil fuel being used in the lime kiln at a kraft mill. Furthermore, at mills limited by recovery boiler capacity, removing some of the lignin from the kraft black liquor would allow an increase in pulp production.

A study was conducted (a) to define the process parameters for a kraft lignin recovery process using carbon dioxide for black liquor acidulation and (b) to evaluate the process economics. As part of this study, a preliminary design was developed for the process equipment, and a capital cost estimate was prepared (1).

Background

Many kraft mills are operating at or above their design capacities. Pulp production is sometimes limited by the amount of black liquor that can be burned in the recovery boiler. Removing some of the lignin from black liquor would decrease the amount of black liquor solids per metric ton of pulp and the heating value of the solids, allowing an increase in pulp production for a limiting load on the recovery boiler. Alternatively, removing some of the lignin from the liquor would provide the option of pulping to a lower kappa number at the same heat load on the recovery boiler. A decrease in the kappa number would generate savings in bleaching chemicals and decrease environmental discharges from the bleach plant.

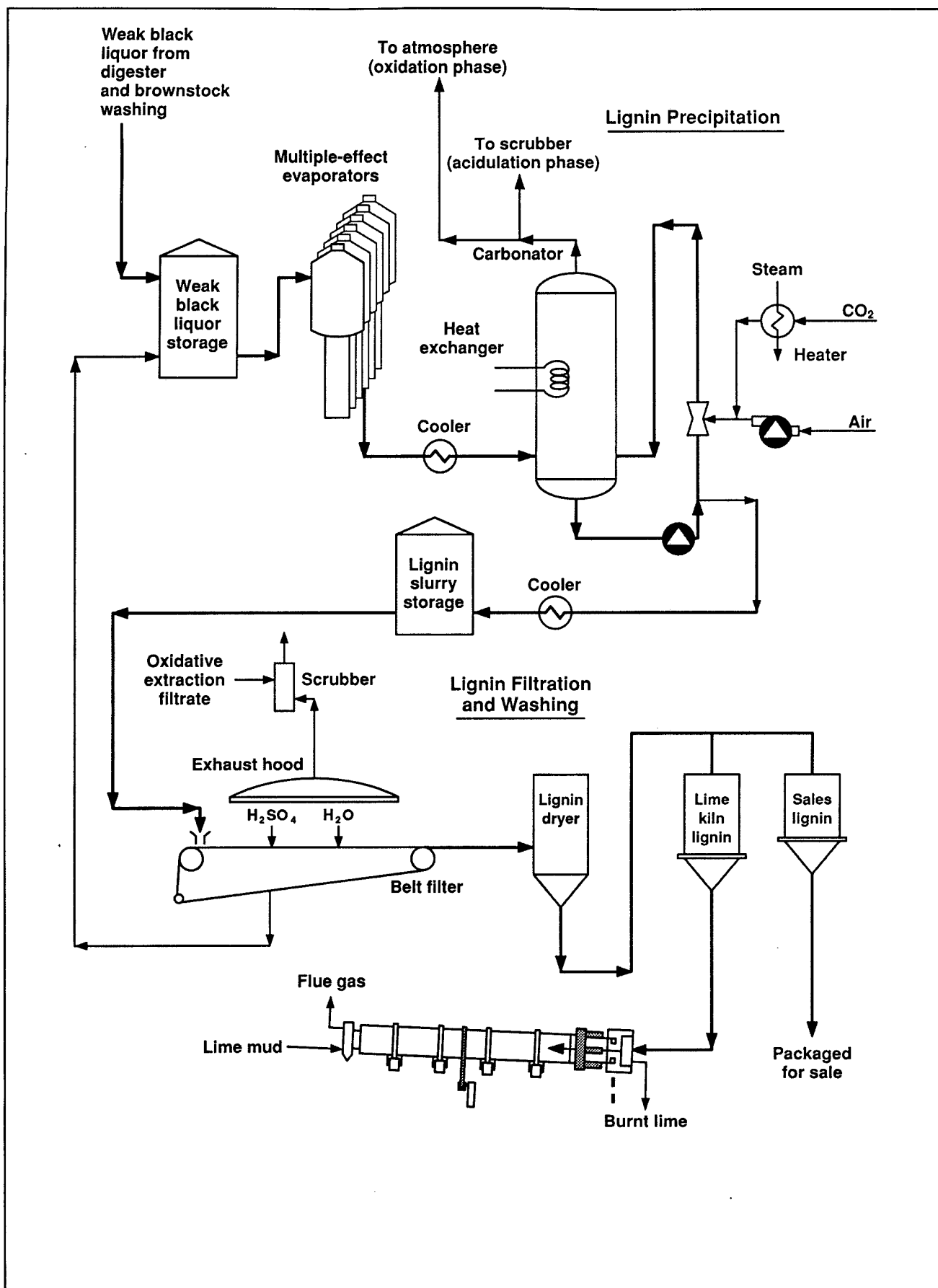
In modern kraft pulping technol-

ogy, fossil fuel requirements for lime kiln operation represent the final barrier to energy self-sufficiency. The lignin removed from the black liquor could replace some or all of the fossil fuel required for the lime kiln (2) and decrease fuel costs. Furthermore, the lignin itself is of value as a specialty chemical. Products produced from kraft lignin include additives for cement, ingredients in printing inks and thickeners, animal feedstock, artificial vanilla, and dispersing agents for dyes (3).

Lignin recovery method

Lignin may be recovered from black liquor by ultrafiltration or acid precipitation. Ultrafiltration is used when lignin of a specific molecular weight is required for a special product (4). In a kraft mill, CO₂ precipitation and H₂SO₄ precipitation are the

1. Flow diagram showing the concept for kraft lignin recovery using carbon dioxide for acidulation



most appropriate methods (4, 5).

H₂SO₄ has been used for removing lignin in some mills (6). At many mills, H₂SO₄ is already added to the black liquor stream in the form of waste acid from older ClO₂ generators. By adding this waste acid to a portion of the black liquor stream, instead of the whole stream, the pH of the black liquor could be lowered enough for lignin to precipitate. However, there is a trend in kraft mills to replace the ClO₂ generator with a methanol-based process, which does not have a waste acid stream. Therefore, lignin recovery based on waste acid from ClO₂ generators acid may not be an available option in the long term. Fresh H₂SO₄ could be used, but the quantity of H₂SO₄ required would upset the liquor cycle chemical balance with excess sulfur. Consequently, CO₂ is preferred over H₂SO₄ for acidulation. Not only does the use of CO₂ minimize the disruption of the mill chemical balance, but it also provides a lignin precipitate that is easier to filter (9).

Initially, the CO₂ could be supplied as a purchased chemical. Once the attractiveness of the process was confirmed, consideration could be given to installing a gas absorption system to recover CO₂ from the lime kiln flue gas; this addition would decrease the operating costs for lignin recovery. It may be possible to contact the black liquor with recovery boiler or lime kiln flue gas to provide a direct source of CO₂ (7). However, the rate of transfer of CO₂ from the flue gas to the liquor probably would limit the effectiveness of the precipitation. A pilot plant utilizing a CO₂ precipitation method was operated with success (8), reportedly, in an eastern Canadian pulp mill.

HCl could be used for the acidulation, but it would result in an intolerably high input of chloride to the liquor cycle. Overall, CO₂ is the best choice for acidulation, especially over the long term.

Process description

Figure 1 illustrates the concept for a kraft lignin recovery process (8) using carbon dioxide for acidulation. In this process, a portion of the black liquor, with a solids content of 30–40%, is diverted to the acidulation reactors from the appropriate effect of the black liquor evaporators. Four batch

reactors would be used, rather than continuous reactors, to allow precise control of the pH–time profile, which is critical in obtaining a filterable lignin precipitate.

To minimize H₂S emissions during the acidulation process, the black liquor would first be oxidized using air. After oxidation, the liquor would be cooled to about 80°C. Then CO₂ would be injected into the reaction vessel to initiate acidulation. As the pH decreased, lignin, with its chemically bound sodium and sulfur, would precipitate out of the solution. A final pH of about 8.5 would minimize CO₂ use per unit weight of lignin precipitated (9, 10).

The resulting liquor slurry would be further cooled and then filtered. The filtrate, containing the soluble parts of the lignin and other dissolved solids, would be recycled back to the chemical recovery cycle. The filter cake would be acidified to remove some of the bound sodium and then would be washed with water. The washed lignin precipitate would be dried to an approximate 5% moisture content. The dry product would be either fired to the lime kiln or packaged for sale on the chemical market.

Researchers have examined the effects of liquor solids concentration, pH, temperature, and pressure of the carbonation stage on the lignin recovery process (5–13). Alen *et al.* (9) found that lignin recovery from pine kraft black liquor at a carbonation pressure of 1500 kPa was maximized at a solids content of approximately 27%, whereas for birch liquor, optimal solids concentration was between 30% and 35% (13). Dilute liquors require large amounts of CO₂ for carbonation, and they generate low lignin yields. Highly concentrated liquors result in a highly viscous liquor in the reactor, causing mixing and handling problems (14). Furthermore, acidulation of stronger black liquor (42% solids) produces a finer lignin precipitate with a greater concentration of chemically bound sodium (5). For black liquor from Canadian wood species, the optimum dissolved solids content for acidulation would be 30–40%.

Lignin yield increases as the pH of the pulping liquor decreases (5, 6, 9). At a pH of 2–3, obtained using H₂SO₄, greater than 90% of the lignin can be recovered (5, 9). However, at such low pH, the lignin is precipitated in a gel-

like form that is difficult to filter and wash (5, 7). Furthermore, acidulation to a low pH produces large quantities of H₂S gas. The filterability can be improved if acidulation at low pH takes place in the presence of small amounts of certain organic liquids (12). A pH of 8–9 appears to be most suitable for the carbonation stage: it results in an easily filtered product, provides a 65–75% yield of precipitated lignin, minimizes chemical costs, and avoids excessive emissions of H₂S (5, 9). In the proposed process, the discharge of H₂S gas would be further minimized by using air to oxidize the Na₂S in the liquor to Na₂S₂O₃, prior to carbon dioxide acidulation.

Temperature also affects the filterability of the lignin (5–8, 10). The filtration rate increases as the temperature increases up to a certain limit, after which the lignin precipitates as a soft, tar-like substance that is difficult to filter. Optimum filterability occurs at carbonation temperatures of 80–85°C, just below the coagulation point of the lignin (10).

The pressure during the acidulation stage also influences yield. Higher pressure results in greater lignin recovery and decreases the carbonation time required to achieve maximum lignin yield (9).

After the lignin has precipitated, the slurry should be cooled to prevent the formation of gummy lignin, which would impede filtration (6, 8). The first filtration phase would separate the precipitated lignin from the liquor. The liquor filtrate, which contains dissolved lignin and other dissolved solids, would be piped to the weak black liquor storage tank. The sodium content of the precipitated lignin could inhibit the use of the product as lime kiln fuel. The presence of sodium compounds in the lime kiln tends to promote the formation of rings and balls, which interfere with kiln operation.

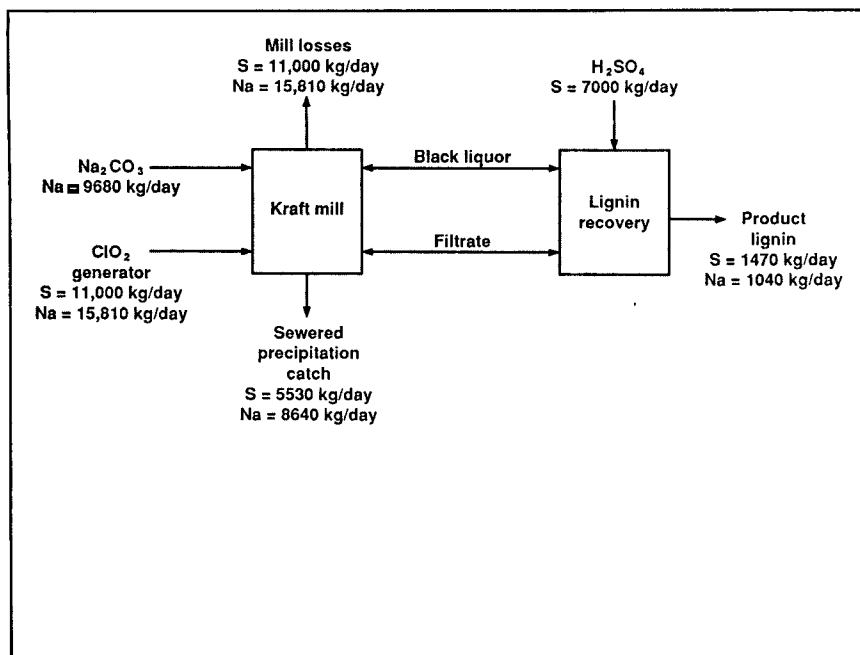
Acid washing would remove some of the sodium chemically bound to the lignin. Studies at the Pulp and Paper Research Institute of Canada have shown that two H₂SO₄ washes reduce the sodium content of lignin (precipitated at pH 8.5 using sulfuric acid) sufficiently that the lignin may be considered as fuel for the lime kiln (5). Acid is consumed in removing the bound sodium from the lignin. Sodium is bound to carboxylic and phenol-

I. Process operating conditions for the lignin recovery process in a mill with production limited by the recovery boiler capacity

Pulp production		Wash water, kg/kg precipitated lignin	4.2
Annual avg. daily production, a.d. tons/day	700	Lignin lost in water, ^e %	20.4
Operating days/year	340	Overall loss, kg/kg precipitated lignin	0.387
Annual prod., a.d. tons/year	238,000	Combined liquor filtrate returned to chemical recovery	
Running rate, a.d. tons/day	770	Dissolved solids in liquor filtrate	22.5
Incremental profit on extra pulp produced, CAD/a.d. ton	300	Increase in evaporation requirements, kg H ₂ O/a.d. ton	378
Weak black liquor ^a		Increase in steam required, In tons/day	65
TDS, ^b kg/a.d. ton	1850	In %	5.8
Gross heating value, kJ/kg	13,900	Lignin drying	
Raw lignin, kg/a.d. ton	780	Moisture content of dried lignin, %	5
Sodium content, % by wt.	10.8	Energy use to dry lignin, kJ/kg washed lignin	3250
Sulfur content, % by wt.	1.1	Natural gas usage	
Free Na ₂ S, kg/a.d. ton	100	m ³ /a.d. ton	6.2
Free NaOH, kg/a.d. ton	105	CAD/a.d. ton	0.44
Free Na ₂ CO ₃ , kg/a.d. ton	91	Equipment sizing	
Multiple-effect evaporators		Lignin recovered, dry basis	
Liquor solids conc., %		In kg/a.d. ton	74
Entering evaporators	18	As % of original lignin in weak black liquor	9.5
In soap skimmer (to lignin precipitation)	30	Normal lignin production, dry basis, tons/day	57
Exiting evaporators	50		
Steam economy, tons of water/ton of steam	4.5	Lime kiln	
Soap solids removed		Product lime, kg/a.d. ton	220
kg/a.d. ton	50	Energy required, GJ/ton kiln prod.	7.4
Percent solids, %	60	Natural gas, gross heating value, kJ/m ³	39,000
Heating value, kJ/kg	32,000	Cost of natural gas, CAD/1000 m ³	90
Liquor oxidation		Lignin fired to kiln, kg/a.d. ton	58
Initial Na ₂ S content of dissolved liquor solids, %	5.4	Lignin gross heating value, kJ/kg	25,000
Degree of oxidation of Na ₂ S	95	Lignin substitution for fuel, % of total heat consumed	89
O ₂ addition, % of theoretical	150	Fuel value of lignin, CAD/kg	0.058
Air usage		Natural gas savings	
In m ³ /s	0.40	m ³ /a.d. ton	37.2
In m ³ /a.d. ton	45	CAD/a.d. ton	3.35
Lignin precipitation		Lignin sales	
Percent of black liquor to precipitation, %	21	Low-quality, kg/a.d. ton	8
Lignin yield, ^c %	70	In metric tons/year	2000
Precipitated lignin, kg/a.d. ton	121	In terms of CAD/kg	0.50
Carbon dioxide		High-quality, kg/a.d. ton	8
Applied		In metric tons/years	2000
kg CO ₂ /kg precip. lignin	0.30	In terms of CAD/kg	1.00
kg CO ₂ /a.d. ton	36	Recovery boiler benefits	
Consumed by NaHS and NaOH		Decrease in black liquor solids	
kg CO ₂ /kg precip. lignin	0.10	In kg/a.d. ton	79
CAD/1000 kg	130	As % of solids at gun	4.0
CAD/a.d. ton	4.68	Decrease in gross heat load, %	7.7
Final pH	8.5	Pulp equivalent increase in recovery boiler capacity, %	7.7
Portion of residual Na ₂ S converted to H ₂ S, mole %	5	Net steam, after soot blowing, based on weak black liquor	
H ₂ S formed, kg/a.d. ton	0.02	after soap skimmer, kg/kg BLS	3.65
H ₂ S to atmosphere, ^d kg/a.d. ton	0.002	Decrease in steam production, kg/a.d. ton	505
Precipitated lignin		Liquor cycle impact, kg/a.d. ton	
Sodium content, % by wt.	10.8	Additional sulfur input with H ₂ SO ₄ , as sulfur	9.1
Sulfur content, % by wt.	2.0	Sulfur loss with product lignin	1.9
Lignin washing		Precipitator catch sewer	34
Cons. of lignin slurry	50	As sulfur	7.2
Washed lignin, kg/a.d. ton	74	As sodium	11.2
Sodium content, % by wt.	1.5	Sodium loss with product lignin	1.3
Sulfur content, % by wt.	2.1	Na ₂ CO ₃ makeup	
Acid applied		As sodium	12.5
kg H ₂ SO ₄ /kg precip. lignin	0.23	As Na ₂ CO ₃	29
kg/a.d. ton	28	Na ₂ CO ₃ cost, CAD/1000 kg	190
Lignin lost by removal of chemicals		Power Boiler	
during acid washing, ^e %	18.5	Additional steam, kg/a.d. ton	590
H ₂ SO ₄ cost, CAD/1000 kg	70	Cost per ton additional steam, CAD	3.00
CAD/a.d. ton	1.96		

^aBefore filtrate is returned from lignin recovery system. ^bTotal dissolved solids. ^cPrecipitated lignin as portion of raw material originally present. ^dScrubbed with (EO) filtrate. ^eAs portion of precipitated lignin.

2. Mass balance for sodium and sulfur



ic groups. The acid reacts with these groups as well as with free NaOH, Na_2CO_3 , and NaHCO_3 in the residual liquor. Subsequent water washes would remove the acid wash residuals. The use of water should be minimized, however, because water solubilizes some of the lignin, resulting in product loss. In contrast, washing with dilute H_2SO_4 does not solubilize the lignin.

The process design was based on drying the lignin to a moisture content of 5% or less, prior to being fired in the lime kiln. Computerized mass and energy balances have indicated that the lignin moisture content should have little impact on kiln temperatures and gas velocities (15). However, firing a lignin slurry has not yet proven technically feasible. A moisture content of 5% or less assures a free flowing powder that would avoid lignin handling problems (15).

Trials on a pilot rotary kiln demonstrated that acid-washed lignin could be used as the sole fuel, while maintaining acceptable lime quality without sacrificing any kiln capacity (16). In fact, there are some indications that lignin would be a superior fuel to natural gas (16).

The marketed lignin would be sold in varying qualities. Low-quality lignin would be used for such purposes as phenolic resin substitutes, asphalt conditioner, and a cattle feed

additive. High-quality lignin would be used to manufacture speciality chemicals, including pharmaceuticals.

Mass balances

A computer simulation using MASSBAL MARKII was used to determine the material balances for the lignin recovery process and to examine the impact of the lignin recovery process on the mill chemical liquor cycle. The mass balance was also used to quantify the potential H_2S gas emissions from the CO_2 acidulation process and to estimate the composition of the filtrate from the wash stages. By altering process parameters in the computer simulation, calculated data were made to fit laboratory washing data derived at Paprican. Using the resulting mass balance data, capital and operating cost estimates for the process were calculated.

The computer model was based on the lignin recovery process described, for a typical kraft mill producing 700 tons/day on an annual average. It was assumed that the capability of the mill to produce pulp was limited by the capacity of its recovery boiler. Process operating conditions are listed in Table I. It was taken that lignin would be used to displace 89% of the fossil fuel in the lime kiln, and an allowance was made for 4000 metric

tons of lignin per year being sold on the chemical market. The amount of lignin recovered would be 74 kg/a.d. ton, representing about 9.5% of the raw lignin originally in the weak black liquor. The equipment was sized for 57 tons/day of product lignin, at a mill running at a rate of 770 a.d. tons/day. Approximately 21% of the black liquor would have to be diverted from the multiple-effect evaporators to the lignin recovery system. The return of the liquor and wash filtrates to the liquor cycle system would increase the evaporation load on the evaporators by about 5.8%.

In laboratory and pilot plant tests, acid washing followed by two or more stages of water washing was found to reduce the sodium content of kraft lignin to an acceptably low level. The acid requirement for washing a CO_2 -precipitated lignin (about 0.23 kg H_2SO_4 /kg of lignin) is higher than for washing lignin precipitated using mineral acids (5). Much of the additional acid is consumed in reacting with entrained carbonates, formed in the reactor by the reaction of CO_2 and the free NaOH released from the lignin during acid washing.

The return of the filtrates from lignin washing to the liquor recovery cycle would disrupt the chemical balance of the liquor cycle at some mills, but process measures could be adopted to overcome this disturbance. Figure 2 illustrates a worst-case situation, in which the mill liquor cycle was assumed to be perfectly balanced without the lignin recovery process. Makeup requirements for sodium and sulfur were met entirely by the return of salt cake from the ClO_2 generator. In such a case, any addition of either sodium or sulfur would disrupt the balance. With the lignin recovery process, there would be a loss of sodium and sulfur with the product lignin, but the sulfur introduced with the H_2SO_4 wash water would exceed that sulfur lost in the product lignin.

The best method to accommodate this would be to sewer some of the recovery boiler precipitator catch, which is mainly Na_2SO_4 , accompanied by small amounts of Na_2CO_3 and NaCl. Sewering about one third of the precipitator catch would allow the sulfur makeup to be balanced with the losses, but it would cause an additional sodium loss. Consequently, 29 kg of

II. Capital cost estimate for a 57-ton/day kraft lignin recovery process, CAD

	Material	Installation	Total
Reactor	1,000,000	150,000	1,150,000
Filter	800,000	200,000	1,000,000
Dryer	850,000	120,000	970,000
Solids handling	150,000	100,000	250,000
Kiln burning	100,000	100,000	200,000
Heat exchangers	150,000	30,000	180,000
Pumps and tanks	300,000	100,000	400,000
Air compressor and fans	250,000	50,000	300,000
Evaporator changes	800,000	200,000	1,000,000
Site preparation and services	400,000	200,000	600,000
Building (new construction)	750,000	550,000	1,300,000
Equipment foundations	400,000	200,000	600,000
Instrumentation	700,000	250,000	950,000
Piping	600,000	900,000	1,500,000
Electrical	500,000	400,000	900,000
Subtotal	7,750,000	3,550,000	11,300,000
<i>Indirect costs</i>		<i>Summary of Costs</i>	
Eng. and construction management	1,800,000	Total indirect cost	3,000,000
Owner's overhead	900,000	Plus total direct cost	11,300,000
Capital spares @ 5% of process material equipment cost	300,000	Total direct and indirect cost	14,300,000
Subtotal	3,000,000	Plus contingency (approx. 30%)	4,700,000
		Total installed capital cost	19,000,000

These figures are based on a new facility in a central B.C. location. Costs not taken into account include: (i) any additional research and development costs, (ii) any startup and commissioning expenses, (iii) allowances for crew training.

Na_2CO_3 /a.d. ton would have to be added to the liquor cycle to compensate for the sodium lost in the sewer recovery boiler precipitator catch and the sodium lost in the product lignin. Sewering some of the precipitator catch would have the added advantage of purging some NaCl from the liquor cycle.

The Pulp and Paper Research Institute of Canada is studying methods to decrease the H_2SO_4 requirements for lignin washing, to minimize the disruption to the liquor cycle chemical balance, and so to minimize the Na_2CO_3 makeup costs.

Energy balances

By burning lignin in the lime kiln, 89% of the fossil fuel requirements would be displaced. A small amount of fossil fuel would be needed to dry the lignin before firing it to the lime kiln. With less lignin in the black liquor fed to the boiler, the amount of steam generated in the recovery boiler per air-dried ton of pulp would decrease. By increasing the pulp production as proposed, the steam generation in the recovery boiler, in tons per hour,

would be maintained. However, with 7.7% more pulp production, up to 7.7% more steam would be required, in tons per hour. This extra steam would have to come from the power boiler. In addition, 5.8% more steam would be needed in the multiple-effect evaporators because of the recycling of filtrate from washing the lignin.

In total, the necessary steam production from the power boiler would increase by 590 kg/a.d. ton, or 19,000 kg/h. For the feasibility study, it was assumed that the power boiler would have the capacity to provide this extra steam using hogged fuel, at a cost of CAD 3.00 (Canadian dollars) per ton of steam. If the power boiler did not have sufficient capacity to make the steam using hogged fuel, fossil fuel would have to be used to generate the additional steam, resulting in a steam cost of about CAD 8.00 per ton of steam.

Equipment design and capital cost

The process parameters and values listed in Table I were used to size the major pieces of equipment. The equip-

ment design was partially based on work conducted by students in a project at the University of British Columbia (1). Table II summarizes the material, installation, and total installed costs of the proposed process.

The belt filter was sized based on a cake thickness of 10 mm and a travel speed of 8 m/min. The resulting filter size was 1 m in width by 10 m in length.

Pilot plant trials on a pulse combustion dryer indicated that this technology could decrease the energy requirements by 30–40%, compared to the energy required in a conventional spray dryer. While capital costs for a pulse combustion dryer and a spray dryer were similar, the reduced size and energy consumption for a pulse combustion dryer should result in lower installed capital costs and lower operating costs.

The capital cost estimate includes a CAD 1,000,000 allowance for improvements to the multiple-effect evaporators to accommodate a 5.8% increase in evaporation requirements. At mills where there is excess evaporator capacity, this outlay would not be required.

The new equipment for the lignin recovery system would be installed in

III. Economic summary for the proposed process

Benefits, CAD/year	
Lignin sales	3,000,000
Lime kiln fuel savings (89% replacement)	800,000
Increase in pulp produced 54 a.d. tons/day	5,500,000
Subtotal	9,300,000
Operating costs, CAD/ year	
CO ₂	1,110,000
H ₂ SO ₄	470,000
Na ₂ CO ₃	1,310,000
Steam	420,000
Natural gas*	100,000
Maintenance	500,000
Labor	700,000
Subtotal	4,610,000
Net benefits, CAD/year	4,690,000
Payback period, years	4.0
Capital cost, CAD	19,000,000

*For drying lignin.

a building roughly 30 m by 35 m. This building would house the reactors, the belt filter, the pulse combustion dryer, and the solids handling equipment. The cost of this new building was included in the estimate.

As shown in Table II, the total installed capital cost for a lignin recovery system with 57 metric tons per day capacity was estimated to be CAD 19,000,000.

Process economics

The economic benefits of kraft lignin recovery are summarized in Table III, together with the anticipated system operating costs. The estimates are based on the chemical costs included in Table I and the assumption that pulp production is limited by the capacity of the recovery boiler. It was taken that 21% of the recovered lignin would be sold on the chemical market, at an average price of CAD 0.75/kg, and that the remainder would be burned in the lime kiln, displacing 89% of the fossil fuel. The pulp equivalent capacity of the recovery boiler would increase by up to 7.7%, allowing an increase in pulp production of 54

tons/day in a 700-a.d. ton/day mill. If pulp production were not increased to the maximum amount, then pulp could be delignified to a greater extent in the digester. Thus, pulp would enter the bleaching stage at a lower kappa number, thereby decreasing chemical usage in the bleach plant.

The total process benefits would be about CAD 9.3 million per year. Operating costs, including chemicals, maintenance, and labor were estimated at about CAD 4.6 million annually, leaving net benefits of about CAD 4.7 million per year. The estimated payback time of four years and net benefits of about CAD 19 per metric ton of pulp make the process economically interesting. Decreasing the sodium carbonate makeup requirements by reducing the H₂SO₄ use for lignin washing would increase net process benefits.

The belt filter washer had a considerable impact on the capital cost for the system. Therefore, work has been initiated with a Canadian manufacturer specializing in washers for the food, pulp and paper, and ore processing industries to optimize the washer design. The capital cost estimate for the washer used in Table II is based on a preliminary cost estimate from that manufacturer.

The revenue from lignin sales has a large impact on the process economics, basically determining the amount of net benefits. Overall economic benefits per kilogram of lignin product, including incremental pulp production, were CAD 1.06/kg for lignin sales compared to only CAD 0.36/kg for the use of the lignin as lime kiln fuel. Therefore, in future work, it is important that potential markets and the lignin selling price be defined as well as possible.

Conclusions

The process of precipitating part of the lignin from kraft pulp mill liquor offers several economic and technical opportunities. The preliminary cost estimates are valuable in identifying areas where continued research might reduce capital or operating costs. One area of opportunity is in washing the lignin. The material balances indicated that all of the liquor from washing the lignin could be accommodated in the mill chemical

cycle, provided that measures are adopted to maintain the chemical balance in the liquor cycle.

In evaluating the process economics, it was assumed that some of the recovered lignin would be sold as a specialty chemical and that the remainder would be burned in the lime kiln, displacing most of the kiln's fossil fuel as normally required. The possibilities of energy self-sufficiency and replacement of petroleum products by a renewable biomass-based fuel are attractive features of the proposed process. In addition, this approach would decrease the heat load on the recovery boiler, permitting more pulp to be produced in mills where the production capacity is limited by the recovery boiler capacity.

The preliminary capital and operating cost estimates developed indicate a pay-back period of about four years, provided that pulp production capacity increases as a result of kraft lignin recovery and that suitable markets can be developed for the lignin that is to be sold. □

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This feasibility study began as a design project for eight students in chemical engineering at the University of British Columbia (1). These students provided a focus for the co-operative effort. They and their faculty advisor, Paul Friedman, brought up issues that may have been missed otherwise. The initial work at UBC on equipment selection and equipment sizing and layout was largely carried through as a basis for this paper. Subsequent to the university work, computers at Sandwell were used to prepare a detailed MASSBAL MARKII process simulation to further develop the liquor cycle process considerations and the capital costs were reassessed. The Prince George laboratory of the Pulp and Paper Institute of Canada also conducted laboratory work on lignin washing.

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