

Using a recovery boiler computer simulation to evaluate process alternatives for obtaining incremental recovery capacity

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Computer models of two kraft recovery furnaces were used to examine the effect of several process modifications on recovery furnace operating conditions, efficiencies, and capacity.

A recovery boiler is one of the most complex and expensive pieces of equipment in a kraft mill. It performs three primary functions: chemical recovery, steam production, and pollution control. Each of these functions imposes operating constraints on the furnace, and satisfying these sometimes opposing parameters further complicates recovery furnace operation. Because recovery boiler capacity cannot be augmented easily, the recovery boiler is often the mill production bottleneck, limiting chemical recovery and, consequently, mill production capacity (1, 2).

Recovery boiler capacity can be limited by plugging of flue-gas passages, by limitations in the boiler's steam generating capacity, or by regulations relating to the emission of odorous reduced sulfur gases (1). Fouling and plugging rates are governed, in large measure, by dust fusion and flue-gas temperatures (3-6). Thus, two of the three most common restrictions on recovery furnace capacity relate directly to the total heat input in the furnace.

Methods of reducing the heat load to the recovery furnace, including kraft lignin recovery (7, 8) and tall-oil soap recovery (9), have been studied as methods to obtain incremental recovery capacity. Other researchers (10-14) have found that higher temperatures in the lower furnace can improve recovery boiler operating efficiency. Techniques such as high-solids firing have been found to increase lower furnace temperatures, recovery furnace reduction efficiency, and recovery boiler capacity (14, 15). It is expensive to directly test the effect of process modifications such as kraft lignin recovery on recovery furnace operations and capacity. Some means of prediction is needed to justify such testing. Sorting out the complex interactions in a recovery furnace is a task for which computer

modeling has been effectively employed (16-18).

In this study, a general energy and material balance system (GEMS) modular recovery furnace simulation package was used for detailed analyses of recovery furnace operations. Validated models of two recovery furnaces were used to examine the effect of process modifications including kraft lignin recovery, oxygen enrichment of combustion air, and flue-gas recirculation or "gas tempering" on recovery-furnace efficiencies and capacity. The magnitude of the incremental gain in recovery furnace capacity resulting from each process modification was assessed, together with any potential operating problems and methods to circumvent the new capacity limitations for each process modification.

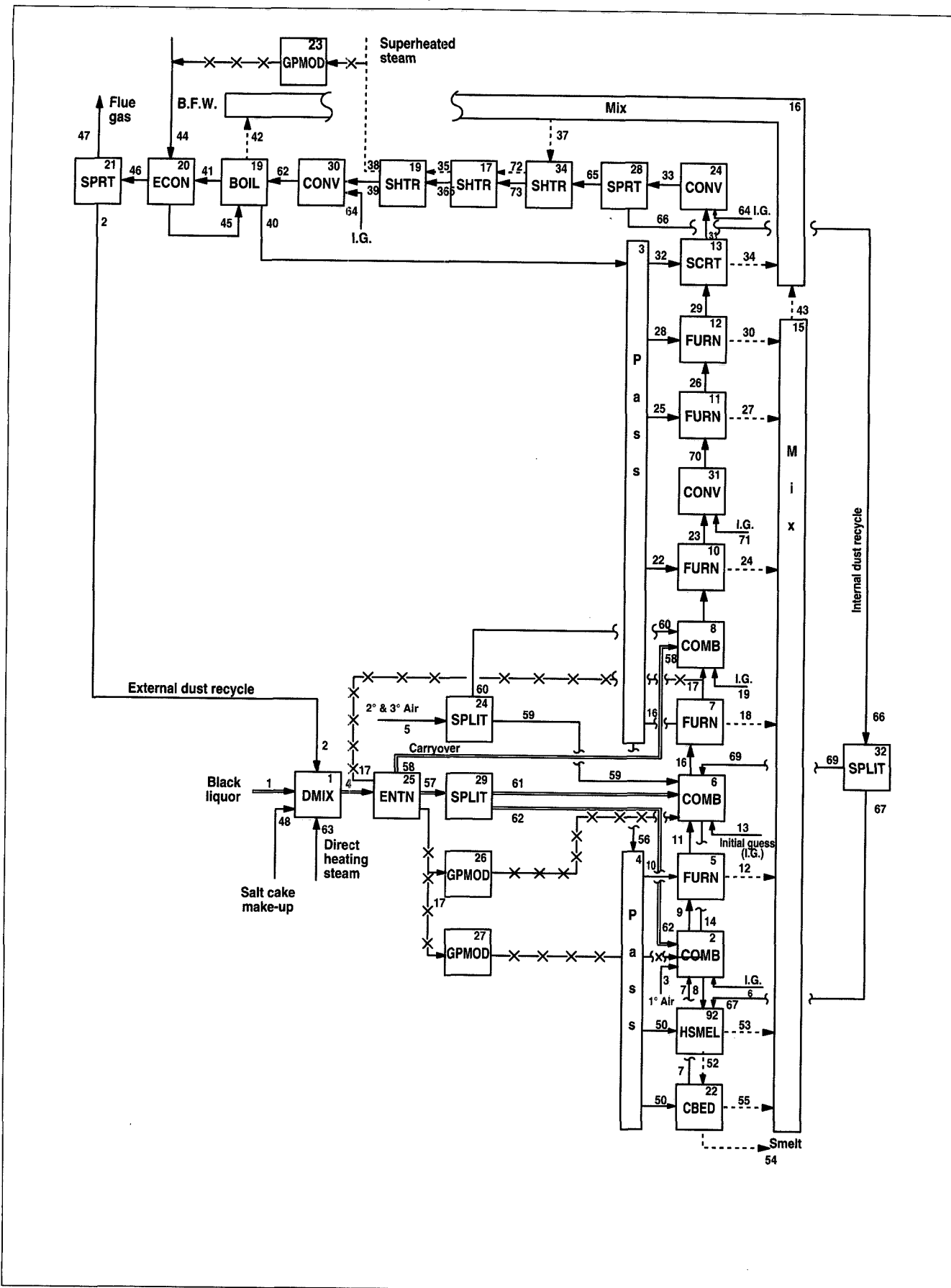
Model description

The recovery boiler model used in these studies was developed at the University of Idaho and has been described previously (3, 16-20). Predictions from the model have been verified against full-scale data (16). Validated models of both a Combustion Engineering (CE) and a Babcock and Wilcox (B&W) recovery boiler were supplied with the GEMS software. These "base case" models were used to evaluate the effects of kraft lignin recovery and of oxygen enrichment of combustion air on recovery furnace capacity. A GEMS computer simulation of a Paprican Member Company recovery furnace was then developed and validated using extensive lower furnace and convection section temperature measurements and pressure drop data taken as part of another project. This model was used to re-examine the effects of kraft lignin recovery, high-solids firing, and gas tempering on recovery boiler operation and capacity.

The GEMS recovery boiler model is a modular program. The block diagram for the GEMS simulation of a Member Company B&W recovery furnace is illustrated in Fig. 1. Each module describes the performance of one piece of equipment or zone of the boiler. These modules can be connected in a variety of ways to model any particular boiler layout. Information concerning flow, concentration, and thermodynamic and chemical state is passed between the modules in a uniform

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1. GEMS block diagram of a Member Company B&W recovery furnace



I. Recovery boiler mechanical design specifications

	University of Idaho CE recovery furnace (3, 16)		Member Company B&W recovery furnace	
Furnace geometry	m	ft	m	ft
Firebox width	7.3	23.9	8.5	28.0
Firebox depth	7.6	24.9	8.4	27.5
Liquor gun elevation	5.7	18.8	7.7	25.25
Bullnose elevation	24.2	79.5	27.6	90.5
Roof elevation	34.8	114.0	40.5	132.75
Heat transfer area	m ²	ft ²	m ²	ft ²
Furnace walls	1,232	13,250	1,251	13,565
Screen tubes	1,060	11,400	83.8	902
Superheater, total	969	10,430	2,471	26,584
Boiler bank	1,919	20,640	3,265	35,130
Economizer	3,889	41,860	4,922	52,955
Furnace volume	55,000 ft ³		84,000 ft ³	

II. Effect of kraft lignin recovery on black liquor composition and heating value in the CE recovery furnace simulation

	Weight %		
	Original liquor	With 10% of the total lignin removed	With 15% of the total lignin removed
Elemental analysis			
Carbon	35.3	34.2	33.6
Hydrogen	3.7	3.6	3.6
Oxygen	35.8	36.1	36.3
Sodium	20.5	21.3	21.7
Sulfur	4.7	4.8	4.8
Heating value			
MJ/kg dry solids	14.63	13.46	12.87
Btu/lb dry solids	6290	5787	5525
Weight % dry liquor solids in feed liquor	70.0	70.0	70.0

format. This allows the individual modules to be tested and validated before the total model is used to predict recovery boiler performance. The primary, secondary, and tertiary air regions of the furnace are treated as separate zones, each having a combustion module and a heat transfer module. In the upper furnace, screen tubes, superheater, boiler bank, and economizer are all separate modules. All heat transfer modules take into account both radiation and convection. Radiation properties are based on local gas composition. Convective heat transfer coefficients are based on appropriate correlations for tube banks and on local gas properties (17). Plug flow is assumed between the reactor modules. Complete details of the model assumptions and physical and mathematical basis can be found elsewhere (20).

Combustion and heat release are distributed among the two or three combustion levels, depending on the particular furnace, according to the amount of air added to each level, the amount of liquor delivered to the level, and the prescribed degree of mixing. Passage of liquor to the bed or upper furnace depends on prescribed entrainment, mixing, and combustion parameters. Adiabatic free-energy minimization calculations are used in each combustion stage on the mixed portion of the gaseous products, and in the upper furnace below the convection section tube banks, to estimate equilibrium gas and solid species concentrations. The unmixed portion bypasses these equilibrium calculations.

Carbon gasification and smelt reduction occur in the char bed. Temperature-dependent kinetic corrections are applied for smelt reduction and sulfur emission (20). The effects of operating conditions on liquor droplet size and carry-over are predicted, relative to a base case, using correlations for mean droplet diameter (21) and a simple force balance between gravity and drag. Fuming chemicals generated in the lower furnace are converted to dust by equilibrium calculations; the fractional conversion of SO₂ and H₂S into dust can be specified by the user to account for incomplete gas mixing or kinetic limitations.

Despite the simplification of breaking the furnace into a relatively few modules, there are still a number of parameters which must be specified. The technical and computational difficulties in specifying liquor droplet flight and air jet trajectory (22) are, however, avoided. Entrainment, flow patterns, degree of mixing, etc. must, however, be prescribed. Fortunately, the model is very robust and heat transfer results, which are the main focus here, are not very sensitive to the values selected (17). This is not true for chemical aspects of recovery boiler performance such as smelt reduction, flue-gas CO, total reduced sulfur (TRS), and SO₂ concentrations. For this reason, the process alternatives examined were run on both validated base-case models supplied with the software and the Member Company recovery boiler simulation. Data from only one of the supplied simulations will be presented in this paper since results from these two models agreed closely.

Boiler models and process modification simulation

The first simulation used in these studies was adapted from a model supplied and validated (3, 16) by researchers at the University of Idaho. This validated base case corresponded to a low-odor CE furnace with two levels of combustion air and an

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extended economizer. Mechanical design specifications for the boiler are summarized in **Table I**. We increased liquor flows for the given base case until further flow increases resulted in TRS emissions exceeding 6 ppm, a typical Canadian regulatory limit. Our base-case furnace produced 188 metric tons (413,600 lb) of superheated steam per hour at 3160 kPa (450 psig) and 292°C (558°F) by firing 1164 metric tons (2.56 million lb) of black liquor solids per day. Black liquor characteristics in the base case and kraft lignin recovery simulations are summarized in **Table II**. In this lignin recovery simulation, pine lignin with a typical molecular formula (23) of $C_9H_{7.9}O_{2.1}S_{0.1}(OCH_3)_{0.83}$ was assumed to comprise 40% of total liquor solids. Analysis of commercial kraft lignins and lignins precipitated in our Prince George laboratory from interior British Columbia kraft liquors indicated lignin elemental composition was typically very close to that derived using this formula, once entrained and chemically bound sodium (1.5% by weight) and sulfur (2.0% by weight) were included. The effect of lignin recovery on feed liquor composition and liquor heating value was calculated using these analyses and laboratory data (?). With 10% and 15% lignin recovery, black liquor flow rates in the simulations were increased by up to 10% and 15%, respectively, to compensate for the reduced liquor calorific value.

In simulating oxygen enrichment of combustion air, total combustion air flow was held constant at 210 metric tons/h (462,000 lb/h) while oxygen injection was used to increase the oxygen concentration in the combustion air from 20.9% to 22.0% and 23.0% by weight. Air distribution (85% to the primary air ports) was not modified. The effects of O_2 injection on furnace operating conditions and temperature profiles were determined as liquor flow rates were gradually increased.

A GEMS model of a Member Company recovery furnace, illustrated in Fig. 1, was also developed. This model corresponded to a low-odor B&W furnace with three levels of combustion air and a triple-pass cross-flow economizer. This furnace was rated at 1090 metric tons (2.4 million lb) of black liquor solids per day but typically operated about 10% over design capacity. The model was tuned using extensive lower-furnace and convection-section temperature measurements, steam flow, and reduction efficiency data taken as part of a second project. The model was then validated using gas and steam temperature, steam production, and furnace draft loss data taken under a variety of operating conditions, for example, at different firing rates with various levels of fouling in the convection-section banks.

Base models for a "clean" furnace and a "moderately fouled" furnace were developed. Actual operating data and model predictions for the "moderately fouled" furnace used in the simulations reported in this paper are compared in **Table III**.

A great deal of difficulty was encountered in modeling heat transfer in the economizer. When deposit thickness was increased to give exit gas temperatures comparable to those measured in the furnace, draft losses were very high (43 in. of water pressure versus the 5 in. actually measured). Since we were interested primarily in gas temperatures through the convection section and their potential effect on plugging, we used this model to examine the effects of kraft lignin recovery.

In the flue-gas recirculation or "gas tempering" simulations, draft loss was expected to be more restrictive than gas temperatures in defining the incremental throughput pos-

III. Comparison of measured parameters and temperatures with input and output data for the simulated recovery furnace

	Averaged measured data	Simulation data	
		Model for lignin recovery studies	Model for air tempering studies
Liquor firing rate, metric tons/h (wet)	75.0	75.0	75.0
Million lb of dry solids/day	2.78	2.78	2.78
Combustion air flow, metric tons/h	214.0	214.0	214.0
Air distribution 1°/2°/3°, %	36/48/16	36/48/16	36/48/16
Gas temperatures, °C			
Just above the bed	883	883	883
At the secondary air level	931	931	931
At the tertiary air level	879	879	879
At the inlet to the screen tubes	852	852	852
At the inlet to the boiler bank	625	600	603
At the economizer inlet	435	378	379
At the economizer outlet	225	223	189
Water/steam side temperatures, °C			
Into the economizer	140	140	140
Into the generating bank	233	195	206
Into the screen tubes	255	255	255
Into the primary superheater	300	289	284
Into the secondary superheater	359	360	355
Product steam			
Temperature, °C	378	379	373
Pressure, 100 kPa (or atmosphere)	42.9	42.9	42.9
Flow, metric tons/h	206	198	203
Total draft loss,			
in. H_2O	5.0	43.1	4.91
Flue -gas O_2 , %	2.5	2.5	2.5
Flue -gas H_2S , ppm	0	0.08	0.08
Flue -gas SO_2 , ppm	0 (<20)	2.4	2.4
Reduction efficiency	96.1	95.1	95.1

IV. Effect of kraft lignin recovery on black liquor composition and heating value in the B&W recovery furnace simulation

	Weight %	
	Original liquor	With 10% of the total lignin removed
Elemental analysis		
Carbon	33.0	32.4
Hydrogen	3.65	3.6
Oxygen	39.3	39.7
Sodium	19.6	20.1
Sulfur	4.15	4.2
Heating value		
MJ/kg dry solids	14.36	13.22
Btu/lb dry solids	6175	5682
Weight % dry liquor solids in feed liquor	70.1	70.1

V. The effect of oxygen enrichment and increased liquor throughput on recovery furnace operating parameters and temperature profiles (GEMS simulation of a CE recovery boiler)

Oxygen in combustion air, %	20.9	22.0	22.0	22.0
Total oxygen feed rate, kg O ₂ /kg black liquor solids	0.905	0.952	0.892	0.886
% Increase in liquor throughput	0	0	6.8	7.5
Gas temperatures, °C				
Just above the char bed	1073	1123	1115	1114
At the secondary air level	1440	1403	1466	1471
At the inlet to the screen tubes	889	883	898	898
At the inlet to the superheater	718	726	739	735
At the inlet to the boiler bank	620	627	636	633
At the economizer outlet	186	187	188	188
Composition of flue gas out of the economizer				
H ₂ S, ppm	0	0	0	10
O ₂ , % by weight on a wet basis	1.9	2.9	1.7	1.6
Total flue-gas flow to the screen tubes, metric tons/h	260.3	261.2	264.7	265.1
Dust content, kg/metric ton of wet gas	7.5	10.8	11.4	11.5
Smelt				
Flow, metric tons/h	21.8	21.6	23.0	23.1
Na ₂ SO ₄ , kg/metric ton of smelt	22.0	21.8	16.3	15.9
Reduction efficiency, %	94.7	94.7	95.8	95.9
Superheated steam flow, metric tons/h				
	188.0	186.0	200.8	201.6
Superheated steam temp., °C				
	291.7	294.1	292.9	292.0

sible, and a second simulation, also summarized in Table III, was used. In this second simulation, draft losses were close to those measured in the furnace, while gas temperatures exiting the economizer were much lower than those measured.

The difficulties observed in modeling this triple-pass cross-flow economizer may simply reflect a limitation of the current GEMS economizer models; only simple countercurrent, one-pass cross-flow or multipass longitudinal economizers are currently modeled. We used the one-pass cross-flow economizer model, although a triple-pass economizer with higher gas velocities throughout would be expected to have higher heat transfer coefficients and higher draft loss. The lack of agreement between mill measurements around the economizer and the computer model might also result from errors in the heat transfer and pressure drop correlations used in the GEMS model. The data from our studies have, thus, been forwarded to the University of Idaho to be used to check the validity of the existing cross-flow economizer block or to develop a triple-pass cross-flow economizer model.

In the kraft lignin recovery simulations, calculations similar

to those outlined above were used to estimate the effect of kraft lignin recovery on liquor composition and heating value. Analyses indicated that lignin comprised about 32% of the total liquor solids in the original liquor, and this value was used as a basis for the calculations. The calculated effect of 10% lignin recovery on black liquor composition and heating value is summarized in Table IV.

Flue-gas temperatures entering the generating bank of a recovery furnace are thought by many to be critical in preventing plugging (1, 3-6, 11-18) because of their effect on dust melting and stickiness and because of the narrower tube spacing in this bank relative to upstream convection section banks. The gas temperature entering the generating bank can be reduced by recycling part of the stack gas. In the utility industry, flue-gas recycling or "tempering gas" is commonly used to control reheat and superheat temperatures. To simulate "gas tempering" in the Member Company B&W furnace, a split block was inserted in the GEMS block diagram. Twenty percent or 40% of the gas exiting the precipitator (stream 47 in Fig. 1) was returned to a mix block inserted just below the

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screen tubes (block 13 in Fig. 1) to simulate perfect mixing of cool tempering gas and hot flue gases. The effect of this practice on gas temperatures, furnace draft loss, and flue-gas velocities at the generating bank inlet was then examined as the liquor flow rate was increased.

Results and discussion

Oxygen enrichment of combustion air

Recovery furnace throughput is most often limited by temperature-related plugging problems in the convection sections or by TRS emissions exceeding regulatory limits (1, 2). Oxygen enrichment of the combustion air fed to a recovery furnace has been proposed as a method to eliminate the furnace bottleneck. Oxygen enrichment would be expected to increase lower furnace temperatures and radiative heat transfer to the lower furnace walls. Gas temperatures entering the convection sections should therefore be lowered. Plugging problems in those sections would thus be expected to be reduced, while the increased oxygen mass flow rate would ensure oxidation of the odorous reduced sulfur gases and permit increased solids throughput.

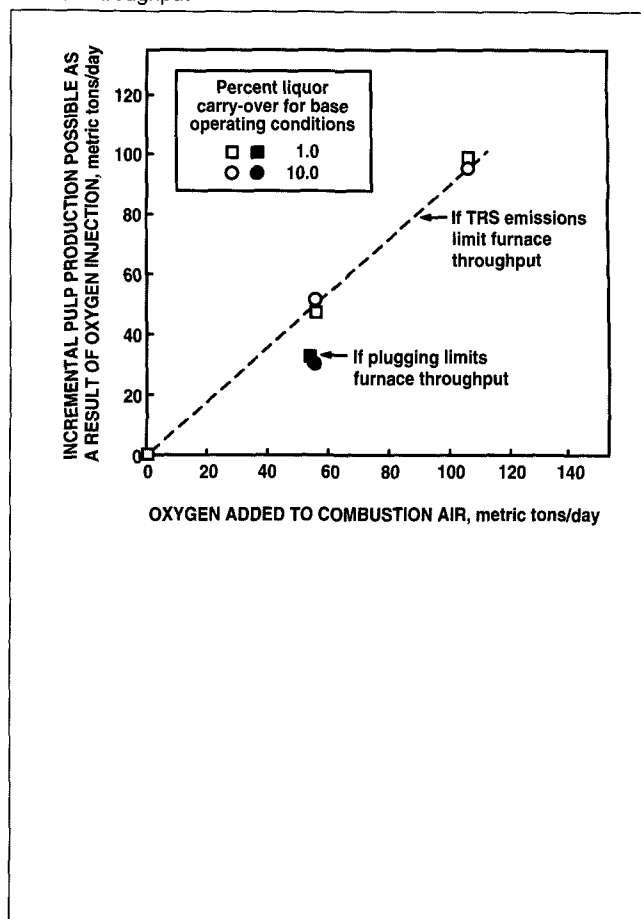
Typical results from the GEMS simulations of oxygen enrichment of the combustion air in a CE recovery furnace are summarized in Table V. With oxygen injection and no increase in the liquor firing rate (compare columns 1 and 2), bed temperature increased significantly, and the gas temperature at the inlet to the screen tubes decreased slightly. The higher bed temperatures resulted in increased sodium fuming, however, and increased dust load at the screen tube inlet by as much as 40%. Increased dust load due to higher bed temperatures has been corroborated by mill trials of "high-solids combustion" (24).

Conversion of the greater amounts of sodium fume to Na_2CO_3 and Na_2SO_4 (in a calculation module following the screen tubes in this particular simulation) resulted in gas temperatures, at both the superheater and boiler bank inlet, which were slightly higher than those in the base case. If complete conversion of sodium fume occurred lower in the furnace, gas temperatures throughout the convection sections would be very close to those in the base case. In GEMS simulations of a B&W recovery furnace, Haynes *et al.* (17) similarly found that large changes in the excess air flow, air distribution, or flow patterns had significant effects on lower furnace temperatures but little effect on the gas temperatures at the superheater and boiler bank inlets. Gas temperatures at the boiler bank changed by only 12°C for a wide range of air flows and air distributions.

As the liquor flow rate was increased in conjunction with oxygen enrichment (columns 3 and 4 in Table V), lower furnace temperatures increased further, and a slight increase in reduction efficiency was predicted. The higher temperatures in the lower furnace again resulted in increased fuming and increased flue-gas dust content. Conversion of the extra sodium fume and liquor burning in the upper furnace increased gas temperatures. Gas temperatures in the superheater and boiler bank were increased by 15–20°C compared to the base case.

While increased lower furnace temperatures resulted in a

2. Effect of oxygen enrichment of the combustion air on recovery furnace throughput



dust with higher carbonate content (and consequently a higher melting point), if temperature-related plugging problems in the superheater or boiler bank limit furnace throughput, these test results suggest that oxygen injection would neither alleviate the plugging problems nor permit large increases in furnace throughput. Assuming that a 5–6°C increase in the gas temperature at the inlet to the superheater or boiler bank could be tolerated in a furnace limited by plugging problems (as a result of higher dust carbonate content and higher dust melting points), further simulations predicted that furnace throughput could only be increased by about 4% as a result of oxygen injection (Fig. 2). Lowering the amount of liquor entrained in the combustion gases from 10% to 1% of the input in each base case had little effect on the relative gas temperature profiles through the convection sections (Fig. 2). In simulating oxygen enrichment of the combustion air supply in a B&W recovery furnace, Haynes *et al.* (17) similarly found the effect in the upper furnace was fairly mild and might permit an increase of about 5% in the liquor firing rate in a recovery furnace limited by plugging problems.

If sulfur gas (TRS) emissions restrict furnace throughput, the results summarized in Table V and Fig. 2 suggest that oxygen enrichment could increase throughput by about 1 metric ton of pulp per metric ton of oxygen supplied. However, steam production increases as the liquor firing rate is increased; a steam-side circulation study should be undertaken

VI. Effect of lignin removal and increased combustion air flow on recovery furnace capacity and operating conditions (computer simulation of a CE recovery boiler)

Case	Base	1	2	3	4
% Lignin recovery	0	10	10	15	15
% Increase in liquor flow rate	0	10	10	15	15
% Increase in air flow rate	0	0	4.3	0	7.6
Gas temperatures, °C					
Just above the char bed	1076	1014	1042	975	1028
At the secondary air level	1416	1365	1357	1287	1308
At the inlet to the screen tubes	883	841	870	811	865
At the inlet to the superheater	712	670	709	656	711
At the inlet to the boiler bank	612	585	618	576	621
At the economizer outlet	185	185	189	186	191
Composition of flue gas out of the economizer					
H ₂ S, ppm	0	1790	0	2250	0
O ₂ , % by volume on a wet basis	2.7	1.9	2.2	1.9	2.2
Total flue-gas flow to the screen tubes, metric tons/h	259	262	272	263	280
Dust content, kg/metric ton of gas	7.3	5.9	7.7	4.9	8.4
Smelt					
Flow, metric tons/h	21.3	24.8	24.5	26.8	26.2
Na ₂ SO ₄ content, kg/metric ton smelt	21.5	59.7	43.6	84.9	57.2
Reduction efficiency, %	94.9	83.3	88.8	73.0	85.0
Superheated steam flow, metric tons/h	182.6	175.8	180.5	164.2	177.1
Superheated steam temperature, °C	292.1	287.5	293.0	289.1	249.9

to establish how high steam production might be safely increased. Precipitator capacity would also need to be assessed to determine if the much higher dust load could be effectively handled.

Kraft lignin recovery

In earlier studies (7, 8), we found that removal of 10% of the lignin from black liquor for sale or separate combustion could reduce the calorific value of that liquor by 8%. Removal of 10% to 15% of the total lignin had no effect on the swelling properties of the liquor solids—an indication that combustibility would not be impaired. Assuming pulp production increased by 8–10% as a result of the reduction in liquor calorific value (8, 26), economic analyses for three proposed lignin recovery processes indicated payback times between 1 and 4 years (8, 25, 26), if the recovered lignin was burned in a lime kiln or hog-fuel boiler. More recent trials and analyses indicate that lignin might be used as a slurry (27) or dry powder (28) to fire a lime kiln and that lignin has some advantages as a fuel compared to natural gas (28). The effect of lignin removal on recovery furnace temperature profiles, operating conditions, and capacity was thus examined using simulations of two recovery furnaces.

Typical results from the GEMS simulations of the CE recovery furnace are summarized in Table VI. With lignin

recovery and increased liquor flow rates, but with no other changes in furnace operating conditions (cases 1 and 3 versus the base case), H₂S emissions increased dramatically. Steam production and furnace reduction efficiency decreased. These adverse effects can be traced directly to reduced operating temperatures in the lower furnace (10, 11).

To increase the lower furnace operating temperatures, combustion air flow was increased stepwise until TRS emissions were brought under control (cases 2 and 4 versus the base case in Table VI). Increased combustion air brought the furnace temperature profiles closer to the base case and increased the furnace reduction efficiency and steam production. With 10–15% lignin recovery, 10–15% increases in recovery furnace capacity appear feasible if the combustion air flow can be increased by 4–8%. With 10% and 15% lignin recovery and no increase in combustion air flow, liquor firing rates could be increased by 6% and 8%, respectively, before TRS emissions exceeded typical regulatory limits (6 ppm).

The dust load entering the convection section in the lignin recovery simulations in Table VI increased by only 5–15%, even with the increased combustion air flows. Dust deposition is more directly related to the dust fusion and flue-gas temperatures than to the quantity of dust emitted (2–6). Predicted flue-gas temperatures entering the superheater were lower for the modified operating conditions than for the base case. Temperatures entering the boiler bank were only marginally

VII. Effect of lignin removal on recovery furnace capacity and operating conditions (computer simulation of a "moderately-fouled" B&W recovery boiler)

Case	Base	1	2	3	4
% Lignin recovery	0	0	10	10	10
% Increase in liquor flow rate	0	6	10	6	6
Solids content of fired liquor, wt%	70.1	75.0	70.1	70.1	75.0
Liquor firing rate, % of design	116	123	127	123	123
Gas temperatures, °					
Just above the char bed	883	873	854	864	872
At the secondary air level	931	944	852	865	894
At the tertiary air level	879	887	764	796	830
At the inlet to the screen tubes	852	854	793	814	826
At the inlet to the superheater	735	812	766	764	766
At the inlet to the boiler bank	600	659	656	623	620
At the economizer outlet	223	227	235	229	224
Composition of flue gas out of the economizer					
H ₂ S, ppm	0.07	0.11	623	0.18	0.14
O ₂ , % by volume on a wet basis	2.51	1.57	1.53	1.91	2.06
Total flue-gas flow to the screen tubes, metric tons/h	268.2	266.4	271.8	270.1	265.3
Dust content, kg/metric ton of gas	13.9	15.1	10.3	10.2	12.7
Smelt					
Flow, metric tons/h	268.2	266.4	271.8	270.1	23.6
Na ₂ SO ₄ content, kg/metric ton smelt	6.6	6.8	26.2	20.5	14.1
Reduction efficiency, %	95.1	95.0	80.2	84.8	89.6
Superheated steam flow, metric tons/h	197.8	210.1	186.4	184.6	189.6
Superheated steam temperature, °C	379	401	406	403	396.5

higher. Therefore, the additional dust load should be readily handled by existing sootblowers, although sootblowing steam consumption may be increased slightly.

With kraft lignin recovery and increased liquor firing rates, the reduction efficiency was lower than that in the base case. While cooking chemical production would still be adequate to support the additional pulping capacity desired, the increased liquor deadload (Na₂SO₄) would greatly affect liquor composition and resulting operating conditions, particularly evaporator scaling and recovery furnace operations. While many mills operate with reduction efficiencies in the mid-'80s, other process modifications were examined, using the Member Company recovery furnace simulation, in an attempt to bring the reduction efficiency closer to that in the base case.

High-solids firing

The results of simulations for the "moderately fouled" B&W

recovery furnace are summarized in **Table VII**. In the base case, the furnace was already running 16% over design capacity. Further increases in the liquor firing rate were predicted to result in high TRS emissions. When the liquor solids content was increased from 70% to 75%, lower furnace temperatures increased slightly, sulfur retention in the bed improved, and throughput could be increased by up to 6% (case 1) before TRS emissions began to exceed regulatory limits. With high-solids firing and increased throughput, however, gas temperatures into the superheater and boiler bank exceeded those in the base case by 60–80°C and flue-gas dust load increased by 10%, indicating a strong possibility of accelerated furnace plugging. Furthermore, superheated steam production increased by 6%, possibly exceeding boiler circulation rates. The increases in furnace throughput dust load and steam production predicted to result from high-solids firing are similar to those observed when high-solids firing has been

VIII. Effect of tempering gas flow on recovery furnace capacity and operating conditions (computer simulation of a "moderately-fouled" B&W recovery furnace)

Case	Base	1	2	3	4
% Stack gas recycle	0	40	20	20	0
% Increase in liquor flow rate	0	0	0	10	-14
% Increase in air flow rate	0	0	0	10	-14
Gas temperature, °C					
Just above the char bed	883	883	883	894	866
At the secondary air level	931	931	931	943	911
At the tertiary air level	879	879	879	896	852
At the inlet to the screen tubes	852	852	852	880	806
At the inlet to the superheater	742	623	680	706	698
At the inlet to the boiler bank	603	548	576	600	563
At the inlet to the economizer	379	402	391	410	352
At the outlet to the economizer	189	224	204	212	180
Composition of flue gas out of the economizer					
TRS, ppm	0.08	0.05	0.06	0.06	0.08
O ₂ , % by volume	2.51	2.50	2.50	2.51	2.51
Total flue gas flow to screen tubes, metric tons/h					
	268	448	336	369	231
Dust content, kg/metric ton					
	13.9	8.5	11.1	11.3	13.7
Dust flow, kg/h					
	3725	3785	3715	4150	3165
Smelt					
Flow, metric tons/h	21.5	21.5	21.5	23.6	18.5
Reduction efficiency, %	95.1	95.1	95.1	95.2	94.9
Superheated steam flow, metric tons/h					
	203	208	204	222	177
Superheated steam temperature, °C					
	373	349	361	364	367
Total pressure drop across convection section					
Tube banks, in. water	4.91	12.8	7.58	9.23	3.55
Gas velocity at the inlet to the generating bank, m/s					
	7.91	12.31	9.70	10.94	6.50

implemented (24, 29).

High-solids firing and kraft lignin recovery

With 10% lignin recovery and liquor firing rates increased by 10% at the original liquor solids concentration (Table VIII, case 2), TRS emissions again exceeded regulatory limits. With 10% lignin recovery and liquor firing rates increased by 6% (case 3), TRS emissions remained low, as observed in the CE recovery furnace simulation. Flue-gas dust load dropped by 30% compared to the base case and gas temperatures at the superheater and boiler bank inlet increased by 25–30°C compared to the base case. However, lower furnace temperatures were again substantially lower than those in the base case, and the furnace reduction efficiency fell from 95% to 85%. When the fired liquor solids content was then increased from 70% to 75% (case 4), the reduction efficiency was increased to a more tolerable 90%.

Dust loading remained 10% below that in the base case, while gas temperatures at the superheater and boiler-bank inlet were increased by 25–30°C over those in the base case. Steam production remained 4% below that in the base case, although steam temperature was substantially increased. Increases in combustion air flow (not shown in Table VII), were found to further increase the reduction efficiency, the flue-gas dust load, steam production, etc. in a manner similar to that shown in Table VI for the CE furnace. With 10% kraft lignin recovery, the liquor solids increased from 70% to 75% and combustion air flow increased by 4%. The reduction efficiency, dust load, and steam production were all very similar to those in the base case; gas temperatures into the superheater and boiler bank were, however, 20–35°C higher than in the base case, indicating that plugging would probably determine the incremental throughput possible in this furnace as a result of kraft lignin recovery. Increased tube cleanliness (i.e., more effective sootblowing) upstream of the generating bank

Kraft Recovery

could decrease gas temperatures entering the boiler bank (17), permitting greater capacity gains.

Tempering gas

Tempering gas is used in the utility industry to control reheat and superheat temperature. Stack gas is recycled and injected into the furnace in the bullnose area. The cooler tempering gas dilutes the furnace combustion gases and reduces the gas temperature at the inlet to both the superheater and the boiler bank. Earlier recovery boiler simulation studies (17) indicate that recycling 40% of the stack gas as tempering gas can reduce the gas temperature at the generating bank inlet as much as reducing the liquor firing rate by 14%. Rapid gas cooling might also change the size, surface morphology, and shape of crystals formed from fume and the sintering tendency of resulting particles (30), thereby further reducing furnace plugging problems.

Our simulations of the "moderately fouled" B&W recovery furnace were used to examine the effect of gas tempering on furnace operating conditions and capacity. Typical results with the model developed for these gas tempering simulations (i.e., with correct furnace draft losses but with economizer exit gas temperatures much lower than those actually measured, as summarized in Table III), are summarized in Table VIII.

With 40% stack gas recycle (case 1) the temperature at the inlet to the superheater and the generating bank decreased, as expected, by 119°C and 55°C, respectively. This resulted in a gas velocity at the generating bank inlet of 12.3 m/s, which could result in dust deposition by impaction (31) and very high plugging rates despite the lower gas temperature.

With 20% stack gas recycle (case 2), gas velocity at the generating bank inlet was 9.7 m/s, and a 27°C drop in the gas temperature at the generating bank inlet was predicted. This allowed us to increase liquor and combustion air flow rates by 10% (case 3), resulting in a gas velocity at the boiler bank inlet of 10.9 m/s and an overall furnace draft loss of 23.4 cm (9.2 in.) of water. Since high-efficiency induced-draft fans can be used to pull air with pressure losses through the convection section tube banks of up to 50–64 cm (20–25 in.) of water (31), draft loss is not expected to limit the incremental throughput possible as a result of gas tempering. Gas velocity at the inlet to the generating bank is likely to be more restrictive since gas velocities at the generating bank inlet should not exceed 10–12 m/s if deposition by impaction is to be avoided (31).

With 20% stack gas recycle and a 10% increase in the air and liquor flow rates (case 3), the gas temperature at the generating bank inlet was slightly lower than in the base case. When the model with correct economizer exit gas temperatures was used to simulate 20% stack gas recycling, liquor and combustion air flow rates could be increased by only 7% before the gas temperature at the generating bank inlet equaled that in the base case. This result is similar to that reported by Haynes *et al.* (17) and suggests that gas temperature and velocity at the generating bank inlet are likely to jointly define the incremental throughput possible in a given furnace as a result of gas tempering.

Ten percent higher liquor firing rates resulted in an 11.4% increase in the dust load to the screens and a 10% increase in

steam production. The capacity of the steam circuit and of the precipitators to handle these increases would also need to be verified.

Tempering was also observed to reduce superheated steam temperature from 373°C to 364°C. Lower steam temperature would result in less electricity production in a turbine generator. A 9°C decrease in average steam temperature in this furnace would reduce electricity production by 0.6 MW per day, increasing purchased electricity costs by about US\$ 200,000 per year (assuming electricity costs US\$ 0.04/kw h).

Conclusions

Computer models of two kraft recovery furnaces were used to examine the effect of several process modifications on recovery furnace operating conditions, efficiencies, and capacity. With 10% kraft lignin recovery, simulation results indicate that liquor firing rates could be increased by 6% before TRS emissions exceeded regulatory limits. If combustion air flow could be increased by 4%, liquor firing rates could be increased by 10% before gas temperatures at the superheater and generating bank inlet were equal to those in the base case. The dust load entering the convection sections would only be increased by about 5% as a result of these changes, but furnace reduction efficiency could decrease from 95% to 89%. High-solids firing in conjunction with kraft lignin recovery could increase the reduction efficiency to 95%, but gas temperatures into the superheater and boiler bank would increase by 20–35°C relative to the base case, indicating that furnace plugging could limit the incremental throughput possible as a result of kraft lignin recovery.

Simulation results indicate that oxygen enrichment of the combustion air fed to a recovery furnace could increase recovery throughput by about 1 metric ton of pulp per metric ton of oxygen supplied, if sulfur gas emissions currently limit production. Oxygen enrichment would not, however, significantly reduce gas temperatures entering the boiler convection sections or fouling and plugging problems. In a furnace limited by plugging problems, oxygen injection might permit a 4–5% increase in throughput with corresponding increases in steam production. Dust loading would, however, be increased by about 30% as a result of increased lower furnace temperatures, and precipitator capacity would also need to be assessed and perhaps increased.

Gas velocity at the generating bank inlet was found to limit recovery furnace capacity when stack gas recycle or gas tempering was simulated. With 20% flue-gas recycle, liquor firing rates could be increased by about 7% before gas temperatures at the boiler bank inlet exceeded those in the base case; furnace draft losses (23 cm or 9 in. of water) were not excessive at this level of recycle and flue-gas velocity at the generating bank inlet was 11 m/s, midway into the 10–12 m/s range where dust deposition by impaction and consequent rapid plugging would become a concern. Gas tempering did reduce product steam temperatures, however, and could reduce electrical production in cogeneration facilities. **■**

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