ABSTRACT

The kraft process dominates pulp and paper production worldwide. Black liquor, a mixture of lignin and inorganic chemicals, is generated in this process as fiber is extracted from wood. At most kraft mills today, black liquor is burned in Tomlinson boilers to produce steam for on-site heat and power and to recover the inorganic chemicals for reuse in the process. Globally, the black liquor generation rate is about 85,000 MW (or 0.5 million tonnes of dry solids per day), with nearly 50% of this in North America. The majority of presently-installed Tomlinson boilers will reach the end of their useful lives during the next 15 to 20 years. As a replacement for Tomlinson-based cogeneration, black liquor gasifier/gas turbine cogeneration promises higher electrical efficiency, with prospective environmental, safety, and capital cost benefits for kraft mills. Several companies are pursuing commercialization of black liquor gasification for gas turbine applications. This paper presents results of detailed performance modeling of gasifier/gas turbine cogeneration systems using different black liquor gasifiers modeled on proposed commercial designs.

INTRODUCTION

Black liquor is the lignin-rich byproduct of fiber extraction from wood in kraft pulp production. In 1994, the U.S. pulp and paper industry consumed 1.2 EJ (10^18 J), or 38,000 MW, of black liquor. This exceeded the 1.0 EJ of total fossil fuel used by the industry (AFPA, 1996). The industry burns black liquor today in Tomlinson recovery boilers that feed back-pressure steam turbine cogeneration systems supplying process steam and electricity to mills. A critical second objective of Tomlinson boilers is the recovery of pulping chemicals (sodium and sulfur compounds) from the black liquor for reuse (Adams et al., 1997). Technologies for gasifying black liquor are under development to replace Tomlinson boilers. This paper assesses the prospective energy performance of black liquor gasifier/gas turbine technologies that are targeted for commercial application by early in the next decade.

This work is motivated by the interest of the pulp and paper industry in black liquor gasification technology arising from prospective improvements in energy, environmental, safety, and capital-investment characteristics compared to Tomlinson technology (Larson et al., 1996; Industria, 1996). The industry sees some urgency in realizing the promise of gasification technology because its fleet of Tomlinson boilers is aging. Some 80% of all currently-operating recovery boilers in the U.S. were built or rebuilt before 1980 (Fig. 1). Typical recovery boiler lifetimes are around 30 years (Weyerhaeuser et al., 1995), so most of the recovery boilers in the U.S. (and Canada as well) will need major attention or replacement over the next 15 to 20 years. As the most capital-intensive of U.S. manufacturing industries, the paper industry is keenly interested in minimizing capital expenditures for replacements, but would like to do so while simultaneously meeting increasingly stringent environmental regulations, improving energy performance, and reducing the risk of major recovery boiler explosions that occur periodically (Grant, 1996). The need to replace Tomlinson boilers over the next two decades provides a once-in-thirty-years window of opportunity for the industry to make a major technology change (to gasification) to address these concerns.

Black liquor gasifier development received considerable attention in the U.S. in the early 1980s (Keller, 1985; Empie, 1991), but the prospects for commercializing technology appear considerably improved at present. In fact, the first fully-commercial black liquor gasifier (Smith et al., 1996) has begun startup testing at Weyerhaeuser's New Bern mill in North Carolina. This unit will provide the mill with incremental capacity for chemicals recovery from black liquor, with the gasifier product gas being burned in a boiler. For the longer-term, there is industry interest in full replacement of Tomlinson-based cogeneration with gas turbine-based systems.

Black liquor gasification technologies under development can be classified according to operating temperature, or equivalently, according to the physical state in which the majority of the inorganic content of the feedstock is recovered. High-temperature gasifiers operate at 900°C or higher and produce a molten slurry of inorganic chemicals. Low-temperature gasifiers operate at 700°C or lower in order to insulate that the inorganics leave as dry solids. Kvaerner and Noell are two companies developing high-temperature gasifiers for gas turbine applications. ABB Beglin and Noell are two companies developing high-temperature gasifiers for gas turbine applications. ABB Beglin and Noell are two companies developing high-temperature gasifiers for gas turbine applications.

Stefano Consonni
Departamento di Energia
Politecnico di Milano
20133 Milano, Italy

Eric D. Larson
Center for Energy and Environmental Studies
School of Engineering and Applied Science
Princeton University, Princeton, NJ 08544

Niklas Berglin
Department of Heat and Power Technology
Chalmers University of Technology
412 96 Goteborg, Sweden

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and MTCI are developing low-temperature gasifiers for such applications. Pressurized operation (at about 25 bar) is being pursued for the high-temperature gasifiers. ABB has proposed a milder pressurization (perhaps up to 5 bar), and MTCI is an atmospheric-pressure design. Additional details regarding these gasifier technologies are available elsewhere (Larson et al., 1996; Grace and Timmer, 1995; Stiggson and Hessebom, 1995; Larson et al., 1996; Dahlquist and Jacobs, 1992; Aghamohammadi et al., 1993). Gasifier development by companies other than these four is at a much-reduced level of intensity (Industra, 1996).

Results of calculations of the full-load performance of gasifier/gas turbine systems incorporating gasifier designs based on three of the four designs under commercial development are reported here. Calculations based on the MTCI gasifier design have yet to be completed.

**MODELING BLACK-LIQUOR COGENERATION**

**General Approach**

The starting point for the calculations is a computation model originally developed to predict the full-load, design-point performance of complex gas-steam power cycles (Consonni, 1992). In the model, the system of interest is defined as an ensemble of components, which can be of twelve basic types: pump, compressor, combustor, gas turbine expander, heat exchanger, mixer, splitter, steam cycle (which includes HRSG, steam turbine, pumps and all auxiliaries), air separation plant, shaft (which accounts for turbomachinery interconnections and electric losses), saturator, and chemical converter. Operating characteristics and mass and energy balances of each component are calculated sequentially and iteratively until the conditions (pressure, temperature, mass flow, etc.) at all interconnections converge toward stable values. The basic model is extensively described elsewhere (Consonni et al., 1991; Lozza, 1990; Lozza et al., 1993; Chiesa et al., 1993; Chiesa et al., 1994).

The model accurately simulates full-load performance of plant components that are crucial to the energy and mass balances (e.g., the gasifier, the gas turbine, the steam cycle, the heat exchanger network), but treats simplistically other components which, despite their technological relevance, have minor impacts on the plant mass and energy flows (e.g., a hot gas filter). The modeling work reported here ignores altogether the important impacts that gasification of black liquor might have on chemical recovery steps outside of a pulp mill's cogeneration plant, e.g. on the lime kiln (NUTEK, 1992; Industra, 1996; Larson et al., 1996). These impacts have little or no bearing on the heat and mass balances of the gasifier/gas turbine system, but would need to be considered at any actual mill.

**Reference Black Liquor**

Mass and energy balance calculations with the model require only a correct estimate of the heating value, the dry solids content and ultimate composition of the input fuel (Consonni and Larson, 1996; Larson et al., 1996). The dry solids content and ultimate composition determine the flux of atomic species into the power plant. The heating value allows closing the heat balances. Black liquor characteristics adopted for all calculations are shown in Table 1. Dry solids can include up to 1% chlorine (not shown in Table 1). Minor chlorine content has practical relevance, but a negligible impact on heat and mass balances, and so is neglected here.

**Table 1. Reference black liquor, with 75% dry solids (ds) content**

<table>
<thead>
<tr>
<th>Ultimate Composition (dry weight %)</th>
<th>Heat Value (MJ/kg ds)</th>
</tr>
</thead>
<tbody>
<tr>
<td>C</td>
<td>H</td>
</tr>
<tr>
<td>----</td>
<td>----</td>
</tr>
<tr>
<td>37.2</td>
<td>3.6</td>
</tr>
</tbody>
</table>

(a) The net heating value (NHV) is commonly used outside of North America. It accounts for unrecovered latent heat of water vapor and unrecovehed heat tied up in reduced sulfur compounds leaving a reactor in practice. The NHV is calculated here using the formulas of Adams et al. (1997): NHV (MJ/kg) = HHV (MJ/kg) • 2.44(18/2)xH - 12.9x(78/32)xS - n. Where H and S are the mass fractions of hydrogen and sulfur in the original black liquor and n is the assumed efficiency of sulfur reduction. 100% sulfur reduction is assumed in the above table.
Gasification

The high sodium and potassium content of black liquor help catalyze gasification reactions, so that even at relatively low temperatures, the composition of the product gas is close to equilibrium, except for methane and hydrogen sulfide (H₂S), e.g., see Table 2. This simplifies modeling the gasifiers for heat and mass balance purposes. Since discrepancies in H₂S are much smaller than those with CH₄, and the concentrations of H₂S are small regardless, ignoring the discrepancy in H₂S yields a negligible error in the overall heat and mass balance of the plant. Corrections for non-equilibrium methane content are made based on compositions reported by Berglin (1996) and Backman and Solmenoja (1994). While there is some uncertainty as to what the methane concentration would be in practice, sensitivity calculations confirm that variations around the concentrations assumed here have little impact on overall plant heat and mass balances.

Gas Turbine

An important feature of the computation model used here is its capability to predict the full-load design-point performance of actual commercial gas turbines without the need for proprietary manufacturers data. For specificity, the Siemens KWU 64.3a turbine is selected here for all calculations. The crucial parameters that determine turbomachinery efficiencies and cooling flows in the model are calibrated such that the model reproduces published performance of the actual commercial engine operating on natural gas fuel (Table 3). The good match with quoted performance provides a basis for confidence in the results for the gasification-based calculations.

The same gas turbine is selected here for all calculations to help illustrate intrinsic differences among alternative gasifier designs. The black liquor fuel requirements for a 64.3a turbine are approximately to the black liquor flow at typical modern kraft mills having production capacities ranging from 1100-1800 air-dry metric tonnes of final product per day. A perfect matching between the quantity of gasified black liquor available at a mill and the fuel requirements of a specific gas turbine will be rare, because the fuel availability is largely determined by considerations related to pulp and paper production (and, possibly, the cost-effectiveness of generating excess power for external sale), while the size of the gas turbine is determined by the few models available on the market. A practical operating strategy at a mill (not considered here) might involve supplementing the available gasified black liquor with natural gas or gasified biomass to provide the full fuel requirement of a gas turbine.

While results are shown for only one gas turbine, the significance of the results is not restricted to this turbine alone: very similar results would be obtained for turbines of other manufacturers that are of the same basic type (heavy-duty industrial), same power output class, and same generation of technology. For example, the General Electric 6001FA is very similar to the Siemens KWU64.3a.

The use of gasified black liquor would require some adjustment in the operating parameters of an actual turbine. In particular, the fuel flow required to reach the design turbine inlet temperature will be higher than with natural gas, due to the lower volumetric energy density. All other factors constant, this larger fuel flow increases the flow through the turbine-expander, and thereby the required expander inlet pressure. Increasing the engine's compression ratio to achieve this may cause compressor stall. To avoid simulating unrealistic compressor operation, the engine's compression ratio is allowed to increase modestly (from 16.6 to 17), but further increase is limited by simulated closing of the compressor inlet guide vanes, which reduces air flow to the engine to avoid stall.

Some hardware modifications may also be required, including enlarging the fuel nozzle flow area. The calculations assume that such hardware changes can be made without significantly affecting thermodynamic performance.

The accurate integrated modeling of actual commercial gas turbines is
when scrubbing $H_2S$. Given the much higher concentration of $CO$ than specifications and also recovering sulfur for recycle in the condensed phase. In practice, some $CO_2$ would also be removed $H_2S$ in the raw gas, we have assumed that 2 moles of $CO$, are removed for $H_2S$ must be removed from the raw syngas—all sodium and potassium leave to be demonstrated. Within the model adopted here for the gasifier, only biomass boiler flue oxygen 4% (dry volume) Steam cycle deaerator pressure 2 to $B$iomass boiler flue oxygen 

Table 4. Tomlinson performance.

<table>
<thead>
<tr>
<th>INPUT ASSUMPTIONS</th>
<th>This work</th>
<th>Adams, 1997</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ambient air temp., °C</td>
<td>15.0</td>
<td>27.0</td>
</tr>
<tr>
<td>Feedwater temp., °C</td>
<td>89</td>
<td>110</td>
</tr>
<tr>
<td>Air prereheat temp., °C</td>
<td>150</td>
<td>150</td>
</tr>
<tr>
<td>Smelt exit temp., °C</td>
<td>850</td>
<td>850</td>
</tr>
<tr>
<td>Carbon conv. to gas, %</td>
<td>99.00</td>
<td>98.48</td>
</tr>
<tr>
<td>Heat loss, % liquor HHV</td>
<td>2.0</td>
<td>2.5</td>
</tr>
<tr>
<td>Stack temp., °C</td>
<td>177</td>
<td>210</td>
</tr>
<tr>
<td>Flue $O_2$ (% dry vol.)</td>
<td>3.0</td>
<td>2.8</td>
</tr>
<tr>
<td>Soot blow steam, lbs</td>
<td>0.18</td>
<td>0.18</td>
</tr>
<tr>
<td>Blowdown, lbs</td>
<td>0.11</td>
<td>0.11</td>
</tr>
<tr>
<td>Steam pressure, bar</td>
<td>60</td>
<td>62</td>
</tr>
<tr>
<td>Steam temp., °C</td>
<td>450</td>
<td>482</td>
</tr>
</tbody>
</table>

CALCULATED

| Steam produced, lbs | 3.36 | 3.11 |
| Heat to steam, %* | 65.0 | 62.5 |

(a) Heat in steam delivered to turbine less heat in soot-blow steam divided by input liquor HHV.

Table 5. Calculated gasifier heat and mass balances.

<table>
<thead>
<tr>
<th></th>
<th></th>
<th></th>
<th></th>
<th></th>
</tr>
</thead>
<tbody>
<tr>
<td>Assumptions</td>
<td>Reactor temp., °C</td>
<td>700</td>
<td>950</td>
<td>1000</td>
</tr>
<tr>
<td>Reactor pressure, bar</td>
<td>2</td>
<td>25</td>
<td>25</td>
<td>25</td>
</tr>
<tr>
<td>Liquid feed, % dry solids</td>
<td>75</td>
<td>75</td>
<td>75</td>
<td>75</td>
</tr>
<tr>
<td>Liquid feed temp., °C</td>
<td>115</td>
<td>n.a.</td>
<td>115</td>
<td>115</td>
</tr>
<tr>
<td>Air feed, lbs/hr</td>
<td>1.67</td>
<td>1.91</td>
<td>n.a.</td>
<td>n.a.</td>
</tr>
<tr>
<td>$O_2$ feed, lbs/hr</td>
<td>n.a.</td>
<td>n.a.</td>
<td>0.471</td>
<td>0.580</td>
</tr>
<tr>
<td>Oxidant temp., °C</td>
<td>268</td>
<td>390</td>
<td>135</td>
<td>135</td>
</tr>
<tr>
<td>Calculated Results</td>
<td>Raw gas, lbs</td>
<td>2.543</td>
<td>2.784</td>
<td>1.344</td>
</tr>
<tr>
<td>Ar (mol %)</td>
<td>0.498</td>
<td>0.555</td>
<td>0.570</td>
<td>1.062</td>
</tr>
<tr>
<td>$CH_4$</td>
<td>0.908</td>
<td>2.324</td>
<td>2.151</td>
<td>2.192</td>
</tr>
<tr>
<td>$CO$</td>
<td>11.615</td>
<td>10.532</td>
<td>23.059</td>
<td>21.894</td>
</tr>
<tr>
<td>$CO_2$</td>
<td>12.405</td>
<td>11.452</td>
<td>18.571</td>
<td>20.740</td>
</tr>
<tr>
<td>$CO$</td>
<td>0.019</td>
<td>0.028</td>
<td>0.051</td>
<td>0.021</td>
</tr>
<tr>
<td>$H_2$</td>
<td>19.083</td>
<td>10.307</td>
<td>22.371</td>
<td>12.503</td>
</tr>
<tr>
<td>$H_2O$</td>
<td>13.005</td>
<td>17.585</td>
<td>31.522</td>
<td>40.927</td>
</tr>
<tr>
<td>$H_2S$</td>
<td>0.637</td>
<td>0.594</td>
<td>1.081</td>
<td>0.264</td>
</tr>
<tr>
<td>$NH_3$</td>
<td>0.007</td>
<td>0.009</td>
<td>0.002</td>
<td>0.000</td>
</tr>
<tr>
<td>$N_2$</td>
<td>41.817</td>
<td>46.615</td>
<td>0.321</td>
<td>0.393</td>
</tr>
<tr>
<td>Condensed phase, lbs</td>
<td>0.463</td>
<td>0.491</td>
<td>0.460</td>
<td>0.436</td>
</tr>
<tr>
<td>Steam or dry solid</td>
<td>dry solid</td>
<td>dry solid</td>
<td>dry solid</td>
<td>dry solid</td>
</tr>
<tr>
<td>$Na_2CO_3$ (weight %)</td>
<td>0.000</td>
<td>0.004</td>
<td>0.002</td>
<td>0.118</td>
</tr>
<tr>
<td>$Na_2CO_3$</td>
<td>81.873</td>
<td>79.945</td>
<td>79.525</td>
<td>52.206</td>
</tr>
<tr>
<td>$Na_2S$</td>
<td>7.668</td>
<td>8.005</td>
<td>7.901</td>
<td>17.398</td>
</tr>
<tr>
<td>$NaOH$</td>
<td>0.084</td>
<td>1.906</td>
<td>2.119</td>
<td>9.222</td>
</tr>
<tr>
<td>$K_2SO_4$</td>
<td>0.000</td>
<td>0.005</td>
<td>0.002</td>
<td>0.130</td>
</tr>
<tr>
<td>C</td>
<td>0.801</td>
<td>0.905</td>
<td>0.806</td>
<td>0.450</td>
</tr>
</tbody>
</table>

Table 5. Detailed assumptions used in models.

<table>
<thead>
<tr>
<th>Gasifier heat losses</th>
<th>0.5% of black liquor, HHV</th>
</tr>
</thead>
<tbody>
<tr>
<td>Carbon conversion to gas</td>
<td>99%</td>
</tr>
<tr>
<td>Gas &amp; condensed phase compositions</td>
<td>equilibrium, except $CH_4$</td>
</tr>
<tr>
<td>Syngas coolers minimum $\Delta T$</td>
<td>10°C</td>
</tr>
<tr>
<td>Gas-side AP</td>
<td>2%</td>
</tr>
<tr>
<td>Heat exchanger minimum gas-liquid $\Delta T$</td>
<td>10°C</td>
</tr>
<tr>
<td>Minimum gas-gas $\Delta T$</td>
<td>25°C</td>
</tr>
<tr>
<td>AP</td>
<td>2%, unless otherwise shown</td>
</tr>
<tr>
<td>Heat losses</td>
<td>Zero</td>
</tr>
<tr>
<td>HRSG minimum $\Delta T$ at pinch points</td>
<td>10°C</td>
</tr>
<tr>
<td>Minimum $\Delta T$ at approach points</td>
<td>25°C</td>
</tr>
<tr>
<td>Minimum subcooling $\Delta T$ at drum inlet</td>
<td>10°C</td>
</tr>
<tr>
<td>Superheater AP</td>
<td>10%</td>
</tr>
<tr>
<td>Economizer AP</td>
<td>25%</td>
</tr>
<tr>
<td>Maximum steam superheat</td>
<td>500°C</td>
</tr>
<tr>
<td>Gas-side AP</td>
<td>300 mm $H_2O$</td>
</tr>
<tr>
<td>Heat losses</td>
<td>0.7% of heat released by gas</td>
</tr>
<tr>
<td>Condensate return from mill</td>
<td>80% at 110°C</td>
</tr>
<tr>
<td>Steam cycle degasser pressure</td>
<td>2 to 4 bar</td>
</tr>
<tr>
<td>Condenser pressure</td>
<td>0.056 bar (35°C)</td>
</tr>
<tr>
<td>Heat rejection parasitic power</td>
<td>2% of rejected heat</td>
</tr>
<tr>
<td>Pump efficiencies</td>
<td>(volume flow)</td>
</tr>
<tr>
<td>Steam turbine efficiencies</td>
<td>(volume flow, D, rpm, etc.)</td>
</tr>
<tr>
<td>Electric generator efficiency</td>
<td>(power output)</td>
</tr>
<tr>
<td>Biomass boiler flue oxygen</td>
<td>4% (dry volume)</td>
</tr>
</tbody>
</table>

an important distinguishing feature of the present work relative to other studies that have examined black liquor gasifier/gas turbine applications in the kraft pulp and paper industry, e.g. Berglin (1996), Ihen (1994), and McKeough et al. (1995). Product Gas Cleaning

For all plant configurations, cleaning of the syngas is assumed to be carried out by an isothermal scrubbing at 110°C. A variety of commercially proven processes are available for $H_2S$ removal from syngas, but the practical feasibility of scrubbing gasified black liquor to meet turbine specifications and also recovering sulfur for recycle to the pulp mill remains to be demonstrated. Within the model adopted here for the gasifier, only $H_2S$ must be removed from the raw syngas—all sodium and potassium leave in the condensed phase. In practice, some $CO_2$ would also be removed when scrubbing $H_2S$. Given the much higher concentration of $CO_2$ than $H_2S$ in the raw gas, we have assumed that 2 moles of $CO_2$ are removed for each mole of $H_2S$, which may overestimate the $CO_2$ removal that would result in practice (McKeough et al., 1995). The $CO_2$ removal rate has an important impact on carbonizing requirements during recovery of the pulping chemicals (NUTEK, 1992), but has a negligible impact on cogeneration performance, the focus of this work.

Heat integration

For each plant configuration modeled here, an effort has been made to optimize the heat integration among components so as to maximize efficiency within practical cost (and material) constraints. With this in mind, the network of heat exchangers in each system has been designed following two guidelines. First, high-temperature gas streams transfer heat only to water or steam-water mixtures (evaporators): due to the high heat transfer coefficients achievable with water and two-phase mixtures, this arrangement guarantees acceptable heat exchanger metal temperatures. Second, to the extent possible in practice, heat is transferred across relatively small temperature differences and between flows having similar thermal capacities. This reduces heat transfer irreversibilities and increases overall system efficiency.

Thermodynamic Data

For all computations requiring thermodynamic property data, the JANAF tables are used (Stull and Prophet, 1971; Gardner, 1984) in
Tomlinson Boiler

To provide a consistent comparison between gasification-based systems and Tomlinson boiler cogeneration systems, the Tomlinson technology is modeled at a comparable level of detail. Table 4 compares model results with an industry-standard performance estimate. The difference in overall performance between the two is due primarily to the difference in assumed stack gas temperature and black liquor heating value (14.363 GJ/tds in this work and 13.950 GJ/tds in Adams et al.). In all calculations here, the Tomlinson steam pressure is set at 60 bar, a common level in practice to minimize corrosion in the furnace and superheater. Pressures of 80 bar or higher are found at a few mills worldwide. With pressures above 60 bar, power production efficiencies would be higher than calculated here, but the comparisons between calculated performance of Tomlinson-based systems and gasifier-based systems would not be qualitatively different.

Process Steam Demands

The demand for process steam at a market-pulp or integrated pulp and paper mill sets the requirement for steam to be supplied by the cogeneration plant. Depending on many factors (end product, mill location and age, installed process equipment, etc.), mill steam demands can vary significantly (Fig. 2). Calculations are carried out here for a range of process steam demands. For steam demand levels that cannot be met using black liquor alone, supplemental consumption of biomass in a boiler is included.

A combined-cycle configuration is considered with each gasifier. Steam at 90 bar feeds the bottoming back-pressure steam cycle. The requisite amount of 10 bar process steam is extracted, with the balance of steam exhausting at 4 bar. (One case without a steam bottoming cycle is also considered. In this case, process steam is generated directly at 10 bar and 4 bar.) In all cases, substantial quantities of low grade heat are rejected to the environment. No effort is made to find useful applications of this heat, e.g. as process hot water for the mill.

PROCESS FLOW DESCRIPTIONS

Performance calculations are reported here for cogeneration systems based around Tomlinson boilers and three different gasifier designs, including pressurized, air-blown and pressurized, oxygen-blown high-temperature reactors (based on designs of Kvaerner and Noell) and one low-pressure, low-temperature air-blown reactor (based on the design of ABB). Figures 3-6 show sample flow schematics.

High-Temperature Air-Blown Gasifier

The basic plant configuration with the pressurized, air-blown, high-temperature gasifier is shown in Figure 4. Black liquor fed at 115°C is gasified at 25 bar in air bled from the gas turbine compressor. The product gas passes through an integral quench bath and is further cooled by raising low pressure steam and preheating makeup and condensate return water. Water condenses from the product gas in this process and is recirculated to the quench bath. The quench bath water preheats the recirculated condensate. Weak wash (containing 30 grams/liter of NaOH) is used in the low-temperature scrubber to capture H$_2$S. The clean syngas is preheated before being fired in the gas turbine. Preheating does not appreciably improve cycle efficiency, but because of the low heating value of the fuel gas it is important in increasing combustion stability. Steam is raised at 90 bar in the HRSG from the relatively clean turbine exhaust.

High-Temperature, Oxygen-Blown Gasifier

This plant configuration (Fig. 5) is conceptually similar to the previous one. Oxygen is used in place of air, permitting higher reactor temperatures.
they are dissolved in water to form green liquor. This recovery of heat from the discharged bed solids before the size of all components to be reduced. One notable feature included in this process configuration is the flow along the syngas path is cut by about 50%, allowing load relative to compressing a gas, and by pumping black liquor feed, which is a small parasitic injected into the gas turbine combustor is accomplished of the compression work required for the fuel gas to be.

Gasifier cases), which improves heat recovery, atmospheric pressure: considered, which offers some advantages over ABB technology). A mild pressurization (2 bar) is blown fluidized-bed gasifier (modeled after the proposed Noel\] gasifier design, which includes from gas turbine compressor bleed air. The gasifier modeled in this case is assumed to be produced from ambient air rather than in integrated fashion to be reached and the use of smaller-capacity equipment. Oxygen is assumed to be produced from ambient air rather than in integrated fashion from gas turbine compressor bleed air. The gasifier modeled in this case is non-adiabatic (based on the proposed Noel gasifier design, which includes steam recovery from a reactor cooling jacket). Compared with the system in Fig. 4, the higher fuel gas heating value allows a saturator to be included upstream of the gas turbine to heat the syngas and increase its water vapor content. In effect, the saturator enables use of low temperature heat in the brayton cycle, rather than in the less efficient rankine cycle as with the air-blown gasifier. The use of oxygen in place of air provides for the possibility of substantially higher gasification temperature—1400°C is considered in addition to 1000°C.

Low-Temperature, Air-Blown Gasifier

This plant configuration (Fig. 6) involves an air-blown fluidized-bed gasifier (modeled after the proposed ABB technology). A mild pressurization (2 bar) is considered, which offers some advantages over atmospheric pressure: (i) scrubbing of the raw syngas can be done at elevated temperature (110°C, as in the other gasifier cases), which improves heat recovery, (ii) part of the compression work required for the fuel gas to be injected into the gas turbine combustor is accomplished by pumping black liquor feed, which is a small parasitic load relative to compressing a gas, and (iii) the volumetric flow along the syngas path is cut by about 30%, allowing the size of all components to be reduced. One notable feature included in this process configuration is the recovery of heat from the discharged bed solids before they are dissolved in water to form green liquor. This

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DISCUSSION OF RESULTS

Performance results are reported for cogeneration systems at kraft market pulp mills, where black liquor is assumed to be produced at a rate of 1.74 tds per air-dry metric tonne of pulp (tds/admt). Black liquor feed rates for the gasifiers were set to the requirements of the gas turbine, as discussed earlier. The feed rate for the Tomlinson boiler (90 tds/hour) was selected to fall in the mid-range of the gasifier feed rates. Table 5 shows key assumptions used in the calculations. Gasifier heat and mass balance results are reported in Table 6. Figures 3-6 show sample total-plant heat and mass balances, with corresponding details in Table 7.

Figure 7 summarizes all of the heat and mass balance results. Power production per air-dry metric tonne of pulp (kWht-ddmt) for four gasifier-based systems and for a Tomlinson-based system are shown there. For reference, Mill #1, #2, and #3 process steam and power demands correspond to those shown in Fig. 2. For each technology in Fig. 7, power output is given for a range of process steam production levels. The lower bound of each shaded region represents the power output assuming that equal masses of medium-pressure (MP, 10-bar) and low-pressure (LP, 4-bar) process steam are generated. The upper bound represents a situation in which all process steam is LP steam. Black liquor fuel alone is sufficient to provide the level of total process steam demand at the left endpoints of each curve. For higher levels of steam demand, biomass fuel is burned in a boiler whose steam production is integrated into the steam production from black liquor. The different markers in Fig. 7 indicate different levels of biomass fuel input. For comparison, the waste bark and hog fuel at a typical kraft

Fig. 6. Heat and mass balance for low-temperature, air-blown gasifier in combined cycle with an integrated biomass boiler. Process steam demand of 16.3 GJ/admt is met. See Table 7 for details.

mill converting logs into pulpal chips might amount to 0.25 dry tonnes of biomass per tonne of final pulp, or about 5 GJ/admt. Many mills may have access to much more biomass. One detailed study around a Weyerhaeuser market-pulp mill in North Carolina identified a sustainable supply of up to 3 dry tonnes of biomass per tonne of pulp (or 60 GJ/admt) at reasonable cost in the form of harvest residues and self- and externally-generated mill residues (Weyerhaeuser et al., 1995).

The curves for the gasification-based technologies in Fig. 7 are steeper than for the Tomlinson technology because of a higher biomass boiler pressure—the pressures are assumed to match those of the gas turbine HRSG (90 bar) and the Tomlinson boiler (60 bar). Also, as the biomass share of total energy input increases, differences in power output from one technology to the next diminish in most cases. (At the highest level of biomass input shown, the biomass energy input approaches the black liquor energy input.)

The gasifier/gas turbine systems produce considerably greater kWh/admt than the Tomlinson-based system. However, more supplemental biomass must be consumed with these systems to meet the same process steam demand. At lower levels of process steam demand (e.g., as at Mill #1), the gasification-based systems produce two (with simple cycle gas turbine) to three (with combined cycle) times the kWh/admt as the Tomlinson system. At higher process steam production (e.g., as needed at Mill #3), the simple-cycle gas turbine provides comparable kWh/admt as the Tomlinson, while the combined-cycle systems produce about twice as much power. In all cases, including with the Tomlinson-based system, power production is in excess of a typical pulp mill's process steam needs. In contrast, most North American pulp mills today use Tomlinson plus supplemental-fuel boilers with back-pressure steam turbines, but operate relatively inefficiently and so generate little or no excess power.

Several interesting comparisons among alternative gasifier-based systems are illuminated by a more detailed examination of the results in Fig. 7 for a fixed mill process steam demand. Table 7 summarizes performance calculations for alternative cogeneration systems that provide a level of process heat and power per admt that characterizes Mill #2.

One interesting comparison is between a simple-cycle and a combined-cycle system with the same gasifier (high-temperature, air-blown is considered here). With a high back-pressure turbine (4-bar) only about 20% of the power output of the combined cycle fired only with gasified black liquor is provided by the steam turbine (Fig. 7). The fraction is larger than this for Mill #2 (see Fig. 4) due to the use of biomass as a supplemental boiler fuel. Correspondingly, the difference in power output between the simple and combined cycles meeting Mill #2's steam demand is larger than would be the case without supplemental firing. The marginal efficiency of producing additional power with the combined cycle in place of the simple cycle is about 60%. On the other hand, the simple cycle requires only about half the biomass fuel of the combined cycle to meet the same process steam demand.

A second comparison is between two combined cycles using high-temperature gasifiers, one air-blown (operating at 950°C) and the other oxygen-blown (operating at 1000°C). The O₂-blown system is a more efficient electricity producer, but a considerably less efficient steam producer (Table 7). This can be attributed largely to the use of the saturator with the O₂-gasifier in place of a low-pressure evaporator. (The difference in electricity output would be still greater if oxygen were produced in an integrated fashion from air bled from the gas turbine compressor.) The net result is that to meet the same process steam demand, the O₂-blown system must consume considerably more biomass than the air-blown system. Electricity production per admt is correspondingly much higher (Fig. 7).

A third comparison is between two O₂-blown systems operating with different gasification temperatures (Table 7). The steam production efficiency of the system operating at 1400°C is comparable to that for the 1000°C system, but the electricity production efficiency is considerably lower. This can be attributed primarily to the greater irreversibilities generated in the raw syngas quench step from 1400°C to 200°C, instead of 1000°C to 200°C. In the range 1000F-1400C, higher gasification temperature may provide benefits relating to chemical recovery due to the lower sodium carbonate levels in the smelt (Table 6) (Larson et al., 1996), but energy efficiency will suffer.

A fourth comparison is between combined cycles with high-temperature/high-pressure gasification and one using low-temperature/low-pressure gasification. The latter system provides the highest electrical efficiency of any systems considered here (Fig. 7 and Table 7). This results
primarily because with a lower gasifier outlet temperature, a larger fraction of the gasifier output enters the inherently more-efficient Brayton cycle rather than the rankine bottoming cycle. (With higher gasifier outlet temperatures, a larger fraction of the gasifier output is recovered as steam, which can only be used in the rankine cycle. Stated another way, more of the syngas fed to a high-temperature gasifier must be fully oxidized to reach reaction temperature, with the result that less gas is converted into chemical energy in the product gas.) Also, the irreversibilities involved in cooling the gasifier product gas are smaller with the low-temperature design. Pressurization of the high-temperature gasifiers partly offsets the electric-efficiency advantage enjoyed by the low-temperature system.

<table>
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<th>Table 7. Summary results for cogeneration with black liquor and biomass that meets steam demands per air-dry metric tonne of pulp (adm) corresponding to those for “Mill #2” in Fig. 2. Figs. 3-6 show detailed flows for 4 of the 6 cases summarized here.</th>
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<tr>
<td><strong>Gascification - Biomass Boiler</strong></td>
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<td><strong>Air-bl.</strong></td>
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<td><strong>Corresponding Figure #</strong></td>
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<td><strong>Mill production, adm/day</strong></td>
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<tr>
<td><strong>Black liquor tons/adm</strong></td>
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<td><strong>LHV, kJ/mol</strong></td>
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<td><strong>Process steam, total</strong></td>
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<td><strong>GJ/adm, 10-bar</strong></td>
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<td><strong>GJ/adm, 4-bar</strong></td>
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<td><strong>kW/adm</strong></td>
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<td><strong>Excess Power, MW,</strong></td>
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A final observation is that adiabatic flame temperatures (AFTs) for the clean syngas streams fed to the gas turbine combustor in all cases (Table 7) are well above current turbine inlet temperature limits. So that future increases in firing temperatures can probably be achieved with gasified black liquor. An expected correlation between AFT and syngas heating value (molar basis) is observed in Table 7. Interestingly, the 1400°C O2-blown gasifier gives the lowest AFT, while the low-temperature, air-blown gasifier gives the highest. With the O2-blown gasifier the turbine fuel is rich in H2O due to the sauzaor (Fig. 5), bringing down the AFT, while nitrogen-dilution with the air-blown gasifier is minimized by the low reaction temperature, thereby raising the AFT.

The final two rows in Table 7 are of interest in a situation where a mill is considering replacing an existing Tomlinson-based cogeneration system, a likely common situation over the next two decades (Fig. 1). A baseline alternative in this situation would be the installation of a new Tomlinson recovery boiler plus a supplemental-biomass boiler to augment steam delivery to a back-pressure steam turbine. If the mill has an opportunity to export power, then the incremental fuel chargeable to power (IFCP) shown in Table 7 provides a measure of the marginal fuel costs associated with the production of exported power using each of the gasification options shown in Table 7 in lieu of the Tomlinson baseline. The numerator of the IFCP is the biomass consumption required in excess of that in the Tomlinson case to meet the same process steam demand. The denominator of the IFCP is the amount of power generated in excess of the Tomlinson power production. The high IFCP values indicate that export power from any of the five systems (especially those based on air-blown gasification) could be generated at low (or negative) marginal fuel costs. Obviously capital investment and operating and maintenance costs would also be considered in any full evaluation of alternative cogeneration technology options.

**CONCLUSIONS**

Black liquor gasification systems offer the possibility for Kraft-based market pulp or integrated pulp and paper mills to generate far more electricity than at present while still meeting process steam demands. Depending on the gasification technology and cycle design, power-to-steam ratios can vary widely, providing flexibility in meeting mill requirements. The present work has focused on better understanding the prospective energy benefits of black liquor gasifier/gas turbine systems. Prospective benefits of gasification with regard to capital-investment, environmental impact, flexibility in preparation of pulping chemicals, safety, and reliability have not been addressed.

Black liquor gasification for gas turbine applications is not yet proven commercially. The level of development of several gasification technologies appears advanced sufficiently that large-scale demonstrations could be launched toward demonstrating commercial viability of gasifier/gas turbine systems. In particular, key features that must be successfully demonstrated include gas cleanup to meet gas turbine specifications, stable gas turbine combustion of syngas, heat recovery from syngas streams and (in the case of the low-temperature gasification process) from solids, cost-effective recovery of pulping chemicals, and overall-plant thermal integration.

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REFERENCES


