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Performance and cost analysis of future, commercially mature gasification-based electric power generation from switchgrass

Haiming Jin, SNC-Lavalin, Houston, TX, USA
Eric D. Larson, Princeton Environmental Institute, Princeton University, NJ, USA
Fuat E. Celik, University of California, Berkeley, CA, USA

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Abstract: Detailed process designs and mass/energy balances are developed using a consistent modeling framework and input parameter assumptions for biomass-based power generation at large scale (4536 dry metric tonnes per day switchgrass input), assuming future commercially mature component equipment performance levels. The simulated systems include two gasification-based gas turbine combined cycles (B-IGCC) designed around different gasifier technologies, one gasification-based solid oxide fuel cell cycle (B-IGSOFC), and a steam-Rankine cycle. The simulated design-point efficiency of the B-IGSOFC is the highest among all systems (51.8%, LHV basis), with modestly lower efficiencies for the B-IGCC design using a pressurized, oxygen-blown gasifier (49.5% LHV) and for the B-IGCC design using a low-pressure indirectly heated gasifier (48.6%, LHV). The steam-Rankine system has a simulated efficiency of 33.0% (LHV). Detailed capital costs are estimated assuming commercially mature ('Nth plant') technologies for the two B-IGCC and the steam-Rankine systems. B-IGCC systems are more capital-intensive than the steam-Rankine system, but discounted cash flow rate of return calculations highlight the total cost advantage of the B-IGCC systems when biomass prices are higher. Uncertainties regarding prospective mature-technology costs for solid oxide fuel cells and hot gas sulfur clean-up technologies assumed for the B-IGSOFC performance analysis make it difficult to evaluate the prospective electricity generating costs for B-IG-SOFC relative to B-IGCC. The rough analysis here suggests that B-IGSOFC will not show improved economics relative to B-IGCC at the large scale considered here. © 2009 Society of Chemical Industry and John Wiley & Sons, Ltd

Keywords: switchgrass; IGCC; gasification; SOFC; Rankine cycle

Correspondence to: Eric D. Larson, Princeton Environmental Institute, Princeton University, Princeton, NJ 08544-1003, USA.
E-mail: elarson@princeton.edu

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Introduction

We report here on the prospective performance and cost of stand-alone facilities for electricity production from switchgrass. We include plant designs that incorporate prospectively commercial biomass gasification technologies integrated with either a gas turbine combined cycle (B-IGCC) or a solid-oxide fuel cell (B-IGSOFC). For comparison, we also include a conventional biomass power plant designed around a steam turbine (Rankine) cycle. For all systems, our analysis assumes that all research, development, and demonstration hurdles have been overcome and that the technologies under consideration have reached commercially mature levels in terms of performance, reliability, and cost. The background for this assumption is discussed by Lynd *et al.*[1]

We develop detailed process simulations (using Aspen Plus software) for systems with input switchgrass capacities of 4536 metric tonnes per day (tpd) dry basis (5000 dry short tons/day), which corresponds to an energy input rate of 983 MW_{HHV} (893 MW_{LHV}, considering the as-received moisture content of 20% (McLaughlin S, 2003, personal communication)). A companion paper[2] discusses the current commercial status of key component technologies. Table 1 gives key characteristics of the feedstock.

Table 1. Characteristics of switchgrass assumed in this analysis.

As-received proximate analysis	
Fixed carbon (wt%)	17.1
Volatile matter (wt%)	58.4
Ash (wt%)	4.6
Moisture content (wt%)	20.0
Lower Heating Value (MJ/kg)	13.6
Higher Heating Value (MJ/kg)	15.0
Ultimate analysis (dry basis)	
Carbon (wt%)	47.0
Hydrogen (wt%)	5.3
Oxygen (wt%)	41.4
Nitrogen (wt%)	0.5
Sulfur (wt%)	0.1
Ash (wt%)	5.7
Lower heating value (MJ/kg)	17.0
Higher heating value (MJ/kg)	18.7

The switchgrass feed rate is considerably larger than has been considered previously for biomass conversion facilities in most published analyses, although a few commercial lignocellulosic-biomass processing facilities this size are in operation today, for example, a number of facilities processing sugarcane bagasse (Suleiman JH, 2004, personal communication). The primary incentive for building large plants is more favorable economics. While it is feasible and economically desirable to build large-scale biomass conversion facilities, most analysis and commercial implementation of bioenergy to date has focused on relatively small-scale conversion plants. This is likely due to the prevailing thinking that low-cost biomass feedstocks (residues and wastes) are necessary for the viability of projects in the near term. Residues and wastes are relatively dispersed resources, so large quantities often cannot be cost-effectively brought to single sites. When large, high-efficiency biomass conversion facilities (of the types being analyzed here) are considered, it becomes more feasible to pay higher prices for biomass feedstocks, since higher biomass prices could be more than offset by capital-cost scale economies and reduced feedstock requirements per unit of final product.[3]

Based on equipment sizing derived from our process simulations, we build up capital cost estimates for our simulated power plants by major plant area drawing on literature sources, extensive discussions with industry experts, and our own prior work. We then present overall financial performance results based on discounted cash flow rate of return calculations.

Gasification-based power plants – design and simulation

In a gasification-based power plant, switchgrass would be brought from a short-term storage area into the feed preparation area where it would be chopped and conveyed to the gasifier feeder. The as-received moisture content of the switchgrass considered here is low enough that active drying is not required before gasification. This saves capital cost and imposes little if any system-efficiency penalty compared to actively drying the switchgrass to a lower moisture content. In the gasifier, the switchgrass is converted into a mix of light combustible species (primarily CO and H_2, with some CH_4), heavy hydrocarbons (tars and oils), and generally unwanted minor contaminants (H_2S, NH_3, HCN, and others). The gas is cooled and cleaned before going to the power island for conversion to electricity. Waste heat generated in the power island and elsewhere in the process is recovered to improve overall efficiency. Depending on the operating pressure of the gasifier and the pressure of the clean gas required at the power island, some intermediate gas compression may be required.

There has been considerable research, analysis, development, and demonstration work over the past 20 years focused on power generation from gasified biomass coupled with gas turbine/steam turbine combined cycles, such that performance and cost projections for such systems can now be made with reasonable confidence. Systems in which gasified biomass is converted to electricity in fuel cells have received less focused effort. Performance and cost projections for such systems necessarily involve a greater level of uncertainty.

An extensive literature search found no consideration given to the assessment of performance or cost of biomass power plants at the large scale we are considering for our baseline designs. (The largest-scale design for a dedicated biomass power plant of any type found in previously published work was for an electricity output of 215 MW_e.[4]) However, one paper[5] suggests that at the scale we are considering, the most attractive gasifier design from an overall cost perspective will likely be some type of pressurized fluidized bed (Fig. 1). We have designed and simulated in detail the performance of three gasification-based switchgrass power-generating systems. One biomass integrated gasification combined cycle (B-IGCC) and one biomass integrated gasification solid oxide fuel cell (B-IGSOFC) are designed around a pressurized, oxygen-blown, fluidized-bed gasifier. A second B-IGCC is designed around an indirectly heated, atmospheric-pressure fast-fluidized-bed gasifier that has been the focus of considerable development efforts in the United States.[6] Appendix A provides detailed assumptions incorporated into our Aspen plus process simulations described below.

B-IGCC with oxygen-blown gasifier

Pressurized air-blown gasification has been demonstrated in a B-IGCC application at pilot scale,[7,8] and feasibility studies have been carried out for commercial-scale demonstrations.[9,10] These and most other efforts involving pressurized partial-oxidation gasification for B-IGCC applications have involved air as the oxidant. An alternative is oxygen, which brings with it the benefit of a smaller gasifier and downstream equipment sizes since nitrogen is not present. At the relatively small scales (under about 80 MW_e) that have been the focus of most B-IGCC development efforts to date, the cost savings from smaller physical equipment sizes would probably not justify the use of oxygen, since oxygen costs are fairly scale sensitive. At the large scale under consideration here, however, performance and cost benefits can be gained by using oxygen-blown gasification and including an air separation unit as an integral component of the B-IGCC system.

Little or no consideration has been given in the past to pressurized oxygen-blown fluidized-bed gasification for power generation, but there have been some development efforts relating to this type of gasification for fuels or chemicals production from biomass. For such applications, oxygen-blown gasification is economically preferred

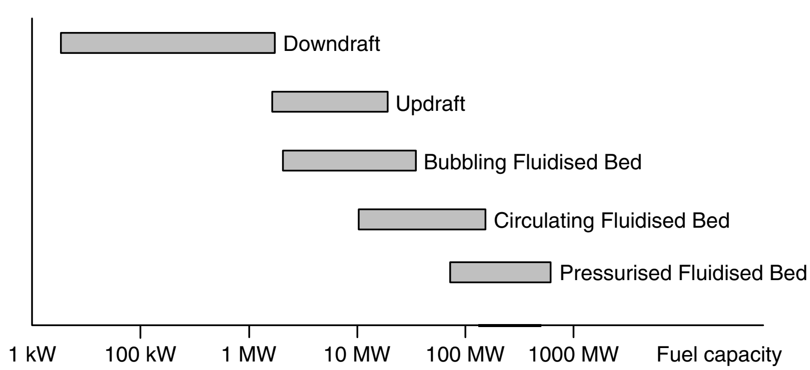

Figure 1. Preferred gasification technology for power generation as a function of input biomass rate.[5]

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Appendix A: Detailed assumptions for Aspen Plus process simulations

Fuel preparation and handling power consumption

- Chopping and conveying: 10 kJ/kg of as-received biomass.
- Lock-hoppers: 10 kJ/kg dry matter (greater work required with pressurized gasifier is accounted for by pressurization of N_2 accounted elsewhere (see ASU below).

Gasifier

- Heat loss as a fraction of biomass HHV: 1% (O_2 gasifier); 2% (indirectly heated gasifier).
- Injection pressures: steam 6% overpressure, oxidant 5%, biomass 2%.
- Tar generation: 1% (1.5%) of dry switchgrass mass flow for oxygen gasifier (indirectlyheated gasifier).
- Tar destruction: O_2 injection in freeboard of O_2 gasifier gives 90% tar cracking. Following indirectly heated gasifier, a catalytic tar cracker converts 100% of tar to light gases.

Filter pressure drops

- Fabric filter 4%; ceramic filter 5%, cyclone separator 0.5%.

Gas turbine

- Simulating frame 7 technology (e. g., General Electric 7FB)
- Compressor: pressure ratio = 19 (19.5) for indirectly heated gasifier case (O_2 gasifier case); polytropic eff = 0.87; mechanical eff = 0.9865; leakage at exit = 0.001 of mass flow.
- Combustor: pressure drop = 3%; heat loss = 0.5% of fuel LHV. Syngas pressure = 1.4x compressor exit.
- Expander: isentropic eff = 0.8977; mechanical eff = 0.9865; cooling flow bypass = 5.2%; turbine inlet temperature = 1370°C; generator eff = 0.986.

Gas expanders

- Isentropic eff * mechanical eff * generator eff = 0.80; no blade cooling; no leakage.

Compressors

- Polytropic eff = 0.80; mechanical eff x electrical eff = 0.90; no leakage.

Heat exchangers

- Pressure drop = 2%; minimum temperature difference = 15°C. Heat lost = 2% of heat transferred.

HRSG and steam cycle

- Pinch point temperature difference = 15°C; three steam pressure levels: 160bar/550°C, 20.5 bar/550°C, 3.5 bar/310°C.
- Steam turbine efficiencies: 1st HP stage 0.75 (160→36 bar), 2nd HP stage 0.78 (36→20.5 bar), IP stage, 0.82 (20.5→3.5 bar); LP stage 0.85 (3.5→1.5bar), and condensing stage 0.82 (1.5→0.05 bar).
- Steam extraction from turbine at 1.5 bar for deaerator; deaerator pressure = 1 bar.
- Condenser pressure = 0.05 bar (32.9 °C).

ASU (Air Separation Unit)

- Oxidant to gasifier or SOFC: 95.0% O_2 (volume), 2.0% N_2, and 3.0% Ar.
- Nitrogen-rich stream: 98.5% N_2, 1.1% O_2, and 0.4% Ar.
- One-third of nitrogen stream is consumed in the molecular sieve and lost to atmosphere; flow of N_2 to lock-hopper for pressurized gasifier is ~10% of dry biomass mass flow; all remaining N_2 goes to GT combustor.
- Cold box distillation pressures:
  - Stand-alone ASU: high pressure column 5.9 bar, low-pressure column 1.5 bar.
  - Integrated ASU/GT high pressure column 8.3, low-pressure column 3.4 bar.

Solid Oxide Fuel Cell (SOFC)

- Full internal reforming of CH_4 and C_2H_6 to $CO+H_2$.
- SOFC working temperature of 800°C
- Conversion of $C+H_2$ = 85%
- Pressure loss = 7%
- Heat loss = 2% of the syngas LHV input
- Cell voltage = 0.7 V
- AC/DC electrical efficiency = 95%.

Other assumptions

- Ambient air: temperature = 25°C, pressure =1.01325 bar, relative humidity = 60%
- Ambient water: temperature = 25°C, pressure = 1.01315 bar
- Minimum stack temperature: 90°C
- Efficiency of pumps = 65% (including mechanical and electrical losses)

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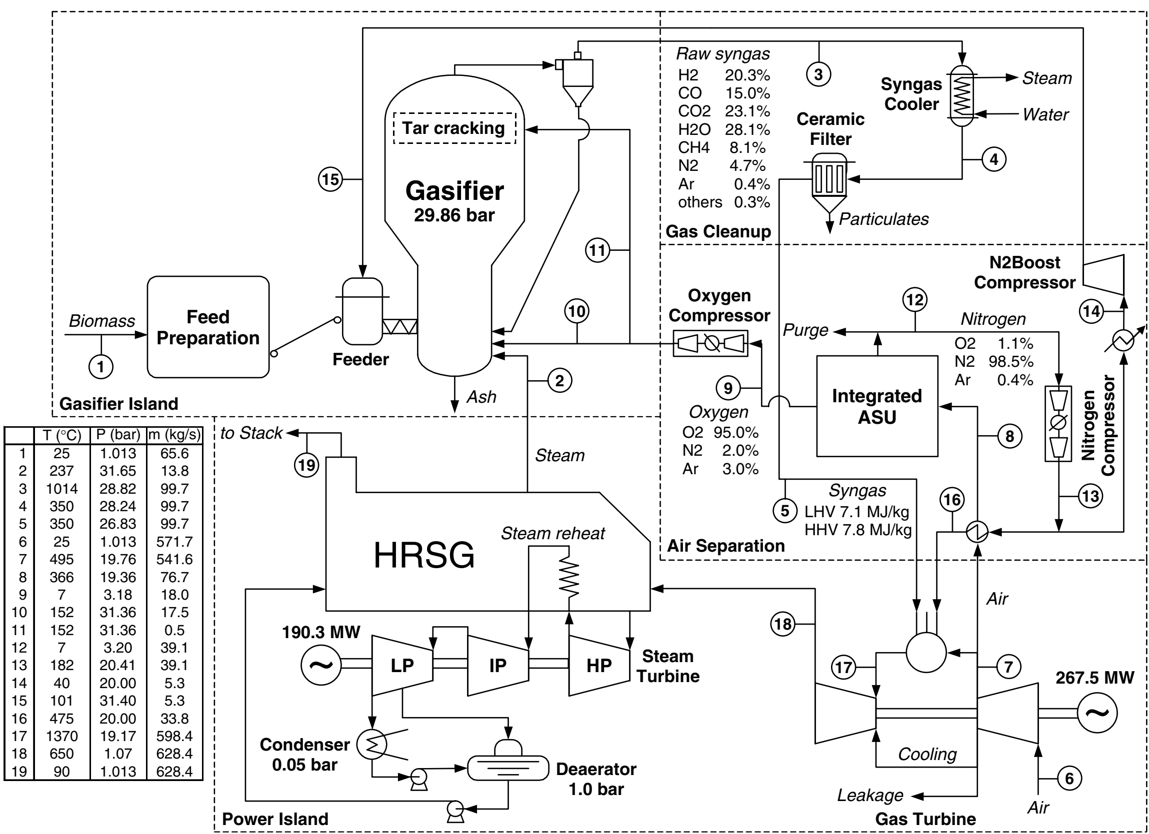
to air-blown gasification at most scales.* Development and pilot-plant demonstration efforts with oxygen-blown fluidized-bed gasification date to the early 1980s in Sweden[11,12] and the mid-1980s in the USA.[13,14] Most such efforts were curtailed when world oil prices fell in the late 1980s. With growing interest in hydrogen as an energy carrier, there has been some recent re-assessment of pressurized, oxygen-blown gasification.[15] The knowledge base relating to pressurized, oxygen-blown gasification is sufficiently developed to enable reasonably accurate performance estimates here.

Process summary

To simulate a B-IGCC system using a pressurized, oxygen-blown gasifier, we have modeled the gasifier after the design of the Gas Technology Institute (GTI).[16,†] As shown in Fig. 2, switchgrass is fed into the pressurized gasifier, from which raw synthesis gas (syngas) leaves at about 1000°C. After cooling and cleaning, the gas is burned in a gas turbine to generate power. The hot exhaust gases from the turbine, together with heat recovered from elsewhere in the process, generate steam in a heat-recovery steam generator (HRSG) that expands through a steam turbine to generate additional power. An air separation unit (ASU), integrated with the gas turbine, provides the oxygen for the gasifier.

* The synthesis of fuels or chemicals from gasified biomass will typically involve reactions under pressure that are driven by the partial pressures of the CO, H_2, and other reacting species in the gas. Inert nitrogen, which would be present in a gas from an air-blown gasifier, is a diluent that would raise compression requirements and reduce the partial pressures of the reacting species and lead to lower synthesis rates. Moreover, fuels and chemicals production will often involve separation, recompression, and recycle of unconverted gas back to the pressurized synthesis reactor. The energy penalty and added equipment cost associated with separating and recycling large amounts of nitrogen and other inert compounds would be substantial.

† The license for the GTI technology is currently owned by Carbona/Andritz.



	T (°C)	P (bar)	m (kg/s)
1	25	1.013	65.6
2	237	31.65	13.8
3	1014	28.82	99.7
4	350	28.24	99.7
5	350	26.83	99.7
6	25	1.013	571.7
7	495	19.76	541.6
8	366	19.36	76.7
9	7	3.18	18.0
10	152	31.36	17.5
11	152	31.36	0.5
12	7	3.20	39.1
13	182	20.41	39.1
14	40	20.00	5.3
15	101	31.40	5.3
16	475	20.00	33.8
17	1370	19.17	598.4
18	650	1.07	628.4
19	90	1.013	628.4

Figure 2. Process simulation results for B-IGCC using pressurized, oxygen-blown gasification. The input biomass moisture content is 20% by weight.

An important design consideration for maximizing the efficiency of switchgrass conversion to electricity is effective heat integration, whereby heat from various process steams that require cooling is transferred to others that require heating. To provide a consistent basis for optimal heat integration within practical constraints, we have carried out a pinch analysis[17] for each process configuration examined in our work. The temperature and mass flows of all process streams requiring heating and all process streams requiring cooling are used as inputs to this analysis. The pinch methodology matches streams that need heating with those that need cooling. Typically, in thermochemical conversion processes, there is waste heat available in excess of that required to meet process heat demands. In the modeling here, this excess heat is assumed to be used for steam raising in an HRSG, with most of the steam being expanded through a steam turbine to generate additional power. (Heat that is available at too low a temperature to generate useful steam is rejected to cooling water.) The pinch analysis enables an accurate estimate of steam generation potential at different steam pressure levels.

As summarized in Table 2, the net power output for this B-IGCC design (Fig. 2) is 442 MW_e, and the electricity-from-switchgrass efficiency is 45.0% on a higher heating value (HHV) basis or 49.5% on a lower heating value (LHV) basis. Details of the plant design and performance simulation are given in the next sections.

Feed preparation and handling

Switchgrass is transported from short-term onsite storage to the feed preparation area, where it is chopped for feeding to the gasifier. The only commercial technology today for feeding low-moisture-content biomass into a high-pressure (> 20 bar) reactor is a lock-hopper. This technology is well proven, but it suffers from high consumption of inert pressurizing gas and the associated gas-compression work required. The feeder model adopted in our simulation assumes successful development of double lock-hopper or hybrid lock-hopper/plug-feed concepts that would considerably reduce the consumption of inert gas without significant added cost.[15] Nitrogen (98.5% purity) is available at low pressure (3.2 bar) from the air separation unit (ASU), and is used to pressurize the feeder.

Table 2. Performance summary for B-IGCC using pressurized, oxygen-blown gasifier.

Switchgrass input, MW_{th}	Higher heating value	983
	Lower heating value	893
Internal power use, MW_e	ASU power[a]	−5.8
	O_2 compressor power	5.3
	N_2 compressor power	10.8
	N_2 boost compressor power	0.3
	Steam cycle pumps, total	3.5
	Fuel handling	0.7
	Lock hopper/Feeder	0.5
Total internal power use, MW_e		15.4
Gross gas turbine output, MW_e		267.5
Gross steam turbine output, MW_e		190.3
NET POWER OUTPUT, MW_e		442.4
Net Electricity Efficiency, %	Higher heating value	45.0%
	Lower heating value	49.5%

[a] For a stand-alone ASU, the air compressor is a large parasitic power load. In the integrated ASU/GT used in this case, the air is compressed in the gas turbine compressor (to 19 bar), and an air expander is added upstream of the ASU to recover some power as electricity while reducing the air pressure to the level needed for the ASU (11 bar in this case). The power term here is the power recovered in the expander.

Gasifier

The gasifier performance is based on empirical data for pilot-scale operation of the GTI technology.[18,19,‡] Switchgrass is

‡ Since biomass gasification is a kinetically controlled process, and kinetic parameter values are not well known, we have developed an approach to modeling biomass gasifier heat and mass balances that relies on empirical data. We use a combination of Aspen reactor modules. Since Aspen Plus is not able to model solid biomass explicitly, we first convert the biomass into fictional components using Aspen's RYIELD reactor. There, the biomass is converted into gaseous H_2, O_2, N_2, H_2O, S, and solid C, as well as ash. These components, together with 98.5% pure nitrogen (used for feeder pressurization), 95% pure oxygen, and steam are fed to an RGIBBS reactor. The steam flow rate is set to simulate the overall dry biomass-to-moisture input ratio indicated in empirical data and the oxygen rate is set to achieve a target reactor temperature (1003°C). We allow the RGIBBS module to calculate a product composition at chemical equilibrium, subject to the following constraints. We specify the output of tar (modeled as abietic acid, $C_{20}H_{30}O_2$) to be 1% by weight of the dry biomass, and we specify the following volume fractions in the product gas based on empirical data: CH_4 (8.2%), C_2H_4 (0.15%) and C_2H_6 (0.15%). We assume 1% of the biomass higher heating value is heat loss. Following the RGIBBS reactor, an RSTOIC reactor is used to adjust the product H_2/CO ratio to be 0.72 to match empirical data.

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injected near the bottom of the reactor together with 0.61 m^3/s of oxidant from the ASU (containing 95% O_2) and 0.90 m^3/s of steam from the HRSG.[§] The simulated gasifier operates at 29.9 bar and 1003°C. The pressure is set such that the clean syngas arriving at the gas turbine combustor has sufficient pressure for injection into the combustor. The feasibility of gasifying switchgrass at the temperature assumed here has not been demonstrated. A switchgrass ash fusion temperature (under oxidizing conditions) of 1016°C has been reported in the literature.[20] If this temperature is also representative for reducing conditions (gasification), then it may be necessary to include an additive in the gasifier bed material to suppress ash fusion. Additives have been demonstrated to be able to raise ash fusion temperatures in biomass combustors.[21, 22]

The primary gasification reactions produce a mix of light combustible gases, heavy hydrocarbons (condensable tars and oils), unconverted carbon (char), and ash, as well as small amounts of hydrogen sulfide, ammonia, alkali compounds, and other gaseous polar impurities. Additional oxygen is injected into the freeboard of the gasifier (above the bed) to promote cracking of the tars and oils to lighter molecules. The literature suggests that 90% conversion of tars and oils to CO and H_2 can be achieved by this oxygen injection.[23] The heat released in the exothermic tar cracking reactions raises the temperature of the gasifier raw product mix to 1014 °C. A cyclone separates the gas from entrained solids (ash and unconverted char), and the gas goes to the clean-up area for treatment. The solids are re-circulated to the gasifier, where the char is assumed to be consumed completely.

Gas clean-up

The temperature of the syngas leaving the cyclone is reduced to 350°C in a syngas cooler that raises steam to augment steam generation in the HRSG. The syngas cooler is a vertical fire-tube design (hot gas inside the tubes) to minimize deposition of condensed alkali species and particulates. Subsequent clean-up of the gas is carried out at 350°C.[¶]

When the syngas is cooled to 350°C, alkali vapors condense onto particulate matter suspended in the gas. Tar and other impurities remain as vapor. By keeping the syngas temperature sufficiently high to avoid tar condensation, the risk of clogging downstream equipment with condensed tars is reduced. Similarly, condensation of ammonia and formic acid, which could damage downstream equipment, is avoided. At 350°C, the syngas retains considerable water vapor, which carries the benefit of added mass flow through the gas turbine expander.

For clean-up, the gas at 350°C passes through a barrier filter (ceramic or sintered metal) which removes particles, along with the alkali species condensed on the particles. The gas is then sent to the gas turbine combustor, where it (including tars and impurities) burns completely before expanding through the turbine.[**]

The simulated 'cold gas efficiency' (chemical energy in the cleaned syngas divided by energy in the input switchgrass), is 79.8% on a lower heating value basis.

Air separation

The ASU produces one stream containing 95.0% (by volume) oxygen and one containing 98.5% nitrogen. The oxygen is compressed in a three-stage intercooled compressor and sent to the gasifier. One-third of the nitrogen is consumed internally in the ASU and is ultimately purged to the atmosphere. The rest of the nitrogen is compressed in a two-stage intercooled compressor. Most of this is sent to the gas turbine combustor to maintain the mass balance between the compressor and turbine (discussed below), to increase power output, and to control thermal NO_x. A small amount (14%) is further compressed in a boost compressor and is used to pressurize the switchgrass feeding system.

The ASU operation in this design is integrated with the gas turbine operation, as contrasted with a stand-alone ASU.

§ Unless otherwise noted, all volumetric gas flows are expressed in this paper in terms of actual volume (not at standard or normal conditions).

¶ Gas cleaning at 350°C, which is more technologically challenging than lower-temperature clean-up, is advantageous thermodynamically only when pressurized gasification is used. If low-pressure gasification were used, the syngas would need to be compressed after cleaning for injection into the gas turbine combustor, which requires cooling the gas to well below 350°C for efficient compression.

** Ammonia in the gas will be converted to NOx in the combustion process. In this case, NO_x emission levels may exceed allowed levels in geographic regions where these are regulated, e.g., many urban areas. Most biomass power plants are likely to be located in rural areas, however, where NO_x regulations are less stringent. To control NO_x emissions, ammonia can be removed at high temperature by placing a catalytic ammonia decomposition reactor between the gasifier and the syngas cooler.[24] The unit decomposes NH_3 into N_2 and H_2. The pressure drop through the reactor would reduce only slightly the overall system efficiency.

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A stand-alone design takes ambient air and compresses it in a dedicated air compressor. Product oxygen and nitrogen are further compressed after the ASU to required pressures. An integrated ASU takes pressurized air from the gas turbine compressor and delivers oxygen and nitrogen at slightly higher pressure than a stand-alone ASU. Because of the integration, operating pressures within an integrated ASU are higher than for a stand-alone unit (the stand-alone air compressor delivers 6 bar air, whereas the gas turbine compressor delivers 19 bar air), which reduces the size of the 'cold box' equipment[††] and reduces the need for post-ASU compression of the oxygen and nitrogen. Moreover, the integrated ASU itself actually produces some electricity (5.9 MW_e in the design shown in Fig. 2).

Removing air from the gas turbine compressor for use with an integrated ASU disturbs the mass balance between the gas turbine compressor and expander. This is corrected by returning to the turbine through several streams most of the air sent to the ASU: the nitrogen in the air is cycled directly to the gas turbine combustor, the nitrogen used to pressurize the gasifier feeder returns as a component of the syngas, as does the oxygen supplied to the gasifier. Only a small amount of nitrogen (accounting for roughly 25% by mass of the input air to the ASU) is used internally in the ASU and does not return to the turbine.

The power-generating efficiency of a B-IGCC with an integrated ASU is not significantly different than with a stand-alone ASU. The power output of the gas turbine itself is lower with the integrated ASU arrangement (due to reduced mass flows through the gas turbine expander), compensating for the reduced auxiliary power consumption compared with a stand-alone ASU (including required compression of air, N_2, and O_2 streams). However, an integrated ASU should decrease the net cost of power generation (due to decreased capital cost) compared to a case with a stand-alone ASU.[25]

Power island

The power island consists of a gas turbine/steam turbine combined cycle and associated generator, piping, ducting, and auxiliary equipment. The gas turbine performance here is based on that for a General Electric MS7001FB, which is in the class of most-advanced gas turbines currently in commercial use with natural gas firing. We assume that this gas turbine operating on synthesis gas will reach the same performance as achieved today on natural gas. Historically, natural-gas-fired gas turbine performance has improved steadily over time, so some future improvements can be expected. However, 7FB-class technology is not commercially mature for syngas firing, so assuming that future syngas-fired turbines achieve performance comparable to that of today's natural-gas-fired turbines may be a reasonable estimate for Nth plant B-IGCC systems. To simulate the performance of the 7FB, we first matched the performance of the 7FB on natural gas quoted by the manufacturer by tuning the expander and compressor efficiencies to match power output and overall thermal efficiency for a fixed natural-gas fuel rate and turbine inlet pressure. (We also fixed values of some other parameters, as shown in Table 3.) To simulate the performance of the gas turbine firing syngas, which has a considerably lower volumetric energy density than natural gas, we assumed that the turbine inlet temperature (TIT) with syngas will be the same as with natural gas.[‡‡] We also assumed that the turbine expander operates with a 'choked'

[††] Pressurized air is first cleaned to remove CO_2 and moisture in a molecular sieve unit, and then passes to the 'cold box', where distillation takes place at low temperature. Input air is cooled against cold product streams by heat exchange. An isentropic cryogenic expander further cools the air. Liquefied air is separated into purified oxygen and nitrogen streams (and argon if needed) in cryogenic distillation towers. The high pressure tower reduces the nitrogen content of the oxygen, allowing pure oxygen to be drawn from the bottom of the low pressure tower. Nitrogen is removed from the top of the low pressure column to purge the molecular sieve and for other uses as needed. Liquid products may also be drawn directly from the columns.

[‡‡] Due to the different flow rate and thermo-physical properties of syngas with respect to natural gas, maintaining the same TIT as with the natural-gas version implies higher temperatures throughout the expansion and thus – everything else equal – higher blade metal temperatures and shorter lifetime for the hot parts of the engine. Running a syngas-fired gas turbine at the same TIT as rated for natural gas implies an increase in blade metal temperatures of 20–25°C (when the ambient temperature is 20°C) and an increase in turbine outlet temperature (TOT) of 10–20°C.[26] This is why syngas-fired gas turbines today are typically de-rated (TIT lower by 20–30°C) to match the lifetime and reliability of natural gas-fired versions. However, by the time the Nth B-IGCC plant is realized in practice, TIT and TOT of state-of-the-art gas turbines will be significantly higher than those adopted today, so our assumption of TIT equal to today's TIT with natural gas firing is reasonable.

Table 3. Comparison of quoted and simulated performance of the General Electric 7 FB gas turbine. Quoted and simulated results are compared for natural gas. Simulation results are shown for syngas from both the oxygen-blown gasifier and the indirectly heated gasifier.

	General Electric 7FB Quoted Performance[a]	Our Aspen Simulations		
	Natural gas[a]		Oxygen-blown gasifier	Indirectly heated gasifer
Natural gas flow (kg/s)	9.985	9.985	0	0
Syngas flow rate (kg/s)	0	0	99.71	47.98
Air mass Flow (kg/s)	438.07	438.07	571.64	617.13
Compressor pressure ratio	18.5	18.5	19.5	19
Net electric output (MW_e)	184.4	184.43	267.5	264.83
Exhaust temperature, C	623	623	650.1	636.9
Thermal Efficiency (%)	36.92	36.92	37.9	36.6
Exhaust flow (kg/s)	448.06	448.06	628.4	665.1
Turbine inlet temperature (°C)	1370	1370	1370	1370
GT air filter pressure drop, bar	N/A	0	0	0
GT compressor polytropic efficiency, % [b]	N/A	87	87	87
GT Compressor mech efficiency, %	N/A	98.65	98.65	98.65
Air leakage, %	N/A	0.1	0.1	0.1
Cooling flow bypass%	N/A	5.161	5.161	5.161
Combustion heat loss, % fuel LHV	N/A	0.5	0.5	0.5
GT turbine isentropic efficiency, % [b]	N/A	89.769	89.769	89.769
GT turbine mech efficiency, %	N/A	98.65	98.65	98.65
Generator efficiency	N/A	98.6	98.6	98.6
GT exhaust pressure, bar	N/A	1.01	1.065	1.065

[a]Quoted performance.[44,45] and (personal communication with J. Cerorski, GE Power Systems (2003)). For the parameters shown below turbine inlet temperature, representative values have been assumed and set based on personal communication with Stefano Consonni at the Politecnico di Milano (2003).

[b]These parameter values were tuned to match as closely as possible the quoted net power output and exhaust temperature.

condition at its inlet. Under this condition, the mass flow into the expander cannot be increased (as would be required since a higher mass flow of fuel is required in the turbine combustor to achieve the same energy input as with natural gas firing) without increasing the pressure upstream of the inlet, i.e., raising the compressor pressure ratio and/or lowering the flow rate of combustion air. We increased the compression ratio over the quoted value for a natural gas machine by about 5%, a percentage increase that is within the range typically tolerated by compressors of heavy-duty gas turbines. Table 3 shows the resulting performance prediction for the gas turbine.

The hot exhaust from the gas turbine enters the HRSG, wherein it is cooled to 90°C and then vented to the atmosphere. (Lower exhaust temperatures would enable greater steam production for power generation, but may cause corrosion problems from formation of sulfuric or other acids.) Steam from the syngas cooler is integrated into the steam side of the HRSG. The HRSG produces steam at 160 bar to drive a steam turbine and a small amount of steam at 31.65 bar for injection into the gasifier. Steam for expansion is produced at only one pressure level because there is insufficient low-grade waste heat available to produce intermediate- or low-pressure steam.

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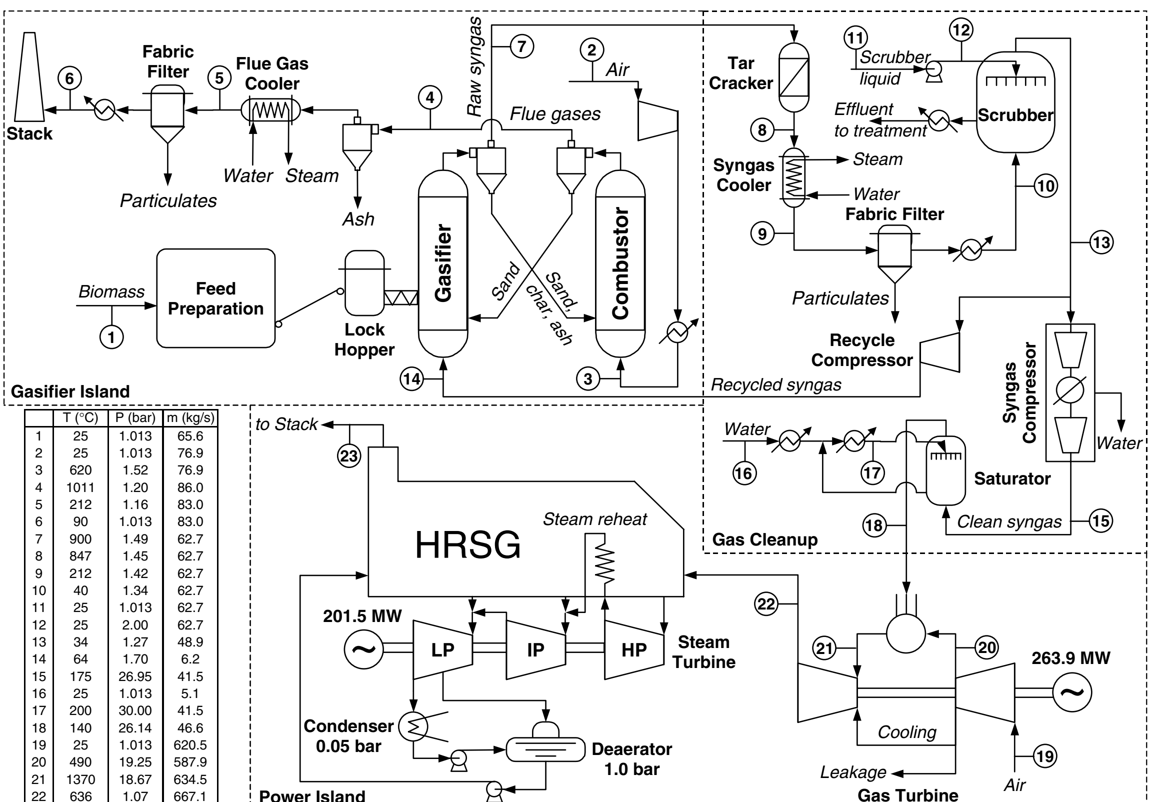
After expanding through the first stage of the steam turbine, the steam is reheated in the HRSG before passing to the second expansion stage. Steam leaving the turbine is condensed and the resulting water is pumped to modest pressure (1.5 bar) before traveling to the deaerator, after which it is further pressurized and returned to the HRSG. A bleed of steam from a low-pressure stage of the turbine is used for deaerator heating.

B-IGCC with indirectly heated gasifier

Alternatives to air or oxygen-blown, partial oxidation gasifier designs that have been the focus of considerable development and demonstration efforts for B-IGCC applications in the USA are indirectly heated gasifiers, in which the heat needed to gasify the biomass is provided through a heat exchanger[27] or by direct contact with an inert heat-carrying material, such as sand.[28, 29] For smaller-scale B-IGCC systems, such designs are attractive because they can produce a gas undiluted by nitrogen without the use of costly oxygen. Without pressurization, however, they are likely to offer less-favorable economics at the large scale of plant considered here. Indirectly heated gasifier designs developed to date are for atmospheric-pressure operation, with poor prospects for pressurization. Nevertheless, for comparison we examine one system that incorporates an indirectly heated gasifier.

Process summary

In this B-IGCC design (Fig. 3 and Table 4) the gasifier is an indirectly heated, fast, fluidized-bed gasifier. In contrast to the B-IGCC with O_2 gasifier, the design in Fig. 3 uses a gasifier operating at nearly atmospheric pressure; it uses no ASU; it uses low-temperature gas clean-up; and it uses a compressor to pressurize the clean syngas for gas turbine use. The steady-state net power output for this design is 431 MW_e, and the electricity from switchgrass efficiency is 43.8 % on a higher heating value (HHV) basis or 48.2 % on a lower heating value (LHV) basis (Table 5). Details of the



	T (°C)	P (bar)	m (kg/s)
1	25	1.013	65.6
2	25	1.013	76.9
3	620	1.52	76.9
4	1011	1.20	86.0
5	212	1.16	83.0
6	90	1.013	83.0
7	900	1.49	62.7
8	847	1.45	62.7
9	212	1.42	62.7
10	40	1.34	62.7
11	25	1.013	62.7
12	25	2.00	62.7
13	34	1.27	48.9
14	64	1.70	6.2
15	175	26.95	41.5
16	25	1.013	5.1
17	200	30.00	41.5
18	140	26.14	46.6
19	25	1.013	620.5
20	490	19.25	587.9
21	1370	18.67	634.5
22	636	1.07	667.1
23	90	1.013	667.1

Figure 3. Process simulation results for B-IGCC using near atmospheric-pressure, indirectly heated gasification. The input biomass moisture content is 20% by weight.

Table 4. Process stream compositions for process streams shown in Fig. 3.

Stream number >>>	2	6	7	8	13	15	18	19	22
Mass, kg/s	76.9	83.0	62.7	62.7	48.9	41.5	46.6	620.5	667.1
Moles, kmol/s	2.66	2.66	2.93	3.00	2.25	1.88	2.18	21.46	23.01
Composition, mol-%									
H_2			13.6%	16.0%	21.6%	22.3%	19.4%		
CO			30.6%	31.5%	42.3%	43.8%	38.1%		
CO_2		19.0%	8.0%	7.8%	10.5%	10.9%	9.4%		6.7%
H_2O			31.2%	28.7%	4.2%	0.8%	13.7%		6.7%
CH_4			11.6%	11.3%	15.2%	15.7%	13.7%		
O_2	21.0%	2.0%						21.0%	12.7%
N_2[a]	78.1%	78.1%	0.4%	0.4%	0.5%	0.5%	0.4%	78.1%	72.9%
Ar	0.9%	0.9%						0.9%	0.9%
C_2H_4			3.9%	3.8%	5.1%	5.3%	4.6%		
C_2H_6			0.5%	0.5%	0.7%	0.7%	0.6%		
NH_3[a]			2 ppm	2 ppm					
Sulfur (as H_2S)[b]			0.1%	0.1%	0.1%	0.1%	0.1%		
Tars[c]			0.1%						
SO_2[d]									71 ppm
LHV, MJ/kg			13.3	13.2	17.0	17.4	15.6		
HHV, MJ/kg			14.2	14.2	18.3	18.8	16.8		

[a] In reality, the NH_3 content of the raw syngas (stream number 7) would be higher. The biomass N would become NH_3 or HCN rather than N_2, and some N would be carried to the combustor in the char. N_2 was chosen for modeling purposes, and choosing another form would not change overall modeling results (efficiency etc.) significantly.

[b] Some biomass sulfur would form COS in the gasifier. Only H_2S was modeled, given the very low overall amount of S in the biomass.

[c] Tar was modeled as abietic acid, $C_{20}H_{30}O_2$ (molecular weight 302.5). While the molar flow would vary with molecular weight the mass flow of tars is independent of the species chosen to model tars, 0.35 kg/s in the raw syngas.

[d] SO_2 emission regulations limit gas turbine exhaust (stream number 22) to 150 ppm at 15% excess oxygen.

energy and mass balances are shown in Fig. 3 and Table 4 and discussed below.

Feed preparation and handling

Switchgrass is prepared as in the previously discussed case, but a simple dual-bin lock hopper feeder is used. A screw feeder moves the biomass from the first, ambient pressure bin to the second bin, which is slightly pressurized with inert gas (nitrogen). A second feed-screw moves the biomass into the gasifier.

Gasifier island

The gasifier operates at near-atmospheric pressure (1.5 bar) and 900 °C in the absence of air (pyrolytic gasification). Heat to drive the gasification is provided directly by hot sand. The sand and biomass are entrained in an upward flow of 4.64 m^3/s of recycled syngas. Syngas, condensable tars and oils, solid char, and ash form in the gasifier. Small amounts of hydrogen sulfide, ammonia, alkali vapors, and other gaseous polar impurities also form.

At the exit of the gasifier a cyclone separator removes most of the entrained solids and the gas travels to the gas clean-up section. The solids removed in the cyclone, which include sand, ash and char, travel to the char combustor vessel through an 'L' valve, and are entrained in 130 m^3/s of combustion air preheated to 620°C. The char burns completely, heating the sand to 1011°C. An additive can be supplied to the combustor to suppress the softening/fusion of ash that might otherwise occur at this temperature.[21, 22]

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Table 5. Performance summary for B-IGCC with indirectly heated gasifier.

Switchgrass input, MW_{th}	Higher heating value (HHV)	983
	Lower heating value (LHV)	893
Internal power use, MWe	Air compressor power	4.2
	Recycle compressor power	0.3
	Syngas compressor power	25.6
	N_2 boost compressor power	0.1
	Steam cycle pumps, total	3.4
	Fuel handling	0.7
	Lock hopper/Feeder	0.5
Total internal power use, MWe		34.8
Gross gas turbine power output, MW_e		263.9
Gross steam turbine power output, MW_e		201.5
Net power output, MWe		430.6
Net Electric Efficiency, %	HHV	43.8%
	LHV	48.2%

A cyclone separates the hot sand leaving the combustor from flue gases and fly ash, and the sand returns to the gasifier. The flue gas then passes through a second cyclone that removes the bulk of the entrained ash. The gas is then cooled by releasing heat in a vertical fire-tube boiler. Alkali vapors condense on the particulate matter present in the flue gas and are removed with the particulates by a fabric filter placed after the cooler. Steam from the flue gas cooler is integrated into the steam cycle in the power island. The flue gas undergoes a final cooling from 212°C to 90°C after the fabric filter before being vented to the atmosphere. The recovered heat is also integrated into the power island.

Gas clean-up

The syngas leaving the cyclone separator first enters a catalytic tar cracking reactor that is assumed to convert over 99% of condensable tars to CO and H_2.[§§] Tar destruction occurs before cooling the gas in order to prevent tar deposition from clogging downstream equipment at lower temperatures. The catalytic reforming of tar is facilitated by the presence of moisture in the raw syngas. The tar-cracking reactions are endothermic, cooling the gas from 900°C to 847°C.

The gas then enters a syngas cooler, exiting at 212°C. The cooler is a vertical fire-tube boiler, similar in design to the combustor flue-gas cooler. Cooling the gas condenses the alkali vapors onto particulate matter in the gas. Any remaining tar or polar impurities with high boiling points may also condense. The particles and condensed matter are then removed in a fabric filter, after which the gas is further cooled (to 40°C).

The final step of the gas cleaning is wet scrubbing. The cooled gas enters the scrubber at the bottom and 25°C liquid is sprayed in near the top of the vessel. The gas cools, and most of the moisture in it is condensed along with the majority of polar impurities, such as ammonia and formic acid. Any remaining particulate matter is also removed. The clean gas exits the top of the scrubber at 34°C. The scrubber liquid effluent passes to a treatment plant, after which it is recycled to the scrubber. Following the gas cleaning, a portion of the clean syngas (12.6%) is recycled via modest compression as the fluidizing gas for the gasifier.

The cold gas efficiency (chemical energy in clean syngas divided by chemical energy in the biomass) is 81% on a lower heating value basis (or 79% on HHV basis), which is approximately the same as in the B-16CC case described earlier. The gas is actually cold and at low pressure here, however, so overall syngas production efficiency is somewhat lower than with the pressurized, oxygen-blown gasifier.

The fuel gas for the gas turbine, after the scrubber, is compressed to 27 bar in a three-stage intercooled compressor wherein 80% of the remaining syngas moisture condenses and is removed. The low temperature of the syngas following the scrubber reduces the amount of compression work needed to pressurize the fuel gas for the gas turbine.

Before combustion in the gas turbine, the syngas is humidified by passing it through a saturator. The moisture content of the syngas is thereby increased, adding mass flow through the gas turbine expander and enhancing power output and overall efficiency. The saturator design is similar to a

§§ Paisley and Overend[30] indicate that up to 90% tar conversion has been achieved in tests with the BCL technology at the Burlington, Vermont pilot plant site. Stevens[24] states that dolomite in an external tar cracker can remove 95–99% of tars from a gas stream at 750–900°C. Finally, Bergman *et al.*[31] describe a new catalytic tar cracking system ('OLGA') that cracks essentially all tars.

scrubber design, with warm water injected countercurrent to rising syngas. A saturator is effective in improving system efficiency when there is low-grade waste heat available for recovery as warm water that would otherwise be wasted.¶¶

Power island

The saturated syngas is burned in a gas turbine to produce power. The gas turbine performance is simulated as described earlier. Hot turbine exhaust and steam (from the combustor flue gas cooler and the syngas cooler) contribute the heat that raises steam in the HRSG. The HRSG produces steam at three pressure levels: 160 bar (HP), 20.5 bar (IP), and 3.5 bar (LP). (Heating of the saturator water and preheating of air for the gasifier-combustor are also integrated into the HRSG.) HP steam enters the HP stage of the steam turbine. After a partial expansion, this steam is reheated in the HRSG before passing through the IP stage of the turbine along with the separately generated IP steam. The IP stage steam exhaust, together with the separately generated LP steam, pass through the LP turbine, and is condensed and pumped to the deaerator and recycled to the HRSG. A small amount of LP steam is bled from the LP turbine and sent to heat the deaerator. The cooled turbine exhaust leaves the HRSG at 90°C and is vented to the atmosphere.

Biomass integrated-gasifier solid oxide fuel cell system

As an alternative to coupling a gas turbine with biomass gasification, we also examine the coupling of a solid oxide fuel cell (SOFC) with gasification. A fuel cell can convert hydrogen-rich gas to electricity with higher efficiency than a gas turbine. Four fuel-cell designs (defined by the electrolyte that separates anode from cathode: phosphoric acid, molten carbonate, proton exchange membrane, and solid oxide) have been the focus of most recent research, development, and demonstration efforts. Among these, the SOFC stands out as the most suitable and promising for gasification-based power generation due to its high tolerance for carbon monoxide in the feed gas, its high operating temperature (which is well matched to integrating with thermochemical gasification processes), and its stable, solid electrolyte.[32]

Process summary

We have chosen to examine a biomass integrated-gasifier SOFC (B-IGSOFC) system utilizing a pressurized, oxygen-blown gasifier (Fig. 4). Because contaminant specifications for feed gas to an SOFC are more stringent than to a gas turbine, a modified gas clean-up strategy is utilized compared to the B-IGCC cases. An external tar cracker is added to ensure complete removal of tars, and the high-temperature filter used in the B-IGCC case is replaced by a process that co-removes particulates and sulfur at elevated temperature. Methane and other trace organic compounds are steam reformed to CO and H_2 at the SOFC anode, which is maintained at the SOFC's assumed operating temperature of 800°C by heating from the exothermic reactions of the reformed gas with oxide ions at the anode. The reformed syngas reacts at the anode with oxygen ions generated at the cathode, producing an external current. (The oxygen is produced in a stand-alone ASU, which also supplies oxygen to the gasifier and tar cracker.) Spent syngas leaves the SOFC and is expanded in a free turbine to generate additional power. The residual combustible components in this spent gas are then burned to generate heat. This heat, along with waste heat from the process, is used to make steam to drive a steam turbine generating additional power. (The low volumetric energy content of the spent gases makes them unsuitable for gas turbine combustion.)

The net power output for this design is 463 MWe, and the electricity from switchgrass efficiency is 47.1 % on a HHV basis or 51.8% on a LHV basis. Table 6 summarizes the performance predicted for the SOFC system shown in Fig. 4. The following sections discuss details of major plant areas following gasification. The feed preparation and gasifier islands for the B-IGSOFC system are identical to those in the first B-IGCC system described above.

Gas clean-up

In contrast to a gas turbine fuel produced by biomass gasification, a fuel produced for an SOFC must be cleaned of

¶¶ No saturator is used in the previously described B-IGCC configuration (with hot gas clean-up) because there is relatively little low-grade waste heat available in the process. Heat could be made available by reducing steam generation elsewhere in the process, but this would reduce the amount of steam available to drive the steam turbine, thereby reducing steam turbine output. The loss in steam turbine output by including a saturator would be greater than the gain in gas turbine output.

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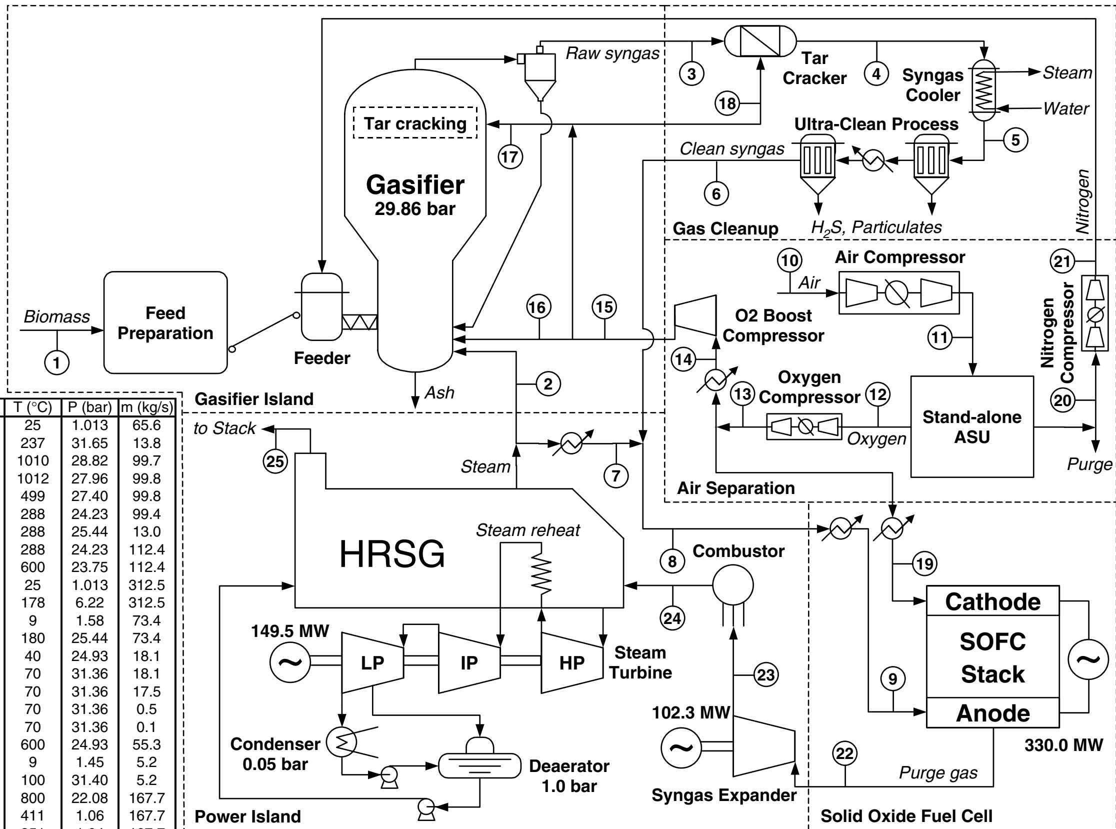

	T (°C)	P (bar)	m (kg/s)
1	25	1.013	65.6
2	237	31.65	13.8
3	1010	28.82	99.7
4	1012	27.96	99.8
5	499	27.40	99.8
6	288	24.23	99.4
7	288	25.44	13.0
8	288	24.23	112.4
9	600	23.75	112.4
10	25	1.013	312.5
11	178	6.22	312.5
12	9	1.58	73.4
13	180	25.44	73.4
14	40	24.93	18.1
15	70	31.36	18.1
16	70	31.36	17.5
17	70	31.36	0.5
18	70	31.36	0.1
19	600	24.93	55.3
20	9	1.45	5.2
21	100	31.40	5.2
22	800	22.08	167.7
23	411	1.06	167.7
24	851	1.04	167.7
25	90	1.013	167.7

Figure 4. Process simulation results for B-IGSOFC using pressurized, oxygen-blown gasification. The input biomass moisture content is 20% by weight.

sulfur. The maximum allowable sulfur (as H_2S and COS) in the feed gas to a SOFC is 1 ppmv.[33] Without active sulfur removal, the sulfur concentration in clean syngas from switchgrass is greater than 350 ppmv in our plant designs. Thus, for an SOFC application, some sulfur removal is required. Sulfur removal at an elevated temperature is desirable for maximizing overall system efficiency, but currently commercial technologies for warm sulfur removal cannot achieve the same sulfur removal levels as low-temperature techniques (e.g., Rectisol). However, ongoing developments promise better warm-gas sulfur removal in the future.

One warm-gas clean-up strategy under development is the Ultra-Clean Dry Cleanup (UCDC) process.[33] This process is being developed primarily for control of sulfur species in coal-derived syngas for applications in synthesis of fuels/chemicals and for SOFCs. It consists of two reactor-filters injecting sorbents that absorb sulfur species (mainly H_2S and COS) and hydrochloric acid (HCl). The reactors operate at about 500°C and 290°C, respectively. The sorbents are filtered out, along with particulates, as the syngas leaves the reactors. The UCDC process is designed as a polishing step to follow bulk sulfur removal. It is intended to bring total sulfur to less than 60 ppbv (also HCl and other halides to less than 10 ppbv and particulates to less than 0.1 ppmw). Because the unmitigated concentration of sulfur in biomass-derived syngas is already relatively low, a bulk-removal step is not required (as it would be with a coal-derived gas). In order to ensure no condensation of tars in the second stage of the UCDC process, an external catalytic tar cracker is included in the gas clean-up system immediately downstream of the gasifier. This is followed by a syngas cooler and first stage of the UCDC. Along with capture of the contaminants, some CO and H_2 (approximately 1.5% of the feed amounts) are lost in the UCDC process. These losses are included in our simulation results.

Table 6. Performance summary for B-IGSOFC with pressurized, oxygen-blown gasifier.

Switchgrass input, MW_{th}	Higher heating value (HHV)	983
	Lower heating value (LHV)	893
Internal power use, MW_e	ASU air compressor power	84.3
	O_2 compressor power	27.2
	O_2 boost compressor power	0.5
	N2 compressor power	2.5
	Steam cycle pumps, total	2.9
	Fuel handling	0.7
	Lock hopper/Feeder	0.5
Total internal power use, MW_e		118.6
Gross SOFC power output, MW_e		330.1
Gross syngas expander power output, MW_e		102.3
Gross steam turbine power output, MW_e		149.5
Net power output, MW_e		463.2
Net Electric Efficiency, %	HHV	47.1%
	LHV	51.8%

Air separation

The ASU supplies oxygen for the gasifier, the SOFC, and the tar cracker. A two-stage intercooled compressor delivers air to the ASU, which produces 95.0% (volume) purity oxygen and 98.5% purity nitrogen. The oxygen is compressed and 75% of it goes to the SOFC cathode. This flow rate ensures that the SOFC tail gas will contain sufficient oxygen for when the gas is subsequently combusted. The remaining 25% of oxygen is further compressed before delivery to the gasifier and cracker. The nitrogen from the ASU, with the exception of a small (2%) stream used for biomass feeder pressurization, is vented to the atmosphere.

Power Island

Electricity is generated in the power island by the SOFC, a gas expansion turbine, and a steam turbine. Clean syngas is mixed with steam and heated to 600°C for delivery to the SOFC, which is operated at 25 bar pressure and 800°C. At the SOFC anode, all methane, ethane, and other light hydrocarbons are reformed into CO and H_2, and the resulting gas mixture reacts at the anode with oxide ions (generated at the cathode) to produce CO_2, water, some residual syngas, and an external current.*** The SOFC is assumed to consume 85% of the CO and H_2 in the fuel.[34,35] The unconsumed CO and H_2 leave the SOFC in a mix with CO_2, H_2O, and oxygen. The SOFC produces 330 MW of electricity, which is 47.2% (42.8%) of the energy in the SOFC feed gas on a LHV (HHV) basis.

The spent gas leaving the SOFC at a pressure of 22 bar is expanded in a free turbine to generate 102 MW of additional electricity. The spent gas is then burned in a catalytic combustor,††† the exhaust from which enters an HRSG raising steam to drive a steam turbine. This turbine produces 149 additional MW of power.

State-of-the-art steam-Rankine cycle – design and simulation

For reference we have simulated a state-of-the-art system for switchgrass power, based on a steam-Rankine cycle (Fig. 5). The assumed boiler design is based on a stoker combustor, where the switchgrass fuel is distributed over a moving grate by a pneumatic stoker. The combustion chamber is separated into a primary and secondary stage. Air passes upwards through the grate and inwards from the walls, causing primary (partial) combustion of the switchgrass in the lower stage and over-fire air introduced in the upper stage of the boiler completes the combustion. In a fixed-grate boiler, ash would collect on the grate and require periodic removal, but the moving-grate arrangement allows ash to be continuously removed and collected for disposal.[36] A forced-draft fan provides air for complete combustion of the biomass plus enough excess air to limit the gas-side temperature to which the boiler tubes are exposed to 870°C. Steam is generated in the boiler tubes at 160 bar, 550°C. One reheat is included in the plant design. Table 7 summarizes the predicted system performance. Net power output is 296 MW_e, with lower and higher heating value efficiencies of 33% and 30%, respectively.

*** The basic anode reactions are $H_2 + O^{2-} \rightarrow H_2O + 2e^-$ and $CO + O^{2-} \rightarrow CO_2 + 2e^-$. The cathode reaction is $\frac{1}{2} O_2 + 2e^- \rightarrow O^{2-}$

††† The spent gas has a very low heating value (0.8 MJ_{HHV}/kg), so a catalyst is needed to sustain combustion.

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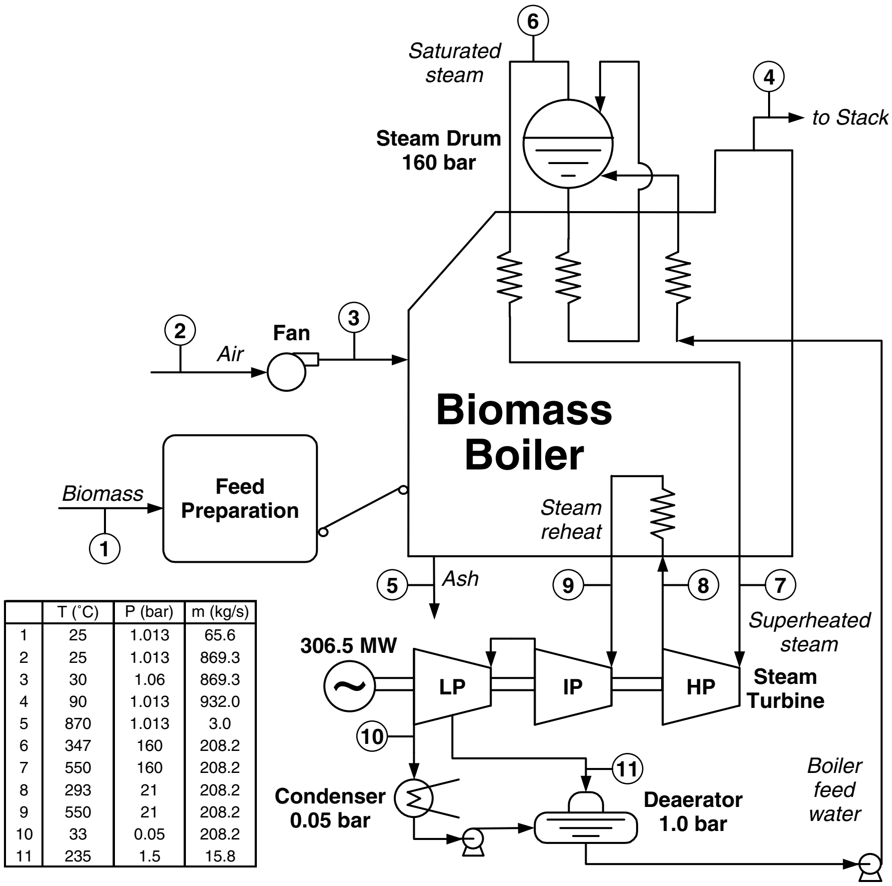

	T (˚C)	P (bar)	m (kg/s)
1	25	1.013	65.6
2	25	1.013	869.3
3	30	1.06	869.3
4	90	1.013	932.0
5	870	1.013	3.0
6	347	160	208.2
7	550	160	208.2
8	293	21	208.2
9	550	21	208.2
10	33	0.05	208.2
11	235	1.5	15.8

Figure 5. Simulation results for switchgrass-fired steam-Rankine cycle. The input biomass moisture content is 20% by weight.

Table 7. Performance summary of conventional steam-Rankine power plant.

Switchgrass input, MW_{th}	Higher heating value (HHV)	983
	Lower heating value (LHV)	893
Internal power use, MW_e	Air blower fan	4.8
	Steam cycle pumps, total	5.5
	Fuel handling	0.7
Total internal power use, MW_e		11.0
Gross steam turbine power output, MW_e		306.5
Net power output, MW_e		295.5
Net Electric Efficiency, %	HHV	30.0%
	LHV	33.0%

Cost analysis

Our cost analysis entails estimating capital costs for commercial (‘Nth plant’) versions of the two B-IGCC and the steam-Rankine plant configurations described above and using these estimates as input to discounted cash flow rate of return (DCFROR) calculations of overall financial performance. (We develop a different type of cost analysis for the B-IGSOFC design, for reasons discussed later.)

For the two B-IGCC systems and the steam-Rankine system, we build up capital cost estimates from the sub-unit level for each major plant area. We draw on a variety of peer-reviewed papers, other reports, industry experts, and our own prior work for cost inputs. We have scrutinized numbers from different sources and made adjustments to original figures as appropriate to develop as self-consistent a set of capital cost estimates as possible. Notes to tables that appear later in this paper give details of our assumptions and sources. We estimate the uncertainty in our capital cost estimates to be ±30%, based on the level of detail (factored estimates for major plant areas), the nature of literature and industry sources (including some equipment-vendor quotes) from which we derived costs, and the inherent uncertainties in projecting ‘Nth plant costs given the pre-commercial status of some of the key system components.

For our DCFROR calculations, we assume the annual operating and maintenance cost for all of our plant designs will be 4% of the overnight installed capital cost, which is a widely used assumption for gasification-based, coal-fired power plants[37] and thus is probably also appropriate for large-scale, gasification-based biomass plants. We specify the price of switchgrass as a parameter with values of $2/GJ$_{LHV}$ ($34/dry metric ton), $3/GJ$_{LHV}$ ($51/dry metric ton), or $4/GJ$_{LHV}$ ($68/dry metric ton). Prospective costs of switchgrass delivered to conversion facilities are discussed by Sokhansanj *et al.*[38] We express all cost inputs and results in constant 2003 dollars, using the GDP implicit price deflator[39] to adjust cost inputs from other-year dollars when needed.[‡‡‡]

[‡‡‡] Some literature sources used for capital cost estimates did not specify the cost basis year. In these cases we assumed a cost basis year one calendar year prior to the publication date. (For example, costs given in a report published in 1996 were assumed to be 1995 dollars.)

Table 8. Reference capacities, capital costs, and scaling factors for major plant areas and sub-units used to develop capital cost estimates for the B-IGCC cases.*

		Capacities (in indicated units)			Cost (in million 2003 $)	
		Base	Max. unit	Unit of Capacity	Base	Scaling exponent
Plant Area	Sub-Unit	S_o	S_{max}		C_o[a]	f
Oxygen-blown gasifier upstream areas						
Gasifier island	Feed preparation[b]	64.6	n.a.	wet tonne/hr biomass	9.84	0.77
	Pressurized O_2 gasifier[c]	41.7	120	dry tonne/hr biomass	6.41	0.7
	Ash cyclone[d]	68.7	180	actual m^3/s gas feed	0.91	0.7
	Lock-hopper N_2 boost comp[h]	10	n.a.	MW_e consumed	4.14	0.67
Gas clean-up	Syngas cooler[e]	77	n.a.	MW_{th} heat duty	25.4	0.60
	Ceramic filter[f]	14.4	n.a.	actual m^3/s gas feed	18.6	0.65
ASU	Integrated ASU[g]	76.6	n.a.	tonne/hr pure O_2	22.7	0.5
	O_2 compressor[h]	10	n.a.	MW_e consumed	5.54	0.67
	N_2 compressor[h]	10	n.a.	MW_e consumed	4.14	0.67
Indirectly heated gasifier upstream						
Gasifier island	Feed preparation[b]	64.6	n.a.	wet tonne/hr biomass	6.79	0.77
	Indirectly heated gasifier[i]	32.3	83	dry tonne/hr biomass	6.03	0.7
	Air compressor[h]	10	n.a.	MW_e consumed	4.14	0.67
	Combustor flue gas cooler[j]	20.4	n.a.	MW_{th} heat duty	2.98	0.6
	Combustor flue gas filter[k]	12.1	64	m^3/s gas feed	1.90	0.65
Gas clean-up	Tar cracker[l]	47.1	n.a.	m^3/s gas feed	0.73	0.7
	Syngas cooler[e]	77.0	n.a.	MW heat duty	25.4	0.6
	Fabric filter[k]	12.1	64	M^3/s gas feed	1.90	0.65
	Scrubber[m]	20.9	64	M^3/s gas feed	0.66	0.7
	Recycle compressor[h]	10	n.a.	MW_e consumed	4.83	0.67
	Syngas compressor[h]	10	n.a.	MW_e consumed	4.83	0.67
	Saturator[n]	20.9	64	M^3/s gas feed	0.22	0.7

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Table 8. *Continued*

Power Island (for both B-IGCC systems)

Power island	Gas turbine[o]	266	334	GT MW_e	56.0	0.75
	HRSG + heat exchangers[p]	355	n.a.	MW_{th} heat duty[q]	41.2	1.0
	Steam turbine[r]	136	n.a.	ST gross MW_e	45.5	0.67

* Column headings: S_o = capacity of unit as given by original source; S_{max} = maximum unit size, C_o = installed cost of unit as given by original source; *f* = scaling exponent. See electronic annex for additional discussion of cost estimating methodology.

Notes to Table 8:

[a] The GDP implicit price deflator[39] has been used to convert to constant 2003 dollars from other year dollars when necessary.

[b] Weyerhaeuser[46] gives the following installed N^{th} plant equipment costs (in 1999$) for feed preparation to feed 64.6 wet tonnes/hr wood chips to a near-atmospheric pressure gasifier: conveyor, $851,000; dried wood chip storage, $561,000; feed bin, $233,000; rotary air lock, $329,000; and water cooled feed screw, $54,000. We have multiplied these by a factor of 3.1 to account for the lower bulk density of chopped switchgrass compared to wood chips.[47] (We assume equipment cost scales linearly with bulk density.) Additionally, for the B-IGCC with pressurized O_2 gasifier, we have multiplied the resulting total by 1.45[42] to account for the cost of a feed preparation system rated for 30 bar. We derived the value of *f* by calculating an overall scaling factor for a feed preparation system with a base capacity of 33.5 wet tonnes/hour from Hamelinck and Faaij[48] consisting of conveyors (C_o = $\$_{2001}$ 0.35M, *f* = 0.8), storage ($1.0M, 0.65), and a feeding system such as a lock hopper and feed screw ($0.41M, 1). Scaling the resulting total feed preparation cost for values of S/S_o between 1 and 3.5 gives an overall scaling factor of 0.77. The cost estimates from Hamelinck and Faaij[48] were not used directly here for C_o because they represent first-of-a-kind plant cost estimates, whereas Weyerhaeuser[47] gives estimates for an N^{th} plant.

[c] From personal communication with Evan Hughes of the Electric Power Research Institute (2003), the cost of an oxygen-blown GTI gasifier operating at 7.93 bar processing 400 dry tpd of bagasse (20% moisture content) is $5 million in 2002 dollars. This price includes installation labor and is given for an N^{th} plant design. Based on Guthrie,[42] the cost of a pressure vessel rated for 30 bar (the gasifier used in the simulation) is 1.26 times that of a pressure vessel rated for 7.93 bar. Therefore the reference cost of $5 million is multiplied by 1.26 to obtain a base cost, C_o, of $6.3 million. According to personal communication with Francis Lau of the Gas Technology Institute (2003), the maximum capacity of an oxygen-blown GTI gasifier operating at atmospheric pressure is roughly 625 dry tpd, at 25 bar is roughly 2500 tpd, and at 30 bar is roughly 15% greater than that at 25 bar (i.e. 2875 tpd). All three of these units would have the same physical dimensions; the increased capacity comes from the smaller volume of gas per tonne of biomass produced in the higher pressure gasifiers. From this information, the maximum capacity of a GTI gasifier in tpd can be calculated at any pressure by 77.8x + 549, where x is the pressure in bar. The maximum capacity at 7.93 bar is 1165 tpd, roughly 40% of the capacity at 30 bar. This means that a 400 tpd gasifier (7.93 bar) could process 2.5 x 400 tpd, or 1000 tpd, if it were pressurized up to 30 bar. For this reason, 1000 tpd (41.7 tonnes/hr) is used as the value of S_o corresponding to a 30 bar gasifier with the same dimensions as the 7.93 bar gasifier noted above. S_{max} is based on personal communication with Francis Lau of the Gas Technology Institute (2003), who indicates that the maximum capacity of a 30 bar GTI gasifier is roughly 2875 tpd or 119.8 tonnes/hr. *f* for a GTI gasifier is assumed to have a value of 0.7 based on personal communication with Evan Hughes of the Electric Power Research Institute (2003) and several other references.[18,48,49,50,51]

[d] Weyerhaeuser[47] gives a cost of $584k (1999 dollars) for a cyclone separator. The ash cyclone is similar to a combustor primary cyclone, and the cost scales with the volumetric flow rate of gas through the unit. Installation labor is included in this cost. This cost is multiplied by 1.45[42] to account for the cost of a cyclone separator rated for 30 bar. The combustor primary cyclone in the Weyerhaeuser[46] design is only rated for near-atmospheric pressure operation. Cost is given for an N^{th} plant design. S_o is volumetric gas feed to the combustor primary cyclone in the Weyerhaeuser[46] design, calculated from the gas mass flow using the temperature, pressure, average molecular weight, and the ideal gas law (for simplicity). S_{max} and *f* are taken from Hamelinck *et al*.[52] for a cyclone separator.

[e] Based on Simbeck,[53] who gives $310/$kW_{th}$ of saturated steam produced in a fire-tube syngas cooler producing 77 MW of such steam from 1000°C input syngas, or $24 million (year 2000$). This cost excludes BOP and indirects. The latter are included (using methodology discussed in online supporting material of Larson, *et al*.[2]) when total plant costs are estimated later in this paper. Scaling exponent for high-temperature heat exchangers is taken from Hamelinck *et al*.[52]

[f] Newby *et al*.[54] give a cost of $14.6 M (1997 dollars) for an N^{th} plant ceramic candle filter ($36.0/kW for a 406 MW coal IGCC plant). This cost does not include installation labor, so 15% is added.[55] S_o is the gas feed rate to the ceramic filter evaluated in Newby *et al*. [54] *f* is taken to have the same value as for a fabric filter in Hamelinck *et al*.[52]

[g] Kreutz *et al*. [40] indicate a cost estimate of $40.4 million (in 2002$) for a standalone air separation unit with an air compressor. This cost includes installation, BOP, engineering and contingency, and is given for an N^{th} plant design. In order to achieve a consistent basis in accounting for indirect costs, engineering and contingency costs were removed from the costs reported in Kreutz *et al*. [40] to obtain C_o. In Kreutz *et al*., [40] the total direct cost (TDC) is the sum of the installed equipment cost and BOP. For the ASU, engineering is 10% of TDC, and contingency is 5% of TDC + engineering. Therefore, to subtract engineering and contingency from Kreutz *et al*.'s cost we divide by (1.1 x 1.05), yielding $35.0 million (or $35.6 million when converted to 2003$). Thus BOP is included in C_o for the ASU, while indirect costs are not. Values of S_o and *f* are also taken from Kreutz *et al*. [40] The ASU used in this study is integrated with the gas turbine, and so no separate air compressor is needed. It has been estimated (using our own Aspen Plus model of a standalone ASU) that a 76.6 tonne/hr O_2 ASU plant requires a 25.7 MW air compressor which costs $7.7M (see note h below for derivation of air compressor costs). Thus, the cost of the integrated ASU (w/o air compressor) is estimated at $35.0 M less $7.7M, or $27.3M. The ASU used in this study also operates at a higher pressure than a standalone ASU due to the integration with the gas turbine and thereby higher input air pressure. This means that the dimensions of the ASU 'cold box' (which includes the high-pressure and low-pressure distillation columns and the low-temperature heat exchangers – essentially everything except the air compressor) are smaller than those needed for a standalone plant producing the same rate of pure oxygen.

Table 8. *Continued*

According to personal communication with R Moore, retired from Air Products and Chemicals (2003), doubling the pressure in the cold box would reduce the physical size of the required distillation columns and other cold box equipment by about half while producing the same rate of pure oxygen. For the ASU, the cost scales with physical size according to a scaling exponent of 0.5 (Kreutz *et al.*[40] and R. Moore, personal communication, 2003), so halving the size leads to a cost of (1/2)^0.5 = 0.707 times the larger size. In our case, the pressure of cold box units are on average 1.5 times those in a standalone ASU, so the roughly 1.5x reduction in size leads to a cost that is (1/1.5)^0.5 = 0.81 times the cost of the lower pressure unit, or a final value of $22.3M (2002 dollars) for C_o ($22.7M in 2003 dollars).

[h] Kreutz *et al*. [40] give the following original costs, in millions of 2002 dollars, for compressors: for oxygen compressors, $6.3M; nitrogen compressors, $4.7M; and PSA purge gas compressors, $6.28M. Here, air compressors are assumed to cost the same as nitrogen compressors, and syngas compressors are assumed to cost the same as purge gas compressors. The former assumption rests on the fact that air is mostly nitrogen. The latter assumption rests on the fact that both the biomass gasification-based syngas in this study and the PSA purge gas in Kreutz *et al*. are low heating value gases consisting mainly of CO, CO_2, H_2O, and H_2. These costs include installation, BOP, engineering and contingency, and are given for an N^{th} plant design. In order to achieve a consistent basis in accounting for indirect costs, engineering and contingency costs were removed from the costs reported in Kreutz *et al*. to obtain C_o. In Kreutz *et al*., the TDC is the sum of the installed equipment cost and BOP. For oxygen and nitrogen compressors, engineering is 10% of TDC, and contingency is 5% of TDC + engineering. For purge gas compressors, engineering is 15% of TDC, and contingency is 15% of TDC + engineering. Therefore, to obtain C_o for oxygen and nitrogen compressors, the cost from Kreutz *et al*. is divided by (1.1x1.05), and to obtain C_o for purge gas compressors, the cost is divided by (1.15 x 1.15). Thus BOP is included in C_o for compressors, while indirect costs are not. Values of S_o and *f* are also taken from Kreutz *et al*.

[i] Weyerhaeuser[46] gives the following costs, in thousands of 1999 dollars, for an atmospheric pressure, indirectly heated BCL gasifier and combustor. The BCL gasifier cost includes the gasifier, $365k; gasifier hot gas line, $132k; gasifier cyclone, $384k; gasifier dashpot, $169k; combustor L valve, $74k; air blower, $586k; combustor, $460k; combustor hot gas lines, $194k; combustor primary cyclone, $584k; combustor J valve, $115k; combustor secondary cyclone, $584k (NB: the secondary cyclone is not included in the original report, and the cost is assumed here to be equal to that of the primary cyclone); sand cooler, $11k; sand silos, $57k; magnesium oxide storage, $7k; expansion joints, $1097k; and misc. ceramic pipe $771k. Installation labor is included in all costs, which are given for an N^{th} plant design. S_o is dry biomass feed to B-IGCC plant as given by Weyerhaeuser.[46] S_{max} is based on Bechtel,[56] Tijmensen,[51] and Hamelinck *et al*.[52] The value of *f* for a BCL gasifier is assumed to be 0.7, based on several references.[18,49,50,51]

[j] Weyerhaeuser[46] gives the following costs, in thousands of 1999 dollars, for a high-temperature flue gas cooler: combustor flue gas cooler, (which is similar to a flue gas HRSG) $2076k, and heat pipe air heater, $684k. Installation labor is included in these costs, which are given for an N^{th} plant design. S_o is the heat duty of the flue gas HRSG of B-IGCC plant in Weyerhaeuser, calculated from the change in enthalpy of the boiler feed-water. *f* for a high-temperature heat exchanger (such as a syngas cooler) is taken from Hamelinck *et al*.[52]

[k] Hamelinck and Faaij[48] indicate a cost of $1.6M (2001 dollars) for a fabric filter. Both the combustor flue gas filter and the syngas fabric filter are similar to a baghouse filter. This cost does not include installation labor, so 15% is added.[55] It is given for a first-of-a-kind plant, though the cost of a well-established technology such as a fabric baghouse filter should not change much as the technology "matures". For this reason, no adjustment is made in an attempt to reach an N^{th} plant value. S_o, $S_{max,}$ and *f* are also taken from Hamelinck and Faaij.[52]

[l] Weyerhaeuser[46] gives a cost of $678k (1999 dollars) for a tar cracker. The tar cracker cost includes installation labor and is given for an N^{th} plant design. S_o is volumetric gas feed to the tar cracker in Weyerhaeuser, calculated from the gas mass flow using the temperature, pressure, average molecular weight, and the ideal gas law (for simplicity). *f* for a tar cracker similar to the one used in Weyerhaeuser is taken from Hamelinck *et al*.[52]

[m] Weyerhaeuser[46] gives the following costs, in thousands of 1999 dollars, for a scrubber. A scrubber includes venturi $23k; scrubber, $206k; scrubber cooler, $201k; skimmer settling tank, $80k; scrubber recirculation tank, $67k; and flare, $34k. Installation labor is included in all costs, which are given for an N^{th} plant design. S_o is volumetric gas feed to the scrubber in Weyerhaeuser, calculated from the gas mass flow using the temperature, pressure, average molecular weight, and the ideal gas law (for simplicity). S_{max} and *f* for a scrubber similar to the one used in Weyerhaeuser are taken from Hamelinck *et al*.[52]

[n] A saturator and a scrubber have essentially the same basic equipment and layout, consisting of a vertical vessel with upward gas flow and spray jets for downward liquid flow. For this reason, the cost for a saturator is assumed to be the same as a scrubber in Weyerhaeuser,[46] $206k (1999 dollars). Although S_o has the same value as for the scrubber, the flow rate of gas exiting the saturator determines the unit size. This is in contrast to the scrubber, where the inflow determines the size, and is because in a saturator the vaporization of water increases the volumetric flow rate of gas through the unit relative to the inlet. The cost of the scrubber in Weyerhaeuser[46] includes installation labor. This cost is multiplied by 1.35 (based on Guthrie[42]) to account for the cost of a saturator rated for 27 bar. The scrubber in Weyerhaeuser[46] is only rated for near-atmospheric pressure operation. Cost is given for an N^{th} plant design. As with the scrubber, S_{max} and *f* are taken from Hamelinck *et al*.[52]

[o] Kreutz *et al*.[40] indicate an installed cost of $72.8 million (2002$) for a Siemens V94.3A gas turbine (266 MW_e output). This cost includes BOP, installation, engineering (15% of installed gas turbine +BOP), and contingencies (15% of installed gas turbine +BOP +engineering). The value in this table for base cost excludes the engineering and contingencies and has been converted to 2003$. Engineering and contingencies are included (using methodology discussed in the online supporting material of Larson, *et al*.[2]) when total plant cost is estimated later in this paper. The assumed scaling exponent was obtained by taking a power-series regression of equipment-only costs vs. power output for all simple cycle gas turbine generators in Gas Turbine World's 2003 Handbook.[57] The value for S_{max} is the power rating of the largest simple-cycle gas turbine generator in Gas Turbine World's 2003 Handbook.[57] For some of the plant designs discussed later, the included gas turbine capacity is considerably smaller than the capacity of a Siemens V94.3A turbine. Our process simulation assumes the unit performance will be the same in all cases, but our cost estimation methodology accounts for higher costs for smaller units.

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Table 8. *Continued*

[p] Based on Simbeck[53] who indicates a cost of \$110/$kW_{thermal}$ of superheated steam for an "HRSG Boiler" (year 2000\$), or \$39 million for a unit producing 355 MW superheated steam. Simbeck excludes BOP and indirects. We include the latter (using methodology discussed in the online supporting material of Larson, *et al*.[2]) when total plant costs are estimated later in this paper. We apply this cost to the heat transfer duty (based on our pinch analysis) that takes place in all heat exchangers throughout the plant (except for the syngas cooler and any water gas shift reactors, since we account for the cost for the syngas cooler and WGS heat exchangers separately). Most of the heat transfer in the process is associated with raising superheated steam. (For example, in the B-IGCC case with O_2 gasifier, approximately 80% of total system heat transfer is accounted for as superheated steam.) Residual heat transfer includes raising warm water for saturators in some cases and heat rejection to cooling water. Our calculation method here implicitly costs heat exchangers needed for these low-temperature heat transfer functions at the same cost per kW transferred as for raising superheated steam. This may overestimate the costs of these lower-temperature heat exchangers. We assume a scale factor of unity, since this cost element for a full plant will typically include several heat exchangers.

[q] Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler and (when present in a system) water gas shift reactors. See previous note for additional discussion.

[r] Kreutz *et al*. [40] indicates an installed cost of \$59.2 million (2002\$) for a steam cycle (steam turbine and condenser) with a gross output of 136 MW_e. This cost includes installation BOP, engineering (15% of installed gas turbine +BOP), and contingencies (15% of installed gas turbine +BOP +engineering). The value in this table for base cost is expressed in 2003\$, excluding the engineering and contingencies. The latter are included (using methodology discussed in the online supporting material of Larson, *et al*.[2]) when total plant cost is estimated later in this paper. Scale factor is from Kreutz *et al.*

Table 9. Capital cost estimate for the B-IGCC with pressurized, oxygen-blown gasifier and having performance characteristics indicated in Table 2.

		Capacities (in indicated units)				Costs (in million 2003 \$)		
Plant Area	Sub-Unit	Required capacity S_r	Number of units n^a	Capacity per unit S^b	Unit of Capacity	Cost per unit C^c	Train cost $C_m{}^d$	Overnight Cost OC^e
Gasifier island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Pressurized O_2 gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash cyclone	16.3	2	8.1	actual m^3/s gas feed	0.21	0.38	0.61
	N_2 boost compressor	0.33	1	0.33	MW_e consumed	0.42	0.42	0.54
Gas clean-up	Syngas cooler	124	2	62	MW_{th} heat duty	21.9	40.9	65.3
	Ceramic filter	8.1	2	4.0	actual m^3/s gas feed	8.1	15.2	24.2
ASU	Integrated ASU	61.3	1	61.3	tonne/hr pure O_2	20.3	20.3	25.7
	O_2 compressor	5.3	1	5.3	MW_e consumed	3.6	3.6	4.6
	N_2 compressor	10.8	1	10.8	MW_e consumed	4.4	4.4	5.5
Power island	Gas turbine	268	1	268	GT MW_e	56.2	56.2	71.4
	HRSG + heat exchangers	433	1	433	MW_{th} heat duty[f]	50.3	50.3	77.2
	Steam turbine	190	1	190	ST gross MW[e]	57.0	57.0	72.4
Total overnight capital cost (million 2003\$)						428		
Total overnight capital cost per unit capacity (\$/kW_e)						968		

[a] Number of units determined by rounding $n = S_r / S_{max}$ up to the nearest integer. Values for S_{max} are given in Table 8. This gives n = 1 whenever $S_r \le S_{max}$ and leads to multiple units whenever $S_r > S_{max}$. We assume each gasifier is coupled to its own feed preparation area and gas clean-up island. Thus, the number of feed preparation/clean-up trains equals the number of gasifiers as determined by the equation above. The values of S_r are determined from the process simulation results for a 4536 tonnes/day plant.

[b] Capacity of each unit given by $S = S_r / n$. If $n = 1$, $S = S_r$.

[c] Cost per unit given by $C = C_o \cdot (S/S_o)f$. See Table 8 for values of C_o, S_o, and f for individual process units. For some units, C will include BOP (see Table 8 notes g, h, and o,), but indirect costs are not included in any values of C (see note e below for discussion of indirect costs).

Table 9. *Continued*

[d] C_m is the cost after accounting for scale economies involved with multiple trains. The cost for all n multiples of a unit is given by $C_m = C \cdot n^m$, where $m = 0.9$.

[e] Overnight capital cost (OC) for each capital unit is the installed cost, plus the balance of plant, plus the indirect costs, or OC = C_m + BOP + IC. Some installed costs (C_m) in Table 8 already include BOP (see note c). For components that do not already include BOP, we estimate BOP as 20.9% of installed cost. This percentage was determined as follows. Hamelinck and Faaij[48] note that the total cost of a plant as a whole grows more quickly with capacity than the BOP cost, so BOP as a % of total cost will be smaller the larger the plant size. We have estimated the % BOP as a function of energy input (biomass HHV MW_{th}) to a B-IGCC plant from a best fit of several literature sources' estimates[9,18,40,46,48,58] for similar plants at varying scales. BOP (%) = 0.8867 / {(biomass $MW_{th})^{0.2096}$}. For the 4536 tonne/day scale investigated in this project, the thermal input (HHV) is 983 MW, so that BOP is 20.9 % of the installed cost. Indirect costs (IC) were estimated from review of several literature sources.[9,18,40,46,52,58,59,60] For gasifier island and gas clean-up units, IC = 32% of total direct costs (TDC = C_m + BOP), and for power island and air separation unit units, IC = 27% of TDC. Thus for a typical gasifier island or gas clean-up unit, OC = C_m x (1.209) x (1.32) = 1.60 C_m. For the air separation unit, compressors, or gas turbine, since BOP is already included in C_o, OC = C_m x (1.27).

[f] Total heat transfer rate considering all heat exchangers in the process, except for the syngas cooler, which is dealt with separately. See note [p] of Table 8.

Capital cost estimates

Table 8 gives detailed reference values from which our capital cost estimates for B-IGCC plants are built up. The resulting capital costs are shown in Table 9 for the B-IGCC plant with an oxygen-blown gasifier and in Table 10 for the B-IGCC with an indirectly heated gasifier. The total capital investment for the two B-IGCC systems is similar: \$428 million and \$456 million, respectively, or \$968/kW_e and \$1059/kW_e. Figure 6 shows a breakdown of these costs by major plant area. Our estimates of total installed cost are lower than most previously published cost estimates for B-IGCC plants. This apparent discrepancy is largely resolved when one accounts for the scale of the B-IGCC plants examined in the different cost studies. Figure 7 shows our installed overnight capital cost estimates for both B-IGCC designs as a function of plant scale. (Discontinuities in the B-IGCC curves reflect the scales at which the number of gasifier or gas turbine trains needed to be increased or decreased.) Overnight capital costs estimated by others for Nth plant B-IGCC systems (with various equipment configurations) are also shown (as points). In the scale range for which there are other literature estimates, our costs fall within the span of costs in the literature,[§§§] thereby providing some basis for confidence in our estimates for larger-scale plants (for which prior literature estimates are not available).

One further calibration of our B-IGCC capital cost estimates is a comparison with estimated Nth plant costs for coal-IGCC (C-IGCC). Kreutz *et al.*[40] have developed such estimates for several C-IGCC plant configurations with net electricity generation of about 400 MW_e. (A number of the major component costs in our B-IGCC estimates are derived from the work of Kreutz *et al.*[40]) For specificity, we consider their plant design that uses a GE-type quench gasifier and generates 390 MW_e of electricity. Kreutz *et al.*'s estimated overnight capital cost for this plant is shown in Table 11 alongside our results for the B-IGCC with an oxygen-blown gasifier scaled (using our capital cost methodology) to 390 MW_e capacity. The C-IGCC total overnight capital cost is \$470 million. Our B-IGCC cost estimate is \$391 million. The difference between these costs can be explained as follows:

- Feed preparation for biomass is more costly than for coal due primarily to the lower bulk density of biomass.
- The biomass gasifier operates at a lower pressure (30 bar) than the coal gasifier (70 bar), and without ash slagging, but the resulting cost savings are largely offset by the high cost for both a syngas cooler and hot-gas particle filter in the B-IGCC case. (Neither of these is present in the C-IGCC design.)
- The low sulfur content of switchgrass obviates the need for any sulfur capture and recovery in the B-IGCC case, since sulfur emission standards can be met without sulfur removal. In contrast, the C-IGCC cost includes \$70 million in sulfur-related capital costs.

[§§§] Our cost curves were developed assuming electricity generating efficiency is the same at all scales. In practice, the equipment design (and hence efficiency) for a specific plant size will reflect a compromise between plant efficiency and capital cost, with higher capital cost being justified by resulting efficiency gains. Since efficiency would actually decrease somewhat with decreasing scale, our cost estimates may underestimate actual costs at the lower end of the scale range.

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Table 10. Capital cost estimate for the B-IGCC with atmospheric-pressure, indirectly heated gasifier and having performance characteristics indicated in Table 5.

		Capacities (in indicated units)				Costs (in million 2003 $)		
		Required capacity	Number of units	Capacity per unit	Unit of Capacity	Cost per unit	Train cost	Overnight Cost
Plant Area	Sub-Unit	S_r	n^a	S^b		C^c	$C_m{}^d$	OC^e
Gasifier island	Feed preparation	236	3	78.7	wet tonne/hr biomass	7.9	21.2	33.9
	Indirectly heated gasifier	189	3	63.0	dry tonne/hr biomass	9.6	25.9	41.3
	Air compressor	4.2	3	1.4	MW_e consumed	1.1	3.0	3.9
	Combustor flue gas cooler	76.1	3	25.4	MWth heat duty	3.4	9.1	14.6
	Combustor flue gas filter	92.1	3	30.7	actual m^3/s gas feed	3.5	9.3	14.9
Gas clean-up	Tar cracker	191	3	63.5	actual m_3/s gas feed	0.90	2.4	3.9
	Syngas cooler	77.8	3	25.9	MW_{th} heat duty	13.2	35.5	56.6
	Fabric filter	85.1	3	28.4	actual m_3/s gas feed	3.3	8.9	14.2
	Scrubber	43.7	3	14.6	actual m_3/s gas feed	0.51	1.4	2.2
	Recycle compressor	0.31	3	0.10	MWe consumed	0.22	0.60	0.79
	Syngas compressor	25.6	1	25.6	MW_e consumed	9.1	9.1	12.0
	Saturator	3.0	1	3.0	actual m_3/s gas feed	0.06	0.06	0.09
Power island	Gas turbine	264	1	264	GT MW_e	55.6	55.6	70.6
	HRSG + heat exchangers	627	1	627	MW_{th} heat duty[f]	72.8	72.8	111.7
	Steam turbine	201	1	201	ST gross MW_e	59.2	59.2	75.2
Total overnight capital cost (million $)						456		
Total overnight capital cost per unit capacity ($/kW$_e$)						1059		

a, b, c, d, e, f. See notes with same lettering in Table 9.

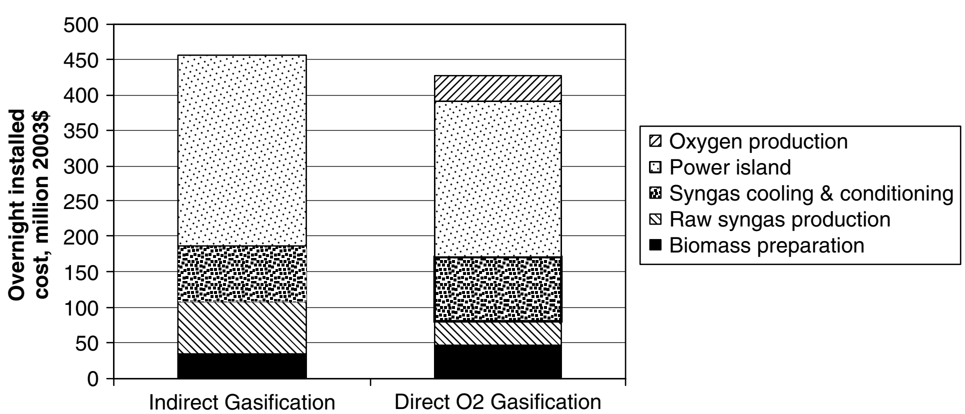

Figure 6. Breakdown by major area of the overnight installed capital cost for two B-IGCC designs.

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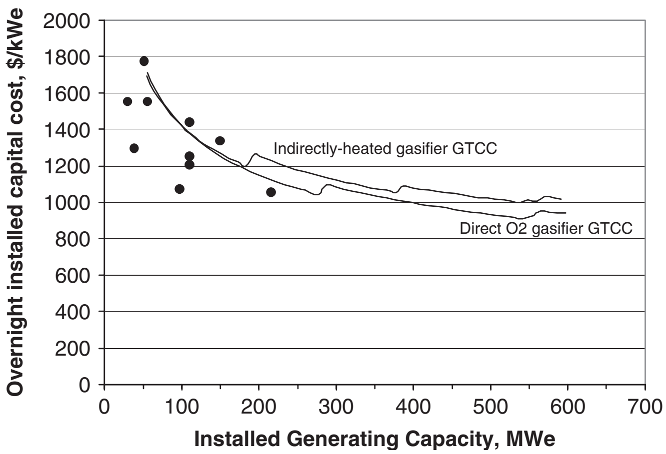

Figure 7. Overnight installed capital cost per kWe as a function of installed generating capacity. The lines are costs developed in this study. Points are estimates for Nth plant (mature-technology) given in the literature.[4, 6,36]

Table 11. Comparison of overnight installed capital costs for a 390 MWe coal-IGCC plant as estimated by Kreutz *et al.*[40] (case EVQ[a]) and a 390 MW_e B-IGCC (based on design in this paper with direct O_2 gasification). Cost are given in 2003 $.

Plant area	Coal-IGCC	B-IGCC
Feed preparation, handling	36.0	42.5
Gasifier	76.3	31.7
Syngas cooler, gas clean-up		83.3
Air separation unit	47.0	24.2
Oxygen compressor	9.0	4.2
Nitrogen compressor	10.7	5.1
Sulfur control	41.5	0.0
Sulfur recovery	28.2	0.0
Gas turbine	74.0	64.9
HRSG and heat exchangers	72.4	68.4
Syngas expander	3.0	0.0
Power island BOP	72.2	66.5
Total	470	391
Total, $/kW_e$	1,205	1,003

[a]GE-type gasifier, with quench gas cooling, Selexol H2S removal, Claus/SCOT sulfur recovery, stand-alone ASU, Siemens 94.3A gas turbine combined cycle. Costs originally in 2002$ converted to 2003$ using GDP deflator.

- Differences in the scale and design of the ASU account for the difference in ASU costs between the two systems. Unlike coal, biomass contains some oxygen and therefore requires less added oxygen for gasification. Also biomass is gasified at lower temperatures than coal, further reducing the required capacity of the ASU. For the systems in Kreutz *et al.*,[40] the coal gasifier consumes 101 t/hr of O_2. Our biomass gasifier uses 61 t/hr O_2. Since the cost for a stand-alone ASU scales with the square root of capacity (Table 8), the B-IGCC ASU could be expected to cost 77% of the C-IGCC ASU, if the ASU designs were otherwise the same, which they are not. The ASU in the C-IGCC case is a stand-alone unit. In the B-IGCC case it is integrated with the gas turbine, which eliminates the large air compressor needed with a stand-alone ASU. Based on Smith *et al.*,[25] the air compressor accounts for about 20% of the total ASU cost at the ASU scale considered here. Moreover, with an integrated ASU, the pressure of air delivered to the ASU (from the gas turbine) will be higher than that from a stand-alone ASU's air compressor. The elevated pressure would result in cost savings due to smaller equipment in the ASU cold box. As detailed in Note (g) of Table 8, a higher-pressure cold box is estimated to cost 81% of a lower-pressure cold box. The smaller oxygen requirement, the elimination of the air compressor, and the smaller cold box for the B-IGCC case explain the $23 million difference in ASU costs compared to Kreutz *et al.*[¶¶¶] Also, because of the integrated ASU design, lower pressure ratio (less costly) oxygen and nitrogen compressors can be used in the B-IGCC case compared with the C-IGCC case.
- The power island components in the BIGCC design are less costly than their counterparts in the C-IGCC design in absolute terms, but individual costs for the gas turbine and the steam turbine are consistent between the two designs. For the C-IGCC, the gross output of the gas turbine is 294 MW, which gives a unit cost of $252/kW. For the B-IGCC design, the gross gas turbine output is 236 MW, for a unit cost of $275/kW. Similarly, for the steam turbine and condenser, the gross output in the C-IGCC case is 179 MW, for a unit cost of $403/kW. For the BIGCC case, the gross steam turbine output is 168 MW, for a unit cost of $396/kW.

[¶¶¶] The coal-IGCC ASU cost is $47.2 million.[40] Accounting for the smaller required oxygen capacity reduces this to (0.77*47.2 =) $36.3 million. Eliminating the air compressor reduces this to (0.80*36.3 =) $29.1 million. Finally, reflecting the cost reduction with a higher-pressure cold box reduces this to (0.81*29.1 =) $23.6 million.

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Thus, our B-IGCC cost estimate appears consistent with Nth plant cost estimates for C-IGCC technology at the scale of about 400 MW$_e$.

For comparison with our B-IGCC cost estimates, we have also developed a capital cost estimate for the conventional steam-Rankine system with performance given in Table 7.

The total installed overnight capital investment of $258 million (Table 12) is considerably lower than that for the B-IGCC systems. However, because of the much lower efficiency of the steam-Rankine system, the total investment per unit of installed capacity is only modestly lower than for the B-IGCCs.

Table 12. Capital cost estimate for conventional steam-Rankine cycle having performance characteristics indicated in Table 7.

				Costs (in million 2003 $)		
	Base Cost (10^6 $, 2003)	Base Scale (MW_{th} biomass input)	Scale exponent	Cost per unit	Train cost	Overnight Cost
Plant area	C_o[a,b]	S_o[a]	f[a]	C[c]	C_m[d]	OC[e,f]
Feed preparation[g]	19.3	543	0.7	29.2	29.2	38.5
Boiler[h]	47.4	543	0.7	71.9	71.9	94.9
Baghouse & cooling tower[i]	3.11	543	0.7	4.71	4.71	6.22
Boiler feed-water & deaerator[j]	6.01	543	0.7	9.1	9.1	12.0
Steam turbine[k]	15.74	543	0.7	23.9	23.9	31.5
Cooling water system[l]	7.04	543	0.7	10.7	10.7	14.1
Balance of plant[m]	29.2	543	0.7	44.3	44.3	58.4
Total overnight capital cost (million $)					256	
Total overnight capital cost per unit capacity ($/kW$_e$)					868	

Notes to Table 12:

[a] The base cost and capacity and the scaling exponent are taken from DOE/EPRI.[36] Only one equipment train is used, so the unit cost, *C*, equals the train cost C_m, for all units.

[b] The GDP implicit price deflator[39] has been used to convert to constant 2003 dollars from other year dollars.

[c] See Table 9, note c.

[d] See Table 9, note d.

[e] See Table 9, note e.

[f] Overnight cost includes indirect costs (at 32% of direct costs). Balance of plant is included as a separate unit.

[g] Fuel preparation cost for a 184 MW$_e$ plant is given as $93/kW (1996 $), or $17.1 million, by DOE/EPRI.[36] Converted to 2003$, this is $19.3 million.

[h] Boiler cost for a 184 MW$_e$ plant is given as $229/kW (1996 $), or $42.1 million, by DOE/EPRI.[36] Converted to 2003$, this is $47.4 million.

[i] Baghouse and cooling tower cost for a 184 MW$_e$ plant is given as $15/kW (1996 $), or $2.8 million, by DOE/EPRI.[36] Converted to 2003$, this is $3.11 million.

[j] Boiler feed-water treatment and deaerator cost for a 184 MW$_e$ plant is given as $29/kW (1996 $), or $5.3 million, by DOE/EPRI.[36] Converted to 2003$, this is $6.01 million.

[k] Steam turbine cost for a 184 MW$_e$ plant is given as $76/kW (1996 $), or $14.0 million, by DOE/EPRI.[36] Converted to 2003$, this is $15.74 million.

[l] Cooling water system cost for a 184 MW$_e$ plant is given as $34/kW (1996 $), or $6.3 million, by DOE/EPRI.[36] Converted to 2003$, this is $7.04 million.

[m] Balance of plant cost for a 184 MW$_e$ plant is given as $141/kW (1996$), or $25.9 million, by DOE/EPRI.[36] Converted to 2003$, this is $29.2 million.

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Discounted cash flow rate of return analysis

We combine our estimated capital costs with O&M and biomass costs to estimate the plant-gate cost of producing power as a function of the required real internal rate of return on equity for plants consuming 4536 dry metric tons per day of switchgrass. We use the discounted cash flow rate of return (DCFROR) methodology described in Laser *et al.*[41] As discussed there, we use an annual average capacity factor of 80%, a value appropriate for plants designed with no spare equipment capacity.[****]

Table 13. Financial parameter assumptions.

Debt fraction	40%
Equity fraction	60%
Interest rate on debt, %/year	7.5%
Baseline internal rate of return on equity, %/year	12%
Federal + state taxes	39%
Property taxes and insurance	0%
Economic lifetime, years	25
Depreciation method	20-yr MACRS[a]
Construction time, years	3
Operating capacity factor, *CF*	80%
Annual cost maintenance as % of TOC, m_a	4%
Alternative biomass prices, P_b ($\$/GJ_{LHV}$)	2, 3, 4

[a] Modified Accelerated Cost Recovery System (MACRS) is a property depreciation system defined by the Internal Revenue Service that applies to assets placed in service after 1986. It results in more rapid depreciation than straight-line depreciation.

B-IGCC costs

For a rate of return on equity of 12% per year (real) and other financial parameter values given in Table 13, the plant-gate costs of electricity from the two B-IGCC designs are similar, ranging from 4.4 to 4.7 ¢/kWh when biomass costs $\$2/GJ_{LHV}$ and 5.9 to 6.3 ¢/kWh when biomass costs $\$4/GJ_{LHV}$ (Table 14). The steam-Rankine system produces power at a cost of 4.9 and 7.1 ¢/kWh for biomass prices of $\$2/GJ_{LHV}$ and $\$4/GJ_{LHV}$, respectively. The efficiency disadvantage for the steam-Rankine system leads to the larger difference in generating cost compared to the B-IGCC when the biomass price is higher.

Figure 8 shows, for two different switchgrass purchase prices, the internal rate of return (IRR) on equity as a function of various plant-gate electricity sale prices for the two B-IGCC systems and the steam-Rankine system. With the lower biomass price ($\$2/GJ_{LHV}$) the spread in IRR among the different technologies is relatively small, though the oxygen-blown B-IGCC is the most attractive option until electricity sale price exceeds 10 ¢/kWh. With the higher biomass price, there is a wider spread in IRRs among the

**** Capacity factor is the fraction of time during a year that a plant is operating at the equivalent of its rated full capacity.

Table 14. Levelized cost of electricity generation at the reference plant scale (4536 metric tons/day switchgrass input, dry basis) for B-IGCC and steam-Rankine systems. [a]

	B-IGCC						Steam Rankine		
	Pressurized O_2			Indirectly heated					
Capital cost, $/kW		967			1059			868	
O & M cost, $/kW·yr		38.7			42.4			34.7	
Efficiency, LHV %		49.5%			48.2%			33.0%	
Heat rate, $MJ_{th,LHV}/kWh_e$		7.27			7.47			10.9	
Switchgrass price, $\$/GJ_{LHV}$	2	3	4	2	3	4	2	3	4
Electricity generating cost, $/kWh									
Capital charges		0.024			0.027			0.022	
O & M charges		0.006			0.006			0.005	
Biomass charges	0.015	0.022	0.029	0.015	0.022	0.030	0.022	0.033	0.044
Total cost	0.044	0.052	0.059	0.047	0.052	0.059	0.049	0.060	0.071

[a] Calculated for a real internal rate of return on equity of 12%/year and other financial parameter values given in Table 13

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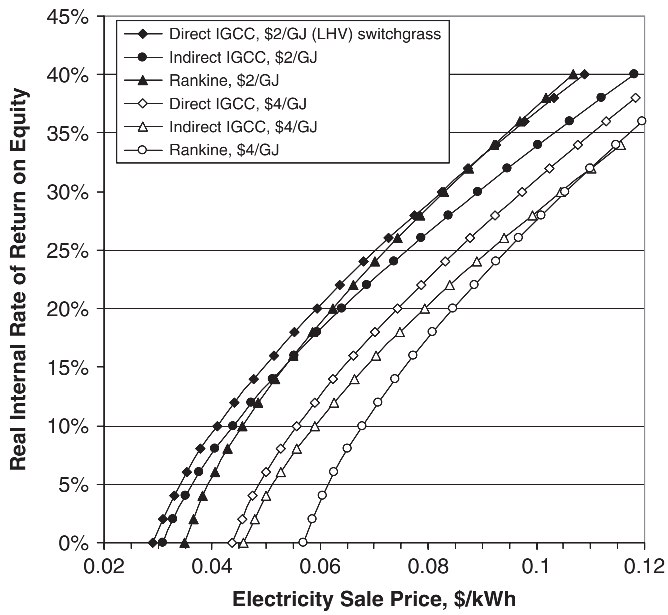

Figure 8. Internal rate of return on equity as a function of electricity sale price for three switchgrass power-generating systems and for two switchgrass purchase prices. The switchgrass prices correspond to $33.9/dry metric ton ($2/GJLHV) and $67.9/dry metric ton ($4/GJLHV).

three technologies, and the oxygen-blown B-IGCC is the most attractive option over the entire electricity price range shown.

B-IGSOFC costs

Because of inherently greater uncertainty in cost projections for B-IGSOFC systems, we have chosen a different approach to the cost analysis than that used with the B-IGCC systems. The key elements of the B-IGSOFC with uncertain costs are the SOFC itself and the high-temperature UCDC sulfur-removal process. Our B-IGSOFC cost analysis seeks to identify a capital cost target that the SOFC and UCDC units must meet if a B-IGSOFC system is to yield the same IRR as estimated for a B-IGCC system. We use the oxygen-blown B-IGCC for this comparison, both because it is based on the same gasifier design as in the B-IGSOFC and because it gives the more attractive financial performance of the two B-IGCC systems analyzed.

Table 15 shows our partial capital cost estimate, developed with the same framework and assumptions as the B-IGCC estimates, for the B-IGSOFC design. This estimate excludes costs for the SOFC and UCDC. With this partial capital cost, the cost for the SOFC plus UCDC subcomponents cannot exceed $58 million if the B-IGSOFC is to be able to sell electricity at the same price as it could be sold from a B-IGCC system, with both systems achieving an IRR of 12% and both systems purchasing biomass for $\$2/GJ_{LHV}$ (Table 16). For a biomass price of $\$4/GJ_{LHV}$, the SOFC plus UCDC can have a higher capital cost ($66 million, Table 16), since the efficiency advantage of the B-IGSOFC grows in importance with biomass price.

The maximum allowable capital cost for the SOFC plus UCDC sub-systems corresponds to $175 to $200 per kW of installed SOFC electricity generating capacity (Table 16). This cost range can be compared with cost estimates found in the literature. One estimate of the cost range for a UCDC system alone (excluding the SOFC) at the scale of plant considered here is $\$25$ - $102/kW_{e,SOFC}$.[††††] Considering allowable costs of $\$175–\$200/kW_{e,SOFC}$ for the SOFC plus UCDC, the corresponding allowable cost range for the SOFC alone is $\$73/kW_{e,SOFC}$ to $\$175/kW_{e,SOFC}$. This cost range can be compared against a National Research Council (NRC) estimate that 250 kW SOFCs today cost $3500/kW and could reach $1800/kW by 2020.[43] If scaled up to 330 MW_e (the SOFC capacity in our analysis) by assuming, simplistically, a 0.7 scaling exponent, the NRC estimate of $1800/kW would fall to $208/kW. (The NRC report also quotes a US Department of Energy cost target for 2010 of $400/kW for SOFC technology of unspecified capacity.) Thus, it seems

[††††] Newby *et al*.,[33] who are involved in developing UCDC technology, indicate a cost of $27 million (in 2001$), or $28 million (2003$), for a hot-gas clean-up process that includes syngas cooling, bulk desulfurization, and a UCDC unit. This is designed for a coal gasification system producing 81,140 kg/hr of raw gas that is cleaned for an SOFC application. Our B-IGSOFC design utilizes two gas clean-up trains, each processing 180,000 kg/hr raw gas. Scaling the Newby *et al*. cost using a scaling exponent of 0.7 yields a cost of $49 million per train, and a dual-train cost of $91 million (assuming 0.9 scale factor for multiple trains, as the capital cost estimating methodology used for all systems in this paper). Per kW of SOFC output, this translates to 275 $\$/kW_{e,SOFC}$. Subtracting from this the cost of a syngas cooler (which, from Table 15, is 173 $\$/kW_{e,SOFC}$) and assuming that bulk desulfurization represents half the remaining cost (since it is one of the three major vessels in the process), the resulting cost of the UCDC unit would be about 51 $/kW. This is a highly uncertain estimate. Considering an uncertainty of -50%/+100% gives an UCDC system cost of $\$25–\$102/kW_{e,SOFC}$.

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Table 15. Partial capital cost estimate for the B-IGSOFC with pressurized, oxygen-blown gasifier having performance characteristics indicated in Table 6.

		Capacities (in indicated units)				Costs (in million 2003 $)		
		Required capacity	Number of units	Capacity per unit	Unit of Capacity	Cost per unit	Train cost	Overnight Cost
Plant Area	Sub-Unit	S_r	n[a]	S[b]		C[c]	C_m[d]	OC[e]
Gasifier island[f]	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Pressurized O_2 gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash cyclone	16.3	2	8.1	m^3/s gas feed	0.21	0.38	0.61
Gas clean-up	Tar cracker[g]	16.8	2	8.4	m^3/s gas feed	0.32	0.59	0.94
	Syngas cooler[h]	98	2	49	MW_{th} heat duty	19.3	36.0	57.4
	Ultra-Clean process[i]	??	??	??		??	??	??
ASU	Stand-alone ASU[j]	250.0	1	250.0	tonne/hr pure O_2	64.2	64.2	81.6
	O_2 compressor[k]	27.2	1	27.2	MW_e consumed	10.8	10.8	13.8
	O_2 boost compressor[k]	0.5	1	0.5	MW_e consumed	0.7	0.7	0.9
	N_2 compressor[k]	2.52	1	2.52	MW_e consumed	1.64	1.64	2.09
Power island	SOFC[i]	??	??	??		??	??	??
	Syngas expander[l]	102	1	102	MW_e produced	11.6	11.6	14.8
	HRSG + heat exchangers[m]	149	1	149	MW_{th} heat duty	56.6	56.6	87.0
	Steam turbine[m]	149	1	149	ST gross MW_e	24.2	24.2	30.7
Partial overnight cost (million $)[n]							294	

[a] Same as note (a) in Table 9.

[b] Same as note (b) in Table 9.

[c] Same as note (c) in Table 9.

[d] Same as note (d) in Table 9.

[e] Same as note (e) in Table 9.

[f] Same gasifier island cost as for B-IGCC using pressurized, oxygen-blown gasifier (Table 9).

[g] The base cost (C_o) for a tar cracker (Table 8) has been multiplied by a pressure factor of 1.45[42] to obtain the base cost for a pressurized (30 bar) tar cracker of $1.06 million for the base scale shown in Table 8. Although the tar cracker in Table 8 carries out reforming-type reactions to crack tars, the cost is assumed to be the same for the partial-oxidation type tar cracker used in the B-IGSOFC design. The base cost is scaled with capacity using scaling exponent of 0.7 to obtain the cost per unit shown in this table.

[h] Same basis for syngas cooler cost as for B-IGCCs (Table 8).

[i] The Ultra-Clean syngas polishing process and the SOFC are at too early a stage of development to predict costs with as much confidence as for other components.

[j] ASU cost is scaled from base cost (C_o) for a stand-alone ASU of $35.6 million (2003$), with base capacity of 76.6 t/hr of pure O_2, and using scaling exponent, f, of 0.5. See note (g) of Table 8.

Table 15. *Continued*

[k] See note (h) in Table 8 for explanation of compressor cost estimates.

[l] Kreutz *et al*.[40] indicate a cost of $3.14M (2002 dollars) for a syngas expander. Both the SOFC purge syngas in this study and the syngas in Kreutz et al. are low-heating-value gases consisting mainly of CO, CO_2, H_2O, and H_2. These costs include installation, BOP, engineering and contingency, and are given for an Nth plant design. In order to achieve a consistent basis in accounting for indirect costs, engineering and contingency costs were removed from the costs reported in Kreutz *et al*. to obtain C_o. In Kreutz et al. the TDC is the sum of the installed equipment cost and BOP. For syngas expanders, engineering is 15% of TDC, and contingency is 15% of TDC + engineering. Therefore, to obtain C_o , the cost from Kreutz *et al*. is divided by (1.15x1.15). Thus BOP is included in C_o for compressors, while indirect costs are not. Values of S_o and *f* are also taken from Kreutz *et al*.

[m] Same cost basis as for B-IGCCs (Table 8). The purge gas combustor present in the B-IGSOFC design is a minor cost item, and its cost is assumed to be included as part of the BOP costs for power island components.

[n] Excludes costs for the SOFC and Ultra-Clean process.

likely that electricity generated at large-scale by a B-IGSOFC system like the one designed here will be more costly than electricity generated with a B-IGCC system at the same large scale.

Conclusions

Table 17 summarizes the design-point performance simulation results and cost results for all the biomass power-generating systems evaluated in this paper.

The B-IGSOFC gives the highest calculated electricity-generating efficiency, followed closely by the oxygen-blown B-IGCC and the indirectly heated B-IGCC. The steam-Rankine system is considerably less efficient. A number of other analysts have estimated design-point performance of B-IGCC systems. Our B-IGCC efficiencies are somewhat higher than results given in most other studies in the literature. The main reasons for this are that, unlike most systems considered in the literature, our efficiencies are calculated for systems that i) use 20% moisture content biomass as the fuel and no pre-gasification drying, as compared to 50% moisture content feedstocks used in most other studies; ii) assume 100% carbon conversion in the gasifier and 100% cracking to light gases of the tars generated during gasification (which avoids discarding any of the energy contained in the tars), as compared to incomplete tar cracking assumed in many other studies ; iii) assume use of the best performing gas turbines on the market today, as permitted by the large-scale of the installations; and iv) assume tight process heat integration, enabling maximum heat recovery that boosts electricity generation.

Regarding electricity generating costs, the B-IGCC systems are estimated to be able to sell electricity at a lower cost (for the same IRR) as the steam-Rankine system. The higher generating efficiencies for the B-IGCC plants (leading to lower biomass costs per kWh generated) more

Table 16. Maximum allowable capital costs for SOFC + UCDC sub-system in a B-IGSOFC having performance characteristics indicated in Table 6, if the B-IGSOFC is to produce electricity at the same cost as estimated for electricity from a B-IGCC (using oxygen-blown gasifier).[a]

Biomass Price	B-IGCC Electricity Cost	Allowable Overnight Capital Cost For SOFC + Ultra Clean Sub-Systems		Resulting Overnight Cost for B-IGSOFC
$\$/GJ_{LHV}$	$/kWh	million $	$\$/kW_{e,SOFC}$	million $
2	0.044	58	175	459
3	0.052	62	188	463
4	0.059	66	200	467

[a] Financial parameter assumptions as indicated in Table 13.

Table 17. Summary of performance and installed overnight capital cost estimates for the biomass power systems evaluated in this paper.

	B-IGCC	B-IGCC	B-IGSOFC	Rankine
Gasifier design >>>>>	Indirectly heated	Pressurized, oxygen-blown		(no gasifier)
Biomass input				
As-received metric tonnes/day	5,670	5,670	5,670	5,670
MW_{th} (lower heating value)	983	983	983	983
MWt_h (higher heating value)	893	893	893	893
Electricity output				
Net production, MW_e	431	442	463	295
Net efficiency, % (LHV basis)	43.8	45.0	47.1	30.0
Net efficiency, % (HHV basis)	48.2	49.5	51.8	33.0
Installed overnight capital cost				
Total, million 2003$	428	456	not estimated	256
Specific cost, $/kW$_e$ (2003$)	968	1059		868
Electricity generation (with financial parameters in Table 13 & $3/GJ$_{LHV}$ biomass)				
Total cost, ¢/kWh	5.5	5.2	–	6.0

than compensate for the modestly higher specific capital costs of the B-IGCC systems. Uncertainties regarding prospective mature-technology costs for SOFC and hot-gas sulfur clean-up technologies assumed for the B-IGSOFC performance analysis make it difficult to evaluate the prospective electricity generating costs for B-IGSOFC relative to B-IGCC, but the rough analysis developed in this work suggests that it is unlikely that B-IGSOFC will ultimately show improved economics relative to B-IGCC at the large scales considered here.

Final note

The work reported in this paper was submitted for publication three years after it was completed in order to appear with other papers comprising the RBAEF study. During this time, one of the authors (EDL) has been involved with major new analyses at Princeton University that have led to new process simulation results and updated capital cost estimates for some of the process designs described in this paper. (Some of this new work has been reported by Kreutz *et al.*[61]) The new Princeton results reflect technology insights and updated and improved data sources for capital cost estimates compared with those reported in this paper. In general the new Princeton results indicate lower energy conversion efficiencies than those reported in this paper, and higher capital costs. For example, for the power generating system shown in Fig. 2 in this paper, the new Princeton analysis predicts a net higher heating value electricity generating efficiency of 43% (starting with 15% moisture content switchgrass) compared with 45% (starting with 20% moisture content biomass) reported in this paper. The new Princeton analysis estimates an overnight installed capital cost of $1570/kW (in 2007$), compared with $968/kW (in 2003$) reported in this paper. Escalating the latter to 2007$ using the *Chemical Engineering Magazine* Plant Cost Index yields a cost of $1270/kW.

Because the new Princeton analysis did not re-examine all of the process designs described in this paper and because some of the other papers in this issue rely on the earlier results, we have chosen to report these earlier results in this paper. While the newer analyses project somewhat higher costs and lower efficiencies, the broad conclusions regarding relative performance and cost among different technology options is not likely to be significantly different from those reported here.

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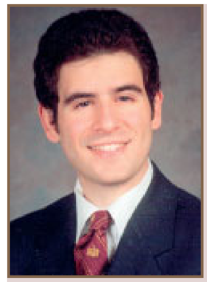
Haiming Jin

Jin is a senior process engineer at SNC-Lavalin, Houston, Texas focusing on coal, natural gas, and biomass conversion to liquid fuels both thermochemically and biologically. A former consultant working on petcoke conversion to methanol and electric power, Jin also has experience in carbon capture and sequestration and has been involved in several CCS engineering design and studies. Jin also spent a few years doing research on cellulosic biomass conversion to ethanol and thermochemical fuels at Princeton University and Dartmouth College.



Fuat Celik

Fuat Celik is currently pursuing a PhD in chemical engineering at the University of California, Berkeley. His research in the field of catalysis involves the investigation of acid-catalyzed Koch carbonylation to form carbon-carbon bonds from carbon monoxide and small organic molecules using solid superacids. He previously worked at the Princeton Environmental Institute on the design, performance, and cost analysis of large-scale systems for converting energy crops and coal to electricity and/or liquid fuels.

Eric D. Larson

Larson is a senior member of the Energy Systems Analysis Group within the Princeton Environmental Institute and an affiliated faculty member in Princeton's Science, Technology, and Environmental Policy Program. Research interests include engineering, economic, and policy-related assessments of advanced clean-energy systems, especially for electric power and transport fuels production from carbonaceous fuels (biomass, coal, natural gas) and for efficient end use of energy. He was the task leader for thermochemical conversion technologies in the RBAEF project.