

SUPPORTING INFORMATION to

Large-Scale Gasification-Based Co-Production of Fuels and Electricity from Switchgrass

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1 Introduction

This annex includes an overview of the current global status of the production of the synthetic fuels examined in the journal paper, namely Fischer-Tropsch fuels (diesel and gasoline blendstocks), dimethyl ether (DME), and hydrogen. Also included here is a detailed description and simulation results for the synthetic fuels production processes summarized in the paper. Finally, this annex also includes details regarding the methodology and data sources for the development of capital cost estimates for the production processes included in our analysis.

2 Synthetic Fuels Included in the Analysis

2.1 Fischer-Tropsch Fuels

The product of Fischer-Tropsch (FT) synthesis is a mixture of straight-chain hydrocarbons (olefins and paraffins) that can be refined into “clean diesel” and naphtha fractions, the latter of which can be upgraded to a gasoline blendstock. FT fuels were first produced commercially in the 1930s when Germany started production from coal syngas as vehicle fuel (Dry, 2002). Subsequently a coal-to-fuels program was started in South Africa and has been operating there since the early 1950s. Starting in the 1990s, there has been renewed interest globally in F-T synthesis to produce liquids from large reserves of remote “stranded” natural gas that have little or no value because of their distance from markets (Oukaci, 2002; Rahmim, 2003). Of particular interest today is the production of middle distillate fuels (diesel-like fuels) with unusually high cetane numbers and containing little or no sulfur or aromatics. Such fuels (derived by natural gas F-T conversion) are now beginning to be blended with conventional diesel fuels in some

countries, to meet increasingly strict vehicle fuel specifications designed to reduce tailpipe emissions.

Major expansion in global capacity for FTL production is occurring. In addition to Shell's gas-to-liquids (GTL, used synonymously with gas-to-FT liquids) plant in Malaysia (14,500 barrels per day FTL) and the PetroSA (formerly Mossgas) plant in South Africa (23,000 bpd) that started up in 1993, there are additional large commercial GTL facilities nearing startup or at advanced planning stages, including:

- 34,000 barrels per day (bpd) plant project of Qatar Petroleum that will use Sasol FT synthesis technology and is slated to come on line in 2006.
- 66,000 bpd expansion of the Qatar Petroleum project to startup in 2009.
- 34,000 bpd Chevron project in Nigeria, also using Sasol FT technology, expected on line in 2009.
- 30,000 bpd BP project in Colombia using BP's FT synthesis technology to come on line in 2011.
- 36,000 bpd project in Algeria to come on line in 2011.
- 140,000 bpd Shell project in Qatar using Shell's FT technology; to come on line in two phases in 2009 and 2011.
- 154,000 bpd ExxonMobil project in Qatar using ExxonMobil FT technology; to come on line in 2011

There is also a growing resurgence of interest in F-T fuels from gasified coal. Coal-based FT fuel production was commercialized beginning with the Sasol I, II, and III plants (175,000 b/d

total capacity) built between 1956 and 1982 in South Africa. (Sasol I is now retired). The U.S. Department of Energy is cost-sharing a \$0.6 billion demonstration project in Gilberton, Pennsylvania, that will make from coal wastes 5,000 bpd of F-T liquids and 41 MW_e of electricity. Also, there are proposals for 33,000 bpd and 57,000 bpd facilities for FT fuels production from coal in the state of Wyoming. China's first commercial coal-FT project is under construction in Inner Mongolia. The plant is slated to produce 20,000 bpd when it comes on line in 2007. China has also signed a letter of intent with Sasol for two coal-FT plants that will produce together 120,000 bpd.

The process for converting biomass into F-T liquids is similar in many respects to that for converting coal. Preliminary technical/economic analyses on biomass conversion was published by Larson and Jin (1999a and 1999b). More recently, there have been several detailed technical and economic assessments published (Bechtel, 1998; Tijmensen, 2000; Tijmensen *et al.*, 2002; Hamelinck *et al.*, 2003; Boerrigter and van der Drift, 2003; Hamelinck *et al.*, 2004). There is considerable current interest in Europe in production of FT fuels from biomass, motivated by large financial incentives. For example, in the UK a 20 pence per liter (\$1.40/gal) incentive for biomass-derived diesel fuel has been in place since July 2002. Incentives are also in place in Germany, Spain, and Sweden. Such incentives have been introduced in part as a result of European Union Directive 2003/30/EC, which recommends that all member states have 2% of all petrol and diesel consumption (on an energy basis) be from biofuels or other renewable fuels by the end of 2005, reaching 5.75% by the end of 2010. The Shell Oil Company, which offers one of the leading commercial entrained-flow coal gasifiers, recently announced a partnership with Choren, a German company with a biomass gasification system, with plans for constructing a commercial biomass to FT liquids facility in Germany (Shell, 2005).

2.2 Dimethyl Ether

Dimethyl ether (DME) is a colorless gas at ambient temperature and pressure, with a slight ethereal odor. It is relatively inert, non-corrosive, non-carcinogenic, almost non-toxic, and does not form peroxides by prolonged exposure to air (Hansen *et al.*, 1995). Today, DME is used primarily as an aerosol propellant in hair sprays and other personal care products, but its physical properties (Table 2) make it a suitable substitute (or blending agent) for liquefied petroleum gas (LPG). It is also an excellent diesel engine fuel due to its high cetane number and absence of soot production during combustion.

DME is produced globally today at a rate of only about 150,000 tons per year (Naqvi, 2002), but this production level will increase dramatically soon. In 2006 a DME plant with production capacity of 110,000 t/yr of DME from natural gas will start up in Sichuan Province, China, and a gas-to-DME facility producing 800,000 tons per year will come on line in Iran (Halder Topsoe, 2004). Most of the DME made at these facilities will be used as an LPG substitute. There is also discussion of a facility to be built in Australia to produce between one and two million tonnes per year of DME from natural gas, and China's State Development Planning Commission has approved plans for the first large-scale coal-to-DME project, to be located in Ningxia province (Lucas, 2002). The first phase would have a capacity of 210,000 tons per year, and the second phase would have a capacity of 630,000 tons per year. Construction has not yet started on this plant. Other DME projects are also under development in China.

The leading commercial developer of fixed-bed DME synthesis reactor designs is Halder-Topsoe.¹ Mobil and Snamprogetti S.p.A. hold patents for DME synthesis processes (Zahner, 1977; Pagani, 1978), but at present are not pursuing commercial development of the technology. Leading private developers of slurry-bed DME synthesis reactors (a technology of focus in our analysis presented later) are Air Products and Chemicals, Inc. (APCI) (Lewnard *et al.*, 1990; Brown *et al.*, 1991; Lewnard *et al.*, 1993; Air Products and Chemicals, 1993; Peng *et al.*, 1999a; Peng *et al.*, 1999b) and the NKK Corporation (Adachi *et al.*, 2000; Fujimoto *et al.*, 1995). The Institute of Coal Chemistry of the Chinese Academy of Sciences (CAS) (Niu, 2000) has also been developing slurry-phase DME synthesis technology since 1995. The CAS Institute of Chemical Physics (Dalian) has done some work on fixed-bed DME synthesis technology (Xu *et al.*, 2001).

The DME reactor design of APCI is derived from its liquid-phase methanol (LPMEOH) synthesis process that was developed in the 1980s. A commercial-scale LPMEOH demonstration plant (250 tonnes per day methanol capacity) has been operating since 1997 with gas produced by the Eastman Chemical Company's coal gasification facility in Kingsport, Tennessee (Eastman Chemical and Air Products and Chemicals, 2003). The construction of this facility was preceded by extensive testing in a 10 tpd capacity process development unit (PDU) in LaPorte, Texas. The PDU was operated in 1999 to generate test data on direct DME synthesis (Air Products and Chemicals, 2001 and 2002).

¹ The fixed-bed design of Halder-Topsoe includes three stages of synthesis reactors with cooling between each stage and recycle of unconverted syngas (Hansen *et al.*, 1995). The patent for this process specifies a feed gas CO concentration of less than 10% and a recycle volume of unconverted syngas ranging from 93% to 98% of the total unconverted syngas (Voss *et al.*, 1999). The fraction of CO converted on a single pass through each reactor stage (assuming a three-stage intercooled set of reactors) ranges from 16% to 34%, depending on the H₂/CO ratio.

DME Development, Inc., a Japanese consortium of nine companies led by NKK and Nippon Sanso, is currently in the construction phase of a 100 tpd DME slurry-phase reactor in Kushiro, Hokkaido. This effort builds on initial testing of a five ton per day capacity reactor that was completed in 1999 by NKK (2003), who prior to that (with support from the Japanese Ministry of International Trade and Industry) worked with the Taiheiyo Coal Mining Co., Sumitomo Metal Industries, and Japan's Center for Coal Utilization to develop the DME slurry reactor technology, with coal applications in mind.

2.3 Hydrogen

Technology for production of hydrogen from syngas is well established commercially (Chiesa *et al.*, 2005; Kreutz *et al.*, 2005), with applications found most commonly in petroleum refining and ammonia production. Globally, natural gas is the most commonly used feedstock to make syngas for hydrogen production, but in China the predominant feedstock is coal.

3 Detailed Process Descriptions and Mass/Energy Balances

In this section we present detailed descriptions of the five process designs developed in this work. Table 1 (reproduced from the paper) summarizes the process designs and gives short-hand identifying names used elsewhere in this annex.

3.1 Producing Clean Synthesis Gas

The design of the upstream section (for producing synthesis gas) is largely the same for all of our five plants. Switchgrass is transported from a short-term storage area to the feed preparation area, where it is chopped for feeding to the gasifier. 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 (Lau et al., 2003). Carbon dioxide available from the downstream Rectisol unit used to pressurize the biomass feeding system.

The gasifier in all cases is a pressurized (29.9 bar) oxygen-blown fluidized bed reactor. We have modeled the gasifier based on empirical data for pilot-plant operation of the gasifier design of the Gas Technology Institute^{2,3} (Katofsky, 1993; OPPA, 1990). Switchgrass is injected near the bottom of the reactor together with 0.61 m³/s of oxygen (95% purity) and 0.90 m³/s of steam.⁴

The gasifier produces a mix of light combustible gases, unconverted carbon (char), ash, heavy hydrocarbons (condensable tars and oils), and small amounts of hydrogen sulfide, ammonia, alkali compounds, and other gaseous polar impurities. A cyclone separates the gas from entrained ash and unconverted char, and the latter is recirculated to the gasifier, where it is assumed to be gasified to extinction. Ash is removed from the bottom of the gasifier.

We simulate the cracking of tars and oils to light gases as a two-step process. In the first step, a small amount of oxygen is injected into the freeboard of the gasifier (above the bubbling fluidized bed) to promote cracking of the tars and oils to lighter molecules. The literature

² The license for the Gas Technology Institute (GTI) technology is currently owned by Carbona.

³ Since biomass gasification is a kinetically controlled process, and kinetic parameters 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 H₂, O₂, N₂, H₂O, S, and solid C, as well as ash. These components, together with some 98.5% pure nitrogen (used for feeder pressurization), some 95% pure oxygen, and some steam are fed to an RGIBBS reactor. The steam input 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, C₂₀H₃₀O₂) to be 1% by weight of the dry biomass, and we specify the following volume fractions in the product gas based on empirical data: CH₄ (8.2%), C₂H₄ (0.15%) and C₂H₆ (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 H₂/CO ratio to be 0.72 to match empirical data.

⁴ Unless otherwise noted, all volumetric gas flows are expressed in this paper in terms of actual volume (not at standard or normal conditions).

suggests that 90% conversion of tars and oils to CO and H₂ can be achieved by this oxygen injection (Pan et al., 1999). The heat released in these reactions raises the temperature of the raw syngas to about 1000°C as it leaves the gasifier. A fusion temperature for switchgrass ash (under oxidizing conditions) of 1016°C has been reported in the literature (McLaughlin et al., 1999). If this temperature is also representative for reducing conditions (gasification), then it may be necessary to introduce 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 (Steenari and Lindqvist, 1998; Ohman and Nordin, 2000).

After the gasifier we include an external catalytic tar cracker to ensure that all tars are converted to light gases. This reactor is assumed to operate adiabatically. The heat needed to drive the endothermic cracking reactions is drawn from the gas itself, resulting in an exit gas temperature of about 800°C.

The gas is subsequently cooled to 350°C in a syngas cooler, a vertical fire-tube design (with hot gases flowing inside the tubes) that minimizes deposition of small particles still in the gas, as well as of alkali species that condense during cooling. Subsequent particle removal by a barrier filter (ceramic or sintered metal) is carried out at 350°C. The resulting syngas has a molar composition of 18% CO, 31% H₂, 4% CH₄, 24% CO₂, and 22% H₂O. Our calculated “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 (or 79.1% on a higher heating value basis).

All five process designs that we have developed share all of the features described above. The processing of the syngas from this point on varies depending on the products being made.

3.2 Fischer-Tropsch Fuels Production

For production of Fischer-Tropsch fuels, the clean syngas is first cooled to 40°C in preparation for the acid-gas removal (AGR) area (Figure 1). AGR consists of a Rectisol™ unit that removes all H₂S and CO₂. H₂S must be removed to avoid poisoning of downstream catalysts.⁵ CO₂ removal is not essential for process reasons, but doing so provides for modestly higher synthesis rates and decreases downstream equipment sizes (and costs).

The cool, clean syngas (with molar composition 58% H₂, 33% CO, 8% CH₄, and less than 1% each of Ar and N₂...) is fed to a slurry-phase Fischer-Tropsch reactor, where the CO and H₂ combine exothermically in the presence of a catalyst under moderate pressure and temperature conditions (ranging from 20-35 bar and 180-350°C, depending on the specific design) to produce a mixture of straight-chain hydrocarbons, namely paraffins (C_nH_{2n+2}) and olefins (C_nH_{2n}), ranging from methane to high molecular weight waxes. The reactor is cooled by raising steam to maintain the desired reaction temperature. The output from the synthesis reactor includes a mix of different-length hydrocarbons, CO₂, H₂O, inert species in the feed gas, and other minor compounds.

The relative proportion of different hydrocarbons produced by the F-T reactions is determined primarily by feed gas composition, catalyst type and loading, reaction temperature, pressure, and residence time, the net result of which can be characterized simplistically in terms of the "alpha" number for a given reactor design and operation. This characterization derives from a single-parameter (α) model that can relatively accurately predict empirically-observed carbon number

⁵ The sulfur content of the switchgrass considered in this analysis is 0.1% by dry weight, which is high for biomass (but very low compared to most coal). The H₂S content of the raw gasifier product is about 100 ppm(v). The acceptable H₂S content in synthesis gas depends on the downstream catalyst to be used, but is generally in the 100 to 500 ppb range (Spath and Dayton, 2003).

distributions (Anderson, 1984). The model accounts for the fact that longer hydrocarbons are formed by the linear addition of $-\text{CH}_2-$ segments to shorter hydrocarbons in the synthesis process. The carbon number distribution predicted by this model is called the Schulz-Flory or Anderson-Flory-Schulz distribution⁶ (Eilers *et al.*, 1990). Commercial F-T synthesis reactors are characterized by $0.65 < \alpha < 0.95$. Lower α will give a lower average molecular weight synthesis product compared to higher- α synthesis.

In GTL plants operating today, the F-T synthesis reactor is typically operated at high-end α values using a cobalt-based catalyst. The resulting heavy waxy product is easily and with high selectivity formed into desired lighter products by subsequent hydrocracking, which involves the breaking up of the large hydrocarbon molecules into desired final products in a hydrogen-rich environment. Hydrocracking of large straight-chain hydrocarbons can be done under much less severe temperature conditions (350-400°C for cracking to C_5 - C_{18} range) than is required for hydrocracking of aromatic molecules in conventional petrochemical refining. Lighter hydrocarbons leaving the hydrocracker can be recycled for further conversion or burned to produce co-product electricity, e.g., using a gas turbine.

In our simulation, we use an iron-based catalyst because the H_2/CO ratio of the syngas produced by the gasifier (1.7) is less than the optimal value of about 2.2 for F-T synthesis, but iron-based catalysts promote the water gas shift reaction (converting some CO to H_2), bringing the effective H_2/CO ratio closer to the optimum for the F-T reactions.

⁶ The distribution of carbon number species (C_n) is given by $\log(C_n) = \log[(1-\alpha)/\alpha] + n\log(\alpha)$.

We have modeled the F-T synthesis reactor as a slurry-phase design based on the triple- α model of Fox and Tam (1995), as discussed in detail by Larson, *et al.* (2005). The simplicity of a single- α model⁶ is appealing, but multiple- α models can provide better matching to empirical data. Building on Fox and Tam's model, we developed our own multi- α model, drawing on more recent empirical results (than used by Fox and Tam) for the kinetics of F-T reactions over iron-based catalysts in slurry-bed reactors (Bukur *et al.*, 2004; Raje *et al.*, 1997).

Our assumed synthesis reactor conditions are 260°C and 22 bar pressure, with an average gas hourly space velocity of 5800 liters/kg_{catalyst}/hr (standard liters input syngas per hour per kilogram of catalyst). These assumptions, together with the syngas volume flowrate and density, enable calculation of the required catalyst mass, which in turn provides the basis for calculating the overall reaction results, with our multiple- α model.

The synthesis step produces a raw mix of products that must be refined to finished products. This refining can be done offsite (e.g., at a petroleum refinery) or onsite. We have chosen the latter option for our design. The design of our separation/upgrading area (Figure 2) is based on a Bechtel design (Bechtel, 1998). The liquid product streams leaving the F-T synthesis reactor are separated into light-gas, naphtha, distillate and heavy-wax fractions. A small amount of syngas is bypassed around the synthesis reactor and processed through a water gas shift reactor. Then, together with a hydrogen-rich stream from the naphtha reformer located in the upgrading area, the gas passes to a pressure-swing adsorption (PSA) unit to generate the hydrogen needed for product upgrading. The product upgrading area includes five major sub-units: (1) a wax hydrocracking plant cracks the raw waxes into naphtha, distillate and light gas; (2) a distillate-fraction hydrotreater and (3) a naphtha hydrotreater that stabilize these F-T fractions by

hydrogen addition (saturating the olefins). The distillate product is then ready for use, e.g., for blending with petroleum diesel; (4) a catalytic reforming area that treats the naphtha fractions (from the wax hydrocracker and the naphtha hydrotreater) and produces a high-octane gasoline blending component and a hydrogen-rich stream (sent to the PSA unit); and (5) an isomerization unit that increases the octane number of the pentane/hexane stream produced by the naphtha hydrotreater, producing additional high-grade gasoline blending component. Some light gases are unavoidably produced in each of the five sub-units, and these are collected and fed with unconverted syngas to a gas turbine combined cycle in the power island.

We have modeled the gas turbine in the power island on the most advanced generation of operating machines now available on the market (“F” technology), as discussed in Paper #xx (Jin, et al., 2006).

The hot gas turbine exhaust passes to a heat recovery steam generator (HRSG), wherein steam is raised to run a steam turbine. As noted earlier, where feasible the recovery of waste heat generated elsewhere in the process is integrated with the HRSG to augment steam generation for the steam turbine. Steam is generated at three different pressure levels (160, 21, and 3.5 bar), and the condenser operates at 0.05 bar.

Some air is taken from the gas turbine compressor exit to provide feed the air separation unit (ASU) generating the oxygen needed for gasification. The pressurized nitrogen available at the ASU is expanded through a free turbine to contribute to overall electricity generation from the plant.

Overall, our F-OT design converts 34% of the input biomass energy (LHV basis) into liquid fuels (62/38 diesel/gasoline mix) and 23% into exportable electricity. To facilitate comparisons with other fuels production designs, we calculate an “effective efficiency” of fuels production as the ratio of the liquid fuel energy produced divided by the following quantity: the total amount of biomass consumed less an amount of biomass that would be consumed in producing the same amount of power in a stand-alone biomass power plant.⁷ The effective efficiency in this case is slightly above 64%, the highest among the liquid fuel process designs we have developed here.

3.3 DME Production

For production of DME with once-through synthesis (D-OT, Figure 3), the design has many similarities to the F-OT design. The clean syngas is cooled to 40°C after which the AGR unit removes all H₂S and most of the CO₂, leaving a CO₂ volume fraction of 3% leaving the Rectisol unit. CO₂ removal increases reaction rates considerably in the DME synthesis reactor, since the presence of CO₂ acts to reduce the partial pressures of the main reacting gases. However, a small amount of CO₂ is necessary to ensure sufficient catalyst activity in a liquid-phase DME synthesis reactor (Larson and Ren, 2003), the reactor design we have assumed here.

Following AGR, the syngas (with molar composition 56% H₂, 32% CO, 7% CH₄, 3% CO₂, and less than 1% each of Ar and N₂) is compressed and heated in preparation for delivery to the DME synthesis reactor. The synthesis of DME (CH₃OCH₃) has similarities to synthesis of

⁷ For all cases, we assume an efficiency for stand-alone biomass power generation of 49.5%, based on the Jin, *et al.* (2006) results for a biomass integrated gasifier combined cycle operating on 20% moisture content switchgrass and using the same pressurized oxygen-blown fluidized bed gasifier as incorporated in the process designs described in this paper.

methanol (CH₃OH), the production technology for which is well-established commercially.⁸

Methanol synthesis is carried out over a catalyst, typically CuO/ZnO/Al₂O₃, and can be (simplistically) represented by the following principal reaction:



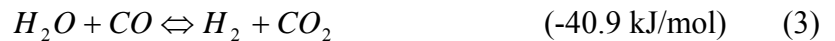
DME (CH₃OCH₃) is produced today exclusively in small-scale facilities by dehydration of methanol over a γ -alumina catalyst:



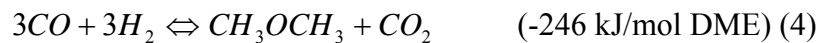
By combining some methanol catalyst and dehydration catalyst in the same reactor, reactions (1) and (2) proceed simultaneously, resulting in direct synthesis of DME. The idea of direct synthesis of DME from syngas was first reported in the literature long ago (Brown and Galloway, 1929), but efforts to commercialize direct synthesis technology did not begin in earnest until around 1990.

Syngas conversion to methanol (equation 1) can be accomplished today to nearly the extent predicted by chemical equilibrium, i.e., to the theoretical maximum extent. Substituting methanol dehydration catalyst for some of the methanol synthesis catalyst results in methanol being reacted away (by Equation 2) as it forms. This effectively by-passes the equilibrium limits of Equation 1. Methanol catalyst also promotes the water gas shift reaction:

⁸ Most commercial methanol is produced today from syngas derived from natural gas (except in China, where most domestic production of methanol – 3.3 million metric tons used in 2001) is via coal gasification (Larson and Ren, 2003).



The overall single-step DME synthesis chemistry can be represented as a combination of Equations 1, 2, and 3:



This reaction suggests that the optimum H₂:CO ratio in the feed gas for DME synthesis is 1:1. In practice, modest departures from this ratio (1.35 for the process configurations considered in this work), do not significantly change the synthesis yields.⁹

Our simulation of the reaction kinetics in a liquid phase DME synthesis reactor is based on a model developed by Larson and Ren (2003). That model is based on rate equations for methanol synthesis developed by Graaf [Graaf, et al. (1988a and 1988b) and Graaf and Beenackers (1996)] from laboratory measurements with a batch liquid-phase reactor and a CuO/ZnO/Al₃O₃ catalyst. Among the rate equations in the literature for which complete information is provided by authors, Graaf's equations appear to be relatively conservative in their prediction of the fractional conversion of CO to methanol. For the DME synthesis model, Larson and Ren added to these reactions a kinetic expression for methanol dehydration (over a γ -alumina catalyst) developed by Ng, *et al.* (1999). By appropriate selection of gas space velocity (6,000 lit/hr-kg_{MeOH cat}) and ratio of dehydration catalyst to methanol catalyst (0.3 was used), Larson and Ren obtained overall synthesis reactor performance predictions that compared well with the

⁹ Our sensitivity analyses indicate about a 10% higher DME production with H₂/CO of 1 instead of 1.7.

predictions of synthesis models developed internally at Air Products and Chemicals, Inc. The Air Products' models were based on typical lifecycle reactor performance, including an assumed average catalyst activity level of 50% of the level for fresh catalyst.

After synthesis, the raw gas-phase product leaving the liquid-phase reactor is sent to the downstream separation area, where a series of flash tanks and cryogenic distillation steps are used to produce separate streams of DME, by-product methanol, a CO₂-rich gas stream, and unconverted synthesis gas. After the initial flash separation steps, one stream containing mostly DME and methanol plus a small amount of other species undergoes distillation. A CO₂-rich stream comes off the top of this distillation tower, and a concentrated DME/methanol liquid mixture leaves at the bottom. The liquid mixture is distilled in a second tower, producing a 99.9% pure DME product stream at the top. The bottom methanol-rich stream passes to a third distillation tower where remaining water is separated out, leaving a pure stream of methanol exiting the top of the tower. A small amount of this methanol is used as make-up for the Rectisol plant. The remaining methanol is passed to a catalytic reactor, where 80% of it is converted to additional DME by dehydration, and the balance is recycled to the third distillation column.

The downstream DME separation area is based on equipment configurations proposed by Air Products and Chemicals, Inc. (Air Products and Chemicals, Inc., 1993), as implemented by Celik *et al.* (2004) in Aspen-Plus models of DME production from coal. Celik *et al.* updated Larson and Ren's model of the downstream area to improve DME separation effectiveness, and hence DME recovery. Among other modifications made by Celik *et al.* was the addition of the methanol dehydration reactor to convert methanol by-product to DME, rather than recycling the

methanol to the synthesis reactor. Separately dehydrating the methanol provides for somewhat improved overall DME yield.

The unconverted syngas remaining after once-through synthesis is rich in methane (since methane is essentially inert in the synthesis reactor). This gas passes through a saturator where it picks up moisture before reaching the gas turbine combustor. The saturator serves three purposes. It provides a use for low-grade waste heat generated elsewhere in the process, it adds mass flow to help maintain the mass balance between gas turbine compressor and expander, and its diluting effect contributes to keeping NO_x emissions below regulated levels.

One additional gas flow is sent to the power island. As in the F-OT design, the mixture of H₂S and CO₂ stripped from the gas at the Rectisol unit is compressed and delivered to the gas turbine combustor. This serves three purposes. It helps maintain the requisite mass balance between the gas turbine compressor and expander, its diluting effect ensures that NO_x emissions are kept below regulated levels, and by converting the small amount of H₂S in the gas into SO₂, (giving acceptably low concentrations in the exhaust stack) it obviates the need for any dedicated H₂S capture system. The portion of the acid gas mixture that is not compressed for gas turbine use is used to pressurize the biomass gasifier feeder.

Overall, the D-OT design converts 30% of the input biomass energy into DME and 24% into electricity, for a total efficiency of 54% and an effective efficiency of 62% (LHV basis).

It is informative to compare the results for the D-OT case with the parallel case F-OT. As noted earlier, the details of the upstream portion of the design are similar in both cases, including

gasification, gas cleanup, acid gas removal, and CO₂ delivery from the AGR unit to the gas turbine. However, there are some significant differences downstream. These are reflected in part in the ratio of fuels-to-electricity produced, which is considerably higher for the FT case (1.5) than for DME (0.8) as a result of the much higher one-pass conversion achieved with FT synthesis. Also, because of the greater exothermicity of the FT reactions as compared to DME reactions, waste heat recovery plays a more significant role in the FT design. Cooling the FT reactor (by steam raising) extracts 105 MW of thermal energy from the synthesis reactor, compared to 40 MW in the case of DME synthesis. The greater availability of waste heat from synthesis reactor cooling in the FT case contributes to more steam being generated to drive the steam turbine in the power island. The result is that the steam turbine accounts for 62% of the gross power generated in the FT case, while it accounts for only 49% in the DME case.

The second plant design for DME production (Figure 4) involves recycle of 97% of the unconverted synthesis gas from the product separation area to the synthesis reactor. (Even higher recycle percentages are possible in theory, but the not-insignificant level of methane in the syngas would lead to prohibitively large recycle volumes in that case.) There are two key design differences between the D-RC and D-OT designs. First, in the D-RC design all CO₂ in the syngas is removed at the Rectisol plant (rather than leaving 3% CO₂ in the syngas); because the post-Rectisol feed gas to the synthesis reactor is a mixture of fresh and recycled syngas (with molar composition of the mixture being 42.4% H₂, 35.3% CH₄, 10.4% CO, 4.6% N₂, 3.5% CO₂, and 3.3% Ar), the CO₂ needed for maintaining synthesis catalyst activity is provided by the recycled syngas. Second, in the D-RC design most of the unconverted syngas leaving the product separation area is compressed and recycled to the synthesis reactor. The lower flow of

un-recycled gas results in a smaller power island, but one that produces sufficient electricity to provide for on-site needs and a modest level of net exports .

The net result is that the D-RC design produces more than twice as much DME as the D-OT design (514 MW vs. 249 MW, HHV basis), but much less electricity (79 MW vs. 270 MW).

The effective efficiency for the D-RC design is 64% (LHV basis).

3.4 Hydrogen Production

Hydrogen production from biomass has been the subject of several detailed studies by others (e.g., Katofsky, 1993; Hamelinck and Faaij, 2002; and Lau *et al.*, 2003). Here we have carried out process design and simulation of hydrogen production using a consistent framework and set of assumptions as for DME and FT fuels analyses and also for the stand-alone power production results reported by Jin, *et al.* (2006). We have developed two basic process designs, one of which produces about as much hydrogen as electricity co-product (on an energy basis) and the second of which produces mostly hydrogen, but with enough electricity co-production to meet onsite needs and export a small amount.

In the design producing mostly hydrogen (H-MAX), the clean synthesis gas leaving the particle filter is passed directly to a two-stage water gas shift (WGS) reactor with sulfur-tolerant catalyst (Figure 5). The first stage of the WGS is operated adiabatically and the second is operated isothermally to convert as much of the CO as possible to CO₂, which correspondingly maximizes the H₂ content in the syngas. Steam is injected in the first stage. Following the WGS, all CO₂ and sulfur species are removed together at the Rectisol™ plant and passed to a downstream boiler for combustion (to fully oxidize the sulfur species). The CO₂-free gas leaving the Rectisol area

is passed to a pressure-swing adsorption (PSA) system that separates hydrogen (99.999% purity) from residual gas components. The hydrogen is compressed to 60 bar for storage or pipeline transport.

Typically, a single PSA unit will remove 70-90% of the hydrogen from a gas stream, with the highest removal possible when the initial H₂ concentration is high (Weist, 2005), as in our designs (90.4% H₂ in feed gas to PSA). We have modeled the PSA assuming an 87% one-pass H₂ recovery, but with recycle of some of the tail gas (containing residual H₂ plus other minor components) to achieve an overall H₂ capture of 95% of the available H₂. The heating value of the remaining tail gas is insufficient for gas turbine combustion, so it is passed to a boiler where it is burned to raise steam to drive a steam turbine generator. Since no gas turbine is present in this design, the air separation unit is a stand-alone design drawing in ambient air and venting nitrogen to the atmosphere.

The second hydrogen production configuration (H-5050, Figure 6) is designed such that approximately equal quantities (MW HHV) of electricity and hydrogen are produced. In this design, exactly half of the clean syngas from the particle filter bypasses the WGS area, going directly to a gas turbine in the power island. Also, the WGS consists of a single, adiabatically-operated reactor. CO₂ is removed following the WGS, a PSA system (with no purge gas recycle) purifies the remaining H₂-rich stream, and the H₂ is compressed for storage or transport. The PSA purge gas is mixed with the WGS bypass stream and passes through a saturator before being burned in a gas turbine to generate electricity. The ASU is integrated with the gas turbine. Steam raised using gas turbine exhaust and waste heat streams from elsewhere in the process powers a three-stage steam turbine to produce additional electricity.

The H-MAX configuration converts nearly 60% of the input biomass energy into hydrogen (LHV basis), with exportable electricity accounting for another 5% of the biomass energy input. Because of the modest amount of electricity generated, the effective efficiency (65%) is only modestly higher than the fraction of biomass energy converted to hydrogen. The H-5050 case produces much less hydrogen (32% of input biomass on LHV basis), but much more electricity (27%), and as a consequence yields the highest effective efficiency (74%) of any of the fuel production configurations we have examined.

4 Cost Analysis

4.1 Capital Cost Estimating Methodology

To develop our capital cost estimates, we first divided each plant into major process areas and sub-units within these areas and then identified from the literature or in consultation with experts installed costs for identical or similar equipment.¹⁰ We refer to these original estimates as the base cost (C_o) for that type of equipment. The base cost includes installation, but generally excludes balance of plant (BOP) and indirect costs (IC) associated with that piece of equipment. BOP includes instrumentation and controls, buildings, civil works, electrical connections, piping, insulation, and site preparation. Indirect costs include engineering and contingencies. Our approach for estimating BOP and IC are discussed below.

¹⁰ If the operating pressure of the reference equipment differed from the pressure used in this study, the reference cost was multiplied by a standard pressure factor to arrive at a new reference cost. Pressure multipliers for the cost of vessels and heat exchangers are from Guthrie (1969).

The base cost refers to a particular equipment size (capacity), S_o , which in most cases is different from the required size (S_r) determined from our process simulations. We scale the base cost (C_o) to estimate the cost of the equipment (C) at the scale of our design.¹¹

In some cases, there are maximum allowed equipment sizes (S_{max}). These might be determined by structural limits for the operation or construction of a unit, or by the size of equipment that can be practically transported, e.g., by truck or by rail, to the construction site. Maximum scales assumed here are for truck delivery in the U.S. If S_r exceeds S_{max} then multiple units must be installed. The number of units required (n) is calculated by rounding the right side of Equation 5 up to the nearest integer.

$$n = S_r / S_{max} \quad (5)$$

For our process designs where multiple gasifiers are required, we have assumed each active gasifier would have associated with it a separate feed preparation area and separate gas cleaning train (gasifier, tar cracker, cyclone, syngas cooler, and ceramic filter). We refer to each set of identical units in sequence surrounding the gasifier as a gasifier train.

The capacity of each piece of equipment in a train, S , is given by Equation 6.

$$S = S_r / n \quad (6)$$

¹¹ For vessels handling gases or liquids, the volumetric flow rate of the fluid defines the required size (assuming that residence time is constant or nearly so for similar units of different sizes). For solids handling, the solid mass feed rate defines the size. The size parameter of a compressor or pump is the amount of electricity consumed, and that of a heat exchanger is the amount of heat removed. The cost of power island equipment is assumed to scale with the electricity generating capacity.

The cost of a unit of size S is determined from the base cost and size for that unit:

$$C = C_o \cdot \left(\frac{S}{S_o} \right)^f \quad (7)$$

where f is a scaling exponent that ranges between zero and one in value, and is close to 0.67 for many types of equipment.¹²

For multiple unit trains of equal size, the installed cost of each additional train will be somewhat less than the cost of the first train, since typically multiple trains will share some auxiliary equipment, the labor required to install a second, third or fourth train is generally less than labor the required for the first train, and the special machining and shop costs to construct the first unit are not all duplicated to construct the second. We capture this idea using the trained cost, C_m , of a unit, given by Equation 8:

$$C_m = C \cdot n^m \quad (8)$$

where m is the scaling exponent for multiple trains. We assume a value for m of 0.9.

As noted earlier, for most of the equipment costs, the values of C_m were developed excluding BOP costs.¹³ In these cases, we estimated the BOP cost for each unit as a percentage of C_m .

¹² While the value of 0.67 turns out to apply, approximately, for many types of equipment, its origin is most apparent when considering a spherical reaction vessel. For such a vessel, the cost scales roughly with the amount of material needed to make the vessel, which in turn is closely related to its surface area: $\text{cost} \sim 4\pi r^2$. Thus, for two vessels with different capacities, the ratio of costs is $\text{cost}_1/\text{cost}_2 \sim (r_1/r_2)^2$. Since the capacity of a spherical vessel is proportional to its volume $(4/3)\pi r^3$, we can write $\text{capacity}_1/\text{capacity}_2 \sim (r_1/r_2)^3$ or rearranging, this gives $(r_1/r_2) \sim (\text{capacity}_1/\text{capacity}_2)^{1/3}$. Putting this latter expression into the cost ratio gives $\text{cost}_1/\text{cost}_2 \sim (\text{capacity}_1/\text{capacity}_2)^{2/3}$.

Hamelinck and Faaij (2001) have noted that the absolute cost of a power plant will grow more quickly with capacity than the BOP fraction of the cost, so BOP as a % of total cost will be smaller the larger the plant size. We confirmed this by a careful review of literature cost studies for similar plants at varying scales. We have estimated an overall BOP percentage as a function of the higher heating value biomass energy input from a best fit of several literature estimates (Katofsky 1993; DeLong 1995; Stone and Webster *et al.* 1995; Weyerhaeuser 2000; Hamelinck and Faaij 2001; and Kreutz *et al.* 2005):

$$\text{BOP (\%)} = 0.8867 / \{(\text{biomass MW}_{\text{th}})^{0.2096}\} \quad (9)$$

For the reference plant size in this project (4,536 tonne/day), the thermal input is 983 MW_{HHV}, so BOP is 20.9 % of the installed cost. (At one-tenth this scale, the figure would be 33.9%).

The sum of C_m and BOP gives the total direct costs (TDC).

Indirect costs (IC) are also incorporated into the analysis as percentages of TDC: 15% for engineering and head office, 5% for startup costs (including initial catalyst loadings), 1% for spares, and 1% for royalties. These values were obtained by comparing literature estimates for similar plants (Katofsky 1993; DeLong 1995; Stone and Webster *et al.* 1995; Mann and Spath 1997; Washington Group and Southern Company 2000; Weyerhaeuser 2000; Hamelinck *et al.* 2003; and Kreutz *et al.* 2005). Contingency costs are also included. Contingencies will vary with the level of experience with a technology and its manufacture and installation. For new and

¹³ When an original equipment cost included BOP, these costs were stripped out (where the original data sources were sufficiently disaggregated to do this) for the purpose of developing consistent values of C . In some cases, the BOP fraction of an original cost estimate could not be disaggregated. These cases have been noted in tables presented in the next section.

unproven technologies, contingencies could be 50% or more of TDC. Since we are considering future commercially mature systems, contingencies will be lower. We assign the power island and air separation unit equipment contingencies of 5%. The gasifier and gas cleanup areas involve inherently larger uncertainties associated with construction, so we assign a contingency of 10% for these areas. Thus, the total percentages of TDC used to calculate indirect costs were 32% for gasifier and gas cleanup islands and 27% for the ASU and power island areas. Table 3 summarizes the assumptions we used to estimate indirect costs.

Finally, we define the sum of TDC and IC as the total overnight capital cost (TOC) for the plant.

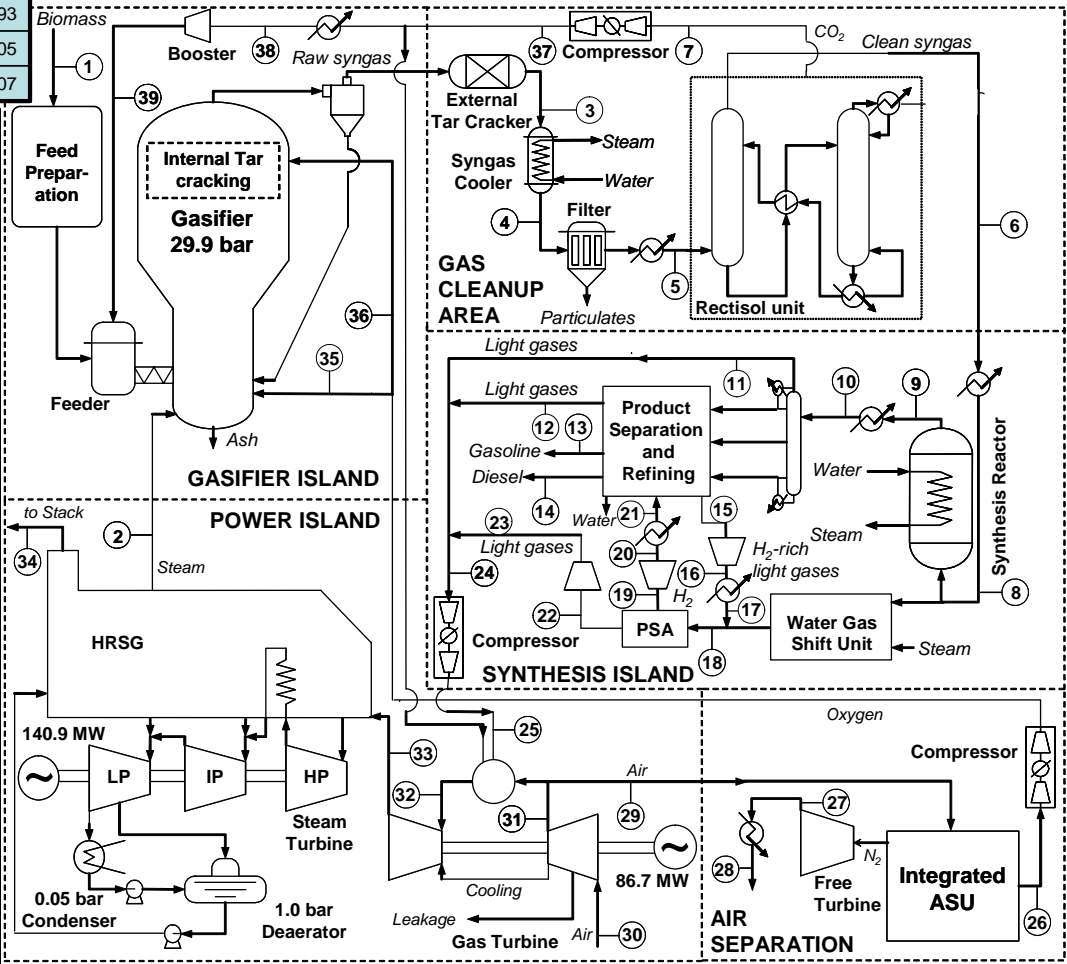
4.2 Capital Cost Estimates

Table 4 gives our reference values of S_o , S_{max} , C_o , and f used to develop estimates of TOC for each of the five fuels production plants described earlier. The costs for components upstream of synthesis or H₂ purification (i.e., the gasifier, ASU, WGS, etc.) and for the power island are consistent with the estimates used for these components in the cost estimates for stand-alone power reported by Jin, *et al.* (2006). Costs for the fuel synthesis and hydrogen purification areas are based primarily on Celik *et al.* (2004), Larson and Ren (2003), discussions with industry experts, and a few other sources, as detailed in the notes to Table 4.

Using the reference costs shown in Table 4, we have estimated TOC as described in Table 5, Table 6, Table 7, Table 8, and Table 9. It may be noted that these capital cost estimates assume no spare equipment capacity.

Switchgrass, MW _{LHV}	893
FT out, MW _{LHV}	305
Net electricity, MW	207

	T (°C)	p (bar)	m (kg/s)
1	25	1.01	65.6
2	236.6	31.65	13.8
3	800	28.82	99.7
4	350	28.25	99.7
5	40	26.3	99.7
6	26.3	23.53	30.9
7	25	1.8	50.2
8	245	23.06	30.9
9	260	22	30.6
10	40	21.56	30.6
11	40	3.5	11.7
12	45.8	3.23	0.6
13	58.6	2.58	2.8
14	131	1.4	4.4
15	40	7.35	0.2
16	184.1	21.05	0.2
17	39.9	20.63	0.2
18	39.7	20.63	0.6
19	39.7	20.13	0.1
20	-53.3	4.08	0.1
21	25	4	0.1
22	39.7	1.5	0.5
23	119.9	3.23	0.5
24	42.9	3.23	12.7
25	164.3	26.83	12.7
26	29.6	3.18	18.0
27	-34.3	1.1	39.1
28	25	1.08	39.1
29	495.4	19.76	76.7
30	25	1.01	305.8
31	495.4	19.76	305.8
32	1370	19.17	270.7
33	658.6	1.06	286.8
34	90	1.01	286.8
35	152.1	31.36	17.5
36	152.1	31.36	0.5
37	136.4	26.83	5.3
38	40	26.3	5.3
39	57.4	31.4	5.3



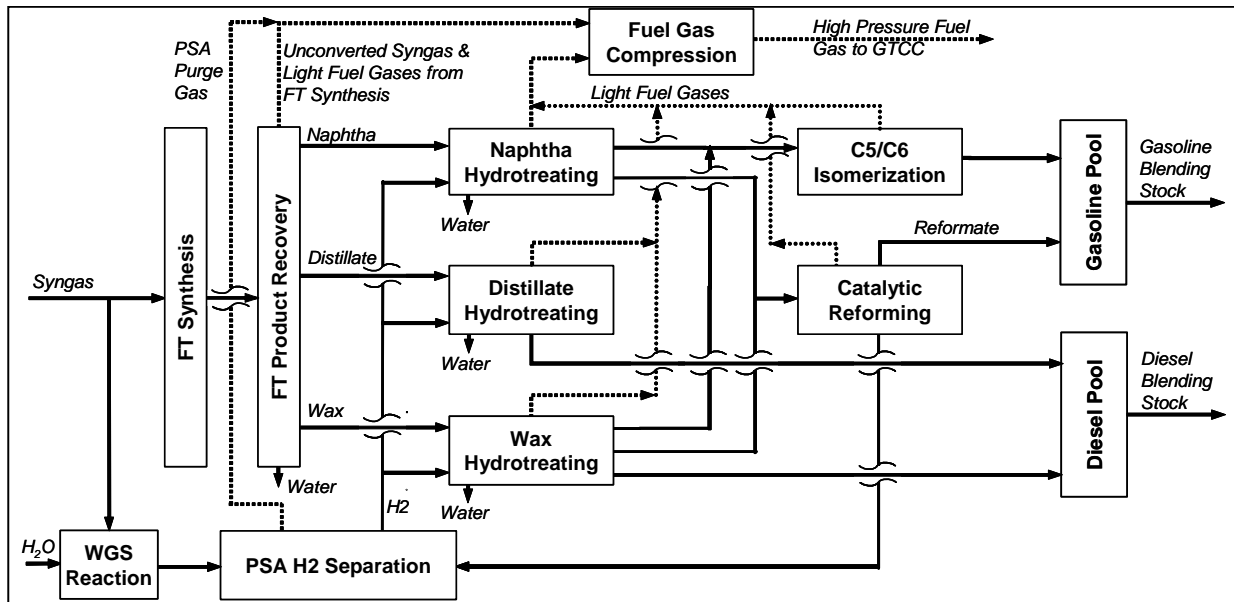


Figure 2. Schematic of the separation and refining area of the F-T production facilities simulated in

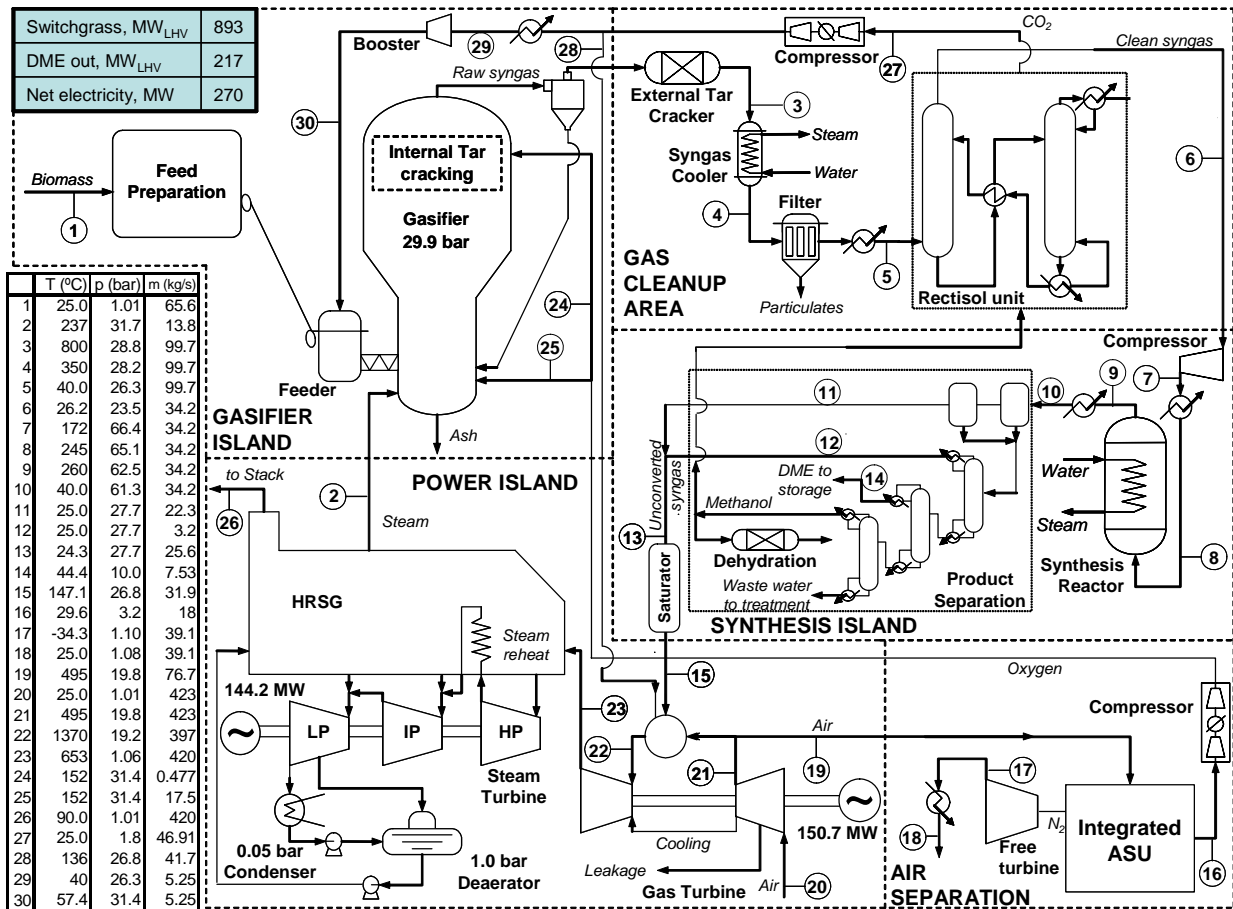


Figure 3. Detailed energy and mass balance for DME production from switchgrass with once-

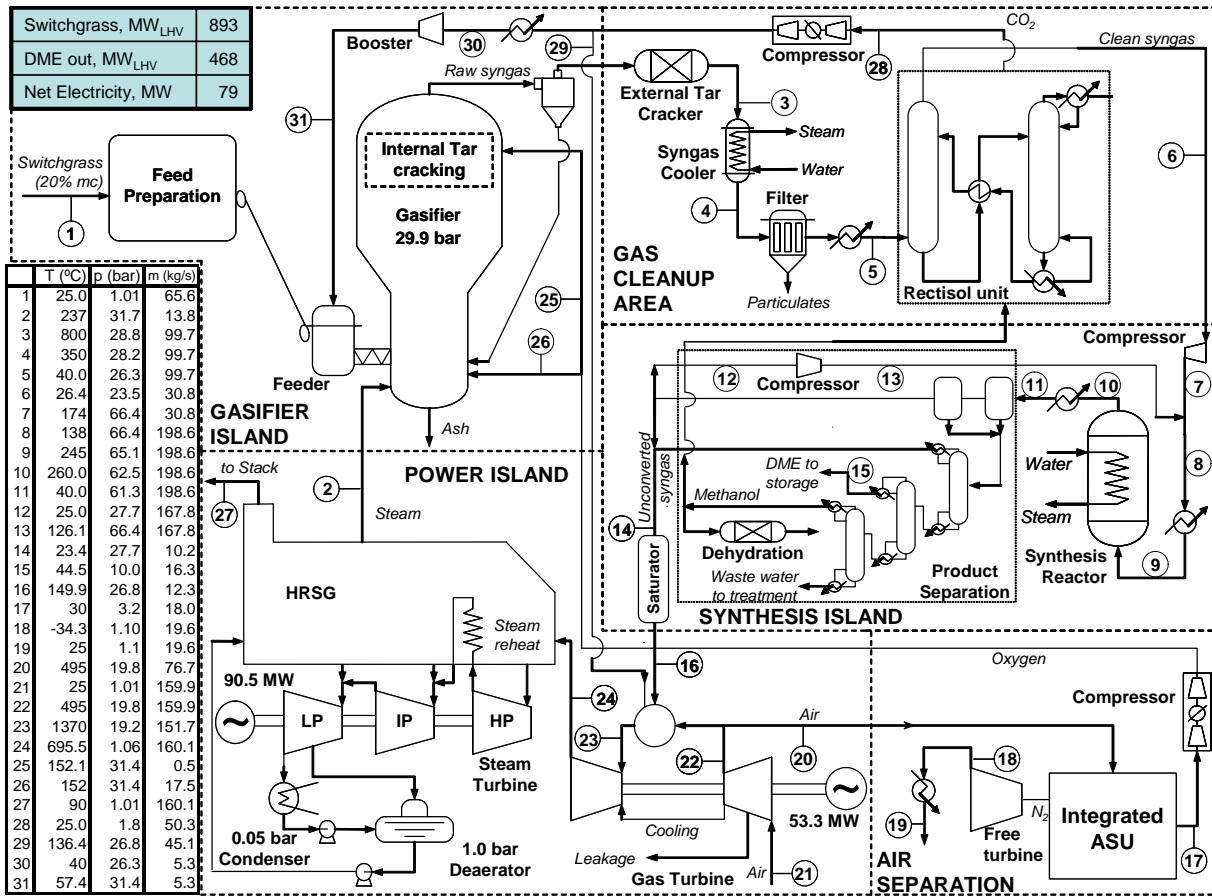


Figure 4. Detailed energy and mass balance for DME production from switchgrass with recycle processing of syngas through a liquid-phase synthesis reactor and production of some electricity

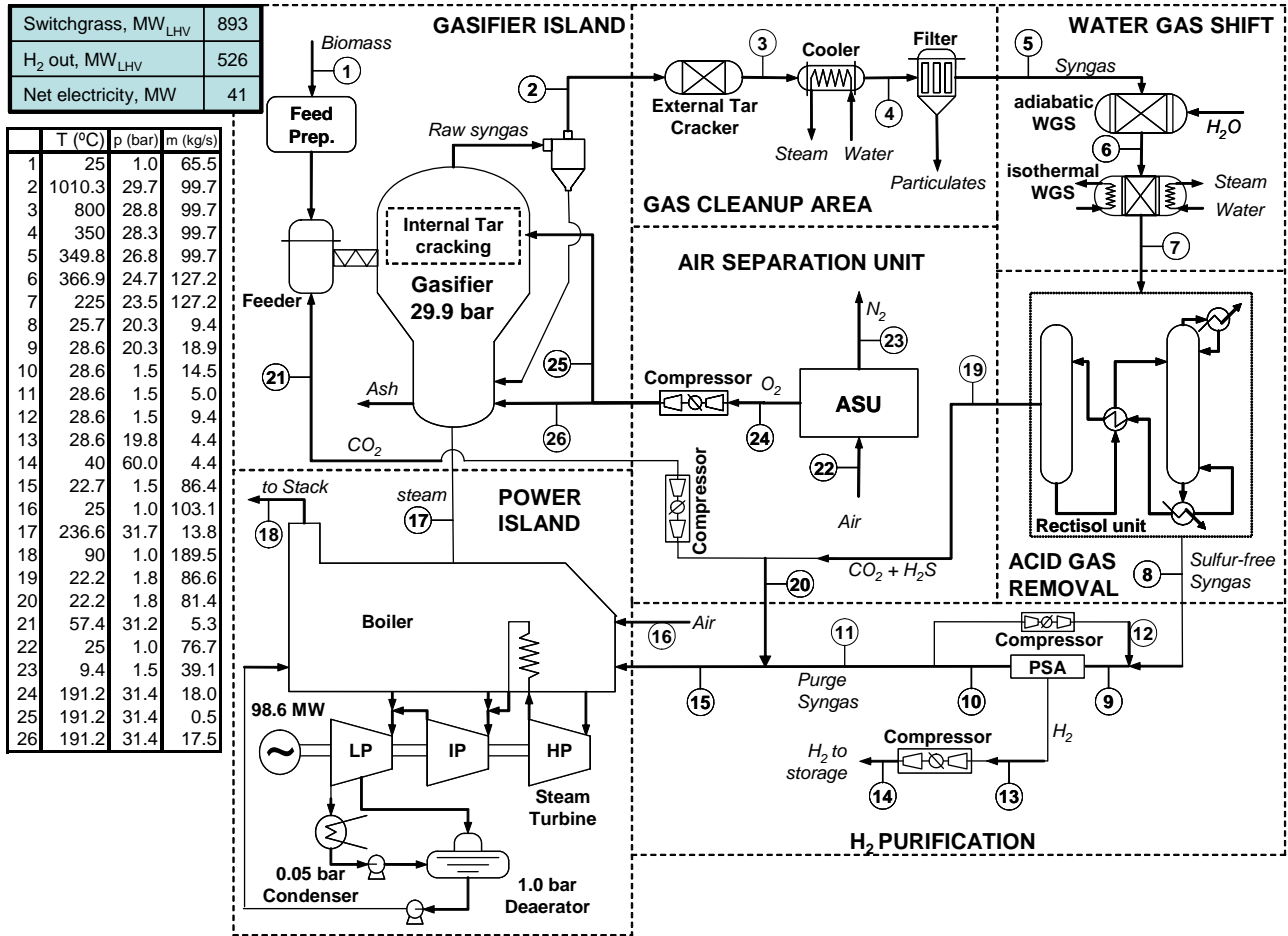


Figure 5. Detailed energy and mass balance for production of hydrogen from switchgrass. This

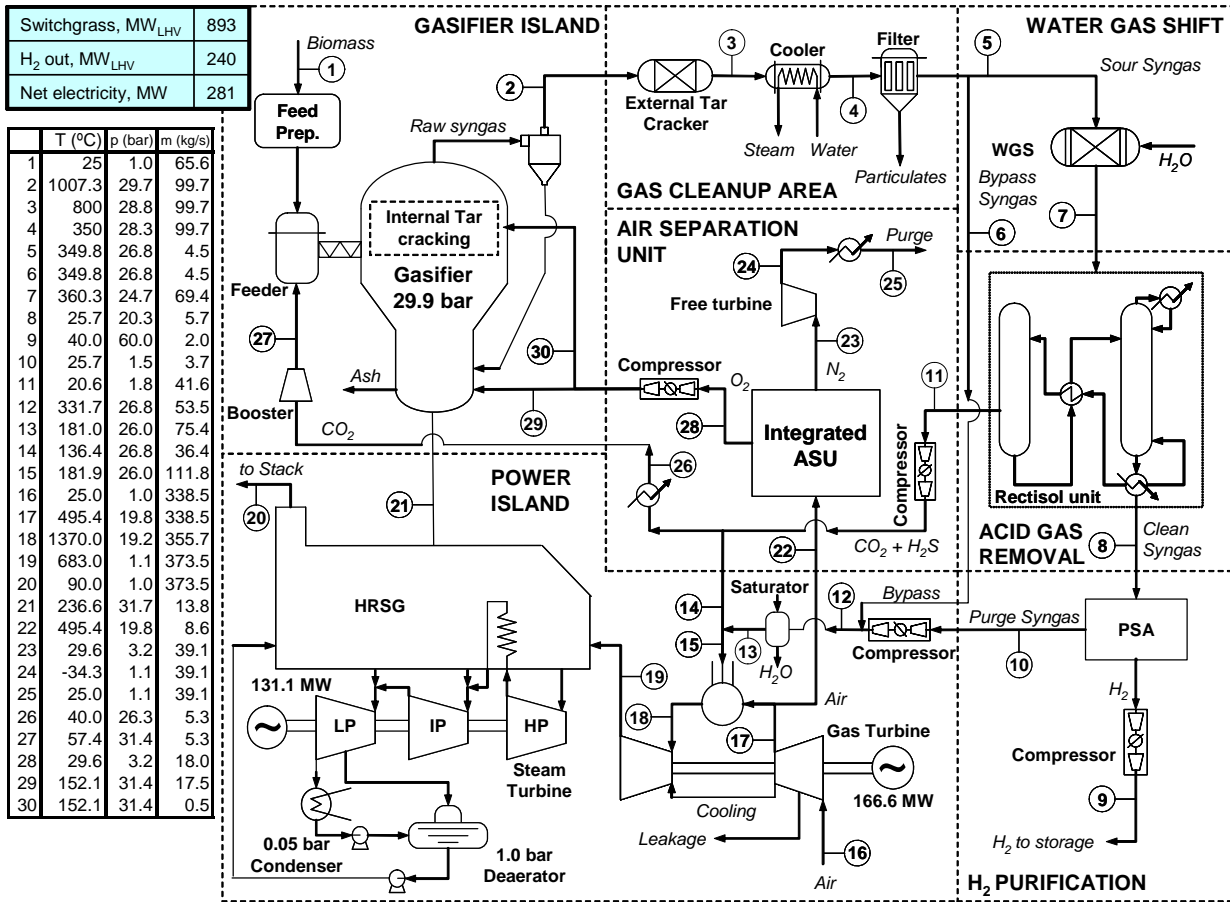


Figure 6. Detailed energy and mass balance for production of approximately equal MW (HHV) of hydrogen and electricity from switchgrass (H-5050). In this design, exactly half of the clean

Table 1. Process designs developed in this work.

Abbreviation for Process Configuration*	Products			
	Electricity	F-T Fuels	DME	H ₂
Dimethyl Ether				
D-OT	✓		✓	
D-RC	low ✓		✓	
Fischer-Tropsch Fuels				

Table 2. Physical Properties of DME, petroleum diesel, and constituents of LPG.

Property	DME	Diesel	Propane	Butane
Cetane number	55-60	40-55	na	na
Vapor Pressure @ 20 deg C [bar]	5.1	< 1	8.4	2.1

Table 3. Our assumptions regarding indirect costs, expressed as percent of total direct cost.

	Gasifier and gas cleanup islands	Air separation unit and power island
Engineering & head office	15%	15%
Contingency	10%	5%

Table 4. Reference capacities, capital costs, and scaling factors for for fuels production plants.

Plant Area	Sub-Unit	Capacities (in indicated units)			Cost (in million 2003 \$)	
		Base	Max. unit	Unit of Capacity	Base	Scaling exp.
		S_o	S_{max}		C_o^a	f
Gasifier Island	Feed preparation ^b	64.6	n.a.	wet tonne/hr biomass	3.17	0.77
	GTI 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
Gas Cleanup	External tar cracker ^e	47.1	52	Actual m ³ /s gas feed	0.732	0.7
	Syngas cooler ^f	77	n.a.	MW _{th} heat duty	25.4	0.6
	Ceramic filter ^g	14.4	n.a.	actual m ³ /s gas feed	18.60	0.65
	Rectisol AGR ^h	200000	n.a.	Nm ³ /hr gas feed	20.00	0.65
	AGR compressor ⁱ	10	n.a.	MW _e consumed	4.83	0.67
	CO ₂ compression ^j	10	n.a.	MW _e consumed	4.75	0.67
FT Synthesis, Fuel Upgrading and Refinery	Slurry phase F-T reactor ^k	2.52	n.a.	million scf/hr feed gas	10.5	0.72
	Hydrocarbon recovery unit ^l	14.44	200	thousand lbs/hr feed	0.56	0.7
	H ₂ recovery unit ^m	0.033	0.1	million scf/hr H ₂ prod	0.65	0.7
	Wax hydrocracker ⁿ	8.984	575	thousand lbs/hr feed	7.21	0.7
	Distillate hydrotreater ^o	2.871	650	thousand lbs/hr feed	1.93	0.7
	Naphtha hydrotreater ^p	2.05	650	thousand lbs/hr feed	0.58	0.7
	Naphtha reformer ^q	3.43	750	thousand lbs/hr feed	4.02	0.7
	C ₅ /C ₆ isomerization ^r	1.158	250	thousand lbs/hr feed	0.74	0.7
	CO shift reactor ^s	0.040	0.080	million scf/hr feed gas	0.79	0.7
Fuel gas compressor ^t	10	n.a.	MW compressor power	4.83	0.67	
DME Synthesis and Separation	Once-through LP synthesis ^u	2.91	n.a.	kmol/sec feed gas	15.8	0.65
	Recycle LP synthesis ^v	8.68	n.a.	kmol/sec total feed gas	88.8	0.65
	DME distillation plant ^w	6.75	n.a.	kg/s DME product	21.3	0.65
	MeOH dehydration ^x	2.91	n.a.	kmol/s MeOH feed	15.8	0.65
	Syngas expander ^y	10	n.a.	MW _e generated	2.41	0.67
	Syngas compressor ^t	10	n.a.	MW compressor power	4.83	0.67
Hydrogen	Water Gas Shift ^z	1377	n.a.	MW _{LHV} biomass input	30.6	0.67

production	PSA ^{aa}	0.294	n.a.	kmol/s purge gas flow	5.46	0.74
	PSA purge gas compressor ^{bb}	10	n.a.	MW _e compressor power	4.83	0.67
	H ₂ -rich gas compressor ^{cc}	10	n.a.	MW _e compressor power	4.83	0.67
Air Separation Area	ASU, if stand-alone ^{dd}	76.6	n.a.	tonne/hr pure O ₂	35.6	0.50
	ASU, if integrated ^{dd}	76.6	n.a.	tonne/hr pure O ₂	22.7	0.50
	O ₂ compressor ^{ee}	10	n.a.	MW _e consumed	5.54	0.67
	N ₂ compressor ^{ee}	10	n.a.	MW _e consumed	4.14	0.67
	N ₂ expander ^y	10	n.a.	MW _e generated	2.41	0.67
Power Island	Saturator ^{ff}	20.9	n.a.	actual m ³ /s gas feed	0.30	0.70
	Gas turbine ^{gg}	266	334	GT MW _e	56.0	0.75
	HRSG + heat exchangers ^{hh}	355	n.a.	MW _{th} heat duty ⁱⁱ	41.2	1
	Steam cycle (turbine + cond.) ^{jj}	136	n.a.	ST gross MW _e	45.5	0.67

Notes to Table 4

- (a) The GDP implicit price deflator (Council of Economic Advisors, February 2004) has been used to convert to constant 2003 dollars from other year dollars when necessary.
- (b) Weyerhaeuser (2000) gives the following installed “Nth 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 1.45 (Guthrie, 1969) to calculate the cost of a feed preparation system rated for 30 bar, and by a factor of 3.1 to account for the lower bulk density of chopped switchgrass compared to wood chips. [The ratio of the average bulk density for softwood chips to chopped straw is given by Jenkins (1989) to be 3.1. We assume equipment cost scales linearly with bulk density.] 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 (2001) 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 (2001) were not used directly here for C_o because they represent first-of-a-kind plant cost estimates, whereas Weyerhaeuser (2000) gives estimates for an Nth plant.
- (c) From Hughes (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 Nth plant design. From Guthrie (1969), 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. Following Lau, *et al.* (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 2.5×400 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 from Hughes (2003). S_{max} is taken from the above discussion of Lau, *et al.* (2003) where 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 Katofsky (1993), Williams *et al.* (1995), Tijmensen (2000), Hamelinck and Faaij (2001), Tijmensen *et al.* (2002), and Hughes (2003).

- (d) From Weyerhaeuser (2000), a cost of \$584k (1999 dollars), was extracted for a cyclone separator. The ash cyclone is similar to a combustor primary cyclone, and the cost scales with the volumetric flowrate of gas through the unit. Installation labor is included in this cost. This cost is multiplied by 1.45 (Guthrie, 1969) to calculate the cost of a cyclone separator rated for 30 bar. The combustor primary cyclone in Weyerhaeuser (2000) 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 Weyerhaeuser (2000), 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.* (2003) for a cyclone separator.
- (e) Cost is based on Weyerhaeuser (2000), who gives a cost of \$0.678 million (in 1999\$) for a tar cracker passing 50.2 metric tonnes per hour of gas (or $47.1 \text{ m}^3/\text{s}$ of gas at the actual temperature and pressure in the Weyerhaeuser study, assuming ideal gas). The estimated maximum capacity and scaling factor are taken from Hamelinck, *et al.* (2003).
- (f) Based on Simbeck (2004), who gives \$310/kW_{th} of saturated steam produced in a firetube 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 text) when total plant costs are estimated later in this paper. Scaling exponent for high-temperature heat exchangers is taken from Hamelinck *et al.* (2003).
- (g) From Newby *et al.* (1998), a cost of \$14.6 M (1997 dollars) was extracted for a ceramic candle filter (\$36.0/kW for a 406 MW coal IGCC plant). This cost does not include installation labor, so 15% is added (Craig and Mann, 1996). It is given for a N^{th} plant. S_o is the gas feed rate to the ceramic filter evaluated in Newby *et al.* (1998). f is taken to have the same value as for a fabric filter in Hamelinck *et al.* (2003).

- (h) This installed cost estimate is from a personal communication (in 2003) with U. Koss of Lurgi, which licenses Rectisol technology. We assume this estimate excludes BOP and indirect costs. The latter are included (using methodology discussed in text) in estimating total installed capital cost for the B-IGCC-CCS and other cases utilizing a Rectisol unit.
- (i) The Rectisol system captures small amounts of CO, H₂, and CH₄ in addition to CO₂ and H₂S (as discussed in text). The CO, H₂, and CH₄ are separated from the CO₂ and H₂S and fed back to the syngas stream downstream of the Rectisol plant. This compressor raises the pressure of the CO, H₂, and CH₄ to the main syngas stream pressure. The cost for this compressor is modeled after that for a syngas compressor, as indicated by Kreutz *et al.* (2005). See note (ee) for discussion of compressor costs.
- (j) This compressor is used to pressurize the CO₂ stream captured by the Rectisol plant up to a pressure of 31.4 bar (as required for use in the feed lockhopper). Cost is based on Kreutz *et al.* (2005), who give a cost of \$6.28 million (in 2002\$) for a 3-stage intercooled purge-gas compressor at a scale of 10 MW_e compressor capacity. The \$6.28 million includes BOP, indirects (15%), and contingencies (15%). The cost in this table includes BOP, but indirects and contingencies have been factored out (to enable a consistent percentage indirects and contingencies to be included in the cost estimates for specific plants). See note (ee) for additional discussion of compressor costing.
- (k) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a slurry-phase F-T reactor operating at an average temperature and pressure of 253°C and 21.4 bar, with a feed gas H₂/CO ratio of 0.52, and having a feed gas capacity of 2.52 million standard cubic feet per hour (mmSCF/hr). Bechtel’s cost estimate is \$9.5 million (in 1997\$). They indicate a maximum single-reactor capacity of 8.1 mmSCF/hr, which is approximately the capacity needed in the systems described in this report. The cost scaling factor is from Hamelinck *et al.* (2003).
- (l) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a hydrocarbon recovery plant with a feed of 14.44 thousand lbs/hour of \$0.507 million (in 1997\$). The maximum capacity for such a plant is indicated by Bechtel to be 200 thousand lbs/hr feed rate. We have assumed the indicated scale factor.
- (m) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a hydrogen recovery plant utilizing pressure-swing adsorption. Bechtel indicates a cost for a unit producing 0.033 million SCF/hr of H₂ of \$0.583 million (in 1997\$) and a maximum capacity for such a system of 0.1 million SCF/hr. We have assumed the indicated scale factor.
- (n) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a wax hydrocracker with a feed rate of 8.984 thousand lbs/hr of \$6.509 million (in 1997\$). Bechtel indicates a maximum unit capacity for such a hydrocracker to be 575 thousand lbs/hr. We have assumed the indicated scale factor.

- (o) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a distillate hydrotreating unit with a feed rate of 2.871 thousand lbs/hr of \$6.509 million (in 1997\$). Bechtel indicates a maximum unit capacity for such a hydrotreater to be 650 thousand lbs/hr. We have assumed the indicated scale factor.
- (p) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a naphtha hydrotreating unit with a feed rate of 2.05 thousand lbs/hr of \$0.528 million (in 1997\$). Bechtel indicates a maximum unit capacity for such a hydrotreater to be 650 thousand lbs/hr. We have assumed the indicated scale factor.
- (q) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a naphtha reformer with a feed rate of 3.43 thousand lbs/hr of \$3.628 million (in 1997\$). Bechtel indicates a maximum unit capacity for such a reformer to be 750 thousand lbs/hr. We have assumed the indicated scale factor.
- (r) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a C₅/C₆ isomerization plant with a feed rate of 1.158 thousand lbs/hr of \$0.669 million (in 1997\$). Bechtel indicates a maximum unit capacity for this type of unit to be 250 thousand lbs/hr. We have assumed the indicated scale factor.
- (s) Based on an “inside battery limits” cost estimate (assumed here to exclude BOP and indirect costs) from Bechtel (1998) for a CO shift reactor with a feed rate of 0.04 million scf/hr of \$0.715 million (in 1997\$). Bechtel indicates a maximum unit capacity for this type of unit to be 0.080 million scf/hr. We have assumed the indicated scale factor.
- (t) For plant designs involving F-T synthesis, this compressor is used to deliver unconverted syngas plus purge gases from the F-T product upgrading area to the gas turbine. For DME synthesis designs, this compressor raises the pressure of the syngas after the Rectisol unit to the required synthesis reactor pressure. In both cases, the cost of this compressor is modeled on that of a purge gas compressor, as discussed in note (ee).
- (u) This cost is based on a projected scaled-up cost for Nth-plant version of the liquid-phase synthesis reactor demonstrated at the Eastman Chemical facility in Kingsport, Tennessee, as provided by Moore (2003). See also Larson and Ren (2003). The original estimate provided by Moore was \$15.76 million (2002\$) for a reactor with a gas feed rate of 2.91 kmol/sec. The scale factor was recommended by Moore (2003). We assume BOP (15%, according to Moore) and indirect costs (assumed to be 32%) are not included in the cost shown in this table. These latter costs are added in later when the total installed capital costs for the DME cases are developed.
- (v) Larson and Ren (2003) indicate a cost for recycle-based synthesis (including recycle compressor) of \$87.37 million (in 2002\$) for a total feed gas flow (fresh feed plus recycled unconverted gas) of 8.69 kmol/s.
- (w) A cost estimate of \$21.26 million (2002\$) for a DME distillation plant delivering 6.75 kg/sec of DME was provided by Moore (2003). See also Larson and Ren (2003). The scale factor was recommended by Moore (2003). We assume BOP (15%, according to Moore) and indirect costs (assumed to be 32%) are not included in the cost shown in this table. These latter costs are added in later when the total installed capital costs for the DME cases are developed.

- (x) Since equipment for once-through liquid-phase DME synthesis has many similarities to MeOH dehydration equipment, we assume that the installed cost per kmol of feed to the MeOH dehydration area would be the same as that for once-through LPDME synthesis. BOP (15%) and indirect costs (32%) are not included in the cost shown in this table. These are added in later when the total installed capital costs for the DME cases are developed.
- (y) The cost for a syngas expander is based on Kreutz *et al.* (2005), who give an estimate of \$3.14 million (in 2002\$) for the installed cost of a unit producing 10 MW_e. We assume the cost for a nitrogen expander is the same as for a syngas expander. We assume Kreutz, *et al.*'s cost includes BOP, engineering, and contingencies. The latter two items are assumed to be 15% each, giving a base cost (including BOP) of \$2.37 million, or \$2.41 million in 2003\$. Engineering and contingency costs are included (using methodology discussed in text) when total plant costs are estimated later in the paper. Scale factor is from Kreutz, *et al.* (2005).
- (z) This estimate is for a two-stage water gas shift system (adiabatic first stage, isothermal second stage, with heat recovery), as configured in the process design, H-MAX. The cost shown is based on Kreutz *et al.* (2005), who give a cost of \$39.8 million (in 2002\$) for a 2-stage water gas shift system (including heat recovery) processing all syngas from a Texaco-type coal gasifier with input of 1377 MW coal (LHV). Kreutz's cost includes BOP, indirects (15%), and contingencies (15%). The number in this table excludes the indirects and contingencies and is expressed in 2003\$. (Indirects and contingencies are included in the total plant costs calculated later in this report.) For a single-stage (adiabatic) WGS (as configured in the process design, H-50/50), Moore (2003) indicates that the cost would be about 40% of the cost for the two-stage system.
- (aa) Kreutz *et al.* (2005) gives an estimated cost for a PSA unit recovering 85% of hydrogen in a shifted syngas stream of \$7.1 million (in 2002\$). The unit has a purge gas flow rate of 0.294 kmol/sec. Factoring out 15% indirect cost and 15% engineering costs, gives \$5.37 million, or \$5.46 million when converted to 2003\$.
- (bb) The cost for a PSA purge gas compressor is modeled as described in note (ee).
- (cc) The cost for a hydrogen compressor, based on Kreutz *et al.* (2005), is taken to be the same as the cost for a syngas compressor, as described in note (ee).
- (dd) From Kreutz *et al.* (2005), we extracted 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 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.* (2005) to obtain C_o . In Kreutz *et al.* (2005), 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's cost we divide by (1.1x1.05), 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.* (2005). 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 O₂ 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 as \$35.0M 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. According to Moore (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.* 2005, Moore 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).

- (ee) From Kreutz *et al.* (2005), the following original costs, in millions of 2002 dollars, were extracted 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.* (2005) are low heating value gases consisting mainly of CO, CO₂, H₂O, and H₂. 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.* (2005) to obtain C_o . In Kreutz *et al.* (2005), the total direct cost (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.* (2005) is divided by (1.1×1.05) , and to obtain C_o for purge gas compressors, the cost is divided by (1.15×1.15) . 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.* (2005).
- (ff) The cost for a saturator has been modeled from that for a gas cleanup scrubber (Weyerhaeuser, 2000), since the technologies are similar. Weyerhaeuser (2000) gives a cost of \$0.206 million (in 1999\$) for a scrubber processing 50.2 tonnes of synthesis gas at close to atmospheric pressure (or 47.1 m³/s of gas at the actual temperature and pressure in the Weyerhaeuser study, assuming ideal gas). The cost in the table above is a result of converting to 2003\$ and multiplying by

a pressure factor (derived from Guthrie, 1969) of 1.35 to account for operating at 27.6 bar instead of atmospheric pressure.

The scaling factor is based on Hamelinck, *et al.* (2003).

- (gg) Kreutz *et al.* (2005) indicates an installed cost of \$72.8 million (2002\$) for a Siemens V94.3A gas turbine (266 MW_e output), the turbine design on which our simulation is based. 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 text) 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 (Anonymous, 2003). The value for S_{max} is the power rating of the largest simple-cycle gas turbine generator in Gas Turbine World's 2003 Handbook. 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.
- (hh) Based on Simbeck (2004), 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 text) 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-VENT case, 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.
- (ii) 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.
- (jj) Kreutz *et al.* (2005) 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 text) when total plant cost is estimated later in this paper. Scale factor is from Kreutz *et al.* (2005).

Table 5. Capital cost estimate for DME and electricity co-production with D-OT design.

Plant Area	Sub-Unit	Capacities (in indicated units)			Unit of Capacity	Costs (in million 2003 \$)		
		Required capacity	Number of units	Capacity per unit		Cost per unit	Train cost	Overnight Cost
		S_r	n^a	S^b		C^c	C_m^d	OC^e
Gasifier Island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash Cyclone	15.6	2	7.9	actual m ³ /s gas feed	0.20	0.37	0.59
Gas Cleanup	Tar cracker	15.6	2	7.8	actual m ³ /s gas feed	0.21	0.39	0.62
	Syngas cooler	81.7	2	40.8	MW _{th} heat duty	17.3	32.3	51.6
	Ceramic filter	9.1	2	4.5	actual m ³ /s gas feed	8.79	16.4	26.2
	Rectisol AGR	363314	1	363314	Nm ³ /hour gas feed	27.4	27.4	43.7
	AGR Compressor	0.094	1	0.094	MW _e consumed	0.21	0.21	0.28
	Syngas compressor	11.6	1	11.6	MW _e consumed	5.35	5.35	7.06
	CO ₂ compressor	11.6	1	11.6	MW _e consumed	5.34	5.34	7.04
	CO ₂ boost compressor	0.07	1	0.1	MW _e consumed	0.17	0.17	0.23
DME Synthesis	LPDME synthesis	2.61	1	2.61	kmol/s feed gas	14.67	14.67	22.3
	DME distillation	7.53	1	7.53	kg/s DME output	22.83	22.83	34.7
	MeOH dehydration	0.082	1	0.082	kmol/s MeOH feed	1.55	1.55	2.36
	Syngas Expander	1.68	1	1.7	MW _e consumed	0.73	0.73	0.97
	Syngas Compressor	0.024	1	0.024	MW _e consumed	0.08	0.08	0.11
Air Separation	ASU	61.3	1	61.3	tonne/hr pure O ₂	20.28	20.28	25.8
	O ₂ compressor	5.44	1	5.44	MW _e consumed	3.69	3.69	4.68
	N ₂ expander	2.57	1	2.57	MW _e produced	0.97	0.97	1.23
Power Island	Saturator	2.65	1	2.65	actual m ³ /s gas feed	0.07	0.07	0.11
	Gas Turbine	151	1	151	GT MW _e	36.55	36.55	46.4
	HRSG + heat exchangers	477	1	477	MW _{th} heat duty ^f	55.42	55.42	85.1
	Steam cycle (turb. + cond.)	144	1	144	ST gross MW _e	47.32	47.32	60.1
Total Overnight Installed Capital Cost (million \$)				502				

(a) This is the number of units required, which is determined by rounding $n = S_r / S_{max}$ up to the nearest integer. This gives $n =$

1 whenever $S_r \leq S_{max}$ 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 4 for values of C_o , S_o , and f for individual process units. For some units, C includes BOP (see notes dd, ee, and gg in Table 4), but indirect costs are not included in any values of C (see note e below for discussion of indirect costs).
- (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 4 already include BOP (see note c above). For the power island, BOP is estimated as a separate component based on Kreutz *et al.* (2005). For all other components, we estimate BOP to be 20.9% of installed cost. This percentage was determined as follows. Hamelinck and Faaij (2001) 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 (Katofsky 1993, DeLong 1995, Stone and Webster *et al.* 1995, Weyerhaeuser 2000, Hamelinck and Faaij 2001, Kreutz *et al.* 2005) for similar plants at varying scales. $BOP (\%) = 0.8867 / \{(\text{biomass } MW_{th})^{0.2096}\}$. For the 4536 tonne/day scale investigated in this project, the thermal input (HHV) is 983.2 MW, so that BOP is 20.9 % of the installed cost. Indirect costs were estimated from review of several literature sources (Katofsky 1993, DeLong 1995, Stone and Webster *et al.* 1995, Mann and Spath 1997, Washington Group and Southern Company 2000, Weyerhaeuser 2000, Hamelinck *et al.* 2003, Kreutz *et al.* 2005). For gasifier island and gas cleanup units, $IC = 32\%$ of TDC, and for power island and air separation unit units, $IC = 27\%$ of TDC. Thus for a typical gasifier island or gas cleanup unit, $OC = C_m \times (1.209) \times (1.32) = 1.60 C_m$. For an air separation unit or power island unit, since BOP is already included in C_o or accounted for as a separate item, $OC = C_m \times (1.27)$.
- (f) Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler. See note (hh) of Table 4 for additional discussion.

Table 6. Capital cost estimate for DME and electricity co-production with D-RC design.

Plant Area	Sub-Unit	Capacities (in indicated units)			Unit of Capacity	Costs (in million 2003 \$)		
		Required capacity	Number of units	Capacity per unit		Cost per unit	Train cost	Overnight Cost

		S_r	n^a	S^a		C^a	C_m^a	OC^a
Gasifier Island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash Cyclone	15.6	2	7.9	actual m ³ /s gas feed	0.20	0.37	0.59
Gas Cleanup	Tar cracker	15.6	2	7.8	actual m ³ /s gas feed	0.21	0.39	0.62
	Syngas cooler	81.7	2	40.8	MW _{th} heat duty	17.3	32.3	51.6
	Ceramic filter	9.1	2	4.5	actual m ³ /s gas feed	8.79	16.4	26.2
	Rectisol AGR	294202	1	294202	Nm ³ /hour gas feed	25.7	25.7	41.0
	AGR Compressor	0.094	1	0.094	MW _e consumed	0.21	0.21	0.28
	Syngas compressor	11.34	1	11.34	MW _e consumed	5.25	5.25	6.93
	CO ₂ compressor	12.46	1	12.46	MW _e consumed	5.59	5.59	7.38
DME Synthesis	CO ₂ boost compressor	0.068	1	0.068	MW _e consumed	0.17	0.17	0.23
	LPDME synthesis, DME distillation, recycle comp.	14.38	1	14.38	kmol/s total feed gas (fresh + recycled)	123.3	123.3	187.2
	MeOH dehydration	0.32	1	0.32	kmol/s MeOH feed	3.76	3.76	5.71
	Syngas Expander	12.1	1	12.1	MW _e consumed	2.74	2.74	3.62
Air Separation	Syngas Compressor	0.043	1	0.043	MW _e consumed	0.12	0.12	0.16
	ASU	61.3	1	61.3	tonne/hr pure O ₂	20.28	20.28	25.8
	O ₂ compressor	5.44	1	5.44	MW _e consumed	3.69	3.69	4.68
	N ₂ compressor	6.69	1	6.69	MW _e consumed	3.16	3.16	4.01
Power Island	N ₂ expander	1.29	1	1.29	MW _e produced	0.61	0.61	0.78
	Saturator	0.81	1	0.81	actual m ³ /s gas feed	0.03	0.03	0.05
	Gas Turbine	53.3	1	53.3	GT MW _e	16.75	16.75	21.3
	HRSG + heat exchangers	466	1	466	MW _{th} heat duty ^b	54.13	54.13	83.1
	Steam cycle (turb. + cond.)	90.5	1	90.5	ST gross MW _e	34.63	34.63	44.0
Total Overnight Installed Capital Cost (million \$)				596				

(a) See corresponding note in Table 5 for descriptions of the column-heading parameters in this table.

(b) Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler. See note (hh) of Table 4 for additional discussion.

Table 7. Capital cost estimate for F-T and electricity co-production with F-OT design.^a

Plant Area	Sub-Unit	Capacities (in indicated units)			Unit of Capacity	Costs (in million 2003 \$)		
		Required capacity	Number of units	Capacity per unit		Cost per unit	Train cost	Overnight Cost
		S_r	n^a	S^a		C^a	C_m^a	OC^a
Gasifier Island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash Cyclone	15.6	2	7.9	actual m ³ /s gas feed	0.20	0.37	0.59
Gas Cleanup	Tar cracker	15.6	2	7.8	actual m ³ /s gas feed	0.21	0.39	0.62
	Syngas cooler	81.7	2	40.8	MW _{th} heat duty	17.3	32.3	51.6
	Ceramic filter	9.1	2	4.5	actual m ³ /s gas feed	8.79	16.4	26.2
	Rectisol AGR	323748	1	323748	Nm ³ /hour gas feed	27.35	27.35	43.7
	AGR Compressor	0.094	1	0.094	MW _e consumed	0.21	0.21	0.28
	CO ₂ compressor	12.44	1	12.44	MW _e consumed	5.59	5.59	7.37
	CO ₂ boost compressor	0.068	1	0.068	MW _e consumed	0.17	0.17	0.23
FT and Upgrading	Slurry phase F-T reactor	8.04	1	8.04	million scf/hr feed gas	24.29	24.29	38.8
	Hydrocarbon recovery unit	242.7	2	242.7	thousand lbs/hr feed	2.49	4.65	7.42
	H ₂ recovery unit	0.15	2	0.15	million scf/hr H ₂ prod	1.17	2.19	3.49
	Wax hydrocracker	43.4	1	43.4	thousand lbs/hr feed	21.71	21.71	34.7
	Distillate hydrotreater	6.81	1	6.81	thousand lbs/hr feed	3.53	3.53	5.63
	Naphtha hydrotreater	12.86	1	12.86	thousand lbs/hr feed	2.11	2.11	3.37
	Naphtha reformer	18.79	1	18.79	thousand lbs/hr feed	13.21	13.21	21.1
	C ₅ /C ₆ isomerization	5.80	1	5.8	thousand lbs/hr feed	2.29	2.29	3.65
	CO shift reactor	0.070	1	0.070	million scf/hr feed gas	1.17	1.17	1.86
Fuel gas compressor	5.70	1	5.70	MW _e consumed	2.93	2.93	3.87	
Air Separation	ASU	61.3	1	61.3	tonne/hr pure O ₂	20.28	20.28	25.8
	O ₂ compressor	5.44	1	5.44	MW _e consumed	3.69	3.69	4.68
	N ₂ expander	2.57	1	2.57	MW _e produced	0.97	0.97	1.23
Power Island	Gas Turbine	86.7	1	86.7	GT MW _e	24.13	24.13	30.7
	HRSG + heat exchangers	476	1	476	MW _{th} heat duty ^p	55.2	55.2	84.8

Steam cycle (turb. + cond.)	141	1	141	ST gross MW _e	46.59	46.59	59.2
Total Overnight Installed Capital Cost (million \$)				541			

- (a) See corresponding note in Table 5 for descriptions of the column-heading parameters in this table.
- (b) Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler. See note (hh) of Table 4 for additional discussion.

Table 8. Capital cost estimate for hydrogen and electricity co-production with H-MAX design.^a

Plant Area	Sub-Unit	Capacities (in indicated units)			Unit of Capacity	Costs (in million 2003 \$)		
		Required capacity	Number of units	Capacity per unit		Cost per unit	Train cost	Overnight Cost
		S_r	n	S		C	C_m	OC
Gasifier Island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash cyclone	15.6	2	7.9	actual m ³ /s gas feed	0.20	0.37	0.59
Gas Cleanup	Tar cracker	15.6	2	7.8	actual m ³ /s gas feed	0.21	0.39	0.62
	Syngas cooler	81.7	2	40.8	MW _{th} heat duty	17.3	32.3	51.6
	Ceramic filter	9.10	2	4.5	actual m ³ /s gas feed	8.79	16.4	26.2
Water Gas Shift	Saturator	9.10	1	9.10	actual m ³ /s gas feed	0.17	0.17	0.27
	2-stage WGS	887	1	887	MW _{LHV} biomass input	22.78	22.78	30.1
CO ₂ Removal	Rectisol AGR	359305	1	359305	Nm ³ /hour gas feed	29.27	29.27	46.7
	AGR compressor	0.052	1	0.052	MW _e consumed	0.14	0.14	0.19
	CO ₂ compressor	1.38	1	1.38	MW _e consumed	1.28	1.28	1.69
PSA area	PSA	1.0	1	1.0	kmol/s purge gas flow	13.79	13.79	18.20
	PSA purge gas compressor	7.27	1	7.27	MW _e consumed	3.90	3.90	5.15
	H ₂ -rich gas compressor	9.74	1	9.74	MW _e consumed	4.74	4.74	6.26
Air Separation	ASU	61.3	1	61.3	tonne/hr pure O ₂	31.82	31.82	40.4
	O ₂ compressor	7.26	1	7.26	MW _e consumed	4.47	4.47	5.68
Power Island	HRSG + heat exchangers	201	1	201	MW _{th} heat duty ^b	23.36	23.36	35.9
	Steam cycle (turb. + cond.)	98.6	1	98.6	ST gross MW _e	36.67	36.67	46.57
Total Overnight Installed Capital Cost (million \$)				397				

(a) See corresponding note in Table 5 for descriptions of the column-heading parameters in this table.

(b) Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler and that associated with the WGS reactors. See note (hh) of for additional discussion.

Table 9. Capital cost estimate for hydrogen and electricity co-production with H-50/50 design.

Plant Area	Sub-Unit	Capacities (in indicated units)			Unit of Capacity	Costs (in million 2003 \$)		
		Required capacity	Number of units	Capacity per unit		Cost per unit	Train cost	Overnight Cost
		S_r	n^a	S^a		C^a	C_m^a	OC^a
Gasifier Island	Feed preparation	236	2	118	wet tonne/hr biomass	15.7	29.2	46.6
	Gasifier	189	2	94.5	dry tonne/hr biomass	11.4	21.2	33.9
	Ash cyclone	15.6	2	7.9	actual m ³ /s gas feed	0.20	0.37	0.59
Gas Cleanup	Tar cracker	15.6	2	7.8	actual m ³ /s gas feed	0.21	0.39	0.62
	Syngas cooler	81.7	2	40.8	MW _{th} heat duty	17.3	32.3	51.6
	Ceramic filter	9.10	2	4.5	actual m ³ /s gas feed	8.79	16.4	26.2
Water Gas Shift	Saturator	9.10	1	4.89	actual m ³ /s gas feed	0.11	0.11	0.17
	Single-stage WGS	443	1	443	MW _{LHV} biomass input	5.73	5.73	7.56
CO ₂ Removal	Rectisol AGR	176789	1	176789	Nm ³ /hour gas feed	18.46	18.46	29.5
	AGR compressor	0.026	1	0.026	MW _e consumed	0.090	0.090	0.12
	CO ₂ compressor	6.44	1	6.44	MW _e consumed	3.59	3.59	4.74
	CO ₂ boost compressor	0.068	1	0.068	MW _e consumed	0.17	0.17	0.23
PSA area	PSA	0.33	1	0.33	kmol/s purge gas flow	5.94	5.94	7.85
	PSA purge gas compressor	3.39	1	3.39	MW _e consumed	2.34	2.34	3.09
	H ₂ -rich gas compressor	4.43	1	4.43	MW _e consumed	2.8	2.8	3.69
Air Separation	ASU	61.3	1	61.3	tonne/hr pure O ₂	20.28	20.28	25.8
	O ₂ compressor	5.44	1	5.44	MW _e consumed	3.69	3.69	4.68
	N ₂ expander	2.57	1	2.57	MW _e produced	0.97	0.97	1.23
Power Island	Saturator	5.5	1	5.5	actual m ³ /s gas feed	0.12	0.12	0.18
	Gas Turbine	166.6	1	166.6	GT MW _e	39.39	39.39	50.03
	HRS + heat exchangers	317	1	317	MW _{th} heat duty ^b	36.73	36.73	56.4
	Steam cycle (turb. + cond.)	131.1	1	131.1	ST gross MW _e	44.4	44.4	56.39
Total Overnight Installed Capital Cost (million \$)				411				

(a) See corresponding note in Table 5 for descriptions of the column-heading parameters in this table.

- (b) Total heat transfer rate considering all heat exchange in the process, except for heat transfer in the syngas cooler and that associated with the WGS reactors. See note (hh) of Table 4 for additional discussion.

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