Lignite-plus-Biomass to Synthetic Jet Fuel with CO_2 Capture and Storage

Design, Cost, and Greenhouse Gas Emissions Analysis for a Near-Term First-of-a-Kind Demonstration Project and Prospective Future Commercial Plants

Final Report
to
The National Energy Technology Laboratory
U.S. Department of Energy
(contract number DE-FE0023697)

by

Chris Greig[a]*
Thomas G. Kreutz[b]
Eric D. Larson[b]
Johannes C. Meerman[b]
Robert H. Williams[b]

a) UQ Energy Initiative and Dow Center for Sustainable Engineering Innovation,
The University of Queensland, St. Luicia,
QLD, Australia

b) Energy Systems Analysis Group,
Andlinger Center for Energy & the
Environment, School of Engineering and
Applied Science, Princeton University,
Princeton, NJ, USA

12 June 2017

REVISED, 1 September 2017

---

* Prior to joining the University of Queensland in 2011, Chris Greig had a 25-year career in the private sector that began with the founding of a successful process technology and contracting company and continued with senior project and executive roles in the mining and energy industries internationally. The latter included being CEO of Australia's ZeroGen project, a 400 MW integrated coal gasification combined cycle plant with CO_2 capture and storage. After leading several years of design, engineering, and CO_2 storage appraisal efforts, Greig recommended cancelation of the project to the Australian government because of high projected costs.

Table of Contents

Acknowledgments and Disclaimer

For important contributions to this work, the authors thank:

- David Brubaker, Harvey Goldstein, Erich Mace, Lora Pinkerton, and Qinghua Xie (WorleyParsons Group, Reading, PA)
- Steve Baxley, Chao Chen, and Randy van Sickle (Southern Company Services, Birmingham, AL)
- Ed Gray, Tim Rooney, and Anneliese Schmidt (Antares Group)
- Cristina Elsido and Emanuele Martelli (Politecnico di Milano)

For review and discussion of an early draft of this report, the authors thank the following individuals at The National Energy Technology Laboratory: Jim Fisher, Kristen Gerdes, Jeff Hoffman, Matt Jamieson, Dale Kearns, Tim Skone, Morgan Summers, Tom Tarka, Charles White, Mark Woods, and Anthony Zinn.

Primary funding was provided by the U.S. Department of Energy's National Energy Technology Laboratory under award number DE-FE0023697. Southern Company Services and the University of Queensland provided cost-sharing contributions.

Abstract

We report on a 30-month design study for a first-of-a-kind (FOAK) demonstration plant that would be built at a site near Meridian, Mississippi, to coprocess lignite coal and woody biomass into jet fuel. The design uses an oxygen-blown TRIG™ gasifier developed by KBR and Southern Company. Fischer-Tropsch conversion of the syngas produces synthetic paraffinic kerosene (SPK) as the primary product, plus naphtha. Other co-products include electricity sold to the grid and CO_2 sold for use in enhanced oil recovery (EOR). Previous studies have identified coprocessing of various coals and biomass with CO_2 capture as promising options for cost-competitive production of low net lifecycle greenhouse gas (GHG) emissions synthetic fuels. The effort reported here goes beyond earlier studies in the level of detail in process design and cost estimation with the aim of improving the understanding of the economic prospects for lignite and woody biomass coprocessing systems. Key objectives in the design of the FOAK lignite/biomass-to-jet (LBJ) plant were to achieve net lifecycle GHG emissions for the SPK that are less than for conventional petroleum-derived jet fuel and to make process design and equipment selections such that the plant could be built and operated in the near term, e.g., before 2025. The process design was developed by researchers at Princeton University and the University of Queensland and validated by engineers at the WorleyParsons Group (WP). Commercial vendors provided design and cost estimates for several major plant components. Bare-erected capital costs and operating and maintenance costs were estimated by WP.

The primary objective in building and operating the FOAK plant would be to demonstrate the technical viability of the LBJ concept as an essential first step toward launching commercial-scale plants in the longer term. With this in mind, the project team developed a set of principles to help guide the process design: the total plant cost should be less than $2 billion (to limit investment risk); the level of input biomass should be both proportionally significant to reflect biomass/lignite co-gasification and sufficient to achieve GHG emissions goals; process design decisions and vendor/equipment selections should prioritize the likelihood of technical success over minimizing the cost of jet fuel production.

The resulting FOAK plant design capacity is 1,551 metric t/d (45.5% moisture) lignite and 556 t/d (43.3% moisture) biomass, for a total input of 295 MW_{HHV}, of which 25% is biomass. The design output is 1,252 actual barrels per day of liquids, of which 80% is SPK (62.3 MW_{LHV}) and 20% is naphtha (13.9 MW_{LHV}). It exports 15 MW_e of electricity after satisfying a 38 MW_e onsite auxiliary load. Additional products are 1,326 t/d of pressurized pipeline-quality CO_2 and 49 t/day of sulfuric acid (93 wt% H_2SO_4). With thinnings from sustainably-managed southern pine plantations as the biomass, and with captured CO_2 stored underground via EOR, the net lifecycle emissions for the SPK product are estimated to be about one-quarter of those for petroleum-derived jet fuel.

The bare-erected cost (BEC) estimated by WP for this plant is $588 million (2015$). The authors' best estimate of total plant cost (TPC) is $1,230 million, arrived at by assuming engineering, procurement and construction management services (20% of BEC), process

contingencies (35% of BEC), and project contingencies (35% of the sum of all other costs, i.e., 35% of 1.55xBEC).

Not surprisingly, an annual discounted cash flow (DCF) analysis determined that it would be impossible to generate a positive net present value (NPV) over a wide range in key input assumptions. An SPK production cost subsidy of nearly $400/bbl over a 20-year plant life would be required to achieve zero NPV under a baseline set of assumptions (including 3% real weighted average cost of capital and a levelized crude oil price of $80/bbl). Alternatively, a capital grant in excess of the TPC value would also achieve zero NPV. The poor financial results reflect the small scale of the plant, the design principle to prioritize technical success, the levels of contingencies appropriate for the relatively early stage of project development, and the first-of-a-kind nature of the plant.

Technology innovations, learning via construction and operating experience, alternative plant configurations, and larger scale should improve economics of future plants. To help understand these prospects, a preliminary analysis of Nth-of-a-kind (NOAK) plants was developed, but with the limitation that plants would use only equipment components that for the most part are already commercial today: consideration of advanced, not-yet-commercial technologies and of R&D-driven improvements in existing technologies were beyond the scope of this analysis. The analysis found that a variety of NOAK plant designs that coprocess lignite and woody biomass to make jet fuel are unlikely to be economically competitive without subsidy even in the presence of a high future carbon tax or equivalent greenhouse gas mitigation policy. This conclusion applies to process configurations and input biomass/lignite ratios that result in net GHG emissions as high as those for petroleum-derived jet fuel and as low as zero.

In contrast, encouraging results were found for plants processing only biomass. The economics of these “BECCS” plants (biomass energy with CCS) improve dramatically with the strength of carbon mitigation policies because of their strongly negative net GHG emissions. These findings do not imply that coal/biomass coprocessing strategies for making synfuels with CCS are not economically promising – only that, in the case of lignite, much more than a 25% biomass coprocessing rate would be needed.

However, our analysis shows that all such systems are unlikely to be economic in the absence of a strong carbon mitigation policy. Future R&D driven technological innovations could modify this conclusion. Among other R&D priorities, an emphasis on better understanding and reducing plant auxiliary loads is warranted.

Executive Summary

Introduction

The U.S. is endowed with vast coal resources of varying rank that offer the potential to enhance the long-term energy supply security of the United States. Enabling coal-to-liquids options in the near term would enhance energy security over the longer term as low-cost liquid hydrocarbon supplies become more scarce. Yet there is a risk that in a carbon-constrained world coal resources will become economically stranded assets without low-carbon-intensity options for conversion and utilization. Previous research publications have identified co-processing of various types of coal with biomass and CO_2 capture as a promising technology strategy for cost-competitive production of lower-carbon synthetic liquid substitutes for petroleum-derived jet fuel, diesel, and gasoline. The effort documented in this report aimed to go beyond earlier studies in terms of the level of detail in process design and cost estimation to further improve understanding of the economic prospects for systems coprocessing lignite with woody biomass.

This project assessed the prospective energy, carbon, and cost performance for a first-of-a-kind (FOAK) demonstration project that would co-gasify lignite coal and woody biomass in a Transport Integrated Gasification (TRIG^{TM}) reactor and synthesize Fischer-Tropsch jet fuel from the resulting syngas. The synthetic fuel would be a substitute for petroleum-derived jet fuel. Byproduct CO_2 would be captured and sold for use in enhanced oil recovery (EOR). Synthetic naphtha would be sold as a co-product, and some electricity in excess of onsite needs would be exported into the electric grid for some of the options considered. Key objectives in the design of this FOAK lignite/biomass-to-jet (LBJ) plant are that the net lifecycle greenhouse gas (GHG) emissions for the primary product be less than for conventional petroleum-based jet fuel and that plant process design and equipment selections be such that the plant plausibly could be built and operated in the near term, e.g., before 2025.

The process design for the FOAK facility was developed by Princeton University and University of Queensland researchers and validated by engineers at the WorleyParsons Group (WP). Designs for several key plant components were supplied by commercial vendors. The bare-erected capital cost and prospective operating and maintenance costs for the FOAK plant were developed by WP.

The primary objective in building and operating a FOAK plant would be to demonstrate in the near term the technical viability of an LBJ system with CO_2 capture and storage (CCS), as an essential first step toward launching commercial-scale plants in the longer term. With this in mind, the project team developed a set of principles at the start of the project to help guide the FOAK plant process design effort. Key among the guiding principles were the following:

- ***Plant Scale***. The total capital investment required should not exceed $2 billion in order to maximize the prospect for securing financing for the project. Within this constraint, the plant was designed to be of sufficient scale that the lessons learned from the construction and operation of this project would provide the information and confidence needed to move forward with a commercial-scale plant.

- ***Greenhouse Gas Emissions.*** The plant should be designed to produce jet fuel having fuel-cycle greenhouse emissions less than for petroleum-derived jet fuel.
- ***Biomass Utilization***. Biomass should be a non-negligible fraction of the input feedstock, but the fraction will not exceed 30%, the highest fraction successfully co-gasified to date in a pilot-scale TRIG^{TM} gasifier. (In the course of this project, engineers involved in the original design and development of the TRIG^{TM} system suggested that the technology would likely be capable of operating with 100% biomass with some further development.)
- ***Plant Outputs***. The primary product will be a synthetic jet fuel designed for 50/50 splash-blending with petroleum-derived jet fuel to meet specifications for semi-synthetic jet fuel approved under ASTM Standard D1655-14 for use in commercial aviation. Co-product naphtha will be processed onsite only to the extent needed for sale as a refinery feedstock. A combined cycle power island will supply all on-site electricity needs and export a non-negligible amount of electricity.
- ***Plant Process and Equipment Selections***. In process design trade-off decisions and in vendor/equipment selections will be to maximize the likelihood of technical success rather than to minimize cost of jet fuel production: *(i)* the use of novel or yet-to-be commercially deployed processes or equipment will be minimized, as will one-off specially designed plant items; *(ii)* off-the-shelf equipment packages or packages that could plausibly become off-the-shelf in commercial deployments following a FOAK demonstration will be preferred; *(iii)* field fabrication/erection of equipment will be minimized, and preference will be given to shop-assembled modules or modules with the potential for shop-assembly for future commercial deployments of LBJ systems; *(iv)* the plant will be designed for steady-state, fixed capacity operation, and plant availability is expected to increase with time over the operating life as issues causing loss of production are identified and corrected.
- ***Design / Operating Lifetime***. The plant will be designed for a 20-year life, but an operating life of 5 to 10 years is anticipated to be sufficient to demonstrate technical viability and provide key information for design of future commercial plants.

The selected site for the FOAK plant is a few miles due south of Meridian, Mississippi, at the site of Mississippi Power Company's Plant Sweatt generating station that is planned for retirement before 2020. The 650 acre site, largely forested, is about 30 miles south of the Kemper County IGCC project and will draw lignite from the Wilcox Group lignite formation that supplies the IGCC project. Woody biomass supplies are abundant in a 100-mile radius around the Plant Sweatt site. Residues from sustainable forest harvesting would be the preferred biomass feedstock based on cost, but consistently sizing these to the feed specifications for a TRIG^{TM} is currently challenging, and so to minimize technological risks with the FOAK LBJ plant debarked pulpwood-quality southern pine logs were selected for the feedstock. With additional development of feeding systems, milled residues could become a viable option.

FOAK Plant Process Design and Performance

Figure ES1 is an overview of the FOAK LBJ plant design. The plant receives lignite and pulpwood-grade logs that are separately subjected to sizing and drying before being fed to a pressurized oxygen-blown TRIG^{TM} gasifier. The resulting syngas is cooled, filtered, and scrubbed before a partial sour water gas shift (WGS) to set the H_2/CO ratio at the Fischer-Tropsch (F-T) synthesis inlet to the required value. CO_2, H_2S, and trace impurities are then removed using a chilled methanol solvent (Rectisol$^{®}$) at the acid gas removal (AGR) island. The CO_2 is compressed and delivered via pipe to the plant fence line. The captured H_2S is converted to wet sulfuric acid (WSA) for sale. Guard beds downstream of the AGR add a final layer of protection for the cobalt-based F-T synthesis catalyst. The analysis here assumes the captured CO_2 will be able to be sold into the currently-existing pipeline located about three miles from the LBJ site. This pipeline transports CO_2 from the Kemper Co. IGCC to an existing CO_2 pipeline network in the Jackson Dome, MS, area, which in turn delivers CO_2 to EOR operations on the Gulf Coast.

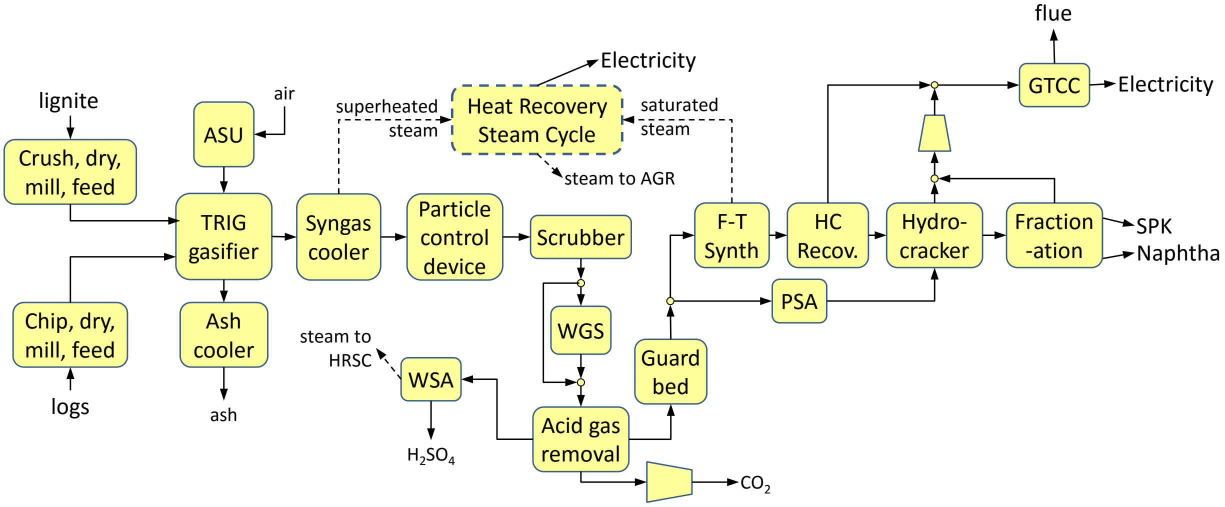


Figure ES1. FOAK LBJ plant simplified process configuration.

The F-T synthesis island includes a tubular fixed-bed reactor with a cobalt-based catalyst, followed by hydrocarbon recovery and processing to produce the primary liquid product (synthetic paraffinic kerosene, SPK) plus a synthetic naphtha coproduct.

The FOAK process design also includes coproduction of electricity. Fuel gas leaving the F-T synthesis island is the fuel for a gas turbine combined cycle. A separate steam cycle operates using process heat recovered primarily from syngas cooling after gasification and from cooling the exothermic FT reactions. This design for power generation was selected because it would enable use of an "off-the-shelf" GTCC package, and the GTCC could be fired with natural gas during plant start up or whenever the syngas production system is down for maintenance.

Tables ES1, ES2, and ES3 summarize the simulated performance of the FOAK plant. The plant processes 1551 t/d of lignite and 556 t/d of biomass (as-received basis) to produce 1250 actual barrels per day of product, of which 80% is SPK and 20% is naphtha. The plant exports 15 MW of electricity after satisfying a 38 MW onsite auxiliary load. Two-thirds of the latter is accounted for by air separation, CO_2 compression, and acid gas removal. Additional products are 49 t/day of sulfuric acid and 1326 t/d of pipeline-quality CO_2. The plant's energy conversion efficiency is a modest 35.6% (LHV basis), a result in part of the significant auxiliary load.

Table ES1. Summary of LBJ FOAK plant mass and energy balances.

Feedstock inputs		
Total feedstock input	MW, HHV	295
Biomass % of feedstock	% of HHV	25.0%
	metric t/d A.R.	1,551
	MW, HHV	221
	MW, LHV (A.R. basis)	193
	metric t/d A.R.	556
	MW, HHV	74
	MW, LHV (A.R. basis)	63
FT liquid outputs		
	actual bbl/day	1,004
	bbl/day equiv. petroleum jet	900
	MW, LHV	62.3
	actual bbl/day	248
	bbl/day equiv. petro-naphtha	223
	MW, LHV	13.9
	actual bbl/day	1,252
	bbl/day equiv. petroleum	1,123
	MW, LHV	76.2
Other products		
CO_2 to pipeline	metric t/day	1,326
Sulfuric acid (93 wt% H_2SO_4)	metric t/day	49
Electricity		
GTCC generation	MW_e	42.9
HRSC generation	MW_e	10.0
Total generation	MW_e	52.9
On-site use, MW_e	MW_e	38.2
Net electricity production	MW_e	14.7
Net electricity fraction	% of energy outputs, LHV	16.3%

Table ES2. LBJ FOAK plant auxiliary loads.

	MW_e
Lignite prep.	2.98
Biomass prep.	1.07
Gasifier	0.03
Acid gas removal system	5.18
Sour water stripper	0.02
CO_2 compressor	6.84
Wet Sulfuric Acid plant	0.12
H2 recovery	0.13
F-T island	0.76
GT fuel gas compressor	1.56
Air separation unit	14.84
Tempered water system	0.08
Cooling water system	2.05
Water treatment system	0.47
Wastewater treatment	0.39
Miscellaneous	1.72
TOTAL	38.21

Table ES3. LBJ FOAK plant carbon flows.

	metric tCO_{2e}/day
Lignite input	1,792
Biomass input	622
TOTAL INPUT	2,414
C in SPK product	378
C in naphtha	83
CO_2 captured	1,326
CO_2 vented	538
C in gasifier ash	57
TOTAL OUTPUT[a]	2,382

a) Carbon balance closure is within bounds allowed by gasifier performance estimates available to the authors.

FOAK Project Greenhouse Gas Performance

One goal for the LBJ project is to produce synthetic jet fuel (SPK) having net lifecycle greenhouse gas (GHG) emissions less than for petroleum-derived jet fuel. Figure ES2 shows the GHG flows that were considered in estimating the net lifecycle emissions. Emissions to the atmosphere include CO_2 in flue gases at the plant and in F-T fuels combustion products. Also included are GHG emissions associated with *(i)* biomass land preparation, harvesting and transporting, *(ii)* lignite mining and delivery, *(iii)* delivering the liquid products to users, *(iv)* delivering CO_2 to the pipeline network that currently carries natural CO_2 from the Jackson Dome are to oil fields. Emissions from combustion of fuels derived from the petroleum produced via EOR are not charged as emissions of the LBJ system. Biomass growth results in net removal of CO_2 from the atmosphere, including some transfer of carbon into the soil. Some of the latter will return as CO_2 to the atmosphere via decomposition of roots. Net electricity exported to the grid results in avoided grid emissions when that electricity displaces fossil fuel generated electricity on the grid.

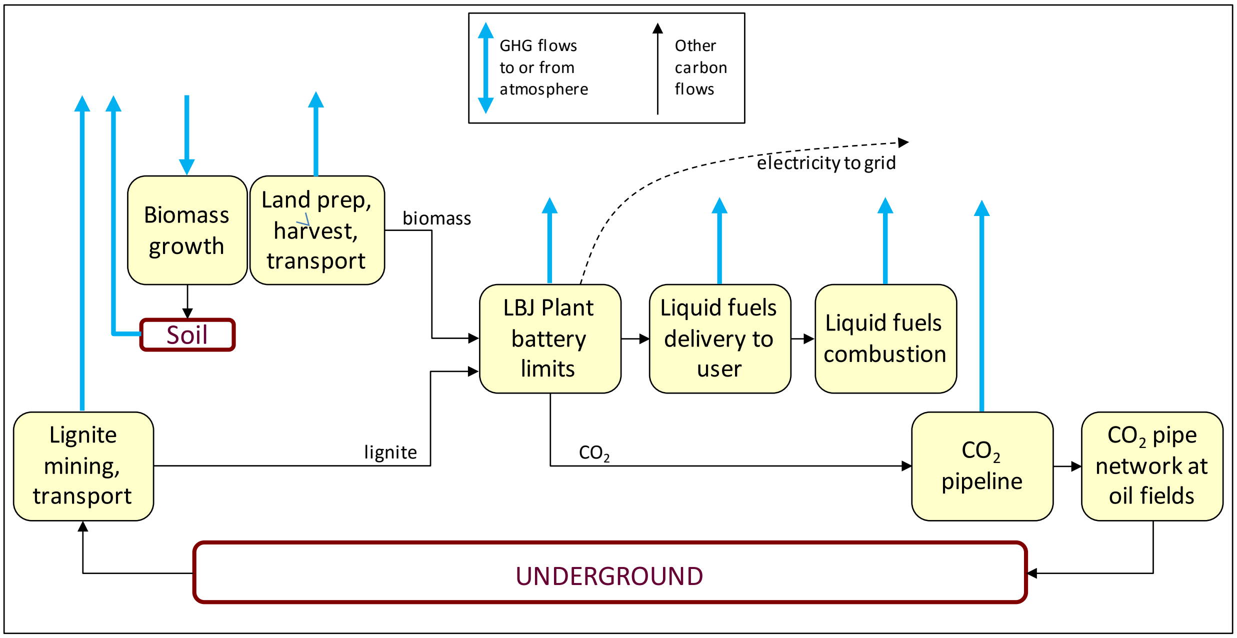


Figure ES2. Greenhouse gas flows considered in estimating GHG emissions of the SPK produced by the LBJ FOAK plant.

System GHG emissions were quantified with two metrics: Greenhouse Gas Emissions Index (GHGI) and GHG emissions charged to SPK in units of $kgCO_{2e}/GJ_{SPK,LHV}$. The dimensionless GHGI is defined as the net GHG emissions for the LBJ system divided by the lifecycle emissions for a reference system that produces an equivalent amount of products by conventional means, with 2005 taken as the reference year for the denominator, since U.S. national carbon mitigation goals as of 2016 were expressed relative to emissions in 2005:

$$GHGI_{2005} = \frac{\text{Net LBJ system emissions}}{\text{Average U.S. lifecycle emissions in 2005 for equivalent products}}$$

GHGI is a particularly informative metric when comparing processes that have different mixes of co-products because it does not require allocating emissions among co-products. The second metric, GHG emissions charged to SPK having units of $kgCO_{2e}/GJ_{SPK,LHV}$, is defined as follows:

$$SPK\ emissions\ \left[\frac{kgCO_{2eq}}{GJ_{SPK,LHV}}\right] = \frac{\text{Net LBJ system emissions - emissions allocated to (electricity + naphtha + CO2 for EOR)}}{\text{energy content of LBJ SPK}}$$

This metric allocates a fraction of the net LBJ system emissions to each co-product and attributes the remaining emissions to the main product, SPK. We chose to allocate co-product emissions equivalent to emissions that would be avoided when the coproducts displace an equivalent amount of conventionally-produced products. For the electricity co-product, we allocate emissions associated with grid-generated electricity that it would displace. The naphtha coproduct is assumed to be used for producing gasoline blending components and assigned lifecycle emissions associated with an energy-equivalent amount of petroleum-derived gasoline. The CO_2 for EOR is assumed to displace natural dome CO_2 that would otherwise have been used.

A key component of lifecycle GHG emissions is emissions of biogenic origin. In contrast to biomass feedstocks that grow on an annual cycle (e.g., perennial grasses or agricultural residues), estimating emissions for the tree-derived feedstock for the LBJ plant is complicated by the slower regrowth time for trees. The biomass feedstock supply strategy for the LBJ project is based on analysis by the Antares Group of the forests that cover about 200 million acres of the U.S. South. About 60% of this area are family forests (non-industrial private forests, NIPFs), and the proposed LBJ feedstock supply strategy relies on NIPF owners intensifying their management practices to provide feedstock for the project. NIPFs make up about 70% of forest acreage in Mississippi and Alabama, and at least half of these NIPF acres are judged to be good candidates for making cost-effective productivity enhancements. These improvements lead to additional biomass production from stands that would otherwise not produce as much biomass under “business-as-usual” management practices. This result provides the potential for GHG emissions benefits when some or all of the additional biomass is used for the LBJ project.

Productivity-improvement measures assessed by Antares were mid-rotation thinning followed by fertilizer application, control of competing vegetation, and improvement of planting stocks. Three alternative combinations of these measures were quantified and labeled according to the level of productivity enhancement achieved: low, medium, and high. Each of the three scenarios involves thinning the stand in year 15 of its 25-year rotation, with the thinnings providing the LBJ feedstock. There are wide variations in potential productivity gains across different regions of the South due to intrinsic biogeophysical and climate differences. Antares quantified low, medium, and high scenarios for each of four regions of the U.S. South. Mississippi and Alabama constitute one of these regions, and results for this region provided the basis for estimating biogenic emissions for the LBJ FOAK project. The analytical framework proposed by the U.S. Environmental Protection Agency for assessing biogenic CO_2 emissions from stationary sources was adapted to quantify the LBJ project’s biogenic emissions.

For all three pine plantation management-intensity scenarios, $GHGI_{2005}$ is well-below one (Figure ES3). For the Low scenario, the net SPK emissions per unit of energy are about one-quarter of those for conventional jet fuel, and they are negative for the Medium and High scenarios (Figure ES4).

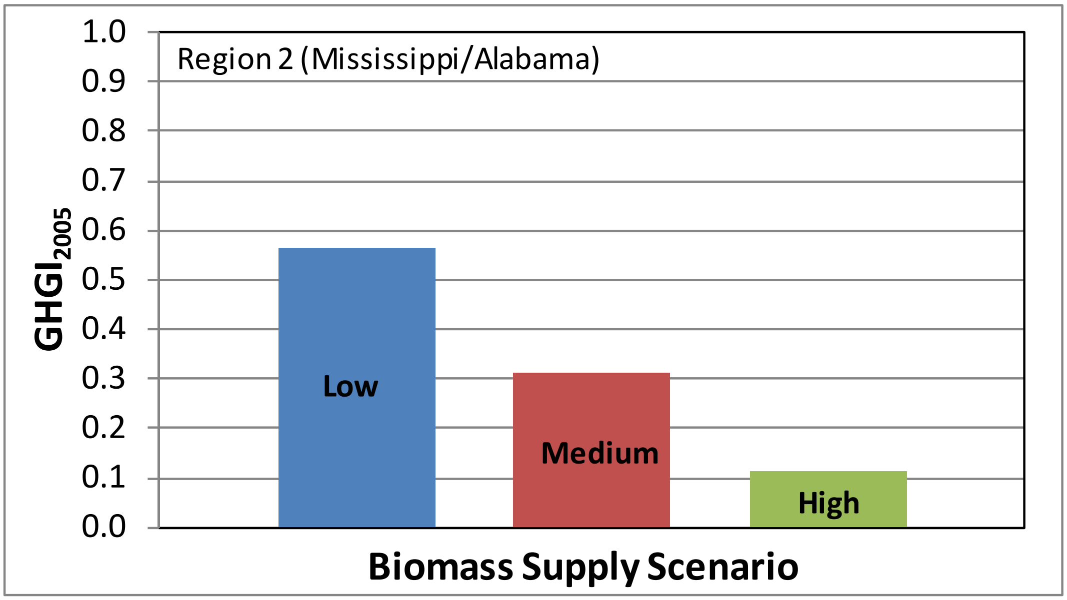


Figure ES3. $GHGI_{2005}$ for the FOAK LBJ system.

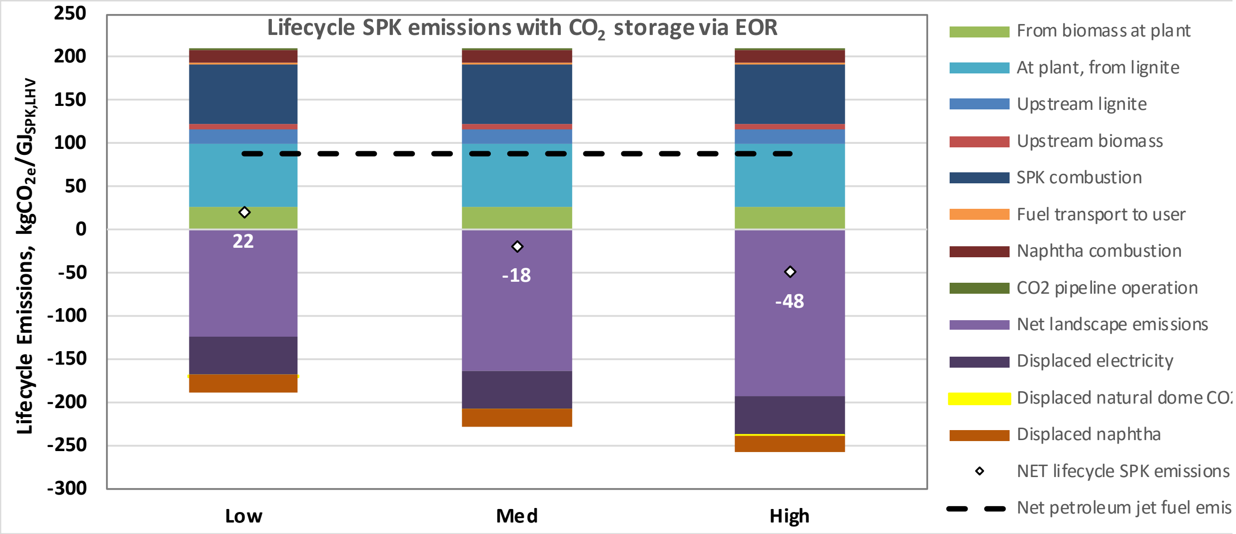


Figure ES4. Net lifecycle emissions (white diamonds and superimposed white numbers) and component emissions associated with SPK production by the FOAK LBJ system for Low, Medium, and High pine plantation management-intensity scenarios. Net lifecycle emissions for conventional jet fuel are shown for comparison.

LBJ FOAK Project Financial Analysis

The WorleyParsons Group (WP) was contracted to prepare design documentation and develop capital and operating cost estimates for the LBJ FOAK plant. They estimated a bare-erected cost (BEC_{WP}) of \$588 million (2015\$) to which they added allowances for EPCM (20%

of BEC_{WP}), process contingencies (15% of BEC_{WP}), and project contingences [25% of (BEC_{WP} + EPCM + process contingencies)]:

$$TPC_{WP} = 1.7 \text{ x } BEC_{WP} = \$ \text{ } 992 \text{ million}$$

Two additional TPC estimates were included in the FOAK plant financial analysis to help assess the impact of the inherent, but difficult to quantify, scope risk and uncertainties in capital cost estimation.

The TPC considered the most likely TPC by the authors (TPC_{LBJ}) assumed allowances, as a percentage of BEC_{WP}, of 20% for EPCM and 35% for process contingencies, plus a 35% project contingency applied to the sum of other costs:

$$TPC_{LBJ} = [(1+0.2+0.35) \text{ x } (1+0.35)] \text{ x } BEC_{WP} = 2.1 \text{ x } BEC_{WP} = \$ \text{ } 1{,}230 \text{ million}$$

The third TPC estimate includes an additional 20% allowance for “supplementary funds” (SF) intended to cover risks of unforeseen events or of outcomes that add to the most-likely cost outcome (TPC_{LBJ}). The SF may not need to be spent, but technology performance, gross scope growth arising from local or market issues, regulatory changes, process integration challenges, slow plant ramp-up to full capacity, redesign and rectification requirements, or other issues may arise, and the project would be unable to proceed if SF are not available to address such issues.

$$TPC_{LBJ \text{ w/SF}} = [(1+0.2+0.35) \text{ x } (1+0.35)\text{x}(1+0.2)] \text{ x } BEC_{WP} = 2.5 \text{ x } BEC_{WP} = \$ \text{ } 1{,}475 \text{ million}$$

The above TPC estimates were incorporated, with other costs, into an annual discounted cash flow (DCF) analysis to assess the financial viability of a FOAK LBJ project. That analysis determined that it would be impossible to generate positive cash flows from the anticipated revenues for a baseline set of assumptions with any of the three TPC estimates in the absence of a subsidy.

To achieve a zero net present value (NPV) under the baseline assumptions and with TPC = TPC_{LBJ} would require an SPK production cost subsidy over the assumed 20-year life of the project of \$388/bbl above and beyond the market revenues of \$107/bbl (levelized) assumed to be received for the SPK. The required production subsidy to achieve zero NPV would be \$306/bbl with TPC = TPC_{WP} and \$472/bbl for TPC = $TPC_{LBJ, \text{ w/SF}}$. Alternatively, zero NPV could be achieved with a grant or other mechanism to reduce the capital cost to investors. With zero production cost subsidy, the required capital grant would exceed the TPC with any of the three TPC estimates. The poor FOAK-project economics are not surprising. They reflect the small scale of the plant, the guiding design principle to prioritize technical success over minimized cost, the levels of contingencies and other costs appropriate for the relatively early-stage of project design, and the first-of-a-kind nature of the project.

Preliminary Analysis for Nth-of-a-Kind (NOAK) Commercial Plants

Technology innovations, learning via construction and operating experience, and scaling-up of plant size would be expected to improve the economics of future plants. To help understand the extent to which economics might improve, some analysis of Nth-of-a-kind (NOAK) plants was undertaken. NOAK process designs and preliminary capital cost estimates were developed for plants using equipment components similar to those selected for the FOAK design: consideration of advanced, not-yet-commercial technologies and of R&D-driven improvements in existing technologies were considered beyond the scope of this analysis. Instead, the focus was on understanding the implications for performance and economics of *(i)* increased plant scales, *(ii)* capital and operating cost reductions resulting from "learning by doing", *(iii)* alternative process configurations, *(iv)* alternative lignite:biomass input ratios, and *(v)* how the economics of systems will vary with crude oil price and cost of GHG emissions (the two most important exogenous parameters influencing NOAK economics). The process simulations developed for the FOAK analysis provided a starting point for simulating NOAK process designs, and the FOAK bare-erected cost estimates provided a quantitative basis for scaling to estimate cost components for NOAK plants.

One design objective for the NOAK plants was to make them as large as possible to capture scale economy benefits. For plant designs that use some biomass, the plant scale was determined by two constraints: total annual biomass input to the plant would be limited to one million tonnes (dry basis) per year, and the amount of lignite co-feed would be set according to the desired GHG footprint for the system, as quantified by its $GHGI_{2005}$. For NOAK GHGI calculations, biogenic carbon emissions were estimated as the average for the Southeastern region as a whole under the medium scenario for improved productivity of NIPF stands. One set of NOAK plants was designed for $GHGI_{2005} = 1.0$, a second set was designed for $GHGI_{2005} = 0$, and a third set was designed to use only biomass as the feedstock, which maximizes negative GHG emissions (negative $GHGI_{2005}$).[a]

Three NOAK plant configurations were assessed, identified with acronyms OT (for "once-through"), RC (for "syngas recycle"), and OTA (for "once through with autothermal reforming"). See Figure ES4a. The three vary by the design of their "light ends" processing (LEP) area, i.e. by how H_2- and hydrocarbon-rich offgas streams from the F-T island are handled (Figure ES4b). OT is the same configuration as employed in the FOAK plant design (Figure ES1). It is a relatively simple design that produces both synfuels and power but maximizes neither synfuels production nor CO_2 capture. The OTA designs are similar to the OT designs except that more than 70% of the carbon input as biomass or coal is captured. OTA is a configuration designed to maximize CO_2 capture by removing carbon from the GTCC fuel gas stream prior to combustion in the gas turbine. The RC design increases the yield of synfuels relative to the OT and OTA designs at the expense of power generation. In all RC plants considered here, the recycle rate was set such that net output power (after meeting the onsite auxiliary load) is zero.

Aspen Plus process simulations were developed for eight NOAK plant designs. For all of these, the unit performance of the upstream components (feed preparation through acid gas removal ahead of the synfuels island) are identical to those for the FOAK design, with the

---

[a] The TRIG^{TM} gasifier, which was originally developed for low-rank coals, was subsequently demonstrated to work well with low-rank coals and a co-feed of up to 30% biomass. In the course of this project, informal discussions with engineers involved in the original design, scale-up, and coal/biomass co-processing trials of the TRIG^{TM} suggested that the technology would be capable of operating solely using biomass. Testing would be required to verify this.

exception of the auxiliary power consumption of the ASU, which was reduced by about 25% from that for the FOAK plant based on discussion with engineers at Southern Company Services. The simulations of downstream processes were adjusted as demanded by the LEP configurations. Efficiency increases were included as a function of scale for gas turbines and steam turbines in the power island.

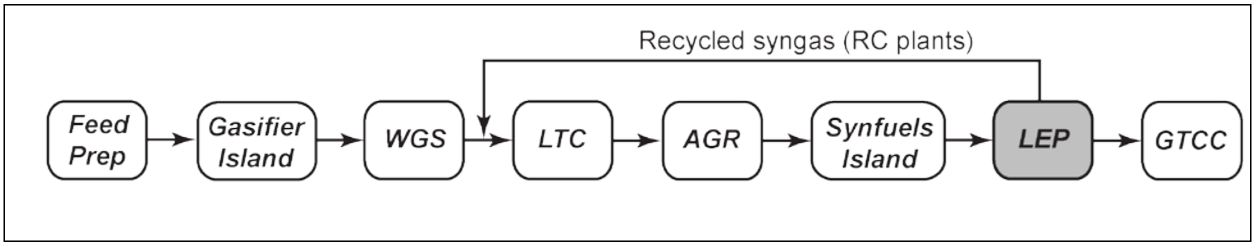


Figure ES4a. Simplified schematic of NOAK plant designs. The OT and OTA configurations exclude the recycled syngas stream.

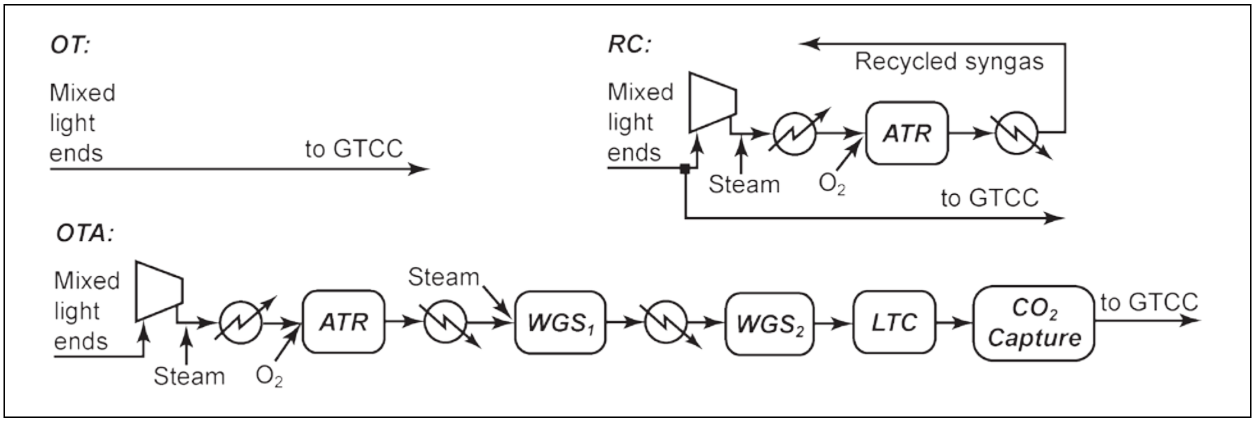


Figure ES4b. Design of light ends processing (LEP) for OT, RC, and OTA configurations.

Performance simulation results are shown in Table ES4. The plants with $GHGI_{2005}$ = 0 are moderately sized (11-16 $kbbl_{eq}$/day of F-T liquids) due to the scale constraints imposed by the biomass input fractions required to achieve the target GHGI value. The group of plants with $GHGI_{2005}$ = 1 are much larger plants (48-61 $kbbl_{eq}$/day of FTL) because the relaxed GHG emissions constraint allows much more lignite to be consumed for the given fixed biomass input. The plants using 100% biomass as input are all relatively small (2.7-3.5 $kbbl_{eq}$/day of FTL) and all have strongly negative GHGI values.

To assess the prospective economics of NOAK plants, capital and operating costs were developed using a simplified scaling approach that also assumes significant learning (cost reduction) in evolving from FOAK to NOAK plants: 1) EPCM fees are assumed to fall to 12% of the bare-erected cost (BEC) from 20% for the FOAK plant; 2) process contingencies fall from 35% to zero; 3) project contingencies fall from 35% to 10% of *BEC*; and 4) supplementary funds are not needed. The result is that TPC is 1.22×BEC, compared with 1.7×BEC to 2.5×BEC for the FOAK plant (excluding owner's costs).

Because vender quotes and WorleyParson's internal capital-equipment cost database used to construct the FOAK estimate represented the most credible cost estimates available to the authors, and because all components used in the NOAK plants are already commercially mature with the exception of the TRIG^{TM} gasifier, component-level BEC for the NOAK plants were estimated by appropriately scaling component costs estimated for the FOAK design. The cost-scaling exercise involved an overall scale up by a factor of about 2 for the biomass-only designs, about 10 for the GHGI = 0 designs, and 32 to 50 for the GHGI = 1 designs. Table ES5 shows the resulting BEC_{NOAK} estimates.

Table ES4. Performance simulation results for eight NOAK plant designs.

	$GHGI_{2005}$ = 0			***$GHGI_{2005}$ = 1***		***Biomass-only cases***		
Plant Configuration	***OT-0***	***OTA-0***	***RC-0***	***OT-1***	***RC-1***	***OT-B***	***OTA-B***	***RC-B***
Dry biomass, Mtonne/yr	1	1	1	1	1	1	1	1
Biomass fraction, % HHV	25.3	17.5	23.8	4.6	7.8	100	100	100
Performance Metrics								
Liquids fraction, % LHV	73.6	86.8	100	73.2	100	75.1	89.8	100
Electricity fraction, % LHV	26.4	13.2	0	26.8	0	24.9	10.2	0
First law efficiency, % LHV	41.5	35.2	41.1	41.8	41.4	40.2	33.6	39.6
$GHGI_{2005}$	0	0	0	1	1	-3.94	-5.94	-5.14
Plant Inputs								
AR lignite, MW_{th} HHV	2,101	3,342	2,281	14,723	8,357	0	0	0
AR biomass, MW_{th} HHV	711	711	711	711	711	711	711	711
Total AR feedstock, t/day	20,121	28,831	21,383	108,747	64,046	5,365	5,365	5,365
Plant Outputs								
SPK, MW_{th} LHV	609.4	881.3	872.5	3,375.9	2,675.5	149.1	149.1	195.7
Naphtha, MW_{th} LHV	136.0	196.7	194.7	753.5	597.1	33.3	33.3	43.7
Total FTL, bbl_{eq}/day	10,986	15,886	15,728	60,854	48,228	2,687	2,687	3,529
Gross power output, MW_e	598.2	761.9	392.4	3317.2	1189.8	145.3	127.1	93.1
Auxiliary load, MW_e	-331.0	-598.6	-392.3	-1,806.7	-1,190.0	-84.7	-106.4	-93.0
Net power output, MW_e	*267.1*	*163.4*	*0.1*	*1,510.6*	*-0.1*	*60.5*	*20.7*	*0.2*
CO_2 captured, tonne/day	12,635	24,114	14,317	68,129	42,941	3,392	4,455	3,573
Carbon disposition (% of input)								
In CO_2 vented at plant	23.6	5.4	13.2	23.7	12.9	23.1	5.4	14.4
In FTL	19.6	19.7	26.3	19.9	26.8	18.4	18.4	24.1
In CO_2 captured	54.8	72.8	58.4	54.3	58.2	56.5	74.2	59.5
In gasifier ash and tar	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0

Table ES5. Estimated NOAK plant bare erected costs (2015$).

	OT-0	***OTA-0***	***RC-0***	***OT-1***	***RC-1***	***OT-B***	***OTA-B***	***RC-B***
Lignite handling & drying	363	520	383	1,826	1,117	0	0	0
Biomass handling & drying	179	192	179	268	233	138	138	138
Air Separation Unit (ASU)	373	641	472	1,709	1,226	120	140	133
Gasifier & ash handling	622	837	647	2,714	1,688	204	204	204
Gas cleanup (AGR, SWS, WSA)	658	1,131	706	2,949	1,932	205	238	211
CO_2 compression	80	145	87	351	235	28	34	29
FT synthesis & refining	343	439	436	1,078	923	133	133	160
Light ends processing (LEP)	0	116	55	0	117	0	35	19
GT + GT HRSC	240	287	132	1,170	366	69	58	40
Custom HRSC	39	51	41	132	90	14	14	14
Balance of plant (BOP)	324	415	338	1,028	704	147	151	148
Total Plant BEC, M$	*3,220*	*4,775*	*3,474*	*13,224*	*8,629*	*1,059*	*1,146*	*1,098*

Financial performances of the NOAK options were quantified in terms of alternative cost-effectiveness metrics, including the breakeven crude oil price (BEOP), which is the market price for crude oil that would result in a refiner's wholesale price of equivalent petroleum-derived fuels being equal to the levelized FTL production cost. For the GHGI = 1 plants, the estimated BEOP is \$125/bbl or higher across the full range of GHG emission prices considered (0 to \$200/tCO_{2e}). Plants with GHGI = 0 have higher BEOPs than GHGI = 1 plants when the GHG emissions price is zero. However, because their GHG emissions are significantly lower than for petroleum-derived fuels, BEOP declines as GHG emissions price increases. In the RC-0 case, the BEOP reaches \$100/bbl for a GHG emissions price of about \$135/tCO_{2e}. BEOPs for the 100% biomass plants are even higher than those of the GHGI = 0 plants when the GHG emissions price is zero. However, because of the strongly negative GHGI values, the BEOPs fall much more quickly with increasing GHG emissions price, reaching zero BEOP for any of the three designs (OT-B, OTA-B, RC-B) as the GHG emissions price approaches \$100/tCO_{2e}, i.e., at this price, CO_2 storage with these "BECCS" options is worthwhile even if crude oil has zero value.

Figure ES5 encapsulates the main findings of the NOAK cost analysis. It shows BEOP for RC-0 and RC-B designs with CO_2 storage via both enhanced oil recovery (EOR) and in deep saline formations (DSF). The latter are available across much of the U.S., unlike more geographically-limited EOR opportunities. At zero GHG emissions price, the BEOP for RC-0/DSF is considerably lower than for RC-B/DSF due to the larger scale of RC-0 and the lower average cost per unit feedstock, but the BEOP for RC-0 still exceeds \$200/bbl at zero GHG emissions price and falls only modestly at higher prices; the RC-0 plant is unlikely to be a competitive option for the foreseeable future. Although the BEOP for RC-B with DSF storage is very high at zero GHG emission price, it drops rapidly with increasing GHG emissions price, reaching \$80/bbl when the emissions price is about \$80/tCO_{2e} a price in the 2030-2050 period that would be consistent with limiting average global temperature rise by 2100 due to greenhouse gases to 3°C above pre-industrial. Thus, RC-B and alternative biomass-only plant designs that perform similarly may be viable options for synthetic jet fuel production even in a relatively modestly carbon-constrained future.

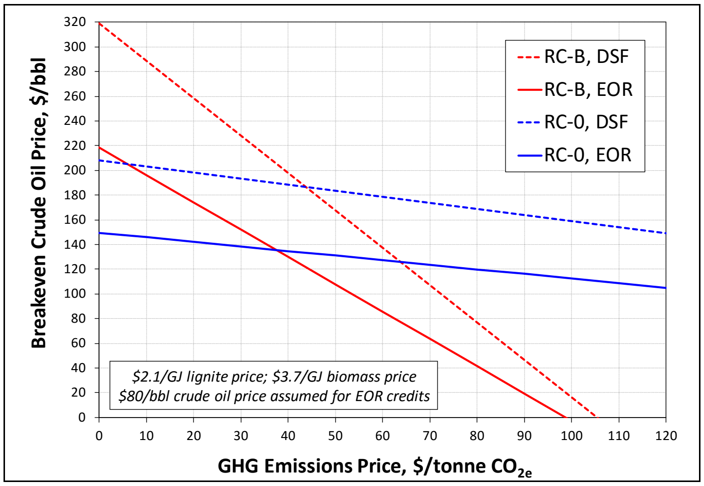


Figure ES5. Breakeven crude oil prices for RC-0 and RC-B plants with CO_2 storage in deep saline formations (DSF) or with enhanced oil recovery (EOR).

Comparisons of Large NOAK Coal-to-Liquids Plants

A number of design studies for 50,000 bbl/day coal-to-liquids (CTL) plants employing CCS have been published over the past decade. Seeking to gain insights into the FOAK and NOAK results from this project, two NOAK CTL plant designs, one using lignite and one using sub-bituminous Powder River Basin (PRB) coal, were developed for comparison with prior studies. The comparisons proved informative. In particular, auxiliary load estimates for our NOAK CTL plants were considerably higher than those in other published studies: the ratio of the auxiliary load to the coal input rate ($MW_e/MW_{coal,HHV}$) in our designs were 0.118 and 0.125, compared with a range from 0.071 to 0.088 in prior studies. In addition, the bare-erected cost (BEC) for our PRB-based plant was 70% higher than the BEC from a prior study for a plant that also used PRB coal and TRIG gasification. The difference in auxiliary power needs at least partially explains the difference in BEC estimates, since for a given liquids output more feedstock must be processed in order to meet onsite power needs. Differences in equipment choices in the two studies, especially for acid gas removal, provided a partial, but not full resolution of the difference in auxiliary load estimates. For future design studies, careful assessment of auxiliary loads is recommended.

Conclusions

This study described and assessed the design of a first-of-a-kind (FOAK) process that would be technically feasible to construct today to coprocess lignite and woody biomass in a 75/25 mix into synthetic Fischer-Tropsch jet fuel and capture byproduct CO_2 for sale for enhanced oil recovery (EOR). For the resulting jet fuel to have a smaller greenhouse gas (GHG) footprint than petroleum-derived jet fuel requires that well-understood productivity enhancements be implemented on the sustainably managed tree stands from which the plant's biomass feedstock supply will originate. The productivity enhancements ensure that additional biomass is produced from these stands compared with "business-as-usual" management practices.

For the design of the FOAK LBJ facility, the approach taken was to make the plant large enough to give investors confidence in the performance of subsequent commercial-scale plants but small enough that the plant might actually be financed and built – even though it was expected *a priori* that such a project would not be undertaken by a private firm without a substantial government subsidy. The analysis found that without a subsidy, the levelized cost of producing synthetic jet fuel in the FOAK LBJ facility would approach $500 per barrel.

Beyond the detailed analysis conducted for a FOAK demonstration project, this study also made a preliminary assessment of the financial viability of future Nth-of-a-kind (NOAK) commercial LBJ plants. For this analysis advanced, not-yet-commercial technologies and R&D-driven improvements in existing technologies were not considered.

A major finding is that NOAK plants that coprocess lignite and biomass to make jet fuel and have zero or higher net greenhouse gas emissions are unlikely to be economically competitive based on any plausible combinations of future crude oil and greenhouse gas emissions prices. This finding arises in large part because even for NOAK plants capital costs are high, in large part because of high auxiliary power needs – suggesting the strategic importance of R&D aimed at reducing plant auxiliary loads.

In contrast, encouraging economic results were found for plants processing only biomass in the presence of a strong carbon mitigation policy – even though such plants were also found to have large auxiliary power needs. The economics of these "BECCS" (biomass energy with CCS) plants improve dramatically with the strength of carbon mitigation policies because of their strongly negative net greenhouse gas emissions. Pure biomass synfuel plants were found to be economically interesting at crude oil and GHG emissions prices consistent with carbon mitigation goals of limiting global warming to no more than 2°C to 3°C – not only for CO_2 EOR applications, but also when captured CO_2 is stored in deep saline formations, and even in the absence of "optimistic" NOAK costing assumptions. Although the analysis shows that such systems are unlikely to be economic in the absence of a strong carbon mitigation policy, future R&D-driven technological innovations could modify this conclusion.

These findings do not imply that coal/biomass coprocessing strategies for making synfuels with CCS are not economically promising – only that, in the case of lignite, much more than a 25% biomass coprocessing rate would be needed. Such lignite/biomass coprocessing options, as well as options coprocessing biomass with higher ranked coals (which will generally have lower auxiliary loads than with lignite) should be investigated in future studies.

1 Introduction

The U.S. is endowed with vast coal resources of varying rank that offer the potential to have a substantial impact in enhancing the long-term energy security of the United States. Conversion options include making electricity (the dominant US coal conversion today), liquid fuels, and/or chemicals. In particular, enabling coal-to-liquids options in the near term would enhance energy security over the longer term as low-cost liquid hydrocarbon supplies become scarcer. Yet there is a risk that in a carbon-constrained world coal resources will become economically stranded assets in the absence of low-carbon-intensity options for conversion and utilization. Previous work published at the National Energy Technology Laboratory (NETL),[1,2,3] at Princeton University,[4,5,6] and elsewhere[7,8] identified co-processing of coal and biomass with CO_2 capture as a promising technology strategy for cost-competitive production at scale of lower-carbon synthetic liquid substitutes for petroleum-derived jet fuel, diesel, and gasoline.

The 30-month effort documented in this report goes beyond these earlier studies in terms of the level of detail in process design and cost estimation to further improve understanding of the economic prospects for co-processing of coal and biomass (specifically lignite coal and woody biomass in this work) to co-produce liquid transportation fuels and electricity while capturing byproduct CO_2 for geologic storage. The effort assessed the prospective energy and carbon performance and costs for a first-of-a-kind (FOAK) demonstration project that would use component technologies that are nearly all commercially proven or demonstrated, i.e., of a project that could be built and operated in the near term. Such a demonstration project would be an essential first step toward launching commercial-scale plants in the longer term.

The goal of a FOAK project would be to demonstrate the technical viability of a facility that would co-gasify lignite coal and woody biomass using Transport Integrated Gasification (TRIG^{TM}) technology and produce as its primary product synthetic Fischer-Tropsch jet fuel suitable for neat use or for blending with petroleum-derived jet fuel while capturing byproduct CO_2 to be sold for use in enhanced oil recovery operations in the Gulf Coast region. Synthetic naphtha would be sold as a co-product, and some electricity in excess of onsite needs would be generated for export into the electric grid.

A key objective in the design of this FOAK lignite/biomass-to-jet (LBJ) plant is that the lifecycle greenhouse gas (GHG) emissions for the primary product be less than for conventional petroleum-based jet fuel, as stipulated in Section 526 of the Energy Independence and Security Act of 2007 for all alternative fuels purchased by federal agencies, including the Department of Defense.

The process design for the FOAK facility was developed by Princeton University and University of Queensland researchers and validated by engineers at the WorleyParsons Group (WP). Designs for several key plant components were supplied by commercial vendors. The bare-erected capital cost and prospective operating and maintenance costs for the FOAK plant were developed by WP. The intended site for the FOAK plant is near Meridian, Mississippi, at the Mississippi Power Company's Plant Sweatt generating station. The generating equipment currently at this site is planned to be retired before 2020.

The financial viability of the FOAK project was analyzed for estimated total investment cost levels that reflect levels of contingencies and other costs appropriate for the relatively early-stage of project design and its first-of-a-kind nature. The expectation at the start of the work was that the FOAK project would not be financially viable without subsidy because of its relatively small scale and FOAK nature. However, technology innovations, learning via operating experience, and scaling-up of plant size would be expected to improve the economics of future plants. While the bulk of the work described in this report was concerned with the design and analysis of the FOAK project, preliminary projections of the technical performance, carbon footprint, and economics of potential future commercial-scale N^{th}-of-a-kind (NOAK) plants were also developed. This latter work included exploring the impact of plants re-configured for different lignite:biomass input mixes and/or different fuel:electricity output mixes.

Taken as a whole, the present work represents an important step forward in understanding the prospects for evolving to commercial production and utilization of low-carbon-intensity coal-derived jet fuels.

2 Guiding Principles for Project Design

Near the start of the project, the project team developed a set of principles to help guide the process design work and achieve clarity and alignment among different stakeholders. The principles were intended to help establish the approach to take in project design, including process design, main equipment selection, level of automation and optimization, feedstock supply, product off-takes, etc. They also helped guide decision making during the design process, especially in relation to cost trade-offs against performance and risk, e.g., cost vs. equipment availability, cost vs. operating flexibility, cost vs. product flexibility, cost vs. efficiency, capital vs. operating costs, etc.

The following guiding principles were developed for the FOAK demonstration project:

Plant Scale. The judgment was made that it would be prudent to constrain the scale of the LBJ project such that the total capital investment required would be no more than about $2 billion. More capital than this for a FOAK demonstration project would be difficult to secure. As a point of reference, the total investment in the Cool Water FOAK integrated gasification combined cycle project was under $600 million (converted to 2012$).[9] Within this capital investment constraint, the plant will be designed to be of sufficient scale that the lessons learned from the construction and operation of this project will provide the information and confidence needed to move forward with a commercial-scale plant. The FOAK plant should have a commercially-relevant jet fuel production capacity, i.e., at least 1,000 barrels per day, with a likely upper limit of about 5,000 bpd.

Greenhouse Gas Emissions. The plant will be designed such that the jet fuel product will have fuel-cycle greenhouse emissions that are less than for petroleum-derived jet fuel, as required for federal government procurement under Section 526 of the Energy Independence and Security Act of 2007.

Biomass Utilization. Biomass will be co-fed with lignite in a fixed, non-negligible fraction not to exceed 30%. It was expected (and confirmed by further analysis) that this level of biomass would be more than sufficient for the resulting jet fuel to satisfy EISA Section 526 requirements for GHG emissions. The maximum of 30% was set because this is the highest fraction demonstrated to date in a TRIG^{TM} gasifier (co-fed with sub-bituminous coal[10]) at the National Carbon Capture Center in Wilsonville, AL. Co-feeding of biomass with lignite has also been demonstrated, with up a biomass input fraction of 17%[11]. In the course of this LBJ project, informal discussion between the authors and the engineers involved in the original design, scale-up, and coal/biomass co-processing trials of the TRIG^{TM} gasifier suggested that the technology would be capable of operating using 100% biomass. Testing would be required to verify this.

An additional guideline regarding biomass supply was that the annual biomass quantities required at the plant will not exceed the prospectively-available quantities at reasonable cost within a 100 mile radius of Plant Sweatt, as identified by Antares in a 2010 study for the Electric Power Research Institute.[12]

Plant Products. The primary product will be a synthetic jet fuel designed for 50/50 splash-blending with petroleum-derived jet fuel to meet specifications for semi-synthetic jet fuel approved under ASTM Standard D1655-14 for use in commercial aviation.

Co-product naphtha and diesel will be processed onsite only to the extent needed for sale as chemical or petroleum refinery feedstocks. Coproduct LPG will be used onsite or minimally processed for sale.

A combined cycle power island using a commercial gas turbine (qualified for syngas firing) will supply all on-site electricity needs and export a non-negligible amount of electricity. The plant will thereby serve to demonstrate a new technological strategy for coal-based grid electricity supply (co-production). Previous analysis of prospective commercial co-production of fuels and power suggests that the profitability of such plants in a carbon-constrained world has the potential to exceed that of the most profitable conventional stand-alone power generation options, including natural gas combined cycles.[13]

Plant Process and Equipment Selections. Given the stipulated capital investment cap ($2B) and expected plant operating life (5 to 10 years), the priority in process design trade-off decisions and in vendor/equipment selections will be to maximize the likelihood of technical success rather than minimize cost of jet fuel production. Technical success will be defined by demonstrating consistent plant operability and product quality, and by generation of knowledge and data that helps identify potential performance and cost improvements in future plant deployments at commercial scale. Given these objectives,

- The use of novel or yet-to-be commercially deployed processes or equipment will be minimized.
- The use of one-off, specially designed plant items will be minimized. Preference will be given to off-the-shelf equipment packages or packages that could plausibly become off-the-shelf in the first commercial deployments following the FOAK demonstration project.

- The use of field fabrication and field erection of equipment will be minimized. Preference will be given to shop-assembled modules or modules with the potential for shop-assembly when adopted in the first commercial deployments following the FOAK demonstration project.
- The plant will be designed for steady-state, fixed capacity operation and for fixed feedstock and product mixes. The ability to vary product mix or biomass/coal input ratio may be beneficial for the economics of future LBJ plants in some settings, but demonstrating these capabilities in the FOAK plant are considered of secondary importance.
- The plant will be designed to accommodate intermittent periods of off-design operation. Future commercial deployments of LBJ plants are envisioned for base-load operation.
- The plant will be designed to facilitate equipment modifications or replacements, e.g., by physical layout that simplifies access.
- Availability of the plant is expected to increase with time over the operating life as issues causing loss of production are identified and corrected. The experience of first-generation coal IGCC plants (see Figure 1) will be used as a guide in evaluating the prospective economics of the FOAK demonstration project.

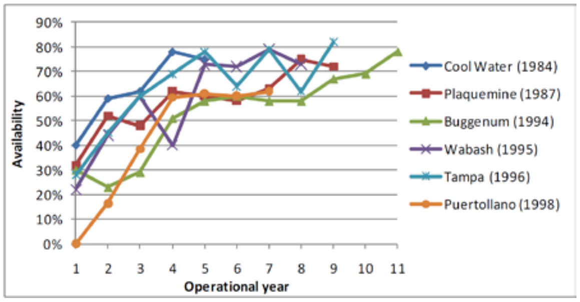


Figure 1. Availabilities of some early coal IGCC plants.[14]

Design / Operating Lifetime. The plant will be designed for a 20-year life. Based on historical precedents, we anticipate the operating life will be 5 to 10 years, which we expect will be sufficient to demonstrate technical viability and to provide critical information for design of future commercial plants.[b] The plant may continue to operate beyond the anticipated operating life if, and as long as, revenues exceed production costs.

Other. The project will be designed to minimize all non-plant capital investment requirements to the extent that this will not compromise the chances for the technical success of the project. For example, for CO_2 transport the option of a short CO_2 pipeline connecting to the pipeline running from the Kemper County IGCC project to the CO_2 pipeline network near Jackson Dome, MS, may be explored, as well as the possibility of Denbury or similar company taking responsibility for CO_2 off-take at the plant gate.

---

[b] For example, the Cool Water project,[9] which successfully demonstrated the technical feasibility of integrated gasification combined-cycle (IGCC) technology with bituminous coal at a scale of 120 MWe (nominal), was operated for approximately 5 years, as a result of which it proved the technical feasibility of IGCC and paved the way for subsequent commercial-scale IGCC projects using bituminous coal (e.g., Tampa Electric, Wabash River, and other projects around the world).

3 First-of-a-Kind (FOAK) Plant Process Design

Process design basis information is provided in the design report prepared by the WorleyParsons Group (WP) as part of this project – see Appendix A. Here we discuss design basis issues associated with the project site, and we describe the design feedstocks.

3.1 Project Site

The prospective site for the LBJ project is in Lauderdale County, Mississippi, the current site of Plant Sweatt, owned by Mississippi Power Company (MPC), a 2 x 40 MW gas-fired facility originally installed in 1951 (when it was oil-fired). MPC's current planning envisions that Plant Sweatt will be retired before 2020, leaving a brownfield site that would facilitate permitting and public acceptance of an LBJ demonstration project. Plant Sweatt sits on a 650 acre plot of largely forested land owned by MPC a few miles south of the city of Meridian and 27 miles south of the Kemper County IGCC project (Figure 2, left). The Plant Sweatt location overlies the same Wilcox Group lignite formation that will supply the Kemper IGCC plant (Figure 2, right).

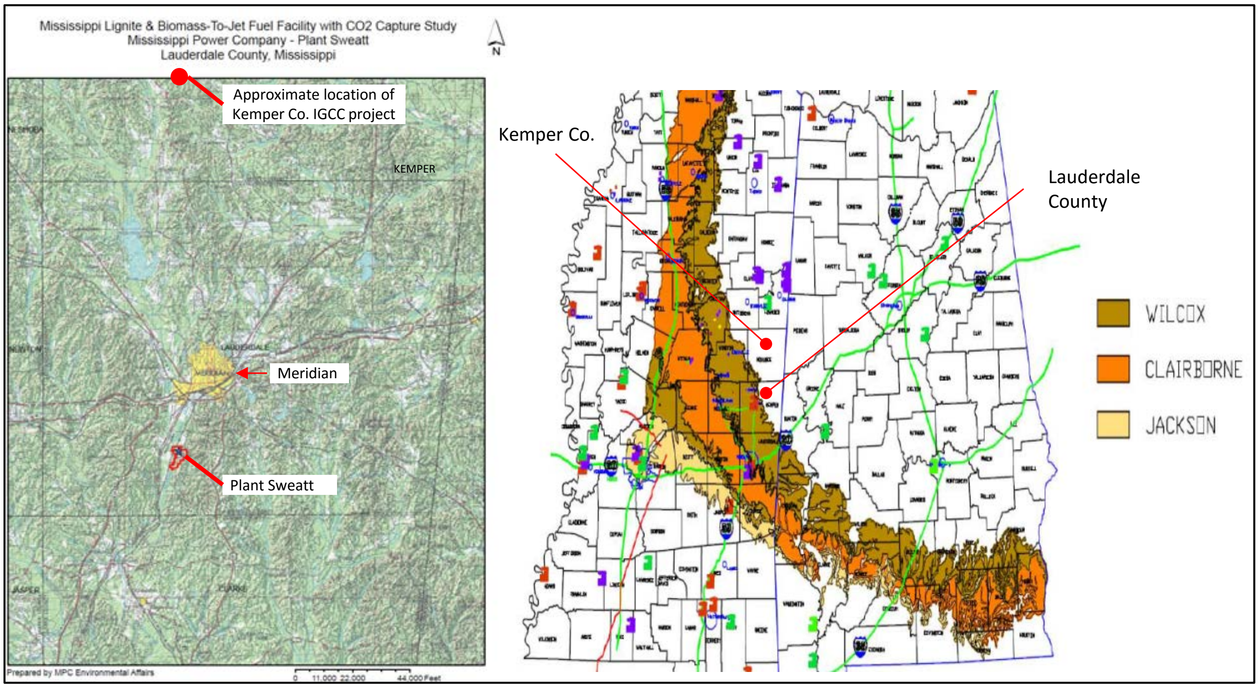


Figure 2. The project host site (Plant Sweatt) is a few miles south of Meridian, MS (left) and overlays the Wilcox Group lignite seam (right) that is also the source of lignite for the Kemper County IGCC project situated 27 miles north of Plant Sweatt.

An understanding of current infrastructure and utilities available on and around the Plant Sweatt site helped inform the cost estimate developed for the LBJ plant. The water flow of the nearby Okatibbee Creek defining the western edge of MPC's property is insufficient for the plant to be granted permits to use this as the water source for the plant. However, there are five existing ground water wells that can supply approximately 500 gpm each. These will provide makeup water needed by the plant. Early in the project the option of contracting for grey water from the Meridian water treatment plant was considered by the project team before it learned that the Kemper County IGCC plant has already contracted for all of Meridian's grey water. Plant

Sweatt currently has a National Pollutant Discharge Elimination System (NPDES) permit for discharge of wastewater into Okatibbee Creek, but for a new facility it is likely that a zero liquid discharge (ZLD) system would be required, as is the case at the nearby Kemper County IGCC. Accordingly, the LBJ plant has been designed with a ZLD system.

Two natural gas pipelines feed the Plant Sweatt site, which provides the LBJ plant with the ability to use gas for plant startup and other intermittent uses. Plant Sweatt currently feeds power to the grid at 115/230 kV. Only minor upgrading of the connection to the main transmission line is assumed to be required to export power from the LBJ facility. A CO_2 pipeline will be required to deliver CO_2 from the LBJ facility to the CO_2 pipeline network in the Jackson Dome area in southwest Mississippi. The CO_2 pipeline from the Kemper County IGCC plant to the Jackson Dome area passes within 3.5 miles of Plant Sweatt. The project team was unable to confirm whether this pipeline has any excess capacity for accepting CO_2 from the LBJ plant. In any case, the assumption is made (in the financial analysis) that the CO_2 off-taker will be responsible for its transport from the LBJ plant gate to the EOR user.

Plant Sweatt currently generates no ash, but an ash management plan would be developed for the LBJ facility. The current plan is modeled on the plan at the Kemper County IGCC. The Kemper site includes a 60-acre ash management unit (AMU) sufficient to store 5 years' worth of Kemper's ash production, during which time efforts will be made to identify beneficial uses of the ash, e.g., as soil amendment, road construction material, or cement ingredient. If the state rejects a Beneficial Use Determination (BUD) permit application from Kemper, the two fallback options are to *(a)* prove to the state that the ash by itself is not hazardous and use it as additional backfill at the adjacent lignite mine or *(b)* obtain permits to increase the size of the AMU to enable it to accommodate ash for the life of the plant. If other options are not available, the land area at Plant Sweatt should be sufficient to house a large enough AMU for the FOAK LBJ project.

3.2 Feedstock Supply

The design feedstocks for the FOAK plant will be Mississippi lignite and woody biomass with the properties shown in Table 1. A key attraction of the Plant Sweatt site is its proximity to both lignite and biomass supplies.

North American Coal (NAC), the company contracted to supply the lignite to the Kemper County IGCC, suggested two options for lignite supply to the LBJ facility at the Plant Sweatt site:[15] build a new lignite mine adjacent to the site or truck lignite from the existing mine adjacent to the Kemper Co. IGCC plant .[c] Costs would likely be similar in either case because trucking from the Kemper mine would add approximately as much cost as the added cost for mining near the LBJ site, which would incur higher mining costs than at Kemper due to the larger overburden on the lignite seam in that area. Since the lignite would be originating from the same region of the Wilcox Group lignite formation, its quality would be similar with either option. The estimated cost for lignite delivered to the LBJ site from the Kemper mine is about $\$3/GJ_{HHV}$ (\$2.8/MBTU), about 20% of which is transport cost.[15]

---

[c] The latter would require an agreement with Mississippi Power Co., who owns the lignite in the Liberty Mine.

Table 1. Design feedstocks.

	Lignite[21]			Southern Pine[3]		
Weight %	***Dry***	***To Gasif.***	***As Rec'd.***	***Dry***	***To Gasif.***	***As Rec'd.***
Carbon	57.84	45.13	31.53	53.89	45.81	30.56
Hydrogen	3.63	2.83	1.98	5.33	4.53	3.02
Nitrogen	0.88	0.69	0.48	0.41	0.34	0.23
Chlorine	0.02	0.02	0.01	0.00	0.00	0.00
Sulfur	1.82	1.42	0.99	0.04	0.03	0.02
Oxygen	13.89	10.83	7.57	39.25	33.36	22.25
Ash	21.92	17.10	11.95	1.09	0.93	0.62
Moisture		22.00	45.50		15.00	43.30
HHV, MJ/kg	22.58	17.61	12.30	20.19	17.17	11.45
LHV, MJ/kg	21.79	16.46	10.76	19.03	15.81	9.73

The Plant Sweatt region is heavily forested (74%),[16] and biomass supplied from sustainably managed pine plantations will be the LBJ feedstock. Southern Company considered the possibility of repowering Plant Sweatt with biomass several years ago, and as a result detailed biomass supply and cost studies were undertaken.[17,18] These studies found that as much as 580,000 tonnes (dry basis) per year of delivered green wood chips would be available for \$51 per dry tonne (\$2.7/$\mathrm{GJ}_{\mathrm{HHV}}$) or less.

In support of the LBJ study, Southern Company Services independently funded additional work by the Antares Group to update the biomass supply and cost analysis in the Plant Sweatt region.[16,19,20] Antares estimated there are currently 4.95 million tonnes/yr of pulpwood quality southern pine generated in excess of pulpwood demands within a 100-mile radius of the Plant Sweatt site. Additionally, 0.82 million tonnes/yr of harvest residues are generated in the region, but do not have a market. An additional 2.3 million tonnes/year of residues are generated from land clearing/site preparation and wood processing activities in the region, but these are fully utilized for energy or other purposes. As detailed in Section 3.4, the LBJ plant will require a maximum of about 0.1 million tonnes/yr of biomass feedstocks. Antares projects that this relatively small supply required by the LBJ plant would not materially impact prevailing market prices for biomass, which Antares estimated as shown in Table 2.

Forest residues would be the preferred biomass feedstock because of their lower cost, but engineers at Southern Company's Power Systems Development Facility (PSDF) involved with development of biomass feeding systems for the TRIG gasifier, the technology selected for the LBJ plant, indicated that the gasifier input biomass should have a maximum particle dimension of ⅛ inch (3,000 µm) and moisture content no higher than 15% to ensure flowability. Biomass processing experts at the Antares Group indicated that to assure a consistent feedstock supply that meets these specifications would likely require starting with debarked pulpwood-quality logs. The option of using residues would not be able to guarantee achievability of these conditions since slash is characterized by a

Table 2. Estimated range in delivered prices for biomass in the Meridian, MS, region in 2014.[16]

	\$ per dry tonne			\$ per $\mathrm{GJ}_{\mathrm{HHV}}$		
	Base	High	Low	Base	High	Low
Forest residues	42	65	32	2.1	3.2	1.6
Mill chips	64	84	57	3.3	4.2	2.9
Log length pulpwood	65	81	44	3.1	4.0	2.2
Chipped pulpwood	73	91	51	3.6	4.5	2.5
Pellets (premium)	209	223	178	10.5	11.2	8.9

high bark-to-wood ratio and bark is difficult to grind to uniform ⅛" size. With additional development of feeding systems, residues may become a viable biomass option, but to minimize technological risks with the FOAK LBJ plant, pulpwood-quality southern pine logs will be the biomass feedstock.

Log-length pulpwood was recommended by Antares as the form of biomass delivered to the LBJ site to be able to meet the gasifier feed specifications and to facilitate on-site storage. Antares estimated that the cost for logs delivered to the LBJ facility would be about \$59 per tonne, or \$2.9/GJ$_{\mathrm{HHV}}$. The sustainability and net greenhouse gas emissions of this biomass supply are discussed in Section 4.2.

3.3 Key Technology Selections

3.3.1 Plant Overview

Figure 3 is an overview of the FOAK LBJ plant configuration. The plant receives lignite and pulpwood-grade logs that are separately subjected to sizing and drying before being fed to a pressurized oxygen-blown TRIG gasifier using 99.5% purity oxidant from a dedicated air separation unit (ASU). The resulting syngas is cooled, filtered, and scrubbed before a partial sour water gas shift (WGS) to set the H_2/CO ratio at the Fischer-Tropsch (F-T) synthesis inlet to the desired value to optimize CO conversion. CO_2, H_2S, and trace impurities are then removed using a methanol solvent (Rectisol®) at the acid gas removal (AGR) island. The CO_2 is compressed and delivered via pipe to the plant fence line.[d] The H_2S is converted to wet sulfuric acid (WSA) for sale. A guard bed downstream of the AGR protects the cobalt-based F-T synthesis catalyst.

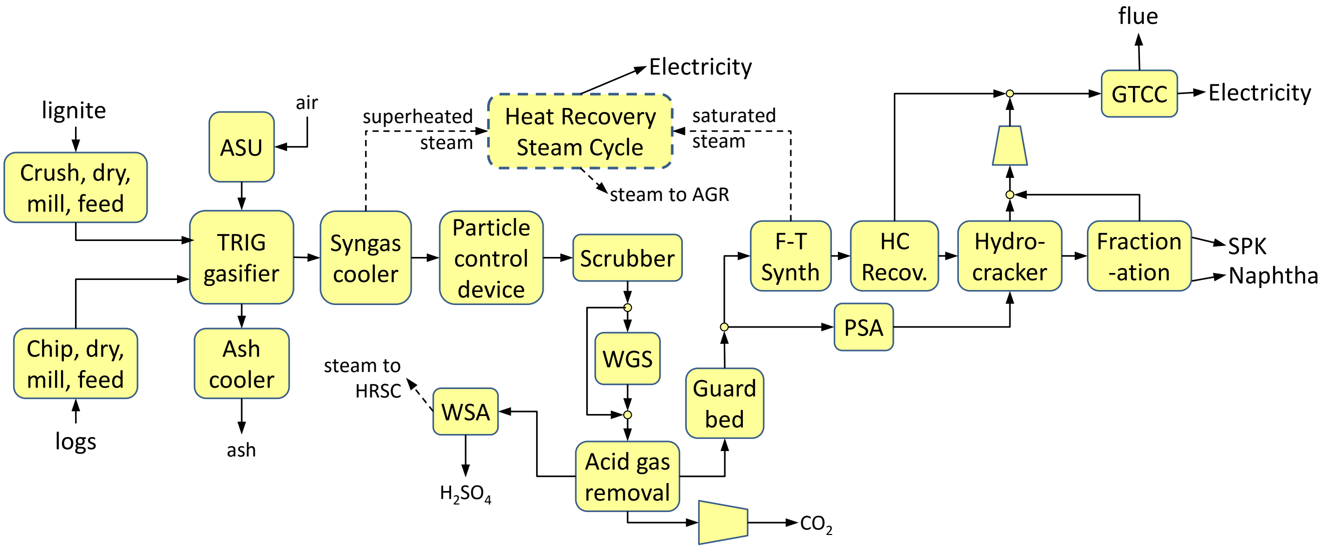


Figure 3. FOAK LBJ plant simplified process configuration.

---

[d] For purposes of this study, it is assumed that the CO_2 will be sold into the pipeline transporting CO_2 from the Kemper Co. IGCC facility, which passes within approximately 3 miles of the LBJ site. That pipeline delivers CO_2 to an existing CO_2 pipeline network in the Jackson Dome, MS, area which in turn delivers CO_2 to enhanced oil recovery operations on the Gulf Coast.

The F-T synthesis island includes a tubular fixed-bed reactor with a cobalt-based catalyst, followed by hydrocarbon recovery and processing to produce the primary liquid product (synthetic paraffinic kerosene, SPK) plus a synthetic naphtha coproduct.

Following F-T synthesis, hydrocarbon recovery separates liquids from permanent gases. The latter are collected for use as fuel in a gas turbine combined cycle (GTCC). Crude F-T liquids are subjected to hydrocracking and fractionation, resulting in synthetic paraffinic kerosene (SPK) and naphtha that are transferred to storage tanks to await off-loading from the plant by truck.[e] Hydrogen for the hydrocracker is supplied from a slipstream of post-AGR syngas via pressure swing adsorption (PSA). The PSA raffinate and the light ends from the hydrocracking and fractionation are collected and used as additional GTCC fuel. Electricity from the GTCC is supplemented with electricity from a separate heat recovery steam cycle (HRSC) that utilizes process heat primarily from syngas cooling and F-T synthesis. Process heat is also used for feedstock drying and some other needs.

The F-T island is designed without recycle of the light (gaseous) fraction leaving the synthesis reactor, which includes unconverted syngas, methane, and other light hydrocarbons formed during synthesis. This eliminates the cost for the reformer that would be needed in a recycle loop and also provides the bulk of the fuel for the onsite GTCC that generates sufficient power to meet auxiliary loads and to export some power to the grid. The fuel gas fraction of synthesis products is fixed by the design of the FT synthesis reactor and so the choice of gas turbine in the power island determines the overall scale of the plant.

Key design features and decisions are discussed in the following sections, followed by a summary of the plant's simulated energy and carbon performance.

3.3.2 Feedstock preparation

Separate equipment trains prepare the lignite and the woody biomass for the gasifier.

Pre-crushed, 2-inch minus lignite is delivered by truck to an unloading station at the plant site consisting of a truck weighing system and three 40-ton hoppers. A belt conveyer transfers the lignite from the hopper into Transfer Tower #1, where a magnetic separator will be used to separate ferrous metal objects from the lignite, after which it is conveyed to 2x100% crushers located in Transfer Tower #2. The crushers produce ½-inch lignite that is conveyed to one of the two 2,300 tonne storage silos. Figure 4 shows a lignite handling system at an existing Louisiana site. The LBJ design will be similar to this.

From storage the lignite is fed to a fluid-bed dryer of the same design used at the Kemper County IGCC project, which dries the lignite from 46 wt% moisture to 22 wt% for feeding to the gasifier (Table 1). The drying gas is a mixture of nitrogen and air containing approximately 10 mol% or less oxygen. Nitrogen from the ASU is added as a diluent as needed. The drying air and dryer bed are heated with coils containing hot (tempered) water.

---

[e] The SPK will be transported to the nearby Meridian delivery station of Kinder-Morgan's Plantation Pipeline, which delivers fuels to airports in Washington DC and many points in between. The SPK is suitable for splash blending in a 50/50 mix with petroleum-derived jet fuel. The naphtha will be sold as a petrochemical feedstock to a nearby refinery (e.g., the 75k bbl/d Hunt refinery in Tuscaloosa, AL, or the 330k bbl/d Chevron refinery in Pascagoula, MS).

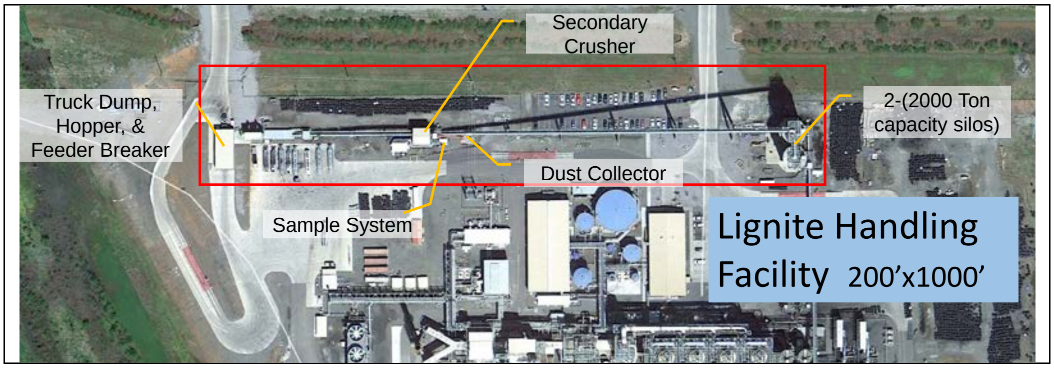


Figure 4. Lignite handling system at the ADA Carbon Solutions facility in Coushatta, LA,[15] with similar plot size as would be the case with the FOAK LBJ plant.

The moist dryer exhaust passes through a multi-cyclone dryer exhaust filter device to remove entrained fine particles before entering the dryer exhaust condenser. The condenser contains a venturi nozzle which removes any remaining entrained fine particulates before the gas is cooled in a direct contact cooler. Condensed water is recovered and utilized as makeup water in the plant. The cooled gas is recycled back to the lignite dryer where it is reheated and reused for drying additional lignite. The dried lignite is combined with fines from the dryer exhaust filter in a nitrogen-swept mill and sized to 500-µm minus and delivered to the gasifier lock hopper surge bin.

The biomass preparation system was designed to deliver a maximum particle dimension for biomass feed to the TRIG gasifier of 3,000 µm with moisture content no greater than 15 wt%, as discussed in Section 3.2.

Biomass will be delivered to the plant as pulpwood-quality southern yellow pine logs. The logs arrive by truck and are stacked in an open area with a storage capacity of approximately 17,000 tonnes, or 30 days of full-load feedstock supply. The onsite debarking and chipping system is sized to produce ½-inch minus wood chips at a rate of 77 t/h, or about three times the feedstock input rate to the gasifier, such that the chipper would operate about 8 hours/day, thereby minimizing after-hours noise issues. The chipping line consists of an infeed conveyor, a drum debarker, a discharge conveyer and a drum chipper. Chipped biomass is conveyed to an open chip pile storage area. Hog from the debarker is screened to 2" minus and transferred to the bark pile by a radial bark stacker. A stoker is used to reclaim the material from the radial stacker chip storage pile. The bark is hauled offsite by trucks and sold for landscaping and other uses.[f] Figure 5 illustrates the anticipated layout of the biomass processing up through this processing step.

The chipped biomass is fed to a low-temperature belt dryer, where its moisture content is reduced from 43 wt% to 15 wt% using air pre-heated by tempered water. The moist dryer exhaust passes through a multi-cyclone dryer exhaust filter device to remove any entrained fine particles before entering the dryer exhaust condenser. The condenser contains a venturi nozzle which removes any remaining entrained fine particulate before the gas is cooled in a direct contact cooler. Condensed water is recovered and utilized as makeup water in the plant. The

---

[f] Revenue from bark sales are neglected in the financial analysis later in the report.

cooled gas is recycled back to the biomass dryer where it is reheated and reused for drying. The dry biomass is conveyed to a hammer mill where it is sized to 3,000-µm minus. The milled biomass is conveyed to the gasifier lock hopper surge bin.

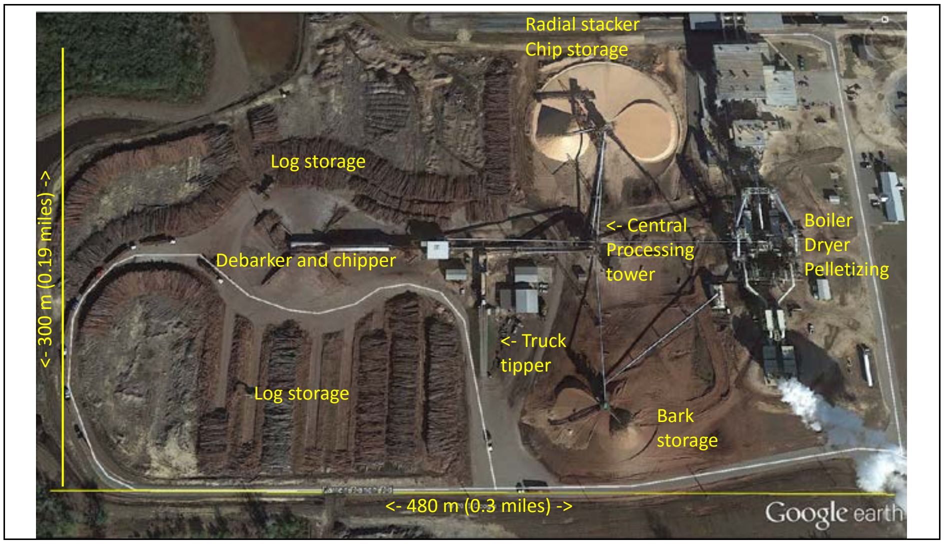


Figure 5. Example of log storage and preparation system similar to that for the LBJ plant (Green Circle Bioenergy in Florida), except that the LBJ facility will not include the "Boiler/Dryer/Pelletizing") and will process about one-fifth as much biomass as this facility.

3.3.3 Gasifier island

The gasifier technology that was pre-selected for the LBJ plant is the TRIG technology developed jointly by KBR and Southern Company. A key reason for this selection is that the capability of the technology to co-gasify biomass and coal feedstocks has been demonstrated at pilot scale.[10,11] Additionally, the first commercial TRIG gasifiers operating on lignite alone have recently been commissioned at the Kemper County IGCC project in Mississippi.[23] There are no commercial-scale TRIG gasifiers co-processing biomass and lignite. However, at Southern Company's Power Systems Development Facility (PSDF), up to 30 wt% biomass has been co-gasified successfully with Powder River Basin subbituminous coal in both air-blown and oxygen-blown configurations. (The TRIG gasifier for the LBJ plant will be oxygen blown to minimize inert components in the syngas.) The PSDF testing found no significant biomass preparation or feeding issues; negligible effect on carbon conversion and syngas heating value (Figure 6), no evidence of agglomeration or deposition

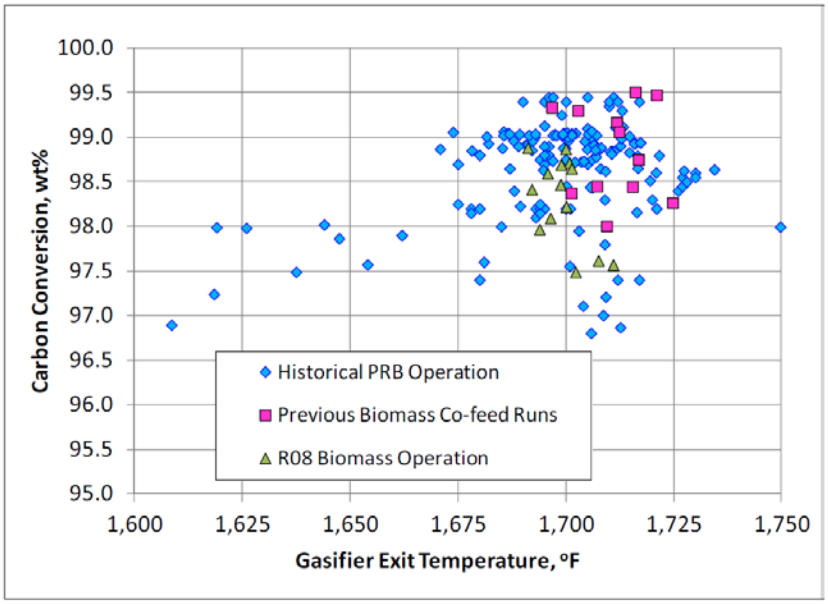


Figure 6. Carbon conversion results for co-gasification of coal and biomass at the PSDF TRIG test facility. R08 refers to test run R08 at PSDF.[22]

of solids in the gasifier, no indication of excessive tar generation, and stable performance of the particulate control device. Based on the levels of biomass co-processed at the PSDF, and consistent with the LBJ project's guiding principle that there should be a non-negligible biomass input fraction to the plant (see Section 2), the decision was made that the biomass co-feed fraction for the LBJ project would be 25% on a higher heating value basis.

A schematic of a TRIG gasifier is shown in Figure 7. Feedstock is added via a lock-hopper system to the top of the mixing zone in the gasifier's riser section. The fuel is rapidly heated by hot recirculating ash from the gasifier's standpipe section. The standpipe connects to the riser via a non-mechanical, J-leg valve that allows ash to flow into the mixing zone against the pressure gradient. Oxygen is fed into the lower part of the mixing zone to drive combustion of carbon returning with the ash. As the hot mixture rises, it provides heat for the endothermic gasification reactions in the riser. The large inventory of ash and its intimate contact with the reacting feedstock promote high carbon conversion at gasification temperatures well below those for oxygen-blown entrained-flow coal gasifiers. The low temperature also leads to a larger methane concentration in the syngas compared with entrained flow gasifiers.

At the top of the riser, the solids are separated from the syngas in a primary cyclone. The larger particles fall into the gasifier seal leg, which acts as a second non-mechanical valve to convey solids into the standpipe. Overhead syngas from the primary cyclone enters a secondary cyclone, which removes finer particulate matter that falls into the gasifier standpipe. The solids level in the standpipe is controlled by the continuous removal of ash from the seal leg and/or standpipe. Syngas from the gasifier's recycle compressor is injected into the non-mechanical valves in order to improve flow.

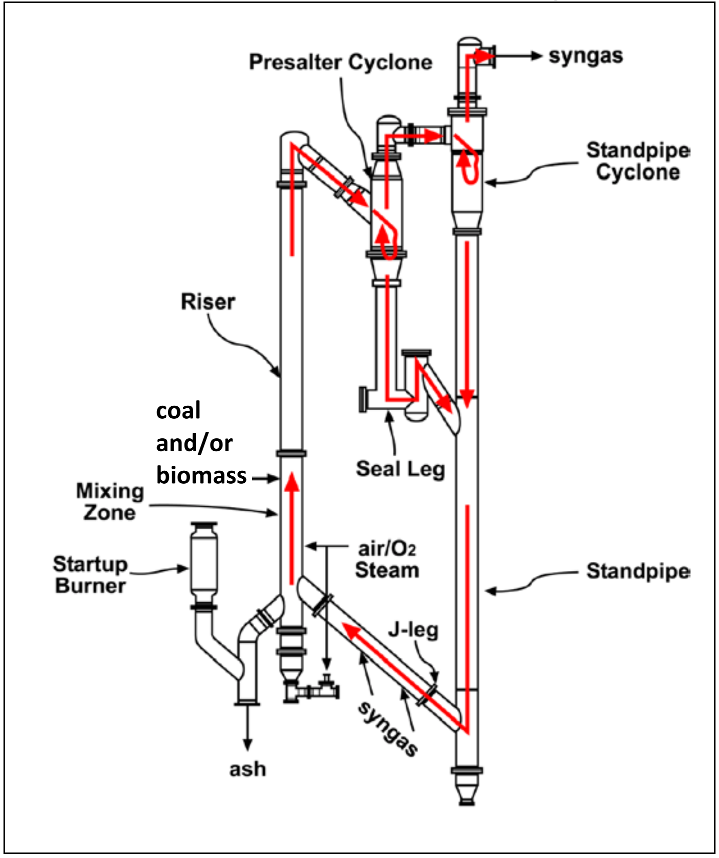


Figure 7. TRIG gasifier simplified schematic.[24]

Following the secondary cyclone, the raw product gas travels through a syngas cooler that reduces its temperature to about 300°C by evaporating water or superheating steam. This is followed by a particulate control device that removes the remaining fine ash. Finally, a wet scrubber serves to provide the clean-up of syngas within the gasifier island battery limits and also to load the syngas with the water content needed for the downstream water-gas shift reaction. See Figure 8.

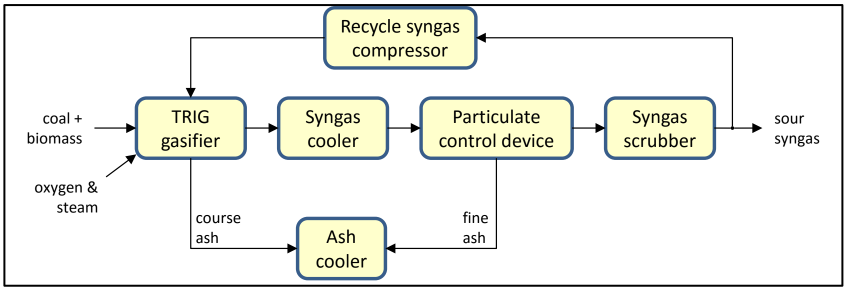


Figure 8. Gasifier island block flow diagram.

KBR provided the project team with a preliminary heat and mass balance for an oxygen-blown TRIG gasifier operating with 100% lignite, but KBR lacks experience with the design of TRIG gasifiers co-processing biomass, so engineers with this experience at the PSDF were consulted to provide additional critical inputs for completing the design and cost estimate for the gasifier island. Table 3 shows the estimated syngas composition following cooling, particulate filtration, and wet scrubbing steps. The relatively high methane fraction (compared with syngas from an entrained flow gasifier) is due to the relatively low operating temperature of the gasifier, and reduces the achievable yield of Fisher-Tropsch liquids because it is unreactive in Fisher-Tropsch synthesis.

3.3.4 Acid Gas Removal and Wet Sulfuric Acid Unit

Acid gas removal (AGR) is needed to remove all sulfur and reactive nitrogen species from the syngas to protect the downstream F-T synthesis catalysts and to remove CO_2 for compression and delivery to a pipeline. Two commercial technology options were considered for acid gas removal, Selexol and Rectisol. Considering that one of the guiding principles for plant design is to prioritize technology selections and design decisions that maximize the likelihood of demonstrating technical success with the first-of-a-kind plant, the decision was made to adopt Rectisol (the higher capital cost option) because of its greater ability to remove sulfur and other contaminants to the extremely low levels required to avoid poisoning of the expensive cobalt-based catalyst used in the FT reactor.

Table 3. Syngas composition leaving gasifier island.

	Volume %
CO	25.79
H_2	19.44
CH_4	3.46
CO_2	19.38
H_2O	31.21
H_2S	0.38
COS	0.02
N_2	0.27
Ar	0.05

Air Liquide was contracted to provide an engineering design and cost estimate for a Rectisol system that was the basis for the design of the acid gas removal island in this project – see Appendix B. Unlike Rectisol system designs that would be used in conjunction with entrained flow coal gasifiers, Air Liquide's design includes a prewash step to ensure that any trace level of tars (heavy hydrocarbons) in the syngas are removed prior to removal of H_2S and CO_2. This pre-wash step is not necessary when treating syngas from entrained-flow gasifiers because the high-temperature at which these operate ensures complete destruction of tars. A consequence of including the pre-wash step is a high parasitic steam demand for recovery of the pre-wash solvent.

The main function of the wet sulfuric acid (WSA) process is to produce a concentrated sulfuric acid product stream by condensation of gases generated from the incineration of sour waste gas streams. The WSA process was selected (rather than Claus/SCOT process to produce elemental sulfur) because there is a ready local market for sulfuric acid and because the Kemper County IGCC plant also uses this process. Haldor Topsoe provided engineering inputs to help design and estimate the cost of the WSA system. The WSA process includes oxidation of sulfur-containing gases to SO_2, followed by catalytic conversion of SO_2 to SO_3, vapor-phase hydration of SO_3 to H_2SO_4, and finally condensation to liquid H_2SO_4. The WSA handles several acid gas streams, the largest of which is acid gas from the AGR hot regenerator. Other streams include pre-wash acid gas from the AGR, mixed sour gas from the sour-water stripper, vent gas from low-temperature cooling condensate collection, syngas from the gasifier ash handling system, and other small vents. Emissions regulations prevent emitting any of these streams to the atmosphere without treatment.

3.3.5 Fischer-Tropsch Synthesis and Upgrading

Several design alternatives were considered for the Fischer-Tropsch synthesis island, including tubular fixed-bed, slurry-bed, and compact designs that utilize fixed catalyst beds. Traditional tubular fixed-bed reactors (such as the Shell reactors at the Pearl facility in Qatar) are very well understood, scalable technology and hence relatively low risk for LBJ. Slurry-bed systems (such as Sasol's technology used at the Oryx project in Qatar) are generally considered better suited to larger scales than envisioned for the LBJ plant. Several companies are pursing commercialization of compact reactor designs with the goal of shop-fabricating modules to minimize high field erection costs.

Traditional FT catalysts are either cobalt- or iron-based. Cobalt provides higher activity, better selectivity, and longer lifetimes, but it is more sensitive to poisons in the syngas and much more expensive. The advantages of cobalt are considered to outweigh their disadvantages in the dominant market for FT technology today – gas to liquids – and most FT technology providers today offer only cobalt-based systems.

Nine potential FT technology providers were approached to explore their interest in participating in the LBJ project. Considering the small-scale of the LBJ project and the guiding principle to maximize modular plant components (for ease of scale up and, potentially, reduced capital costs), Princeton's evaluation of FT technologies focused on modular options. Of the nine companies contacted, Compact GTL, Greyrock, SCG Energia, Shell, and Sasol declined to consider participating. Initial conversations were pursued with CCTSA (a South African company), Emerging Fuels Technology (EFT, an American company), and Velocys (also a U.S. company). CCTSA's technology seemed to be still very much in a development stage, and considering the guiding principle for the LBJ project to adopt proven technology wherever possible, CCTSA was dropped from further consideration. Negotiations were pursued with Velocys, which offers a unique micro-channel fixed-bed modular FT reactor design (Figure 9) and with EFT, which offers a more traditional tubular fixed-bed design fabricated in modules (Figure 10).

Consistent with the project guiding principle to select established technologies where possible, the EFT technology was selected because tubular fixed-bed reactors are proven commercially. EFT has over 100-person years of experience in Fischer-Tropsch and related fields, including prior management experience at Exxon, Sasol, ConocoPhillips, Chevron, Syntroleum, and Rentech. They offer traditional tubular fixed-bed reactor technology with a proprietary low-temperature cobalt catalyst. The target market for their technology is small-scale gas-to-liquids conversion. As such their reactor design is modular to facilitate shop fabrication and minimize field erection costs. For the LBJ project, the system designed by EFT has a nominal reactor module size of 500 barrels/day. At this capacity, EFT estimates that more than 100 shops across the U.S. have the capability to build these modules. The large number of shops should contribute to competitive pricing. EFT also has expertise in the design of the back-end FT syncrude upgrading area and included this in the engineering design and cost estimation work they completed for the LBJ project.

Figure 9. Example of a modular FT reactor. This is a Velocys microchannel reactor module with a production capacity of 175 barrels per day.

Appendix C details the design and cost information provided by EFT for the project. The FT island comprises three parallel synthesis reactors, followed by a hydrocarbon recovery unit, a hydrocracker, and a fractionator. Unconverted syngas and light ends collected from the upgrading steps could be recycled via an autothermal reformer (ATR) to increase FT liquids output. This recycle approach was not used in the LBJ design. Instead, the unconverted syngas and light ends were collected for use as fuel in the power island, thereby avoiding the cost of an ATR in the FT island and enabling some export of electricity from the plant.

3.3.6 Power Island

The project guiding principles call for the power island to include a gas turbine combined cycle for the purpose of demonstrating the technical feasibility of a design that would be implemented in future commercial-scale plants. Gas turbines made by Solar, Siemens, and General Electric were considered. A key feature of the gas turbine selected is its ability to use as fuel unconverted syngas combined with light hydrocarbons produced in the FT reactor and upgrading areas. In the smaller size range suitable for the LBJ project, several turbines that appear to have this capability were investigated. These generally use diffusion flame combustors, for which NO_x emissions typically exceed regulated levels. As a result, it is anticipated that selective catalytic reduction will be required to meet emission limits. In consultation with WorleyParsons engineers, the Siemens SGT-700 gas turbine was selected. This machine has a nominal simple-cycle output on natural gas of 33 MW_e.

Figure 10. EFT's Fischer-Tropsch synthesis module concept.

In addition to fuel for power generation, the LBJ process also generates considerable process heat, primarily from syngas cooling after gasification and from cooling of the exothermic FT reactions. Two approaches to using the process heat to augment power generated by the gas turbine combined cycle were analyzed. In design A, the recovery of process heat was integrated with the gas turbine exhaust heat recovery steam generator. In design B, the process heat was used in a heat recovery steam cycle (HRSC) decoupled from the gas turbine combined cycle. Design A would require only a single steam turbine in the power island, while design B would require two smaller steam turbines, for which efficiencies would be lower and unit costs would be higher than for the single large turbine in design A. On the other hand, benefits of decoupling the GTCC from the process heat recovery steam cycle, as in design B, are the ability to use an "off-the-shelf" GTCC package, reduced complexity in the design and operation of the heat recovery steam generator, and the ability to fire the GTCC with natural gas during plant start up or when the syngas production system is down for maintenance.

To determine how much of a thermodynamic benefit might result with Design A, experts at the Politecnico di Milano under the guidance of Prof. Emanuele Martelli were contracted to develop detailed process designs for each approach. Using process heat availability predicted by Princeton's Aspen Plus process simulations as inputs, Politecnico's analysis utilized their in-house process optimization methods to design heat exchange networks and estimate performance of designs A and B. As expected, Politecnico found that design A would provide greater total steam turbine power output, but only about 10% more than design B – see Appendix D. The other benefits of design B, especially the ability to operate the GTCC on natural gas when needed, were deemed to outweigh the modest efficiency gain provided with design A, and design B was adopted for the final plant design, as shown in Figure 3.

3.4 Plant Energy Performance Summary

Figure 11, Figure 12, and Figure 13 constitute a summary level block flow sheet for the LBJ plant, and the accompanying stream tables (included after Figure 13) give the results of the steady-state heat and mass balance simulations developed for the LBJ plant. These balances were developed collaboratively by WorleyParsons engineers and Princeton University researchers using Aspen Plus software. Detailed process flow diagrams developed by WorleyParsons for purposes of estimating bare erected plant capital costs are available in the report prepared by WorleyParsons under contract to this project (Appendix A).

As designed, the plant processes 1551 t/d of lignite and 556 t/d of biomass (both as-received basis) to produce 1250 actual barrels per day of product, of which 80% is SPK and 20% is naphtha (Table 4). The plant also exports 15 MW of electricity after satisfying a 39 MW onsite auxiliary load. Two-thirds of the latter is accounted for by air separation (15 MW), CO_2 compression (7 MW), and acid gas removal (5 MW) (Table 5). Additional products are 49 t/day of sulfuric acid and 1326 t/d of pipeline-quality CO_2. The plant converts 19% of the input carbon into fuels, and 55% is captured for sale (Table 6). The corresponding energy conversion efficiency for the plant is 35.6% (LHV basis), a modest value that results primarily because more than 70% of the gross power generated onsite must be used to meet auxiliary power demands.

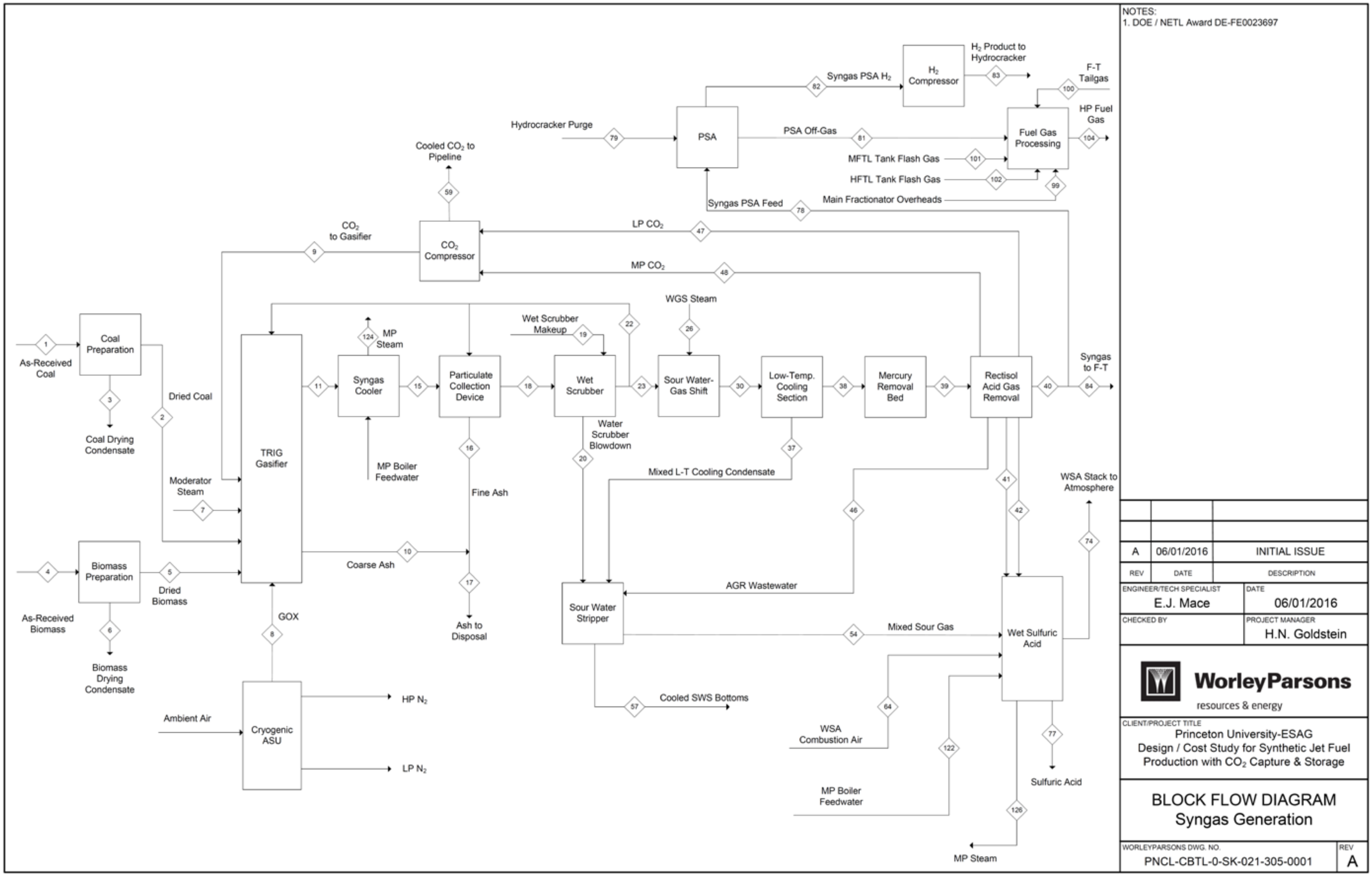


Figure 11. LBJ plant block flow diagram, syngas generation area. See stream tables following Figure 13.

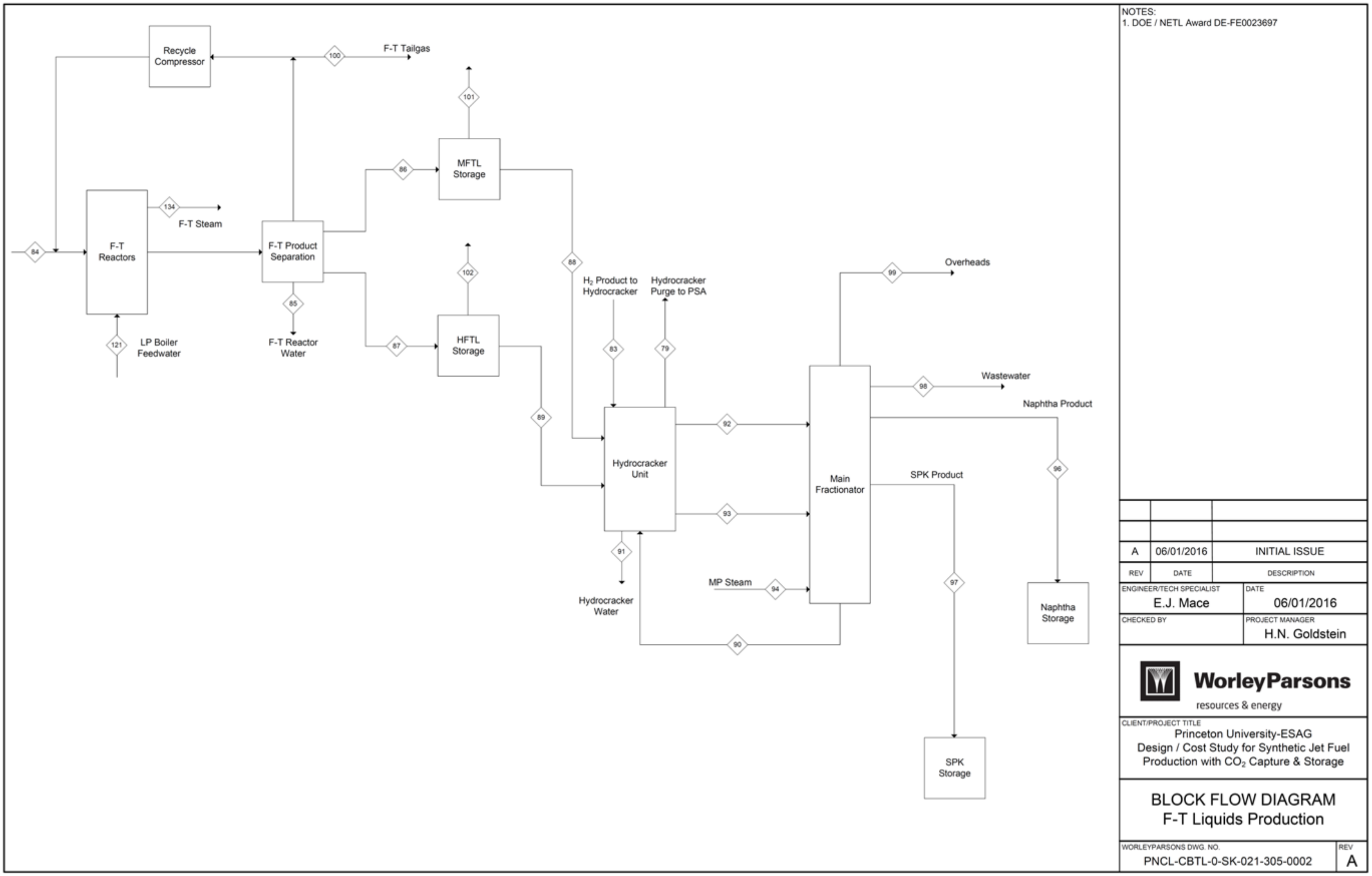
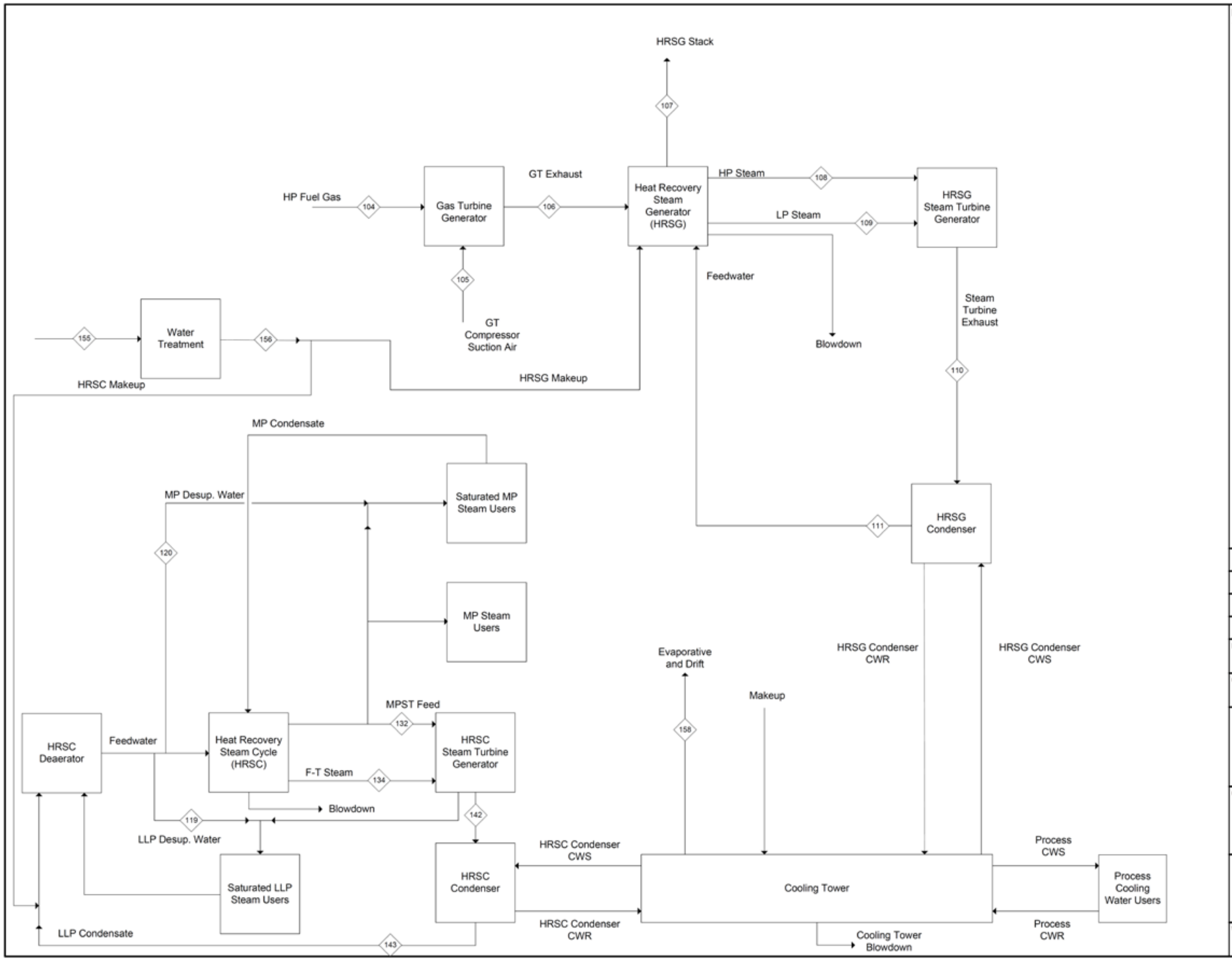


Figure 12. LBJ plant block flow diagram, F-T liquids production area. See stream tables following Figure 13.



NOTES:
1. DOE / NETL Award DE-FE0023697

REV	DATE	DESCRIPTION
A	06/01/2016	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: E.J. Mace
DATE: 06/01/2016
CHECKED BY:
PROJECT MANAGER: H.N. Goldstein

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
Princeton University-ESAG
Design / Cost Study for Synthetic Jet Fuel Production with CO_2 Capture & Storage

BLOCK FLOW DIAGRAM
Power Generation

WORLEYPARSONS DWG. NO.: PNCL-CBTL-0-SK-021-305-0003
REV: A

Figure 13. LBJ plant block flow diagram, power generation area. See stream tables following Figure 13.

Stream ID number		***1***	***2***	***3***	***4***	***5***	***6***	***7***	***8***	***9***	***17***
Temperature	C	30.0	62.0	101.9	30.0	62.0	101.9	424.5	15.6	25.0	1034.5
Pressure	BARA	1.01	1.01	1.01	1.01	1.01	1.01	55.00	51.70	44.82	44.80
Mass Flow	KG/SEC	17.9625	12.5507	5.4118	6.4351	4.2926	2.1425	5.5505	8.1453	2.7164	2.3385
CO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2	KG/SEC	0.00000	0.35559	0.00000	0.00000	0.19438	0.00000	0.00000	0.00000	0.00000	0.00000
CO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	2.71635	0.00000
H2O	KG/SEC	0.00000	2.76115	5.41176	0.00000	0.64389	2.14252	5.55049	0.00000	0.00000	0.00000
H2S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NH3	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00209	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.00000	1.35947	0.00000	0.00000	1.43207	0.00000	0.00000	8.10772	0.00000	0.00000
AR	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03049	0.00000	0.00000
N2	KG/SEC	0.00000	0.08620	0.00000	0.00000	0.01480	0.00000	0.00000	0.00713	0.00000	0.00000
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.17779	0.00000	0.00000	0.00129	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	5.66236	0.00000	0.00000	1.96629	0.00000	0.00000	0.00000	0.00000	0.15257
CH4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H24	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30H62	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH3OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H5OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H7OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H9OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H11OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H13OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H15OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H17OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H19OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H21OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H41OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-BUTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-PENTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEXANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-OCTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-NONANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COAL	KG/SEC	17.96245	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	6.43515	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	2.14605	0.00000	0.00000	0.03991	0.00000	0.00000	0.00000	0.00000	2.18596
Vapor Fraction			0.732	1.000		0.988	1.000	1.000	1.000	1.000	
Liquid Fraction			0.268	0		0.012	0	0	0	0	

Stream ID number		***1***	***2***	***3***	***4***	***5***	***6***	***7***	***8***	***9***	***11***
Mole Flow	KMOL/HR		1351	1081		639	428	1109	916	222	4901
Component Mole Fraction											
CO			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.2664
H2			0.4701	0.0000		0.5434	0.0000	0.0000	0.0000	0.0000	0.2008
CO2			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	1.0000	0.2002
H2O			0.4084	1.0000		0.2014	1.0000	1.0000	0.0000	0.0000	0.2892
H2S			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0039
COS			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0002
NH3			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL2			0.0001	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O2			0.1132	0.0000		0.2522	0.0000	0.0000	0.9960	0.0000	0.0000
AR			0.0000	0.0000		0.0000	0.0000	0.0000	0.0030	0.0000	0.0006
N2			0.0082	0.0000		0.0030	0.0000	0.0000	0.0010	0.0000	0.0028
NO			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO2			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0358
C2H6			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H12			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H14			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H16			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H18			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H20			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H22			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C11H24			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C12H26			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C13H28			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C14H30			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C15H32			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C16H34			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C17H36			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C18H38			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C19H40			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H42			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C30H62			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H10			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H12			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H14			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H16			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H18			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH3OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H5OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H7OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H9OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H11OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H13OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H15OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H17OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H19OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H21OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H41OH			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-BUTANE			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-PENTAN			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEXANE			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEPTAN			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-OCTANE			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-NONANE			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-DECANE			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H6			0.0000	0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Volume Flow (1 atm, 0 C)	CUM/HR		27174.0	32995.6		17341.4	13063.0	1081.5	405.2	86.8	11989.1
Average Molecular Weight			12.16	18.02		12.88	18.02	18.02	32.02	44.01	22.71

Stream ID number		***15***	***17***	***19***	***20***	***22***	***23***	***26***	***30***	***37***	***38***
Temperature	C	250.0	1034.5	15.0	190.9	1034.5	189.9	424.5	188.8	112.3	37.7
Pressure	BARA	42.75	44.80	43.58	42.74	44.80	42.70	55.00	40.56	37.56	37.56
Mass Flow	KG/SEC	30.9170	2.3385	1.7344	0.9234	3.4352	31.7260	4.3172	33.4163	7.1033	26.3130
CO	KG/SEC	10.15789	0.00000	0.00000	0.00016	1.12865	10.15773	0.00000	5.83718	0.00017	5.83700
H2	KG/SEC	0.55102	0.00000	0.00000	0.00001	0.06122	0.55116	0.00000	0.83634	0.00003	0.83631
CO2	KG/SEC	11.99635	0.00000	0.00000	0.00127	1.33293	11.99508	0.00000	18.64992	0.00963	18.64030
H2O	KG/SEC	7.09334	0.00000	1.73439	0.91946	0.78815	7.90700	4.31723	7.06930	7.03162	0.03769
H2S	KG/SEC	0.18263	0.00000	0.00000	0.00005	0.02029	0.18208	0.00000	0.17917	0.00110	0.17807
COS	KG/SEC	0.01358	0.00000	0.00000	0.00000	0.00151	0.01358	0.00000	0.00787	0.00001	0.00786
NH3	KG/SEC	0.00060	0.00000	0.00000	0.00002	0.00007	0.00000	0.00000	0.08323	0.06052	0.02271
HCN	KG/SEC	0.00002	0.00000	0.00000	0.00000	0.00000	0.00002	0.00000	0.01016	0.00002	0.01014
HCL	KG/SEC	0.00214	0.00000	0.00000	0.00244	0.00024	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	KG/SEC	0.03049	0.00000	0.00000	0.00000	0.00339	0.03049	0.00000	0.02503	0.00001	0.02502
N2	KG/SEC	0.10763	0.00000	0.00000	0.00000	0.01196	0.10764	0.00000	0.03159	0.00000	0.03159
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	0.15257	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	KG/SEC	0.78127	0.00000	0.00000	0.00001	0.08681	0.78126	0.00000	0.64730	0.00010	0.64719
C2H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H24	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30H62	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH3OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H5OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H7OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H9OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H11OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H13OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H15OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H17OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H19OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H21OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H41OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-BUTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-PENTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEXANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-OCTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-NONANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03915	0.00008	0.03907
COAL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	2.18596	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Vapor Fraction		1.000		0.000	0.000	1.000	1.000	1.000	1.000	0.000	1.000
Liquid Fraction		0.000		1.000	1.000	0.000	0.000	0.000	0.000	1.000	0.000

Stream ID number		***15***	***17***	***19***	***20***	***22***	***23***	***26***	***30***	***37***	***38***
Mole Flow	KMOL/HR	4901		347	184	545	5063	863	5374	1419	3955
Component Mole Fraction											
CO		0.2664		0.0000	0.0001	0.2664	0.2579	0.0000	0.1396	0.0000	0.1897
H2		0.2008		0.0000	0.0001	0.2008	0.1944	0.0000	0.2779	0.0000	0.3776
CO2		0.2002		0.0000	0.0006	0.2002	0.1938	0.0000	0.2839	0.0006	0.3856
H2O		0.2892		1.0000	0.9978	0.2892	0.3121	1.0000	0.2629	0.9903	0.0019
H2S		0.0039		0.0000	0.0000	0.0039	0.0038	0.0000	0.0035	0.0001	0.0048
COS		0.0002		0.0000	0.0000	0.0002	0.0002	0.0000	0.0001	0.0000	0.0001
NH3		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0033	0.0090	0.0012
HCN		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0003
HCL		0.0000		0.0000	0.0013	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL2		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O2		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
AR		0.0006		0.0000	0.0000	0.0006	0.0005	0.0000	0.0004	0.0000	0.0006
N2		0.0028		0.0000	0.0000	0.0028	0.0027	0.0000	0.0008	0.0000	0.0010
NO		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO2		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4		0.0358		0.0000	0.0000	0.0358	0.0346	0.0000	0.0270	0.0000	0.0367
C2H6		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H12		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H14		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H16		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H18		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H20		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H22		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C11H24		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C12H26		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C13H28		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C14H30		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C15H32		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C16H34		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C17H36		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C18H38		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C19H40		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H42		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C30H62		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H10		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H12		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H14		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H16		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H18		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH3OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H5OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H7OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H9OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H11OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H13OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H15OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H17OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H19OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H21OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H41OH		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-BUTANE		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-PENTAN		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEXANE		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEPTAN		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-OCTANE		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-NONANE		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-DECANE		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H6		0.0000		0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0005
Volume Flow (1 atm, 0 C)	CUM/HR	4913.9		6.2	3.8	1332.1	4419.4	841.2	4957.3	28.5	2634.4
Average Molecular Weight		22.71		18.02	18.05	22.71	22.56	18.02	22.39	18.02	23.95

Stream ID number		***40***	***41***	***42***	***46***	***47***	***48***	***54***	***57***	***59***	***64***
Temperature	C	32.0	41.1	69.2	39.6	31.9	32.0	86.0	48.9	41.1	25.0
Pressure	BARA	33.86	2.00	1.77	6.00	1.10	3.10	1.59	1.89	150.00	1.01
Mass Flow	KG/SEC	7.3255	0.7665	0.1880	1.7099	16.0685	1.9243	0.1044	9.6297	17.9929	2.2931
CO	KG/SEC	5.81199	0.00000	0.00002	0.00000	0.00780	0.01719	0.00033	0.00000	0.02500	0.00000
H2	KG/SEC	0.83631	0.00000	0.00000	0.00000	0.00000	0.00000	0.00005	0.00000	0.00000	0.00000
CO2	KG/SEC	0.00225	0.58818	0.11839	0.00070	16.04237	1.88810	0.01159	0.00000	17.93048	0.00105
H2O	KG/SEC	0.00000	0.01116	0.00768	1.70750	0.00000	0.00000	0.02898	9.62961	0.00000	0.00371
H2S	KG/SEC	0.00000	0.15945	0.01859	0.00001	0.00000	0.00000	0.00115	0.00000	0.00000	0.00000
COS	KG/SEC	0.00000	0.00735	0.00051	0.00000	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000
NH3	KG/SEC	0.00000	0.00000	0.02087	0.00146	0.00000	0.00000	0.06197	0.00000	0.00000	0.00000
HCN	KG/SEC	0.00000	0.00000	0.00764	0.00006	0.00000	0.00000	0.00008	0.00000	0.00000	0.00000
HCL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.52952
AR	KG/SEC	0.02502	0.00000	0.00000	0.00000	0.00000	0.00000	0.00001	0.00000	0.00000	0.02913
N2	KG/SEC	0.03159	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	1.72971
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	KG/SEC	0.61534	0.00000	0.00002	0.00000	0.01356	0.01827	0.00012	0.00000	0.03183	0.00000
C2H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H22	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C11H24	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C12H26	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C13H28	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C14H30	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C15H32	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C16H34	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30H62	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH3OH	KG/SEC	0.00297	0.00033	0.00413	0.00017	0.00478	0.00076	0.00004	0.00013	0.00554	0.00000
C2H5OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H7OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H9OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H11OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H13OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H15OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H17OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H19OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H21OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H41OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-BUTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-PENTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEXANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-OCTANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-NONANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.01017	0.00000	0.00000	0.00000	0.00009	0.00000	0.00000	0.00000
COAL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Vapor Fraction		1.000	1.000	1.000	0.000	1.000	1.000	1.000	0.000	0.000	1.000
Liquid Fraction		0.000	0.000	0.000	1.000	0.000	0.000	0.000	1.000	1.000	0.000

Stream ID number		***40***	***41***	***42***	***46***	***47***	***48***	***54***	***57***	***59***	***64***
Mole Flow	KMOL/HR	2385	68	20	342	1317	161	20	1924	1478	285
Component Mole Fraction											
CO		0.3131	0.0000	0.0001	0.0000	0.0008	0.0137	0.0021	0.0000	0.0022	0.0000
H2		0.6261	0.0000	0.0000	0.0000	0.0000	0.0000	0.0040	0.0000	0.0000	0.0000
CO2		0.0001	0.7111	0.4946	0.0002	0.9965	0.9602	0.0471	0.0000	0.9926	0.0003
H2O		0.0000	0.0330	0.0784	0.9988	0.0000	0.0000	0.2876	1.0000	0.0000	0.0026
H2S		0.0000	0.2489	0.1003	0.0000	0.0000	0.0000	0.0061	0.0000	0.0000	0.0000
COS		0.0000	0.0065	0.0016	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3		0.0000	0.0000	0.2253	0.0009	0.0000	0.0000	0.6507	0.0000	0.0000	0.0000
HCN		0.0000	0.0000	0.0520	0.0000	0.0000	0.0000	0.0005	0.0000	0.0000	0.0000
HCL		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.2088
AR		0.0009	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0092
N2		0.0017	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.7791
NO		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4		0.0579	0.0000	0.0002	0.0000	0.0023	0.0255	0.0013	0.0000	0.0048	0.0000
C2H6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H8		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H10		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H12		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H14		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H16		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H18		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H20		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H22		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C11H24		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C12H26		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C13H28		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C14H30		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C15H32		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C16H34		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C17H36		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C18H38		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C19H40		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H42		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C30H62		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H10		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H12		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H14		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H16		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H18		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH3OH		0.0001	0.0005	0.0237	0.0001	0.0004	0.0005	0.0002	0.0000	0.0004	0.0000
C2H5OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H7OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H9OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H11OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H13OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H15OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H17OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H19OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H21OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H41OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-BUTANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-PENTAN		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEXANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEPTAN		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-OCTANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-NONANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-DECANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H6		0.0000	0.0000	0.0239	0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000
Volume Flow (1 atm, 0 C)	CUM/HR	1808.7	874.8	311.7	6.4	30196.5	1296.2	375.1	35.8	98.7	6976.2
Average Molecular Weight		11.06	40.78	34.57	18.02	43.93	43.07	18.67	18.02	43.83	28.93

Stream ID number		***74***	***78***	***79***	***81***	***82***	***84***	***85***	***86***	***87***	***88***
Temperature	C	100.0	32.0	43.3	43.3	39.5	32.0	43.8	43.1	160.7	32.2
Pressure	BARA	1.00	33.86	49.28	1.01	33.86	33.86	4.46	11.36	11.36	1.01
Mass Flow	KG/SEC	3.3520	0.4018	0.2358	0.1584	0.1164	6.9237	3.1675	0.5568	1.5338	0.5393
CO	KG/SEC	0.00000	0.31876	0.00003	0.00003	0.00000	5.49323	0.00004	0.00097	0.00105	0.00001
H2	KG/SEC	0.00000	0.04587	0.09110	0.01366	0.11642	0.79044	0.00000	0.00002	0.00004	0.00000
CO2	KG/SEC	0.77931	0.00012	0.00047	0.00047	0.00000	0.00212	0.00125	0.00118	0.00055	0.00010
H2O	KG/SEC	0.29313	0.00000	0.00188	0.00188	0.00000	0.00000	3.14679	0.00010	0.00259	0.00001
H2S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NH3	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.10336	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	KG/SEC	0.02913	0.00137	0.00000	0.00000	0.00000	0.02365	0.00000	0.00008	0.00008	0.00000
N2	KG/SEC	1.80185	0.00173	0.00000	0.00000	0.00000	0.02986	0.00001	0.00006	0.00007	0.00000
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	KG/SEC	0.34523	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	KG/SEC	0.00000	0.03375	0.00025	0.00025	0.00000	0.58159	0.00000	0.00428	0.00310	0.00014
C2H6	KG/SEC	0.00000	0.00000	0.04525	0.04525	0.00000	0.00000	0.00000	0.00039	0.00014	0.00006
C3H8	KG/SEC	0.00000	0.00000	0.04050	0.04050	0.00000	0.00000	0.00000	0.00110	0.00024	0.00042
C4H10	KG/SEC	0.00000	0.00000	0.00206	0.00206	0.00000	0.00000	0.00000	0.00485	0.00064	0.00333
C5H12	KG/SEC	0.00000	0.00000	0.00192	0.00192	0.00000	0.00000	0.00000	0.01327	0.00115	0.01168
C6H14	KG/SEC	0.00000	0.00000	0.00148	0.00148	0.00000	0.00000	0.00000	0.02620	0.00164	0.02513
C7H16	KG/SEC	0.00000	0.00000	0.00090	0.00090	0.00000	0.00000	0.00000	0.03654	0.00196	0.03605
C8H18	KG/SEC	0.00000	0.00000	0.00034	0.00034	0.00000	0.00000	0.00000	0.04423	0.00252	0.04403
C9H20	KG/SEC	0.00000	0.00000	0.00014	0.00014	0.00000	0.00000	0.00000	0.04674	0.00354	0.04667
C10H22	KG/SEC	0.00000	0.00000	0.00005	0.00005	0.00000	0.00000	0.00000	0.04843	0.00565	0.04840
C11H24	KG/SEC	0.00000	0.00000	0.00004	0.00004	0.00000	0.00000	0.00000	0.04776	0.00914	0.04775
C12H26	KG/SEC	0.00000	0.00000	0.00002	0.00002	0.00000	0.00000	0.00000	0.04331	0.01370	0.04331
C13H28	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.03777	0.02119	0.03777
C14H30	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.02846	0.02957	0.02846
C15H32	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.02390	0.03931	0.02390
C16H34	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01605	0.04438	0.01605
C17H36	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01039	0.04785	0.01039
C18H38	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00640	0.04924	0.00640
C19H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00414	0.05139	0.00414
C20H42	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00241	0.05657	0.00241
C30H62	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00001	0.00342	1.09783	0.00342
C2H4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00012	0.00005	0.00001
C3H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00222	0.00050	0.00077
C4H8	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00530	0.00075	0.00340
C5H10	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00468	0.00044	0.00400
C6H12	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00730	0.00048	0.00692
C7H14	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01087	0.00057	0.01071
C8H16	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01159	0.00063	0.01153
C9H18	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.01116	0.00082	0.01114
C10H20	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00922	0.00101	0.00921
C20H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.02382	0.02046	0.02382
CH3OH	KG/SEC	0.00000	0.00016	0.00000	0.00000	0.00000	0.00280	0.01702	0.00003	0.00010	0.00003
C2H5OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00209	0.00444	0.00022	0.00424
C3H7OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00012	0.00088	0.00004	0.00086
C4H9OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00006	0.00019	0.00001	0.00019
C5H11OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00007	0.00015	0.00001	0.00015
C6H13OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00002	0.00015	0.00001	0.00015
C7H15OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00034	0.00003	0.00034
C8H17OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00031	0.00005	0.00031
C9H19OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00188	0.00046	0.00188
C10H21OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00206	0.00095	0.00206
C20H41OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00747	0.02097	0.00747
I-BUTANE	KG/SEC	0.00000	0.00000	0.02776	0.02776	0.00000	0.00000	0.00000	0.00018	0.00003	0.00011
I-PENTAN	KG/SEC	0.00000	0.00000	0.01284	0.01284	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEXANE	KG/SEC	0.00000	0.00000	0.00548	0.00548	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00207	0.00207	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-OCTANE	KG/SEC	0.00000	0.00000	0.00078	0.00078	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-NONANE	KG/SEC	0.00000	0.00000	0.00030	0.00030	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00011	0.00011	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COAL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Vapor Fraction		1.000	1.000	1.000	1.000	1.000	1.000	0.000	0.033	0.023	0.000
Liquid Fraction		0.000	0.000	0.000	0.000	0.000	0.000	1.000	0.967	0.977	1.000

Stream ID number		***74***	***78***	***79***	***81***	***82***	***84***	***85***	***86***	***87***	***88***
Mole Flow	KMOL/HR	388	131	175	37	208	2255	631	16	18	14
Component Mole Fraction											
CO		0.0000	0.3131	0.0000	0.0001	0.0000	0.3131	0.0000	0.0076	0.0076	0.0001
H_2		0.0000	0.6261	0.9301	0.6661	1.0000	0.6261	0.0000	0.0026	0.0039	0.0000
CO_2		0.1645	0.0001	0.0002	0.0010	0.0000	0.0001	0.0002	0.0059	0.0025	0.0006
H_2O		0.1512	0.0000	0.0022	0.0103	0.0000	0.0000	0.9965	0.0013	0.0289	0.0001
H_2S		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH_3		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL_2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2		0.0300	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
AR		0.0068	0.0009	0.0000	0.0000	0.0000	0.0009	0.0000	0.0005	0.0004	0.0000
N_2		0.5975	0.0017	0.0000	0.0000	0.0000	0.0017	0.0000	0.0004	0.0005	0.0000
NO		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO_2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2		0.0501	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4		0.0000	0.0579	0.0003	0.0015	0.0000	0.0579	0.0000	0.0588	0.0390	0.0022
C_2H_6		0.0000	0.0000	0.0310	0.1479	0.0000	0.0000	0.0000	0.0028	0.0009	0.0005
C_3H_8		0.0000	0.0000	0.0189	0.0903	0.0000	0.0000	0.0000	0.0055	0.0011	0.0024
C_4H_{10}		0.0000	0.0000	0.0007	0.0035	0.0000	0.0000	0.0000	0.0184	0.0022	0.0143
C_5H_{12}		0.0000	0.0000	0.0005	0.0026	0.0000	0.0000	0.0000	0.0405	0.0032	0.0403
C_6H_{14}		0.0000	0.0000	0.0004	0.0017	0.0000	0.0000	0.0000	0.0669	0.0038	0.0726
C_7H_{16}		0.0000	0.0000	0.0002	0.0009	0.0000	0.0000	0.0000	0.0803	0.0039	0.0896
C_8H_{18}		0.0000	0.0000	0.0001	0.0003	0.0000	0.0000	0.0000	0.0852	0.0044	0.0960
C_9H_{20}		0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0802	0.0056	0.0906
$C_{10}H_{22}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0749	0.0080	0.0847
$C_{11}H_{24}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0673	0.0118	0.0761
$C_{12}H_{26}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0560	0.0162	0.0633
$C_{13}H_{28}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0451	0.0231	0.0510
$C_{14}H_{30}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0316	0.0300	0.0357
$C_{15}H_{32}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0248	0.0373	0.0280
$C_{16}H_{34}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0156	0.0395	0.0176
$C_{17}H_{36}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0095	0.0401	0.0108
$C_{18}H_{38}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0055	0.0390	0.0063
$C_{19}H_{40}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0034	0.0385	0.0038
$C_{20}H_{42}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0019	0.0403	0.0021
$C_{30}H_{62}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0018	0.5227	0.0020
C_2H_4		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0010	0.0004	0.0001
C_3H_6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0116	0.0024	0.0045
C_4H_8		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0208	0.0027	0.0151
C_5H_{10}		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0147	0.0013	0.0142
C_6H_{12}		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0191	0.0012	0.0205
C_7H_{14}		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0244	0.0012	0.0272
C_8H_{16}		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0227	0.0011	0.0256
C_9H_{18}		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0195	0.0013	0.0220
$C_{10}H_{20}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0145	0.0015	0.0164
$C_{20}H_{40}$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0187	0.0147	0.0211
CH_3OH		0.0000	0.0001	0.0000	0.0000	0.0000	0.0001	0.0030	0.0002	0.0006	0.0002
C_2H_5OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0212	0.0010	0.0229
C_3H_7OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0032	0.0001	0.0036
C_4H_9OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0000	0.0006
$C_5H_{11}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0004	0.0000	0.0004
$C_6H_{13}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0003	0.0000	0.0004
$C_7H_{15}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0006	0.0001	0.0007
$C_8H_{17}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0005	0.0001	0.0006
$C_9H_{19}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0029	0.0006	0.0032
$C_{10}H_{21}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0029	0.0012	0.0032
$C_{20}H_{41}OH$		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0055	0.0141	0.0062
I-BUTANE		0.0000	0.0000	0.0098	0.0469	0.0000	0.0000	0.0000	0.0007	0.0001	0.0005
I-PENTAN		0.0000	0.0000	0.0037	0.0175	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEXANE		0.0000	0.0000	0.0013	0.0063	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-HEPTAN		0.0000	0.0000	0.0004	0.0020	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-OCTANE		0.0000	0.0000	0.0001	0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-NONANE		0.0000	0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-DECANE		0.0000	0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_6H_6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Volume Flow (1 atm, 0 C)	CUM/HR	12007.7	99.2	95.3	950.4	128.6	1709.5	12.2	4.1	9.4	2.8
Average Molecular Weight		31.14	11.06	4.85	15.56	2.02	11.06	18.07	122.56	308.77	134.30

Stream ID number		***89***	***90***	***91***	***92***	***93***	***94***	***96***	***97***	***98***	***99***
Temperature	C	121.1	121.2	44.3	43.5	232.7	123.5	37.8	37.8	43.3	43.3
Pressure	BARA	1.01	4.46	4.46	3.91	3.56	2.19	1.01	1.01	1.70	1.70
Mass Flow	KG/SEC	1.5210	1.1823	0.0039	1.4620	1.6574	0.3432	0.3129	1.4206	0.3368	0.2099
CO	KG/SEC	0.00002	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
H2	KG/SEC	0.00000	0.00000	0.00000	0.00102	0.00118	0.00000	0.00000	0.00000	0.00000	0.00220
CO2	KG/SEC	0.00003	0.00000	0.00000	0.00006	0.00002	0.00000	0.00000	0.00000	0.00000	0.00008
H2O	KG/SEC	0.00017	0.00080	0.00393	0.00025	0.00009	0.34322	0.00007	0.00152	0.33685	0.00432
H2S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NH3	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
AR	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
N2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
SO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	KG/SEC	0.00010	0.00000	0.00000	0.00001	0.00001	0.00000	0.00000	0.00000	0.00000	0.00002
C2H6	KG/SEC	0.00001	0.00000	0.00000	0.01172	0.00207	0.00000	0.00041	0.00000	0.00000	0.01338
C3H8	KG/SEC	0.00004	0.00000	0.00000	0.03227	0.00317	0.00000	0.00321	0.00000	0.00000	0.03224
C4H10	KG/SEC	0.00019	0.00000	0.00000	0.00503	0.00031	0.00000	0.00129	0.00000	0.00000	0.00404
C5H12	KG/SEC	0.00054	0.00000	0.00000	0.01406	0.00066	0.00000	0.00727	0.00000	0.00000	0.00745
C6H14	KG/SEC	0.00106	0.00000	0.00000	0.03072	0.00150	0.00000	0.02380	0.00010	0.00000	0.00832
C7H16	KG/SEC	0.00155	0.00000	0.00000	0.04559	0.00277	0.00000	0.04107	0.00219	0.00000	0.00509
C8H18	KG/SEC	0.00223	0.00000	0.00000	0.05383	0.00467	0.00000	0.00338	0.05497	0.00000	0.00015
C9H20	KG/SEC	0.00333	0.00000	0.00000	0.05674	0.00726	0.00000	0.00000	0.06400	0.00000	0.00000
C10H22	KG/SEC	0.00548	0.00000	0.00000	0.05609	0.01074	0.00000	0.00000	0.06683	0.00000	0.00000
C11H24	KG/SEC	0.00900	0.00000	0.00000	0.12153	0.03479	0.00000	0.00000	0.15632	0.00000	0.00000
C12H26	KG/SEC	0.01359	0.00000	0.00000	0.10592	0.04532	0.00000	0.00000	0.15123	0.00000	0.00000
C13H28	KG/SEC	0.02111	0.00004	0.00000	0.08819	0.05686	0.00000	0.00000	0.14501	0.00000	0.00000
C14H30	KG/SEC	0.02952	0.00042	0.00000	0.06872	0.06956	0.00000	0.00000	0.13786	0.00000	0.00000
C15H32	KG/SEC	0.03927	0.00228	0.00000	0.05442	0.07739	0.00000	0.00000	0.12953	0.00000	0.00000
C16H34	KG/SEC	0.04436	0.01013	0.00000	0.04205	0.08595	0.00000	0.00000	0.11788	0.00000	0.00000
C17H36	KG/SEC	0.04784	0.03919	0.00000	0.03329	0.10048	0.00000	0.00000	0.09458	0.00000	0.00000
C18H38	KG/SEC	0.04923	0.10412	0.00000	0.02849	0.12480	0.00000	0.00000	0.04917	0.00000	0.00000
C19H40	KG/SEC	0.05139	0.14980	0.00000	0.02210	0.13672	0.00000	0.00000	0.00902	0.00000	0.00000
C20H42	KG/SEC	0.05657	0.13711	0.00000	0.01293	0.12464	0.00000	0.00000	0.00046	0.00000	0.00000
C30H62	KG/SEC	1.09783	0.73840	0.00000	0.01935	0.71910	0.00000	0.00000	0.00004	0.00000	0.00000
C2H4	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H6	KG/SEC	0.00007	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H8	KG/SEC	0.00020	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H10	KG/SEC	0.00019	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H12	KG/SEC	0.00029	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H14	KG/SEC	0.00044	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H16	KG/SEC	0.00055	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H18	KG/SEC	0.00077	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H20	KG/SEC	0.00098	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H40	KG/SEC	0.02041	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH3OH	KG/SEC	0.00005	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C2H5OH	KG/SEC	0.00012	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C3H7OH	KG/SEC	0.00003	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C4H9OH	KG/SEC	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C5H11OH	KG/SEC	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C6H13OH	KG/SEC	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C7H15OH	KG/SEC	0.00003	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H17OH	KG/SEC	0.00005	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H19OH	KG/SEC	0.00046	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C10H21OH	KG/SEC	0.00095	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H41OH	KG/SEC	0.02096	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
I-BUTANE	KG/SEC	0.00001	0.00000	0.00000	0.05072	0.00351	0.00000	0.01032	0.00000	0.00000	0.04391
I-PENTAN	KG/SEC	0.00000	0.00000	0.00000	0.07354	0.00372	0.00000	0.03326	0.00000	0.00000	0.04399
I-HEXANE	KG/SEC	0.00000	0.00000	0.00000	0.08615	0.00416	0.00000	0.06141	0.00011	0.00000	0.02879
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00000	0.09116	0.00524	0.00000	0.08074	0.00253	0.00000	0.01313
I-OCTANE	KG/SEC	0.00000	0.00000	0.00000	0.09061	0.00719	0.00000	0.04669	0.04830	0.00000	0.00282
I-NONANE	KG/SEC	0.00000	0.00000	0.00000	0.08624	0.00995	0.00000	0.00001	0.09618	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00000	0.07918	0.01358	0.00000	0.00000	0.09275	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COAL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Vapor Fraction		0.000	0.000	0.000	0.076	0.148	1.000	0.009	0.000	0.000	1.000
Liquid Fraction		1.000	1.000	1.000	0.924	0.852	0.000	0.991	1.000	1.000	0.000

Stream ID number		***89***	***90***	***91***	***92***	***93***	***94***	***96***	***97***	***98***	***99***
Mole Flow	KMOL/HR	16	12	1	47	25	69	13	31	67	17
Component Mole Fraction											
CO		0.0002	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H2		0.0001	0.0000	0.0001	0.0390	0.0853	0.0000	0.0004	0.0000	0.0000	0.2326
CO2		0.0002	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0004
H2O		0.0021	0.0128	0.9999	0.0011	0.0007	1.0000	0.0011	0.0099	1.0000	0.0512
H2S		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
AR		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4		0.0015	0.0000	0.0000	0.0001	0.0001	0.0000	0.0000	0.0000	0.0000	0.0003
C2H6		0.0001	0.0000	0.0000	0.0301	0.0100	0.0000	0.0039	0.0000	0.0000	0.0950
C3H8		0.0002	0.0000	0.0000	0.0565	0.0105	0.0000	0.0208	0.0000	0.0000	0.1560
C4H10		0.0007	0.0000	0.0000	0.0067	0.0008	0.0000	0.0064	0.0000	0.0000	0.0148
C5H12		0.0017	0.0000	0.0000	0.0150	0.0013	0.0000	0.0288	0.0000	0.0000	0.0221
C6H14		0.0027	0.0000	0.0000	0.0275	0.0025	0.0000	0.0790	0.0001	0.0000	0.0206
C7H16		0.0034	0.0000	0.0000	0.0351	0.0040	0.0000	0.1173	0.0026	0.0000	0.0108
C8H18		0.0043	0.0000	0.0000	0.0364	0.0060	0.0000	0.0085	0.0564	0.0000	0.0003
C9H20		0.0058	0.0000	0.0000	0.0341	0.0082	0.0000	0.0000	0.0585	0.0000	0.0000
C10H22		0.0086	0.0000	0.0000	0.0304	0.0110	0.0000	0.0000	0.0551	0.0000	0.0000
C11H24		0.0128	0.0000	0.0000	0.0600	0.0324	0.0000	0.0000	0.1172	0.0000	0.0000
C12H26		0.0178	0.0000	0.0000	0.0480	0.0388	0.0000	0.0000	0.1041	0.0000	0.0000
C13H28		0.0255	0.0001	0.0000	0.0369	0.0449	0.0000	0.0000	0.0922	0.0000	0.0000
C14H30		0.0331	0.0006	0.0000	0.0267	0.0511	0.0000	0.0000	0.0815	0.0000	0.0000
C15H32		0.0411	0.0031	0.0000	0.0198	0.0531	0.0000	0.0000	0.0715	0.0000	0.0000
C16H34		0.0436	0.0129	0.0000	0.0143	0.0553	0.0000	0.0000	0.0610	0.0000	0.0000
C17H36		0.0443	0.0471	0.0000	0.0107	0.0609	0.0000	0.0000	0.0461	0.0000	0.0000
C18H38		0.0430	0.1181	0.0000	0.0086	0.0714	0.0000	0.0000	0.0227	0.0000	0.0000
C19H40		0.0426	0.1611	0.0000	0.0064	0.0742	0.0000	0.0000	0.0039	0.0000	0.0000
C20H42		0.0445	0.1401	0.0000	0.0035	0.0643	0.0000	0.0000	0.0002	0.0000	0.0000
C30H62		0.5776	0.5042	0.0000	0.0035	0.2478	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H6		0.0004	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H8		0.0008	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H10		0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H12		0.0008	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H14		0.0010	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H16		0.0011	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H18		0.0014	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20		0.0015	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40		0.0162	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH3OH		0.0003	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H5OH		0.0006	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C3H7OH		0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C4H9OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H11OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H13OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H15OH		0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H17OH		0.0001	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H19OH		0.0007	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C10H21OH		0.0013	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H41OH		0.0156	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-BUTANE		0.0000	0.0000	0.0000	0.0673	0.0088	0.0000	0.0508	0.0000	0.0000	0.1612
I-PENTAN		0.0000	0.0000	0.0000	0.0787	0.0075	0.0000	0.1319	0.0000	0.0000	0.1301
I-HEXANE		0.0000	0.0000	0.0000	0.0771	0.0070	0.0000	0.2039	0.0001	0.0000	0.0713
I-HEPTAN		0.0000	0.0000	0.0000	0.0702	0.0076	0.0000	0.2305	0.0030	0.0000	0.0280
I-OCTANE		0.0000	0.0000	0.0000	0.0612	0.0092	0.0000	0.1169	0.0496	0.0000	0.0053
I-NONANE		0.0000	0.0000	0.0000	0.0519	0.0113	0.0000	0.0000	0.0879	0.0000	0.0000
I-DECANE		0.0000	0.0000	0.0000	0.0429	0.0139	0.0000	0.0000	0.0764	0.0000	0.0000
C6H6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Volume Flow (1 atm, 0 C)	CUM/HR	7.6	5.8	0.0	31.5	52.2	1008.2	4.6	7.1	1.2	254.6
Average Molecular Weight		338.38	341.35	18.01	112.83	241.48	18.02	89.52	166.54	18.02	44.81

Stream ID number		***100***	***101***	***102***	***104***	***105***	***106***	***107***	***155***
Temperature	C	43.3	32.2	121.1	117.3	15.0	559.4	90.0	47.5
Pressure	BARA	23.52	1.01	1.01	30.32	1.01	1.01	1.01	1.70
Mass Flow	KG/SEC	1.6656	0.0175	0.0127	2.4263	84.5734	86.9997	86.9997	13.1380
CO	KG/SEC	0.43736	0.00095	0.00103	0.75814	0.00000	0.00005	0.00005	0.00004
H2	KG/SEC	0.02348	0.00002	0.00004	0.04628	0.00000	0.00000	0.00000	0.00000
CO2	KG/SEC	0.09472	0.00108	0.00052	0.09699	0.03870	5.47549	5.47549	0.00125
H2O	KG/SEC	0.00719	0.00009	0.00242	0.01591	0.54605	3.82394	3.82394	13.11720
H2S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
NH3	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCN	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
HCL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CL2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
O2	KG/SEC	0.00000	0.00000	0.00000	0.00009	19.44775	13.03670	13.03670	0.00000
AR	KG/SEC	0.02349	0.00008	0.00007	0.02502	1.07735	1.10237	1.10237	0.00000
N2	KG/SEC	0.02972	0.00006	0.00007	0.03158	63.46358	63.43800	63.43800	0.00001
NO	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.12104	0.12104	0.00000
NO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00212	0.00212	0.00000
SO2	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
S	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
CH4	KG/SEC	0.84167	0.00414	0.00300	0.88283	0.00000	0.00000	0.00000	0.00000
C2H6	KG/SEC	0.01720	0.00033	0.00013	0.07628	0.00000	0.00000	0.00000	0.00000
C3H8	KG/SEC	0.01613	0.00068	0.00020	0.08974	0.00000	0.00000	0.00000	0.00000
C4H10	KG/SEC	0.02342	0.00152	0.00045	0.03149	0.00000	0.00000	0.00000	0.00000
C5H12	KG/SEC	0.02209	0.00160	0.00061	0.03367	0.00000	0.00000	0.00000	0.00000
C6H14	KG/SEC	0.01555	0.00107	0.00058	0.02697	0.00000	0.00000	0.00000	0.00000
C7H16	KG/SEC	0.00808	0.00049	0.00041	0.01495	0.00000	0.00000	0.00000	0.00000
C8H18	KG/SEC	0.00356	0.00019	0.00029	0.00451	0.00000	0.00000	0.00000	0.00000
C9H20	KG/SEC	0.00147	0.00007	0.00022	0.00186	0.00000	0.00000	0.00000	0.00000
C10H22	KG/SEC	0.00060	0.00002	0.00018	0.00082	0.00000	0.00000	0.00000	0.00000
C11H24	KG/SEC	0.00023	0.00001	0.00014	0.00038	0.00000	0.00000	0.00000	0.00000
C12H26	KG/SEC	0.00009	0.00000	0.00011	0.00018	0.00000	0.00000	0.00000	0.00000
C13H28	KG/SEC	0.00003	0.00000	0.00008	0.00007	0.00000	0.00000	0.00000	0.00000
C14H30	KG/SEC	0.00001	0.00000	0.00005	0.00002	0.00000	0.00000	0.00000	0.00000
C15H32	KG/SEC	0.00000	0.00000	0.00004	0.00001	0.00000	0.00000	0.00000	0.00000
C16H34	KG/SEC	0.00000	0.00000	0.00002	0.00000	0.00000	0.00000	0.00000	0.00000
C17H36	KG/SEC	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000
C18H38	KG/SEC	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000
C19H40	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H42	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C30H62	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00001
C2H4	KG/SEC	0.00784	0.00011	0.00005	0.00800	0.00000	0.00000	0.00000	0.00000
C3H6	KG/SEC	0.03794	0.00146	0.00043	0.03983	0.00000	0.00000	0.00000	0.00000
C4H8	KG/SEC	0.03038	0.00190	0.00055	0.03282	0.00000	0.00000	0.00000	0.00000
C5H10	KG/SEC	0.00939	0.00068	0.00025	0.01031	0.00000	0.00000	0.00000	0.00000
C6H12	KG/SEC	0.00540	0.00038	0.00019	0.00596	0.00000	0.00000	0.00000	0.00000
C7H14	KG/SEC	0.00254	0.00016	0.00013	0.00282	0.00000	0.00000	0.00000	0.00000
C8H16	KG/SEC	0.00112	0.00006	0.00008	0.00126	0.00000	0.00000	0.00000	0.00000
C9H18	KG/SEC	0.00041	0.00002	0.00005	0.00048	0.00000	0.00000	0.00000	0.00000
C10H20	KG/SEC	0.00012	0.00001	0.00003	0.00016	0.00000	0.00000	0.00000	0.00000
C20H40	KG/SEC	0.00005	0.00000	0.00006	0.00000	0.00000	0.00000	0.00000	0.00000
CH3OH	KG/SEC	0.00005	0.00000	0.00005	0.00027	0.00000	0.00000	0.00000	0.01715
C2H5OH	KG/SEC	0.00285	0.00020	0.00010	0.00314	0.00000	0.00000	0.00000	0.00209
C3H7OH	KG/SEC	0.00024	0.00002	0.00001	0.00027	0.00000	0.00000	0.00000	0.00012
C4H9OH	KG/SEC	0.00002	0.00000	0.00000	0.00002	0.00000	0.00000	0.00000	0.00006
C5H11OH	KG/SEC	0.00001	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000	0.00007
C6H13OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00002
C7H15OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C8H17OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C9H19OH	KG/SEC	0.00000	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000	0.00000
C10H21OH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
C20H41OH	KG/SEC	0.00000	0.00000	0.00001	0.00000	0.00000	0.00000	0.00000	0.00000
I-BUTANE	KG/SEC	0.00116	0.00007	0.00002	0.07292	0.00000	0.00000	0.00000	0.00000
I-PENTAN	KG/SEC	0.00000	0.00000	0.00000	0.05682	0.00000	0.00000	0.00000	0.00000
I-HEXANE	KG/SEC	0.00000	0.00000	0.00000	0.03425	0.00000	0.00000	0.00000	0.00000
I-HEPTAN	KG/SEC	0.00000	0.00000	0.00000	0.01518	0.00000	0.00000	0.00000	0.00000
I-OCTANE	KG/SEC	0.00000	0.00000	0.00000	0.00359	0.00000	0.00000	0.00000	0.00000
I-NONANE	KG/SEC	0.00000	0.00000	0.00000	0.00029	0.00000	0.00000	0.00000	0.00000
I-DECANE	KG/SEC	0.00000	0.00000	0.00000	0.00010	0.00000	0.00000	0.00000	0.00000
C6H6	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
COAL	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
BIOMASS	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
ASH	KG/SEC	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000	0.00000
Vapor Fraction		1.000	1.000	0.999	1.000	1.000	1.000	1.000	0.000
Liquid Fraction		0.000	0.000	0.001	0.000	0.000	0.000	0.000	1.000

Stream ID number		***100***	***101***	***102***	***104***	***105***	***106***	***107***	***155***
Mole Flow	KMOL/HR	317	2	2	435	10553	10945	10945	2623
Component Mole Fraction									
CO		0.1776	0.0645	0.0781	0.2241	0.0000	0.0000	0.0000	0.0000
H2		0.1324	0.0220	0.0407	0.1901	0.0000	0.0000	0.0000	0.0000
CO2		0.0245	0.0467	0.0249	0.0182	0.0003	0.0409	0.0409	0.0000
H2O		0.0045	0.0099	0.2846	0.0073	0.0103	0.0698	0.0698	0.9992
H2S		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
COS		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
NH3		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCN		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCL		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CL2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O2		0.0000	0.0000	0.0000	0.0000	0.2073	0.1340	0.1340	0.0000
AR		0.0067	0.0039	0.0039	0.0052	0.0092	0.0091	0.0091	0.0000
N2		0.0121	0.0038	0.0051	0.0093	0.7728	0.7448	0.7448	0.0000
NO		0.0000	0.0000	0.0000	0.0000	0.0000	0.0013	0.0013	0.0000
NO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO2		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH4		0.5967	0.4902	0.3960	0.4556	0.0000	0.0000	0.0000	0.0000
C2H6		0.0065	0.0207	0.0088	0.0210	0.0000	0.0000	0.0000	0.0000
C3H8		0.0042	0.0292	0.0095	0.0168	0.0000	0.0000	0.0000	0.0000
C4H10		0.0046	0.0497	0.0165	0.0045	0.0000	0.0000	0.0000	0.0000
C5H12		0.0035	0.0420	0.0180	0.0039	0.0000	0.0000	0.0000	0.0000
C6H14		0.0021	0.0235	0.0143	0.0026	0.0000	0.0000	0.0000	0.0000
C7H16		0.0009	0.0093	0.0087	0.0012	0.0000	0.0000	0.0000	0.0000
C8H18		0.0004	0.0032	0.0054	0.0003	0.0000	0.0000	0.0000	0.0000
C9H20		0.0001	0.0010	0.0036	0.0001	0.0000	0.0000	0.0000	0.0000
C10H22		0.0000	0.0003	0.0026	0.0000	0.0000	0.0000	0.0000	0.0000
C11H24		0.0000	0.0001	0.0019	0.0000	0.0000	0.0000	0.0000	0.0000
C12H26		0.0000	0.0000	0.0014	0.0000	0.0000	0.0000	0.0000	0.0000
C13H28		0.0000	0.0000	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
C14H30		0.0000	0.0000	0.0006	0.0000	0.0000	0.0000	0.0000	0.0000
C15H32		0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000
C16H34		0.0000	0.0000	0.0002	0.0000	0.0000	0.0000	0.0000	0.0000
C17H36		0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
C18H38		0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
C19H40		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C20H42		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C30H62		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C2H4		0.0032	0.0074	0.0036	0.0024	0.0000	0.0000	0.0000	0.0000
C3H6		0.0103	0.0658	0.0215	0.0078	0.0000	0.0000	0.0000	0.0000
C4H8		0.0062	0.0641	0.0207	0.0048	0.0000	0.0000	0.0000	0.0000
C5H10		0.0015	0.0184	0.0075	0.0012	0.0000	0.0000	0.0000	0.0000
C6H12		0.0007	0.0085	0.0048	0.0006	0.0000	0.0000	0.0000	0.0000
C7H14		0.0003	0.0030	0.0027	0.0002	0.0000	0.0000	0.0000	0.0000
C8H16		0.0001	0.0011	0.0016	0.0001	0.0000	0.0000	0.0000	0.0000
C9H18		0.0000	0.0003	0.0009	0.0000	0.0000	0.0000	0.0000	0.0000
C10H20		0.0000	0.0001	0.0005	0.0000	0.0000	0.0000	0.0000	0.0000
C20H40		0.0000	0.0000	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000
CH3OH		0.0000	0.0002	0.0036	0.0001	0.0000	0.0000	0.0000	0.0007
C2H5OH		0.0007	0.0082	0.0044	0.0006	0.0000	0.0000	0.0000	0.0001
C3H7OH		0.0000	0.0005	0.0004	0.0000	0.0000	0.0000	0.0000	0.0000
C4H9OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C5H11OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H13OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C7H15OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C8H17OH		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C9H19OH		0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
C10H21OH		0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
C20H41OH		0.0000	0.0000	0.0001	0.0000	0.0000	0.0000	0.0000	0.0000
I-BUTANE		0.0002	0.0023	0.0007	0.0104	0.0000	0.0000	0.0000	0.0000
I-PENTAN		0.0000	0.0000	0.0000	0.0065	0.0000	0.0000	0.0000	0.0000
I-HEXANE		0.0000	0.0000	0.0000	0.0033	0.0000	0.0000	0.0000	0.0000
I-HEPTAN		0.0000	0.0000	0.0000	0.0013	0.0000	0.0000	0.0000	0.0000
I-OCTANE		0.0000	0.0000	0.0000	0.0003	0.0000	0.0000	0.0000	0.0000
I-NONANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
I-DECANE		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C6H6		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Volume Flow (1 atm, 0 C)	CUM/HR	343.7	47.2	54.8	459.8	249520.0	747752.0	326006.0	49.7
Average Molecular Weight		18.94	33.17	26.93	20.09	28.85	28.62	28.62	18.03

Table 4. Summary of LBJ FOAK plant mass and energy balances.

Feedstock inputs		
Total feedstock input	MW, HHV	295
Biomass % of feedstock	% of HHV	25.0%
	metric t/d A.R.	1,551
	MW, HHV	221
	MW, LHV (A.R. basis)	193
	metric t/d A.R.	556
	MW, HHV	74
	MW, LHV (A.R. basis)	63
FT liquid outputs		
	actual bbl/day	1,004
	bbl/day equiv. petroleum jet	900
	MW, LHV	62.3
	actual bbl/day	248
	bbl/day equiv. petro-naphtha	223
	MW, LHV	13.9
	actual bbl/day	1,252
	bbl/day equiv. petroleum	1,124
	MW, LHV	76.2
Other products		
CO_2 captured & sold for EOR	metric t/day	1,326
Sulfuric acid (93 wt% H_2SO_4)	metric t/day	49
Electricity		
GTCC generation	MW_e	42.93
HRSC generation	MW_e	10.04
Total generation	MW_e	53.0
On-site use, MW_e	MW_e	38.2
Net electricity production	MW_e	14.8
Net electricity fraction	% of energy outputs, LHV	16.3%

Table 5. LBJ FOAK plant auxiliary loads.

	MW_e
Lignite prep.	2.98
Biomass prep.	1.07
Gasifier island	0.03
Acid gas removal system	5.18
Sour water stripper	0.02
CO_2 compressor	6.84
Wet Sulfuric Acid plant	0.12
H2 recovery	0.13
F-T island	0.76
GT fuel gas compressor	1.56
Air separation unit	14.84
Tempered water system	0.08
Cooling water system	2.05
Water treatment system	0.47
Wastewater treatment	0.39
Miscellaneous	1.72
TOTAL	38.21

Table 6. LBJ FOAK plant carbon flows.

	metric tCO_{2e}/day
Lignite input	1,792
Biomass input	622
TOTAL INPUT	2,414
C in SPK product	378
C in naphtha	83
CO_2 captured	1,326
CO_2 vented	538
C in gasifier ash	57
TOTAL OUTPUT[a]	2,382

a) Carbon balance closure is within bounds allowed by gasifier performance estimates available to the authors.

4 FOAK System-Wide Greenhouse Emissions Analysis

A key design goal for the LBJ plant is to produce synthetic jet fuel (SPK) having lifecycle greenhouse gas (GHG) emissions less than for petroleum-derived jet fuel. Figure 14 shows the GHG flows that were considered in estimating the net full fuel-cycle emissions associated with production and use of the SPK. Most of the GHG flows are in the form of CO_2, but emissions of CH_4, N_2O, and other radiatively-active gases are also included in the analysis. Emissions to the atmosphere include CO_2 from combustion in the gas turbine and WSA at the plant and from combustion of the produced fuels. Also included are GHG emissions associated with *(i)* biomass land preparation, harvesting and transporting, *(ii)* lignite mining and delivery, *(iii)* delivering the liquid products to users, and *(iv)* delivering CO_2 to the pipeline network that currently carries natural CO_2 from the Jackson Dome to oil fields. Biomass growth results in net removal of CO_2 from the atmosphere, including some transfer of carbon into the soil under the growing trees. Some of the latter will return as CO_2 to the atmosphere via decomposition of roots.

The lifecycle emissions analysis considers that net exported electricity displaces an equivalent amount of grid electricity and avoids emissions associated with the latter. Also, CO_2 delivered to the pipeline network serving oil fields is assumed to displace CO_2 that would otherwise have come from the Jackson Dome natural CO_2 formation. This is an appropriate assumption for the near-term if the existing pipeline network for CO_2 delivery to oil fields is already operating at capacity. In the longer term, once CO_2 pipeline capacity will have been enlarged, it would be more appropriate to assume that incremental oil production resulting from EOR using CO_2 from an LBJ plant would displace an equivalent amount of conventional oil production. This latter approach is used in the Nth-of-a-kind (NOAK) plant analysis in Section 6.

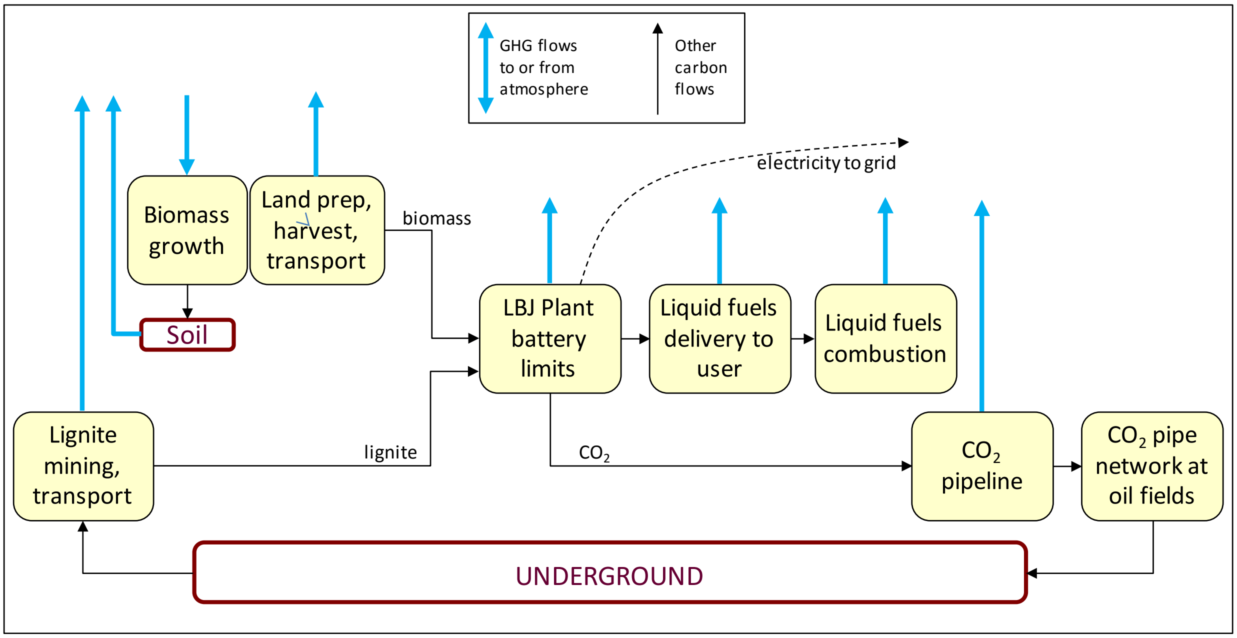


Figure 14. Greenhouse gas flows considered in estimating GHG emissions of the SPK produced by the LBJ FOAK plant.

4.1 Greenhouse Gas Emissions Metrics

We characterize GHG emissions for the LBJ FOAK project using two metrics. One metric is the Greenhouse Gas Emissions Index (GHGI), first introduced by Liu *et al.*,[25] and defined as the net GHG emissions for the LBJ system divided by the lifecycle emissions for a reference system that produces an equivalent amount of products by conventional means:

$$GHGI_{2005} = \frac{\text{Net LBJ system emissions}}{\text{Average U.S. lifecycle emissions in 2005 for equivalent products}}$$

The numerator includes all GHG flows shown in Figure 14 except for the avoided grid emissions. The reference system is taken to be the average U.S. emissions in 2005 associated with producing and using an equivalent amount of liquid fuels from crude oil, an equivalent amount of electricity generated with the grid-average GHG emissions per kWh in 2005, and an equivalent amount of natural dome CO_2 for EOR. (For the F-T naphtha, we assume it would be used for producing gasoline blending components and we assign lifecycle emissions associated with an energy-equivalent amount of petroleum-derived gasoline.) The reference year of 2005 is selected because U.S. national carbon mitigation goals as of 2016 (28% reduction by 2025 and 80% reduction by 2050) are expressed relative to emissions in 2005,[26,27] and thus $GHGI_{2005}$ provides a measure by which to gauge the ambitiousness of the emissions reductions achieved by the LBJ system relative to national goals. A $GHGI_{2005}$ value less than unity means that the LBJ system provides emissions reductions relative to the reference system. GHGI is a particularly informative metric when comparing processes that have different mixes of co-products because it does not require allocating emissions among co-products.

For the second metric used to characterize LBJ FOAK GHG emissions, we allocate a fraction of the net LBJ system emissions to each co-product and attribute the remaining emissions to the main product, SPK. SPK is our "functional unit" in the language of the lifecycle analysis literature.[28] The emissions estimated for SPK in this manner can be compared with the lifecycle emissions estimated for petroleum-derived jet fuel and other alternative jet fuel options.

$$SPK\ emissions \left[\frac{kgCO_{2e}}{GJ_{SPK,LHV}}\right] = \frac{\text{Net LBJ system emissions - emissions allocated to (electricity+naphtha+CO2 for EOR)}}{\text{energy content of LBJ SPK}}$$

For a process like LBJ, which has several co-products, different methods can be considered for allocating emissions among the co-products.[29] We have chosen to allocate using a "displacement" approach, whereby each coproduct is allocated an amount of emissions that would be avoided when it displaces a conventionally-produced equivalent product. This displacement approach is considered reasonable when the main product accounts for the major part of total output.[30] In our design SPK accounts for 61% of LBJ energy outputs, while electricity and naphtha co-products account for 25% and 14%, respectively. Sulfuric acid is a very minor co-product, and we choose to allocate zero emissions to it. For consistency, we allocate emissions to electricity, F-T naphtha, and CO_2 for EOR the same lifecycle emissions as used in calculating the reference system emissions for the $GHGI_{2005}$ metric.

Table 7 gives values of input parameters used in the calculation of the two GHG metrics. For consistency in our analysis, parameter values are based to the extent possible on lifecycle analyses published by the National Energy Technology Laboratory (NETL). The most important departure from NETL sources relates to the estimation of the net biogenic emissions, as described in the next section.

Table 7. Parameter values for calculating GHG emission metrics.

	Value	Source
LBJ System Parameters		
Lignite mining and delivery, $kgCO_{2e}/GJ_{lignite,HHV}$	4.93	[31]
Biomass land preparation, harvest, delivery, $kgCO_{2e}/GJ_{biomass,HHV}$	3.95[a]	[32, 33]
Net biogenic emissions, $kgCO_{2e}/GJ_{SPK,LHV}$ (scenario dependent)	see Table 10	
Plant emissions of lignite origin, $kgCO_{2e}/GJ_{SPK,LHV}$	75.1	b
SPK transport to user, $kgCO_{2e}/GJ_{SPK,LHV}$	0.84[c]	[34]
FT naphtha transport to user, $kgCO_{2e}/GJ_{naphtha,LHV}$	0.91[c]	[34]
Emission during 250 km CO_2 transport to oil-field pipe network, $kgCO_2/tCO_2$ delivered	6.83	[35]
Oil field emissions with water-alternating-gas EOR, $kgCO_{2e}/tCO_2$ stored[d]	204[e]	[36]
Oil field emissions for conventional oil production, $kgCO_{2e}/GJ_{oil,LHV}$[d]	6.66	[34]
Reference System Parameters (2005 U.S. average lifecycle emissions)		
Electric grid, $kgCO_{2e}/MWh$	661	f
Crude-oil derived jet fuel, $kgCO_{2e}/GJ_{jet\ fuel,LHV}$	88	[34]
Crude-oil derived gasoline (naphtha), $kgCO_{2e}/GJ_{jet\ fuel,LHV}$	91	[34]
Natural dome CO_2 extraction and delivery to EOR site, $kgCO_{2e}/kgCO_2$ sequestered	0.074	[37,38]

(a) Excludes emissions associated with biomass debarking, chipping, drying, and milling of the biomass. These are accounted for in the net emissions at the plant.

(b) This is based on the carbon input to the plant as lignite multiplied by the fraction of total input carbon that is vented at the plant. See Table 6.

(c) Assumed equal to emissions for transport of petroleum-derived jet fuel (for SPK) or gasoline (for FT naphtha).

(d) This parameter is relevant only for the NOAK plant analysis (Section 6). For the NOAK analysis, emissions from oil field operations associated with EOR are charged to the LBJ system. These are calculated as the difference in emissions per barrel of oil produced via EOR versus via conventional oil production.

(e) Assuming 3 incremental barrels of oil produced per tonne of CO_2 stored, and 68 $kgCO_{2e}$ emitted per incremental barrel produced.

(f) This is an estimate of the U.S. national average lifecycle emissions for electricity generated on the grid in 2005. Emissions at power plants are estimated from EPA's eGRID database,[39] and upstream emissions associated with extraction and delivery of fossil fuels to power plants is based on the GREET model.[40]

4.2 Estimating GHG Emissions of Biogenic Origin

Estimating the emissions of biogenic origin requires especially careful accounting. As discussed in Section 3.2, the biomass feedstocks for the LBJ plant will be pulpwood-quality southern pine logs from managed plantations. Unlike the case with biomass energy feedstocks that grow on an annual cycle such as perennial grasses, assessing the GHG emissions associated with the production and use of tree logs is complicated by the longer time required to grow a tree to replace the harvested tree.

For biomass produced and used on an annual cycle, e.g., crop residues, the carbon emitted in using the biomass, e.g., by combustion for power generation (without CCS), is absorbed from the atmosphere by the following year's re-growth of biomass. Thus, to a first approximation, there is no net accumulation of CO_2 in the atmosphere from using this biomass for energy. Moreover, with crop residues, if they were not used for energy their carbon content would largely oxidize to CO_2 by the following year's growing season, implying that there would be no net change in the CO_2 content of the atmosphere compared to the case when the crop residues are used for energy (without CCS).

Unlike the case with crop residues, however, if a tree is cut and burned for power generation (without CCS), the time required for a replacement tree to absorb as much CO_2 from the atmosphere as was emitted by burning the first tree can be very long. For example, Walker *et al.*[41] estimate that if mature hardwood trees were to be harvested from a forest in Massachusetts and used to generate electricity in utility-scale plants, the net emissions after 40 years (even with replanting of the trees) would be higher than if the power had been generated using coal instead of biomass. The managed pine plantations of the Southeast U.S. represent a very different situation from that of mature hardwood forests of the Northeast. There are a variety of management strategies for planted pine forests in the Southeast region that can avoid the undesirable result found by Walker *et al.* for Massachussetts. The biomass feedstock supply strategy for the LBJ plant recommended by the Antares Group[16,19] is designed specifically to ensure that the production and use of tree-derived biomass provides net GHG emissions benefits within the lifetime of the LBJ project.

4.2.1 Biomass feedstock supply strategy for the LBJ project

Southern forests cover about 200 million acres, or 40% of the land area of the 13 states that make up the Southern Region of the US Forest Service.[42] About one-quarter of this forest acreage is owned by large private companies, including Timberland Investment Management Organizations (TIMOs), Timber Real Estate Investment Trusts (TREITs), and forest product companies. These entities typically employ advanced management practices (including improved planting stocks, fertilization, and thinning) to increase final saw timber and pulpwood yield on managed plantations. Nearly 80% of the forest acres not owned by large private companies are family forests (non-industrial private forests, NIPF). It is uncommon for NIPF owners to fertilize,[43] to thin, or to use other intensified management practices. NIPF stands are typically not managed to optimize merchantable production and are clear cut after a full rotation time between 20 and 35 years. The resulting mix of saw timber and pulpwood is sold and the land is replanted. NIPFs make up 70% of forest acreage in Mississippi, with a similar percentage in Alabama. Antares estimates that at least half of the NIPF acres in these states could produce higher yields cost-effectively[16,19] by improving management practices, and this provides the basis for a sustainable biomass supply strategy that can provide net GHG benefits when used in the LBJ project (and other potential woody biomass energy projects in the future).

Antares quantified these benefits using an analytical framework proposed by the U.S. Environmental Protection Agency (EPA) for assessing biogenic CO_2 emissions from stationary sources.[44] The EPA framework is likely to be revised before being implemented in regulation, but the current version provides a comprehensive and transparent basis for the analysis here. A key feature of the proposed EPA framework is the establishing of a counter-factual or "business-as-usual" (BAU) scenario in which a biomass-using energy project is *not* built. For the LBJ analysis, the assumed BAU scenario is that NIPF stands continue to be managed as most are today, with no use of fertilization, thinning, or improved planting stock aimed at increasing yields. The stands are assumed to be harvested at 25 years of age and produce a mix of saw logs and pulpwood. Although this BAU scenario is a reasonable benchmark, commercial forestry is a dynamic industry in the Southeast region, e.g., with recent pulp and paper mill closures and startups of pellet fuel manufacturing. Taking account of such dynamics is beyond the scope of this analysis, but might be warranted at a future stage of project development.

The biomass feedstock for the LBJ plant is envisioned to be pulpwood-quality (13 to 23 cm in diameter at breast height) southern pine logs that result from thinning of planted NIPF stands in the 15th year of a 25-year rotation. Fertilizer would be applied after thinning to promote increased growth compared to BAU during the last decade of growth before final harvest. Based on studies of managed loblolly pine forests,[45] Antares estimates that the total above-ground biomass (excluding dead or fallen branches or trees) produced per hectare during the decade after thinning would increase on average by 22% compared to a stand that is not thinned or fertilized.[19] Additional management measures, including improving planting stocks and controlling competing vegetation, would increase production still further. Antares defined three scenarios (low, medium, and high) of increased production that would come from increasingly-intensive management (Table 8).

Antares used the estimates in Table 8 together with an industry-standard yield/growth model for loblolly pine plantations[46] to estimate biomass production levels over the full 25-year rotation under a BAU scenario and low, medium, and high management-intensity scenarios. The estimates were made for each of four regions of the U.S. South defined by ecological characteristics (soils, climate, geography, etc.), state borders, and market/transportation infrastructure boundaries (Figure 15). Mississippi and Alabama constitute region 2 (Gulf Coast/Coastal Plain East) in Antares' analysis, and Table 9 shows the biomass production levels estimated by Antares for this region. The estimates in this table are the basis for estimating the GHG emissions associated with the biomass used in the LBJ FOAK project, as detailed in Section 4.2.2.

Table 8. Alternative scenarios for biomass supply to the LBJ project from planted non-industrial private forest stands.[19]

Scenario >>>	Low	Med	High
Increase in total above-ground biomass growth/ha after thinning thru final harvest			
Thinning and fertilizing	22%	22%	22%
Improving planting stock	--	33%	33%
Controlling competing vegetation	--	--	25%
Total due to intensified management practices	22%	55%	80%

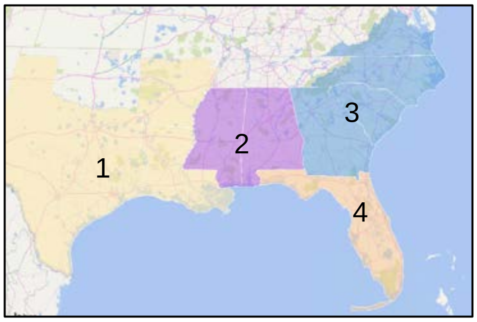


Figure 15. Regions for biomass supply analysis.

Table 9. Representative biomass production levels over a 25-year rotation in the Mississippi/Alabama region from NIPF stands subjected to intensified management scenarios, each of which includes thinning in year 15 followed by fertilization.

dry tonnes/ha	BAU	Low	Med	High
Merchantable biomass				
Pulpwood thinnings	0.0	10.9	10.9	10.9
Pulpwood @ final harvest	80.3	87.7	89.5	90.9
Sawlogs @ final harvest	60.0	65.6	66.9	68.0
Total	**140.3**	**164.2**	**167.3**	**169.8**
Total live above-ground biomass, including tops and branches				
Pulpwood thinnings	0.0	13.1	13.1	13.1
Pulpwood @ final harvest	100.3	99.8	101.7	103.2
Sawlogs @ final harvest	75.1	74.7	76.1	77.2
Total	**175.4**	**187.5**	**190.9**	**193.4**

A key challenge to implementing the proposed biomass supply strategy is convincing NIPF owners to intensify their management practices. A strong case can be made that investments in management improvements, including thinning, will provide good financial returns,[19] regardless of whether the thinnings are sold as LBJ feedstock or as pulpwood. This should facilitate the process of contracting land owners to intensify management. Given the relatively small amount of biomass needed annually for the LBJ plant (115,000 t_{dry}/yr) and the significant acreage of planted NIPF stands in Mississippi and Alabama, we assume that a sufficient number of NIPF owners could be contracted to meet the LBJ biomass supply requirements. The maximum forested area that would be needed to supply the LBJ plant with thinnings over a 20-year LBJ plant life is about 200,000 hectares, considering that any single stand can be thinned only once during the LBJ plant lifetime and that the average expected yield at thinning is 10.9 t/ha (Table 9).

4.2.2 Calculating net biogenic GHG emissions

The EPA's proposed methodology for estimating biogenic GHG emissions[44] was developed specifically for estimating biogenic emissions from a stationary source, such as a biomass power plant. We adapt the EPA approach to estimate net emissions of biogenic origin for the LBJ system.

The EPA methodology involves calculating net biogenic emissions of CO_{2e} (**NBE**) as the difference between system-wide emissions for a scenario in which a conversion facility is built and the scenario in which it is not built (BAU). In the EPA methodology:

$$NBE \equiv BAF \cdot PGE \qquad \text{Eqn. (1)}$$

where **BAF** is a dimensionless biogenic assessment factor, and **PGE** is the potential gross biogenic CO_{2e} emissions. **PGE** is the CO_{2e} content of the biomass at some specified accounting reference point along the bio-carbon's path from production to plant exhaust stack. In our work, we consider the LBJ plant gasifier inlet as the reference point. In the EPA formulation:

$$BAF = [GROW + AVOIDEMIT + SITENC + LEAK] \cdot L \cdot P \qquad \text{Eqn. (2)}$$

where the sum in brackets, EPA's so-called "landscape emissions effect", represents emissions associated with the stand from which the biomass feedstock is supplied. **GROW** accounts for the net growth and harvesting of biomass relative to the BAU scenario. **AVOIDEMIT** accounts for biogenic emissions avoided compared to the BAU scenario (e.g., avoided decomposition or burning of harvest residues). **SITENC** accounts for the net change in total non-feedstock carbon pools (e.g., small branches and roots) relative to the BAU scenario. **LEAK** accounts for indirect impacts, i.e., changes in net emissions due to changes outside the feedstock production area. For example, if the LBJ plant uses biomass that would otherwise have gone to another user in the BAU scenario, the other user would seek an equivalent amount of biomass from elsewhere, and LEAK would capture these impacts.

The parameter **L** in Eqn. (2) is a unit-less adjustment factor (greater than one) that accounts for the feedstock carbon lost between the feedstock production site and the **PGE** accounting reference point. In our case, since **PGE** is defined at the LBJ gasifier inlet, **L** adjusts for losses after harvest and prior to arrival at the gasifier, e.g., biomass lost in transit, by decomposition during storage, by debarking, and by sizing and drying. Thus, **PGE*L** is the carbon content of the biomass grown at the feedstock production site in order to deliver **PGE** of feedstock to the LBJ gasifier. For the LBJ analysis, we assume that 10% of the biomass harvested for the LBJ plant is lost before arriving at the gasifier, i.e., **L** = **PGE*L / PGE** = [1 / (1 – 0.1)] = 1.11.

In the EPA methodology the parameter **P** is a unit-less adjustment factor defined as the fraction of **PGE** carbon that is emitted from the exhaust stack of the plant. EPA includes the **P** term because their methodology was developed specifically for estimating biogenic emissions from the exhaust stack of a stationary bioenergy conversion facility. Exported co-products with embodied carbon do not contribute to emissions from the conversion plant, and **P** excludes such carbon from the biogenic emissions assigned to the conversion facility.

We have chosen to separate the analysis of biogenic carbon accounting for the LBJ system into two steps. In the first step we follow the EPA methodology to estimate what impact landscape (upstream) factors have on net biogenic emissions. In the second step we incorporate the impact of of how the biomass is processed after arriving at the gasifier.

For the first step, we set **P** in Eqn. (2) equal to 1, so that the calculated BAF reflects only the upstream landscape emissions effect, independent of what happens to the biomass after it enters the gasifier. To clarify that our first step considers only upstream factors, we designate the resulting BAF as the "upstream" BAF:

$$UBAF = [GROW + AVOIDEMIT + SITENC + LEAK] \cdot L \qquad \text{Eqn. (3)}$$

The EPA[44] specifies methodologies for calculating each of the terms in the brackets in Eqn. (3) except **LEAK**, for which there is a lack of scientific consensus on appropriate methodology. For the LBJ analysis, following Rooney and Gray,[16] we assume **LEAK** is negligible primarily because the LBJ biomass feedstock supply strategy includes land management changes to increase per-hectare yields of biomass relative to the BAU scenario.[h] According to the EPA methodology and when **LEAK** is neglected, Eqn. (3) can be expanded to:

$$UBAF = \left[\frac{(GROW_{LBJ}^{-} - GROW_{BAU}^{-}) - (GROW_{LBJ}^{+} - GROW_{BAU}^{+})}{(GROW_{LBJ}^{-} - GROW_{BAU}^{-})} + \frac{(DCOMP_{LBJ} - DCOMP_{BAU})}{(GROW_{LBJ}^{-} - GROW_{BAU}^{-})} + \frac{(SITENC_{BAU} - SITENC_{LBJ})}{(GROW_{LBJ}^{-} - GROW_{BAU}^{-})}\right] \cdot L$$

Eqn. (4)

where the subscript LBJ refers to the scenario in which biomass is used at the LBJ plant, the subscript BAU refers to the BAU scenario, and the numerator and denominator of each of the three fractions inside the square brackets have units of tCO_{2e}/(ha-yr) at the biomass production site. In Eqn. (4):

$\boldsymbol{GROW_{XXX}^{-}}$ = Biomass feedstock harvested for energy during the 25-year stand rotation. (Note that $GROW_{BAU}^{-} = 0$),

$\boldsymbol{GROW_{XXX}^{+}}$ = Total net aboveground biomass growth during the 25-yr stand rotation (i.e., growth, including thinnings that are removed, minus losses due to tree mortalities and fallen limbs and foliage),

DCOMP = Emissions of CO_{2e} resulting from above-ground decomposition of dead biomass, including dead trees, fallen branches and foliage, and harvest residues left behind,

SITENC = Below-ground accumulations of CO_{2e} including in the soil litter layer, coarse and fine roots, and soil carbon.

The complexity/completeness of the EPA formalism is somewhat forbidding, but as will be detailed below, the dominant term in **UBAF** [Eqn. (4)] for the biomass supply scenarios we are considering is the one involving $\boldsymbol{GROW^{-}}$ and $\boldsymbol{GROW^{+}}$. To help understand more intuitively the meaning of **UBAF**, if we ignore for the moment the other terms inside the square brackets in Eqn. (4), we can write it in plain-language terms:

$$UBAF = \frac{(Harvested - Extra\ Grown)}{Harvested} \cdot \frac{Harvested}{Used}$$

where *Harvested* is the biomass (or really the equivalent amount of CO_{2e}) harvested at the growing site, *Extra Grown* is the additional biomass grown relative to the BAU scenario, and *Used* is the biomass that is actually used in the gasifier at the conversion site. Then, based on Eqn. (1),

$$Net\ Biogenic\ Emissions = UBAF \cdot Used = Harvested - Extra\ Grown$$

From this simplified representation, we can readily understand the meaning of different values of **UBAF**. For example, a **UBAF** of -½ means that, as a result of improved management of a stand of trees, the extra biomass that grows (relative to the BAU scenario, where management is not improved) is 50% larger than the amount harvested for energy, and thus for every molecule of biogenic CO_2 emitted to the atmosphere in using the harvested biomass, 1.5 molecules were

---

[h] Leakage may, in fact, be negative, since the final harvest (excluding thinnings) of pulpwood and sawlogs is *higher* in the Low, Med, or High scenarios than in the BAU scenario (Table 9).

absorbed from the atmosphere by photosynthesis. A **UBAF** of -1 means that the extra growth is 100% larger than the harvest, and thus two molecules of CO_2 were absorbed from the atmosphere for each one harvested and emitted. A **UBAF** of zero indicates that all extra biomass (above BAU) was harvested, and its conversion to atmospheric CO_2 would have no net climate impact; this is the classic "carbon-neutral" scenario, where additional biomass is grown and it's CO_{2e} content returns to the atmosphere when it is burned for energy. Positive **UBAF** values indicate that the amount of harvested biomass exceeds that grown additionally for the project, and thus its conversion to atmospheric CO_2 increases the level of the latter above BAU. A **UBAF** of +½ would result when the amount of harvested biomass is twice as large as the extra growth, and a **UBAF** of +1 results when no change is made to increase stand productivity relative to BAU. (Note that a **UBAF of +**1 means the impact of biomass use would be equivalent to using fossil fuels instead.)

UBAF determines the landscape's influence on *potential* net biogenic emissions. We now move on to incorporate into the biogenic emissions accounting how the biomass arriving at the plant is processed. In the case of the LBJ plant, some of the biogenic carbon arriving at the gasifier leaves embedded in SPK and naphtha products, some is captured as CO_2 for geologic storage via EOR, some is vented from the plant as CO_2, and some leaves as solid waste, embedded in gasifier ash. The captured CO_2 and the carbon in solid waste are assumed to be permanently sequestered from the atmosphere. All other carbon arriving at the gasifier will ultimately return to the atmosphere when the products are used for energy. Thus, the net biogenic emissions for the LBJ system as a whole can be calculated as the landscape emissions adjusted for carbon at the plant that is sequestered from the atmosphere

$$NBE_{LBJ} = (UBAF \cdot PGE) - (PGE \cdot F) = PGE \cdot [UBAF - F] \qquad \text{Eqn. (5)}$$

where **F** is the fraction of the biomass carbon delivered to the gasifier (**PGE**) that is permanently sequestered from the atmosphere, as determined from the process simulation results (Table 6). The first term in Eqn. 5 quantifies the potential upstream net biogenic emissions (if all harvested biomass is converted to CO_2 and released to the atmosphere); the second term quantifies the mitigating effects of capturing a fraction, **F,** of the biocarbon entering the plant and permanently sequestering it from the atmosphere.[i]

The terms in the **UBAF** equation are calculated based on the cumulative carbon flows over a 45-year period that begins 15 years prior to the start of the LBJ plant and extends through the end of

---

[i] As applied to the LBJ system, Eqn. (5) gives a different numerical result than EPA's formulation for **NBE** [Eqn. (1) and (2)], and we believe Eqn. (5) gives the correct result. That the two formulations will give different results can be readily seen by considering a simplified scenario of a biomass power plant with CCS, where **F** is the fraction of the input biomass carbon captured and sequestered from the atmosphere. The balance of the input carbon is vented from the stack. If we simplify the EPA formulation of **BAF** in Eqn. (2) by assuming **GROW** as the only non-zero term inside the brackets and that **L** = 1, then $\boldsymbol{BAF = GROW \cdot P}$. Substituting this into Eqn. (1) gives $\boldsymbol{NBE = PGE \cdot GROW \cdot P}$. Since **P** is the fraction of **PGE** vented at the plant, we can write **P** = (1 – **F**). Thus,

$$\boldsymbol{NBE_{EPA} = PGE \cdot GROW} \cdot (1 - \boldsymbol{F}) = \boldsymbol{PGE \cdot GROW - PGE \cdot GROW \cdot F}.$$

With the same assumptions for **GROW** and **L**, Eqn. (5) gives

$$\boldsymbol{NBE_{LBJ} = PGE} \cdot [\boldsymbol{GROW - F}] = \boldsymbol{PGE \cdot GROW - PGE \cdot F}.$$

In the EPA formulation, **F**, or equivalently (1 – **P**), effectively scales the landscape effect (**GROW**), even though the landscape effect should be independent of how much carbon is captured and sequestered at the power plant. We believe the EPA formulation yields an incorrect result in this case, whereas the LBJ formulation treats the landscape effect separately and gives a correct result.

the growth rotation for all lands thinned to provide LBJ feedstock (Figure 16). The analysis timeframe extends ten years beyond the 20-year operating life of the plant to cover the full rotation for all forests that provided feedstock to the LBJ plant.

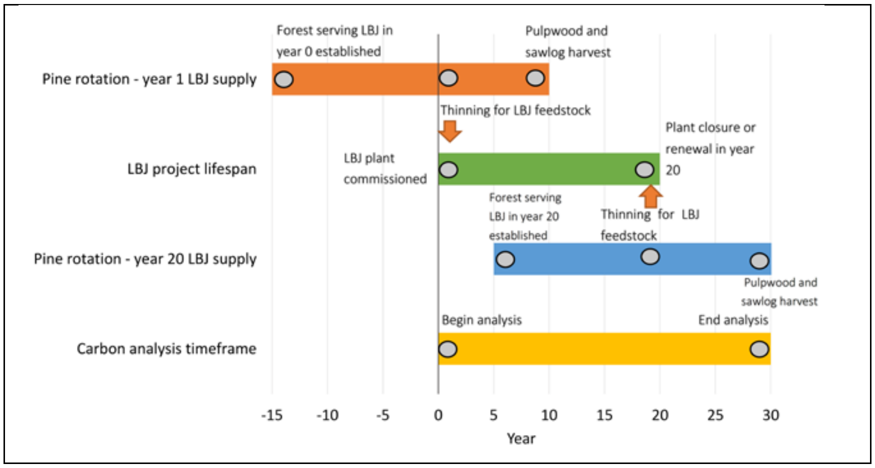


Figure 16. Timeline for estimating biogenic carbon emissions.

Table 10 details the calculation of **UBAF** values. The two **GROW** terms in Eqn. (4) are calculated from the total live above-ground biomass values in Table 9. The **DCOMP** and **SITENC** terms are estimated (using allometric relations) from the merchantable amounts of pulpwood and sawlogs. The LBJ scenarios are characterized by greater decomposition emissions because thinning operations leave tops and branches on the ground, and the larger merchantable tonnage at final harvest (Table 9) generates more harvest residues. The higher decomposition is offset to a large degree by greater in-soil storage of carbon (**SITENC**) in the LBJ scenarios. As a result, the **GROW** terms in Eqn. (4) largely determine the value of **UBAF**.

Because fertilization used in the LBJ scenarios will result in some emissions of N_2O, the **UBAF** values calculated using the methodology described above are adjusted to account for N_2O emissions. The N_2O adjustment assumes a standard fertilizer application rate of 224 kg N per hectare. It assumes that 1.17% of the N is released to the atmosphere as N_2O[47] and that the global warming potential of N_2O is 310 times that of CO_2 on a mass basis.[48]

As shown in Table 10, the **UBAF** values with N_2O adjustments are negative for all three LBJ scenarios, meaning that over the course of the analysis period, the increase in biomass production (compared to the BAU scenario) from the forest stands that supply the LBJ feedstocks is greater than the CO_{2e} content of the biomass harvested for the LBJ project. In effect, there is removal of CO_2 from the atmosphere, net of the BAU scenario, under each of the three LBJ scenarios, even if all of the carbon in the delivered biomass were to be released back to the atmosphere. In fact, since some of the carbon delivered as biomass to the gasifier (**PGE** = 622 tCO_{2e}/day, Table 6) is captured and stored (**F** = [1,326 + 57] / 2,382 = 0.58, based on Table 6), net biogenic emissions are more negative still. The **NBE**$_{\mathbf{LBJ}}$ values in Table 10, calculated using Eqn. (5), take into account the capture and storage of some of the biomass feedstock carbon.

The **UBAF** and **NBE** values in Table 10 were developed for biomass supplies originating from planted forests in Mississippi and Alabama (Region 2), as appropriate for the location of the LBJ demonstration project. Antares also developed estimates for other regions (Table 11). The wide variations in **UBAF** (and corresponding **NBE**$_{\mathbf{LBJ}}$) across regions are due to intrinsic biogeophysical and climate differences. **UBAF** values are negative for every region. Regions 3 and 4 have especially large negative values.

Table 10. Calculation of biomass accounting factor (BAF) and net biogenic emission (NBE) for the LBJ system for biomass production scenarios in Table 9.

	BAU	Low	Medium	High
Variables in Eqn. (3)	**metric tCO_{2e}/hectare**			
GROW+	302	323	329	333
GROW-	-	18.8	18.8	18.8
DCOMP[a]	21.0	25.2	25.7	26.0
SITENC[b]	5.6	10.1	10.4	10.5
UBAF terms in Eqn. (2)	**Dimensionless**			
GROW		- 0.11	- 0.42	- 0.65
AVOIDEMIT		0.23	0.25	0.27
SITENC		- 0.24	- 0.25	- 0.26
UBAF (base)		**- 0.14**	**- 0.46**	**- 0.71**
N_2O adjustment		0.06	0.05	0.04
UBAF with N_2O adjustment		**- 0.07**	**- 0.41**	**- 0.67**
Net Biogenic Emissions (NBE_{LBJ})[c]				
tCO_{2e}/day @ 100% CF		- 406	- 615	- 778
$kgCO_{2e}/GJ_{SPK,LHV}$		- 193	- 292	- 369

(a) Estimated using allometric equations and forest inventory data.[16] Accounts for emissions from decomposition of aboveground coarse woody debris (CWD: litter fall, tops and branches generated by natural breakage and tops and branches generated during harvesting operations).

(b) Estimated using published factors for decomposition rates.[16] Accounts for accumulation of carbon and emissions from below-ground biomass growth and CWD that became part of the soil.

(c) Based on Eqn. (4), using UBAF with N_2O adjustment. These are the total biogenic emissions for the LBJ system, including emissions at the plant and during combustion of the liquid products.

Table 11. Estimated BAF and NBE_{LBJ} for each region and weighted average for all regions combined.

Scenario		Region[a]				South
		1	2	3	4	average
	UBAF w/N_2O adjustment	-0.48	-0.07	-0.85	-1.17	-0.57
	NBE_{LBJ}, $kgCO_{2e}/GJ_{SPK,LHV}$	-313	-193	-423	-517	-340
	UBAF w/N_2O adjustment	-0.53	-0.41	-1.59	-2.43	-1.06
	NBE_{LBJ}, $kgCO_{2e}/GJ_{SPK,LHV}$	-328	-292	-641	-890	-485
	UBAF w/N_2O adjustment	-0.56	-0.67	-2.17	-3.39	-1.43
	NBE_{LBJ}, $kgCO_{2e}/GJ_{SPK,LHV}$	-337	-369	-813	-1173	-594

(a) Region 2 covers Mississippi and Alabama. See Figure 15 for states included in each region.

4.3 LBJ Greenhouse Gas Emissions Results

Using inputs from Table 7 and Table 10, Figure 17 quantifies the individual contributions to net LBJ system GHG emissions for the three biomass supply scenarios for the Mississippi/Alabama region (Region 2) described in Section 4.2. Net emissions (black diamonds) are positive for each scenario, but decrease from the low to medium to high scenarios. Individual contributions that are positive are the same for each scenario. The net landscape emissions (based on UBAF) vary with the plantation management scenario and are negative in all cases, as expected based on the discussion in Section 4.2.2.

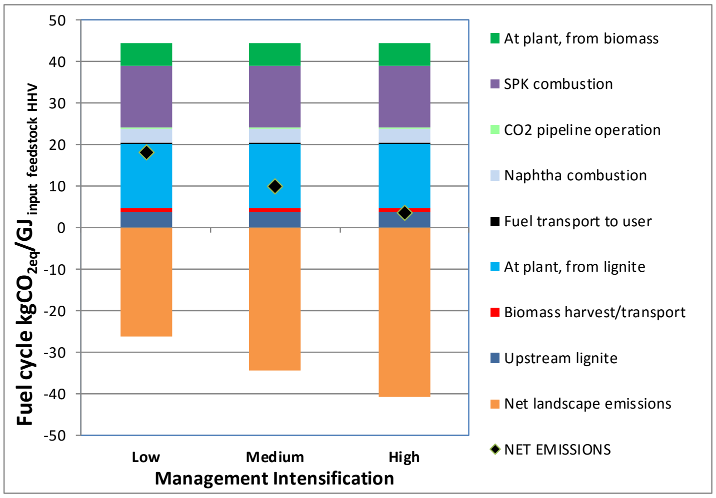


Figure 17. Estimated LBJ system lifecycle GHG emissions per unit of biomass + lignite energy delivered to the LBJ gasifier for the three biomass supply scenarios for Region 2 (Mississippi/Alabama) described in Section 4.2.

Figure 18 shows estimates of $GHGI_{2005}$ for the FOAK plant. For all three pine plantation management-intensity scenarios, $GHGI_{2005}$ is well-below one, indicating LBJ system emissions far below those for the 2005 reference system in all cases.

The values of the second GHG metric, $kgCO_{2e}/GJ_{SPK,LHV}$ are shown in Figure 19 and mirror the $GHGI_{2005}$ values. For the Low scenario, the net SPK emissions are about one-quarter of those of conventional jet fuel, and thus easily meets one of would key project goals, which was to design a facility producing jet fuel with GHG emissions less than those for conventional jet fuel. Net SPK emissions are negative for the medium and high scenarios. The Low scenario emissions estimate is adopted for the financial analysis of the LBJ demonstration project (Section 5.3).

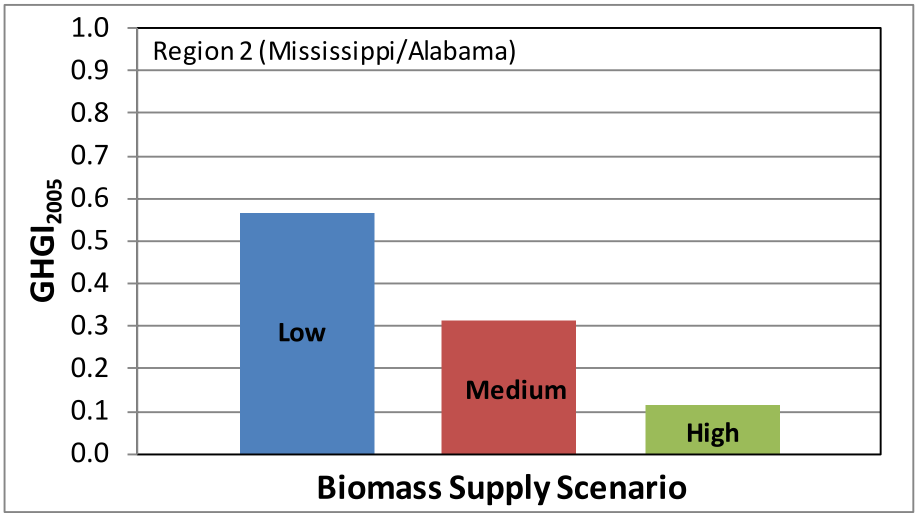


Figure 18. $GHGI_{2005}$ for the LBJ system for the three biomass supply scenarios for the Mississippi/Alabama region described in Section 4.2

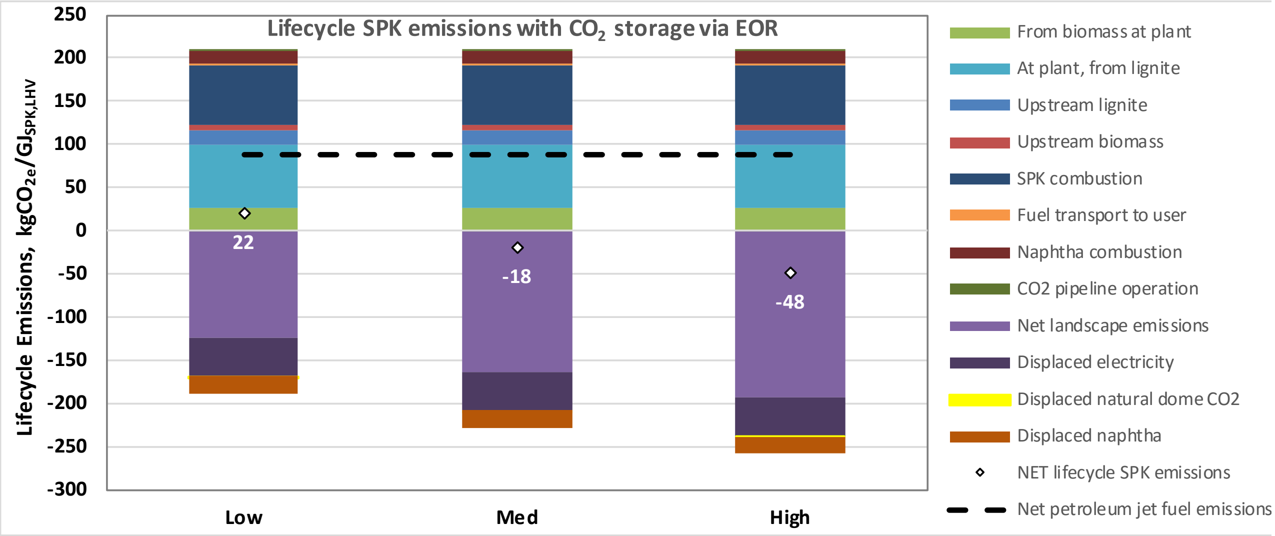


Figure 19. Net lifecycle emissions (white diamonds and superimposed white numbers) and component emissions associated with SPK production by the LBJ system for Low, Medium, and High pine plantation management-intensity scenarios for the Mississippi/Alabama region described in Section 4.2. Net lifecycle emissions for conventional jet fuel (from Table 7) are shown for comparison.

5 FOAK Plant Costs and Financial Viability

WorleyParsons Group, Inc. (WP) was contracted to prepare design documentation and develop capital and operating cost estimates for the LBJ plant. The WP report is available as Appendix A. The funding available for the project as a whole allowed for the design and definition of the project to be advanced to about 1% of completed project engineering. Additionally, while most of the plant components are based on proven technologies, the plant as a whole represents a first-of-a-kind (FOAK) integration of these components. Given these considerations, there is inherent, but difficult to quantify, scope risk and uncertainty in the cost estimates prepared by WP. To help assess the impact of uncertainty, the financial performance of the project (as described in Section 5.3) includes analyses of the impact of a range in final as-built cost scenarios.

WP generated a heat and material balance for the entire LBJ facility to the level of detail necessary to support their cost estimate (see Appendix A). Equipment sizes and other relevant parameters for cost estimation were based on in-house process simulations or project-specific equipment performance quotations. Budget vendor estimates were received for the following major subsystems and components:

- Air separation unit (Air Products) – see Appendix E,
- Biomass preparation & drying (BRUKS Rockwood),
- Lignite drying (ANDRITZ),
- Rectisol AGR unit (Air Liquide) – see Appendix B,
- Pressure swing adsorption unit (Air Liquide),

- CO_2 compressor (Dresser-Rand) – see Appendix F,
- FT synthesis and upgrading units (Emerging Fuels Technology) – see Appendix C,
- Wet sulfuric acid unit (Haldor Topsoe),
- Gas turbine (Siemens SGT-700),
- HRSC steam turbine (Siemens),
- ZLD system (Suez).

Non-confidential elements of the vendor quotations are included in the indicated appendices.

5.1 Capital Cost Estimates

WP prepared an estimate of the FOAK Total Plant Cost (TPC), overnight basis, by estimating the Bare Erected Cost (BEC) and applying their experience-based factors to estimate EPCM (Engineering, Procurement and Construction Management) services and contingencies. The BEC includes estimated costs for equipment, construction materials, and the associated direct and indirect installation labor. To estimate installation labor, WP considered current merit-shop rates for Mississippi with a competitive bidding environment, with adequate skilled craft labor available locally. WP includes contingencies to account for costs that cannot be explicitly determined due to a lack of project definition and engineering, but which experience has shown will be incurred by the end of the project. WP excluded owner's costs from its estimate.

The EPCM estimate of 20% of BEC is a consensus value arrived at by WP in discussion with Princeton and Queensland University researchers and reflects the complexity and FOAK nature of the LBJ project. WP applied a process contingency of 15% of BEC to compensate for uncertainty in cost estimates caused by performance and technology integration uncertainties: the plant represents a collection of mature, or at least commercially demonstrated, technologies and a FOAK integration of these. WP applied a project contingency of 25% to the sum of BEC, EPCM, and process contingency based on the current level of project definition and the factored approach used to develop the significant parts of the estimate. Additional details of WP's capital cost estimate include:

- Costs given in U.S. dollars and represent "overnight" costs for mid-2015, with no forward escalation to period-of-performance;
- An estimate boundary limit of the facility's "fence line," including lignite and wood receiving and water supply systems and transmission and utility connections that lie within the fence line, such as CO_2 pipeline;
- No new switchyard or switchyard modifications: the existing switchyard at Plant Sweatt is assumed to be re-used;
- The assumption that the project site will be furnished in a clean, level condition;
- Costs grouped according to a system-oriented code of accounts: all reasonably allocable components of a system or process are included in the specific system account in contrast to a facility, area, or commodity account structure.

WP adjusted, as required, all of the vendor cost quotations mentioned above to include freight to site, vendor technical direction during installation, incomplete or missing scope items, and/or changes in capacity (from the initial design specifications given the vendors to the final study plant design), as well as conversion to U.S. dollars. WP escalated and scaled costs for lignite receiving and unloading from a quotation for a similar facility. WP initially developed costs for the TRIG gasifier, coal milling/drying/handling, and ash removal by escalating and scaling a high-level budgetary quotation from a previous project. The initial estimates were then reviewed with Southern Company and adjusted based on their recommendations. The balance of the cost accounts were estimated by WP using their in-house database and conceptual estimating model. This database and model are maintained by WP as part of a commercial power plant design base of experience for similar equipment, materials, and construction labor requirements in the company's range of power and process projects.

WP estimates a BEC (BEC_{WP}) of $588 million and a total plant cost of $992 million for the LBJ plant, excluding owner's costs (Table 12). Three systems account for 57% of the BEC: the gasifier island (including ASU), the gas cleanup system (including Rectisol), and the Fischer-Tropsch island (including synthesis and upgrading).

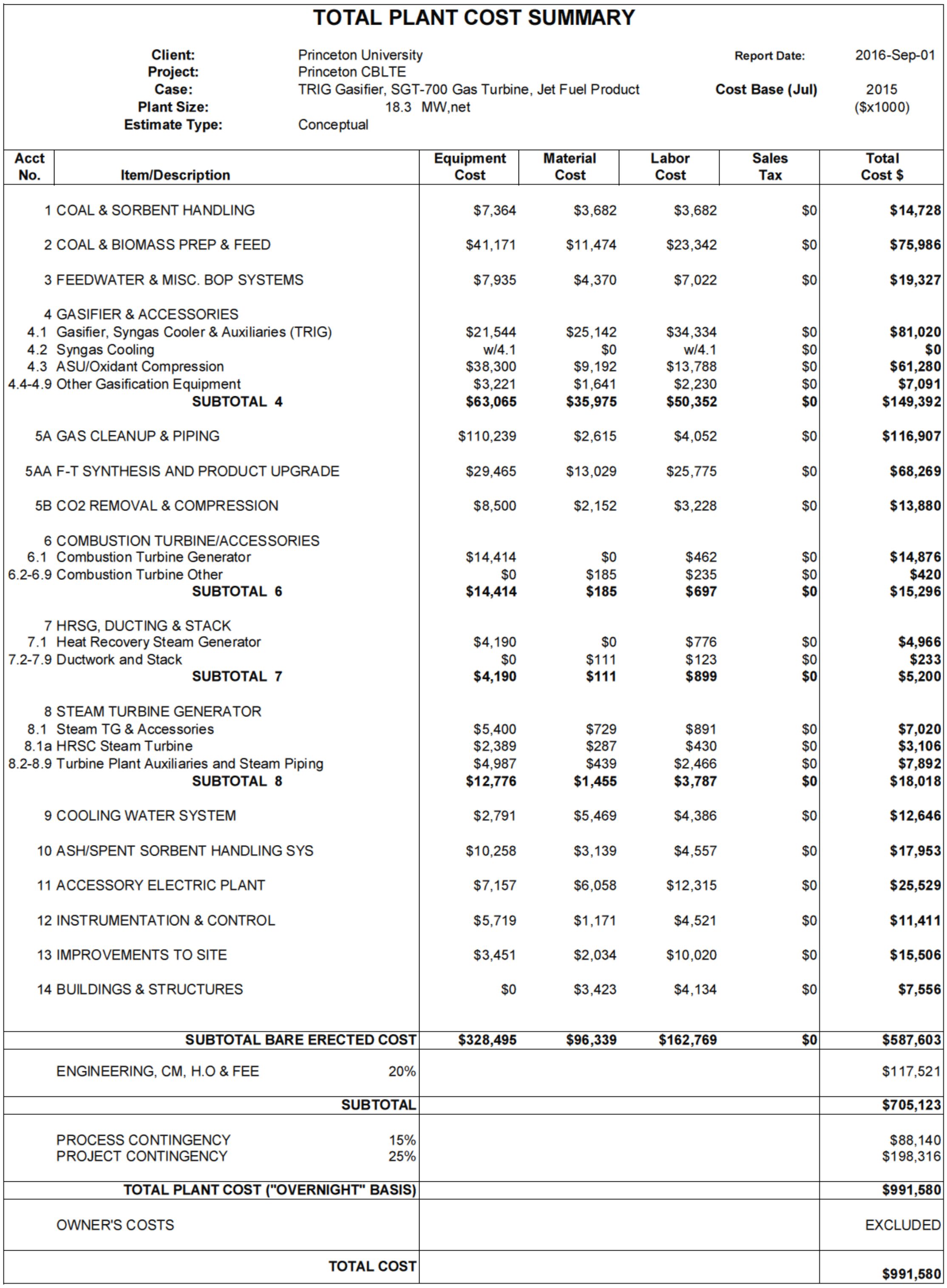
The WP-based total plant cost (TPC_{WP} = $992 million) is well below the $2 billion limit specified in the guiding principles for project design (in Section 2). However, we consider the WP estimate to be a lower-bound estimate, because historical experience,[49,50] and evidence from recent pioneer projects, including coal-IGCC projects at Edwardsport, IN, and Kemper County, MS, and a post-combustion CO_2 capture retrofit at the Boundary Dam pulverized coal plant in Saskatchewan, Canada, support the use of higher contingency allowances than specified by WP for a FOAK project at the early stage of project definition that characterizes the LBJ project. Capital cost estimates for the Edwardsport, Kemper, and Boundary Dam projects based on early conceptual or scoping-level studies were far below the capital actually expended to complete those projects.

Greig, *et al*.,[50] based on their detailed study of estimated versus actual project capital costs, have recommended the guidelines shown in Table 13 for establishing contingency (process and project) allowances. Their recommended values vary with how much engineering has been completed toward development of the project and with the level of commercial maturity of the technologies and processes involved. The values in the table are percentages of an overnight cost (TPC) estimate assumed to have been arrived at using a "conventional" approach to cost estimation. Because different practitioners may define the extent of project definition differently and/or assign different levels of contingency allowances in a "conventional" TPC estimate for the same level of project definition, the values in Table 13 suggested for process and project contingency allowances are guidelines only.

For especially large and complex megaprojects, especially in early stages where project definition is limited, where there may be limited knowledge of site conditions and available services and where the lead time to financial investment decision (FID) is long (e.g. due to resource appraisal durations) with consequent uncertainty in both local and macroeconomic conditions, an additional supplementary funds (SF) allowance might be appropriate to include in the estimate of capital requirements.

The SF allowance is to cover risks of unforeseen events or of outcomes that add to the most-likely cost outcome. The SF may not need to be spent, but technology performance, gross scope growth arising from local or market issues, regulatory changes, process integration challenges, slow plant ramp-up to full capacity, redesign and rectification requirements, or other issues may arise, and the project would be unable to proceed if SF are not available to address such issues. Table 13 provides guidelines for SF allowances.

Table 12. WorleyParsons capital cost estimate summary for the LBJ FOAK demonstration plant. See Appendix A for a more detailed breakdown of this estimate.



TOTAL PLANT COST SUMMARY

Client: Princeton University
Project: Princeton CBLTE
Case: TRIG Gasifier, SGT-700 Gas Turbine, Jet Fuel Product
Plant Size: 18.3 MW,net
Estimate Type: Conceptual
Report Date: 2016-Sep-01
Cost Base (Jul) 2015
($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Cost	Sales Tax	Total Cost $
1	COAL & SORBENT HANDLING	$7,364	$3,682	$3,682	$0	**$14,728**
2	COAL & BIOMASS PREP & FEED	$41,171	$11,474	$23,342	$0	**$75,986**
3	FEEDWATER & MISC. BOP SYSTEMS	$7,935	$4,370	$7,022	$0	**$19,327**
4	GASIFIER & ACCESSORIES					
4.1	Gasifier, Syngas Cooler & Auxiliaries (TRIG)	$21,544	$25,142	$34,334	$0	**$81,020**
4.2	Syngas Cooling	w/4.1	$0	w/4.1	$0	**$0**
4.3	ASU/Oxidant Compression	$38,300	$9,192	$13,788	$0	**$61,280**
4.4-4.9	Other Gasification Equipment	$3,221	$1,641	$2,230	$0	**$7,091**
	SUBTOTAL 4	**$63,065**	**$35,975**	**$50,352**	**$0**	**$149,392**
5A	GAS CLEANUP & PIPING	$110,239	$2,615	$4,052	$0	**$116,907**
5AA	F-T SYNTHESIS AND PRODUCT UPGRADE	$29,465	$13,029	$25,775	$0	**$68,269**
5B	CO2 REMOVAL & COMPRESSION	$8,500	$2,152	$3,228	$0	**$13,880**
6	COMBUSTION TURBINE/ACCESSORIES					
6.1	Combustion Turbine Generator	$14,414	$0	$462	$0	**$14,876**
6.2-6.9	Combustion Turbine Other	$0	$185	$235	$0	**$420**
	SUBTOTAL 6	**$14,414**	**$185**	**$697**	**$0**	**$15,296**
7	HRSG, DUCTING & STACK					
7.1	Heat Recovery Steam Generator	$4,190	$0	$776	$0	**$4,966**
7.2-7.9	Ductwork and Stack	$0	$111	$123	$0	**$233**
	SUBTOTAL 7	**$4,190**	**$111**	**$899**	**$0**	**$5,200**
8	STEAM TURBINE GENERATOR					
8.1	Steam TG & Accessories	$5,400	$729	$891	$0	**$7,020**
8.1a	HRSC Steam Turbine	$2,389	$287	$430	$0	**$3,106**
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$4,987	$439	$2,466	$0	**$7,892**
	SUBTOTAL 8	**$12,776**	**$1,455**	**$3,787**	**$0**	**$18,018**
9	COOLING WATER SYSTEM	$2,791	$5,469	$4,386	$0	**$12,646**
10	ASH/SPENT SORBENT HANDLING SYS	$10,258	$3,139	$4,557	$0	**$17,953**
11	ACCESSORY ELECTRIC PLANT	$7,157	$6,058	$12,315	$0	**$25,529**
12	INSTRUMENTATION & CONTROL	$5,719	$1,171	$4,521	$0	**$11,411**
13	IMPROVEMENTS TO SITE	$3,451	$2,034	$10,020	$0	**$15,506**
14	BUILDINGS & STRUCTURES	$0	$3,423	$4,134	$0	**$7,556**
	SUBTOTAL BARE ERECTED COST	**$328,495**	**$96,339**	**$162,769**	**$0**	**$587,603**
	ENGINEERING, CM, H.O & FEE 20%					$117,521
	SUBTOTAL					**$705,123**
	PROCESS CONTINGENCY 15%					$88,140
	PROJECT CONTINGENCY 25%					$198,316
	TOTAL PLANT COST ("OVERNIGHT" BASIS)					**$991,580**
	OWNER'S COSTS					EXCLUDED
	TOTAL COST					**$991,580**

For purposes of using the Table 13 guidelines, we judge the level of project definition in the current study to be consistent with a scoping-level of effort, and the level of commercial maturity to be cautiously considered “first-of-a-kind”, considering that key areas such as the gasifier (especially with lignite/biomass co-feed) and FT synthesis using this type of syngas, along with overall system integration, are clearly FOAK. An off-setting consideration is the focus on a specific site for which a good deal is known about conditions, services and infrastructure. Also, we note that the contingency allowances used by WP in their TPC estimate in Table 12 took some account of the novelty of the process integration required with this project. Additionally, we recognize that the physical scale of the LBJ project is considerably smaller than the scale of most of the projects reviewed by Greig, *et al*.[50] to arrive at the recommendations in Table 13 and thus arguably lower contingency allowances are warranted.

Considering these factors, we have estimated a "most-likely" TPC level for the LBJ project by adopting a process contingency allowance equal to 35% of the BEC_{WP} shown in Table 12. Projects using very mature technologies and processes, such as pulverized coal power plants, would attract no process contingency. In adopting a 35% process contingency, we note that coal or biomass gasification integrated with combined cycles number less than a dozen globally and none are yet operating integrated with FT liquids production or CO_2 capture.

Table 13. Guidelines recommended for contingency allowances in estimating capital costs for energy megaprojects.[50] All figures are percentages of a "conventional" TPC estimate for the given level of technology maturity and stage of project development.

	Phase of Project Development				
	Concept study	**Scoping study**	**Pre-feasibility**	**Feasibility study**	**Engineering Completed**
First-of-a-kind					
Process contingency	50	35	20	15	10
Project contingency	50	35	20	15	10
Total contingencies	100	70	40	30	20
Supplementary funds	50	40	30	20	10
Early mover					
Process contingency	25	15	10	5	5
Project contingency	40	30	20	15	10
Total contingencies	65	45	30	20	15
Supplementary funds	30	20	15	10	5
N^th^ of a kind					
Process contingency	0	0	0	0	0
Project contingency	30	25	20	10	5
Total contingencies	30	25	20	10	5
Supplementary funds	25	15	10	5	2.5

For the project contingency allowance, we adopted a value of 35% applied to the sum of all other project costs to cover: the likely growth in project scope with further engineering work,[j] allowances for cost changes reflecting the lack of specific information at this early stage on procurement strategies and market conditions, and contingency for events that are statistically likely

---

[j] The engineering work described in this report represents only about 1% of the total engineering costs that will be incurred prior to start of project construction.

to occur but are not included in the base estimate (to cover such items as changes to the execution plan, 'normal' allowances for weather problems, labor problems, etc.). For an EPCM allowance equal to 20% of BEC_{WP} plus the pair of 35% contingency allowances, our best estimate of TPC is

$$TPC_{LBJ} = BEC_{WP} \times (1 + 0.2 + 0.35) \times (1 + 0.35) = 2.1 \times BEC_{WP}.$$

Given the complexity of the LBJ plant, it would not be unreasonable to also consider an additional allowance for supplementary funds. The complexity is countered somewhat by the relatively modest scale of the plant, the high level of knowledge of site conditions and available services for a project at this stage. Based on Table 13, then, we consider a 20% SF allowance applied to TPC_{LBJ}, i.e.,

$$TPC_{LBJ,w/SF} = 2.1 \times BEC_{WP} \times (1 + 0.2) = 2.5 \times BEC_{WP}.$$

We examine in Section 5.3 the financial performance of the project for all three TPC estimates:

$TPC_{WP} = 1.7 \times BEC_{WP}$ = \$ 991,580,000

$TPC_{LBJ} = 2.1 \times BEC_{WP}$ = \$ 1,229,559,000

$TPC_{LBJ\ w/SF} = 2.5 \times BEC_{WP}$ = \$ 1,475,471,000

While these TPC estimates may seem inflated to those not accustomed to implementation of FOAK process projects, the range of TPC/BEC ratios indicated are broadly consistent with the findings of Rand Corporation[49] who analyzed some 44 pioneer-process plant projects undertaken in the U.S. by a variety of petrochemical and mineral processing companies during the 1970's. Based on their analysis, the authors established a model that estimates expected cost growth from initial cost estimates for FOAK projects by assigning weights to the following process/project characteristics:[51]

- Percent of the cost estimate that incorporates technology unproven in commercial use
- Judgment by industry process engineers of difficulties with process impurities encountered during process development
- Block count of all process steps in the plant
- Percentage of items checked on a checklist assessing completeness of estimates
- Levels of site-specific information and engineering included in the estimate
- Level of process maturity, ranging from still in R&D to proven at pre-commercial or commercial scale.

Princeton University of University of Queensland researchers applied their best judgement to the quantification of these process and project characteristics as they apply to the LBJ project. Based on this judgment, the Rand model would indicate even higher likely project costs than the highest TPC estimate described above ($TPC_{LBJw/SF}$).

Owner's costs are an additional capital cost considered in the discounted cash flow analysis in Section 5.3. The following owner's costs are included in that analysis, above and beyond the TPC: a front end engineering design study [\$20 million], owner's team to manage the primary EPCM and subsidiary contracts not included in the main EPCM contract, arrange permits, pay bills, etc. [\$11.8 million (2% of BEC)], initial fills [\$4.2 million, as estimated by WP in Table 14], land (\$0, assuming the Plant Sweatt site is already the owner's property), annual property tax and insurance [2% of TPC, as estimated by WP in Table 14]. Additional components of owner's costs are less definable at this early stage of project development and are assumed to be captured by the project contingency allowance included in the TPC.

5.2 Operating and Maintenance Cost Estimates

WorleyParsons estimated non-feedstock operating and maintenance costs in mid-2015 dollars on a basis consistent with their estimate of capital costs (Table 14). WP presented O&M costs on an average annual basis and excluded initial start-up costs. The O&M costs are split into fixed and variable. Fixed costs are independent of plant capacity factor. Variable costs are proportional to capacity factor, which the numbers in Table 14 assume is 80%. WP included annual property taxes and insurance (at 2% of TPC) as part of fixed operating costs.

The cost for operating labor was based on the estimated staffing required by equipment area. The corresponding hours were converted to equivalent around-the-clock (24/7) operating jobs. Maintenance costs were evaluated on the basis of experience-based ratios of maintenance cost to initial capital cost for similar equipment items and processes. Consumables (water, chemicals, catalysts) costs were determined on the basis of individual rates of consumption, the unit cost of each consumable, and the plant annual operating hours. The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis. Waste disposal costs were estimated on the basis of individual consumption / production rates, the unit costs for each item, and the plant annual operating hours.

5.3 FOAK Discounted Cash Flow Analysis Results

The capital and O&M cost estimates described in the previous section were incorporated in an annual discounted cash flow (DCF) analysis to assess the financial viability of the FOAK LBJ project. A baseline set of input assumptions for the DCF analysis are shown in Table 15, and Table 16 describes eight scenarios for which a DCF analysis was run. These include a scenario for each of the three TPC estimates described in Section 5.1 (TPC_{WP}, TPC_{LBJ}, and $TPC_{LBJ,\ w/SF}$), plus five scenarios that each use the mid-range TPC estimate (TPC_{LBJ}) and examine the sensitivity of the results to several parameter variations.

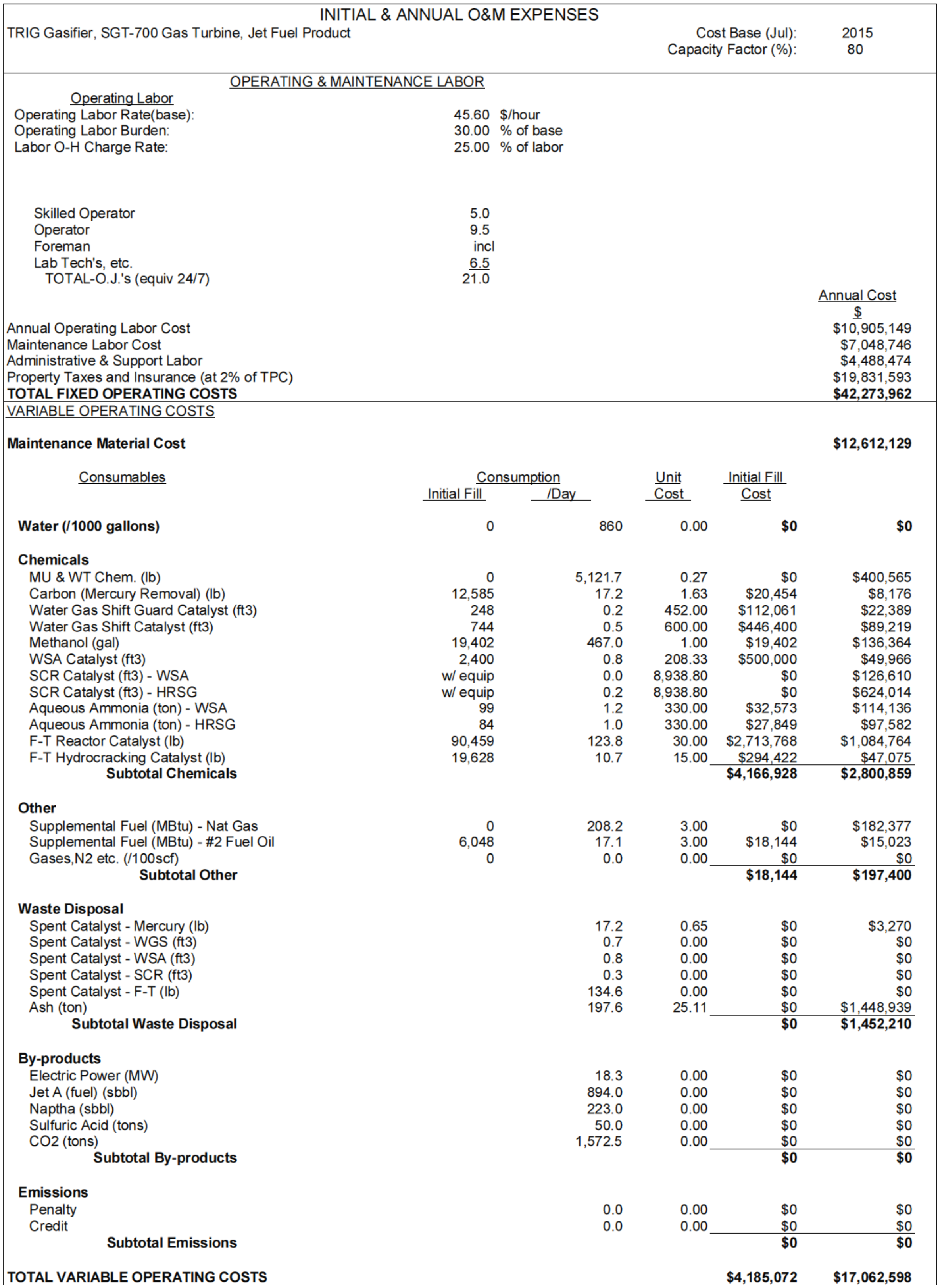
Because of the high capital and fixed O&M costs for this relatively small FOAK demonstration project it is not possible to generate positive cash flows from the anticipated revenues for the baseline set of assumptions with any of the TPC estimates in the absence of some form of additional financial support. With the baseline input assumptions and the mid-range TPC estimate (TPC_{LBJ}), the cost of SPK production approaches $500/bbl, dominated by capital and O&M charges. To achieve a zero net present value (NPV) under the baseline assumptions would require a levelized SPK production cost subsidy over the assumed 20-year life of the project of $388/bbl above and beyond the market revenues of $107/bbl (levelized) assumed to be received for the SPK. The required production subsidy to achieve zero NPV would be $306/bbl with the low-end TPC estimate (TPC_{WP}) and $472/bbl for $TPC_{LBJ,\ w/SF}$, the high-end estimate.

Alternatively, zero NPV could be achieved with a grant or other mechanism to reduce the capital cost to investors. The required capital cost grant would exceed the TPC with any of the three TPC estimates.

Figure 20 shows combinations of production cost subsidy and capital grant that would achieve zero NPV in each of the scenarios described in Table 16. The six scenarios involving TPC_{LBJ} show the sensitivity of the results to several parameter variations. Because of the high capital intensity of the plant, capital-related parameters [weighted average cost of capital (WACC) and plant availability ramp rate] have significant influences on the required subsidies, but the high oil-price assumption shows the largest impact. With an oil price of $120/bbl instead of the baseline $80/bbl, the required SPK production cost subsidy is reduced by about $70/bbl (for any assumed level of capital grant). Higher oil prices make the economics less unfavorable as a result of increased revenues for both SPK and EOR CO_2. Prices assumed for GHG emissions, exported electricity, and input feedstocks individually have relatively minor influences on the required level of financial support – see Table 16, note (a).

The poor FOAK-project economics are not surprising. They reflect the small scale of the plant, the guiding design principle to prioritize technical success over minimized cost, the levels of contingencies appropriate for the relatively early stage of project development, and the first-of-a-kind nature of the plant.

Table 14. WorleyParsons estimate of non-feedstock operating and maintenance costs. See Appendix A for additional discussion of O&M cost estimate.



INITIAL & ANNUAL O&M EXPENSES

TRIG Gasifier, SGT-700 Gas Turbine, Jet Fuel Product	Cost Base (Jul):	2015
	Capacity Factor (%):	80

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate(base):	45.60	$/hour
Operating Labor Burden:	30.00	% of base
Labor O-H Charge Rate:	25.00	% of labor

Skilled Operator	5.0
Operator	9.5
Foreman	incl
Lab Tech's, etc.	6.5
TOTAL-O.J.'s (equiv 24/7)	21.0

	Annual Cost $
Annual Operating Labor Cost	$10,905,149
Maintenance Labor Cost	$7,048,746
Administrative & Support Labor	$4,488,474
Property Taxes and Insurance (at 2% of TPC)	$19,831,593
TOTAL FIXED OPERATING COSTS	**$42,273,962**

VARIABLE OPERATING COSTS

Maintenance Material Cost					**$12,612,129**

Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost	
Water (/1000 gallons)	0	860	0.00	**$0**	**$0**
Chemicals					
MU & WT Chem. (lb)	0	5,121.7	0.27	$0	$400,565
Carbon (Mercury Removal) (lb)	12,585	17.2	1.63	$20,454	$8,176
Water Gas Shift Guard Catalyst (ft3)	248	0.2	452.00	$112,061	$22,389
Water Gas Shift Catalyst (ft3)	744	0.5	600.00	$446,400	$89,219
Methanol (gal)	19,402	467.0	1.00	$19,402	$136,364
WSA Catalyst (ft3)	2,400	0.8	208.33	$500,000	$49,966
SCR Catalyst (ft3) - WSA	w/ equip	0.0	8,938.80	$0	$126,610
SCR Catalyst (ft3) - HRSG	w/ equip	0.2	8,938.80	$0	$624,014
Aqueous Ammonia (ton) - WSA	99	1.2	330.00	$32,573	$114,136
Aqueous Ammonia (ton) - HRSG	84	1.0	330.00	$27,849	$97,582
F-T Reactor Catalyst (lb)	90,459	123.8	30.00	$2,713,768	$1,084,764
F-T Hydrocracking Catalyst (lb)	19,628	10.7	15.00	$294,422	$47,075
Subtotal Chemicals				**$4,166,928**	**$2,800,859**
Other					
Supplemental Fuel (MBtu) - Nat Gas	0	208.2	3.00	$0	$182,377
Supplemental Fuel (MBtu) - #2 Fuel Oil	6,048	17.1	3.00	$18,144	$15,023
Gases,N2 etc. (/100scf)	0	0.0	0.00	$0	$0
Subtotal Other				**$18,144**	**$197,400**
Waste Disposal					
Spent Catalyst - Mercury (lb)		17.2	0.65	$0	$3,270
Spent Catalyst - WGS (ft3)		0.7	0.00	$0	$0
Spent Catalyst - WSA (ft3)		0.8	0.00	$0	$0
Spent Catalyst - SCR (ft3)		0.3	0.00	$0	$0
Spent Catalyst - F-T (lb)		134.6	0.00	$0	$0
Ash (ton)		197.6	25.11	$0	$1,448,939
Subtotal Waste Disposal				**$0**	**$1,452,210**
By-products					
Electric Power (MW)		18.3	0.00	$0	$0
Jet A (fuel) (sbbl)		894.0	0.00	$0	$0
Naptha (sbbl)		223.0	0.00	$0	$0
Sulfuric Acid (tons)		50.0	0.00	$0	$0
CO2 (tons)		1,572.5	0.00	$0	$0
Subtotal By-products				**$0**	**$0**
Emissions					
Penalty		0.0	0.00	$0	$0
Credit		0.0	0.00	$0	$0
Subtotal Emissions				**$0**	**$0**
TOTAL VARIABLE OPERATING COSTS				**$4,185,072**	**$17,062,598**

Table 15. Baseline assumptions for the discounted cash flow analysis. All prices are expressed in 2015$ and represent levelized values over the assumed 20-year (2021-2040) economic lifetime of the plant.

Financial parameters	
Plant capacity factor (CF) after ramp up[a]	80%
Economic lifetime, years	20
Construction time, years	3
Inflation rate	3%
After-tax nominal [real] weighted average cost of capital[b] (WACC)	5% [1.94%]
Federal + state tax rate	38%
Depreciation schedule	15-yr MACRS
GHG emission rates (used for calculating value of GHG emissions)	
Displaced grid electricity lifecycle emissions[c]	661 kgCO_{2e}/MWh
Displaced petroleum-derived jet fuel[d]	88.3 kgCO_{2e}/GJ_{LHV}
Displaced petroleum-derived naphtha[d]	91.4 kgCO_{2e}/GJ_{LHV}
GTCC natural gas lifecycle emissions[e]	53.5 kg/GJ_{HHV}
Prices paid for inputs delivered to the plant gate	
Lignite (\$/GJ_{HHV})[f]	2.68
Biomass (\$/GJ_{HHV})[f]	2.94
Natural gas (\$/GJ_{HHV})[g]	6.0
Prices (baseline) received for outputs at plant gate or busbar	
SPK (\$/bbl)[h]	94
Naphtha (\$/bbl)[h]	90
Electricity to grid (\$/MWh)[i]	70
EOR-CO_2 (\$/tonne)[j]	26
Sulfuric acid (\$/tonne)[k]	140
Other prices	
Crude oil (\$/bbl)[l]	80
GHG emissions (\$/tonne CO_{2e})[m]	25

(a) The steady-state capacity factor of 80% is assumed to be reached in year 5 of plant operation. In the first year, the capacity factor is half this value. In years 2 through 4, it is 70%, 85%, and 95% of this value.

(b) For comparison, some recent average values of WACC (nominal) were 2.6% for Southern Co.; 2.8% Con. Edison; 3% Duke Energy, 3.5% PGE; 4% PSEG; 7.1% NRG 7% BP, 7.8% ExxonMobil; 8.1% Shell, and 9.6% Chevron.[52]

(c) This is the estimated U.S. grid-average lifecycle emissions for electricity generation in 2005.[53]

(d) As estimated for 2005 U.S. production of jet fuel and (as surrogate for naphtha) gasoline.[34]

(e) The GTCC operates on gas (8,000 kJ/kWh heat rate) for the hours that the LBJ plant does not operate at the rated 80% CF.

(f) See Section 3.2.

(g) For comparison, the levelized (7% discount rate) annual average natural gas price to electricity generators in the SERC generating region from 2021 to 2040, as projected in the AEO 2015 Reference Scenario,[54] is \$6.15/GJ_{HHV} (in 2013\$).

(h) Estimated as the baseline crude oil price (\$80/bbl) plus refinery margin. The margins are estimated as the levelized (7% discount rate) difference between annual average wholesale price of jet fuel (or gasoline) in the East South Central region of the U.S. and the WTI crude oil spot price from 2021 to 2040 as projected in the AEO 2015 Reference Scenario.[54] Also, see note (m).

(i) For comparison, the levelized (7% discount rate) annual average electricity generation price in the SERC generating region from 2021 to 2040, as projected in the AEO 2015 Reference Scenario,[54] is \$67/MWh (in 2013\$). Also, see note (m).

(j) Rubin, *et al*.,[55] wrote that "…conventional wisdom suggests that the price that EOR projects can afford to pay for CO_2 (in \$/1000 standard ft^3) is 2% of the oil price in \$/bbl." On this basis, the analysis here assumes this price at the oil field and \$5/t as the cost of transporting the CO_2 to the field. Thus, for an oil price of \$80/bbl (and a specific volume for CO_2 of 19.3 scf/kg), the CO_2 price is (80 x 0.02 x 19.3 – 5) = 25.9 \$/t.

(k) This is at the high-end of the range of prices for sulfuric acid deliveries to the Gulf Coast in mid-2014, according to Argus.[56]

(l) For comparison, the levelized (7% discount rate) annual average crude oil price projected for 2021-2040 in the AEO 2015 Reference, Low, and High Oil Price Scenarios are \$102/bbl, \$67/bbl, and 187/bbl, respectively (in 2013\$).

(m) Plant operator pays this rate for the lifecycle emissions from the system (including its products) and is paid prices for its outputs that include the baseline prices in this table plus the value of the lifecycle GHG emissions of petroleum-derived liquids and grid electricity displaced.

Table 16. Discounted cash flow scenarios examined.

Scenario name in Figure 20	Distinguishing input assumption
With baseline input assumptions (Table 15)	
TPC(LBJ,w/SF)	Capital cost = TPC$_{LBJ,w/SF}$ (includes an allowance for supplemental funds)
TPC(WP)	Capital cost = TPC$_{WP}$ (WorleyParsons estimate)
TPC(LBJ)	Capital cost = TPC$_{LBJ}$ (best guess capital cost estimate)
With TPC = TPC$_{LBJ}$ and all baseline input assumptions (Table 15), except:	
Lo avail. ramp	Plant capacity factor is 80% for operating years 6 to 20. In year 1 to 5, it is 0.25, 0.35, 0.55, 0.75, and 0.90 times 80%.
Hi biomass	Levelized biomass price, 2021-2040 = double the baseline value.
Hi GHG price[a]	Levelized GHG emissions price, 2021-2040 = 100 \$/tCO$_{2e}$.
Hi oil price	Levelized crude oil price, 2021-2040 = 120 \$/bbl.
WACC = 0	Real weighted average cost of capital = 0 (nominal WACC = 3%).

(a) The DCF results would be nearly identical to those for the "Hi GHG price" scenario if the baseline electricity sales price were doubled, or if the lignite price were halved, while keeping the GHG emissions price and all other input parameters at the baseline values in Table 15.

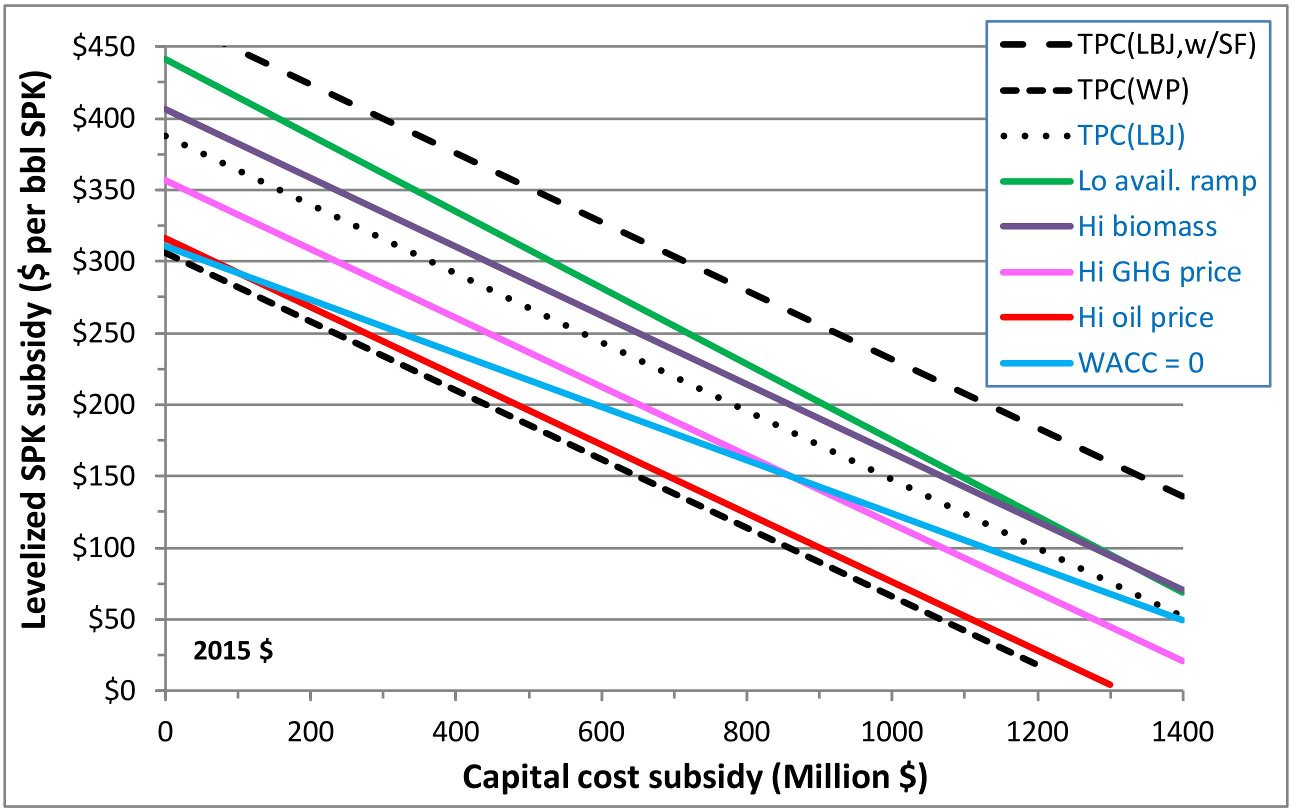


Figure 20. Discounted cash flow analysis results showing the combinations of capital subsidy and/or SPK production subsidy required to achieve zero NPV. Three scenarios identified as TPC(XXX) all use the baseline input assumptions (Table 15) but different TPC. The six scenario names in light blue in the legend all assume TPC = TPC$_{LBJ}$. These scenarios use the baseline input assumptions for all parameters except the one indicated in the scenario name – see Table 16.

6 Analysis of Prospective Nth-of-a-kind (NOAK) Commercial Plants

The economics of LBJ plants built after the FOAK demonstration plant can be expected to improve as a result of building larger scale plants, technological innovation, and "learning-by-doing" that would occur as an LBJ industry matures. To help understand the extent to which economics might improve, preliminary analyses of Nth-of-a-kind (NOAK) plant designs were undertaken, as described in this section. Without actual experience in constructing and operating LBJ plants, there is no rigorous basis on which to estimate NOAK plant costs. Our analysis provides preliminary (and perhaps optimistic) capital cost estimates for plants made up of equipment components similar to those selected for the FOAK design, i.e., only equipment that for the most part is already commercial today is considered for the NOAK plant designs. Consideration of advanced, not-yet-commercial technologies or of R&D-driven improvements in existing technologies that might improve plant performance and/or reduce capital or operating costs are beyond the scope of this project. The focus instead is on understanding the implications for performance and economics of *(i)* increased plant scales, *(ii)* capital and operating cost reductions resulting from "learning by doing", *(iii)* alternative process configurations, *(iv)* alternative lignite:biomass input ratios, and *(v)* how the economics of all system configurations considered varied with crude oil price and value of GHG emissions (the most important exogenous parameters influencing NOAK economics). The detailed Aspen process simulations developed for the FOAK analysis provided a starting point for simulating the process designs examined here, and the FOAK bare-erected cost estimates provided a quantitative basis for scaling analysis to estimate cost components for NOAK designs.

6.1 NOAK Plant Scales, Process Configurations, and Lignite:Biomass Ratios

Key technical variables in the NOAK plant designs are plant scale, plant configuration, and lignite:biomass input ratio. Larger plants have lower capital costs per unit of output, *ceteris paribus*, and thus one design objective for our NOAK plants was to make them as large as possible. We set the maximum plant scale for any process design that includes the use of some biomass by applying two constraints. First, total annual biomass input to the plant should be one million tonnes (dry basis) per year, which earlier Princeton work[57] suggests is a practical maximum for truck-delivered biomass. For an assumed 90% capacity factor for commercially mature plants, this corresponds to an input biomass capacity of 3,044 t/d of biomass. Second, the amount of lignite co-feed in a given design is set according to the desired greenhouse gas footprint for that system, as quantified by its $GHGI_{2005}$ (see discussion of $GHGI_{2005}$ in Section 4.1). The value of $GHGI_{2005}$ is determined by both the input biomass:lignite ratio and by the output fuels:electricity ratio, for a given UBAF factor.[k] The latter is determined by the process configuration.

Plants were designed for a range of target GHGI values. One set of plants was designed for $GHGI_{2005}$ = 1.0, i.e., for net lifecycle GHG emissions exactly equal to lifecycle emissions for the

---

[k] For all GHGI calculations for NOAK plants in this report, the biogenic carbon emissions are calculated using a UBAF value of negative 1.06, the average for the South region as a whole under the medium scenario (see Table 11). A different UBAF can have a significant impact on the GHGI value and on the plant's economics in the presence of a GHG emissions price. A forthcoming journal paper will examine the implications of assuming UBAF = 0.

production and consumption of an equivalent amount of conventional liquid fuels from crude oil and electricity generated with the U.S. grid-average GHG emissions in 2005. A second set of plants was designed with $GHGI_{2005} = 0$, i.e., zero net lifecycle GHG emissions, and a third set of plants was designed to use only biomass as the feedstock, which maximizes negative GHG emissions (negative $GHGI_{2005}$).[1] Two additional plants were designed using only coal as input and having nominal outputs of 50,000 bbl/day of liquids for comparison with other published designs. One of the two additional plants used 100% lignite and the other used 100% Powder River Basin subbituminous coal. These two plants are discussed in Section 6.6.

Drawing on insights from prior Princeton work,[4,5,6] three plant configurations were included in the NOAK analysis (Figure 21a). These are identified here with acronyms OT for "once-through", RC for "syngas recycle", and OTA for "once through with autothermal reforming". The three configurations vary by the design of their "light ends" processing (LEP) area, i.e. by how H_2-rich and hydrocarbon-rich offgas streams ("light ends") from the synfuels island are handled (Figure 21b).

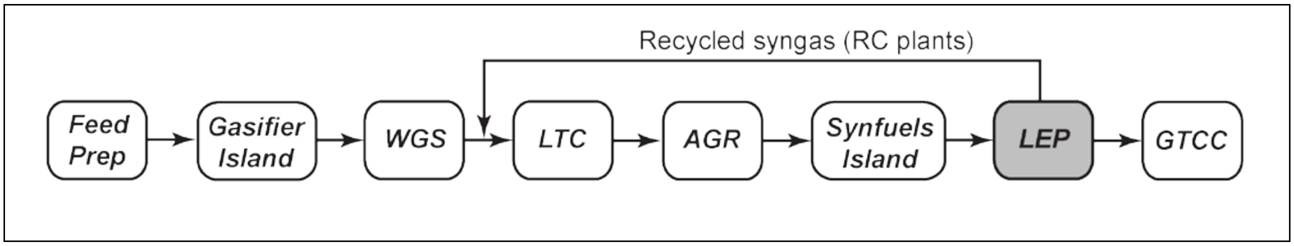


Figure 21a. Simplified schematic of primary plant processes: water-gas shift (WGS), low temperature cooling (LTC), acid gas removal (AGR), "light ends" processing (LEP; see Figure 21b), and gas turbine combined cycle (GTCC). The OT configuration excludes the recycled syngas stream. For simplicity, sour water stripper, wet sulfuric acid plant, and heat recovery steam cycle are not shown. A bypass loop (not shown) around the WGS is used as needed to adjust the H_2/CO ratio of the gas stream entering the synfuels island to equal 2.

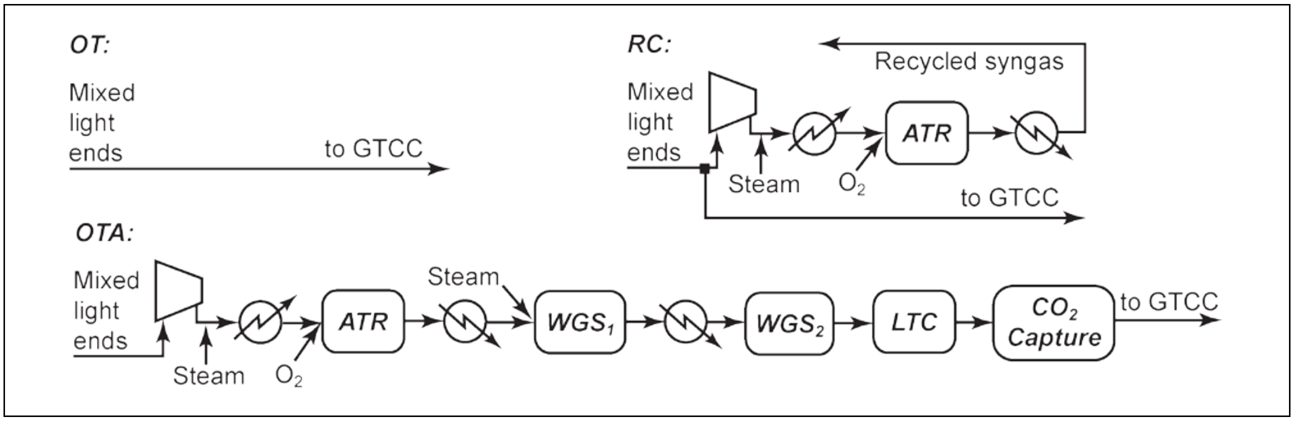


Figure 21b. Alternative designs for light ends processing (LEP).

---

[1] The TRIG™ technology was originally developed to gasify low-rank coals, but was subsequently demonstrated to work well with mixtures of low-rank coals and biomass. In the course of this project, informal discussions with engineers involved in the original design, scale-up, and coal/biomass co-processing trials of the TRIG™ suggested that the technology would be capable of operating solely using biomass. Testing would be required to verify this.

OT configuration. OT is the same configuration employed in the LBJ FOAK plant design. It is a relatively simple design that produces both synfuels and power but maximizes neither synfuels production nor CO_2 capture. As shown in Figure 21, sulfur-free syngas from the acid gas removal unit (AGR) is converted to SPK and naphtha in the synfuels island (comprising the FT synthesis reactor and syncrude refinery), along with a multiplicity of refinery light-end streams that are compressed, heated, mixed, and finally burned in a gas turbine combined cycle (GTCC) to generate electricity. The OT designation indicates that the light ends are directly combusted rather than, for example, being reformed to $CO+H_2$ and returned back to the AGR (in an external recycle loop) to increase the yield of synfuels. (Note that "once through" is technically a misnomer; in fact, all plants investigated here employ a common synfuels island design in which "internal" recycle loops around both the FT synthesis reactor and the hydrocracking unit are used to significantly boost the conversion efficiency beyond "single pass" values. These internal recycle streams are included in the Aspen process simulations, but they are not depicted in this higher level description of syngas flow through each plant.)

RC configuration. Relative to the OT configuration, the RC design increases the yield of synfuels at the expense of GTCC power. As depicted in Figure 21a and Figure 21b, instead of sending all of the mixed light ends to the GTCC, a significant fraction of the stream is split off, heated, mixed with steam and oxygen, reformed to $CO+H_2$ in an autothermal reformer (ATR), cooled, compressed, and recycled back to the primary syngas stream exiting the plant's primary water-gas shift (WGS) unit. The fraction of the mixed light ends stream that is recycled, is a design variable that can vary from zero to 95+%. Zero recycle is equivalent to the OT configuration, the LEP design that yields the highest electricity output fraction among the cases reported here. Higher recycle fractions increase synfuels output and decrease gross electricity generation. For both conceptual and analytical simplicity, all RC plants in this work are designed with a recycle fraction that yields zero net output power, i.e. they export only fuels and CO_2. It is worth noting that although an RC plant's net output power is zero, it still generates a significant amount of electricity to satisfy the auxiliary load.

OTA configuration. While RC is an OT variant that increases synfuels production, OTA is a configuration designed instead to maximize CO_2 capture. Like RC, the light ends are first reformed to $CO+H_2$ in an ATR; in OTA, however, these gases then undergo a secondary WGS to convert most (92.5%) of the CO to CO_2, and then an AGR step to remove the bulk (96.2%) of the CO_2. The downstream AGR includes a Rectisol absorption column, the solvent generation for which is integrated with that for the upstream AGR system. In the LEP design for the OTA configuration, the CO_2-scrubbed gas, which is mostly H_2 (94.4%), serves as a low-carbon fuel gas for the GTCC. It is diluted significantly with N_2 (produced by the ASU) just prior to the GTCC to reduce NO_x formation in the GTCC combustor and decrease heat transfer to the gas turbine expander blades. This configuration reduces CO_2 emissions from the GT exhaust to only about 5% of the plant's input carbon. More than 70% of the input carbon is captured for storage.

6.2 NOAK Process Simulations

Aspen process simulations were carried out for each of the NOAK plant designs. These simulations represent an extrapolation from the FOAK demonstration plant simulation. For all of the NOAK designs, the configuration and unit performance of the upstream components (feed preparation through acid gas removal ahead of the synfuels island) are identical to those for the FOAK design, with the exception of the auxiliary power consumption of the ASU. WP estimated the ASU auxiliary load for the FOAK plant to be 508 kWh per tonne of pure O_2 based on three vendor quotes that they received. Appendix E includes one of these quotes. During review of the performance estimates for the LBJ FOAK plant, engineers at Southern Company Services (SCS) indicated that based on their experience and interactions with ASU suppliers, they would expect a considerably lower ASU auxiliary load, particularly for larger-scale plants than the FOAK plant.[58] Based on the feedback from SCS, we adopted for NOAK plants an ASU auxiliary load of 386 kWh/tO_2 (350 kWh/short ton O_2).

The efficiency of gas turbines (GT) increase with scale. We calculated the power output of the GT for the NOAK plants using a model we developed that combines results of direct GT simulations in Aspen Plus with a correlation based on the published performance[59] of 102 commercial GTs whose power outputs range from 0.2 to 375 MW_e. The correlation captures the rising efficiency of commercial GTs with increasing scale. See additional discussion of GT modeling below and in Appendix G. A similar approach is used to predict improvements in steam turbine efficiencies with increasing scale, as also discussed below and in Appendix G.

No adjustments were made for the NOAK analysis to efficiencies of electric motors generally and of heat losses from reactors and heat exchangers (as a fraction of their duties). These can be expected to improve with scale, but much less significantly than the adjustments made for the ASU, GT, and ST performance.

In addition to the impacts of larger scale, the most significant differences between the NOAK and FOAK process simulations occur as a result of *i)* differences in the configuration of the light-ends processing area, *ii)* derating of gas turbine performance with H_2 firing, and *iii)* design differences in the heat recovery steam cycle.

Light Ends Processing: In the synfuels island, multiple offgas streams from FT synthesis and syncrude refining are heated, compressed, and mixed to create a GT-ready fuel gas at 117°C and 30.3 bara. Its composition, which varies with plant configuration, is primarily CH_4, CO, H_2, and light hydrocarbon gases (below C_6); however, it also contains non-negligible amounts of vaporized, long chain hydrocarbons, C_6-C_{15+}. As shown in Figure 21b, this stream (labeled “mixed light ends”) is processed in the LEP island in various ways, depending on the plant configuration. In OT plants, it goes directly to the GTCC.

In RC and OTA plants, a fraction of the mixed light ends stream is compressed to 46.4 bara (to overcome downstream pressure drops), heated using steam (55 bara, 424.5°C), and mixed with additional steam to provide a 414°C feed gas which is then reformed to synthesis gas in an O_2-blown, Ni-based autothermal reformer (ATR). The reforming process converts hydrocarbons back to CO and H_2, for either syngas recycle (RC design) or for secondary WGS and CO_2 capture for low carbon power generation (OTA design). Following Consonni,[60] the ATR input steam-to-carbon ratio is set to 1.0 and its exit temperature to 1000°C, yielding an H_2/CO mole ratio in the reformed syngas of ~2.1; a bypass around the primary WGS unit (described in the caption for Figure 21a) is adjusted to provide an H_2/CO ratio of 2.0 in the sweet syngas entering the FT synthesis reactor. In order to avoid metal dusting,[61,62] the hot syngas exiting the ATR is first cooled by evaporator tubes (59 bara water/steam) followed by a superheater that raises the steam temperature to that of the plant's primary medium-pressure steam header, 425°C.

In the RC configuration, the reformed syngas is further cooled to 189°C and recycled back to the main syngas flow into the primary low temperature cooling (LTC) section.

In OTA, the ATR reduces the volume fraction of CH_4 from 45.2 to 2.2%; the remainder of the ATR exit stream is 46.4%vol H_2, 23.2% CO, 21.5% H_2O, and 6.1% CO_2. The reformed syngas undergoes two-stage WGS in which the syngas is first cooled to 200°C and mixed with superheated medium-pressure steam to yield a 250°C WGS feed with a steam/CO mole ratio of 2.1 (steam/dry exit gas mole ratio = 5.1).[63] This gas enters the first (adiabatic, Fe/Cr-based) WGS reactor and exits at 403°C with 77.1%vol of its CO converted to CO_2 + H_2. The shifted syngas is cooled to 250°C and sent to a second (adiabatic, Cu/Zn-based) WGS reactor, which increases the CO molar conversion to 92.6%. The shifted syngas is cooled by a series of condensing heat exchangers in the secondary low temperature cooling (LTC) section, followed by a Rectisol CO_2 capture column that extracts 96.2% of the CO_2. The remaining GT fuel gas, which is 92%vol H_2, has 83.9% of the lower heating value (LHV) of the original mixed light ends.

Gas turbine simulation. In the OT and RC configurations, the GT fuel gas is sufficiently similar to natural gas to be burned in a standard dry low-NO_x premixed combustor.[64] In the OTA configuration, the high H_2 content of the OTA fuel gas (92%vol before N_2 dilution) precludes the use of a premixed combustor because of the high flame speed with H_2 combustion and associated risk of flashback; instead, the OTA designs employ fuel gas dilution and assumes non-premixed combustion. (We also assume that no modification to standard commercial GTs are required; we assume the same pressure ratio and expansion nozzle area. Only the position of the compressor inlet guide vanes are assumed to change relative to operation on natural gas.) Following Chiesa[65] and Gazzani,[66] the OTA fuel gas is diluted with compressed N_2 until its stoichiometric flame temperature is 2300 K, which reduces the H_2 concentration to 49.6% and NO_x formation in the GT combustor to ~40 ppmv (down from ~200 ppmv), significantly decreasing the NO_x load on the downstream selective catalytic reduction (SCR) unit. Fuel gas dilution also minimizes changes (relative to natural gas) in the fuel's Wobbe index, and lowers the concentration of water vapor in the hot exhaust gases flowing through the gas turbine expander (from 45.8%vol undiluted to 11.5% with dilution), reducing heat transfer to the turbine blades and thus the rate of their thermal degradation. Following Martelli,[67] in order to further reduce this relatively high heat transfer and thus insure standard blade lifetimes, we additionally derate the turbine inlet temperature (*TIT*) of the GT. The method of Chiesa,[65] which reduces the *TIT* until critical blade temperatures obtained during H_2-rich syngas operation equal those when firing on natural gas (using the same air cooling flows), yields a ~35°C reduction in *TIT*. We employ here instead the correlation of Oluyede,[68] because it is more conservative:

$$\Delta T_{\mathrm{TIT}}\,(°\mathrm{F}) = 13.312\ \Phi_{\mathrm{H2}}^{0.69} \qquad \text{Eqn. (6)}$$

where ΔT_{TIT} is the reduction in *TIT* (°F) and Φ_{H2} is volume percentage of H_2 in the *diluted* syngas fed to the GT combustor. In case OTA (49.6% H_2 in the diluted gas), this correlation derates the *TIT* by 109°C, which leads to a 2.6% reduction in GT output power (relative to no derating: TIT = 1,277°C). Our model for predicting GT performance indicates that diluted H_2-rich OTA syngas yields 8.6% more power from the GT than with an equal LHV flow of natural gas, but after subtracting the power needed for N_2 compression (from a high pressure ASU, 234.3 kJ/kg N_2), the net power output is only 2.7% higher.[m]

Process heat recovery and power generation simulation. For steam cycle power generation from recovered process heat, we used a similar design approach as was pursued in detail for the FOAK plant (see Appendix D), but with some adjustments made for expediency in estimating the performance of the multiple NOAK plant designs. The FOAK plant design, which uses the OT configuration, utilizes two separate heat-recovery steam cycles (HRSC), one dedicated to using the GT exhaust and a second, "bespoke" HRSC using the plant's multiplicity of other sources of both process heat and steam. For the NOAK plant using the OT design, this same arrangement was simulated at larger scale. In the case of the OTA and RC designs, the bespoke HRSC was assumed to use the same process heat sources as in the OT design, and so its performance (and cost) could be scaled from the FOAK design. The HRSC bottoming cycle of the GTCC was assumed to utilize the process heat streams not used by the bespoke HRSC. These streams occur primarily in the light-ends processing area. Details of the power island simulation are provided in Appendix G.

6.3 NOAK Plant Performance Results

Technical performance results for all NOAK cases that use biomass are given in Table 17. The simulated performance of the **FOAK** plant is shown for comparison. The size of all plants except the ***FOAK*** are determined by the biomass supply constraint (1 million t/y) discussed above. For an assumed 90% capacity factor for commercial plants this corresponds to a biomass input capacity of 711 MW_{th} HHV for these NOAK designs. In Table 17, the plants with $GHGI_{2005}$ = 0 are moderately sized (11,000-16,000 bbl_{eq}/day FTL), with biomass input fractions of 18-25%. The fraction of their outputs that is electricity range from 0 (***RC-0***) to 13.2% (***OTA-0***) to 26.4% (***OT-0***).

The group of plants with $GHGI_{2005}$ = 1 are much larger plants (48,000-61,000 bbl_{eq}/day FTL) because the relaxed GHG emissions constraint allows much more lignite to be consumed for the given fixed biomass input.[n] The $GHGI_{2005}$ = 1 plants have a carbon footprint equal to that of the 2005 reference system, so would be unlikely to be built unless they prove to have considerably better economics than the reference system and/or unless the use of domestic lignite and biomass for transportation energy is very highly valued.

---

[m] Firing the GT on OTA fuel gas vs. natural gas, the mass flows of compressor air and of fuel are 1.2% and 27% lower, respectively; however, the added N_2 diluent increases the total mass flow expanded in the turbine by 7.7%. The power needed to compress N_2 (to 32 bara) from a high pressure ASU is reduced by ~40% relative to that from a low pressure ASU.[73]

[n] An OTA-1 case was also simulated, but is not included in Table 17 because there would seem to be little point in adding the complexity of an ATR, WGS, and CO_2 capture to reduce GHG emissions only to the level that achieves GHGI of 1.

The group of plants using 100% biomass as input are all relatively small plants (2,700-3,500 bbl$_{eq}$/day FTL), and all have strongly negative *GHGI* values, i.e., they provide net removal of GHGs from the atmosphere. For each of the three groups of designs, the OT and RC designs have comparable first-law efficiencies, but the OTA design suffers a significant (6 to 7 percentage points) energy efficiency penalty compared to OT, as a result of the added ATR (which requires additional O_2 production), WGS, and CO_2 capture (which requires additional steam consumption for solvent regeneration). Comparing the GHGI$_{2005}$ = 0 cases with the biomass-only cases, Table 17 indicates that the negative GHG footprints of the latter are achieved with relatively modest penalties in first-law efficiency of 1.3 to 1.7 percentage points.

Table 17. Performance simulation results for NOAK plant designs.

		NOAK Plants							
		GHGI$_{2005}$ = 0			***GHGI$_{2005}$ = 1***		***Biomass-only cases***		
Plant Configuration		***OT-0***	***OTA-0***	***RC-0***	***OT-1***	***RC-1***	***OT-B***	***OTA-B***	***RC-B***
Recycled syngas fraction, %	-	-	-	60.5	-	61.9	-	-	53.8
Syngas bypass fraction, %	-	-	-	-	-	-	-	-	-
Dry biomass, Mtonne/yr	0.104	1	1	1	1	1	1	1	1
Biomass fraction, % HHV	25.0	25.3	17.5	23.8	4.6	7.8	100	100	100
Performance Metrics									
Liquids fraction, % LHV	83.4	73.6	86.8	100	73.2	100	75.1	89.8	100
Electricity fraction, % LHV	16.2	26.4	13.2	0	26.8	0	24.9	10.2	0
First law efficiency, % LHV	35.6	41.5	35.2	41.1	41.8	41.4	40.2	33.6	39.6
GHGI$_{2005}$ (UBAF = - 1.06)	0[a]	0	0	0	1	1	-3.94	-5.94	-5.14
Plant Inputs									
AR lignite, MW$_{th}$ HHV	220.9	2,101	3,342	2,281	14,723	8,357	0	0	0
AR biomass, MW$_{th}$ HHV	73.6	711	711	711	711	711	711	711	711
Total AR feedstock, t/day	2,107	20,121	28,831	21,383	108,747	64,046	5,365	5,365	5,365
Plant Outputs									
SPK, MW$_{th}$ LHV	62.3	609.4	881.3	872.5	3,375.9	2,675.5	149.1	149.1	195.7
Naphtha, MW$_{th}$ LHV	13.9	136.0	196.7	194.7	753.5	597.1	33.3	33.3	43.7
Total FTL, bbl$_{eq}$/day	1,124	10,986	15,886	15,728	60,854	48,228	2,687	2,687	3,529
GT power output, MW$_e$	30.5	336.9	441.7	174.9	1,850.7	516.9	84.7	76.8	46.4
GT SC power output, MW$_e$	12.5	158.8	172.0	108.4	899.5	340.2	35.4	25.1	21.5
Bespoke HRSC power, MW$_e$	10.0	102.5	148.2	109.1	567.0	332.7	25.2	25.2	25.2
Auxiliary loads,[b] MW$_e$	-38.2	-331.0	-598.6	-392.3	-1,806.7	-1,190.0	-84.7	-106.4	-93.0
Aux. load, % feedstock HHV	13.0	11.8	14.8	13.1	11.7	13.1	11.9	15.0	13.1
Net power output, MW$_e$	*14.8*	*267.1*	*163.4*	*0.1*	*1,510.6*	*-0.1*	*60.5*	*20.7*	*0.2*
CO_2 captured, tonne/day	1,326	12,635	24,114	14,317	68,129	42,941	3,392	4,455	3,573
H_2SO_4 (93% wt), tonne/day	50.3	478.8	760.9	519.7	3,348.4	1,901.1	1.1	1.1	1.1
Carbon disposition (% of input)									
In CO_2 vented at plant	22.3	23.6	5.4	13.2	23.7	12.9	23.1	5.4	14.4
In FTL	19.1	19.6	19.7	26.3	19.9	26.8	18.4	18.4	24.1
In CO_2 captured	54.9	54.8	72.8	58.4	54.3	58.2	56.5	74.2	59.5
In gasifier ash and tar	2.4	2.0	2.0	2.0	2.0	2.0	2.0	2.0	2.0

a) This is the FOAK demonstration plant design described in earlier sections of this report. The GHGI$_{2005}$ value here is different from the demonstration plant because the UBAF value used for the GHGI$_{2005}$ in this table is the South-region average for the medium scenario in Table 11 (-1.06) and not the Mississippi-specific value (-0.41).

b) Appendix G (Part 2) provides detailed breakdowns of each plant's auxiliary load.

From a carbon-balance perspective, across all plant designs, 20-27% of the input carbon is converted to FTL, 54-74% is captured as CO_2 for EOR, and 5-24% is vented at the plant as CO_2. OT plants capture the smallest fraction (54-57%), RC are intermediate (58-60%), and OTA capture the most (73-74%).

As discussed above, when comparing case ***FOAK*** to ***OT-0***, the reduced ASU specific power consumption and increased gas and steam turbine efficiencies when operating at commercial scales yields a not-insignificant 17% increase in first law efficiency, from 35.6% to 41.5%.

6.4 NOAK Cost Analysis

To assess the prospective economics of commercial-scale NOAK plants, capital and operating costs were developed for the above plant designs using a simplified scaling approach, informed by the detailed estimates made for the FOAK plant and assuming that significant cost-learning occurs from the FOAK design to a NOAK plant. There is relatively little data to suggest what a credible prospective cost-learning rate might be for the type of plants evaluated here. We have chosen to adopt what might be considered optimistic estimates of future plant costs, as explained next.

6.4.1 NOAK capital cost estimation

As mentioned above, it is impossible to accurately forecast a learning rate for complex process technologies associated with megaprojects, especially when the first plant is yet to be subject to detailed engineering, constructed and operated and so the actual FOAK plant cost remains unknown. However, we have constructed a scenario for NOAK plant costs by assuming the current uncertainties reflected in high contingencies and supplementary funding allowances reduce to levels associated with mature technologies. In particular, we assume that over time there will be learning such that: 1) EPCM fees fall to 12% of the bare-erected cost (BEC), which compares with 20% used for our FOAK analysis; 2) process contingencies fall from 35% to zero; 3) project contingencies fall from 35% to 10% of *BEC*; and 4) supplementary funds are not needed. The result of such a scenario is that total plant cost, TPC, would be 1.22×BEC (excluding owner's costs, which are treated separately, as discussed in Section 6.4.3). For comparison, in the FOAK analysis, TPC_{FOAK} (excluding owner's costs) ranged from 1.7×BEC to 2.1×BEC when 35% process and project contingencies are assumed (as suggested by Table 13 guidelines for FOAK projects) and 2.5×BEC if 20% supplementary funds (SF) are also included.

The NOAK assumptions in this scenario are plausible in the context of the guidelines presented in Table 13 for estimating costs as a function of both the level of technology maturity and stage of project development – when adjustments are made for the small scales of the systems considered. The assumed reduction in EPCM fees is reasonable if system designs are shared among successive early-mover projects and no novel components are introduced along the way. The assumption of zero process contingency is consistent with Table 13 (because processes are mature for NOAK plants). However, the assumed 10% project contingency might be considered "optimistic" compared to the 25% project contingency recommended in Table 13 for NOAK cost estimates derived from scoping studies (such as the present study). A mitigating consideration is that Table 13 guidelines are for large plants that typically require a great deal of field construction, whereas much of construction for components of the systems considered here can be carried out in factories because of their modularity – so that much of the required field construction may involve more "bolting together" of modular components; thus, arguably the lower contingency allowance might be warranted. Finally, the assumption of zero SF (compared to the 15% recommended in Table 13) might also be considered optimistic. The risk of exposure to SF (which might never be spent even in the FOAK project) will likely be considerably reduced with the experience of many projects and/or may often be covered by an insurance policy to facilitate project financing arrangements with lenders. As will be discussed below, based on our analysis, the only NOAK options considered that might end up becoming commercial are some of those that involve only biomass as feedstock, in which cases the maximum SF according to Table 13 guidelines would be about $200 million. This is a modest amount in the realm of the megascale projects reviewed that gave rise to the guidelines of that table.

We constructed BEC estimates for most components of the NOAK plants by appropriately scaling component costs estimated for the FOAK design. This involved a scaling up by factors of about 2 for the biomass-only designs, factors of about 10 for the GHGI = 0 designs, and factors of 32 to 50 for the GHGI = 1 designs. The rationale for this approach is that vender quotes and WorleyParson's internal capital-equipment cost database used to construct the FOAK estimate represent the most credible information available to us about current and future costs of component and, with the exception of the TRIG^{TM} gasifier, all components used in the NOAK plants are already quite mature. Thus, no significant reductions in equipment costs per unit throughput would be expected beyond those that accompany larger scale.

BEC values (except for the GT and GT HRSC) were scaled from the FOAK values using component-level power-law scaling exponents. The bare erected cost, C (M$), of a plant component having size S, is related to the bare erected cost, C_o, of a single train of a reference component having size S_o, by the relationship:

$$C = n^e C_o [S/(n S_o)]^f \quad \text{Eqn. (7)}$$

where n is the integral number of equally sized equipment trains having capacity S/n, f is the component-specific power law cost scaling factor taken from Turner and Pinkerton[69] in most cases, and $e = 0.9$ is the cost scaling exponent for multiple trains of equipment. When multiple trains were needed, they were designed for equal sizes. The assumed component-level capital cost scaling parameters are given in Table 18, where the number of trains for five key plant components (labeled $n_1 - n_5$) are given by:

$$n = S / S_{max} \text{ (if result is a non-integer, then round up to nearest integer)} \quad \text{Eqn. (8)}$$

where S_{max} is the largest single train capacity currently employed commercially. For many other components, their number of trains is set equal to the most closely related of those five, creating a series of coherent, parallel process trains.

The BEC for the "bespoke" HRSC is scaled in the same manner as described above. The GT and GT HRSC are estimated differently than for other plant components, as summarized next (and described in more detail in Appendix G). For the GT, capital cost correlations were constructed using price and performance data on commercial GTs in the *Gas Turbine World Handbook*.[59] The costs of GTs operating on natural gas, C_{GTg} (2015 \$/kW$_e$), are found to be well-approximated (Figure 22) by:

$$C_{GTg} = \varepsilon\; 827.3\; {p_{GTg}}^{-0.217} \qquad \text{Eqn. (9)}$$

where p_{GTg} is the output power in MW$_e$ and $\varepsilon = 0.9524$ (the CEPCI cost escalation factor from 2012 to 2015 US dollars[70,71]) For simplicity we use this relation for all GTs, regardless of fuel gas composition. Note that when the GT output exceeds 375 MW$_e$, a 2×1 GT×ST configuration is assumed (and an even number of GTs, n_5). Consistent with our optimistic NOAK cost assumptions, additional costs for spare equipment trains are not included.

Table 18. Component-level capital cost scaling parameters.[a]

Plant component(s)	*Scaling parameter*	*S_o*	*S_{max}*	*n*	*f*	*C_o (M\$)$_{2015}$*
Lignite handling	AR lignite, tonne/day	1,366	-	1	0.62	14.73
Lignite milling & drying	AR lignite, tonne/day	1,366	-	= n_1	0.66	47.63
Biomass prep. & drying	AR biomass, tonne/day	489.3	-	= n_1	0.66	28.36
Feedwater & misc. BOP	AR feed, MW$_{th}$ HHV	259.3	-	= n_1	0.713	19.33
Gasifier, SG cooler & aux.	Dried feed, tonne/day	1,281	5,000	n_1	0.64	88.11
ASU & O_2 compression	Pure O_2 output, tonne/day	617.0	3,750	n_2	0.70	61.28
Gas cleanup & piping	AR feed, MW$_{th}$ HHV	259.3	-	= n_1	0.725	14.72
Rectisol AGR	Captured CO_2, tonne/day	1,167	8,000	n_3	0.63	88.67
CO_2 drying & compr.	Captured CO_2, tonne/day	1,167	-	= n_3	0.67	13.88
WSA unit	H_2SO_4 (93%wt), tonne/day	44.32	-	= n_3	0.67	13.52
FT synthesis & refining	FTL output, bbl$_{eq}$/day	989	70,000	n_4	0.67	68.27
LEP ATR	Syngas output, kmol/hr	31,000	-	= n_4	0.67	99.17
LEP WGS reactor	Syngas input, MWth LHV	1,377	-	= n_4	0.725	42.36
GT and accessories	GT power, MWe	26.82	375	n_5 (even)	-	Eqn. (7)
GT HRSC	Heat recovered, *q*, MWth	31.22	-	= $n_5/2$	-	Eqn. (8)
Bespoke HRSC ST	Bespoke ST net pwr, MW$_e$	7.96	-	1	0.713	6.29
Cooling water system	AR feed, MW$_{th}$ HHV	259.3	-	1	0.598	10.66
Cooling towers	AR feed, MW$_{th}$ HHV	259.3	-	1	0.72	1.99
Ash handling	AR feed, MW$_{th}$ HHV	259.3	-	1	0.64	17.95
Accessory electric plant	Auxiliary power, MW$_e$	33.65	-	1	0.45	24.61
Main power transformers	Net plant power, MW$_e$	77.37	-	1	0.71	0.92
Instruments & controls	AR feed, MW$_{th}$ HHV	259.3	-	1	0.24	8.24
Other I&C equipment	AR feed, MW$_{th}$ HHV	259.3	-	1	0.13	3.17
Improvements to site	AR feed, MW$_{th}$ HHV	259.3	-	1	0.34	15.51
Buildings & structures	AR feed, MW$_{th}$ HHV	259.3	-	1	0.09	5.77
Steam turbine building	AR feed, MW$_{th}$ HHV	259.3	-	1	0.45	1.78

a) Notes: The FOAK design did not include a LEP ATR or LEP WGS. The characteristics in this table for these components were estimated from previous work at Princeton[4] and escalated to 2015 US dollars using the Chemical Engineering Plant Cost Index.[70,71] Most scaling exponents in this table are based on Turner and Pinkerton, but the latter document does not give scaling guidelines for Rectisol or FT synthesis components. For Rectisol, the value of 0.63 was used, based on communications with an experienced Rectisol engineer at Lurgi, as reported by Liu, *et al.*[4] For the FT island, a generic 0.67 scaling exponent was assumed.

The cost of the GT HRSC (i.e. HRSG and ST) is estimated separately by first noting that for commercial natural gas-fired heavy duty GTCCs, for which installed prices are published in the *Gas*

Turbine World Handbook, on average about two-thirds of the GTCC output is provided by the GT and one-third by the ST.[59] The price for the HRSG + ST is then determined as the difference between the price for the GTCC and the price for its GT (as estimated from Figure 22), and plotted as a function of the amount of heat recovered by the HRSG (Figure 23). The resulting best fit to the calculated values is

$$C_{SCg}(q) \; = \; \varepsilon\,(0.378\; q + 8.051) \qquad \text{Eqn. (10)}$$

where C_{SCg} (in 2015 M$) is the price for the GT HRSC, q (MW_{th}) is the HRSG heat-exchange duty, $\varepsilon = 0.9524$ (as before).

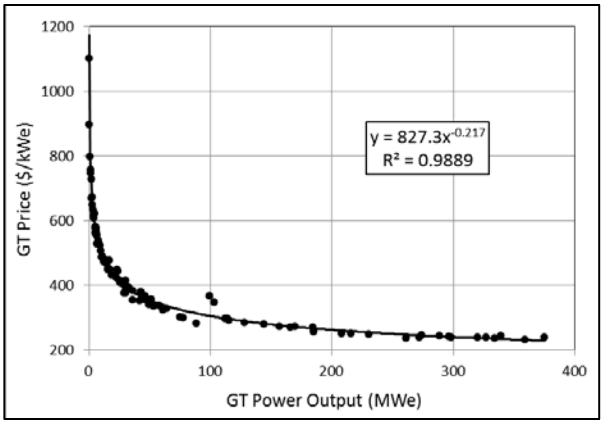


Figure 22. Published prices (installed)[59] for commercial GTs as a function of output power and best curve fit.

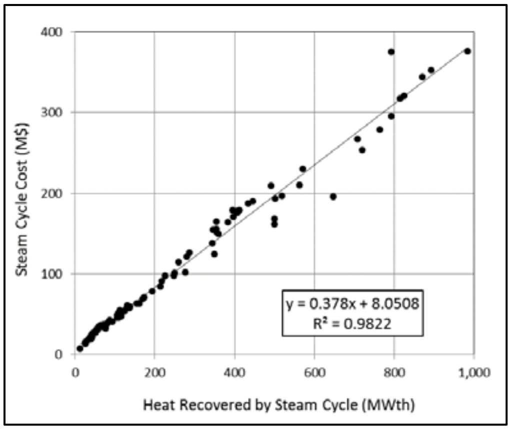


Figure 23. Cost (2015$) of HRSG plus ST as a function of the heat recovered by the HRSG, *q*.

The above cost estimating methodology results in the (partially aggregated) component-level *BEC*s given in Table 19. Economics of scale are evident there. For example, the ***OT-0*** case is roughly ten times larger than **FOAK**, but its *BEC* is larger by only a factor of five.

6.4.2 NOAK operating and maintenance cost estimation

Non-feedstock operating and maintenance (O&M) costs for the NOAK designs were estimated based on scaling up the O&M costs estimated for the LBJ FOAK plant by WorleyParsons (WP), as described here.

Table 19. Estimated NOAK plant bare erected costs (million 2015$).

	FOAK	*OT-0*	*OTA-0*	*RC-0*	*OT-1*	*RC-1*	*OT-B*	*OTA-B*	*RC-B*
Lignite handling & drying	68	363	520	383	1,826	1,117	0	0	0
Biomass handling & drying	31	179	192	179	268	233	138	138	138
Air Separation Unit (ASU)	67	373	641	472	1,709	1,226	120	140	133
Gasifier & ash handling	115	622	837	647	2,714	1,688	204	204	204
Gas cleanup (AGR, SWS, WSA)	127	658	1,131	706	2,949	1,932	205	238	211
CO_2 compression	15	80	145	87	351	235	28	34	29
FT synthesis & refining	74	343	439	436	1,078	923	133	133	160
Light ends processing (LEP)	0	0	116	55	0	117	0	35	19
GT + GT HRSC	32	240	287	132	1,170	366	69	58	40
Custom HRSC	7	39	51	41	132	90	14	14	14
Balance of plant (BOP)	97	324	415	338	1,028	704	147	151	148
Total Plant BEC, M$	*634*	*3,220*	*4,775*	*3,474*	*13,224*	*8,629*	*1,059*	*1,146*	*1,098*

Total annual fixed O&M cost is the sum of fixed operating labor, fixed maintenance labor, and fixed administrative and support labor.

For fixed operating labor, the estimates by WP for the LBJ FOAK plant were used with those also estimated by WP for the study by Shah, *et al.*,[2] for a 50,000 bbl/day coal-to-FT liquids facility. The LBJ FOAK plant requires twenty-one 24/7-equivalent operators (Table 14) and the Shah design (with 45 times the liquids production capacity) requires 64 such operators. The implied scaling exponent is 0.29, which we applied to calculate the number of equivalent 24/7 fixed operators for NOAK plants as a function of the feedstock input capacity, $HHV_{in,NOAK}$:

$$Operators_{NOAK} = Operators_{FOAK} \cdot \left(\frac{HHV_{in,NOAK}}{HHV_{in,FOAK}}\right)^{0.29} \qquad \text{Eqn. (11)}$$

Lora Pinkerton, the lead cost-estimator at WP, in a meeting with the authors confirmed that a scaling exponent between 0.2 and 0.3 would be reasonable. To calculate cost of the operators, we assume an average hourly wage, including benefits, of $59.3, as estimated by WP for the FOAK plant.

WP estimates costs for maintenance labor and materials as a percentage of the bare-erected cost in both the LBJ FOAK and Shah studies, allocating 40% of the resulting number to fixed maintenance labor and 60% to variable maintenance materials. (The latter assumes 80% capacity factor.) Additionally, fixed administrative and support labor is 25% of the total fixed labor (operating and maintenance) in both studies. For the FOAK plant, total maintenance labor and materials is 3.35% of BEC. In the Shah study it is 3.65%. For the NOAK plants we use a fixed 3.5% of BEC independent of plant size. Thus, annual fixed maintenance labor (AFML) and annual fixed administrative and support labor (AFASL) are:

$$AFML = BEC \cdot 0.035 \cdot 0.40 \qquad \text{Eqn. (12)}$$

$$AFASL = 0.25 \cdot (AFML + AFOL) \qquad \text{Eqn. (13)}$$

Where AFOL is annual fixed operating labor calculated as described above.

The total annual variable O&M costs are the sum of the cost of variable maintenance materials and of consumables. The annual variable maintenance materials (AVMM) costs are

$$AVMM_{NOAK} = BEC_{NOAK} \cdot 0.035 \cdot 0.60 \cdot \left(\frac{CF_{NOAK}}{0.80}\right) \qquad \text{Eqn. (14)}$$

where CF_{NOAK} is the capacity factor. The cost for consumables will scale linearly with outputs. For LBJ plant designs, two consumables are connected exclusively with electricity generation: catalyst and ammonia for SCR. For the FOAK plant these cost $0.722 million/yr (assuming 80% capacity factor), and this value is scaled linearly with gas turbine power output to estimate NOAK costs. Two consumables connected primarily with FTL production are the synthesis and hydrocracking catalysts. These cost $1.132 million/yr for the FOAK plant (80% CF) and this value is scaled linearly with total barrels per day (BPD) liquids production to estimate NOAK costs. Finally, the remainder of consumables are essentially independent of the ratio of liquids to electricity outputs. These total $2.6 million/yr for the FOAK plant (80% CF), and this value is scaled linearly with total feedstock input to estimate NOAK costs. To summarize, the total cost for consumables for a

NOAK design (in million \$/yr) is the sum of electricity-related costs (EC), fuels-related costs (FC), and input-related costs (IC) given by:

$$EC_{NOAK} = 0.722 \cdot (CF_{NOAK}/0.80) \cdot \left(MW_{GT,NOAK}/MW_{GT,FOAK}\right) \qquad \text{Eqn. (15)}$$

$$FC_{NOAK} = 1.132 \cdot (CF_{NOAK}/0.80) \cdot (BPD_{NOAK}/BPD_{FOAK}) \qquad \text{Eqn. (16)}$$

$$IC_{NOAK} = 2.6 \cdot (CF_{NOAK}/0.80) \cdot (HHVin_{NOAK}/HHVin_{FOAK}) \qquad \text{Eqn. (17)}$$

6.4.3 NOAK financial parameter assumptions

For the NOAK economic analysis, we utilize the EPRI TAG "revenue requirement" methodology.[72] Key financial parameter assumptions are given in Table 20, where a few relatively optimistic (NOAK-inspired) choices bear particular note, including a real rate of return on invested equity ("hurdle rate") of 10.2%/yr, a 3-year construction time, which leads to an allowance for funds used during construction (*AFUDC*) of 7.16% of *TPC* (assuming equal annual end-of-year payments), and a resulting 20-year levelized capital charge rate (*CCR*) of 15.6%/yr applied to the total plant investment (*TPI* = *TPC*+*AFUDC*).

Table 20. Financial parameter assumptions, following EPRI TAG.[72]

Parameter	Value
Debt:equity ratio, %	55:45
Cost of debt, %/yr	4.4 real, 6.5 nominal
Cost of equity, %/yr	10.2 real, 12.4 nominal
Corporate income taxes	39.2 %/yr
Owner's costs (*OC*)	22.8% of *TPC*
Property taxes & insurance	2% of *TPC+OC*
Construction time, yr	3
Book life / tax life, yr	20 / 20
Depreciation schedule	MACRS
AFUDC	7.16% of *TPC*
Capital charge rate (*CCR*)[a]	15.6%/yr of *TPI*

a) With the EPRI TAG methodology, the CCR factors in all of the parameters in the upper section of this table, except for construction time. The latter is used in calculating the allowance for funds used during construction (AFUDC).

Operation and maintenance (O&M) costs using the approach described in the previous section are approximately 5.5% of *TPC* per year for any of the NOAK plants. We assume a 90% operating capacity factor from day one of operation. For feedstock prices, we assume 20-year levelized costs for lignite and woody biomass of 2.14 and 3.70 \$/GJ HHV, respectively.[o] As in the FOAK financial analysis, the plant gate sales price of CO_2 for EOR, P_{CO2} (\$/tonne), is assumed to increase with the untaxed market price of crude oil, P_{oil} (\$/bbl), as follows.

$$P_{CO_2} = \lambda P_{oil} - \tau \qquad \text{Eqn. (18)}$$

where $\lambda = 0.386$ bbl oil/tonne CO_2 [see note (j), Table 15] and $\tau = 5$ \$/tonne CO_2 is the assumed cost of supercritical pipeline transport from the plant to the EOR site. For example, at a crude oil price of \$80/bbl, the plant sells CO_2 (dried and compressed to 150 bara) for 25.8 \$/tonne. Revenue received for co-product H_2SO_4 is assumed to be the same as in the FOAK analysis, 133 \$/tonne (Table 15).

---

[o] For reference, the estimated marginal price for 8 million tons of pulpwood-quality logs (delivered) is \$4.6/GJ HHV (\$84/short ton, dry), and for woody residues is \$3.3/GJ HHV (\$60/short ton, dry).[16] The biomass price for the NOAK analysis assumes that approximately 70% of the plant feedstock is residues and 30% is logs.

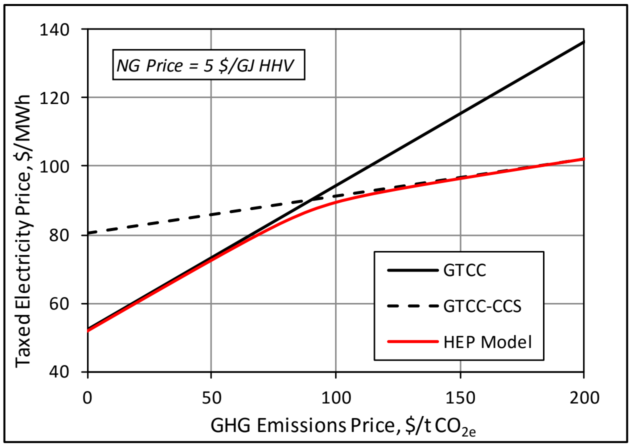


Figure 24. Levelized cost of electricity as a function of GHG emissions price for GTCC, GTCC-CCS and the hyperbolic electricity price (HEP) model.

Electricity is an important co-product for some plant designs. For the economic analysis we assume the price paid for electricity equals the levelized cost of electricity from a natural gas GTCC estimated using the EPRI TAG methodology, performance and capital and O&M cost estimates of NETL,[73] and a levelized natural gas price of 5 $/GJ HHV. With no carbon mitigation policy in place (i.e., zero GHG emissions price), this generating cost is 52 $/MWh. Because commercial LBJ plants capturing CO_2 are unlikely to be built absent any carbon mitigation policy, however, we include the GHG emissions price as a variable in our analysis. For non-zero GHG emission prices, we assume, for analytical simplicity, that the selling price of electricity follows the "hyperbolic electricity price" (HEP) model shown in Figure 24 (red line). At low CO_2 emission prices, a natural gas-fired GTCC is assumed to be the price-setting generator. At high GHG emission prices, the HEP model assumes that GTCC with post-combustion CO_2 capture and underground storage in a deep saline aquifer (GTCC-CCS) would be the price-setting generator. The HEP model provides a convenient analytical form for calculating a GTCC-based wholesale electricity price paid to electricity generators over the full range of CO_2 emission prices examined. The reduced slope of the HEP curve at higher GHG emissions price is an important feature of the model; without it, net power generation would be significantly overvalued at high GHG emissions prices.

In the presence of a GHG emissions price, the plant operator pays for the net lifecycle emissions generated by the system, including upstream emissions associated with feedstock production and delivery, emissions at the plant, transport of liquids to end users, and combustion of the products. Components of the net lifecycle GHG emissions are estimated in the same manner as for the FOAK plant, except for emissions associated with CO_2 supplied for enhanced oil recovery, as explained in note (d) of Table 7. Some NOAK results are also reported for CO_2 storage in deep saline formations (DSF) rather than via EOR. For DSF storage, emissions associated with pipeline transmission of the CO_2 to a DSF site are included, and emissions during injection and long-term storage are assumed negligible. For estimating biogenic emissions for the NOAK systems, the South-region UBAF factor for the medium scenario, -1.06, is used (Table 11).

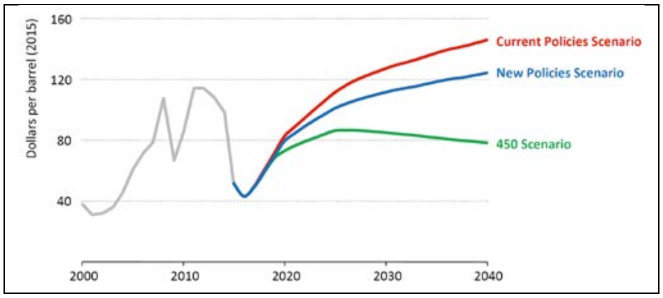


Figure 25. Average crude oil import prices for three scenarios in the 2016 World Energy Outlook.[74]

A final input parameter assumption needed for some of the results reported in the next section is the assumed market price for crude oil. Here we adopt $80 per barrel as a reference case value, which is consistent with the crude oil price projected by the IEA in its 2016 World Energy Outlook for future levels of global crude oil demand consistent with an emissions scenario having a 50% chance of limiting global warming to 2^{o}C – the IEA's "450 scenario" (Figure 25).

6.4.4 NOAK economic performance

Economic performance is assessed here using three metrics: levelized FTL production cost, breakeven crude oil price (BEOP), and internal rate of return on equity.

Table 21 shows total project investment for NOAK plants and a breakdown of the levelized FTL production cost for two different GHG emissions prices: \$25/tonne CO_{2e} and \$100/tCO_{2e}. Comparing results across cases in Table 21 provides some insights. For example, the impact of scale economies is evident by comparing **FOAK** and **OT-0** plants: these have similar configurations but a factor of 10 difference in scale. As a result, the capital charges per unit of FTL produced are half the **FOAK** level for **OT-0**.

In the case of GHGI = 1 plants, scale-economy gains relative to the counterpart GHGI = 0 configuration are augmented by lower average feedstock cost due to the lower input biomass fractions. However, despite these differences, all GHGI = 0 and GHGI = 1 plants have net FTL production costs in a relatively narrow range of \$29-40 \$/GJ$_{LHV}$ (3.4-4.8 \$/gal gasoline$_{eq}$) for a GHG emissions price of \$25/tCO_{2e} and 33-39 \$/GJ$_{LHV}$ (4.0-4.7 \$/gg$_{eq}$) for a GHG emissions price of \$100/tCO_{2e}, because the scale and feedstock cost benefits for the GHGI = 1 cases are partially offset by higher costs for GHG emissions. At this GHG emissions price, the 100% biomass plants, despite diseconomies of scale and high feedstock costs, have very low (and even negative) FTL production costs because they get credit for their negative GHG emissions. Revenues from the sale of CO_2 are moderately important in all cases. In OT plants, revenue from output power is comparable to that from CO_2 sales; in OTA plants, it is a factor of 2 to 3 lower. Revenues from H_2SO_4 are small or zero.

Table 21. Estimated FTL production costs and breakeven crude oil prices for NOAK systems.

	FOAK	***OT-0***	***OTA-0***	***RC-0***	***OT-1***	***RC-1***	***OT-B***	***OTA-B***	***RC-B***
Total Plant Invest. (*TPI*), M\$	829	4,210	6,242	4,542	17,290	11,282	1,385	1,498	1,435
TPI, thousand \$ per bbl/day	738	383	393	289	284	234	515	558	407
Capital @ 15.6%/yr of *TPI*	59.6	31.0	31.7	23.3	23.0	18.9	41.6	45.0	32.9
O&M @ ~5.5%/yr of *TPC*	20.9	10.2	9.9	7.5	7.7	6.1	14.5	15.1	11.2
Lignite @ 2.14 \$/GJ, HHV	6.2	6.0	6.6	4.6	7.6	5.5	0.0	0.0	0.0
Biomass @ 3.70 \$/GJ, HHV	3.6	3.5	2.4	2.5	0.6	0.8	14.4	14.4	11.0
CO_2 for EOR @ 25.8 \$/tCO_2	-5.2	-5.1	-6.7	-4.1	-4.9	-3.9	-5.6	-7.3	-4.5
H_2SO_4 @ 133 \$/tonne	-1.0	-1.0	-1.1	-0.8	-1.3	-0.9	0.0	0.0	0.0
For GHG emissions @ \$25/tCO_{2e}									
Electricity @ \$62.3/MWh	-3.4	-6.2	-2.6	0.0	-6.3	0.0	- 5.7	- 2.0	0.0
GHG emissions	0.0	0.0	0.0	0.0	3.9	2.2	- 14.8	- 16.3	- 11.4
Total FTL cost, \$/GJ$_{LHV}$	**80.7**	**38.4**	**40.4**	**33.1**	**30.3**	**28.7**	**44.5**	**49.0**	**39.2**
Total FTL cost, \$/gal gasoline equiv	**9.7**	**4.6**	**4.8**	**4.0**	**3.6**	**3.4**	**5.3**	**5.9**	**4.7**
For GHG emissions @ \$100/tCO_{2e}									
Electricity @ \$89.4/MWh	-4.8	-8.9	-3.8	0.0	-9.1	0.0	- 8.2	- 2.8	0.0
GHG emissions	0.0	0.0	0.0	0.0	15.6	8.9	- 59.1	- 65.2	- 45.7
Total FTL cost, \$/GJ$_{LHV}$	**79.3**	**35.7**	**39.2**	**33.1**	**39.2**	**35.4**	**- 2.3**	**- 0.7**	**4.9**
Total FTL cost, \$/gal gasoline equiv	**9.5**	**4.3**	**4.7**	**4.0**	**4.7**	**4.2**	**- 0.3**	**- 0.1**	**0.6**

The BEOP is perhaps a more intuitive metric than levelized production cost per GJ. The BEOP is the untaxed market price for crude oil that would result in a refiner's wholesale price of equivalent petroleum-derived fuels equal to the levelized FTL production cost. For this calculation, it is assumed that naphtha will have a market energy value equivalent to that for crude oil-derived gasoline. Under this assumption naphtha is likely to be somewhat overvalued when considering market values of crude oil-derived products, but this assumption is not likely to bias the overall economics much, because naphtha accounts for less than 20% of the liquid fuel output, and synthetic paraffinic naphtha is worth more than crude oil-derived naphtha – e.g., because of higher yields of ethylene and propylene. BEOP is plotted against GHG emissions price in Figure 26 for the five NOAK plants having GHGI values of 1 or 0.

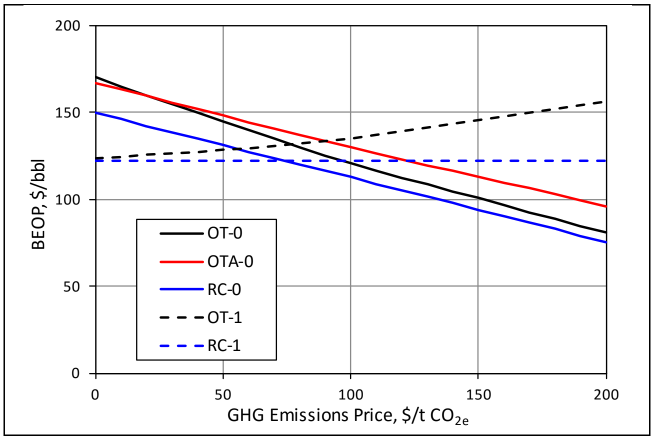


Figure 26. BEOP vs. GHG emissions price for plants with GHGI = 1 and GHGI = 0.

When GHGI = 1, the BEOP curve is either flat (for ***RC-1***) or slightly rising (***OT-1***) with increasing GHG emissions price, but BEOP never falls below about $125/bbl. A flat profile for ***RC-1*** is to be expected, since its output (100% FTL) has (by design) the same GHG emissions as crude oil-derived equivalents. The slightly rising profile of ***OT-1*** arises because the electric power produced by this plant has a carbon intensity equal to that of the reference 2005 U.S. power grid (661 kg CO_{2e}/MWh), which is higher than the value given by the HEP model (natural gas fired GTCC evolving to GTCC-CCS, i.e. 421 → 110 kg CO_{2e}/MWh).

The plants with GHGI = 0 show negative slopes in Figure 26 because their GHG emissions are significantly lower than for petroleum-derived fuels. Among the three GHGI = 0 options, ***RC-0*** offers the lowest BEOP in the absence of a GHG emissions price, but this BEOP is high (about $150/bbl). The economics improve with increasing GHG emissions price: the BEOP reaches $100/bbl for a GHG emissions price of about $135/t$CO_{2e}$. The BEOPs for the 100% biomass plants (Figure 27) are considerably higher than those of the GHGI = 0 plants (Figure 26) when the GHG emissions price is zero or very low. However, because of the strongly negative GHGI values, their BEOPs fall much more quickly with increasing GHG emissions price than for the GHGI = 0 plants, reaching a zero BEOP as the GHG emissions price approaches $100/t$CO_{2e}$ (Figure 26) – i.e., at this price, CO_2 storage via all of these "BECCS" options is worthwhile even if crude oil has zero value.

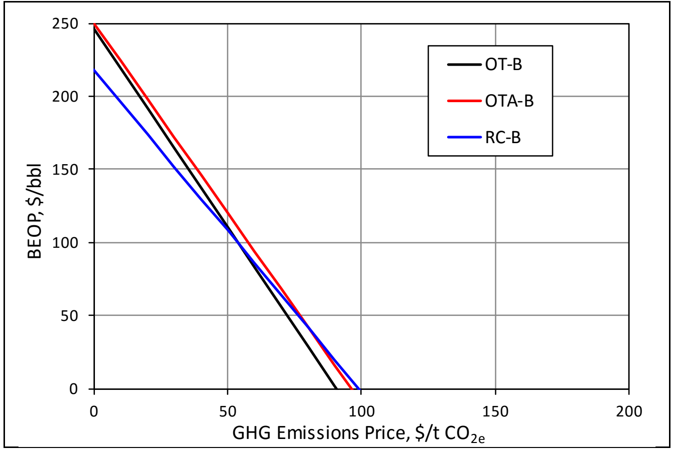


Figure 27. BEOP vs. GHG emissions price for plants with 100% biomass.

The final economic metric is the internal rate of return on invested equity, IRR_e. For an assumed crude oil price, this is the cost of equity (in the framework of the EPRI TAG methodology – see Table 20) that results in a levelized cost of FTL production equal to the (carbon-taxed) price of equivalent crude-oil derived products. IRR_e is plotted against CO$_2$ emissions price in Figure 29 and Figure 30 for the fixed-GHGI plants and the 100% biomass plants, respectively. In both cases, the assumed crude oil price is \$80/bbl. These figures tell a similar story as the BEOP results.

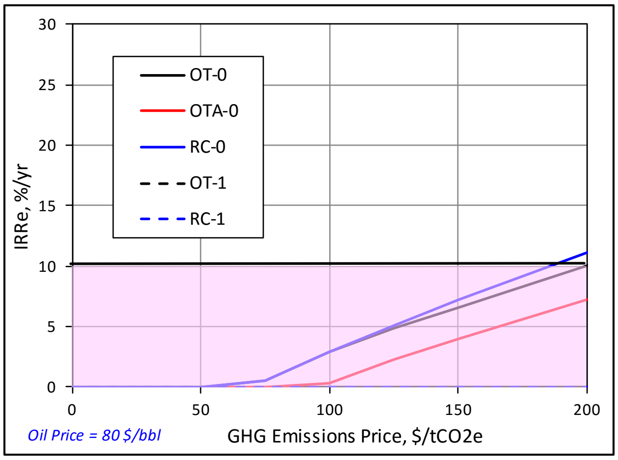


Figure 29. IRR$_e$ vs. GHG emissions price for plants with GHGI = 1 and GHGI = 0. The assumed crude oil price is \$80/bbl. The shaded area represents IRR$_e$ values considered unprofitable using the EPRI TAG methodology.

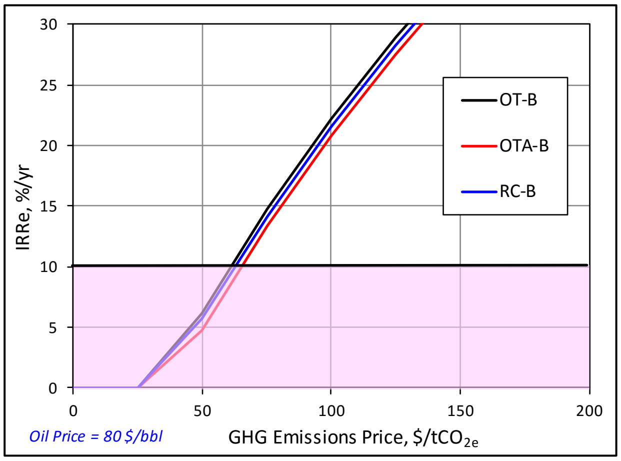


Figure 30. IRR$_e$ vs. GHG emissions price for plants with 100% biomass. The assumed crude oil price is \$80/bbl. The shaded area represents IRR$_e$ values considered unprofitable using the EPRI TAG methodology.

The GHGI = 1 plants do not achieve IRR$_e$ levels that can be considered profitable, regardless of the GHG emissions price. The *GHGI*=0 plants are profitable ($\geq$ 10.2% IRR$_e$) only when the GHG emissions price reaches at least \$130/tonne CO$_{2e}$, at which point the ***RC-0*** plant IRR$_e$ reaches the threshold 10.2% value. IRR$_e$ values for the 100% biomass plants exhibit profitability for GHG emission prices above 55 \$/tonne CO$_{2e}$, and profitability rises rapidly with GHG emissions prices.

All three designs (OT, OTA, and RC) have similar IRR$_e$, although OTA is slightly less profitable than the other two options, for which IRR$_e$ values are nearly identical.

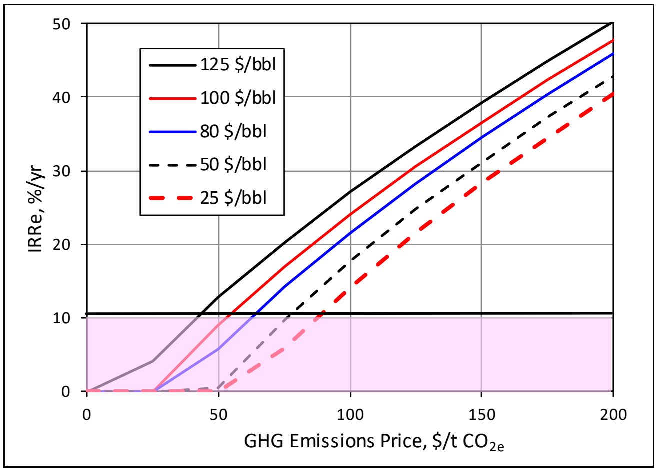


Figure 28. IRR$_e$ vs. GHG emissions price for case *RC-B* at various crude oil prices.

Because OT-B and RC-B offer nearly identical IRR$_e$ values and because creating a viable business model for marketing two products is more challenging than for marketing one, it is suggested that emphasis be given to RC-B as the most promising jet fuel production option based on processing biomass only. Figure 28 presents IRR_e values for RC-B plotted against GHG emissions price for several crude oil prices. For an oil price as low as 25 \$/bbl, RC-B is profitable with a GHG emissions price of \$80/tCO_{2e}. At 125 \$/bbl, viability occurs at about \$50/tCO_{2e}. In all cases, the rate of return rises rapidly with increasing GHG emissions price.

6.5 NOAK Plants with CO_2 Storage in Deep Saline Formations (DSF)

All of the economic results presented thus far assume that captured CO_2 is stored via EOR. The prospective capacity for CO_2 storage via EOR is significant in the U.S.,[75] but it is also of interest to examine the economics of NOAK LBJ systems when captured CO_2 is stored in deep saline formations (DSF), because suitable DSF sites are available across much of the U.S., while EOR opportunities are geographically limited. Additionally, the estimated capacity for CO_2 storage in DSFs is one to two orders of magnitude larger than via EOR,[76] and this greater capacity will likely be needed to achieve ambitious long-term carbon mitigation goals.

To illustrate the impact of DSF storage on the economics of NOAK systems relative to CO_2 EOR applications, we have estimated BEOPs for RC-0 and RC-B systems when CO_2 is stored in DSFs. We estimate CO_2 transport and storage costs for DSF to be about \$17/tCO_2 for RC-0 (storing 4.3 million t/y) and about \$29/tCO_2 for RC-B (storing 1.1 million t/y). These estimates are the sum of separately estimated CO_2 transportation and storage costs. For transportation, we estimate cost per tCO_2 as a function of CO_2 flow rate and transport distance using the algorithm developed by McCollum and Ogden[p] based on their comprehensive review of many different CO_2 transport cost models.[77] We assume a transportation distance of 250 km to be consistent with the "standard" distance assumed in the Rubin *et al*. review of CCS costs.[78] We update the resulting estimate to 2015\$ using the Chemical Engineering Plant Cost Index.[71] For CO_2 storage costs, we adopt for both RC-0 and RC-B the average of high and low costs published by the Global CCS Institute (\$9.6/tCO_2, when converted to \$2015).[79]

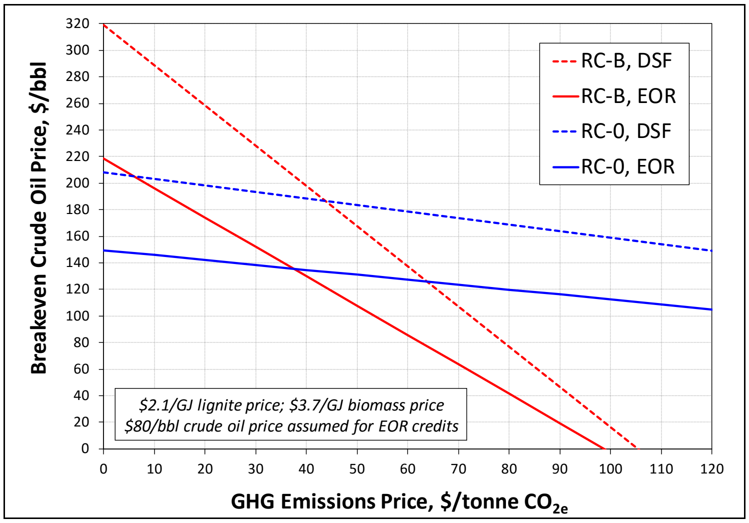


Figure 31. Breakeven crude oil prices for RC-0 and RC-B plants with CO_2 storage in deep saline formations (DSF) or with enhanced oil recovery (EOR).

Figure 31 shows BEOPs for the RC-0 and RC-B systems with DSF storage as a function of GHG emissions price and, for comparison, the BEOPs with EOR storage. BEOP values with DSF are, of course, higher than with EOR. At zero GHG emissions price, the BEOP with DSF storage for the RC-0 case is considerably lower than for RC-B due to the larger scale of RC-0 and the lower average cost per unit input feedstock, but the RC-0 BEOP is still in excess of \$200/bbl. Moreover, the

---

[p] Assuming this McCollum and Ogden algorithm and the financing parameters of the present analysis, the cost of CO_2 transport (in \$2015/t) = $[6.432 \times F^{-0.65} \times L^{1.13}] / CF$, where F = pipeline flow capacity (t/day), L = pipeline length (km), and CF = annual average capacity factor.

BEOP for RC-0 falls only modestly with GHG emission price. Thus, RC-0 is likely to remain an uncompetitive option under foreseeable crude oil prices in the coming decades, irrespective of how successful society is in responding to the carbon mitigation challenge (see Figure 25). Although the BEOP for RC-B with DSF storage is very high at zero GHG emission price, it drops rapidly with increasing GHG emissions price, reaching $80/bbl when the emissions price is about $80/tCO$_{2e}$.

To put this GHG emission price in context, Figure 32 shows CO$_2$ emissions price trajectories from multiple integrated assessment model scenarios assessed in the IPCC's 5th Assessment Report that would achieve different target atmospheric CO$_{2e}$ concentrations by 2100. The dashed lines in the figure indicate corresponding 20-year levelized prices that an RC-B plant starting up in 2031 would experience during its economic lifetime. Based on Figure 32, a levelized CO$_2$ emission price (2031-2050) of $80/t is consistent with limiting the CO$_{2e}$ concentration in 2100 to less than 580 ppm, which corresponds approximately to a global average warming of 3°C above pre-industrial – a less ambitious goal than what was agreed to in Paris. Thus, RC-B and alternative biomass-only plant designs that perform similarly may be viable options for synthetic jet fuel production even in a relatively modestly carbon-constrained future.

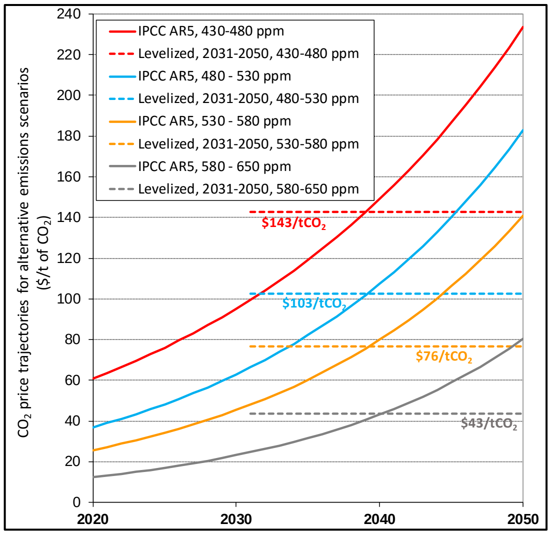


Figure 32. The solid lines represent median estimates of the CO$_2$ emissions price-time trajectories from multiple integrated assessment model scenarios assessed in the IPCC's 5th Assessment Report that achieve the indicated atmospheric CO$_{2e}$ concentrations in 2100. (Based on Figure TS.12 in Edenhofer, *et al.*[80]). The dashed lines indicate corresponding 20-year levelized CO$_2$ prices for 2031-2050, assuming a 7% discount rate.

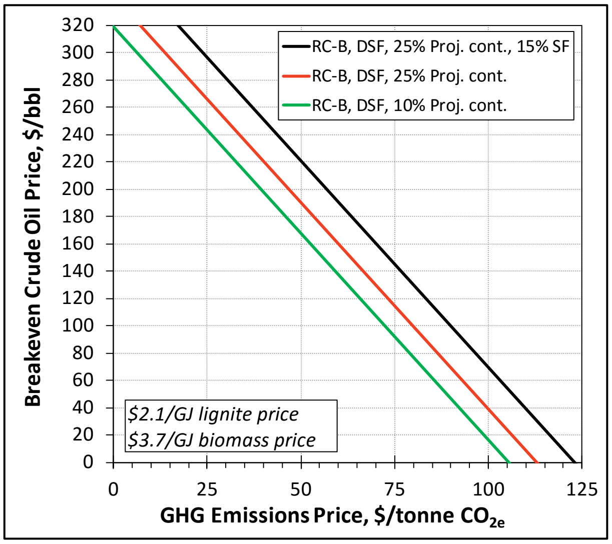


Figure 33. Testing the robustness of BEOP for RC-B with DSF CO$_2$ storage.

Finally, the robustness of the financial performance estimate for RC-B with DSF CO$_2$ storage was tested by estimating BEOP for higher capital costs. In particular, guidelines in Table 13 for Nth-of-a-kind projects, strictly followed, suggest that at the scoping-study stage of project development 25% be allowed for project contingencies and, in cases where supplementary funds are advised, 15% for SF. Applying these factors would give TPC values of 1.37xBEC$_{RC\text{-}B}$ with a 25% project contingency and 1.58xBEC$_{RC\text{-}B}$ with a 25% project contingency and 15% SF. These compare with the 1.22xBEC$_{RC\text{-}B}$ used for the results in Figure 31. With the higher allowances, BEOP values increase (Figure 33), but even the highest TPC case still achieves a BEOP of $80/bbl when the GHG emissions price is less than $\$100/tCO_{2e}$ (a price consistent with limiting global warming to no more than 2 °C – see Figure 32). The value of the strong negative emissions moderates the impact of capital cost on BEOP.

6.6 Comparisons with Other Coal-to-Liquids Design Studies

A number of design/cost studies for large-scale (50,000 bbl/day) coal-to-liquids (CTL) plants employing CCS have been published over the past decade. In an attempt to gain further insights into our results for FOAK and NOAK plants co-processing biomass and lignite, we developed two coal-only NOAK plant designs to compare with other published studies. We assessed the economics of our coal-only designs using the methodology described in Sections 6.4.1 through 6.4.3. We note that our methodology for estimating bare erected costs (BEC) for CTL designs involved scaling up component costs from the FOAK plant design by up to a factor of 17. This introduces more uncertainty than in our BEC estimates for the 100% biomass plants (maximum scale-up by factor of 2) or the GHGI = 0 plants (maximum scale-up by factor of 10), but perhaps less uncertainty than for the GHGI = 1 plants (32x to 50x scale-up).

Our coal-only NOAK designs adopted the RC configuration described in Section 6.1, because this configuration most closely resembles those developed in most other studies. We developed one design using lignite as feedstock and one using sub-bituminous coal from the Powder River Basin (PRB).[q] We refer to these cases as **LIG100** and **PRB100**. Each of these employ oxygen-blown TRIG gasification, Rectisol for acid gas removal, and jet fuel and naphtha production using cobalt-catalyzed tubular fixed-bed FT synthesis reactors. Somewhat less than half the unconverted syngas and refining off-gases are recycled via an ATR to increase liquids production. The rest is used as fuel for the gas turbine combined cycle power island.

We selected three of the more recent (published since 2010) CTL studies for comparison. These include a 2011 study[4] led by the same Princeton University (PU) researchers who led the present study. We label this study **PU-GEE**, because it uses a GE Energy slurry entrained-flow gasifier. We also include comparisons with two studies published by NETL in 2014. One of these is part of NETL's series of Baseline Studies for Fossil Energy Plants, and we refer to it here as **NETL-Shell**, because it uses a Shell gasifier.[2] We refer to the second NETL study[3] as **NETL-TRIG**, because that design uses a TRIG gasfier. Each of the three prior studies includes some distinct differences from the present study.

PU-GEE uses Illinois #6 bituminous coal as feedstock, oxygen-blown slurry-feed entrained-flow GE Energy gasifier, a Rectisol AGR system, iron-catalyzed slurry-bed FT synthesis, and onsite upgrading to produce finished diesel and gasoline. Most of the syngas unconverted in a single pass through the synthesis reactor is recycled through an ATR and mixed with the fresh syngas upstream of the AGR. A recycle-loop purge stream plus refining off-gases fuel a steam Rankine cycle generating electricity to meet onsite needs and export.

---

[q] The composition and heating value assumed for sub-bituminous PRB coal are those in reference 3. Also, the PRB coal was dried from 26% to 18% prior to gasification, as in the plant design in reference 3.

NETL-Shell also uses Illinois #6 coal, but with an oxygen-blown, dry-feed entrained flow Shell gasifier. A Rectisol AGR system is used upstream of an iron-catalyzed, slurry-bed FT synthesis reactor and upgrading that produces diesel and naphtha as final liquid products. Because iron-based FT generates a significant amount of CO_2 (unlike cobalt-based FT), a downstream amine CO_2 removal unit is included in the NETL-Shell design to increase the fraction of input carbon captured. Electricity is generated onsite using off-gases and waste heat recovery to fuel a gas turbine combined cycle. The generated power matches the plant's auxiliary power needs.

NETL-TRIG uses sub-bituminous PRB coal in an oxygen-blown TRIG gasifier, followed by a Selexol AGR system and an iron-catalyzed slurry-bed FT synthesis system that includes a recycle loop with an ATR. An amine CO_2 removal unit is included between the FT reactor and the ATR to capture additional CO_2. The raw FT liquid product is upgraded to produce jet fuel, diesel, naphtha, and LPG. A purge stream of unconverted syngas fuels a GTCC power island generating a significant amount of electricity in excess of onsite needs.

Table 22 shows performance and cost summaries for the above five CTL designs**.** A sixth case, **PRB-TRIG2**, was developed to facilitate comparisons with PRB100. PRB-TRIG2 represents a modified NETL-TRIG design, in which no LPG is produced and instead the LPG-precursor gases are used as additional fuel in the power island.

Despite differences in process designs, PU-GEE, NETL-Shell, NETL-TRIG, and PRB-TRIG2 are generally consistent with each other in terms of overall energy efficiencies and estimated total plant costs. Because PRB-TRIG2 uses the same gasifier technology and same rank of coal as PRB100, we focus here on a more detailed comparison of these two cases. The PRB100 process simulation was developed using the same coal properties as for PRB-TRIG2 and with input coal and FT off-gas recycle rates set such that the total liquids output (2,771 MW_{LHV}) and net electricity output (337 MW_e) exactly matched those of PRB-TRIG2.

Three differences between PRB100 and PRB-TRIG2 stand out in Table 22. The auxiliary electrical load for PRB100 is 66% higher, its bare-erected cost is 70% higher, and its total plant cost is nearly 50% higher. Because of the higher auxiliary load, PRB100 requires a higher fraction of the FT-recycle stream to be diverted to the power island. Additionally, the primary liquid product in PRB100 is SPK, rather than diesel (in PRB-TRIG2), which probably contributes to a lower total liquids yield per unit of syngas in PRB100. Both of these considerations lead to the requirement of greater feedstock input with PRB100 to achieve the same total liquids output as PRB-TRIG2. The larger feedstock input contributes to larger capacity upstream equipment, including the gasifier island, and thereby to a higher bare erected capital cost.

To better understand the large difference in auxiliary loads between PRB100 and PRB-TRIG2, Table 23 shows disaggregated loads. Where cells are blank, no values were given for these line items in the respective study, although in some cases they were included with other line items. For example, in the PRB100 design, the auxiliary load for cooling tower fans was included as part of the cooling water system auxiliaries. A first observation from Table 23 is that of the four largest individual auxiliary loads in PRB100 (coal handling and preparation, ASU, CO_2 compressor, and

AGR) three are significantly higher than for PRB-TRIG2.[r] A second observation is that the sum of all other line items is also considerably higher for PRB-100 (224 MW) than PRB-TRIG2 (114 MW). Auxiliary loads in PRB100 and PRB-TRIG2 were discussed with the lead process design engineer involved in the NETL-TRIG study[81] and led to the following understandings.

Table 22. Comparison of nominal 50,000 bbl/day coal-to-liquids plant designs with CCS.

	PU-GEE	**NETL-Shell**	**NETL-TRIG**	**PRB-TRIG2[a]**	**PRB100**	**LIG100**
Coal type	Illinois #6 bitumin.		PRB sub-bituminous			lignite
Gasifier type	GEE	Shell	TRIG			
FT liquids, actual bbl/d[b]	52,916	49,992	50,000	46,386	45,515	45,514
Coal in, tonne/d, AR	24,087	19,097	27,712	27,712	34,408	59,907
Coal input, MW$_{HHV}$	7,559	5,985	6,375	6,375	7,933	8,532
Liquids out, MW$_{LHV}$	3,159	2,992	2,952	2,771	2,771	2,771
Diesel, MW$_{LHV}$	1,990	2,148	291	291	-	-
SPK/jet fuel, MW$_{LHV}$			1,510	1,510	2,265	
Naphtha, MW$_{LHV}$	1,169	844	970	970	506	506
LPG, MW$_{LHV}$	-	-	181	-	-	-
Gross electricity, MW$_e$	849	427	794	899	1,271	1,407
Auxiliary loads, MW$_e$	554	423	562	562	934	1,071
Net electricity out, MW$_e$	295	5	232	337	337	337
Aux load, % of coal HHV	7.3%	7.1%	8.8%	8.8%	11.8%	12.5%
Liquids eff., LHV$_{liq}$/HHV$_{coal}$	0.42	0.50	0.46	0.43	0.35	0.32
Total eff., (LHV$_{liq}$+El)/HHV$_{coal}$	0.46	0.50	0.50	0.49	0.39	0.36
CO$_2$ captured, tonne/d	29,039	23,970	28,006	28,006	34,589	39,077
Percent of input C	52%	54%	56%	56%	55%	56%
Pure O$_2$ rate, tonne/d	21,634	13,693	16,218	16,218	20,567	22,721
O$_2$ input, t/MW-day coal HHV	2.9	2.3	2.5	2.5	2.6	2.7
FT recycle flow (mass%)	97%	not avail.	54%	52%	47%	41%
Bare Erected Cost, 10^6 2015\$	no est.[c]	3,939	4,394	4,416	6,857	8,015
EPCM & Contingencies	no est.[c]	1,609	1,744	1,753	1,509	1,763
Total Plant Cost, 10^6 2015\$	5,213	5,548	6,138	6,168	8,366	9,778

(a) This is a slight variant of NETL-TRIG developed by the authors of this report to facilitate comparisons with the PRB100 case. NETL-TRIG produces some LPG, whereas PRB100 uses C3 and C4 compounds generated by FT synthesis as part of the fuel gas for the GTCC power island. PRB-TRIG2 adjusts the process design of NETL-TRIG to use the LPG-precursor gases for GTCC power generation (at 58% LHV converted to gross GTCC power[82]). Also, the power island capital cost was up-scaled accordingly using a 0.67 scaling exponent.

(b) These are actual barrels per day, not petroleum-equivalent barrels per day.

(c) The PU-GEE study estimated TPC on a component level basis, without disaggregating BEC, EPCM, or contingencies.

---

[r] Using lignite instead of sub-bituminous coal increases the auxiliary load even more, as can be seen by comparing results for PRB100 and LIG100 in Table 23. Auxiliary loads for solids handling, ASU, AGR, and CO$_2$ compression are all significantly increased as a result of the much larger as-received coal input. The 12% higher coal (HHV) input for LIG100 is explained mainly by two factors: a lower TRIG cold gas efficiency with lignite (82% vs. 79%, HHV basis) due to the higher ash fraction of lignite (22% vs. 11%, dry basis) and the lower FT recycle fraction (41% vs. 47%) which is needed to be able to generate sufficient electricity to meet the larger auxiliary load. The larger coal input necessitates more O$_2$ use in the gasifier, which results in more CO$_2$ being captured, increasing the auxiliary loads for the ASU, AGR, and CO$_2$ compression. The auxiliary load for solids handling is scaled with as-received coal tonnage, which is 82% higher for lignite than for sub-bituminous coal due to higher ash and moisture fractions.

Table 23. Auxiliary loads for (nominal) 50,000 bbl/day coal-to-liquids plant designs.

	PU-GEE	NETL-Shell	NETL-TRIG	PRB-TRIG2	PRB100[a]	LIG100[a]
Coal handling/preparation	21.0	9.9	4.8	4.8	66.0[b]	115.0[b]
Air separation/O_2 & N_2 comp.	324.0	281.2	286.2	286.2	330.6[c]	365.2[c]
CO_2 compressor	109.0	43.5	91.6	91.6	178.4	201.5
Gasifier island			0.5	0.5	0.8	0.9
BFW, circulating water pumps		13.1	18.9	18.9		
Cooling tower fans		3.6	4.9	4.9		
H_2S conversion system		1.8	7.0	7.0	1.9	4.5
Acid gas removal island	40.0	13.6	65.4	65.4	135.1	152.6
Recycle compressor	16.0	2.9			3.8	3.4
FT island	33.0	31.9	13.7	13.7	27.7	27.7
Hydrogen recovery system			48.6	48.6	4.6	4.6
GT fuel gas compressor					57.1[d]	58.3[d]
Tempered water system					2.1	2.3
Cooling water system					51.0[e]	53.9[e]
Water treatment					15.6	16.6
Wastewater treatment					13.0	13.9
Other	11.0	21.0	20.0	20.0	46.7	50.2
Total auxiliary load	**554**	**423**	**562**	**562**	**934**	**1,071**
Auxilliary load per unit processed						
Coal handling/prep, $kWh/t_{coal\ AR}$	21	12	4	4	46	46
Coal handling/prep, $kW/MW_{coal\ LHV}$	2.9	2.1	0.8	0.8	8.6	15.4
ASU and compression, kWh/tO_2	359	493	424	424	386	386
CO_2 compression, $kWh/t\ CO_2$	90	44	78	78	124	124
Acid gas removal, kWh/tCO_2	33	14	56	56	94	94

(a) Except for the coal handling/preparation, ASU, GT fuel gas compressor, and cooling water system, the line-item auxiliary loads were scaled from the FOAK-plant auxiliary load estimates using the following scaling parameters: total AR feedstock HHV (gasifier island, tempered water system, miscellaneous low-voltage users, and sour water stripper – the SWS is included as part of the acid gas removal auxiliary load), CO_2 captured (acid gas removal island and CO_2 compressor), H_2SO_4 product (H_2S conversion system), H_2 used for syncrude hydrocracking (hydrogen recovery system), CO+H_2 feed to FT synthesis (FT island), water handled by zero liquid discharge system (water treatment and wastewater treatment).

(b) The auxiliary load for coal handling/preparation was estimated to be 46 kWh/metric t of AR feedstock, based on an auxiliary load estimate provided to the authors by an Andritz engineer as part of a detailed heat and mass balance for a 1700 metric t/day AR lignite tempered-water/fluid-bed drying system modeled on the one installed at the Kemper County IGCC plant. This estimate does not include lignite crushing (pre-drying) or grinding (post-drying), and thus may underestimate total auxiliary load for coal handling and preparation.

(c) Based on discussions with engineers at Southern Company Services, the ASU auxiliary load was assumed to be 385 kWh/tonne of pure O_2 delivered in an oxidant product having 99.5% purity.

(d) Auxiliary consumption for the GT fuel gas compressor was taken from our Aspen Plus simulation.

(e) Cooling water system auxiliary load was scaled from the FOAK-plant estimate using total AR feedstock HHV as the scaling parameter and then adjusting the load to account for any difference in whole-plant energy efficiency compared with the FOAK plant.

Air Separation Unit. The oxidant provided to the gasifier in PRB100 was 99.5% purity, while 95% purity was used for the PRB-TRIG2 plant. All else equal, the higher purity should require a higher ASU auxiliary load per tonne of O_2 produced. However, other ASU design differences will also affect its auxiliary power requirements. The less-than-10% difference between the specific auxiliary load (kWh/tO_2) for PRB-TRIG2 and PRB100 (shown at the bottom of Table 23) is within the uncertainty range to be expected for the level of engineering design involved in these studies.

Acid Gas Removal. The PRB100 and PRB-TRIG2 plants differ in their AGR equipment. The latter uses a Selexol system upstream of FT synthesis, plus a post-FT amine unit to remove the additional CO_2 generated during iron-catalyzed synthesis. The PRB100 system uses a Rectisol AGR upstream of a cobalt-catalyzed FT unit (and no downstream AGR unit). For equivalent capture fractions (e.g., 90%), a Rectisol system could be expected to have about a 10% higher auxiliary load.[83] In PRB-TRIG2, 90% of available CO_2 was captured, whereas in PRB100 about 97% of the available CO_2 was captured. The higher capture fraction may further contribute to the higher AGR auxiliary load in PRB100. Additionally, the Rectisol design for PRB100 includes a front-end tar washing stage (as discussed in reference to the FOAK design in Section 0), which likely contributes a still further auxiliary load above and beyond that for a "conventional" Rectisol system: the added washing stage increases solvent circulation and cooling requirements. The differences in AGR auxiliary load between PRB100 and PRB-TRIG2 may be largely explained by the above three factors. It remains a question, however, whether or not a Selexol system would provide adequate cleanup for a cobalt-catalyzed FT synthesis reactor.

CO_2 Compression. The difference in power required for CO_2 compression between PRB100 and PRB-TRIG2 can be attributed to three factors. First, the total amount of CO_2 delivered to the plant gate at 150 bara in the PRB100 case was larger (see Table 22). Second, the average initial pressure of CO_2 prior to pressurization to 150 bara is lower in PRB100, necessitating more compression work to achieve the same CO_2 delivery pressure.[s] The average initial CO_2 pressure is lower in PRB100 in part because the higher CO_2 recovery fraction in the AGR system is achieved by flashing the CO_2-containing solvent to lower pressure. Third, the CO_2 compressor in the PRB-TRIG2 design uses relatively aggressive intercooler temperature differences that may create operating difficulties, and moderating this aspect of the design would increase its specific power consumption by about 10% over that shown in Table 23.[81]

Coal handling and preparation. The difference in auxiliary load estimates for coal handling and preparation between PRB100 and PRB-TRIG2 is difficult to reconcile. The PRB100 estimate was arrived at by scaling with tonnage received an estimate of 46 kWh/tonne of as-received (AR) lignite provided by an Andritz design engineer as part of a detailed heat and mass balance for a 1700 tonne/day AR lignite drying system (prepared for the authors of this report) that represented a scaled-down version of the tempered-water/fluid-bed lignite drying system installed at the Kemper County IGCC plant. The scaled-down design was for drying lignite from 45.5% to 20% moisture content. The 46 kWh/t AR lignite did not include additional auxiliary load that would be required for crushing the coal before drying, milling after drying, nor intermediate conveying steps, and it

---

[s] In PRB100, the initial pressures of two streams of CO_2 available for pressurizing to 150 bar for export are 1.1 bara (88% of the exported CO_2) and 3.1 bara (12%). In PRB-TRIG2, there are four CO_2 streams with initial pressures of 1.4 bara (12%), 1.7 bara (39%), 4.1 bara (29%) and 12.4 bara (20%). Self-consistent Aspen Plus calculations were performed to estimate the auxiliary power required to pressurize each of these two sets of CO_2 flows to 150 bara. The resulting auxiliary loads for PRB100 and PRB-TRIG2 were 104 and 83 kWh/tCO_2 delivered, respectively.

thus probably underestimates the total auxiliary requirements for coal handling and preparation.[t] However, the 46 kWh/t AR lignite is consistent with results of a detailed study of fluidized-bed drying of lignite for a 360 MW_e(net) combustion power plant. In that study, lignite was dried from 55% to 12% moisture content using a RWE's WTA drying system.[u] The auxiliary load for the WTA system was 52 kWh/tonne of AR lignite.[84]

For PRB-TRIG2, the auxiliary load for coal handling and preparation was estimated as the sum of separate estimates for transport/conveying, grinding, and fluidized-bed drying.[81] The transport and conveying estimate was based on NETL Baseline power studies for bituminous coal and so may underestimate this component of the auxiliary load for PRB coal. Power for grinding to 600 to 800 micron particle size was based on empirical correlations for power versus particle size derived from studies at the Energy Research Center of the Netherlands (ECN). Fluidization power for drying PRB coal from 26% to 18% moisture content was based on an estimate of auxiliary power requirements of the WTA process. However, given the WTA estimate cited earlier, the PRB-TRIG2 auxiliary load for coal handling and preparation seems to be underestimated. That this is the case is also suggested by a comparison of loads for PRB-TRIG2 (4 kWh/t AR coal) and NETL-Shell (12 kWh/t AR coal). The latter uses bituminous coal that dries from 11% to 5% moisture content as it is milled for feeding to an entrained flow gasifier.

Balance of auxiliary loads. The balance of the auxiliary loads, i.e., all except ASU, AGR, CO_2 compressor, and coal handling, represent 224 MW (24% of total) and 114 MW (20% of total) for PRB100 and PRB-TRIG2, respectively. A large contributor to the 110 MW difference between these estimates is the GT fuel gas compressor (57 MW) used in PRB100: no GT fuel compressor was used in the PRB-TRIG2 design.[v] Also, the PRB-TRIG2 auxiliary load estimate did not factor in wastewater treatment, and the PRB100 waste water treatment includes a zero liquid discharge (ZLD) system, which increases the auxiliary load relative to a non-ZLD wastewater treatment system.

Summary of auxiliary load comparisons. The analysis above indicates that a significant portion of the difference in total auxiliary load for the PRB100 and PRB-TRIG2 plants arises from differences in equipment selections made during plant design, although these do not fully explain the difference in total auxiliary load between the two plants. The choice of a cobalt-FT reactor was made for PRB100 because this was the only option offered (for the FOAK plant) by the multiple commercial vendors who were approached for quotes for the FOAK plant design. This led to the choice of Rectisol for the AGR to ensure protection of the expensive FT catalyst. In turn, the supplier of the design for the Rectisol system included a front-end tar washing stage necessitated by

---

[t] For comparison, the auxiliary load for coal handling and preparation estimated by WorleyParsons for the FOAK plant was 9 kWh/t of AR (45.5% moisture) lignite (see Table 1 in Appendix A), or only about 20% of the level estimated for lignite drying alone by the Andritz engineer. In the final design report for the FOAK plant prepared by WorleyParsons (Appendix A), insufficient detail is provided in support of this auxiliary load estimate to allow an assessment of its accuracy, and so we chose to use the Andritz-based estimate, despite it being limited to the auxiliary load only for drying.

[u] WTA is a fluidized-bed process that incorporates recovery of the latent energy of water evaporated from the coal to increase overall efficiency.

[v] The PRB2-TRIG design assumes that fuel gas from the FT synthesis reactor without a pressure boost would be adequate for reliable operation of the power island. The PRB100 design assumes that the fuel gas pressure is boosted to 1.6 times the GT combustor operating pressure, as was done in the WorleyParsons design study (Appendix A) and which is in line with the 1.45x pressure boost specified by Siemens for the SGT-700 GT firing natural gas. (The SGT-700 was the basis for the GT in the FOAK plant design.)

the risk of trace tar levels arriving in the syngas from a TRIG gasifier, leading to a higher than "normal" Rectisol auxiliary load. Additionally, the high CO_2 recovery fraction specified for the Rectisol system contributed to a relatively low CO_2 recovery pressure that required increased CO_2 compressor work. Also, unlike for PRB-TRIG2, PRB100 included a GT fuel gas compressor (to ensure sufficient feed pressure for fuel to the GT) and a ZLD wastewater treatment system, each with its attendant auxiliary load. The auxiliary load for coal handling and preparation for PRB100 was difficult to reconcile with the much lower auxiliary load estimated for PRB-TRIG2. The latter latter appears to be substantially underestimated, especially in light of the fact that the auxiliary load for coal handling and preparation as a fraction of the coal input energy is much less than for bituminous coal (NETL-Shell case).

In conclusion, this analysis suggests careful attention should be paid in future studies to quantifying auxiliary loads for synfuels plants, especially those using low-rank coals. Accurate estimates of auxiliary loads will provide a better understanding of prospective technical and economic performance. Some R&D emphasis on reducing auxiliary loads is also recommended, since this will help increase overall plant efficiencies and economics. Although historically auxiliary load reduction has been considered a priority for R&D mainly so as to improve major conventional FTL process components (e.g., ASU and AGR units), the findings of this study suggest that additional R&D is needed to address the auxiliary load requirements associated with the ongoing trend toward more stringent environmental performance (e.g, lower NOx emissions, zero water discharge, and, in the future, lower carbon footprints) and interest in making synfuels from low-rank coals.

7 Conclusions

This study described and assessed the process design of a first-of-a-kind (FOAK) demonstration plant that could be constructed today to coprocess lignite and woody biomass in a 75/25 mix (energy basis) into synthetic Fischer-Tropsch jet fuel while capturing byproduct CO_2 for sale for enhanced oil recovery (EOR).

Realization of carbon mitigation benefits using forest biomass for energy is challenging because of the long time required to generate as much new biomass from a new tree as in the harvested tree (in contrast to herbaceous biomass, for which harvested biomass is replenished within a single year). This challenge can be overcome by pursuing forest-management strategies aimed at enhancing forest productivity. This study showed that jet fuel derived from lignite and biomass with CCS can have a lower greenhouse gas (GHG) footprint than crude oil-derived jet fuel if well-understood productivity enhancement strategies are implemented on sustainably managed pine stands in the Southeastern U.S. from which the plant's biomass supply would originate; the identified productivity enhancement approaches ensure that additional biomass is produced from these stands compared to "business-as-usual" management practices.

For the design of the FOAK LBJ facility, the approach taken was to make the plant large enough to give investors confidence in the performance of subsequent commercial-scale plants but small enough that the plant might actually be financed and built – even though it was expected *a priori* that such a project would not be undertaken by a private firm without a substantial government subsidy. (Indeed, without a subsidy, the levelized cost of producing synthetic jet fuel for the 1,250 bbl/day FOAK LBJ plant approaches $500/bbl.) Even if a small-scale FOAK demonstration plant is uneconomic, its construction with heavy government subsidy might be worthwhile if subsequent commercial-scale plants have good economic prospects. Accordingly, this project analyzed the economic prospects for future Nth-of-a-kind (NOAK) commercial-scale plants (LBJ design and several variations thereof) to provide a basis for understanding better whether building a FOAK demonstration-scale plant is worthwhile.

NOAK (as well as FOAK) installed costs were estimated from bare erected costs (BEC) by considering how BEC multipliers that take into account how contingencies, supplemental fund requirements, and other ancillary costs have tended to vary for "megascale" projects – as a function of both technological maturity (for a FOAK, early-mover, or NOAK plant) and level of analytical effort (concept study, scoping study, pre-feasibility study, feasibility study, or analysis carried out when all engineering has been completed). No attempt was made to estimate the rate of cost reduction through experience ("learning by doing") and thus the number of plants that would have to be built to reach NOAK cost levels for commercial plants. Moreover, the NOAK cost estimates arrived at might be considered optimistic. Nevertheless, the estimated NOAK costs can be very useful for policymakers, because they shed light on the question: Is building a small demonstration-scale LBJ plant (or some variant thereof) worthwhile?

For the preliminary assessment of future NOAK commercial plants presented here, advanced (not currently commercial) technologies and R&D-driven improvements in existing technologies were not considered. Instead, the focus was on implications for performance and economics of *(i)* increased plant scales, *(ii)* capital and operating cost reductions resulting from "learning by doing", *(iii)* alternative process configurations, *(iv)* alternative lignite:biomass input ratios, and *(v)* how the economics of systems will vary with crude oil price and value of GHG emissions (the two most important exogenous parameters influencing NOAK economics).

A major finding is that NOAK plants that coprocess lignite and biomass to make jet fuel and have zero or higher net greenhouse gas emissions are unlikely to be economically competitive based on any plausible combinations of future crude oil and greenhouse gas emissions prices. This finding arises in large part because even for NOAK plants capital costs are high, in large part because of high auxiliary power needs – suggesting the strategic importance of R&D aimed at reducing plant auxiliary loads.

In contrast, encouraging economic results were found for plants processing only biomass in the presence of a strong carbon mitigation policy – even though such plants were also found to have large auxiliary power needs. The economics of these "BECCS" (biomass energy with CCS) plants improve dramatically with the strength of carbon mitigation policies because of their strongly negative net greenhouse gas emissions. Pure biomass synfuel plants were found to be economically interesting at crude oil and GHG emissions prices consistent with carbon mitigation goals of limiting global warming to no more than 2^{o}C to 3^{o}C – not only for CO_2 EOR applications, but also when captured CO_2 is stored in deep saline formations, and even in the absence of "optimistic" NOAK costing assumptions. Although the analysis shows that such systems are unlikely to be economic in the absence of a strong carbon mitigation policy, future R&D-driven technological innovations could modify this conclusion.

These findings do not imply that coal/biomass coprocessing strategies for making synfuels with CCS are not economically promising – only that, in the case of lignite, much more than a 25% biomass coprocessing rate would be needed. Such lignite/biomass coprocessing options, as well as options coprocessing biomass with higher ranked coals (which will generally have lower auxiliary loads than with lignite) should be investigated in future studies.

Appendices (in separate document)

Appendix A: WorleyParsons Group, Process Design/Cost Study…

Brubaker, D., Goldstein, H., Mace, E., Pinkerton, L., and Xie, Q. (WorleyParsons Group), *Design/Cost Study for Synthetic Fuel Production from Lignite and Woody Biomass*, prepared for Princeton University's Energy Systems Analysis Group by WorleyParsons Group, Reading, PA, September 2016.

Appendix B: Air Liquide, Rectisol Study…

Princeton University Rectisol Feasibility Study, study number 37750-40, prepared by Air Liquide Global E&C Solutions, Frankfurt am Main, Germany, October 2015.

Appendix C: Emerging Fuels Technology, Fischer-Tropsch Study…

Process Design and Cost Estimate for Fischer-Tropsch Synthesis and Upgrading Equipment, prepared for Princeton University's Energy Systems Analysis Group by Emerging Fuels Technology, August 2015.

Appendix D: Politecnico di Milano, Plant Heat Integration Study…

Elsidoa, C. and Martelli, E. (Politecnico di Milano), and Kreutz, T. (Princeton University), *Heat Integration, Heat Exchanger Network Synthesis and Heat Recovery Steam Cycle Optimization for a CBTLE Demonstration Plant*, October 2016.

Appendix E: Air Products and Chemicals, Air separation unit quote…

Luff, B.A., *Budgetary response for the supply of an Air Separation Unit for the Princeton / DOE coal and biomass to jet fuel project*, provided to WorleyParsons Group by Air Products and Chemicals, Allentown, PA, January 2016.

Appendix F: Dresser-Rand, CO2 compressor quote…

Reif, P., *Budgetary Technical Proposal for WorleyParsons/Princeton University DOE Coal and Biomass to Jet Fuel Project,* Client Ref: 108103-08256, Dresser Rand Corporation, January 2016.

Appendix G: Simulation details for NOAK plant analysis

Kreutz, T., *Details Relating to NOAK Plant Analysis for Synthetic Jet Fuel Production from Lignite and Woody Biomass with CO2 Capture and Storage via EOR*, April 26, 2017.

References

1. Tarka, T.J., “Affordable, Low-Carbon Diesel Fuel from Domestic Coal and Biomass,” DOE/NETL-2009/1349, National Energy Technology Laboratory, 14 January 2009.

2. V. Shah, N.J. Kuehn, M.J. Turner, and S.J. Kramer, "Cost and Performance Baseline for Fossil Energy Plants, Vol 4: Coal-to-Liquids via Fischer-Tropsch Synthesis," DOE/NETL-2011/1477, 15 Oct 2014.

3 Skone, T., T. Eckard, J. Marriott, G. Cooney, J. Littlefield, C. White, D. Gray, J. Plunkett, and W. Smith, "Comprehensive Analysis of Coal and Biomass Conversion to Jet Fuel: Oxygen Blown, Transport Reactor Integrated Gasifier (TRIG) and Fischer-Tropsch (F-T) Catalyst Configurations," DOE/NETL-2012/1563, 19 February 2014.

4. Liu, G., E. Larson, R. Williams, T. Kreutz, X. Guo, “Making Fischer-Tropsch Fuels and Electricity from Coal and Biomass: Performance and Cost Analysis,” *Energy & Fuels*, 25(1): 415-437, 2011.

5. Larson, E.D., R.H. Williams, and T.G. Kreutz, Energy, Environmental, and Economic Analyses of Design Concepts for the Co-Production of Fuels and Chemicals with Electricity via Co-Gasification of Coal and Biomass, final report to NETL under DE-FE0005373, 30 June 2012.

6. Williams, R.H., “Coal/biomass coprocessing strategy to enable a thriving coal industry in a carbon-constrained world,” *Cornerstone*, 1(1): 51-59, 2013

7. National Coal Council, Harnessing Coal’s Carbon Content to Advance the Economy, Environment, and Energy Security, Washington, DC, 22 June 2012.

8. America’s Energy Futures Panel on Alternative Liquid Transportation Fuels, *Liquid Transportation Fuels from Coal and Biomass: Technological Status, Costs, and Environmental Impacts*, National Research Council, 2009.

9. Cool Water Gasification Program and Radian Corporation, Cool Water Coal Gasification Program: Final Report, GS-6806, Electric Power Research Institute, December 1990.

10. Southern Company Services, Inc., “Test Run Summary Report: Test Run R08, June 25 - Sept. 18, 2012,” under DOE Cooperative Agreement DE-NT0000749, December 2012.

11. Southern Company Services, Inc., “Test Run Summary Report: Test Run R04, April 1-28, 2010,” under DOE Cooperative Agreement DE-NT0000749, June 2010.

12. Antares Group Inc., “Conceptual Design Review for Biomass Repowering at Plant Sweatt,” draft final report, Electric Power Research Institute, October 2010.

13. R.H. Williams, “Capture technology cost buydown in CO_2 EOR market applications under an Alternative Energy Portfolio Standard,” *Energy Procedia*: 12th International Greenhouse Gas Control Technologies Conference (Austin, TX, October 2014), forthcoming.

14. This availability data are shown in H. Meerman, “Perspectives on gasification systems to produce energy carriers and other chemicals with low CO_2 emissions: Techno-economic system analysis on current and advanced flexible thermo-chemical conversion of fossil fuels and biomass,” PhD thesis, Utrecht University, Netherlands, 2012. (Meerman cites as the source of these data J. Phillips, Integrated Gasification Combined Cycle (IGCC) Design Considerations for High Availability, Vol. 1 – Lessons from Existing Operations, Rpt 1012226, EPRI, March 2007.)
15. Tipton, Harry B. (“Tres”), Vice President of Engineering for Louisiana and Mississippi Operations, North American Coal Corporation, personal communications, February 2014.
16. Rooney, T. and Gray, E. “Biomass Supply and Biogenic Carbon Impacts Evaluation for Lignite-Biomass to Jet Fuel Project, Final Report,” Antares Group Inc., 9 January 2016.
17. Antares Group, “Plant Sweatt Biomass Assessment Final Report,” MPC, Southern Co, 25 Mar 2009.
18. Comer, K. (Antares Group, Inc.), Conceptual Design Review for Biomass Repowering at Plant Sweatt, Report 1022382, Southern Company Services (Brimingham, AL) and Electric Power Research Institute (Palo Alto, Calif.), December 2010.
19. Rooney, T. Schmidt, A., and Gray, E., “Forest Data Analysis: Overview of Analytical Approach,” slides presented by Antares Group to Southern Co. Services and Princeton LBJ project team, 26 July 2016.
20. Gray, E., “Southern Company Services (SCS) Continuing Technical Assistance to the LBJ Project: Data Sources and Methods,” (accompanying reference 19), 22 September 2016.
21. Nelson, M., Rush, R., Madden, D., Pinkston, T., and Lunsford, L., “Kemper County IGCC^{TM} Project Preliminary Public Design Report,” prepared for US Department of Energy under award number DE-FC26-06NT42391 by Southern Company Services, Birmingham, AL, 30 June 2012.
22. Southern Company Services, “Test Run Summary Report: Test Run R08, June 25, 2012 – September 18, 2012,” under DOE Cooperative Agreement DE-NT0000749, December 2012.
23. Up to date Kemper County project citation.
24. Pinkston, T., “Kemper County IGCC Project Update,” *2013 Gasification Technologies Conference*, October 2013.
25. Liu, G., Larson, E.D., Williams, R.H., Kreutz, T.G., Guo, X., “Making Fischer-Tropsch fuels and electricity from coal and biomass: performance and cost analysis,” *Energy and Fuels*, 25: 415-437, 2011.
26. U.S. “Intended nationally determined contribution” to global greenhouse gas emissions reductions, as reported to the United Nations Framework Convention on Climate Change.
27. Office of the Press Secretary, “Fact Sheet: U.S. Reports its 2025 Emissions Target to the UNFCCC,” The Whitehouse, Wash., DC, March 31, 2015. URL
28. Skone, T., “LCA at the Department of Energy (DOE), National Energy Technology Laboratory (NETL),” presentation, LCA XII, Tacoma, WA, 27 September 2012.
29. Skone, T., “The Challenge of Co-product Management for Large-scale Energy Systems: Power, Fuel and CO2,” presentation, LCA XIII, Orlando, FL, 2 October 2013.
30. Xie, X., Wang, M., and Han, J., “Assessment of Fuel-Cycle Energy Use and Greenhouse Gas Emissions for Fischer-Tropsch Diesel from Coal and Cellulosic Biomass,” *Environ. Sci. Technol.*, (45): 3047-3053, 2011.
31. Pietro, P. "Life Cycle Analysis of Coal and Natural Gas-fired Power Plants," presented at EPRI Coal Fleet, 20 July 2010.
32. G. Cooney, personal communication (22 October 2014) of details behind results published by Skone, *et al.*[33]

33. Skone, T.J., Eckard, R., Marriott, J., Cooney, G., Littlefield, J., White, C., Gray, D., Plunkett, J., Smit, W., *Comprehensive Analysis of Coal and Biomass Conversion to Jet Fuel: Oxygen-Blown, Transport Reactor Integrated Gasifier (TRIG) and Fischer-Tropsch (F-T) Catalyst Configurations*, DOE/NETL-2012/1563, National Energy Technology Lab, 19 February 2014.

34. Skone, T. and Gerdes, K., *Development of Baseline Data and Analysis of Life Cycle Greenhouse Gas Emissions of Petroleum-Based Fuels*, (and companion spreadsheet model, 2009), DOE/NETL-2009/1346, National Energy Technology Lab, 26 November 2008.

35. The Aviation Fuel Life Cycle Assessment Working Group (T.J. Skone, Lead, D.T. Allen, C.Allport, K. Atkins, D. Baniszewski, D.G. Choi, J.S. Cooper, A. Curtright, R.M. Dilmore, R.S. Eckard, A. Elgowiny, W. Gillette, W.M. Griffin, W.E. Harrison III, J.I. Hileman, S. Kennedy, E. Larson, A. Levy, K. Lewis, J. Marriott, D.J. Morgan, C.F. Murphy, M.A. Nippert, M. Pearlson, G. Rhoads, G. Schivley, R.W. Stratton, T.Tarka, P.H. Taylor, V.M. Thomas, M.Q. Wang, and H. Willis), *Life Cycle Greenhouse Gas Analysis of Advanced Jet Propulsion Fuels: Fischer-Tropsch Based SPK-1 Case Study*, AFRL-RZ-WP-TR-2011-2138, Air Force Research Lab, Propulsion Directorate, Wright-Patterson Air Force Base, Ohio, September 2011.

36. Mariott, J., "A Parameterized Life Cycle Analysis of Crude from CO_2-Enhanced Oil Recovery," presentation, National Energy Technology Lab, 2 October 2013.

37. Skone, T.J., James, R., Littlefield, J., Cooney, G., Jamieson, M., Schivley, G., and Marriott, J., *Cradle-to-Gate Life Cycle Analysis Model for Alternative Sources of Carbon Dioxide*, DOE/NETL-2013/1601, National Energy Technology Laboratory, 30 September 2013.

38. National Energy Technology Laboratory, *Life Cycle Inventory Data – Unit Process: CO_2 Compression*, U.S. Department of Energy, NETL, last updated July 2012 (version 01).

39. EPA, "The Emissions & Generation Resource Integrated Database (eGRID)," US Environmental Protection Agency. URL.

40. ANL, "The Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation Model," Argonne National Laboratory. URL.

41. Walker, T., P. Cardellichio, A. Colnes, J. Gunn, B. Kittler, B. Perschel, C. Recchia, and D. Saah, "Biomass Sustainability and Carbon Policy Study," for the Commonwealth of Massachusetts Department of Energy Resources by Manomet Center for Conservation Sciences, Manomet, Massachusetts, 2010.

42. Hanson, C., Yonavjak, L., Clarke, C., Minnemeyer, S., Boisrobert, L., Leach, A., and Schleeweis, K., *Southern Forests for the Future*, World Resources Institute, 2010.

43. Dickens, D., Moorhead, D.J., and McElvany, B., "Pine plantation fertilization," *Better Crops*, 87(1): 12-15, 2003.

44. Office of Air and Radiation, *Framework for Assessing Biogenic CO2 Emissions from Stationary Sources*. U.S. Environmental Protection Agency, Washington, DC, November 2014.

45. Siry, J. P., "Potential Impacts of Increased Management Intensities on Planted Pine Growth and Yield and Timber Supply Modeling in the South," *Proceedings of the 1999 Southern Forest Economics Workshop*. April 18-20 Biloxi, MS, 1999.

46. Mississippi State University Forest and Wildlife Research Center, *Cutover Loblolly Growth/Yield Model*, undated.

47. EPA, "Chapter 14: Greenhouse Gas Biogenic Sources," *AP 42, Fifth Edition Compilation of Air Pollutant Emission Factors, Volume 1: Stationary Point and Area Sources, 1996.*

48. Johnston, A., "Nitrous Oxide Emissions from Fertilizer Nitrogen," *News and Views*, Potash & Phosphate Institute newsletter, February 2005. URL

49. Merrow, E.W., Phillips, K.E., and Myers, C.W., *Understanding Cost Growth and Performance Shortfalls in Pioneer Process Plants*, for DOE (DE-AC01-79PE79978) by Rand Corp., Santa Monica, CA, 1981.

50. Greig, C., Garnett, A., Oesch, J., and Smart, S., *Guidelines for Scoping and Estimating Early Mover CCS Projects*, Milestone 5 Final Report, The University of Queensland, 19 June 2014. URL.

51. Horvath, R.E., *Pioneer Plants Study User's Manual*, R-2569/1-DOE, prepared for U.S. Department of Energy, Rand, Santa Monica, CA, June 1983. URL.

52. Downloaded from http://www.gurufocus.com/, 11 June 2016.

53. Hailey, A.K., Meerman, J.C., Larson, E.D., and Loo, Y-L., "Low-carbon "drop-in replacement" transportation fuels from non-food biomass and natural gas," *Applied Energy*, DOI: 10.1016/j.apenergy.2016.09.068, Oct 2016.

54. Energy Information Administration, *Annual Energy Outlook 2015*, US DOE, April 2015.

55. Rubin, E., Davison, J. and Herzog, H., "The cost of CO_2 capture and storage," *Int. J GHG Control*, **40**: 378-400, 2015.

56. Argus, "FMB North American Sulphur and Sulphuric Acid," Issue 14-29, Argus Media, 17 July 2014.

57. Larson, E.D., Fiorese, G., Liu, G., Williams, R.H., Kreutz, T.K., and Consonni, S. "Co-production of decarbonized synfuels and electricity from coal + biomass with CO_2 capture and storage: an Illinois case study," *Energy & Environmental Science*, 3: 28-42, 2010.

58. Chen, C. (Southern Company Services), personal communication with E. Larson, September 2016.

59. Farmer, R. (Editor-in-Chief), *2012 Gas Turbine World Handbook*, Pequot Publishing, Fairfield, Connecticut, 2012.

60. Consonni, S. and Vigano, F., "Decarbonized hydrogen and electricity from natural gas", *Int. J. Hydrogen Energy*, 30: 701-718, 2005.

61. Grabke, H.J., Krajak, R., and Müller-Lorenz, E.M., "Metal dusting of high temperature alloys," *Werkstoffe und Korrosion* **44**(3), 89–97, 1993.

62. Grabke, H.J. and Spiegel, M., "Occurrence of metal dusting – referring to failure cases," *Materials and Corrosion*, **54**(10): 799–804, 2003.

63. Livia Sierra Llorens (Haldor Topsoe), personal communication to Eric Larson, 22 January 2016.

64. Clement Celestin (Siemens), personal communication to Harvey Goldstein (WorleyParsons Group), 14 July 2015.

65. Chiesa, P. and Lozza, G., "Using Hydrogen as a Gas Turbine Fuel", *ASME Journal of Engineering for Gas Turbines and Power*, 127(1): 73–81, 2005.

66. Gazzani, M., Chiesa, P., Martelli, E., Sigali, S., and Brunetti, I., "Using Hydrogen as Gas Turbine Fuel: Premixed Versus Diffusive Flame Combustors," *ASME Journal of Engineering for Gas Turbines and Power*, 136(5): 051504-1 – 051504-10, 2014.

67. Martelli, E., Kreutz, T.G., Carbo, M., Consonni, S., and Jansen, D. "Shell Coal IGCCs with Carbon Capture: Conventional Gas Quench vs. Innovative Configurations," *Applied Energy*, 88: 3978-3989, 2011.

68. Oluyede, E.O. and Phillips, J.N., "Fundamental impact of firing syngas in gas turbines," *Proceedings of ASME Turbo Expo 2007: Power for Land, Sea and Air*, Vol. GT2007-27385, 2007.

69. Turner, M.J. and Pinkerton, L.L., *Quality Guidelines for Energy System Studies: Capital Cost Scaling Methodology*, DOE/NETL-341/013113, National Energy Technology Laboratory, 31 January 2013.

70. Vatavuk, W.M., "Updating the CE plant cost index", *Chemical Engineering*, pp. 62–70, Jan. 2002.

71. *Chemical Engineering Plant Cost Index*, published monthly. See http://www.chemengonline.com/pci.
72. Electric Power Research Institute, *Technical Assessment Guide (TAG^{TM}) Electricity Supply*, TR-102276-V1R7, EPRI, Palo Alto, CA, 1993.
73. NETL, *Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity Final Report*, Revision 1, DOE/NETL-2007/1281, National Energy Technology Laboratory, August 2007.
74. International Energy Agency, *World Energy Outlook 2016*, Paris, France, 2016.
75. Kuuskra, V. and Ferguson, R., "Storing CO_2 with Enhanced Oil Recovery," DOE/NETL-402/1312/02-07-08, National Energy Technology Laboratory, 7 February 2008.
76. National Energy Technology Laboratory, *Carbon Storage Atlas*, 5th Edition, Office of Fossil Energy, US Department of Energy, 2015.
77. McCollum, D.L. and Ogden, J.M., *Techno-Economic Models for Carbon Dioxide Compression, Transport, and Storage & Correlations for Estimating Carbon Dioxide Density and Viscosity*, UCD-ITS-RR-06-14, Institute for Transportation Studies, University of California, Davis, 1 October 2016.
78. Rubin, E.S., Davison, J.E., and Herzog, H.J., "The Cost of CO_2 Capture and Storage," *International Journal of Greenhouse Gas Control*, 40: 378-400, 2015
79. WorleyParsons Services and Schlumberger Carbon Services, *Economic assessment of carbon capture and storage technologies: 2011 update*, prepared for the Global CCS Institute, Canberra, Australia, 2011.
80. Edenhofer O., R. Pichs-Madruga, Y. Sokona, S. Kadner, J.C. Minx, S. Brunner, S. Agrawala, G. Baiocchi, I. A. Bashmakov, G. Blanco, J. Broome, T. Bruckner, M. Bustamante, L. Clarke, M. Conte Grand, F. Creutzig, X. Cruz-Núñez, S. Dhakal, N.K. Dubash, P. Eickemeier, E. Farahani, M. Fischedick, M. Fleurbaey, R. Gerlagh, L. Gómez-Echeverri, S. Gupta, J. Harnisch, K. Jiang, F. Jotzo, S. Kartha, S. Klasen, C. Kolstad, V. Krey, H. Kunreuther, O. Lucon, O. Masera, Y. Mulugetta, R.B. Norgaard, A. Patt, N.H. Ravindranath, K. Riahi, J. Roy, A. Sagar, R. Schaeffer, S. Schlömer, K.C. Seto, K. Seyboth, R. Sims, P. Smith, E. Somanathan, R. Stavins, C. von Stechow, T. Sterner, T. Sugiyama, S. Suh, D. Ürge-Vorsatz, K. Urama, A. Venables, D.G. Victor, E. Weber, D. Zhou, J. Zou, and T. Zwickel, *Technical Summary. In: Climate Change 2014: Mitigation of Climate Change*, Contribution of Working Group III to the Fifth Assessment Report of the Intergovernmental Panel on Climate Change [Edenhofer, O., R. Pichs-Madruga, Y. Sokona, E. Farahani, S. Kadner, K. Seyboth, A. Adler, I. Baum, S. Brunner, P. Eickemeier, B. Kriemann, J. Savolainen, S. Schlömer, C. von Stechow, T. Zwickel and J.C. Minx (eds.)]. Cambridge University Press, Cambridge, UK and New York, NY, USA, 2014.
81. Charles White (KeyLogic Systems, Inc.), personal communication with Eric Larson, 26 May 2017.
82. NETL, *Cost and Performance Baseline for Fossil Energy Plants Volume 1a: Bituminous Coal (PC) and Natural Gas to Electricity Revision 3*, DOE/NETL-2015/1723, National Energy Technology Laboratory, July 2015.
83. Mohammed, I.Y., Samah, M., Mohamed, A., and Gbadegesin, S., "Comparison of Selexol^{TM} and Rectisol$^{®}$ Technologies in an Integrated Gasification Combined Cycle (IGCC) Plant for Clean Energy Production," *International Journal of Engineering Research*, 3(12), 742-744, 2014.
84. Kakaras, E., Doukelis, A., Grammelis, P., Koumanakos, A., and Agraniotis, M., *Concept study for a 700°C power plant: using poor quality brown coal with ultra supercritical PF boiler*, final report covering 1/1/2006 to 31/10/2006, National Technical University of Athens, Mechanical Engineering Department, Thermal Engineering Section, Laboratory of Steam Boilers and Thermal Plants, Athens, Greece, October 2006.