# Fischer-Tropsch Fuels from Coal and Biomass

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# 1 Introduction

The prospect of sustained high oil prices, the heavy dependence of the US on imports for meeting its oil needs, and Middle East turmoil have together catalyzed intense interest in secure domestic alternatives to oil for satisfying US transportation energy needs. Also, it is now highly likely that the US will soon put into place a serious carbon mitigation policy—in which the transportation sector, accounting for 1/3 of US GHG emissions from fossil fuel burning, is likely to get focused attention. The two most significant domestic supplies that might be mobilized to address these challenges are biomass and coal.

Spurred by farm policy, biomass has long been a focus of development efforts that have focused on using food crops for making biofuels (primarily corn-based ethanol but also biodiesel derived from soybeans and canola). However, concerns about food price impacts [1] and indirect land use impacts of growing biomass for energy on croplands [2,3] have led to growing recognition that emphasis should be shifted instead to exploiting for energy mainly ligno-cellulosic feedstocks that don't require use of food biomass for providing energy—such as various crop and forest residues and energy crops that can be grown on degraded lands. These options include cellulosic ethanol produced biochemically and synthetic fuels derived thermochemically via biomass gasification—so-called biomass to liquids (BTL) technologies. Renewable lignocellulosic biomass provided using modest fossil fuel inputs can be considered a nearly "carbon neutral" feedstock, since $CO_2$ released to the atmosphere is recycled via photosynthesis.

Among BTL options the production of Fischer-Tropsch liquids (FTL) from biomass has been given considerable attention [4,5,6,7,8]. FTL offers as advantages over cellulosic ethanol the prospects that: (*i*) no significant transportation fuel infrastructure changes would be required for widespread use, (*ii*) the technology could plausibly come into widespread use more quickly than cellulosic ethanol, which needs considerably more development before it can be widely deployed, (*iii*) it can probably accommodate more easily the wide range of biomass feedstocks that are likely to characterize the lignocellulosic biomass supply—because gasification-based processes tend to more tolerant of feedstock heterogeneity than biochemical processes.

Recent oil price increases have led to considerable interest in making synthetic fuels from coal—so called coal-to-liquid (CTL) fuels—in light of coal's relatively low prices and the abundance of coal both in the US and in other world regions that are not politically volatile. Much of this attention has been focused on FTL [9,10,11,12]. Coal can do much to improve energy security if it is used to make FTL. Moreover, the synfuels provided would be cleaner than the crude oil products displaced (having essentially zero sulfur and other contaminants and ultra-low aromatic content). Also, for FTL production via modern entrained flow gasifiers, the air-pollutant emissions from the plant are extremely low. But synthetic fuels made from coal without capture and storage of by-product $CO_2$ result in net GHG emissions about double those from petroleum fuels. And even with $CO_2$ capture and storage (CCS), the net GHG emission rate would be no less than for the crude oil products displaced. This would not be an auspicious feature of CTL with CCS technology if society decides to pursue an energy future that avoids dangerous anthropogenic interference with climate—as is required by the UN Framework Convention on Climate Change; there is now near scientific consensus that this will require by mid-century deep reductions in GHG emissions worldwide relative to the current global GHG emission rate [13].

One approach to this challenge is to identify negative GHG emissions opportunities that might offset the CTL emissions and emissions from other difficult-to-decarbonize energy sources. Among these are opportunities to provide FTL from biomass at strong negative GHG emission rates. A striking feature of FTL technology is that a natural part of the process is the production of a stream of pure $CO_2$, accounting for about ½ of the carbon in the feedstock. If this $CO_2$ were captured and stored via CCS for FTL derived from biomass, the biofuels produced would be characterized by strong negative GHG emissions, because of the geological storage of photosynthetic $CO_2$ [14]. However, sustainably-recovered biomass is expensive, and the size of the BTL facilities will be limited by the quantities of biomass that can be gathered in a single location—which implies high specific capital costs ($ per barrel/day).

These challenges posed by the BTL-with-CCS option could be mitigated by co-processing biomass with coal in the same facility. The economies of scale inherent in coal conversion could thereby be exploited, the average feedstock cost would be lower than for a pure BTL plant, and if CCS were carried out at the facility, the negative $CO_2$ emissions associated with the biomass could offset the unavoidable positive emissions with coal, leading to FTLs with low, zero, or negative net emissions [15]. Since this CBTL-with-CCS idea was first introduced, there has been much government and industrial interest in the concept: (*i*) in 2007 an Air Force/National Energy Technology Laboratory study was released exploring the prospects that its 2016 goal for alternative jet fuel supplies[1] might be met via CBTL with CCS to the extent of reducing the GHG emission rate for the FTL so produced to 0.8 times the rate for the crude oil products displaced [17]; (*ii*) the CBTL with CCS concept got focused attention in a recent Western Governors' Association Report on future transportation fuels [18]; (*iii*) the National Energy Technology Laboratory is carrying out a major study comparing a wide range of CTL, BTL, and CBTL options with and without CCS [19], and (*iv*) some synfuel project developers are pursing plans to incorporate biomass as a feedstock along with coal in future FTL projects—including an FTL plant with CCS being planned by Baard Energy on the Ohio River at Wellsville, Ohio, that would eventually produce 50,000 barrels per day of FTL with up to up to 30% biomass by weight [20].

Despite the growing interest in using CCS and biomass along with coal in addressing simultaneously the energy insecurity and climate change challenges posed by fuels for transportation, there is not yet available a comprehensive analytical framework for deciding the most promising ways forward—including a systematic way of assessing: (*i*) BTL vs CBTL vs CTL options, (*ii*) the amounts of biomass that might be accommodated in CBTL systems, (*iii*) the appropriate scales for BTL and CBTL systems, (*iv*) the extent to which $CO_2$ capture might plausibly be pursued for all FTL systems derived from coal and/or biomass, and (*v*) prospective carbon policy impacts on FTL projects.

This paper can be considered a first step toward addressing these issues. We present here a comprehensive analytical framework suitable for addressing these challenges and early results of applying this framework by making comparisons in a self-consistent manner of designs for 16 alternative CTL, BTL, and CBTL plants, with and without CCS, with regard to mass/energy/carbon balances and economics.

# 2 CTL, BTL, and CBTL Process Designs

Sixteen processes configurations were designed and their mass and energy balances were simulated in detail. The acronyms used to describe the key design features that distinguish the 16 cases are defined in Table 1; Table 2 lists the 16 cases analyzed and indicates defining

[1] The US Air Force has established a goal of satisfying by 2016 half of its North American jet fuel demand with alternatives to crude oil.

constraints not apparent in the system acronym. Common feedstock and equipment assumptions have been maintained from one design to the next to the extent possible.

Table 1. **Definitions of acronyms for design features of FTL systems modeled.**

| | |
|---|---|
| CTL | Coal to finished FTL fuels and electricity |
| BTL | Biomass to finished FTL fuels and electricity |
| CBTL | Coal + biomass to FTL fuels and electricity |
| RC | FTL synthesis systems that recycle (RC) unconverted syngas to maximize FTL production |
| OT | FTL synthesis systems that pass syngas only once through (OT) synthesis reactor and use unconverted syngas to make coproduct power in a combined cycle power plant |
| OTA | OT system to which an autothermal reformer and extra $CO_2$ capture equipment are added downstream of synthesis to increase the fraction of feedstock C not in FTL products that is captured/stored as $CO_2$ |
| S | Biomass is mixed prairie grasses grown on C-depleted soils providing substantial soil/root C buildup over the life of the FTL production facility as a biomass carbon storage mechanism complementing underground storage of supercritical photosynthetic $CO_2$ |
| V | Coproduct $CO_2$ is vented |
| CCS | Coproduct $CO_2$ is captured |

**Table 2. Sixteen process configurations for FTL fuels production analyzed in this work.**

| **System Name** | **Constraints that accompany design features encoded in acronym** |
|---|---|
| CTL-RC-V | 50,000 barrels/day of finished FTL |
| CTL-RC-CCS | Same coal input, FTL outputs as CTL-RC-V |
| | |
| BTL-RC-V | CBIR (Common Biomass Input Rate = $10^6$ dry metric tonnes per year processed) |
| BTL-RC-CCS | CBIR, same FTL outputs as BTL-RC-V |
| | |
| CBTL-RC-V | CBIR, same coal input, FTL outputs as CBTL-RC-CCS |
| CBTL-RC-CCS | CBIR, coal/biomass ratio that yields FTL with zero GHG emission rate |
| | |
| CTL-OT-V | Same coal input as CTL-RC-V |
| CTL-OT-CCS | Same coal input as CTL-RC-V, same FTL outputs as CTL-OT-V |
| CTL-OTA-CCS | Same coal input as CTL-RC-V, same FTL outputs as CTL-OT-V |
| | |
| CBTL-OT-V | CBIR, same coal input, FTL outputs as CBTL-OT-CCS |
| CBTL-OT-CCS | CBIR, coal/biomass ratio that yields FTL with zero GHG emission rate |
| CBTL2-OT-CCS | CBIR, same FTL outputs as CTL-OT-CCS |
| CBTL-OTS-CCS | CBIR, coal/biomass ratio that yields FTL with zero GHG emission rate |
| | |
| CBTL-OTA-V | CBIR, same coal input, FTL outputs as CBTL-OTA-CCS |
| CBTL-OTA-CCS | CBIR, coal/biomass ratio that yields FTL with zero GHG emission rate |
| CBTL-OTAS-CCS | CBIR, coal/biomass ratio that yields FTL with zero GHG emission rate |

## 2.1 Common Features of all Designs

Figure 1 is a simplified generic process block diagram representing the systems considered here. The solid feedstock (coal and/or biomass) are first gasified in oxygen and steam, with subsequent gas conditioning that includes cleaning of the raw synthesis gas (syngas) and in some cases adjusting the composition of the syngas in preparation for downstream synthesis of Fischer-Tropsch liquids (FTL). Prior to synthesis, $CO_2$ and sulfur compounds are removed in the acid gas removal step. The $CO_2$ may be vented (-V) or captured and stored underground (-CCS). In the synthesis area, two basic sets of process designs are examined. In one set of designs, most of the syngas that is not converted in the synthesis reactor after a single pass is recycled (RC) to the reactor for additional conversion. In the other set of designs, the syngas is passed only once through (OT) the synthesis reactor. Following synthesis, the crude FTL product (syncrude) is refined into finished diesel and gasoline blendstocks. In either set of designs, light gases generated by syncrude refining and either purge gases (RC designs) or syngas unconverted in a single pass through the synthesis reactor provide fuel for the power island, where net power is generated for export to the grid after meeting onsite power needs.

All sixteen process designs share some common features, as discussed in this section. The subsequent two sections (2.2 and 2.3) describe key features of individual processes.

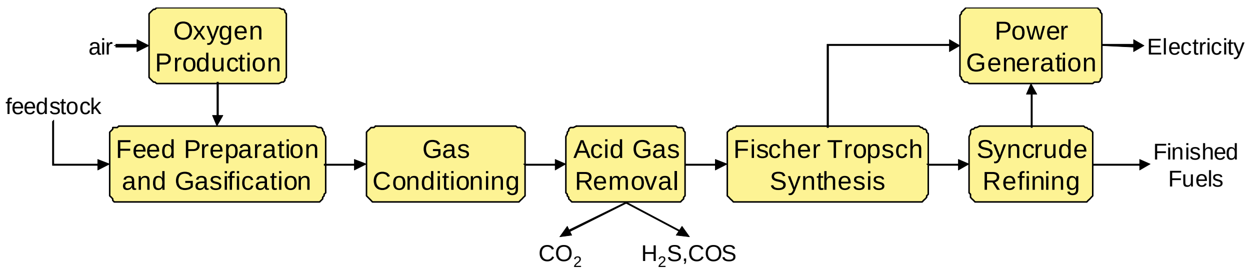


**Figure 1. Generic process block diagram.**

### 2.1.1 Feedstocks

Illinois #6 bituminous coal and an herbaceous biomass (switchgrass or mixed prairie grasses) are the feedstocks. Switchgrass is the assumed biomass for all but two of the designs that use biomass. For once-through (OT) CBTL systems with $CO_2$ capture and storage (CCS options), two cases using mixed prairie grasses (MPGs) grown on carbon-depleted soils are analyzed, as well as switchgrass. Systems utilizing biomass are designed for a Common Biomass Input Rate (CBIR) of one million dry tonnes/year (corresponding to an input capacity of 3,044 dry tonnes/day, assuming 90% capacity factor).

### 2.1.2 Gasifiers

Syngas from coal is provided by a water-slurry-fed, oxygen/steam-blown, entrained-flow gasifier operating at 75 bar, modeled after GE Energy's integral-quench gasifier. Syngas from biomass is provided by a dry-fed, oxygen/steam-blown fluidized-bed gasifier operating at 29.9 bar. The gasifier is modeled based on empirical data for pilot-plant operation from the Gas Technology Institute.[2,3] For all cases $O_2$ is provided by an onsite air separation unit (ASU). RC designs utilize 99.5% purity $O_2$; OT designs utilize 94.3% purity.

For all the options involving biomass gasification, a small amount of secondary oxygen is injected into the freeboard of the gasifier (above the bubbling fluidized bed) to promote cracking of tars and oils (heavy hydrocarbon molecules) that are produced to greater or lesser degree by all low-temperature (< 1000°C) biomass gasifiers. The literature suggests that 90% conversion of tars and oils to CO and $H_2$ can be achieved by this oxygen injection [26]. The heat released in

[2] The license for the Gas Technology Institute (GTI) technology is currently owned by Carbona/Andritz.
[3] There are no large-scale pressurized biomass gasifiers in commercial operation at present, but development and pilot-plant demonstration efforts with oxygen-blown fluidized bed gasification date to the early-1980s in Sweden [21,22] and the mid-1980s in the USA [23,24]. Most such efforts were curtailed when world oil prices fell in the late 1980s. With growing interest in hydrogen as an energy carrier, there has been some recent re-assessment of pressurized oxygen-blown gasification [25]. Thus, a not-insignificant knowledge base already exists relating to pressurized oxygen-blown gasification technology for biomass.

these reactions raises the temperature of the syngas leaving the gasifier to about 1000$^{o}$C.[4] A cyclone subsequently separates the syngas from entrained ash and unconverted char. Some ash is also removed from the bottom of the gasifier.

Following the cyclone, an external catalytic tar cracker is used to convert any residual tar in the biomass-derived syngas to light gases. This reactor is assumed to operate adiabatically. The heat needed to drive the endothermic cracking reactions is drawn from the gas itself, resulting in an exit gas temperature of about 800$^{o}$C. The tar-free gas is then cooled to 350$^{o}$C in a vertical fire-tube syngas cooler (hot gas inside the tubes) that minimizes deposition of small particles still in the gas, as well as of alkali species that condense during cooling. Particulates are then removed by a barrier filter (ceramic or sintered metal) at 350$^{o}$C, with subsequent cooling of the syngas to 40$^{o}$C for processing in the acid gas removal system.

### 2.1.3 Acid gas removal

For all systems the acid gases $CO_2$, $H_2S$, and COS contained in the syngas are removed using a Rectisol® acid gas removal system (using methanol as the working fluid). $CO_2$ removal is required to improve the kinetics and economics of the downstream synthesis process. $H_2S$ removal is required (to much lower levels than is required for power generation applications) to avoid poisoning of the synthesis catalyst. Typically any $CO_2$ removal rate (up to 100%) can be achieved, and nearly all of the sulfur species can also be removed. Solubilities of acid gases in methanol increase with decreasing temperature, so operating temperatures for good Rectisol performance are relatively low, necessitating a refrigeration plant. Aside from refrigeration, the only electricity consumption for a Rectisol plant comes from pumping of solvent and some compression to recover incidentally-removed syngas components.[5]

A Rectisol system (operating at 20 to 70 bar) can be designed for either separate removal of $CO_2$ and sulfur species from a gas stream or for co-removal. Separate removal is required when the sulfur is to be converted to elemental sulfur for disposal or sale. In all of our CTL and CBTL designs, the relatively high levels of sulfur in the syngas necessitate separate $CO_2$ and S removal. A two-column absorber design is used, with each species removed in a separate column. $CO_2$ is recovered from the solvent by flashing at successively lower pressures. The $H_2S$ is removed in a final stripping step (together with residual $CO_2$). For all but the BTL systems the captured $H_2S$ is processed via a Claus/SCOT system into elemental sulfur. In the acid gas removal step for BTL systems $CO_2$ and sulfur species can be co-removed using a single Rectisol absorber column because sulfur concentrations in the syngas are very low. The $H_2S$ is either oxidized and vented (without violating regulated $SO_2$ emission limits) or co-stored (with the $CO_2$) underground.

### 2.1.4 F-T synthesis

In our designs, we utilize a slurry-phase F-T synthesis reactor with an iron catalyst. Our kinetic model of the synthesis reactions is based on a model developed by Lox and Froment [30]

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[4] A fusion temperature for switchgrass ash (under oxidizing conditions) of 1016$^{o}$C has been reported in the literature [27]. If this temperature is also representative for reducing conditions (gasification), then it may be necessary to introduce an additive in the gasifier bed material to suppress ash fusion. Additives have been demonstrated to be able to raise ash fusion temperatures in biomass combustors [28,29].

[5] Due to the small but finite solubilities in methanol of $H_2$, CO, and $CH_4$, some quantities of these syngas components will be incidentally removed with the acid gases. In a two-column design, one may expect a loss rate of 0.45% of input $H_2$, 2.25% of input CO, and 4.5% of input $CH_4$. (In a one-column design, the loss rates would be about half these levels.) It is feasible to recover these species during solvent regeneration and to return them via a compressor to the main syngas flow, such that the net loss rates are essentially zero. This feature is included in our simulations.

and on synthesis performance reported by Fox and Tam [31] and in Bechtel's major study of FTL production in the 1990s [32]. In a slurry-phase reactor, the feed syngas is bubbled through an inert oil in which catalyst particles are suspended (Figure 2). High release rates of reaction heat (i.e., high extents of reaction) can be accommodated without excessive temperature rise due to the effective reactor cooling by boiler tubes immersed in the fluid. The vigorous mixing, intimate gas-catalyst contact, and uniform temperature distribution enable a high conversion of feed gas to liquids in a relatively small reactor volume. Conversion by slurry-phase F-T synthesis can achieve single-pass fractional conversion of CO to FTL as high as 80% [33]. This compares to less than 40% conversion with traditional fixed-bed FTL reactors. High single-pass conversion rates make slurry-phase reactors especially attractive from an overall cost perspective for OT process designs [34]. Slurry-phase reactors are commercially available for F-T synthesis but are not yet widely deployed commercially.

All FTL systems except BTL systems are designed with enough water gas shift of syngas upstream of synthesis that the $H_2$:CO ratio is 1:1 for the total quantity of syngas entering the FTL synthesis reactor. For BTL systems this ratio is instead 1.8—the ratio for the syngas exiting the gasifier with no water gas shift reactor between the gasifier and the synthesis reactor.

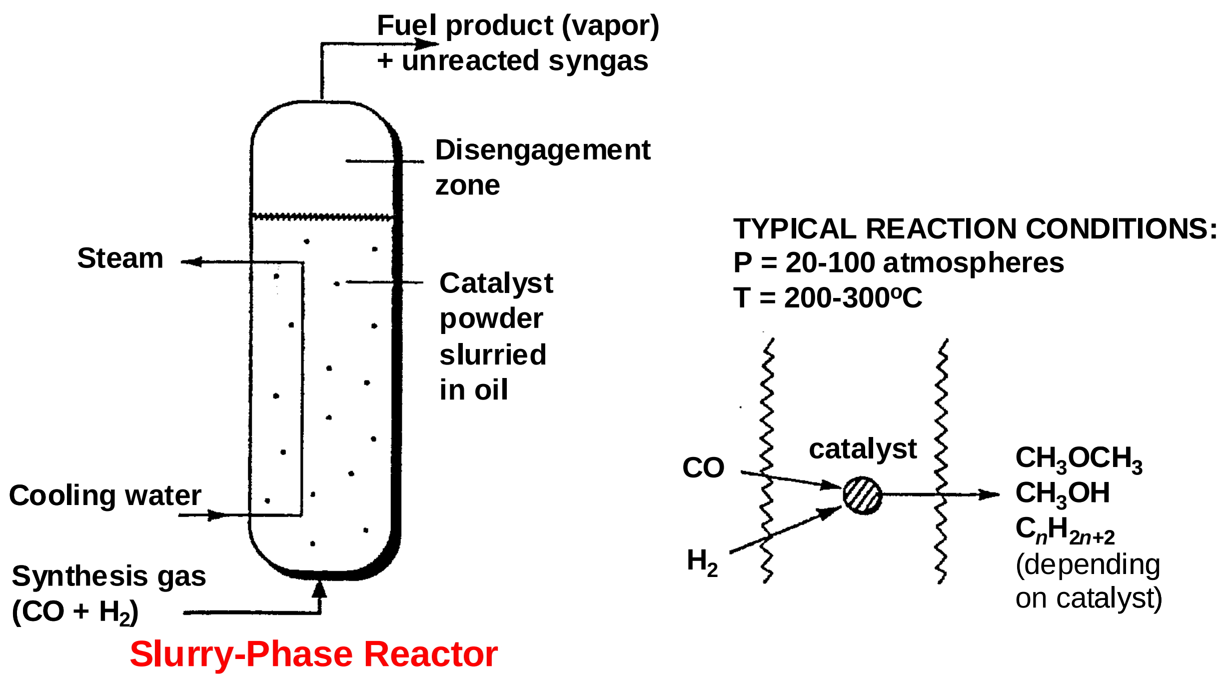


**Figure 2. Simplified depiction of a slurry-phase synthesis reactor, which can yield higher one-pass conversion of ($CO+H_2$) to liquids than traditional gas-phase reactors.**

### 2.1.5 Syncrude Refining

Most FTL systems that have been proposed would produce middle distillates (a mix of jet fuel and heavy diesel) plus naphtha, which would be sold as a feedstock to the chemical process industry. The production of finished gasoline blendstock is not often considered because of the added cost and energy expenditures associated with upgrading naphtha to gasoline. However, because only a few FTL plants would produce enough naphtha to saturate the chemical process industry's demand for naphtha, the present analysis includes upgrading naphtha to gasoline—because the focus is on understanding FTL prospects under widespread deployment conditions. The raw product from the FT reactor contains a mixture of hydrocarbons, the liquid portion of which ("syncrude") is refined into finished diesel and gasoline blendstocks

We have based the design of the refining area, including the upgrading of naphtha to gasoline blendstock (Figure 3), on a detailed Bechtel design [32,35]. Initially, the light gases in the raw FT reactor product (unconverted syngas and $C_1$-$C_4$ gases) are separated from the liquid fraction in the hydrocarbon recovery step (Figure 3). The syncrude is distilled to produce separate streams of naphtha, distillate, and wax. These fractions are processed through a series of refining steps, each of which unavoidably generates some $C_1$-$C_4$ gases that are collected for use elsewhere in the plant (outside of the refinery area).

The three liquid fractions are processed as follows. The naphtha stream is first hydrotreated, resulting in the production of hydrogen-saturated liquids (primarily paraffins), a portion of which are converted by isomerization from normal paraffins to isoparaffins to boost their octane value. Another fraction of the hydrotreated naphtha is catalytically reformed to provide some aromatic content to (and further boost the octane value of) the final gasoline blendstock. The distillate stream is also hydrotreated, resulting directly in a finished diesel blendstock. The wax fraction is hydrocracked into a finished distillate stream and naphtha streams that augment the hydrotreated naphtha streams sent for isomerization and for catalytic cracking.

In all processes we have simulated, the finished FTL products are Diesel and gasoline blendstocks produced in a mixture of 60.6% diesel and 39.4% gasoline (on a LHV basis).

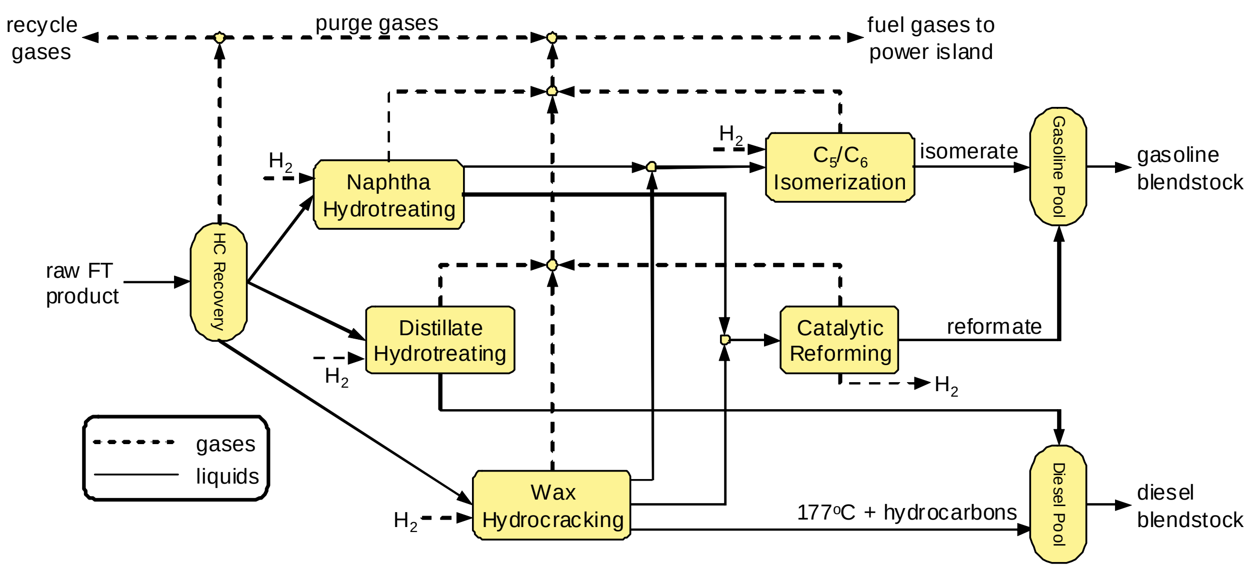


**Figure 3. Hydrocarbon recovery and refining downstream of FT synthesis. This design, based on [35], is utilized in all plant configurations.**

### 2.1.6 Power island

The assumed power island technologies are different for the RC and OT cases. For RC systems, a small amount of byproduct electricity is generated by burning purge gases and $C_1$-$C_4$ gases from the FTL refinery to produce electricity in a steam-turbine-based power plant. In the OT systems a relatively large amount of coproduct power is generated by burning unconverted syngas and $C_1$-$C_4$ gases from the FTL refinery in a gas turbine/steam turbine combined cycle power plant.

On the power island of OT systems the gas turbine performance is based on that for a General Electric MS7001FB turbine. The "7FB" is in the most-advanced class of gas turbine in wide commercial use with natural gas firing today. The hot turbine exhaust raises steam in a three-pressure level heat recovery steam generator to drive a reheat steam turbine bottoming cycle. The HRSG is integrated with process heat sources that provide additional heat inputs for steam raising, as discussed next.

### 2.1.7 Heat integration

An important design consideration for maximizing the energy efficiency of the overall process is effective heat integration, whereby heat from various process streams that require cooling is transferred to others that require heating. In all of our designs, we have integrated waste heat recovery to eliminate the need for heating from external sources and to minimize overall cooling needs.

Effective heat integration leads to very high efficiencies for the extra power provided by OT systems compared to RC systems. The exothermicity of synthesis allows for the production of saturated intermediate-pressure steam. To the extent that this saturated steam can be superheated, it can be used for power generation via expansion through a steam turbine. In the RC cases less high-temperature “waste heat” is available than in the OT cases to superheat the intermediate-pressure steam raised by the FT exotherm (see Section 6.2.2.1 for a quantitative discussion).

### 2.1.8 $CO_2$ transport and storage

For all CCS systems it is assumed that the $CO_2$ is compressed to 150 bar, transported 100 km from the conversion facility, and injected for storage 2 km underground in deep saline formations.

## 2.2 Individual Recycle (RC) Process Designs

Plants that recycle syngas unconverted in a single pass through the synthesis reactor so as to maximize the liquid fuel production per unit of feedstock are the focus of most current FTL interest. All RC systems analyzed here involve placing an autothermal reformer downstream of the recycle compressor so as to convert all $C_1$-$C_4$ gases to CO and $H_2$ and thereby maximize production of the desired FT liquid products (diesel and naphtha that is subsequently refined to gasoline). To the extent that these plants produce electricity it is mostly to meet onsite electricity needs. All RC designs considered here involve producing a small quantity of net electricity for export to the electric grid as a byproduct. Here -V and -CCS variants of RC systems based on CTL, BTL, and CBTL configurations are described.

### 2.2.1 Coal as feedstock

*CTL-RC-V*: This configuration maximizes FTL production per unit of coal input and involves venting of by-product $CO_2$. The system we have designed has a finished FTL product output capacity of 50,000 energy-equivalent barrels per day.[6]

Figure 4 is a simplified schematic of this design. A coal/water slurry is fed with oxygen from a dedicated onsite cryogenic air separation unit (ASU) into the coal gasifier. The syngas leaving the reactor is scrubbed with water before some of it enters a single-stage (adiabatic) water gas shift (WGS) reactor. A syngas bypass around the WGS reactor is adjusted to ensure that a molar $H_2$:CO ratio of 1:1 is realized for the total quantity of syngas entering the FTL synthesis reactor. Downstream of the WGS reactor, the two streams of syngas are combined,

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[6] The “energy equivalent barrel” measure assumed here is a barrel of crude-oil-derived (COD) equivalent products having the same LHV as the FTL fuels. More specifically, on a LHV basis, our FTL plants yield 60.6% synthetic diesel and 39.4% synthetic gasoline. An energy equivalent barrel of FTL has the same energy (but not volume) as a barrel of COD equivalent products containing 60.6% (by LHV) low sulfur diesel and 39.4% (by LHV) reformulated gasoline (a ratio of 1.349:1 by volume). Table 4 provides properties of FTL fuels and COD equivalent products.

cooled, and expanded to 29.5 bar, the pressure at which downstream processing occurs. $CO_2$ and $H_2S$ contained in the syngas are then removed in the Rectisol-based acid gas removal system.

In the CTL-RC-V configuration the captured $CO_2$ is vented to the atmosphere and the captured $H_2S$ is processed via a Claus/SCOT system into elemental sulfur.

Most of syngas following acid gas removal travels to the FT reactor. (A portion of the syngas is diverted for use in producing hydrogen needed for downstream refining of the raw FTL product.)

Most of the unconverted syngas and light gases generated in the hydrocarbon recovery step are compressed and recycled back through the synthesis reactor to increase FTL production after passing through the autothermal reformer, where steam and oxygen are added to produce a mixture of primarily CO, $H_2$, and $CO_2$. The reformed gas stream is mixed with the fresh syngas flow upstream of the acid gas removal island. A small portion of the gases leaving the hydrocarbon recovery unit are drawn off as a purge stream to avoid the buildup of inert gases in the recycle loop. (Note also that in the RC designs, the ASU is designed to produce extra-high purity oxygen – 99.5% molar – in order to minimize inert gas buildup.)

The purge gas is mixed with light gases produced in the FT refining area, and the mixture of gases is burned in a boiler of a steam turbine cycle in the power island. The power island generates all the electricity needed to run the facility plus a modest amount of export electricity.

*CTL-RC-CCS*: This configuration (Figure 5) involves capturing the stream of pure $CO_2$ generated as a natural part of the F-T synthesis process (accounting for 51% of the carbon in the coal or 78% of the carbon in the coal that is not contained in the FTL products). This system design is motivated by the fact that the GHG emission rate for CTL-RC-V is about twice that for the crude oil products displaced (COPD), as discussed further in Section 6.2.1.1. The CTL-RC-CCS design is identical in all respects to the CTL-RC-V design, except that the by-product $CO_2$ is compressed for delivery to a pipeline for transport to an underground storage site. The $CO_2$ captured in the Rectisol® system is recovered from the methanol solvent by flashing at two different pressures. A small amount of the captured $CO_2$ is unavoidably vented when it is stripped from the solvent with the $H_2S$ in the solvent regeneration step. For this system the fuel-cycle-wide FTL GHG emission rate is about the same as for the COPD.

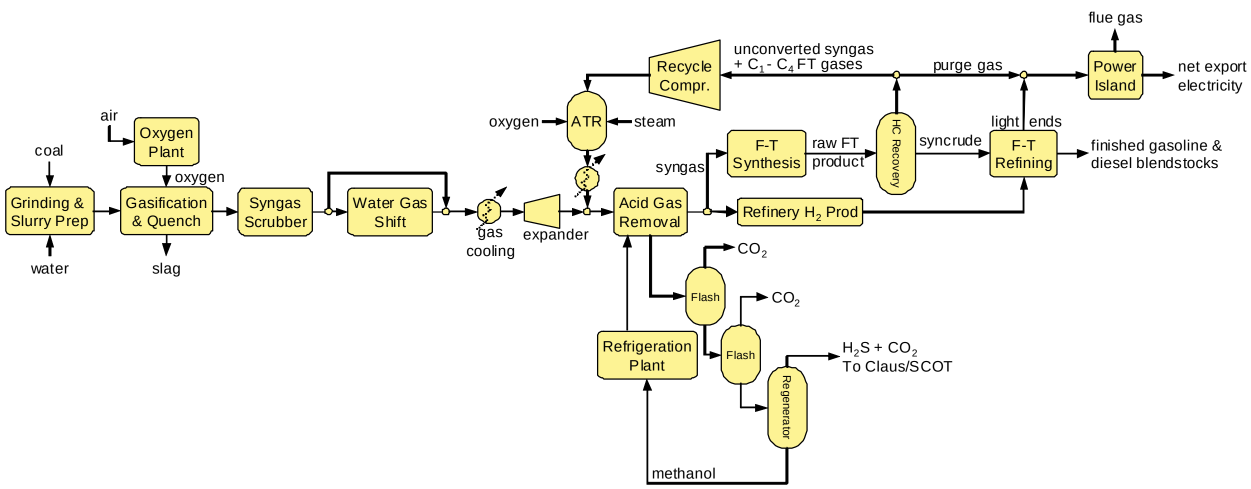


**Figure 4. CTL-RC-V process configuration for maximum FTL production from coal with venting of $CO_2$.**

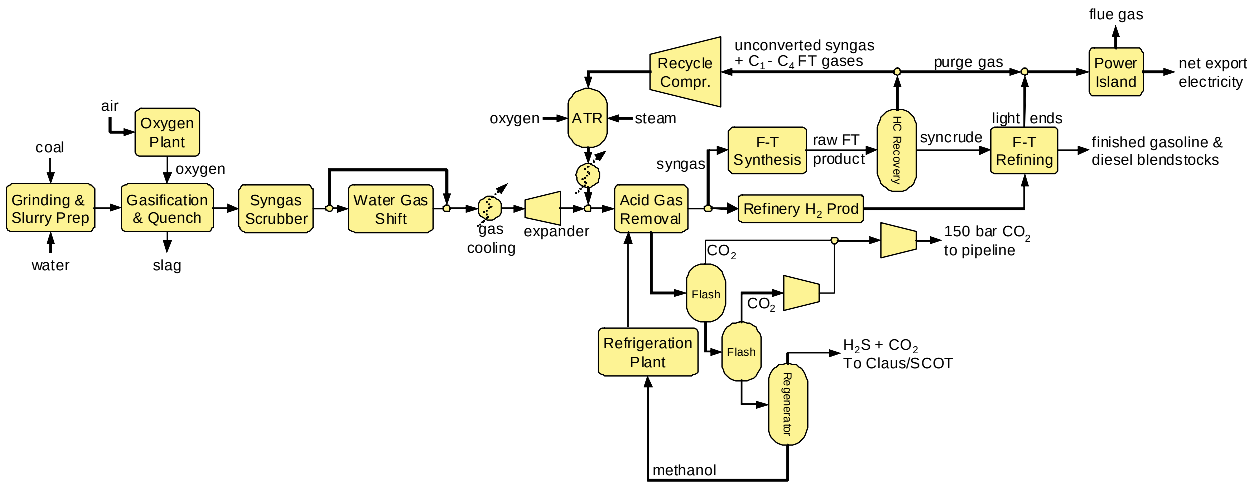


**Figure 5. CTL-RC-CCS process configuration for maximum FTL production from coal with capture and storage of $CO_2$.**

### 2.2.2 Biomass as feedstock

*BTL-RC-V and BTL-RC-CCS*: The design of our BTL options are motivated by: (i) climate change concerns—in a carbon constrained world it will be desirable to reduce the GHG emission rate for liquid fuels to levels far below those for the COPD, (ii) the fact that the emissions rate for the CTL-RC-CCS system is not lower than approximately the rate for the COPD, and (iii) the fact that biofuels produced from biomass grown with low fossil fuel inputs on a sustainable basis can be approximately carbon neutral because the carbon released to the atmosphere from biofuel combustion is offset by the $CO_2$ extracted from the atmosphere in growing new biomass to replenish the consumed supply.

The design FTL output capacity of the BTL-RC systems modeled is 4400 barrels per day—determined by the assumed constraint of the CBIR of 1 million dry tonnes per year.

The equipment configurations for maximum production of FTL fuels from biomass without and with CCS (Figure 6 and Figure 7, respectively) are identical to the coal designs described above, except for the gasification island (upstream of acid gas removal). The gasification island for the BTL designs includes chopping of the herbaceous feedstock followed by feeding via lockhoppers (using $CO_2$ as the pressurizing gas) to the oxygen/steam blown fluidized bed gasifier operating at 29.9 bar. Following gasification, removal of entrained ash and unconverted char in the cyclone, external catalytic tar cracking, cooling, and acid gas removal, the process equipment configurations for the BTL-RC-V and BTL-RC-CCS systems are identical to those for the corresponding CTL systems described above.

### 2.2.3 Coal and biomass as co-feedstocks

*CBTL-RC-V and CBTL-RC-CCS*: As discussed later (Section 6.2.1.2), the shortcomings of the BTL-RC options are the high feedstock cost and the steep scale economies for both the plant capital cost and for $CO_2$ transport to the storage site. These challenges can be mitigated by coprocessing some biomass with bituminous coal (CBTL-RC options).

For systems that utilize both coal and biomass as feedstocks, there are several process equipment configurations possible for the gasification island.

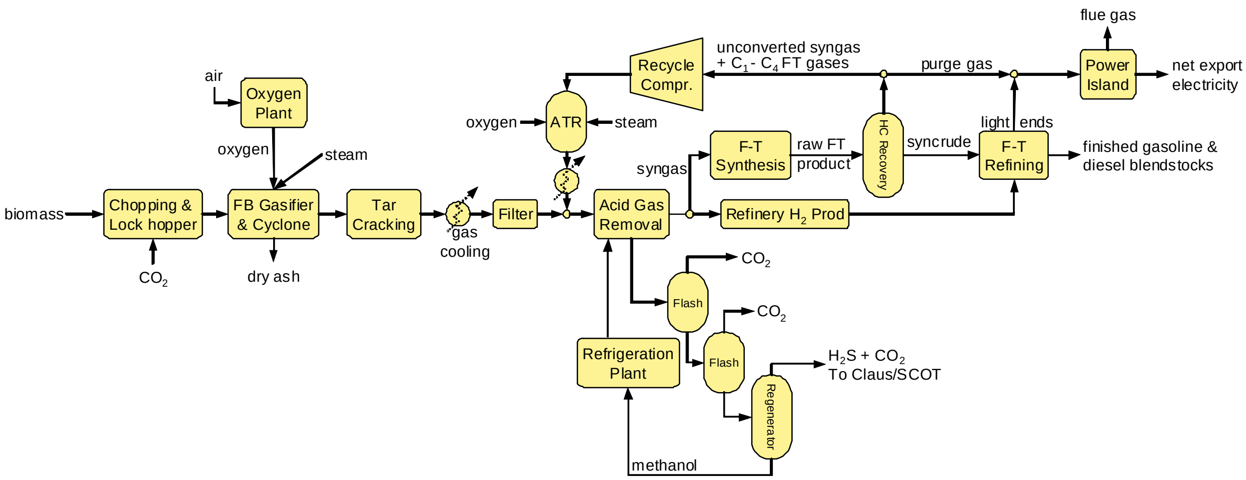


**Figure 6. BTL-RC-V process configuration for maximum FTL production from biomass with venting of $CO_2$.**

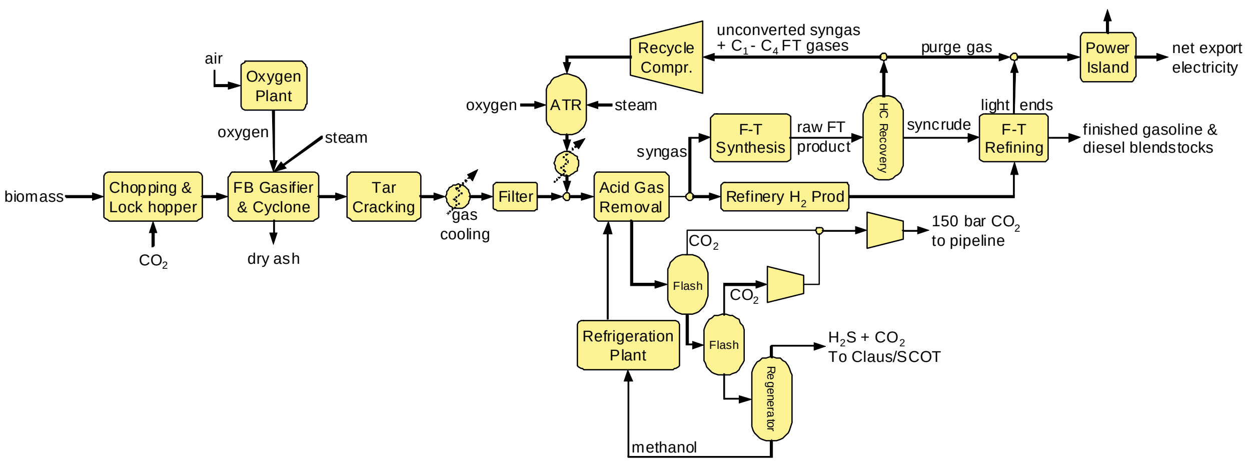


**Figure 7. BTL-RC-CCS process configuration for maximum FTL production from biomass with capture and storage of $CO_2$.**

One option is to feed coal and biomass into the same gasifier (Figure 8, top). This requires the use of a dry-feed coal gasifier (e.g., the Shell gasifier), since the added moisture delivered with the biomass co-feed would significantly reduce the efficiency of a slurry-fed coal gasifier. Co-gasification of coal and biomass in a Shell gasifier has been routinely done since 2006 at one commercial coal-IGCC facility in the Netherlands [36], and at least two proposed projects in the U.S. would utilize a similar approach [20,37]. For this design, the biomass must be milled to very small size (< 1 mm) to ensure that particles are completely gasified. A drawback of this design is that there is no opportunity to recover and return to the field the biomass ash for its inorganic nutrient value.

A second process configuration is gasification of coal and biomass in separate reactors (Figure 8, middle). With this arrangement, gas cleanup (especially tar cracking of biomass syngas) would also be done in separate trains. This arrangement has the benefit of utilizing reactors designed specifically for each feedstock, and of biomass ash recovery for return to the field. A disadvantage compared to co-gasification is the added capital cost that arises from multiple gasifier/gas cleanup trains.

A third process option is a hybrid between the first two. In this design, biomass is gasified in its own reactor, and the resulting raw syngas is fed to an entrained flow coal gasifier (Figure 8, bottom). This configuration enables recovery of the biomass ash and also utilizes the coal gasifier to crack tars formed during biomass gasification, thereby obviating the need (and cost) for a separate train for biomass gas cleanup. Also, either a dry-feed or slurry-feed coal gasifier should be suitable for use in this configuration.

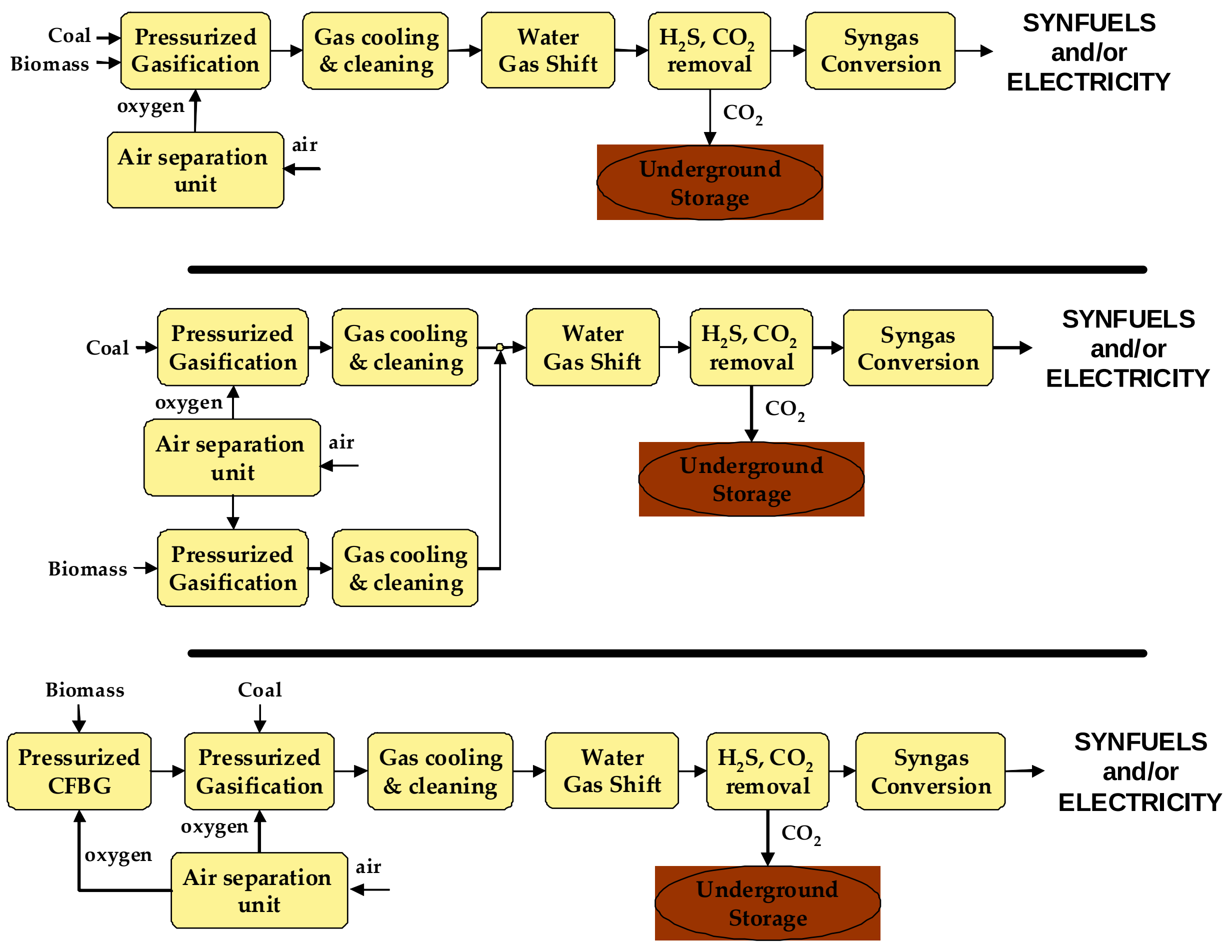


**Figure 8. Alternative process designs for combined processing of coal and biomass feedstocks for production of FTL fuels.**

All of the CBTL systems we have designed utilize separate gasifier/cleanup trains for coal and for biomass (Figure 8, middle). To illustrate the possibilities we have designed a CBTL-RC-CCS plant using just enough biomass to reduce to zero the fuel-cycle-wide GHG emission rate for the production and consumption of the FTL. Under this condition, the scale of the plant (10,000 barrels per day of FTL output capacity) is determined by the condition that the biomass delivery rate to the synfuel plant is 1 million dry tonnes per year. Biomass accounts for 43.4% of the feedstock on a HHV basis for this plant design. The CBTL-RC-V plant has the same coal/biomass inputs and the same FTL outputs as the CBTL-RC-CCS plant.

The gasifier/cleanup trains for each feedstock are as described above for the CTL and BTL systems, respectively, and the downstream processing configurations for cases without and with CCS (Figure 9 and Figure 10, respectively) are identical to the configurations used for CTL-RC-V and CTL-RC-CCS cases, respectively.

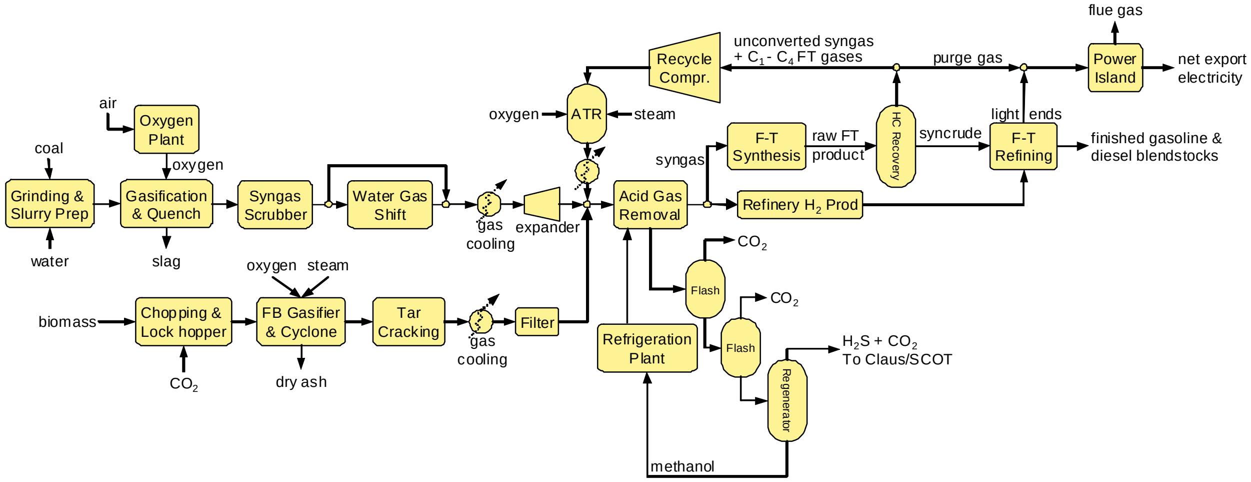


**Figure 9. CBTL-RC-V process configuration for maximum FTL production from coal plus biomass with venting of $CO_2$.**

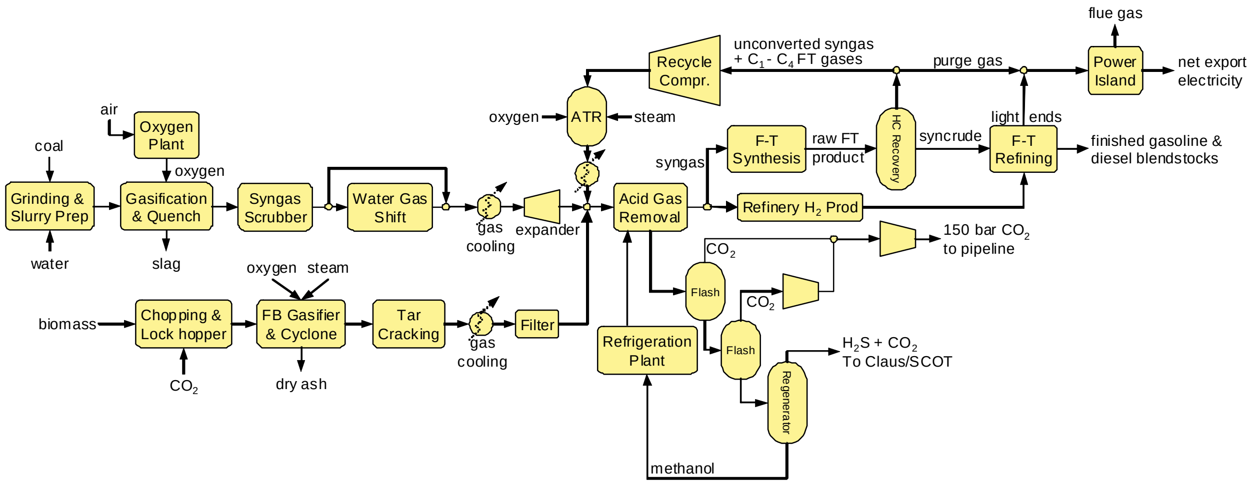


**Figure 10. CBTL-RC-CCS process configuration for maximum FTL production from coal plus biomass with capture and storage of $CO_2$.**

## 2.3 Individual "Once-Through" (OT) Process Designs

Complementing the recycle designs described above are plant designs that co-produce a substantial amount of electricity along with FTL fuels. In these designs, the syngas is passed only once through (OT) the synthesis reactor, and the unconverted syngas is used as fuel for a gas turbine/steam turbine combined cycle for power generation. As a result, the OT systems are able to meet all onsite electricity needs and export substantial amounts of power to the grid. For these systems the large amount of electricity produced suggests labeling the electricity as a coproduct rather than a byproduct. We discuss here the -V and -CCS variants of OT systems in CTL, BTL, and CBTL configurations.

### 2.3.1 Coal as feedstock

Three process designs were developed for OT systems using coal as the only input feedstock. The scale of each of these designs was set by fixing the coal input rate to be the same as the feed rate for the CTL-RC cases.

*CTL-OT-V*: This design (Figure 11) is like the CTL-RC-V design (Figure 4) except for the synthesis and power islands. In CTL-OT-V all of the unconverted syngas and light off-gases from FTL refining are compressed and supplied to a gas turbine combined cycle power island.

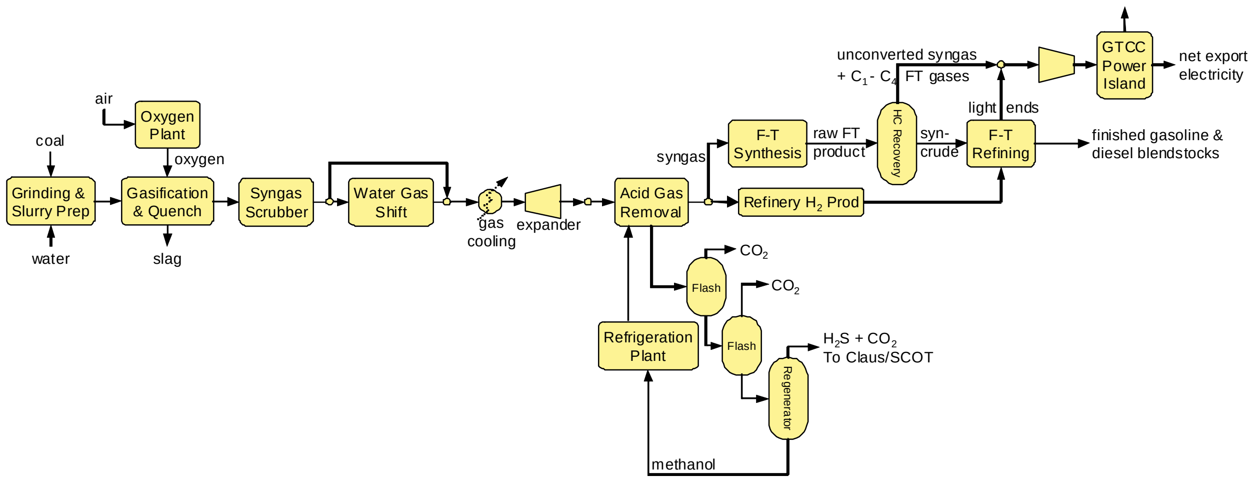


**Figure 11. CTL-OT-V process configuration for co-production of FTL fuels and electricity from coal with venting of $CO_2$.**

*CTL-OT-CCS* and *CTL-OTA-CCS*: Two CTL-OT systems utilizing $CO_2$ capture and storage have been designed. For both of these system designs the coal input level and the FTL outputs are the same as for the CTL-OT-V case. In these designs, $CO_2$ is removed at two points in the system: one is upstream of the FTL synthesis island (as in all designs discussed above) and the second is downstream of the synthesis island. Some cost savings are achieved by using a common Rectisol solvent regeneration system for the two separate absorption systems.)

In the CTL-OT-CCS design (Figure 12) the unconverted syngas and refinery off-gases that constitute the power island fuel are passed through a $CO_2$ absorption column before reaching the power island. Removal of the $CO_2$ increases the calorific value of the power island fuel, necessitating measures (not needed in the CTL-OT-V case) to ensure that thermal $NO_x$ emissions from the power island remain below regulated limits.[7] This is achieved by first passing the gas turbine fuel through a saturator and then utilizing some nitrogen dilution in the gas turbine combustor to effectively further dilute the calorific value of the turbine fuel. Nitrogen for this purpose is taken via a compressor from the plant's oxygen production unit.

The CTL-OT-CCS system captures 68% of the coal carbon that is not contained in the FTL products, compared to 78% for the CTL-RC-CCS system—which implies a higher FTL GHG emission rate for the CTL-OT-CCS system (as discussed further in Section 6.2.2.)

The CTL-OTA-CCS design (Figure 13) is similar to the CTL-OT-CCS design but utilizes additional equipment downstream of the synthesis island to increase the amount of $CO_2$ that can be captured and stored—from ~ 68% of the coal C not contained in the FTL products for CTL-OT-CCS to ~ 85%, thereby reducing the GHG emission rate for the FTL produced (see Section 6.2.2.1). The unconverted syngas and refinery off-gases are processed through an autothermal reformer (ATR) to provide additional CO and $H_2$, a two-stage water gas shift system subsequently converts essentially all CO in the syngas to $H_2$ and $CO_2$, and the latter is captured in additional downstream $CO_2$ capture equipment.

[7] For simulation purposes, we assume that the adiabatic flame temperature in the gas turbine combustor must not exceed 2300 K if $NO_x$ emissions are to remain below 25 ppmv.

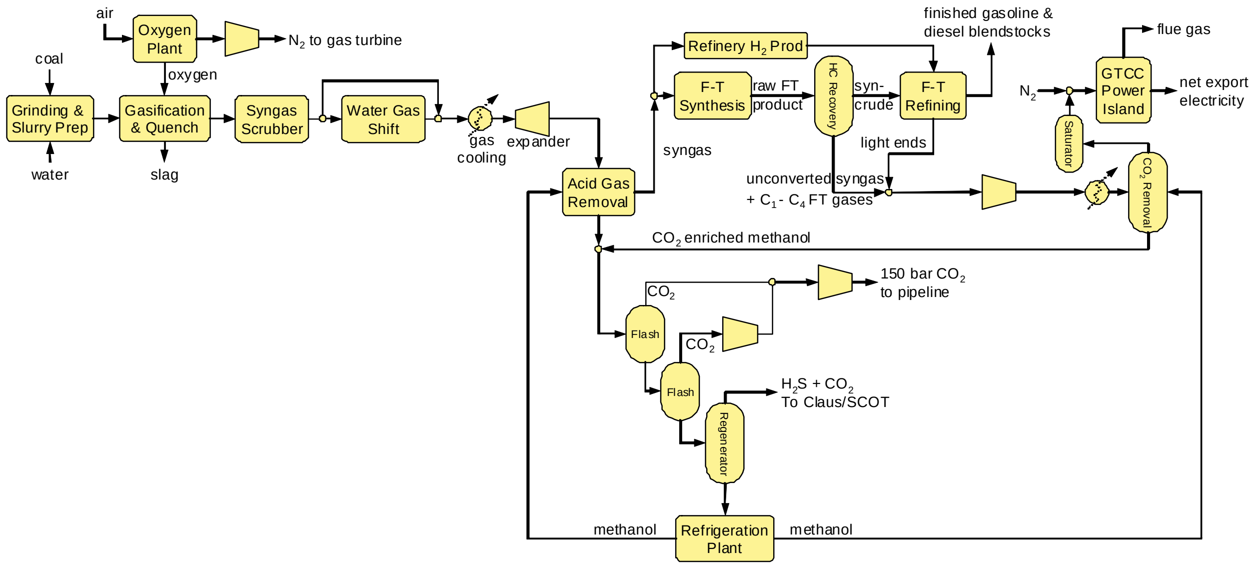


**Figure 12. CTL-OT-CCS process configuration for co-production of FTL fuels and electricity from coal with capture and storage of $CO_2$.**

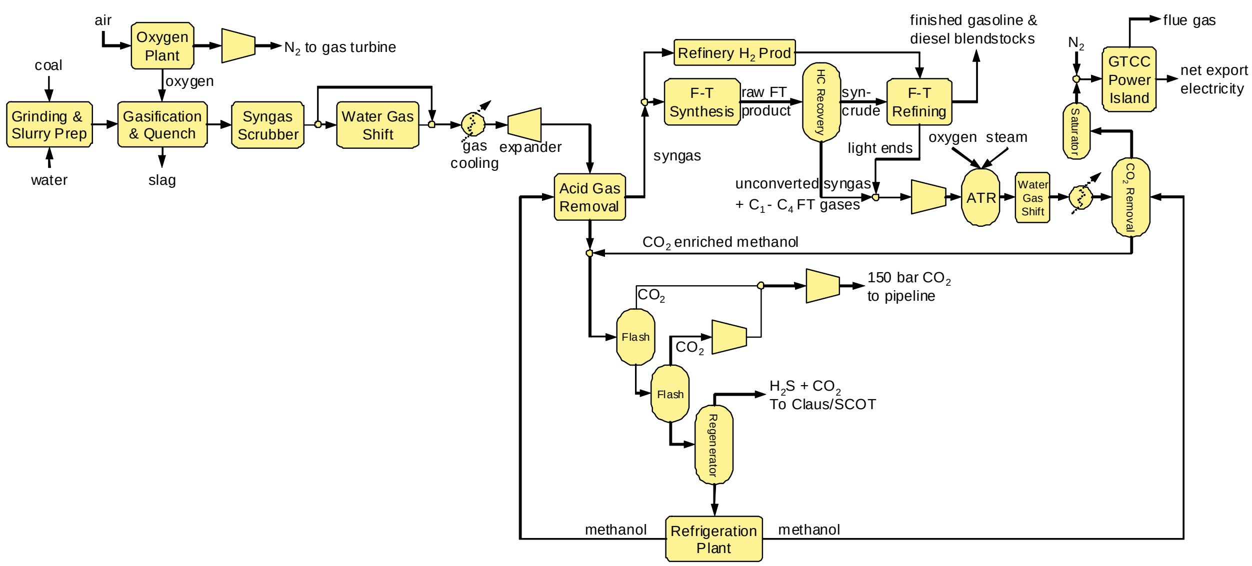


**Figure 13. CTL-OTA-CCS process configuration – an alternative to CTL-OT-CCS for co-production of FTL fuels and electricity from coal with capture and storage of additional amounts of $CO_2$.**

### 2.3.2 Coal and biomass as co-feedstocks

As for the CBTL-RC options, coprocessing some biomass with the coal can give rise to substantial C-mitigation benefits, especially by exploiting photosynthetic $CO_2$ storage opportunities. Here the possibilities for CBTL-OT systems are illustrated:

- First, by describing what might be accomplished in terms of GHG emissions reduction by adding a relatively small amount of biomass (switchgrass) to a bituminous coal-based FTL system so as to create a CBTL2-OT-CCS option with energy output characteristics similar to those for the CTL-OT-CCS option discussed above;
- Second, by describing a CBTL-OT-CCS option that provides FTL for plants consuming bituminous coal and just enough switchgrass to realize a zero net GHG rate for FTL;

- Third, by describing a CBTL-OTS-CCS option that provides FTL with a zero net GHG emission rate for plants consuming bituminous coal and mixed prairie grasses (MPGs), the growing of which on C-depleted soils leads to additional storage of carbon via the buildup of C in soil and roots (S = C storage in soil + roots via the growing of MPGs).
- Fourth, by describing CBTL options with CCS that employ autothermal reformers so as to reduce the amount of biomass needed to realize a zero net GHG emission rate for FTL by increasing the fraction of carbon in the feedstocks that is captured and stored underground as $CO_2$.

*CBTL2-OT-CCS*: As discussed in Section 6.2.2, a CTL-OT-CCS system can produce FTL at much lower costs than a CTL-RC-CCS system, but the FTL GHG emission rate is much higher than for the COPD as a result of the fact that only 68% of the carbon in the feedstocks not contained in the FTL products is captured, compared to almost 78% for CTL-RC-CCS systems. The CTL-OTA-CCS design employs an autothermal reformer and associated additional $CO_2$ capture technologies to substantially increase the $CO_2$ capture fraction (to 85% of the carbon in the feedstocks not contained in the FTL products) so as to reduce the GHG emission rate for FTL. But this option is much more costly than the CTL-OT-CCS option (see Section 6.2.2). The CBTL2-OT-CCS design offers an alternative approach for reducing the FTL GHG emission rate. The CBTL2-OT-CCS plant is designed with the same FTL output capacity as CTL-OT-CCS (36,655 barrels per day) but with some of the coal input replaced by 1 million dry tonnes of switchgrass per year.[8] This system (see Figure 14), fired with 8.6% biomass on a HHV basis, would have an FTL GHG emission rate equal to the rate for the COPD and would be far more cost-effective than the CTL-OTA-CCS option (see Section 6.2.2).

*CBTL-OT-CCS, CBTL-OT-V, CBTL-OTS-CCS*: The CBTL-OT-CCS system (Figure 14) includes a gasification area configuration identical to that for CBTL-RC-CCS system described earlier (Figure 10) and a configuration downstream of the gasification/cleanup area that is identical to the CTL-OT-CCS design (Figure 12). The system is designed in the same spirit as CBTL-RC-CCS—using just enough biomass in the form of switchgrass to reduce to zero the fuel-cycle-wide GHG emission rate for the produced FTL. Under this condition and the assumption that MPGs are provided at a rate of 1 million dry tonnes per year, the scale of the FTL output capacity is 8,100 barrels per day and biomass accounts for 38.1% of the feedstock on a HHV basis (compared to 43.4% in the CBTL-RC-CCS case).

The CBTL-OT-V option (no figure included) is introduced merely to enable a calculation of the cost of GHG emissions avoided for the CBTL-OT-CCS option (see Section 6.2.2.2). Accordingly, the FTL output capacity and the percentage of biomass in the feedstock of the CBLT-OT-V design are the same as for the CBTL-OT-CCS design.

The CBTL-OTS-CCS configuration (S = soil and root carbon storage) is identical to the CBTL-OT-CCS configuration, except that the feedstock is mixed native-prairie grasses (MPGs)

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[8] Among the sixteen process designs described here, only the CBTL2-OT-CCS case was not simulated in detail using Aspen Plus. Instead, the characteristics of this plant were estimated as a linear combination of the CTL-OT-CCS plant and a linearly scaled-up version of CBTL-OT-CCS that assumes the same physical equipment configuration and energy efficiencies of the CBTL-OT-CCS system that was simulated in detail. To estimate the outputs for the CBTL2-OT-CCS design, we scaled all of the outputs and equipment capacities of the CBTL-OT-CCS case such that the FTL fuels output capacity was the same as in the CTL-OT-CCS design: 36,655 barrels per day of FTL fuels. The CBTL2-OT-CCS design represents a linear combination of the characteristics of the scaled-up CBTL-OT-CCS (22% contribution) and those of the CTL-OT-CCS plant (78% contribution). The net result is a plant, CBTL2-OT-CCS, that produces 36,655 bpd using an input coal-to-biomass ratio of 11.1 (lower heating value basis).

rather than switchgrass (Figure 15). This system is again designed using just enough biomass feedstock to reduce to zero the fuel-cycle-wide GHG emission rate for the produced FTL.

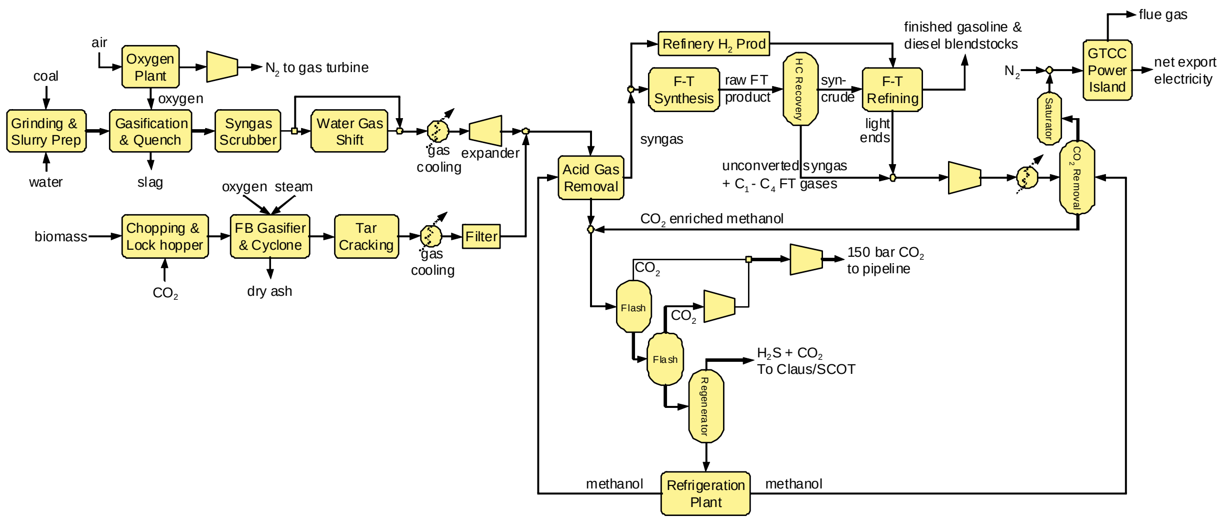


**Figure 14. CBTL-OT-CCS and CBTL2-OT-CCS process configurations for co-production of FTL fuels and electricity from coal plus switchgrass biomass with OT synthesis and with capture and storage of $CO_2$.**

When established on carbon-depleted soils (as assumed here), MPGS can store significant amounts of carbon in the soil and roots over decadal time scales [45]. With soil/root carbon storage included in the lifecycle GHG calculations, less biomass is needed in the CBTL-OTS-CCS system to achieve the same level of net lifecycle GHG emissions associated with the FTL fuels production as in the CBTL-OT-CCS system. Here it is assumed that the feedstock is 16 species of MPGs grown on carbon-depleted soils—e.g., on Conservation Reserve Program (CRP) lands or other degraded lands. Following [45] it is assumed that the average rate of soil and root carbon buildup is 0.6 tonnes of C per tonne of C in the harvested biomass, when averaged over a 30-year period. Under this condition and the assumption that MPGs are provided at a rate of 1 million dry tonnes per year, the scale of the FTL output capacity is 13,040 barrels per day and biomass accounts for 23.9% of the feedstock on a HHV basis.

*CBTL-OTA-CCS, CBTL-OTA-V, CBTL-OTAS-CCS*: As suggested by the discussion of the CTL-OTA-CCS design in Section 2.3.1, it is possible to increase the level of $CO_2$ capture achieved with the CBTL-OT-CCS design by adding an autothermal reformer (ATR) and associated additional $CO_2$ capture technologies downstream of synthesis (Figure 16). This is an especially important consideration for CBTL systems in light of the scarcity of biomass resources—because these designs make it feasible to realize a low net GHG emission rate for FTL with a lower percentage of biomass in the feedstock

The CBTL-OTA-CCS design makes it feasible to increase the percentage of feedstock C not contained in the FTL products that is captured as $CO_2$—from the ~ 67% level for CBTL-OT-CCS to ~ 85% (see Section 6.2.2.2). This makes it possible to reduce the biomass percentage from 38.1% for CBTL-OT-CCS to 23.9% when both systems are characterized by a zero net GHG emission rate for FTL. As with the CTL-OTA-CCS case (Figure 13), a fuel gas saturator and $N_2$ injection to the gas turbine are used to control $NO_x$ emissions.

As in the case of the CBTL-OT-V option, the CBTL-OTA-V option (no diagram included here) is introduced to enable a calculation of the cost of GHG emissions avoided for the CBTL-

OTA-CCS option (see Section 6.2.2.2). Accordingly, the FTL output capacity and the percentage of biomass in the feedstock of the CBLT-OTA-V design are the same as for the CBTL-OTA-CCS design. (Despite the OTA designation, there is no autothermal reformer in the CBTL-OTA-V system, which is merely the –V version of the CBTL-OTA-CCS system.)

Similarly, if the CBTL-OTAS-CCS design (no diagram) is substituted for the CBTL-OTS-CCS design, the biomass percentage of the feedstock input is reduced from 23.9% to 16.4%.

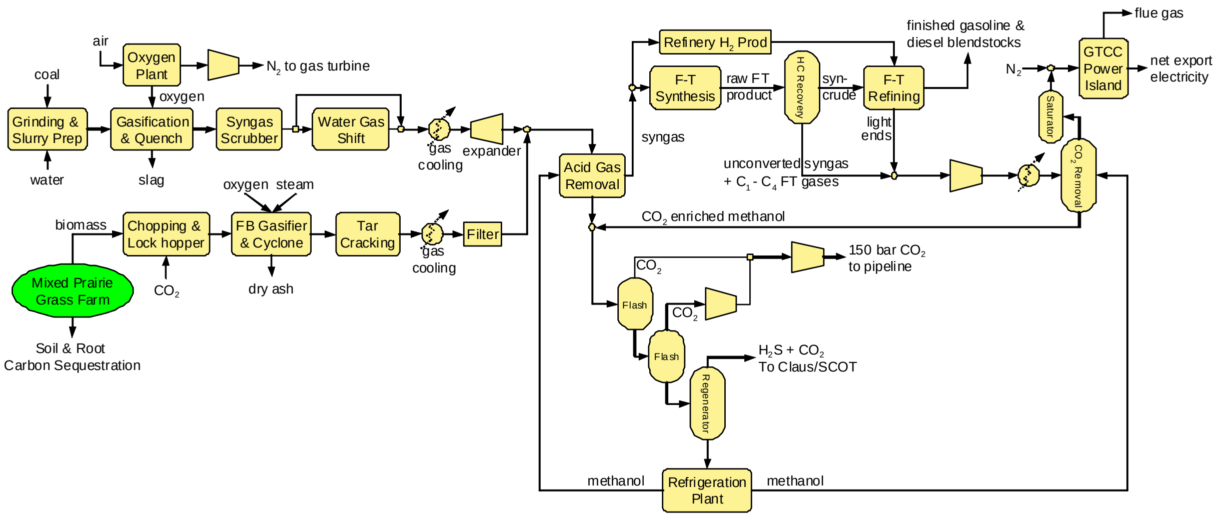


**Figure 15. CBTL-OTS-CCS process configuration for co-production of FTL fuels and electricity from coal plus mixed prairie grasses with once-through synthesis and with capture and storage of $CO_2$.**

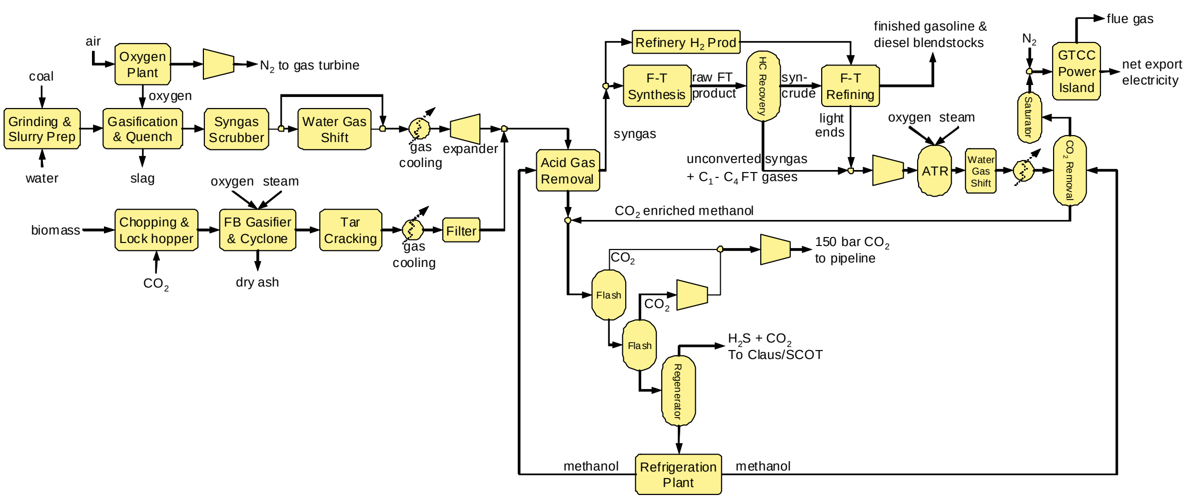


**Figure 16. CBTL-OTA-CCS process configuration for co-production of FTL fuels and electricity from coal plus biomass with once-through synthesis and with greater level of capture and storage of $CO_2$ than the CBTL-OT-CCS design in Figure 14.**

# 3 Performance Simulations

The mass and energy balances for the sixteen process designs described above were simulated in detail using Aspen Plus® chemical process software. In this section we describe quantitative simulation assumptions and results. The latter provide the basis for the capital cost estimates and economic analyses later in this paper.

## 3.1 Simulation assumptions

To ensure consistency from one process simulation to the next, feedstock characteristics and values assigned to most design parameters are kept fixed for all simulations.

The feedstocks considered in our simulations are Illinois #6 coal and herbaceous biomass (switchgrass or mixed prairie grasses), Table 3. For process designs using only coal as feedstock, the processing capacity is fixed for all cases at 24,297 metric tonnes per day (tpd) of "as-received" (AR) coal, corresponding to a dry matter flow of 21,595 tpd. The energy flow corresponding to 24,297 tpd of 11.12% moisture coal is 7,272 $MW_{HHV}$ or 6,937 $MW_{LHV}$. For all process designs using biomass (BTL and CBTL designs), the biomass input capacity is fixed at 3,581 metric tons per day (tpd) of AR biomass, corresponding to a dry matter input capacity of 3,044 tpd (or 1 million dry tonnes per year of actual throughput, assuming a 90% capacity factor). (The 15% moisture level is sufficiently low that active drying of the feed material is not necessary before gasification.) The energy flow corresponding to 3,581 tpd of 15% moisture switchgrass is 601 $MW_{HHV}$ or 548 $MW_{LHV}$.

Table 4 provides, for reference, properties of liquid fuels discussed in this report. Table 5 indicates key design assumptions for each piece of process equipment in our simulations.

**Table 3. Feedstock characteristics for coal (from [55]) and biomass (from [38]) used in process simulations.**

| | Illinois #6 coal | Herbaceous biomass |
|---|---|---|
| **Proximate Analysis (weight%, as-received)** | | |
| Fixed carbon | 44.19 | 18.1 |
| Volatile matter | 34.99 | 61.6 |
| Ash | 9.7 | 5.3 |
| Moisture content | 11.12 | 15.0 |
| Lower Heating Value (MJ/kg) | 25.861 | 14.509 |
| Higher Heating Value (MJ/kg) | 27.114 | 15.935 |
| **Ultimate Analysis (weight%, dry basis)** | | |
| Carbon | 71.72 | 46.96 |
| Hydrogen | 5.06 | 5.72 |
| Oxygen | 7.75 | 40.18 |
| Nitrogen | 1.41 | 0.86 |
| Chlorine | 0.33 | 0 |
| Sulfur | 2.82 | 0.09 |
| Ash | 10.91 | 6.19 |
| Higher Heating Value (MJ/kg) | 30.506 | 18.748 |

To simulate a gas turbine operating on fuel gas with characteristics different from the natural gas fuel that most gas turbines are designed for, we first calibrated our gas turbine model against the quoted performance of the GE 7FB gas turbine operating on natural gas. We tuned expander and compressor efficiencies, with some other parameters fixed (Table 6) to match the power output and overall thermal efficiency quoted by GE. (We fixed values of some other parameters, as shown in Table 6). We then adjusted our calibrated gas turbine model to simulate the performance of the gas turbine firing syngas, which has a considerably lower volumetric heating content than natural gas. We assumed that the turbine inlet temperature (TIT) with syngas will be the same as with natural gas.[9] We also assumed that the turbine expander operates with a "choked" flow condition at its inlet. Under this condition, the mass flow into the expander cannot be increased (as would be required since a higher mass flow of fuel is required in the turbine combustor to achieve the same energy input as with natural gas firing) without increasing the pressure upstream of the inlet, i.e. raising the compressor pressure ratio and/or lowering the flow rate of combustion air. We increased the compression ratio over the quoted value for a natural gas machine by about 5%, a percentage increase that is within the range typically tolerated by compressors of heavy-duty gas turbines. Table 6 shows our resulting performance prediction for the gas turbine fired with a syngas.

**Table 4. Properties of FTL fuels assumed in Aspen simulations and of crude-oil derived fuels.**

| | FTL, as simulated (a) | | Crude oil derived (b) | |
|---|---|---|---|---|
| **Fuel Property** | **Diesel** | **Gasoline** | **Diesel** | **Gasoline** |
| Average chemical composition | C15H32 | C9H20 | | |
| Molecular weight, g/mol | 212 | 128 | | |
| LHV, MJ/kg | 44.0 | 44.3 | 42.6 | 42.4 |
| HHV, MJ/kg | 47.3 | 47.8 | 45.6 | 45.4 |
| Density, kg/liter | 0.600 | 0.658 | 0.847 | 0.747 |
| Volumetric energy density, MJ/liter LHV | 26.4 | 29.2 | 36.1 | 31.7 |
| Volumetric energy density, MJ/liter HHV | 28.4 | 31.4 | 38.6 | 34.0 |
| Carbon intensity, kgC/GJ LHV | 19.3 | 19.0 | 20.4 | 19.8 |

(a) These are values Aspen associates with the mass flows it calculates for FTL diesel and gasoline streams in our simulations. The FTL energy rates (in MW) that we report in Section 3.2 are the product of the Aspen flow rate and the corresponding Aspen heating value. Likewise, carbon emissions from combustion of FTL diesel and gasoline are calculated as the Aspen-based energy rates multiplied by the carbon intensities given here. The volumetric output rates that we report later in Section 3.2 are the FTL energy rates divided by the volumetric heating value of crude-oil derived products (gasoline or diesel) given in the right had side of this table.

(b) From [43] for low-sulfur diesel and reformulated gasoline.

Finally, for the sake of convenience, in all of our simulations involving a gas turbine, we assume that the turbine performance based on our 7FB simulation is achieved regardless of the scale of the turbine.[10] (In our cost analysis discussed later, multiple turbines are utilized as necessary if a single turbine would exceed a specified maximum capacity of 266 $MW_e$.)

Our heat recovery steam generator (HRSG) cools the gas turbine exhaust to 90$^o$C, with steam being generated at 5.0 bar, 24.3 bar, and 132 bar. The HRSG includes a superheating/ reheating section that raises the steam temperature to 566 $^o$C for both the high pressure steam and

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[9] Due to the higher flow rates and higher concentrations of $H_2O$ in the combustion products with syngas relative to natural gas firing, maintaining the same TIT as for the natural-gas version implies higher temperatures and heat transfer to the blades throughout the expansion and thus, *ceteris paribus*, higher blade metal temperatures and shorter lifetime for the hot parts of the engine. Running a syngas-fired gas turbine at the same TIT as rated for natural gas implies an increase in blade metal temperatures of 20-25°C and an increase in turbine outlet temperature of 10-20°C [39]. This is why syngas-fired gas turbines today are typically de-rated (TIT lower by 20-30°C) to match the lifetime and reliability of natural gas-fired versions. However, because natural gas-fired gas turbine performance has historically improved steadily over time, we assume that syngas-fired gas turbines in future FTL plants will be able to achieve performance (i.e., TIT) at least comparable to that of today's natural gas-fired machines. We make the same assumption for the cases where the turbine fuel is essentially hydrogen (CTL-OTA-CCS and CBTL-OTA-CCS), which may require a larger derating of the natural gas TIT – perhaps as much as 50$^o$C [40] – to keep the metal temperatures and lifetime similar to those found during operation on natural gas.

[10] In practice, gas turbines are available commercially only in discrete capacities. Realistically, each plant should be individually scaled to utilize the full design capacity of its gas turbine(s). However, we believe that the added complication of having different scales for each plant is not worth the relatively small gain in modeling accuracy; the overall conclusions we draw from this work will be the same in either case.

the reheated intermediate pressure steam. A bleed of steam from a low-pressure stage of the turbine is used for deaerator heating. The LP steam turbine discharges at 31.7 $^oC$ and 0.0462 bar.

**Table 5. Process design parameter assumptions for Aspen Plus simulations.**

| | |
|---|---|
| Air separation unit | $O_2$ purity = 94.3 (mol) for OT designs, 99.5% (mol) for RC designs.<br>Air input (mol%): $O_2$, 20.95%; $N_2$, 78.12%; Ar = 0.93%.<br>Power (kWh/t pure $O_2$ at 75 bar) = 365 (for 99.5% purity) and 385 (for 94.3% purity). |
| Coal and biomass prep and handling | Coal: Electricity requirement = 0.29% of input coal $MW_{LHV}$.<br>Biomass handling: 10 kJ/kg AR biomass; Lock hopper electricity: 10 kJ/kg dry biomass. |
| Entrained-flow, coal-slurry, integral-quench gasifier | Slurry wt% solids = 66.5; Carbon conversion = 95%; $T_{gasifier}$ = 1371$^oC$ by adjusting $O_2$ in;<br>Heat loss = 1% coal HHV; $P_{reactor}$ = 75 bar; Δp = 3.5 bar. |
| Scrubber | $T_{water, in}$ = 200$^oC$; Simulated as flash drum, $T_{gas,out}$ = ~250°C |
| Pressurized bubbling fluidized-bed biomass gasifier | $O_2$:dry biomass (wt. ratio) = 0.30; Steam:dry biomass (wt. ratio) = 0.25; Carbon conversion = 90.18%; Heat loss = 1% biomass HHV; Pressure = 29.9 bar; $CO_2$ to lock hopper = 0.1 x dry biomass weight. |
| Auto-Thermal Reformer | Chemical equilibrium at $T_{out}$ (Gibbs reactor). $T_{steam, in}$ = 550 $^oC$, $T_{oxygen, in}$ = 300 $^oC$.<br>Steam-to-carbon mol ratio = 1.6 to 3; $T_{out}$ = 900 to 1000$^oC$ by adjusting $O_2$ flow; |
| Partial water gas shift (single stage) | $T_{in}$=250°C; Δp = 2 bar; Heat loss = 0. Chemical equilibrium (Gibbs reactor).<br>No external steam input required. Inert compounds: $CH_4$, $C_2H_4$, $C_2H_6$, Ar, $N_2$ |
| Full water gas shift (two stages) | *First stage*: Adiabatic; $T_{in}$ = 400°C;; Δp = 1 bar; Chemical equilibrium (Gibbs reactor)<br>*Second stage*: Isothermal; $T_{in}$ = 180$^oC$; Δp = 1 bar; Chemical equilibrium (Gibbs reactor).<br>No external steam input required in either stage; Inert compounds: $CH_4$, $C_2H_4$, $C_2H_6$, Ar, $N_2$ |
| Rectisol acid gas removal | 97% $CO_2$ capture; 100% S capture; 3 mols $CO_2$ vented / mol $H_2S$ sent to Claus/SCOT;<br>Utilities: Electricity (other than refrig) = 1900 kJ/kmol($CO_2$+$H_2S$); Refrigeration ($MW_e$) = 3 x $MW_{thermal}$ from cooling input syngas 12$^oC$; 5 bar steam = 6.97 kg/mol ($H_2S$+$CO_2$) |
| Slurry-phase FT synthesis reactor | $T_{in}$ = 245$^oC$; $T_{out}$ = 260$^oC$; Simulated as RCSTR (user-subroutine FT kinetic model). |
| Syncrude refining | Unit operations mass balances as reported by Bechtel [35]. |
| Gas turbine | $P_{fuel}$ =26.83 bar; See Table 6 for other parameter values. |
| Heat recovery steam generator (HRSG) | 3 Pressure Levels: high pressure (HP) steam = 124.1 bar/565.6$^oC$, intermediate pressure (IP) steam (after reheat) = 23.6 bar/565.6$^oC$, low pressure (LP) steam = 2.413 bar; Flue gas $T_{gas,out}$ = 90 to 100°C; $T_{water,in}$ = 15°C; Steam-drums blowdown: 1% of inlet flow; Deaerator vent = 0.5% of inlet flow; Δp/p = 5%, except IP superheaters (≤ 1%). |
| Combustor (feeding power island boiler/ HRSG in RC cases) | Gibbs reactor; $P_{in}$ = 1.2 bar; Δp = 0.1 bar;<br>Excess air coefficient=1.2; Air preheated to ~300$^oC$. |
| Steam turbine | 3 Stages: HP inlet steam @ 124.1 bar/565.6$^oC$, IP steam @ 23.6 bar/565.6$^oC$, LP steam @ 2.413 bar; LP turbine exhaust = 0.0462 bar; Vapor fraction at LP turbine exit ≥ 85%; $\eta_{isentropic}$ for HP, IP, LP = 0.84, 0.89, 0.85; $\eta_{mechanical}$ = 0.98. |
| Heat Exchangers | Δp/p = 2%; $\Delta T_{min}$ = 15°C (gas-liq), 30 °C (gas-gas); heat loss = 1% to 2% of heat transferred |
| Pumps | Centrifugal; $\eta_{polytropic}$ = 0.65 to 0.9; $\eta_{mechanical}$ =1 |
| Compressors | $\eta_{polytropic}$ = 0.80; $\eta_{mechanical}$ = 0.90 |
| Multistage compressors | $\eta_{Isentropic}$ = 0.76 – 0.82; $\eta_{mech}$ = 0.94; $\Delta p_{stage}$ = same for each stage; maximum stages = 5;<br>$T_{out}$ ≤ 250°C for each stage; inter-stage cooling to T = 25-35°C. |

**Table 6. Quoted and simulated performance of the GE 7FB gas turbine.**

| | *GE Quoted*[a] | *Our Aspen Simulations* | |
|---|---|---|---|
| | *Natural gas*[a] | *Natural gas*[a] | *Syngas* |
| Natural Gas Flow (kg/s) | 9.985 | 9.985 | 0 |
| Syngas flow rate (kg/s) | 0 | 0 | 99.71 |
| Air Mass Flow (kg/s) | 438.07 | 438.07 | 571.64 |
| Compressor Pressure Ratio | 18.5 | 18.5 | 19.5 |
| Net Electric Output ($MW_e$) | 184.4 | 184.43 | 267.5 |
| Exhaust Temperature, C | 623 | 623 | 650.1 |
| Thermal Efficiency (%) | 36.92 | 36.92 | 37.9 |
| Exhaust Flow (kg/s) | 448.06 | 448.06 | 628.4 |
| Turbine Inlet Temperature ($^oC$) | 1370 | 1370 | 1370 |
| GT Air Filter Pressure drop, bar | N/A | 0 | 0 |
| GT Compressor Polytropic Efficiency, % [b] | N/A | 87 | 87 |
| GT Compressor Mech Efficiency, % | N/A | 98.65 | 98.65 |
| Air Leakage, % | N/A | 0.1 | 0.1 |
| Cooling flow bypass% | N/A | 5.161 | 5.161 |
| Combustion Heat Loss, % fuel LHV | N/A | 0.5 | 0.5 |
| GT Turbine Isentropic Efficiency, % [b] | N/A | 89.769 | 89.769 |
| GT Turbine Mech Efficiency, % | N/A | 98.65 | 98.65 |
| Generator Efficiency | N/A | 98.6 | 98.6 |
| GT Exhaust Pressure, bar | N/A | 1.01 | 1.065 |

(a) Quoted performance is from Brooks [41], Scholz [42], and Cerovski (GE Power Systems, Schenectady, personal communication, 2003). For the parameters shown below turbine inlet temperature, representative values have been assumed based on discussion with Stefano Consonni (Politecnico di Milano, Italy, personal communication, 2004).

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

## 3.2 Mass and energy balance simulation results

Table 7 and Table 8 summarize the mass and energy balance simulation results for all 6 RC cases and all 10 OT process designs, respectively. (Table 9 and Table 10 provide additional detail on the electricity production and consumption for each design.) Detailed process flow diagrams and tabulations of process flow simulations for several RC and OT cases examined in this paper are presented in subsections 3.2.1 and 3.2.2, respectively.

It should be noted that the convention we use here in reporting volumetric production rates of FTL fuels (e.g., barrels per day) is to report the volumetric rates from displacing an equivalent amount of energy (LHV basis) in crude oil products.[6] Others, e.g. Gray, *et al.* [17] report actual FTL volumetric flow rates. Actual FTL volumetric flow rates are higher than the reported volumetric flow rates of the crude oil products displaced (COPD), because COPDs have higher volumetric energy densities than FTL fuels (Table 4).

It should also be pointed out that, unlike most other studies, we simulate the production of finished gasoline and diesel blendstocks, rather than production of naphtha and diesel (see Section 2.1.5).

### 3.2.1 Recycle FTL systems

Figure 17 and Table 11 describe the detailed material balance for the CTL-RC-V case. Overall, 43.3% of the input coal energy (LHV basis) is converted into liquid fuels (61% diesel and 39% gasoline) and 5.9% into exportable electricity, giving a total 1$^{st}$ law efficiency of 49.1%, LHV basis (Table 7). With $CO_2$ capture and storage incorporated (CTL-RC-CCS), the process flow diagram remains largely unchanged from the CTL-RC-V diagram: fuel outputs are the same, but electricity output falls about 26% due to added $CO_2$ compression requirements for pipeline delivery to a storage site, at a rate of 90.5 kWh per tonne of captured $CO_2$. The resulting 1$^{st}$ law efficiency is 47.6%, LHV basis (Table 7). The amount of carbon captured and stored, 332 tC/hr, represents 51.5% of the carbon in the input coal.

Figure 18 and Table 12 show the detailed material balance for the BTL-RC-V design, which converts 46.2% of the input biomass energy (LHV basis) into FTL fuels and 5.7% into exportable electricity, giving a total 1$^{st}$ law efficiency of 51.9% (Table 7). For the BTL-RC-CCS design, fuel outputs are unchanged from the BTL-RC-V case, but net electricity output falls about 30%, resulting in a total 1$^{st}$ law efficiency of 50.3% (Table 7). The amount of captured carbon, 30.6 tC/hr, is 51.4% of the carbon in the input biomass.

Figure 19 and Table 13 shows the detailed material balance for the CBTL-RC-V case, in which 44.2 % of the input coal+biomass energy (LHV basis) is converted into FTL fuels and 6.8% into exported electricity, for a total 1$^{st}$ law efficiency of 51.1% (Table 7). With $CO_2$ capture and storage (CBTL-RC-CCS), fuel outputs are unchanged from the CBTL-RC-V case, but electricity output falls about 23%, giving a total 1$^{st}$ law efficiency of 49.5% (Table 7). The amount of carbon captured, 71.5 tC/hr, is 54.0% of the carbon in the input feedstocks.

**Table 7. Summary of mass and energy balances for all RC process designs examined in this work.**

| | *CTL-RC-V* | *CTL-RC-CCS* | *CBTL-RC-V* | *CBTL-RC-CCS* | *BTL-RC-V* | *BTL-RC-CCS* |
|---|---|---|---|---|---|---|
| ***INPUTS*** | | | | | | |
| **Coal feed capacity** | | | | | | |
| As-received, metric t/day | 24,297 | 24297 | 2747 | 2747 | - | - |
| Dry, metric t/d | 21,595 | 21,595 | 2,441 | 2,441 | - | - |
| Moisture and ash free (MAF) metric t/d | 19,238 | 19,238 | 2,175 | 2,175 | - | - |
| Coal, MW LHV | 7,272 | 7,272 | 822 | 822 | - | - |
| Coal, MW HHV | 7,625 | 7,625 | 862 | 862 | - | - |
| **Biomass feed capacity** | | | | | | |
| As-received metric t/day | - | - | 3,581 | 3,581 | 3,578 | 3,581 |
| Dry, metric t/d | - | - | 3,044 | 3,044 | 3,041 | 3,044 |
| Moisture and ash free (MAF) t/d | - | - | 2,822 | 2,822 | 2,820 | 2,822 |
| Biomass, MW LHV | - | - | 601 | 601 | 601 | 601 |
| Biomass, MW HHV | - | - | 661 | 661 | 660 | 661 |
| % biomass HHV basis | - | - | 43.4% | 43.4% | 100% | 100% |
| **Oxygen feed capacity** | | | | | | |
| Total Oxidant, metric t/d | 921 | 921 | 165 | 165 | 59 | 59 |
| Oxidant purity (mass% O2) | 99.54% | 99.54% | 99.54% | 99.54% | 99.54% | 99.54% |
| Oxidant purity (mol% O2) | 99.50% | 99.50% | 99.50% | 99.50% | 99.50% | 99.50% |
| Pure O2, metric t/d | 917 | 917 | 164 | 164 | 58 | 58 |
| ***OUTPUTS*** | | | | | | |
| **FTL production capacity** | | | | | | |
| FT diesel, MW LHV | 1,907 | 1,907 | 381 | 382 | 169 | 169 |
| FT gasoline, MW LHV | 1,240 | 1,240 | 248 | 248 | 109 | 109 |
| FT diesel, MW HHV | 2,050 | 2,050 | 410 | 410 | 182 | 182 |
| FT gasoline, MW HHV | 1,336 | 1,336 | 268 | 268 | 117 | 117 |
| FT diesel, bbl/day crude-oil derived diesel equiv | 28,712 | 28,712 | 5,744 | 5,745 | 2,543 | 2,546 |
| FT gasoline, bbl/day crude-oil derived gasoline equiv | 21,288 | 21,288 | 4,262 | 4,262 | 1,867 | 1,869 |
| TOTAL FTL out, MW LHV | 3,147 | 3,147 | 630 | 630 | 278 | 278 |
| TOTAL FTL out, MW HHV | 3,387 | 3,387 | 678 | 678 | 299 | 299 |
| TOTAL FTL out, bbl/day crude oil products displaced (COPD) | 50,000 | 50,000 | 10,006 | 10,007 | 4,409 | 4,415 |
| **Electricity** | | | | | | |
| Gross production, MW | 874 | 874 | 178 | 178 | 66 | 66 |
| On-site consumption, MW | 447 | 557 | 81 | 104 | 32 | 42 |
| Net export to grid, MW | 427 | 317 | 97 | 75 | 34 | 24 |
| ***ENERGY RATIOS*** | | | | | | |
| **Fraction of input energy (LHV) converted to** | | | | | | |
| FTL (LHV) | 43.3% | 43.3% | 44.2% | 44.2% | 46.2% | 46.2% |
| Exportable electricity | 5.9% | 4.4% | 6.8% | 5.2% | 5.7% | 4.0% |
| FTL (LHV) + exported electricity | 49.1% | 47.6% | 51.1% | 49.5% | 51.9% | 50.2% |
| **Fraction of input energy (HHV) converted to** | | | | | | |
| FTL (HHV) | 44.4% | 44.4% | 44.5% | 44.5% | 45.3% | 45.3% |
| Exportable electricity | 5.6% | 4.2% | 6.4% | 4.9% | 5.2% | 3.7% |
| FTL (HHV) + electricity | 50.0% | 48.6% | 50.9% | 49.4% | 50.5% | 49.0% |
| **FTL Yields** | | | | | | |
| Liters gasoline equiv (lge) per dry metric t biomass | | | 565 | 565 | 249 | 249 |
| Yield relative to cellulosic ethanol | | | 2.3 | 2.3 | 1.0 | 1.0 |
| **Gross power from steam and gas turbines** | | | | | | |
| Steam turbine power/gas turbine power | - | - | - | - | - | - |
| Steam turbine MW / FTL MW LHV | 0.257 | 0.257 | 0.273 | 0.273 | 0.237 | 0.237 |
| ***CARBON ACCOUNTING*** | | | | | | |
| **Plant Carbon Flows** | | | | | | |
| C input as feedstock, kgC/second | 179 | 179 | 37 | 37 | 17 | 17 |
| C stored as CO2, % of feedstock C | 0 | 51.5% | 0 | 54.0% | 0 | 51.4% |
| C in char (unburned), % of feedstock C | 5.0% | 5.0% | 7.2% | 7.2% | 9.8% | 9.8% |
| C vented to atmosphere, % of feedstock C | 60.3% | 8.9% | 61.7% | 7.7% | 57.2% | 5.8% |
| C in FTL, % of feedstock C | 33.7% | 33.7% | 32.8% | 32.8% | 32.2% | 32.2% |
| Closure on carbon balance | 99.0% | 99.0% | 101.7% | 101.7% | 99.3% | 99.3% |
| **Carbon Storage Rates** | | | | | | |
| tCO2 per hour | 0 | 1217 | 0 | 262 | 0 | 112 |
| million tCO2 per year (with 90% capacity factor) | 0 | 9.60 | 0 | 2.07 | 0 | 0.88 |
| tCO2 per barrel of FTL | 0 | 0.58 | 0 | 0.63 | 0 | 0.61 |
| C stored as CO2, % of feedstock C not in FTL | | 77.6% | | 80.4% | | 75.8% |

**Table 8. Summary of mass and energy balances for all OT process designs examined in this work.**

| | *CTL-OT-V* | *CTL-OT-CCS* | *CTL-OTA-CCS* | *CBTL-OT-V* | *CBTL-OT-CCS* | *CBTL2-OT-CCS* | *CBTL-OTS-CCS* | *CBTL-OTA-V* | *CBTL-OTA-CCS* | *CBTL-OTAS-CCS* |
|---|---|---|---|---|---|---|---|---|---|---|
| ***INPUTS*** | | | | | | | | | | |
| **Coal feed capacity** | | | | | | | | | | |
| As-received, metric t/day | 24297 | 24297 | 24297 | 3419 | 3418 | 22343 | 6689 | 6702 | 6702 | 10710 |
| Dry, metric t/d | 21,595 | 21,595 | 21,595 | 3,039 | 3,038 | 19,859 | 5,945 | 5,957 | 5,957 | 9,519 |
| Moisture and ash free (MAF) metric t/d | 19,238 | 19,238 | 19,238 | 2,707 | 2,706 | 17,691 | 5,296 | 5,307 | 5,307 | 8,480 |
| Coal, MW LHV | 7,272 | 7,272 | 7,272 | 1,023 | 1,023 | 6,688 | 2,002 | 2,006 | 2,006 | 3,206 |
| Coal, MW HHV | 7,625 | 7,625 | 7,625 | 1,073 | 1,073 | 7,012 | 2,099 | 2,103 | 2,103 | 3,361 |
| **Biomass feed capacity** | | | | | | | | | | |
| As-received metric t/day | 0 | 0 | 0 | 3,579 | 3,578 | 3,581 | 3,581 | 3,581 | 3,581 | 3,581 |
| Dry, metric t/d | 0 | 0 | 0 | 3,042 | 3,041 | 3,044 | 3,044 | 3,044 | 3,044 | 3,044 |
| Moisture and ash free (MAF) t/d | 0 | 0 | 0 | 2,821 | 2,820 | 2,822 | 2,822 | 2,822 | 2,822 | 2,822 |
| Biomass, MW LHV | 0 | 0 | 0 | 601 | 601 | 601 | 601 | 601 | 601 | 601 |
| Biomass, MW HHV | 0 | 0 | 0 | 660 | 660 | 661 | 661 | 661 | 661 | 661 |
| % biomass HHV basis | 0 | 0 | 0 | 38.1% | 38.1% | 8.61% | 23.9% | 23.9% | 23.9% | 16.4% |
| **Oxygen feed capacity** | | | | | | | | | | |
| Total Oxidant, metric t/d | 830.3 | 830.3 | 971.7 | 161.5 | 161.4 | 682.4 | 273.3 | 273.7 | 328.6 | 489.0 |
| Oxidant purity (mass% O2) | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% | 93.43% |
| Oxidant purity (mol% O2) | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% | 94.34% |
| Pure O2, metric t/d | 776 | 776 | 908 | 151 | 151 | 638 | 255 | 256 | 307 | 457 |
| ***OUTPUTS*** | | | | | | | | | | |
| **FTL production capacity** | | | | | | | | | | |
| FT diesel, MW LHV | 1,397 | 1,397 | 1,397 | 309 | 309 | 1,397 | 497 | 497 | 497 | 727 |
| FT gasoline, MW LHV | 910 | 910 | 910 | 201 | 201 | 910 | 324 | 325 | 324 | 475 |
| FT diesel, MW HHV | | | | | | | | | | |
| FT gasoline, MW HHV | | | | | | | | | | |
| FT diesel, bbl/day crude-oil derived diesel equiv | 21,040 | 21,040 | 21,040 | 4,649 | 4,648 | 21,038 | 7,477 | 7,489 | 7,489 | 10,951 |
| FT gasoline, bbl/day crude-oil derived gasoline equiv | 15,616 | 15,615 | 15,615 | 3,453 | 3,453 | 15,617 | 5,561 | 5,570 | 5,569 | 8,156 |
| TOTAL FTL out, MW LHV | 2,307 | 2,307 | 2,307 | 510 | 510 | 2,307 | 821 | 822 | 822 | 1,202 |
| TOTAL FTL out, MW HHV | 2,483 | 2,483 | 2,483 | 549 | 549 | 2,483 | 883 | 885 | 884 | 1,294 |
| TOTAL FTL out, bbl/day crude oil products displaced (COPD) | 36,655 | 36,655 | 36,655 | 8,102 | 8,100 | 36,655 | 13,039 | 13,060 | 13,058 | 19,107 |
| **Electricity** | | | | | | | | | | |
| Gross production, MW | 1,673 | 1,664 | 1,468 | 392 | 399 | 1,696 | 609 | 618 | 541 | 782 |
| On-site consumption, MW | 393 | 589 | 650 | 78 | 124 | 583 | 203 | 131 | 229 | 337 |
| Net export to grid, MW | 1,279 | 1,075 | 818 | 315 | 276 | 1,113 | 406 | 487 | 312 | 445 |
| ***ENERGY RATIOS*** | | | | | | | | | | |
| **Fraction of input energy (LHV) converted to** | | | | | | | | | | |
| FTL (LHV) | 31.7% | 31.7% | 31.7% | 31.4% | 31.4% | 31.7% | 31.5% | 31.5% | 31.5% | 31.6% |
| Exportable electricity | 17.6% | 14.8% | 11.2% | 19.4% | 17.0% | 15.3% | 15.6% | 18.7% | 12.0% | 11.7% |
| FTL (LHV) + exported electricity | 49.3% | 46.5% | 43.0% | 50.8% | 48.4% | 46.9% | 47.1% | 50.2% | 43.5% | 43.3% |
| **Fraction of input energy (HHV) converted to** | | | | | | | | | | |
| FTL (HHV) | 32.6% | 32.6% | 32.6% | 31.7% | 31.7% | 32.4% | 32.0% | 32.0% | 32.0% | 32.2% |
| Exportable electricity | 16.8% | 14.1% | 10.7% | 18.2% | 15.9% | 14.5% | 14.7% | 17.6% | 11.3% | 11.1% |
| FTL (HHV) + electricity | 49.3% | 46.7% | 43.3% | 49.8% | 47.6% | 46.9% | 46.7% | 49.6% | 43.3% | 43.2% |
| **FTL Yields** | | | | | | | | | | |
| Liters gasoline equiv (lge) per dry metric t biomass | | | | 457 | 457 | | 736 | 737 | 737 | 1,078 |
| Yield relative to cellulosic ethanol | | | | 1.8 | 1.8 | | 2.9 | 3.0 | 3.0 | 4.3 |
| **Gross power from steam and gas turbines** | | | | | | | | | | |
| Steam turbine power/gas turbine power | 1.225 | 1.115 | 1.453 | 1.180 | 1.113 | 1.115 | 1.061 | 1.192 | 1.371 | 1.387 |
| Steam turbine MW / FTL MW LHV | 0.384 | 0.366 | 0.361 | 0.408 | 0.404 | 0.375 | 0.372 | 0.398 | 0.369 | 0.365 |
| ***CARBON ACCOUNTING*** | | | | | | | | | | |
| **Plant Carbon Flows** | | | | | | | | | | |
| C input as feedstock, kgC/second | 179 | 179 | 179 | 42 | 42 | 181 | 66 | 66 | 66 | 96 |
| C stored as CO2, % of feedstock C | 0.00% | 51.06% | 64.10% | 0.00% | 50.95% | 51.03% | 50.87% | 0.00% | 64.62% | 64.46% |
| C in char (unburned), % of feedstock C | 5.00% | 5.00% | 5.00% | 6.91% | 6.91% | 5.42% | 6.21% | 6.21% | 6.21% | 5.83% |
| C vented to atmosphere, % of feedstock C | 69.56% | 18.56% | 5.47% | 69.06% | 18.17% | 18.47% | 18.42% | 69.24% | 4.62% | 4.89% |
| C in FTL, % of feedstock C | 24.68% | 24.68% | 24.68% | 23.42% | 23.42% | 24.40% | 23.88% | 23.89% | 23.88% | 24.13% |
| Closure on carbon balance | 99.25% | 99.30% | 99.25% | 99.39% | 99.45% | 99.33% | 99.39% | 99.34% | 99.33% | 99.32% |
| **Carbon Storage Rates** | | | | | | | | | | |
| tCO2 per hour | 0 | 1207 | 1515 | 0 | 280 | 1221 | 442 | 0 | 562 | 812 |
| million tCO2 per year (with 90% capacity factor) | 0 | 9.52 | 11.95 | 0 | 2.21 | 9.63 | 3.49 | 0 | 4.44 | 6.41 |
| tCO2 per barrel of FTL | 0 | 0.79 | 0.99 | 0 | 0.83 | 0.80 | 0.81 | 0 | 1.03 | 1.02 |
| C stored as CO2, % of feedstock C not in FTL | | 67.8% | 85.1% | | 66.5% | 67.5% | 66.8% | | 84.9% | 85.0% |

**Table 9. Electricity balance for RC process designs.**

| | CTL-RC-V | CTL-RC-CCS | CBTL-RC-V | CBTL-RC-CCS | BTL-RC-V | BTL-RC-CCS |
|---|---|---|---|---|---|---|
| ***Electricity Production, MW*** | | | | | | |
| WGS bypass expander | 6.34 | 6.34 | 0.82 | 0.82 | 0 | 0 |
| Pre-rectisol expander | 57.68 | 57.68 | 5.57 | 5.57 | 0 | 0 |
| H2 expander (in FT refinery) | 1.28 | 1.28 | 0 | 0 | 0 | 0 |
| Steam turbine gross output | 808.50 | 808.50 | 171.69 | 171.84 | 65.86 | 65.93 |
| Gas turbine output | - | - | - | - | - | - |
| **Total electricity generation** | **873.80** | **873.80** | **178.34** | **178.49** | **65.99** | **66.05** |
| ***On-site consumption, MW*** | | | | | | |
| Biomass handling | 0.00 | 0.00 | 0.41 | 0.41 | 0.41 | 0.41 |
| Coal handling | 21.09 | 21.09 | 2.38 | 2.38 | 0.00 | 0.00 |
| Lock Hopper | 0.00 | 0.00 | 0.35 | 0.35 | 0.35 | 0.35 |
| Rectisol process power | 16.11 | 16.11 | 3.34 | 3.34 | 1.51 | 1.52 |
| Rectisol refrigeration power | 22.49 | 22.49 | 4.50 | 4.50 | 2.21 | 2.21 |
| Rectisol recovery compressor | 1.50 | 1.50 | 0.33 | 0.33 | 0.14 | 0.14 |
| Fuel gas comp. inside refinery | 1.10 | 1.10 | 0.21 | 0.21 | 0.08 | 0.08 |
| CO2 compression to 3 bar | 0.00 | 16.96 | 0.74 | 3.65 | 0.00 | 1.74 |
| CO2 comp to lockhopper pressure | 0.00 | 0.00 | 0.00 | 13.88 | 0.74 | 6.60 |
| CO2 compression to supercritical | 0.00 | 93.18 | 0.00 | 5.98 | 0.00 | 2.63 |
| H2-feed compressor | 2.03 | 2.03 | 0.41 | 0.41 | 0.18 | 0.18 |
| FT fresh feed compressor | 21.34 | 21.34 | 4.27 | 4.27 | 1.95 | 1.95 |
| GT fuel gas compressor | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| Recycle compressor | 15.88 | 15.88 | 3.12 | 3.12 | 1.21 | 1.21 |
| F-T Refinery | 2.04 | 2.04 | 0.41 | 0.41 | 0.18 | 0.18 |
| Water cycle pumps, total | 4.46 | 4.46 | 0.50 | 0.50 | 0.00 | 0.00 |
| Steam cycle pumps, total | 9.94 | 9.94 | 2.40 | 2.40 | 1.27 | 1.27 |
| ASU power (delivering at 2 bar) | 197.05 | 197.05 | 36.73 | 36.73 | 14.78 | 14.79 |
| HP O2 compressor (for C gasif) | 115.91 | 115.91 | 12.36 | 12.36 | 0.00 | 0.00 |
| IP O2 comp (for B-gasf + ATR) | 15.71 | 15.71 | 8.64 | 8.64 | 6.55 | 6.56 |
| N2 compressor (to GT) | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| **Total electricity consumption** | **446.66** | **556.80** | **81.10** | **103.86** | **31.57** | **41.83** |
| **Total net sales to grid (MW)** | **427.13** | **316.99** | **97.25** | **74.63** | **34.42** | **24.23** |

**Table 10. Electricity balance for OT process designs.**

| | CTL-OT-V | CTL-OT-CCS | CTL-OTA-CCS | CBTL-OT-V | CBTL-OT-CCS | CBTL2-OT-CCS | CBTL-OTS-CCS | CBTL-OTA-V | CBTL-OTA-CCS | CBTL-OTAS-CCS |
|---|---|---|---|---|---|---|---|---|---|---|
| ***Electricity Production, MW*** | | | | | | | | | | |
| WGS bypass expander | 6 | 6 | 6 | 1 | 1 | 6 | 2 | 2 | 2 | 3 |
| Pre-rectisol expander | 55 | 55 | 55 | 7 | 7 | 50 | 15 | 15 | 15 | 24 |
| H2 expander (in FT refinery) | 1 | 1 | 1 | 0 | 0 | 1 | 0 | 0 | 0 | 1 |
| Steam turbine gross output | 886.84 | 844.67 | 833.09 | 207.84 | 206.01 | 864.04 | 304.95 | 326.94 | 303.37 | 438.54 |
| Gas turbine output | 724.0 | 757.2 | 573.4 | 176.1 | 185.1 | 775.0 | 287.5 | 274.2 | 221.2 | 316.1 |
| **Total electricity generation** | | | | | | | | | | |
| ***On-site consumption, MW*** | | | | | | | | | | |
| Biomass handling | 0.00 | 0.00 | 0.00 | 0.41 | 0.41 | 0.41 | 0.41 | 0.41 | 0.41 | 0.41 |
| Coal handling | 21.09 | 21.09 | 21.09 | 2.97 | 2.97 | 19.39 | 5.81 | 5.82 | 5.82 | 9.30 |
| Lock Hopper | 0.00 | 0.00 | 0.00 | 0.35 | 0.35 | 0.35 | 0.35 | 0.35 | 0.35 | 0.35 |
| Rectisol process power | 8.30 | 16.00 | 19.70 | 2.06 | 3.75 | 16.21 | 5.89 | 3.14 | 7.34 | 10.59 |
| Rectisol refrigeration power | 14.46 | 21.17 | 23.31 | 3.23 | 4.73 | 21.22 | 7.41 | 5.06 | 8.35 | 12.29 |
| Rectisol recovery compressor | 1.28 | 1.48 | 1.41 | 0.29 | 0.35 | 1.50 | 0.55 | 0.47 | 0.52 | 0.75 |
| Fuel gas comp. inside refinery | 0.74 | 0.74 | 0.74 | 0.16 | 0.16 | 0.74 | 0.27 | 0.27 | 0.26 | 0.39 |
| CO2 compression to 3 bar | 0.00 | 16.82 | 21.12 | 0.00 | 4.08 | 17.19 | 6.33 | 0.00 | 8.01 | 11.50 |
| CO2 comp to lockhopper pressure | 0.00 | 89.11 | 111.87 | 0.74 | 15.51 | 84.93 | 23.94 | 0.74 | 30.44 | 43.68 |
| CO2 compression to supercritical | 0.00 | 3.37 | 4.15 | 0.00 | 6.58 | 9.21 | 10.10 | 0.00 | 13.20 | 19.07 |
| H2-feed compressor | 1.49 | 1.49 | 1.49 | 0.33 | 0.33 | 1.49 | 0.53 | 0.53 | 0.53 | 0.78 |
| FT fresh feed compressor | 15.58 | 15.58 | 15.58 | 3.44 | 3.44 | 15.58 | 5.54 | 5.55 | 5.55 | 8.12 |
| GT fuel gas compressor | 17.87 | 17.86 | 17.86 | 3.94 | 3.94 | 17.85 | 6.36 | 6.38 | 6.34 | 9.30 |
| Recycle compressor | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| F-T Refinery | 1.50 | 1.50 | 1.50 | 0.33 | 0.33 | 1.50 | 0.53 | 0.53 | 0.53 | 0.78 |
| Water cycle pumps, total | 4.66 | 4.75 | 4.75 | 0.62 | 0.64 | 4.34 | 1.23 | 1.19 | 1.21 | 1.87 |
| Steam cycle pumps, total | 7.63 | 7.22 | 8.84 | 2.01 | 1.64 | 7.27 | 2.97 | 2.98 | 3.74 | 5.13 |
| ASU power (delivering at 2 bar) | 178.38 | 178.38 | 213.62 | 35.51 | 35.49 | 174.45 | 60.07 | 60.18 | 72.23 | 107.50 |
| HP O2 compressor (for C gasif) | 120.29 | 120.29 | 116.13 | 16.34 | 16.33 | 110.03 | 31.97 | 32.04 | 32.04 | 51.19 |
| IP O2 comp (for B-gasf + ATR) | 0.00 | 0.00 | 15.67 | 4.95 | 4.95 | 4.95 | 4.95 | 4.95 | 11.03 | 13.63 |
| N2 compressor (to GT) | 0.00 | 72.09 | 51.35 | 0.00 | 17.88 | 74.04 | 28.05 | 0.00 | 21.25 | 30.28 |
| **Total electricity consumption** | **393.28** | **588.95** | **650.18** | **77.70** | **123.87** | **582.66** | **203.26** | **130.59** | **229.17** | **336.89** |
| **Total net sales to grid (MW)** | **1,279.24** | **1,074.68** | **818.05** | **314.69** | **275.62** | **1,112.85** | **405.88** | **487.29** | **312.30** | **444.83** |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 281.2 | 1.0 | 25.0 |
| 2 | 94.7 | 1.0 | 25.0 |
| 3 | 255.8 | 1.6 | 9.0 |
| 4 | 1164.1 | 71.5 | 251.6 |
| 5 | 150.0 | 71.0 | 250.4 |
| 6 | 150.0 | 68.9 | 417.5 |
| 7 | 989.8 | 71.0 | 250.4 |
| 8 | 525.0 | 29.5 | 40.0 |
| 9 | 489.1 | 28.0 | 39.7 |
| 10 | 491.0 | 28.0 | 39.7 |
| 11 | 225.4 | 1.2 | 25.0 |
| 12 | 112.7 | 3.0 | 25.0 |
| 13 | 33.3 | 1.8 | 25.0 |
| 14 | 491.0 | 20.1 | 27.3 |
| 15 | 483.3 | 20.1 | 27.3 |
| 16 | 483.3 | 24.4 | 245.0 |
| 17 | 483.3 | 23.2 | 260.0 |
| 18 | 483.3 | 23.2 | 73.7 |
| 19 | 483.3 | 23.2 | 40.0 |
| 20 | 366.2 | 19.7 | 40.0 |
| 21 | 5.7 | 3.2 | 45.9 |
| 22 | 28.0 | 2.6 | 58.6 |
| 23 | 43.3 | 1.4 | 129.4 |
| 24 | 1.6 | 7.4 | 40.0 |
| 25 | 1.6 | 21.1 | 184.3 |
| 26 | 1.6 | 20.6 | 40.0 |
| 27 | 1.0 | 20.1 | 24.8 |
| 28 | 1.0 | 4.1 | -64.3 |
| 29 | 1.0 | 4.0 | 25.0 |
| 30 | 8.3 | 1.5 | 24.8 |
| 31 | 8.3 | 3.2 | 118.6 |
| 32 | 21.3 | 1.2 | 100.0 |
| 33 | 358.9 | 19.6 | 122.0 |
| 34 | 358.9 | 26.9 | 159.5 |
| 35 | 40.2 | 29.1 | 185.1 |
| 36 | 213.8 | 38.0 | 246.7 |
| 37 | 613.0 | 29.0 | 980.4 |
| 38 | 613.0 | 28.0 | 40.0 |

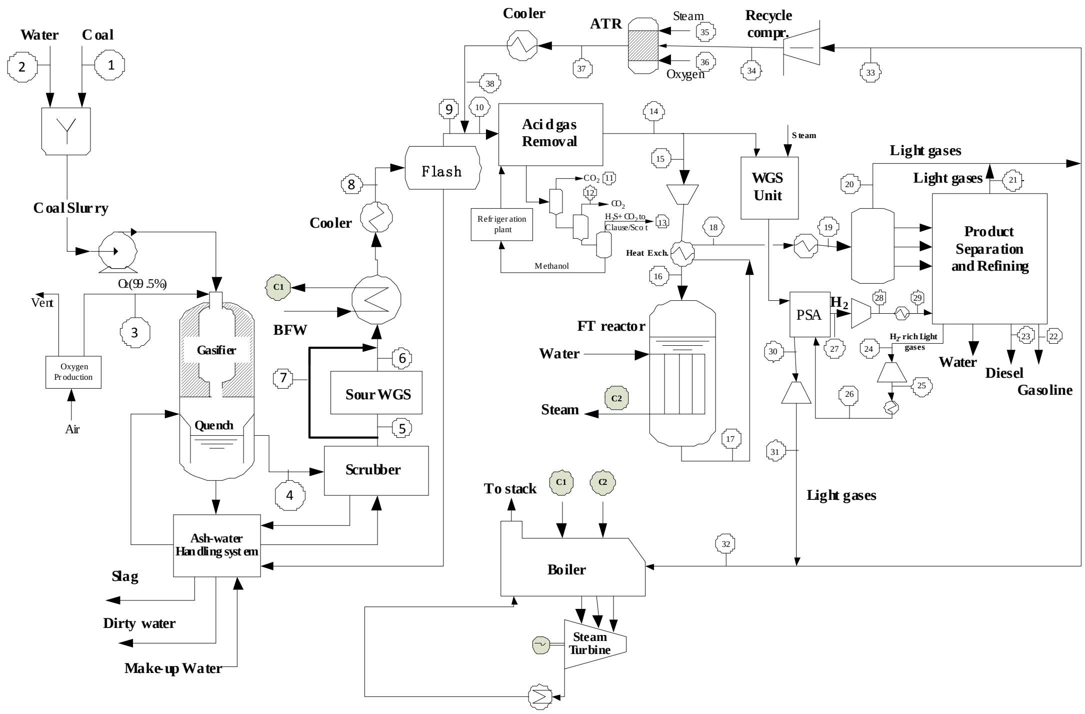


**Figure 17. Process flow simulation results for CTL-RC-V design.**

**Table 11. Stream data for CTL-RC-V simulation. Stream numbers refer to Figure 17.**

| | Scrubber Exit | Water Gas Shift reactor, | Water Gas Shift reactor, Outlet | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Recycle gas feed into | Raw synthesis product | Final Liquid product: Gasoline | Final Liquid product: Diesel | Fuel gas to Gas turbine |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Number | **5+7** | **5** | **6** | **10** | **14** | **16** | **38** | **19** | **22** | **23** | **32** |
| Total Flow kg/sec | 1139.8 | 150.0 | 150.0 | 864.4 | 491.0 | 483.3 | 375.3 | 483.3 | 28.0 | 43.4 | 21.3 |
| Total Flow kmol/sec | 60.1 | 7.9 | 7.9 | 36.5 | 27.9 | 27.5 | 12.6 | 14.7 | 0.2 | 0.2 | 0.7 |
| Mole Frac | | | | | | | | | | | |
| CO | 19.62% | 19.62% | 1.63% | 31.59% | 41.29% | 41.29% | 9.39% | 1.45% | 0.00% | 0.00% | 26.44% |
| H2 | 13.25% | 13.25% | 31.25% | 31.35% | 40.98% | 40.98% | 16.52% | 15.52% | 0.00% | 0.00% | 19.16% |
| CO2 | 3.94% | 3.94% | 21.94% | 23.34% | 0.92% | 0.92% | 37.86% | 32.37% | 0.00% | 0.00% | 14.08% |
| H2O | 62.50% | 62.50% | 44.50% | 0.27% | 0.00% | 0.00% | 0.29% | 14.58% | 0.00% | 0.00% | 0.00% |
| CH4 | 0.01% | 0.01% | 0.01% | 0.02% | 0.03% | 0.03% | 0.00% | 0.51% | 0.00% | 0.00% | 0.23% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.25% | 0.00% | 0.00% | 15.88% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 100.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.36% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.44% | 0.00% | 0.00% | 0.60% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.67% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.20% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.84% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.07% | 0.07% | 0.00% | 0.13% | 0.00% | 0.00% | 0.10% |
| N2 | 0.26% | 0.26% | 0.26% | 12.36% | 16.16% | 16.16% | 34.72% | 30.27% | 0.00% | 0.00% | 22.73% |
| AR | 0.01% | 0.01% | 0.01% | 0.43% | 0.56% | 0.56% | 1.21% | 1.05% | 0.00% | 0.00% | 0.79% |
| H2S | 0.34% | 0.34% | 0.34% | 0.54% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.04% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.06% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 41.4 | 1.0 | 25.0 |
| 2 | 2.6 | 31.7 | 236.6 |
| 3 | 10.6 | 31.4 | 152.1 |
| 4 | 1.0 | 31.4 | 152.1 |
| 5 | 55.3 | 28.8 | 800.0 |
| 6 | 55.3 | 28.2 | 350.0 |
| 7 | 55.3 | 28.0 | 40.0 |
| 8 | 48.0 | 26.8 | 40.0 |
| 9 | 32.7 | 20.1 | 27.2 |
| 10 | 3.5 | 3.0 | 25.0 |
| 11 | 3.5 | 31.4 | 123.0 |
| 12 | 3.5 | 31.4 | 40.0 |
| 13 | 32.7 | 24.4 | 245.0 |
| 14 | 32.7 | 23.2 | 260.0 |
| 15 | 32.7 | 23.2 | 40.0 |
| 16 | 18.6 | 19.7 | 40.0 |
| 17 | 0.5 | 3.2 | 45.9 |
| 18 | 2.5 | 2.6 | 58.5 |
| 19 | 3.8 | 1.4 | 129.9 |
| 20 | 0.1 | 7.4 | 40.0 |
| 21 | 0.1 | 21.1 | 184.3 |
| 22 | 0.1 | 20.6 | 25.0 |
| 23 | 0.1 | 20.1 | 24.9 |
| 24 | 0.1 | 4.1 | -64.2 |
| 25 | 0.1 | 4.0 | 25.0 |
| 26 | 0.7 | 20.6 | 24.9 |
| 27 | 0.6 | 1.5 | 24.9 |
| 28 | 0.6 | 3.2 | 115.2 |
| 29 | 17.7 | 19.6 | 122.0 |
| 30 | 17.7 | 26.9 | 159.8 |
| 31 | 4.7 | 29.1 | 185.1 |
| 32 | 11.2 | 38.0 | 246.7 |
| 33 | 33.6 | 28.2 | 40.0 |
| 34 | 20.4 | 28.0 | 40.0 |
| 35 | 68.4 | 28.0 | 40.0 |
| 36 | 23.1 | 1.2 | 25.0 |
| 37 | 11.5 | 3.0 | 25.0 |
| 38 | 0.5 | 1.8 | 25.0 |
| 39 | 2.0 | 1.2 | 104.0 |

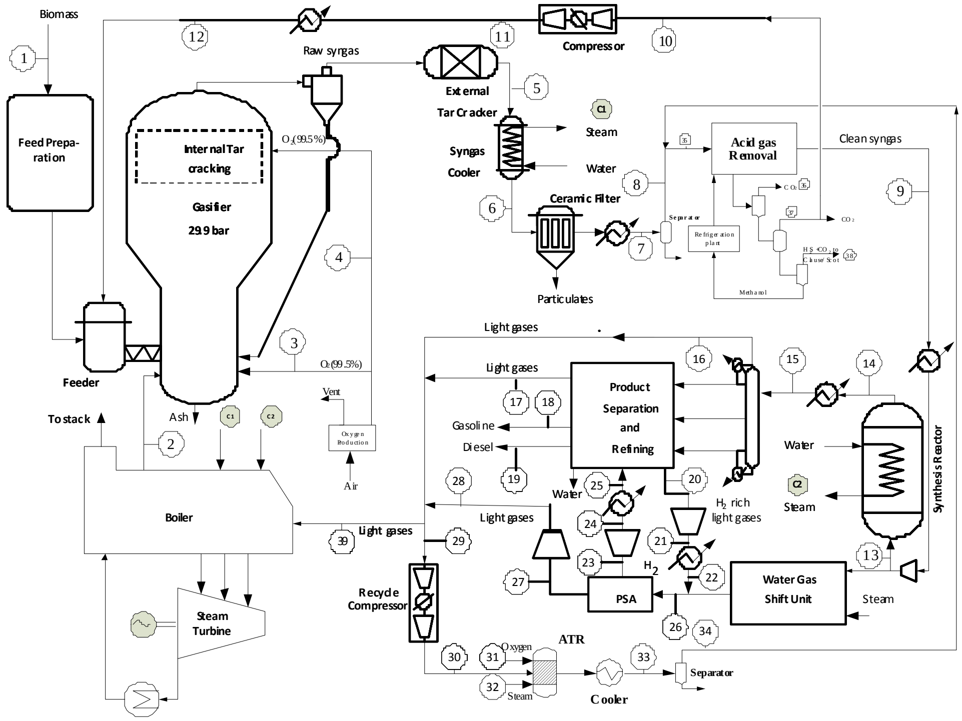


**Figure 18. Process flow simulation results for BTL-RC-V design.**

**Table 12. Stream data for BTL-RC-V simulation. Stream numbers refer to Figure 18.**

| | Raw Syngas | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Recycle gas feed into synthesis | Raw synthesis product | Final Liquid product | | Fuel gas to Gas turbine |
|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | Gasoline | Diesel | |
| Stream Number | **5** | **35** | **9** | **13** | **34** | **15** | **18** | **19** | **39** |
| Total Flow kg/sec | 55.29 | 68.41 | 33.27 | 32.74 | 20.38 | 32.74 | 2.45 | 3.84 | 1.99 |
| Total Flow kmol/sec | 2.58 | 3.36 | 2.55 | 2.51 | 1.18 | 1.38 | 0.02 | 0.02 | 0.08 |
| Mole Frac | | | | | | | | | |
| CO | 21.55% | 23.58% | 31.01% | 31.01% | 19.99% | 0.64% | 0.00% | 0.00% | 15.43% |
| H2 | 32.46% | 42.68% | 56.11% | 56.11% | 50.45% | 29.05% | 0.00% | 0.00% | 34.50% |
| CO2 | 24.96% | 24.37% | 0.96% | 0.96% | 14.80% | 15.12% | 0.00% | 0.00% | 12.94% |
| H2O | 15.84% | 0.28% | 0.00% | 0.00% | 0.28% | 28.99% | 0.00% | 0.00% | 0.00% |
| CH4 | 4.66% | 3.61% | 4.75% | 4.75% | 0.11% | 9.38% | 0.00% | 0.00% | 10.02% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.23% | 0.00% | 0.00% | 11.84% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.24% | 100.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.08% | 0.00% | 100.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.34% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.30% | 0.00% | 0.00% | 1.07% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.61% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.19% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.80% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.09% | 0.09% | 0.00% | 0.16% | 0.00% | 0.00% | 0.17% |
| O2 | 0.01% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| N2 | 0.43% | 5.33% | 7.00% | 7.00% | 14.21% | 12.74% | 0.00% | 0.00% | 13.87% |
| AR | 0.00% | 0.06% | 0.08% | 0.08% | 0.17% | 0.15% | 0.00% | 0.00% | 0.16% |
| H2S | 0.11% | 0.08% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 41.5 | 1.0 | 25.0 |
| 2 | 2.6 | 31.7 | 236.6 |
| 3 | 10.6 | 31.4 | 152.1 |
| 4 | 1.0 | 31.4 | 152.1 |
| 5 | 55.3 | 28.8 | 800.0 |
| 6 | 55.3 | 28.2 | 350.0 |
| 7 | 55.3 | 26.8 | 40.0 |
| 8 | 10.7 | 1.0 | 25.0 |
| 9 | 31.8 | 1.0 | 25.0 |
| 10 | 24.4 | 75.0 | 104.5 |
| 11 | 131.5 | 71.5 | 251.6 |
| 12 | 128.7 | 71.0 | 250.4 |
| 13 | 0.0 | 71.0 | 250.4 |
| 14 | 0.0 | 68.9 | 417.6 |
| 15 | 54.4 | 29.5 | 40.0 |
| 16 | 100.5 | 26.8 | 39.5 |
| 17 | 48.5 | 1.2 | 25.0 |
| 18 | 24.3 | 3.0 | 25.0 |
| 19 | 4.2 | 1.8 | 25.0 |
| 20 | 93.1 | 20.1 | 27.4 |
| 21 | 91.6 | 20.1 | 27.4 |
| 22 | 91.6 | 24.4 | 245.0 |
| 23 | 91.6 | 23.2 | 260.0 |
| 24 | 91.6 | 23.2 | 74.3 |
| 25 | 91.6 | 23.2 | 40.0 |
| 26 | 67.9 | 19.7 | 40.0 |
| 27 | 1.1 | 3.2 | 45.9 |
| 28 | 5.6 | 2.6 | 58.6 |
| 29 | 8.7 | 1.4 | 129.4 |
| 30 | 0.3 | 7.4 | 40.0 |
| 31 | 0.3 | 21.1 | 184.3 |
| 32 | 0.3 | 20.6 | 25.0 |
| 33 | 1.6 | 1.5 | 24.8 |
| 34 | 1.6 | 3.2 | 117.5 |
| 35 | 0.2 | 20.1 | 24.8 |
| 36 | 0.2 | 4.1 | -64.3 |
| 37 | 0.2 | 4.0 | 25.0 |
| 38 | 4.8 | 1.2 | 102.5 |
| 39 | 65.9 | 19.6 | 122.0 |
| 40 | 65.9 | 26.9 | 158.9 |
| 41 | 37.2 | 38.0 | 246.7 |
| 42 | 9.9 | 29.1 | 185.1 |
| 43 | 113.0 | 28.2 | 1000.0 |
| 44 | 69.9 | 28.0 | 40.4 |
| 45 | 3.5 | 3.0 | 25.0 |
| 46 | 3.5 | 31.4 | 142.5 |
| 47 | 3.5 | 30.5 | 40.0 |

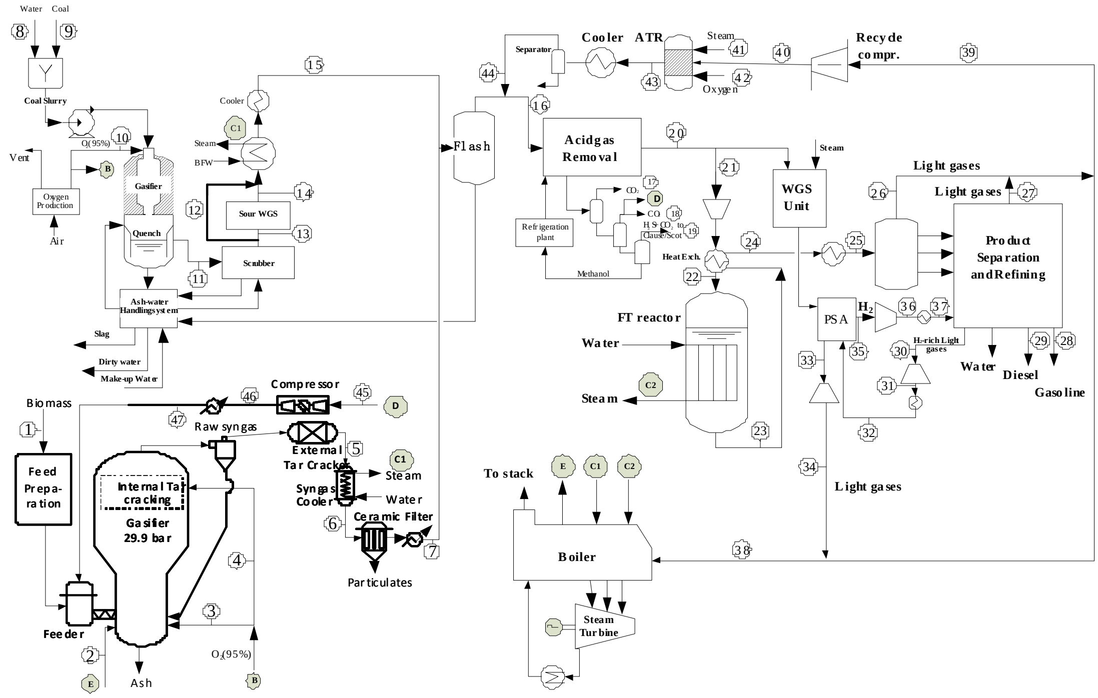


**Figure 19. Process flow simulation results for CBTL-RC-V design.**

**Table 13. Stream data for CBTL-RC-V simulation. Stream numbers refer to Figure 19.**

| | Scrubber Exit | Water Gas Shift reactor, inlet | Water Gas Shift reactor, Outlet | Biomass-derived raw syngas | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Recycle gas feed into synthesis | Raw synthesis product | Final Liquid product: Gasoline | Final Liquid product: Diesel | Fuel gas to Gas turbine |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Number | **11** | **12** | **13** | **5** | **16** | **20** | **22** | **44** | **25** | **27** | **28** | **38** |
| Total Flow kg/sec | 128.75 | 0.01 | 0.01 | 55.34 | 160.87 | 102.67 | 91.63 | 69.94 | 91.63 | 5.60 | 8.67 | 4.76 |
| Total Flow kmol/sec | 6.78 | 0.00 | 0.00 | 2.58 | 8.53 | 4.43 | 5.51 | 2.64 | 2.93 | 0.04 | 0.04 | 0.16 |
| Mole Frac | | | | | | | | | | | | |
| CO | 19.64% | 19.64% | 1.63% | 21.55% | 44.26% | 18.92% | 41.24% | 15.88% | 1.74% | 0.00% | 0.00% | 23.27% |
| H2 | 13.26% | 13.26% | 31.27% | 32.46% | 40.73% | 31.40% | 43.49% | 26.35% | 19.52% | 0.00% | 0.00% | 21.74% |
| CO2 | 3.95% | 3.95% | 21.95% | 24.96% | 11.00% | 20.25% | 0.96% | 33.00% | 32.04% | 0.00% | 0.00% | 17.85% |
| H2O | 62.47% | 62.47% | 44.46% | 15.84% | 0.15% | 0.17% | 0.00% | 0.29% | 15.15% | 0.00% | 0.00% | 0.00% |
| CH4 | 0.01% | 0.01% | 0.01% | 4.66% | 2.84% | 0.01% | 2.17% | 0.01% | 4.64% | 0.00% | 0.00% | 3.68% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.25% | 0.00% | 0.00% | 13.78% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 100.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.36% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.44% | 0.00% | 0.00% | 0.78% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.67% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.20% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.84% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.07% | 0.00% | 0.14% | 0.00% | 0.00% | 0.11% |
| O2 | 0.00% | 0.00% | 0.00% | 0.01% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| N2 | 0.26% | 0.26% | 0.26% | 0.43% | 0.67% | 28.42% | 11.76% | 23.85% | 22.07% | 0.00% | 0.00% | 18.31% |
| AR | 0.01% | 0.01% | 0.01% | 0.00% | 0.01% | 0.75% | 0.31% | 0.63% | 0.58% | 0.00% | 0.00% | 0.48% |
| H2S | 0.34% | 0.34% | 0.34% | 0.11% | 0.29% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.00% | 0.02% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.00% | 0.03% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

### 3.2.2 Once-Through FTL systems

For the OT systems, which produce large amounts of exported electricity, the electricity may be more appropriately called a co-product rather than a by-product. Here, mass and energy balances for the -V and -CCS variants of OT systems based on CTL, BTL, and CBTL configurations are discussed.

Figure 20 and Table 14 show detailed results for the CTL-OT-V case, wherein 31.7% of the input coal (LHV basis) is converted into FTL fuels and 17.6% into exportable electricity, giving a total 1$^{st}$ law efficiency of 49.3% (Table 8).

For the CTL-OT-CCS design (Figure 21 and Table 15), fuel outputs are unchanged from the CTL-OT-V case, but electricity output falls about 16%, resulting in a total 1$^{st}$ law efficiency of 46.5% (Table 8). The amount of carbon captured, 329.5 tC/hr, is 51.1% of the carbon in the input coal. The electricity output declines at a rate of 169.4 kWh per tonne of captured $CO_2$, accounted for by $CO_2$ compression (55.7%), electricity for the additional Rectisol equipment (7.5%), and $N_2$ compression (36.8%). The last item is to provide $N_2$ to the gas turbine combustor to dilute the fuel gas to the extent needed to limit $NO_x$ emissions from the turbine to an acceptable level.

Loss of net electricity output as a result of $N_2$ compression for $NO_x$ control is a key feature of all once-through –CCS systems; $N_2$ is not needed for $NO_x$ control with once-through –V systems because the amount of $CO_2$ in the syngas flowing to the gas turbine is adequate to keep the adiabatic flame temperature to levels such that $NO_x$ emission limits are not violated.

Additional $CO_2$ capture, beyond the CTL-OT-CCS design, is possible by reforming the hydrocarbons in the fuel gases that feed the gas turbine. For this process configuration, CTL-OTA-CCS, the $CO_2$ capture rate increases to 64.1% of input feedstock carbon, the fuel outputs are unchanged from the other CTL-OT cases, but electricity output falls about 36% compared to the CTL-OT-CCS case, resulting in a total 1$^{st}$ law efficiency of 43.0% (Table 8). In this case the electricity output declines at a rate of 304.3 kWh per tonne of captured $CO_2$, accounted by $CO_2$ compression (52.8%), electricity for the additional Rectisol equipment (7.9%), $N_2$ compression (19.7%), and electricity to provide $O_2$ for the autothermal reformer (19.6%).

Figure 22 and Table 16 show the detailed simulation results for the CBTL-OT-V design. In this case, 31.4% of the input feedstock energy (LHV basis) is converted into FTL fuels and 19.4% into exportable electricity, giving a total 1$^{st}$ law efficiency of 50.8% (Table 8).

Figure 23 and Table 17 show the simulations results for the CBTL-OT-CCS design, wherein the carbon capture rate is 76.6 tC/hr (50.9% of input feedstock carbon), and electricity output falls about 12%, resulting in a total 1$^{st}$ law efficiency of 48.4% (Table 8).

The CBTL-OTA-CCS system with an autothermal reforming step can be added ahead of the gas turbine enables a higher $CO_2$ capture rate (64.6% of input feedstock carbon). In this design, the total 1$^{st}$ law efficiency is reduced to 43.5%.

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 281.2 | 1.0 | 25.0 |
| 2 | 94.7 | 1.0 | 25.0 |
| 3 | 230.6 | 1.6 | 9.0 |
| 4 | 1185.0 | 71.5 | 251.4 |
| 5 | 205.7 | 71.0 | 250.3 |
| 6 | 205.7 | 68.9 | 415.8 |
| 7 | 955.2 | 71.0 | 250.3 |
| 8 | 531.6 | 29.5 | 40.0 |
| 9 | 512.6 | 28.0 | 39.7 |
| 10 | 104.8 | 1.2 | 25.0 |
| 11 | 52.4 | 3.0 | 25.0 |
| 12 | 33.4 | 1.8 | 25.0 |
| 13 | 5.1 | 20.1 | 27.8 |
| 14 | 320.1 | 20.1 | 27.8 |
| 15 | 315.1 | 20.1 | 27.8 |
| 16 | 315.1 | 24.4 | 245.0 |
| 17 | 315.1 | 23.2 | 260.0 |
| 18 | 315.1 | 23.2 | 85.2 |
| 19 | 315.1 | 23.2 | 40.0 |
| 20 | 241.5 | 19.7 | 40.0 |
| 21 | 4.1 | 3.2 | 45.9 |
| 22 | 20.5 | 2.6 | 58.6 |
| 23 | 31.8 | 1.4 | 129.4 |
| 24 | 1.2 | 7.4 | 40.0 |
| 25 | 1.2 | 21.1 | 184.3 |
| 26 | 1.2 | 20.6 | 40.0 |
| 27 | 0.8 | 20.1 | 24.9 |
| 28 | 0.8 | 4.1 | -64.2 |
| 29 | 0.8 | 4.0 | 25.0 |
| 30 | 5.5 | 1.5 | 24.9 |
| 31 | 5.5 | 3.2 | 118.2 |
| 32 | 251.1 | 19.6 | 121.4 |
| 33 | 251.1 | 28.6 | 182.0 |

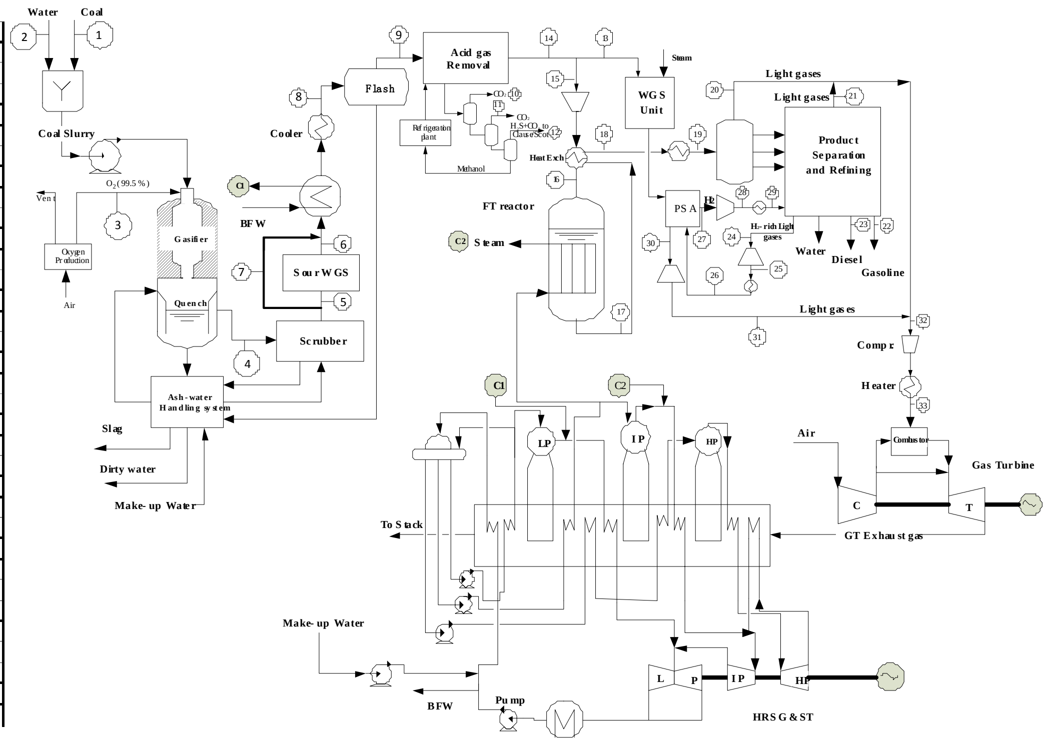


**Figure 20. Process flow simulation results for CTL-OT-V design.**

**Table 14. Stream data for CTL-OT-V simulation. Stream numbers refer to Figure 20.**

| | Scrubber Exit | Water Gas Shift reactor, inlet | Water Gas Shift reactor, Outlet | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Raw synthesis product | Final Liquid product | | Fuel gas to Gas turbine |
|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | Gasoline | Diesel | |
| Stream Number | **5+7** | **5** | **6** | **9** | **14** | **16** | **19** | **22** | **23** | **32** |
| Total Flow kg/sec | 1160.8 | 205.7 | 205.7 | 512.6 | 320.1 | 315.1 | 315.1 | 20.5 | 31.8 | 251.1 |
| Total Flow kmol/sec | 60.8 | 10.8 | 10.8 | 24.8 | 20.3 | 20.0 | 10.5 | 0.2 | 0.1 | 9.7 |
| Mole Frac | | | | | | | | | | |
| CO | 19.36% | 19.36% | 1.56% | 39.69% | 48.38% | 48.38% | 6.80% | 0.00% | 0.00% | 9.00% |
| H2 | 13.04% | 13.04% | 30.84% | 39.66% | 48.34% | 48.34% | 34.68% | 0.00% | 0.00% | 38.39% |
| CO2 | 3.93% | 3.93% | 21.73% | 17.29% | 0.63% | 0.63% | 39.70% | 0.00% | 0.00% | 43.15% |
| H2O | 62.38% | 62.38% | 44.58% | 0.26% | 0.00% | 0.00% | 8.49% | 0.00% | 0.00% | 0.00% |
| CH4 | 0.01% | 0.01% | 0.01% | 0.03% | 0.03% | 0.03% | 1.13% | 0.00% | 0.00% | 1.23% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 0.00% | 0.00% | 1.13% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.27% | 100.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.37% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.49% | 0.00% | 0.00% | 1.62% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.69% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.21% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.86% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.01% | 0.01% | 0.02% | 0.00% | 0.00% | 0.02% |
| N2 | 0.34% | 0.34% | 0.34% | 0.84% | 1.02% | 1.02% | 1.95% | 0.00% | 0.00% | 2.15% |
| AR | 0.53% | 0.53% | 0.53% | 1.29% | 1.58% | 1.58% | 3.01% | 0.00% | 0.00% | 3.32% |
| H2S | 0.34% | 0.34% | 0.34% | 0.79% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.05% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.09% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 281.2 | 1.0 | 25.0 |
| 2 | 94.7 | 1.0 | 25.0 |
| 3 | 230.6 | 1.6 | 9.0 |
| 4 | 1185.0 | 71.5 | 251.4 |
| 5 | 206.3 | 71.0 | 250.3 |
| 6 | 206.3 | 68.9 | 415.8 |
| 7 | 954.5 | 71.0 | 250.3 |
| 8 | 532.1 | 29.5 | 40.0 |
| 9 | 512.7 | 28.0 | 39.7 |
| 10 | 111.8 | 3.0 | 25.0 |
| 11 | 223.6 | 1.2 | 25.0 |
| 12 | 335.3 | 3.0 | 30.0 |
| 13 | 33.4 | 1.8 | 25.0 |
| 14 | 319.9 | 20.1 | 27.8 |
| 15 | 314.9 | 20.1 | 27.8 |
| 16 | 314.9 | 24.4 | 245.0 |
| 17 | 314.9 | 23.2 | 260.0 |
| 18 | 314.9 | 23.2 | 85.4 |
| 19 | 314.9 | 23.2 | 40.0 |
| 20 | 241.3 | 19.7 | 40.0 |
| 21 | 4.1 | 3.2 | 45.9 |
| 22 | 20.5 | 2.6 | 58.6 |
| 23 | 31.8 | 1.4 | 129.4 |
| 24 | 1.2 | 7.4 | 40.0 |
| 25 | 1.2 | 21.1 | 184.3 |
| 26 | 1.2 | 20.6 | 40.0 |
| 27 | 0.8 | 20.1 | 24.9 |
| 28 | 0.8 | 4.1 | -64.2 |
| 29 | 0.8 | 4.0 | 25.0 |
| 30 | 5.5 | 1.5 | 24.9 |
| 31 | 5.5 | 3.2 | 118.2 |
| 32 | 250.9 | 19.6 | 121.4 |
| 33 | 250.9 | 28.2 | 40.0 |
| 34 | 82.7 | 26.8 | 124.4 |
| 35 | 166.7 | 28.0 | 146.3 |
| 36 | 249.4 | 26.8 | 182.0 |

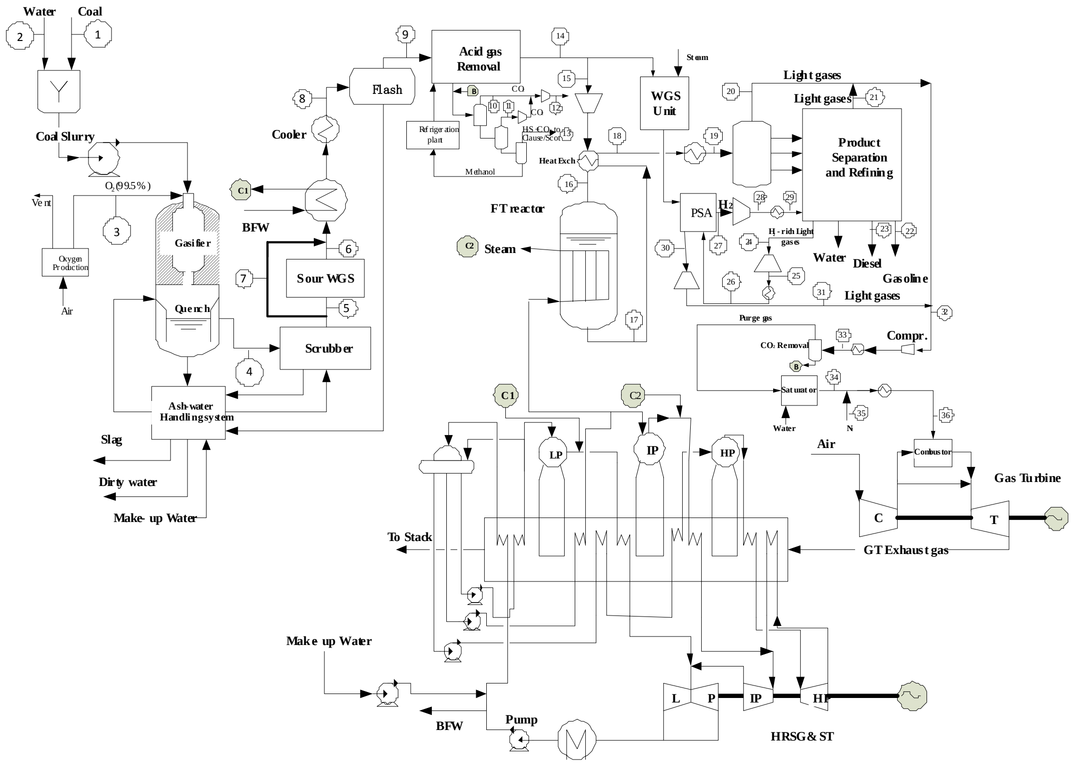


**Figure 21. Process flow simulation results for CTL-OT-CCS design.**

**Table 15. Stream data for CTL-OT-CCS simulations. Stream numbers refer to Figure 21.**

| | Scrubber Exit | Water Gas Shift reactor, inlet | Water Gas Shift reactor, Outlet | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Raw synthesis product | Final Liquid product | | Fuel gas to Gas turbine | Fuel gas to GT combustor |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | Gasoline | Diesel | | |
| Stream Number | 5+7 | 5 | 6 | 9 | 14 | 16 | 19 | 22 | 23 | 32 | 36 |
| Total Flow kg/sec | 1160.8 | 206.3 | 206.3 | 512.7 | 320.0 | 314.9 | 314.9 | 20.5 | 31.8 | 73.3 | 249.4 |
| Total Flow kmol/sec | 60.8 | 10.8 | 10.8 | 24.8 | 20.3 | 20.0 | 10.5 | 0.2 | 0.1 | 5.6 | 12.1 |
| Mole Frac | | | | | | | | | | | |
| CO | 19.36% | 19.36% | 1.56% | 39.66% | 48.35% | 48.35% | 6.78% | 0.00% | 0.00% | 15.40% | 7.17% |
| H2 | 13.04% | 13.04% | 30.84% | 39.68% | 48.37% | 48.37% | 34.70% | 0.00% | 0.00% | 65.96% | 30.71% |
| CO2 | 3.93% | 3.93% | 21.73% | 17.31% | 0.63% | 0.63% | 39.67% | 0.00% | 0.00% | 2.22% | 1.03% |
| H2O | 62.38% | 62.38% | 44.58% | 0.26% | 0.00% | 0.00% | 8.52% | 0.00% | 0.00% | 0.00% | 4.46% |
| CH4 | 0.01% | 0.01% | 0.01% | 0.03% | 0.03% | 0.03% | 1.13% | 0.00% | 0.00% | 2.11% | 0.98% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 0.00% | 0.00% | 1.93% | 0.90% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.27% | 100.00% | 0.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.37% | 0.00% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.49% | 0.00% | 0.00% | 2.77% | 1.29% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.69% | 0.00% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.21% | 0.00% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.86% | 0.00% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.01% | 0.01% | 0.02% | 0.00% | 0.00% | 0.20% | 0.02% |
| DME | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| O2 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.52% |
| N2 | 0.34% | 0.34% | 0.34% | 0.84% | 1.02% | 1.02% | 1.95% | 0.00% | 0.00% | 3.69% | 50.03% |
| AR | 0.53% | 0.53% | 0.53% | 1.29% | 1.58% | 1.58% | 3.01% | 0.00% | 0.00% | 5.71% | 2.87% |
| H2S | 0.34% | 0.34% | 0.34% | 0.79% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.05% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.09% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 41.4 | 1.0 | 25 |
| 2 | 2.6 | 31.7 | 236.6 |
| 3 | 10.6 | 31.4 | 152.1 |
| 4 | 1.0 | 31.4 | 152.1 |
| 5 | 55.3 | 28.8 | 800 |
| 6 | 55.3 | 28.2 | 350 |
| 7 | 55.3 | 26.8 | 40 |
| 8 | 10.7 | 1.0 | 25 |
| 9 | 31.8 | 1.0 | 25 |
| 10 | 24.4 | 75.0 | 104.5 |
| 11 | 131.4 | 71.5 | 251.55 |
| 12 | 128.7 | 71.0 | 250.36 |
| 13 | 0.0 | 71.0 | 250.36 |
| 14 | 0.0 | 68.9 | 417.55 |
| 15 | 54.3 | 29.5 | 40 |
| 16 | 118.1 | 26.8 | 39.49 |
| 17 | 28.2 | 1.2 | 25 |
| 18 | 14.1 | 3.0 | 25 |
| 19 | 5.2 | 1.8 | 25 |
| 20 | 1.1 | 20.1 | 27.41 |
| 21 | 69.2 | 20.1 | 27.41 |
| 22 | 69.2 | 24.4 | 245 |
| 23 | 69.2 | 23.2 | 260 |
| 24 | 69.2 | 23.2 | 74.26 |
| 25 | 69.2 | 23.2 | 40 |
| 26 | 52.7 | 19.7 | 40 |
| 27 | 0.9 | 3.2 | 45.92 |
| 28 | 4.5 | 2.6 | 58.58 |
| 29 | 7.0 | 1.4 | 129.39 |
| 30 | 0.3 | 7.4 | 40 |
| 31 | 0.3 | 21.1 | 184.27 |
| 32 | 0.3 | 20.6 | 25 |
| 33 | 1.2 | 1.5 | 24.83 |
| 34 | 1.2 | 3.2 | 117.48 |
| 35 | 0.2 | 20.1 | 24.83 |
| 36 | 0.2 | 4.1 | -64.25 |
| 37 | 0.2 | 4.0 | 25 |
| 38 | 54.8 | 19.6 | 121.44 |
| 39 | 54.8 | 28.6 | 182 |
| 40 | 3.5 | 3.0 | 25 |
| 41 | 3.5 | 31.4 | 142.53 |
| 42 | 3.5 | 30.5 | 40 |

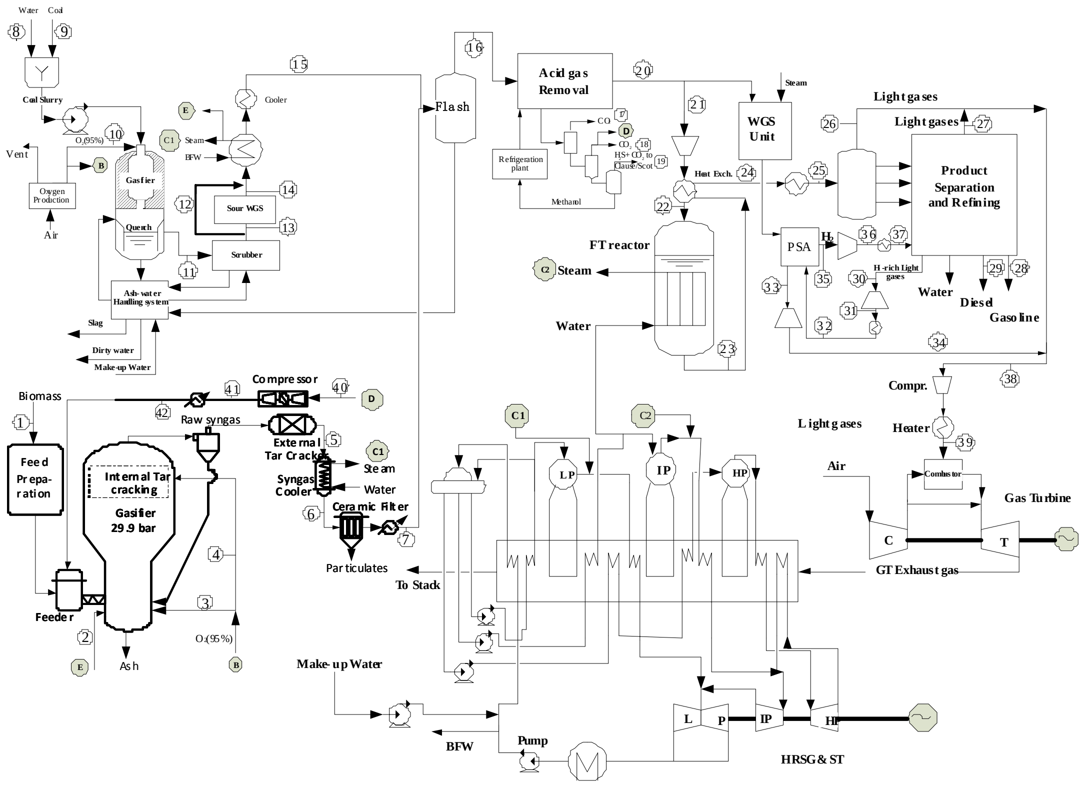


**Figure 22. Process flow simulation results for CBTL-OT-V design.**

**Table 16. Stream data for CBTL-OT-V simulations. Stream numbers refer to Figure 22.**

| | Scrubber Exit | Water Gas Shift reactor, inlet | Water Gas Shift reactor, | Biomass-derived raw syngas | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Raw synthesis product | Final Liquid product | | Fuel gas to Gas turbine |
|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | Gasoline | Diesel | |
| Stream Number | **11** | **12** | **13** | **5** | **16** | **20** | **22** | **25** | **27** | **28** | **38** |
| Total Flow kg/sec | 163.58 | 9.17 | 9.17 | 56.14 | 70.26 | 70.26 | 69.15 | 69.15 | 4.54 | 7.02 | 54.84 |
| Total Flow kmol/sec | 8.56 | 0.48 | 0.48 | 2.68 | 4.50 | 4.50 | 4.43 | 2.33 | 0.04 | 0.03 | 2.13 |
| Mole Frac | | | | | | | | | | | |
| CO | 19.34% | 19.34% | 1.55% | 21.71% | 47.80% | 47.80% | 47.80% | 6.21% | 0.00% | 0.00% | 8.37% |
| H2 | 13.02% | 13.02% | 30.81% | 35.61% | 47.85% | 47.85% | 47.85% | 33.39% | 0.00% | 0.00% | 37.15% |
| CO2 | 3.93% | 3.93% | 21.71% | 24.84% | 0.72% | 0.72% | 0.72% | 39.36% | 0.00% | 0.00% | 42.96% |
| H2O | 62.43% | 62.43% | 44.65% | 14.40% | 0.00% | 0.00% | 0.00% | 8.88% | 0.00% | 0.00% | 0.00% |
| CH4 | 0.01% | 0.01% | 0.01% | 2.77% | 1.67% | 1.67% | 1.67% | 4.20% | 0.00% | 0.00% | 4.64% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 0.00% | 0.00% | 1.13% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.27% | 100.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.37% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.48% | 0.00% | 0.00% | 1.62% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.69% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.21% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.86% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.01% | 0.01% | 0.01% | 0.02% | 0.00% | 0.00% | 0.02% |
| O2 | 0.00% | 0.00% | 0.00% | 0.07% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| N2 | 0.34% | 0.34% | 0.34% | 0.42% | 0.90% | 0.90% | 0.90% | 1.72% | 0.00% | 0.00% | 1.90% |
| AR | 0.53% | 0.53% | 0.53% | 0.07% | 1.05% | 1.05% | 1.05% | 1.99% | 0.00% | 0.00% | 2.21% |
| H2S | 0.34% | 0.34% | 0.34% | 0.10% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

| | m (kg/s) | P (bar) | T (°C) |
|---|---|---|---|
| 1 | 41.4 | 1.0 | 25.0 |
| 2 | 2.6 | 31.7 | 236.6 |
| 3 | 11.3 | 31.4 | 152.1 |
| 4 | 1.1 | 31.4 | 152.1 |
| 5 | 56.1 | 28.8 | 800.0 |
| 6 | 56.1 | 28.2 | 350.0 |
| 7 | 56.1 | 26.8 | 40.0 |
| 8 | 13.3 | 1.0 | 25.0 |
| 9 | 39.6 | 1.0 | 25.0 |
| 10 | 32.4 | 75.0 | 104.5 |
| 11 | 166.3 | 71.5 | 251.3 |
| 12 | 154.3 | 71.0 | 250.3 |
| 13 | 9.3 | 71.0 | 250.3 |
| 14 | 9.3 | 68.9 | 415.6 |
| 15 | 71.2 | 29.5 | 40.0 |
| 16 | 118.1 | 26.8 | 39.4 |
| 17 | 27.1 | 3.0 | 25.0 |
| 18 | 81.4 | 3.0 | 30.0 |
| 19 | 5.2 | 1.8 | 25.0 |
| 20 | 70.2 | 20.1 | 27.4 |
| 21 | 69.1 | 20.1 | 27.4 |
| 22 | 69.1 | 24.4 | 245.0 |
| 23 | 69.1 | 23.2 | 260.0 |
| 24 | 69.1 | 23.2 | 74.3 |
| 25 | 69.1 | 23.2 | 40.0 |
| 26 | 52.7 | 19.7 | 40.0 |
| 27 | 0.9 | 3.2 | 45.9 |
| 28 | 4.5 | 2.6 | 58.6 |
| 29 | 7.0 | 1.4 | 129.4 |
| 30 | 0.3 | 7.4 | 40.0 |
| 31 | 0.3 | 21.1 | 184.3 |
| 32 | 0.3 | 20.6 | 25.0 |
| 33 | 1.2 | 1.5 | 24.8 |
| 34 | 1.2 | 3.2 | 117.5 |
| 35 | 0.2 | 20.1 | 24.8 |
| 36 | 0.2 | 4.1 | -64.3 |
| 37 | 0.2 | 4.0 | 25.0 |
| 38 | 54.8 | 19.6 | 121.4 |
| 39 | 54.8 | 28.2 | 40.0 |
| 40 | 17.5 | 26.8 | 118.5 |
| 41 | 41.4 | 28.0 | 146.3 |
| 42 | 58.9 | 26.8 | 182.0 |
| 43 | 3.5 | 3.0 | 25.0 |
| 44 | 3.5 | 31.4 | 142.5 |
| 45 | 3.5 | 30.5 | 40.0 |

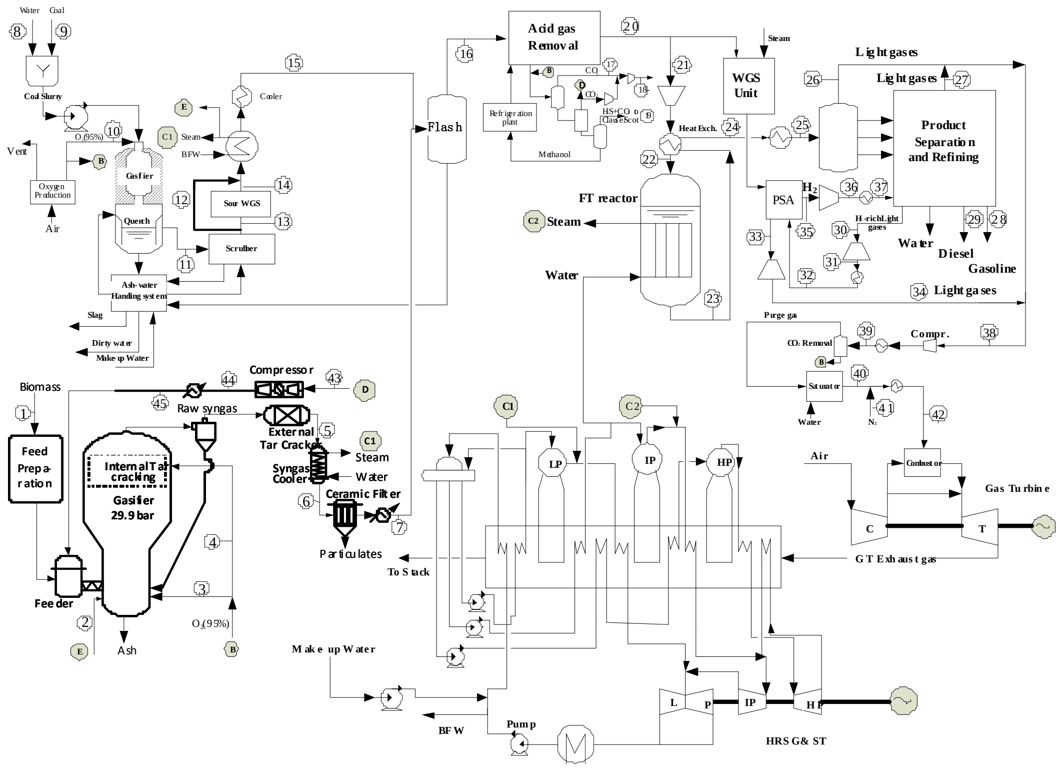


Figure 23. Process flow simulation results for CBTL-OT-CCS design.

## Table 17. Stream data for CBTL-OT-CCS simulations. Stream numbers refer to Figure 23.

| | Scrubber Exit | Water Gas Shift reactor, inlet | Water Gas Shift reactor, Outlet | Biomass-derived raw syngas | H2S removal unit, inlet | H2S removal unit, Outlet | Total feed into synthesis (inc. recy.) | Raw synthesis product | Final Liquid product | | Fuel gas to Gas turbine | Fuel gas to GT combustor |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| | | | | | | | | | Gasoline | Diesel | | |
| Stream Number | 12+13 | 13 | 14 | 5 | 16 | 20 | 23 | 25 | 28 | 29 | 38 | 42 |
| Total Flow kg/sec | 163.51 | 9.26 | 9.26 | 56.12 | 118.12 | 70.21 | 69.11 | 69.11 | 4.54 | 7.02 | 15.80 | 58.89 |
| Total Flow kmol/sec | 8.56 | 0.48 | 0.48 | 2.68 | 5.60 | 4.50 | 4.43 | 2.32 | 0.04 | 0.03 | 1.24 | 2.81 |
| Mole Frac | | | | | | | | | | | | |
| CO | 19.34% | 19.34% | 1.55% | 21.72% | 38.36% | 47.78% | 47.78% | 6.19% | 0.00% | 0.00% | 14.29% | 6.32% |
| H2 | 13.02% | 13.02% | 30.81% | 35.64% | 38.44% | 47.88% | 47.88% | 33.42% | 0.00% | 0.00% | 63.64% | 28.16% |
| CO2 | 3.93% | 3.93% | 21.71% | 24.84% | 19.37% | 0.72% | 0.72% | 39.33% | 0.00% | 0.00% | 2.20% | 0.98% |
| H2O | 62.43% | 62.43% | 44.65% | 14.38% | 0.27% | 0.00% | 0.00% | 8.91% | 0.00% | 0.00% | 0.00% | 3.47% |
| CH4 | 0.01% | 0.01% | 0.01% | 2.76% | 1.33% | 1.66% | 1.66% | 4.19% | 0.00% | 0.00% | 7.92% | 3.51% |
| C4H10-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.26% | 0.00% | 0.00% | 1.93% | 0.86% |
| C9H20-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.27% | 100.00% | 0.00% | 0.00% | 0.00% |
| C15H32 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.09% | 0.00% | 100.00% | 0.00% | 0.00% |
| C21H44 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.37% | 0.00% | 0.00% | 0.00% | 0.00% |
| C4H8-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 1.48% | 0.00% | 0.00% | 2.77% | 1.23% |
| C9H18-1 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.68% | 0.00% | 0.00% | 0.00% | 0.00% |
| C15H30 | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.21% | 0.00% | 0.00% | 0.00% | 0.00% |
| C21OL | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.86% | 0.00% | 0.00% | 0.00% | 0.00% |
| MEOH | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.01% | 0.01% | 0.02% | 0.00% | 0.00% | 0.20% | 0.05% |
| O2 | 0.00% | 0.00% | 0.00% | 0.07% | 0.03% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.56% |
| N2 | 0.34% | 0.34% | 0.34% | 0.42% | 0.72% | 0.90% | 0.90% | 1.72% | 0.00% | 0.00% | 3.25% | 52.96% |
| AR | 0.53% | 0.53% | 0.53% | 0.07% | 0.84% | 1.05% | 1.05% | 1.99% | 0.00% | 0.00% | 3.78% | 1.90% |
| H2S | 0.34% | 0.34% | 0.34% | 0.10% | 0.54% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| COS | 0.02% | 0.02% | 0.02% | 0.00% | 0.03% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |
| HCL | 0.04% | 0.04% | 0.04% | 0.00% | 0.06% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% | 0.00% |

## 3.3 Net fuel cycle GHG emissions

### 3.3.1 Assumptions

Our plant simulations provide the basis for estimating $CO_2$ emissions associated with feedstock conversion and FTL fuel combustion. There are also $CO_2$ and other GHG emissions that occur outside the plant boundaries, and we consider these in estimating total net fuel cycle GHG emissions. Figure 24 summarizes all of the carbon flows considered in the net fuel cycle. The flows include emissions associated with production and delivery of the coal and/or biomass feedstocks to the FTL facility, emissions as flue gas at the facility, emissions associated with delivery of the liquid fuel to vehicle fueling stations, and emissions from combustion of the FTL fuels. The $CO_2$ absorbed during photosynthesis is included as a negative emission. This absorbed carbon may travel with the harvested biomass to the FTL facility or in some cases some of it is assumed to be stored permanently in the soil and roots in the fields where the biomass is harvested.

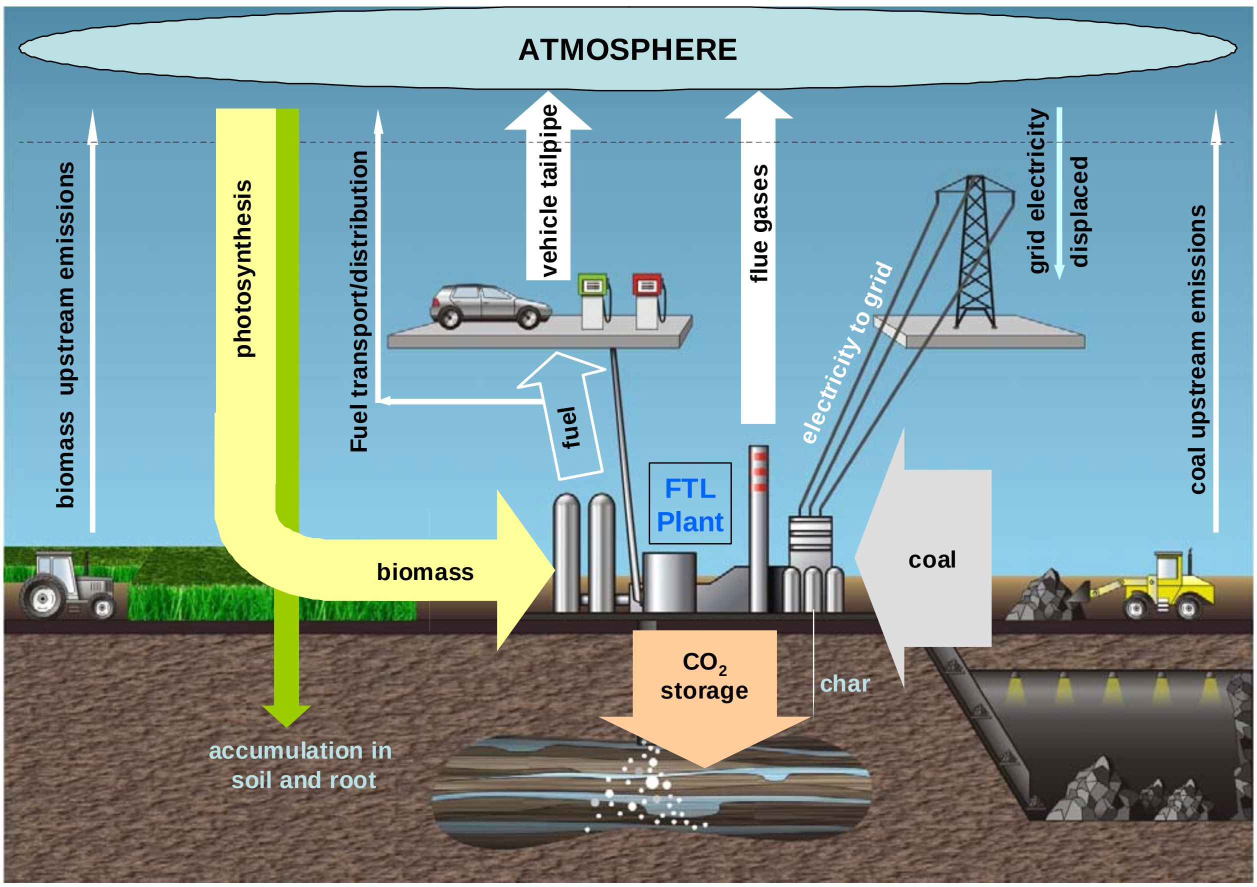


**Figure 24. Schematic of all flows considered in estimating net fuel cycle greenhouse gas emissions with coal, biomass, or coal+biomass conversion to FTL.**

Electricity exported from an FTL facility may generate a GHG emissions credit, since it may displace electricity that would otherwise have been generated elsewhere on the grid. The appropriate level of electricity emissions credit is the amount of GHG emissions actually avoided elsewhere on the grid. However, because in practice it is impossible to know precisely what grid electricity is displaced, one must assign a value arbitrarily. Some alternative values that could be used are given in Table 18.

The emissions upstream and downstream of the FTL plant are based on the GREET model [43], as are fuel cycle emissions for petroleum-derived reformulated gasoline and low-sulfur diesel, against which we compare fuel cycle emissions for FTL fuels (Table 19). For upstream emissions associated with biomass production and delivery, separate values are given for switchgrass biomass and for mixed prairie grass biomass. The latter is assumed to store greater amounts of carbon in the soil and roots than the former (see Table 19 notes).

**Table 18. Full lifecycle carbon-equivalent emissions for power generation.**

| | $kgC_{eq}/MWh$ | $kgCO_{2eq}/MWh$ |
|---|---|---|
| Average U.S. electricity grid generation in 2007 (a) | 174 | 636 |
| Average existing coal power plant in the U.S. (b, c) | 271 | 992 |
| 550 $MW_e$ pulverized coal supercritical steam plant with $CO_2$ vented (b,d) | 227 | 831 |
| 546 $MW_e$ pulverized coal supercrit. steam, 90% flue gas $CO_2$ capture/storage (b,d) | 47 | 171 |
| 640 $MW_e$ coal IGCC with $CO_2$ vented (b, d) | 228 | 833 |
| 556 $MW_e$ coal IGCC with 90% pre-combustion $CO_2$ capture/storage (b, d) | 38 | 138 |
| 560 $MW_e$ natural gas combined cycle with $CO_2$ vented (b, e) | 115 | 421 |
| 482 $MW_e$ natural gas combined cycle with post-combustion 90% $CO_2$ cap/storage (b, e) | 30 | 110 |

(a) Based on GREET [43].
(b) Upstream emissions (e.g., coal mining, natural gas extraction, etc.) are based on GREET (Table 19).
(c) Assuming an average higher heating value efficiency of 32.786%, which is based on 2005 U.S. power sector generation of 1992 TWh from coal and corresponding consumption of 21.878 EJ of coal [44]. We have assumed Illinois #6 coal (Table 3) would be converted at this efficiency to estimate the GHG emission rate for the average existing coal power plant in the U.S.
(d) Emissions during coal conversion are based on NETL [55]; for coal cases, the coal is Illinois #6 (Table 3); for IGCC, the gasifier is the GE radiant-quench design.
(e) Emissions during natural gas conversion are based on NETL [55]; gas turbine is F-class.

**Table 19. $CO_2$-equivalent greenhouse gas emissions from GREET 1.8b expressed in equivalent carbon or $CO_2$ per GJ of delivered energy.**

| | kgCeq/GJ | | kgCO2eq/GJ | |
|---|---|---|---|---|
| | LHV basis | HHV basis | HHV basis | LHV basis |
| **Coal upstream (a)** | | | | |
| Mine to synfuel plant gate | 1.0241 | 0.98 | 3.58 | 3.76 |
| Mine to power plant gate | 1.1986 | 1.14 | 4.19 | 4.40 |
| **Biomass upstream** | | | | |
| Herbaceous biomass (switchgrass) (b) | 1.79 | 1.63 | 5.96 | 6.55 |
| Mixed Prairie grasses, with soil/root carbon storage (c) | -16.2 | -14.8 | -54.18 | -59.51 |
| **Natural gas** | | | | |
| Extraction to power plant gate | 2.48 | 2.24 | 8.22 | 9.10 |
| **Petroleum fuels distribution (refinery gate to vehicle fuel tank)** | | | | |
| Reformulated gasoline | 0.155 | 0.145 | 0.53 | 0.57 |
| Low sulfur diesel | 0.134 | 0.125 | 0.46 | 0.49 |
| **Petroleum fuels full fuel cycle (well to tailpipe)** | | | | |
| Reformulated gasoline | 24.92 | 23.27 | 85.31 | 91.37 |
| Low sulfur diesel | 25.06 | 23.43 | 85.91 | 91.89 |

(a) GREET assumes that coal synfuel plants would be located at the mine mouth, while coal power plants would be remote from the mine and thus require additional coal transportation.
(b) This figure is from the GREET model and includes a carbon credit (0.805 $kgC_{eq}/GJ_{HHV}$ or 0.884 $kgC_{eq}/GJ_{LHV}$) for buildup of carbon in the soil when switchgrass replaces annual row cropping.
(c) This figure is for mixed prairie grasses (MPGs) established on carbon-depleted soils. It includes GHG emissions as in GREET (see note b), but substitutes an independent value for the soil carbon credit included in GREET. The soil and root carbon build up rate is assumed to be 0.3 t carbon per dry t harvested biomass as an average over a 30-year period [45]. We assume 7% of the harvested biomass is lost during delivery to the conversion facility, so that the carbon storage rate [0.3/(1 - 0.07)] = 0.323 t carbon per t dry biomass delivered, or 18.9 $kgC_{eq}/GJ_{LHV}$, 17.2 $kgC_{eq}/GJ_{HHV}$.

### 3.3.2 Results

Figure 25 shows calculated net lifecycle GHG emissions associated with FTL fuels for CTL-RC, BTL-RC, and CBTL-RC plant configurations, with net lifecycle emissions for comparable petroleum-derived fuels shown for comparison. These GHG emissions for FTL fuels in Figure 25 include no credit for exported electricity. FTL made from coal without CCS (CTL-RC-V) results in net GHG emissions more than double those for petroleum-derived fuels. With CCS, the CTL-RC system emits slightly more net GHG emissions than the equivalent petroleum-derived fuels. For the BTL-RC systems, net GHG emissions are slightly negative despite positive emissions associated with production and transportation of the biomass to the plant and with distribution of the FTL fuels to refueling stations. The negative emissions arise because about 10% of the carbon in the input biomass leaves the FTL plant as char in the gasifier ash, and we assume this carbon is sequestered from the atmosphere, e.g., via landfilling. With CCS, net emissions from the BTL-RC system are strongly negative. For the CBTL-RC system, for which 42% of the lower heating value of the feedstocks input is biomass, when no CCS is included, net emissions are about 30% higher than for the equivalent petroleum-derived fuels. With CCS, the net lifecycle emissions are essentially zero.

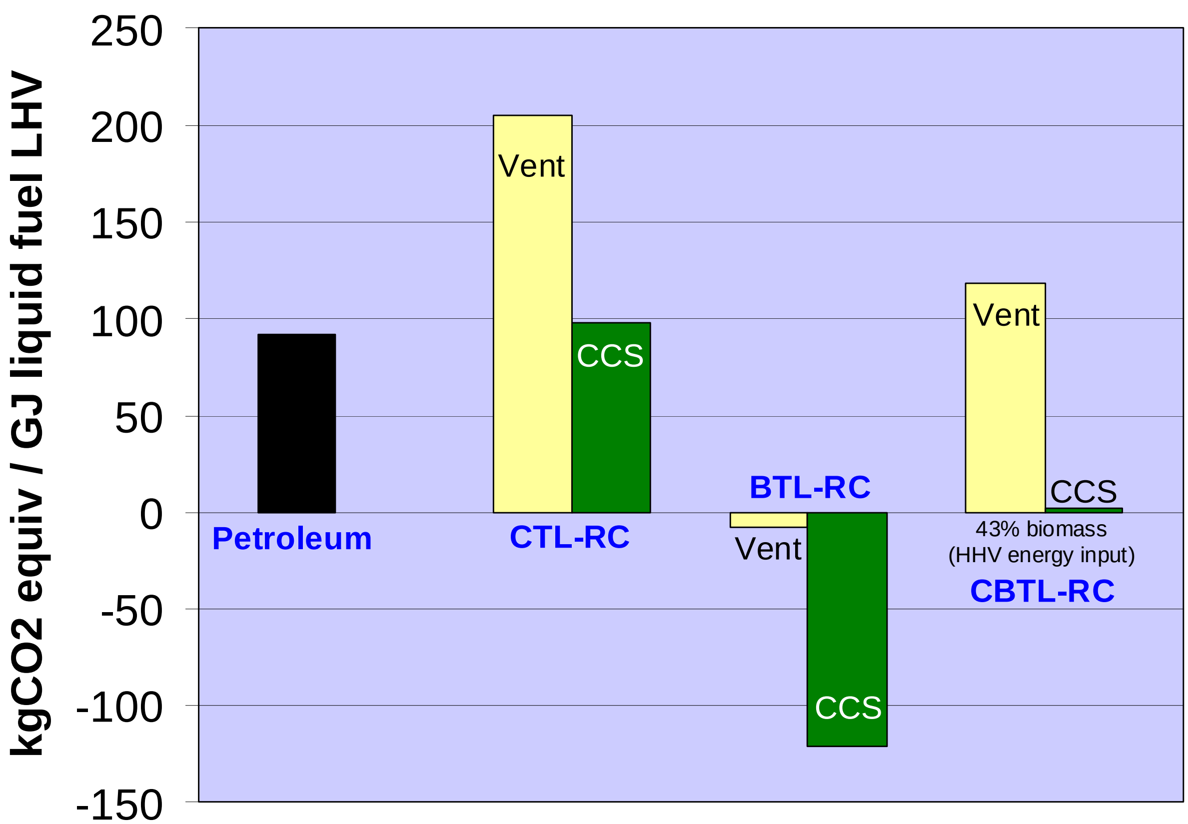


**Figure 25. Net fuel cycle GHG emissions for FTL fuels made via recycle process configurations. The figures here assume no GHG emissions credit for electricity exported from the FTL plants.**

Assigning a GHG emissions credit to the electricity byproduct or co-product has no impact on the overall system economics described in later sections, but various stakeholders are interested in knowing how emissions are allocated between the liquid fuel and electricity products so as to be able to compare emission rates of FTL systems with emission rates for other systems that provide liquid fuels and electricity.

Because our RC designs export relatively little electricity, assigning an emissions credit for exported electricity only modestly changes the numbers shown in Figure 25. This is in contrast to the OT designs, for which including an export electricity credit can significantly change the calculated net lifecycle emissions. Figure 26 shows GHG emissions as a function of the assumed electricity credit. Shown for reference are lifecycle GHG emission rates for integrated coal-gasifier combined cycle power plants either venting $CO_2$ (IGCC-Vent) or capturing and storing $CO_2$ (IGCC-CCS), as well as for the average mix of power generated in the U.S. in 2007. Given the likelihood that few new coal-fired power plants are likely to be built in the future in the United States without CCS, an appropriate emissions credit to consider might be the emission rate for a IGCC-CCS plant (138 $kgCO_{2eq}$/MWh), in which case the net lifecycle FTL emissions for our 16 process designs would be as indicated in Table 20.

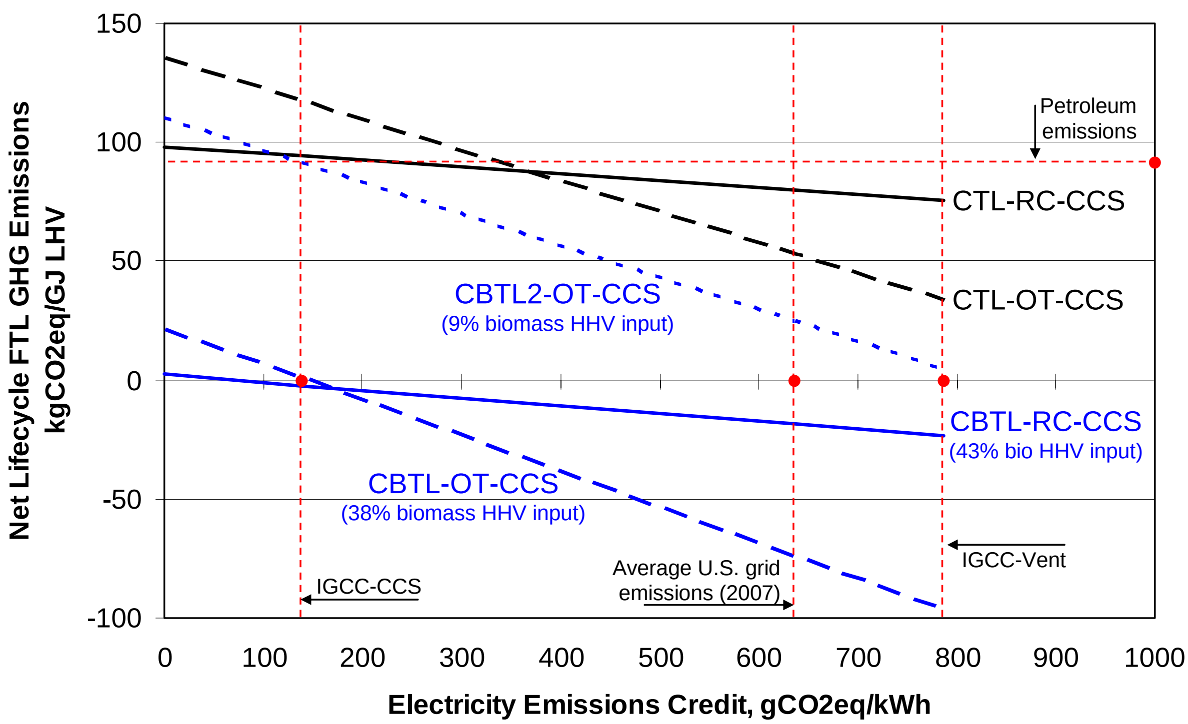


**Figure 26. Net fuel cycle GHG emissions for FTL fuels from coal or coal+biomass with varying level of electricity emissions credit.**

## 3.4 Yields of net zero-GHG liquid fuels from biomass

Biomass is the only carbon-bearing renewable energy resource, which makes it especially valuable for making carbon-bearing liquid transportation fuels. Significant amounts of biomass are potentially available in the U.S. for energy, but the magnitude of available biomass resources is nevertheless much less than available coal and other fossil fuel supplies. The US Departments of Energy and of Agriculture, in a joint study [46], estimate there are some 386 tonnes/yr (about 7.2 EJ/year) of crop residues that could be available for energy uses (without compromising soil quality and other services provided by residues) and 335 tonnes/yr (about 6.9 EJ/yr) of forestry residues and forest-fuel treatment thinnings (that should be removed to reduce fire risks). The sum of these biomass resources (14 EJ/yr) can be compared with U.S. coal use of 24 EJ in 2007 and total primary energy use of 107 EJ. Given the constraints on biomass supplies, its efficient use is important.

**Table 20. Net lifecycle GHG emissions associated with FTL fuels when electricity coproduct is assigned GHG emissions of IGCC-CCS (138 $kgCO_{2eq}$/MWh).**

| | Net Lifecycle GHG Emissions | |
|---|---|---|
| | kgCO2eq/GJ FTL LHV | FTL relative to COPD |
| CTL-RC-V | 200 | 2.2 |
| CTL-RC-CCS | 94 | 1.0 |
| CBTL-RC-V | 112 | 1.2 |
| CBTL-RC-CCS | -2.3 | 0.0 |
| BTL-RC-V | -13 | -0.1 |
| BTL-RC-CCS | -124 | -1.4 |
| | | |
| CTL-OT-V | 259 | 2.8 |
| CTL-OT-CCS | 118 | 1.3 |
| CTL-OTA-CCS | 85 | 0.9 |
| CBTL-OT-V | 151 | 1.6 |
| CBTL-OT-CCS | 1.0 | 0.0 |
| CBTL2-OT-CCS | 92 | 1.0 |
| CBTL-OTS-CCS | -2.3 | 0.0 |
| CBTL-OTA-V | 192 | 2.1 |
| CBTL-OTA-CCS | 10 | 0.1 |
| CBTL-OTAS-CCS | 0.7 | 0.0 |

One measure of how efficiently biomass is used for making low/zero GHG emitting transportation fuels is the effective yield of liquids per unit of biomass input. Figure 27 shows effective yields of low/zero GHG FTL fuels from biomass for four of the process designs and compares these yields with estimates of yields for production of "cellulosic ethanol", another low/zero GHG emitting liquid fuel. The left three bars are for FTL systems that produce essentially zero GHG fuels, as discussed above. The first of these – BTL-RC-V – produces 249 liters of gasoline equivalent per dry tonne of biomass input. With the CBTL systems, because some of the feedstock is coal, the total liquid fuel produced per unit of biomass input is much higher than for the BTL-RC-V design – more than double in the CBTL-RC-CCS case. The fourth bar is for the BTL-RC-CCS system, which produces FTL fuels with strong negative emissions, providing "room" for use of some petroleum fuels without exceeding zero net GHG emissions for the BTL-RC-CCS fuels + the petroleum fuels. This system produces comparable liquid fuel amounts to the CBTL-RC-CCS system, albeit with some of the liquid fuel being petroleum-derived. With cellulosic ethanol production technology projected to be available by 2010, the liquid fuel yields would approach 250 liters gasoline per dry tonne biomass, or roughly half the yields with either the CBTL-RC-CCS or BTL-RC-CCS systems.

An alternative perspective on the effectiveness with which biomass is used for fuels production is given in Figure 28. This shows (dark shading) the effective amount of biomass energy used when producing an energy unit of low/zero GHG liquid fuels. The BTL-RC-V system effectively requires about 1.8 units input to produce one unit of liquid fuel. For the CBTL systems, less biomass is required (0.6 to 0.8 units) since coal provides a substantial portion of the feedstock input. In the case of the BTL-RC-CCS system, the large negative emissions for this system enable some petroleum fuels to be burned, as discussed in the previous paragraph. This effectively lowers the biomass input per unit of low/zero GHG liquid fuels produced to about 0.8. The biomass requirements for cellulosic ethanol are about 2.1 for technology projected for 2010.

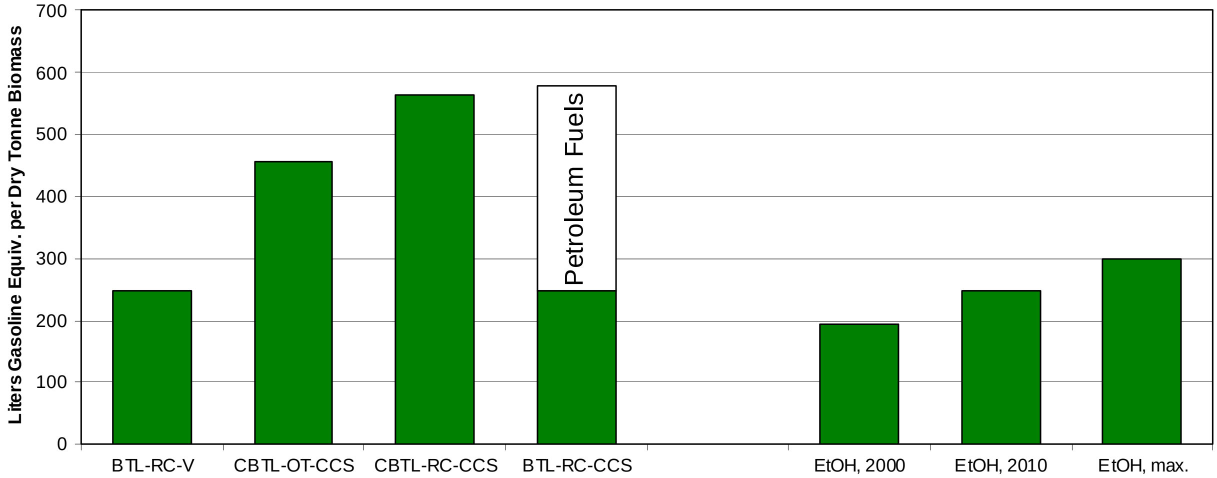


**Figure 27. Yields of low/zero net GHG emitting liquid fuels from biomass. The estimates for cellulosic ethanol are for production from corn stover. The lowest ethanol estimate is for technology considered to be known in 2000 [47]. The middle ethanol estimate is projected to be achievable by 2010 [83]. The right-most ethanol estimate is the maximum theoretically achievable yield from corn stover [47].**

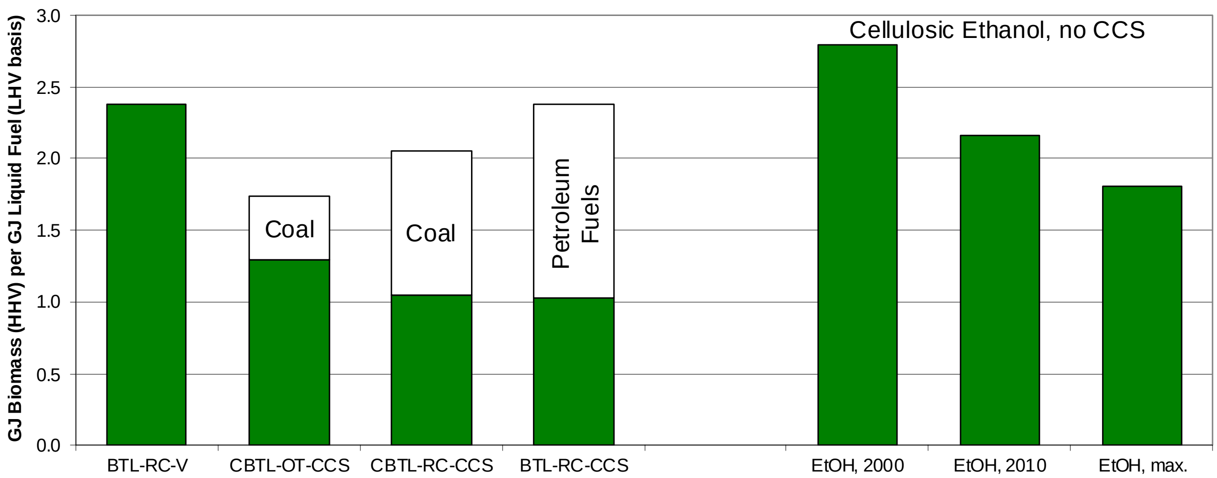


**Figure 28. Biomass input requirements per unit of low/zero net GHG emitting liquid fuels produced. For the CBTL systems, the amount shown for coal is the total coal used less a credit for coal that would be used to produce the same amount of electricity in a coal integrated gasification combined cycle stand-alone power plant with an efficiency of 32.5% (HHV). The ethanol estimates are for corn stover, as described in the caption to Figure 27.**

# 4 Capital Cost Estimates

For each of our plant designs, we developed capital cost estimates using a consistent set of data sources and assumptions.

We estimated capital costs on a disaggregated basis for each major plant area drawing on published literature. We scrutinized the numbers given in these sources and made adjustments to original figures as appropriate (as discussed below) to develop a self-consistent set of capital cost estimates. We have expressed all costs in constant mid-2007 U.S. dollars. To capture the higher than economy-wide average cost escalations observed in recent years in plant materials, equipment, engineering, and construction, we have used the Chemical Engineering Plant Cost

Index [48] to convert from original-year dollars when necessary. Since 2000, the CEPCI has grown considerably more rapidly than the economy as a whole, as represented by the GDP deflator (Figure 29).

## 4.1 Method for Estimating Total Plant Cost (TPC) and Total Plant Investment (TPI)

Following the ERPI-TAG methodology [72], we define the total plant cost (TPC) as the "overnight" capital investment required to construct a plant and the total plant investment (TPI) as the TPC plus interest during construction.

To estimate TPC, we subdivided each plant into major process areas and sub-components within these areas. We estimate the capital cost for each component and sum these to obtain the TPC. (We separately calculate interest during construction, as discussed in Section 5.1.) Estimating component costs involved first estimating from the literature or in consultation with industry experts the "base cost" ($C_o$) for each component. Base costs include installation, but generally exclude balance of plant (BOP) and indirect costs (IC) such as engineering and contingencies. (BOP and IC are discussed further below). The base cost is for a particular component size (capacity), $S_o$, which in most cases is different from the component size required ($S$), as determined from our process simulations. We scale $C_o$ with the physical parameter appropriate to the particular component[11] to estimate the cost of the component ($C$) at the required size.

Since there are maximum physical limits to the size of any given component, if the required capacity or size ($S_r$) from our simulation results exceeds the practical maximum capacity ($S_{max}$) then multiple trains must be installed for the component. For some components, we are able to estimate $S_{max}$. In these cases, the number of trains required ($n$) is calculated as the nearest integer number greater than ($S_r$ / $S_{max}$), and the size of each train is

$$S = S_r / n \qquad (1)$$

For other components, we have simply assumed a number of trains. For example, we assume each gasifier in a plant would have associated with it a separate feed preparation area and separate gas cleaning train.

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

$$C = C_o \cdot (S/S_o)^{\wedge f} \qquad (2)$$

where $f$ is the cost scaling factor that typically ranges between one-half and one, and is close to two-thirds for many types of equipment.[12]

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

[12] While the value of 0.67 turns out to apply, approximately, for many types of equipment, it's origin is most apparent when considering a spherical reaction vessel. For such a vessel, the cost scales roughly with the amount of material needed to make the vessel, which in turn is closely related to its surface area: cost ~ $4\pi r^2$. Thus, for two vessels with different capacities, the ratio of costs is $cost_1/cost_2 \sim (r_1/r_2)^2$. Since the capacity of a spherical vessel is

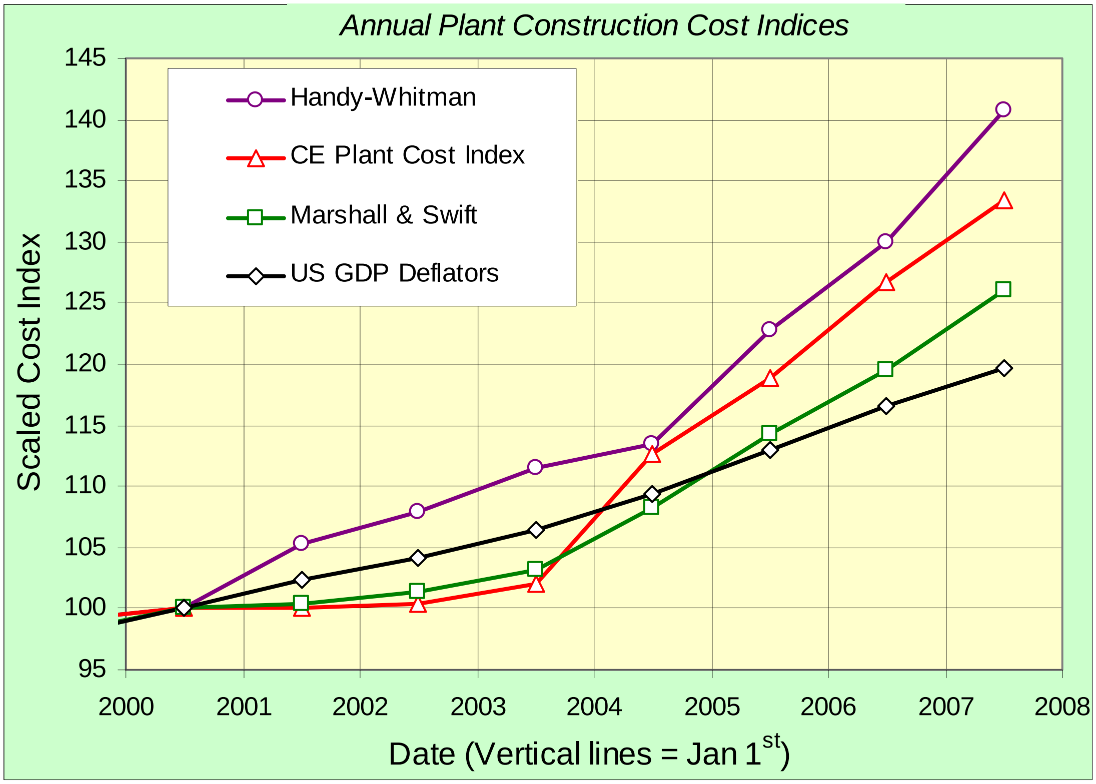


**Figure 29. Cost indices, each normalized to year 2000 value. The Chemical Engineering Plant Cost Index [48], published monthly since 1963, is a composite of indices for equipment costs, construction labor costs, buildings costs, and engineering & supervision costs [[49]]. Marshall & Swift/Boeckh is a leading supplier of construction cost information for construction projects throughout the United States. The Marshall & Swift Index is published monthly in Chemical Engineering [48]. The US GDP Deflator is the annual average implicit price deflator for gross domestic product published monthly by the U.S. Department of Commerce [82]. The Handy-Whitman "Total Plant-All Steam Generation" Index [50] is taken from [51].**

For multiple unit trains of equal size, the installed cost of each additional train will be somewhat less than the cost of the first train. This is because the two units may share some auxiliary equipment, the labor required to install two units is generally less than double that required for one, and the special machining and shop costs to construct the first unit are not all duplicated to construct the second. We capture this idea using the trained cost, $C_m$, of a unit:

$$C_m = C \cdot n^m \tag{3}$$

where m is the scaling exponent for multiple trains, which we assume has value of 0.9.

The original sources from which costs were drawn reported balance of plant (BOP) costs (instrumentation and controls, building, civil works, electrical connections, piping, insulation, and site preparation) and indirect costs (engineering, head office, startup, contingencies, etc.) inconsistently, in some cases explicitly stated, in others not. We adjusted the trained cost, $C_m$, on a component-by-component basis in an effort to ensure that BOP costs and indirect costs were included for all components.

For plant components that did not already include indirect costs in the values of $C_m$, Table 21 summarizes our assumptions, which are based on a review of literature estimates for similar

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proportional to its volume $(4/3)\pi r^3$, we can write capacity$_1$/capacity$_2$ ~ $(r_1/r_2)^3$ or rearranging, this gives $(r_1/r_2)$ ~ (capacity$_1$/capacity$_2$)$^{1/3}$ . Putting this latter expression into the cost ratio gives cost$_1$/cost$_2$ ~ (capacity$_1$/capacity$_2$)$^{2/3}$.

plants. The relatively low percentages for contingencies is due to the fact that we are considering Nth plant (commercially mature) systems.

**Table 21. Indirect cost assumptions, as percent of $C_m$.**

| | Biomass gasifier and gas cleanup components | Power island components |
|---|---|---|
| Engineering & head office | 15% | 15% |
| Contingency | 10% | 5% |
| Startup | 5% | 5% |
| Royalties, fees, etc. | 1% | 1% |
| Spare parts | 1% | 1% |
| **Total** | **32%** | **27%** |

For BOP costs, we used two sets of assumptions. For component costs, $C_m$, that are based on costs given by NETL [55], we assume that BOP costs are 15.5% of $C_m$. For other components that did not already have BOP costs included in $C_m$, we estimated the BOP cost as a percentage of $C_m$ as follows. Hamelinck and Faaij [61] have noted that the absolute cost of a power plant will grow more quickly with capacity than the BOP portion of the cost, so BOP as a % of total cost will be smaller the larger the plant size. We confirmed this by a careful review of literature cost studies for similar plants at varying scales. We have estimated an overall BOP percentage as a function of the higher heating value biomass energy input from a best fit of several literature estimates [52,53,54,56,61]:

$$\text{BOP (\%)} = 0.8867 / \{(\text{biomass MW}_{\text{HHV}})^{0.2096}\} \tag{4}$$

The sum of $C_m$ (adjusted by BOP and indirect cost percentages) gives the total plant (TPC).

We calculate total plant investment assuming a 3-year construction period (which is expected to be characteristic of a fully mature industry) and equal annual expenditures during the construction period. The assumed interest rate during construction is as discussed in Section 5.1.

## 4.2 Base Costs and Cost Scaling Factors

Table 22 through Table 25 give the base costs and cost scaling factors used for all of our estimates of Total Plant Capital costs. The notes given with each table describe the sources and derivation of specific values.

**Table 22. Gasification island reference equipment capacities, costs, and scaling factors.**

| *Plant component* | *Scaling parameter* | *Base capacity, So* | *Max capacity, Smax* | *Cost scaling factor* | *Base cost, Co (2007 M$)* | *BOP cost (%)* | *Indirect cost (%)* | *Notes* |
|---|---|---|---|---|---|---|---|---|
| ***Air separation unit (ASU), and O2 and N2 compression*** | | | | | | | | |
| Delivering at 75 bar pressure | Pure O2, mt/day | 2,202 | | 0.50 | 105 | 15.5% | included | (a) |
| ***Coal handling and gasification*** | | | | | | | | |
| Coal handling | AR coal, mt/day | 5,447 | | 0.67 | 36.4 | 15.5% | included | (a) |
| GE - coal preparation & feeding | AR coal, mt/day | 2,724 | | 0.67 | 31.4 | 15.5% | included | (a) |
| GE - ash/slag dewatering & handling | Coal ash, mt/day | 528 | | 0.67 | 46.0 | 15.5% | included | (a) |
| GE quench gasifier, SG cooler, black H2O, sour gas | AR coal, MW LHV | 815 | | 0.67 | 68.3 | 15.5% | included | (a) |
| GE - gas scrubbing & low temp cooling | AR coal, MW LHV | 815 | | 0.67 | 18.2 | 15.5% | included | (a) |
| ***Biomass handling and gasification*** | | | | | | | | |
| Biomass storage, prep, handling | Biomass feed, wet mt/hr | 64.6 | 110 | 0.77 | 12.3 | Eqn. 4 | 32% | (b) |
| GTI gasifier, gas cooling, cleaning | AR biomass, MW LHV | 815 | 568 | 0.67 | 198.8 | 15.5% | included | (c) |

(a) Air separation and all coal-related component costs are based on NETL [55] for a system utilizing a water-slurry GE-type radiant-quench gasifier. Original NETL costs in beginning-of-year 2007 dollars have been converted to mid-2007 costs using the annual average CE plant cost index. Cost scaling factor of 0.5 for ASU is as indicated by Kreutz, *et al.* [56]. Cost scaling factor for coal-related components is assumed to be 0.67. In the case of the gasifier cost, since we utilize a GE-type quench gasifier in our simulations, we have multiplied the NETL cost for a radiant-quench gasifier by 0.583, which is the ratio of installed costs (i.e. material, labor, indirects, and contingencies) for a quench gasifier + ash handling to the installed cost for a radiant-quench gasifier + ash handling, following Matchak, *et al.* [57].

(b) Biomass storage, preparation, and handling costs are based on Weyerhaeuser [58], which gives the following installed "N^th plant" equipment costs (in 1999$) for feed preparation to feed 64.6 wet tonnes/hr wood chips to a near-atmospheric pressure gasifier: conveyor, $851,000; dried wood chip storage, $561,000; feed bin, $233,000; rotary air lock, $329,000; and water cooled feed screw, $54,000. We multiply these by a factor of 3.1 to account for the lower bulk density of chopped switchgrass compared to wood chips [59]. (We assume equipment cost scales linearly with bulk density.) Additionally, we have multiplied the resulting total by 1.45 [60] to account for the cost of a feed preparation system rated for 30 bar. We escalate the resulting number to 2007 dollars using the CE plant cost index. We derived the cost scaling factor, *f*, by calculating an overall scaling factor for a feed preparation system with a base capacity of 33.5 wet tonnes/hour from Hamelinck and Faaij [61] consisting of conveyors [$C_o$ = $350,000 (2001$) f = 0.8], storage ($1,000,000, 0.65), and a feeding system such as a lock hopper and feed screw ($410,000, 1). Scaling the resulting total feed preparation cost for values of S/So between 1 and 3.5 gives an overall scaling factor of 0.77. The cost estimates from Hamelinck and Faaij were not used directly here for $C_o$ because they represent first-of-a-kind plant cost estimates, whereas [58] gives estimates for an N^th plant.

(c) For consistency with our coal gasification cost estimates, we developed the cost for the steam/oxygen blown fluidized-bed biomass gasifier plus gas cooling and gas cleaning as follows. The ratio of the bare erected capital cost for an oxygen-blown entrained-flow quench coal gasifier system (including 42 bar gasification, slag removal, and gas scrubbing for particulate removal) to that for a fluidized-bed coal gasifier (including 32 bar gasification in steam and oxygen, high-temperature syngas cooling, particulate removal, and ash removal) was determined to be 2.1, based on two self-consistent studies by Fluor Engineers [62,63]. Both studies were for Illinois #6 coal, and the cost for the fluidized-bed system was scaled from 4950 metric tonnes per day coal input to 6095 t/d coal input (using 0.5 scaling exponent) to match the scale of the entrained-flow gasification system. The resulting ratio was 2.72. This ratio was adjusted for the much higher throughput possible in a fluidized-bed reactor of given size when biomass is the feestock compared to coal. Operating experience in Finland with the U-Gas fluidized-bed gasifier operating on coal and separately on biomass suggests that the higher heating value energy throughput with biomass could be 1.6 times the throughput possible with coal for the same vessel size [64]. Assuming a cost-capacity scaling exponent of 0.5, we arrive at the bare erected capital cost ratio between a pressurized $O_2$-blown fluidized-bed biomass gasification system and a pressurized $O_2$-blown entrained-flow coal gasification system of 2.1 ( = $2.72 / 1.6^{0.5}$). To calculate our reference biomass gasification system cost, we apply this multiplier to our reference cost for coal gasifier plus ash/slag system. Lau [65] indicates that the maximum capacity of a 30-bar GTI gasifier is roughly 2875 metric tonnes per day, which we have adopted as the maximum unit size in our calculations.

**Table 23. Water gas shift, acid gas removal, sulfur processing, $CO_2$ compression, and miscellaneous reference equipment capacities, costs, and scaling exponents.**

| *Plant component* | *Scaling parameter* | *Base capacity, So* | *Max capacity, Smax* | *Cost scaling factor* | *Base cost, Co (2007 M$)* | *BOP cost (%)* | *Indirect cost (%)* | *Notes* |
|---|---|---|---|---|---|---|---|---|
| ***Syngas cleanup*** | | | | | | | | |
| Upstream WGS (1 stage) | Corresponding AR coal, MW LHV | 815 | | 0.67 | 3.36 | 15.5% | included | (a) |
| Downstream WGS (2 stages) | AR coal, MW LHV | 815 | | 0.67 | 8.40 | 15.5% | included | (a) |
| Expanders | Expander pwr, MWe | 3.1 | | 0.67 | 4.18 | 15.5% | included | (b) |
| Rectisol for BTL, incl. refrig. | Syngas feed, Nm^3/hr | 200,000 | 700,000 | 0.63 | 28.8 | Eqn. 4 | 32% | (c) |
| Rectisol for CTL, CBTL, incl. refrig. | Syngas feed, Nm^3/hr | 200,000 | 700,000 | 0.63 | 52.3 | Eqn. 4 | 32% | (c) |
| Rectisol recovery compressor | Compressor power, MWe | 10 | | 0.67 | 6.31 | included | 32% | (d) |
| Claus/SCOT | S input, mt/day | 137 | | 0.67 | 33.8 | 15.5% | included | (e) |
| ***CO2 compression*** | | | | | | | | |
| Subcritical outlet pressure | Compressor power, MWe | 10 | | 0.67 | 6.31 | included | 32% | (f) |
| Supercritical outlet pressure | Compressor power, MWe | 13 | | 0.67 | 9.52 | included | 32% | (g) |

(a) The downstream water gas shift reactor cost is based on NETL [55] for a two-stage (adiabatic reactor, followed by heat exchange and isothermal reactor) system. NETL costs in beginning-of-year 2007 dollars have been converted to mid-2007 costs using the annual average CE PCI. A default cost scaling factor of 0.67 is assumed. The upstream WGS is a single-stage (adiabatic reactor) system that is assumed to be 40% of the cost of a two-stage system with the same syngas input rate [66].

(b) Expander cost is based on NETL [55]. NETL costs in beginning-of-year 2007 dollars were converted to mid-2007 costs using the annual average CE PCI. A default cost scaling factor of 0.67 is assumed.

(c) Separate cost estimates are used for acid gas removal (Rectisol® technology) configurations where $CO_2$ is separately removed from sulfur species ($H_2S$ and COS) – two-column design – and where they are co-removed – one column design. Both estimates are based on information provided by Koss [67]. Separate removal is required if the sulfur is to be recovered via a Claus plant. In this design, $CO_2$ is removed using fresh solvent in the second column, and the resulting $CO_2$-rich solvent is used in the first column to remove sulfur species. In either the one- or two-column design, virtually any level of $CO_2$ removal can be achieved, and essentially 100% sulfur removal is possible. The $CO_2$ can be recovered at nearly 100% purity, while the sulfur species can be recovered at 30%-70% volume concentration, with the balance being $CO_2$. Koss' estimate in 2003 of the installed bare equipment costs (including the requisite refrigeration plant for solvent cooling, but excluding BOP and indirect costs) for a 1-column system with an input syngas flow of 200,000 $Nm^3$/hr was 20 million Euros, or $22 million (2003$) considering that the average dollar-euro exchange rate in 2003 was about $1.1/Euro. For a 2-column system, Koss' estimates were $40 million (2003$) for a 200,000 $Nm^3$/hr capacity system and $62 million (2003$) for a 400,000 $Nm^3$/hr system, which imply a cost scaling factor of 0.63. Some of our process designs include an AGR step upstream of the FT synthesis island and a second AGR downstream. Since the two systems would share a solvent regeneration system, the costs for such a dual upstream/downstream Rectisol system can be approximated as the cost a single Rectisol system with capacity equal to the sum of the capacity of the upstream unit and half the capacity of the downstream unit.

(d) Due to small but finite solubilities of $H_2$, CO, and $CH_4$ in Rectisol solvent, some quantities of these syngas components are always incidentally removed along with the acid gases. In our designs, the incidentally-removed gases are separated from the $CO_2$ and $H_2S$ and fed back via a compressor to the main syngas stream leaving the Rectisol plant. The cost for this compressor is based on that given for an $N^{th}$ plant purge gas compressor by Kreutz *et al.* [56]: $6.28 million (2002$) for a 10 MWe unit, including installation, BOP, engineering and contingencies, and are given for an Nth plant design. Engineering and contingency costs are removed from the cost reported in the table (and re-added later, as discussed in the text), and the resulting cost was escalated to 2007$ using the CE PCI. The value assumed for *f* is also from Kreutz et al. [56].

(e) The cost of the Claus/SCOT system for sulfur recovery is based on NETL [55]. Original NETL costs in beginning-of-year 2007 dollars have been converted to mid-2007 costs using the annual average chemical plant cost engineering index. A default cost scaling factor of 0.67 is assumed.

(f) The cost of compressors for $CO_2$ compression to subcritical pressures is based on Kreutz, *et al.* [56], as discussed in note (d) of this table.

(g) The cost and scale factor for $CO_2$ compression to supercritical pressure is based on Kreutz, *et al.* [56], who give a cost of $14.8 million (2002$) for a $CO_2$ dehydration and compression system (to 150 bar) at a scale of 13 $MW_e$. Of this total, 36% is due to the dehydration equipment, which is not needed in our designs because the $CO_2$ emerges dry from the Rectisol process. Kreutz's cost includes BOP, indirects (15%), and contingencies (15%). The cost in this table includes BOP, but indirects and contingencies have been factored out (and re-added later in the analysis, as discussed in the text). Costs were converted to 2007$ using the CE PCI.

**Table 24. Liquid fuel synthesis reference equipment capacities, costs, and scaling exponents.**

| *Plant component* | *Scaling parameter* | *Base capacity, So* | *Max capacity, Smax* | *Cost scaling factor* | *Base cost, Co (2007 M$)* | *BOP cost (%)* | *Indirect cost (%)* | *Notes* |
|---|---|---|---|---|---|---|---|---|
| ***Fischer-Tropsch synthesis; product distillation and refining*** | | | | | | | | |
| FT fresh feed compressor | Compressor power, MWe | 10 | | 0.67 | 6.31 | included | 32% | (a) |
| FT synthesis; slurry reactor and HXs | FT input vol flowrate, MM SCF/hr | 2.5 | 10.0 | 0.75 | 13.60 | Eqn. 4 | 32% | (a) |
| HC recovery plant | FT flow to HC recovery, Mlb/hr | 14 | 200 | 0.70 | 0.724 | Eqn. 4 | 32% | (a) |
| H2 recovery (PSA, etc.) | FT H2 recovery plant, MM CF/hr | 0.0 | 0.2 | 0.70 | 0.833 | Eqn. 4 | 32% | (a) |
| Wax hydrocracking | FT HC inlet flow, Mlb/hr | 9.0 | 575.0 | 0.55 | 9.30 | Eqn. 4 | 32% | (a) |
| Distillate hydrotreating | FT distillate HC inlet flow, Mlb/hr | 2.9 | 650.0 | 0.60 | 2.49 | Eqn. 4 | 32% | (a) |
| Naphtha hydrotreating | FT naphtha HC inlet flow, Mlb/hr | 2.1 | 650.0 | 0.65 | 0.754 | Eqn. 4 | 32% | (a) |
| Naphtha reforming | FT naphtha reforming, Mlb/hr | 3.4 | 750.0 | 0.60 | 5.18 | Eqn. 4 | 32% | (a) |
| C5/C6 isomerization | FT C5/C6 isomerization, Mlb/hr | 1.2 | 250.0 | 0.62 | 0.956 | Eqn. 4 | 32% | (a) |
| H2-feed, PSA-feed comps. | Compressor power, MWe | 10 | 70 | 0.67 | 6.31 | included | 32% | (b) |
| Fuel gas comp. inside refinery | Compressor power, MWe | 10 | 70 | 0.67 | 6.31 | included | 32% | (b) |
| Recycle compressor | Compressor power, MWe | 10 | 70 | 0.67 | 6.31 | included | 32% | (b) |
| Auto-thermal reformer | ATR input SG, MM SCF/day | 365 | 8,000 | 0.67 | 29.9 | Eqn. 4 | 32% | (c) |
| H2 expander in FT refinery | Expander power, MWe | 10 | 70 | 0.67 | 3.16 | included | 32% | (d) |

(a) The costs and cost scaling factors for the FT synthesis reactor and FTL upgrading components (originally in 1994$ and assumed to exclude BOP and indirect costs) are from Bechtel [35], which in turn was based on work done by Bechtel earlier in the 1990s [32]. Original costs were converted to mid-2007 costs using the annual average CE PCI.

(b) Compressor costs are based on Kreutz, *et al.* [56], as discussed in note (d) of Table 23.

(c) The cost (in 1991$) for an autothermal reformer is from Bechtel [68], and is assumed to exclude BOP and indirect costs. Original costs were converted to mid-2007 costs using the annual average CE PCI. A default cost scaling factor of 0.67 is assumed.

(d) For expander cost, see note (b) of Table 23.

**Table 25. Power island reference equipment capacities, costs, and scaling factors.**

| *Plant component* | *Scaling parameter* | *Base capacity, So* | *Max capacity, Smax* | *Cost scaling factor* | *Base cost, Co (2007 M$)* | *BOP cost (%)* | *Indirect cost (%)* | *Notes* |
|---|---|---|---|---|---|---|---|---|
| ***Gas Turbine topping cycle*** | | | | | | | | |
| GT fuel gas compressor | Compressor power, MWe | 10 | | 0.67 | 6.31 | included | 32% | (a) |
| Fuel gas saturator | Saturator SG output, m^3/s | 21 | 64 | 0.70 | 0.374 | Eqn. 4 | 32% | (b) |
| Gas turbine | Net output power, MWe | 266 | | 0.75 | 73.2 | included | 27% | (c) |
| ***Heat recovery and steam cycle*** | | | | | | | | |
| Boiler/steam generator, ductwork, & stack | Boiler duty, MWth | 355 | | 1.00 | 52.0 | Eqn. 4 | 27% | (d) |
| Steam cycle (turbine, condenser, piping, auxiliaries) | ST gross power, MWe | 275 | | 0.67 | 66.7 | 15.5% | included | (e) |

(a) Compressor costs are based on Kreutz, *et al.* [56], as discussed in note (d) of Table 23.

(b) The cost for a saturator is modeled as the cost for a scrubber , which utilizes essentially the same basic equipment and layout: a vertical vessel with upward gas flow and spray jets for downward liquid flow. Unlike a scrubber, the volumetric flow rate of gas exiting a saturator determines the unit size (rather than the volumetric inlet flow of gas). Weyerhaeuser [58] indicates an installed scrubber cost of $206k (1999 dollars) for an inlet mass flow of 21 $m^3$/sec of syngas at near-atmospheric pressure. We multiply this value by 1.35 [60] to account for a required pressure rating of 27 bar. We escalate the result to 2007$ using the CE PCI. The values of $S_{max}$ and *f* are from Hamelinck *et al*. [69].

(c) Kreutz *et al*. [56] indicates an installed cost of $72.8 million (2002$) for a Siemens V94.3A gas turbine (266 $MW_e$ output). This cost includes BOP, installation, engineering (15% of installed gas turbine +BOP), and contingencies (15% of installed gas turbine +BOP +engineering). The value in this table for base cost excludes the engineering and contingencies and has been updated to 2007$ using the CE CPI. Engineering and contingencies are included (using methodology discussed in text) when total plant cost is estimated later in this paper. The assumed scaling exponent was obtained by taking a power-series regression of equipment-only costs vs. power output for all simple cycle gas turbine generators in Gas Turbine World's 2003 Handbook [70]. The value for $S_{max}$ is the power rating of the largest simple-cycle gas turbine generator in Gas Turbine World's 2003 Handbook. For some of the plant designs discussed later, the included gas turbine capacity is considerably smaller than the capacity of a Siemens V94.3A turbine. Our process simulation assumes the unit performance will be the same in all cases, but our cost estimation methodology accounts for higher unit costs for smaller units.

(d) Boiler/steam generator cost is based on Simbeck [71], who indicates a cost of $110/$kW_{thermal}$ of superheated steam for an "HRSG Boiler" (2000 $), or $39 million for a unit producing 355 MW superheated steam. Simbeck's cost excludes BOP and indirects. We apply this cost to the heat transfer duty of the boiler or HRSG and to heat exchange that occurs upstream and/or downstream of the AGR system. (The costs for all other heat exchange equipment in our process configurations are included in the cost for the associated main process equipment, e.g., water gas shift reactor or FT synthesis reactor.) We assume a cost scale factor of unity, which may be conservative.

(e) The steam turbine and condenser cost are based on NETL [55], with original mid-2006 dollars converted to mid-2007 dollars using the annual average CE PCI. A default cost scaling factor of 0.67 is assumed.

## 4.3 Total Plant Costs

The process simulations described in Section 3.2 provide the total required capacities for components within each island (gasification island, acid gas removal island, synthesis island, power island, etc.) of a given system. To apply the base equipment costs and scaling factors given in the tables in the previous section requires a decision on the number of trains to be used to meet the requisite component capacities.

Table 26 and Table 27 provide the training assumptions in our analysis. In general, the number of trains takes into account limits on the practical physical size of different components, based on commercial experience as described in literature sources and industry practitioners that we consulted. Note that the number of trains assumed for some components in the CBTL2-OT-CCS design (which is identical in configuration to CBTL-OT-CCS, but uses a much higher coal-to-biomass input ratio) are non-integers. This arises because for this one case (among the sixteen described here) the characteristics of this plant were estimated from an interpolation between two simulated plants.

**Table 26. Number of equipment trains used for capital cost estimate for each RC design.**

| | CTL-RC-V | CTL-RC-CCS | CBTL-RC-V | CBTL-RC-CCS | BTL-RC-V | BTL-RC-CCS |
|---|---|---|---|---|---|---|
| ***Air separation unit (ASU), and O2 and N2 compression*** | | | | | | |
| ASU (stand alone) | 7 | 7 | 2 | 2 | 1 | 1 |
| O2 compressor for C-gasifier | 2 | 2 | 1 | 1 | 0 | 0 |
| O2 comp for B-gasifier or ATR | 1 | 1 | 1 | 1 | 1 | 1 |
| N2 compressor to GT | 0 | 0 | 0 | 0 | 0 | 0 |
| N2 expander | 0 | 0 | 0 | 0 | 0 | 0 |
| ***Biomass handling and gasification*** | | | | | | |
| Biomass storage, prep, hand. | 0 | 0 | 2 | 2 | 2 | 2 |
| GTI gasifier, gas cooling, cleaning | 0 | 0 | 2 | 2 | 2 | 2 |
| ***Coal handling and gasification*** | | | | | | |
| Coal storage, prep, hand. | 10 | 10 | 2 | 2 | 0 | 0 |
| GEE quench coal gasifier | 10 | 10 | 2 | 2 | 0 | 0 |
| ***Syngas cleanup*** | | | | | | |
| C-WGS reactor (1-stage) and HXs | 1 | 1 | 1 | 1 | 0 | 0 |
| WGS bypass expander | 1 | 1 | 1 | 1 | 0 | 0 |
| Rectisol system | 5 | 5 | 1 | 1 | 1 | 1 |
| Rectisol recovery compressor | 1 | 1 | 1 | 1 | 1 | 1 |
| Downstream WGS (2-stage) & HXs | 0 | 0 | 0 | 0 | 0 | 0 |
| Claus/SCOT | 4 | 4 | 1 | 1 | 0 | 0 |
| ***CO2 compression*** | | | | | | |
| CO2 compressor 1, subcritical | 0 | 1 | 0 | 1 | 0 | 1 |
| CO2 compressor 2, subcritical | 0 | 0 | 1 | 1 | 1 | 1 |
| CO2 compressor 3, supercritical (150 bar) | 0 | 2 | 0 | 1 | 0 | 1 |
| ***Fischer-Tropsch synthesis; product distillation and refining*** | | | | | | |
| FT fresh feed compressor | 1 | 1 | 1 | 1 | 1 | 1 |
| FT synthesis; slurry reactor and HXs | 9 | 9 | 2 | 2 | 1 | 1 |
| HC recovery plant | 9 | 9 | 2 | 2 | 1 | 1 |
| H2 production (PSA, etc.) | 7 | 7 | 2 | 2 | 1 | 1 |
| Wax hydrocracking | 1 | 1 | 1 | 1 | 1 | 1 |
| Distillate hydrotreating | 1 | 1 | 1 | 1 | 1 | 1 |
| Naphtha hydrotreating | 1 | 1 | 1 | 1 | 1 | 1 |
| H2-feed, PSA-feed comps. | 1 | 1 | 1 | 1 | 1 | 1 |
| Fuel gas comp. inside refinery | 1 | 1 | 1 | 1 | 1 | 1 |
| FT recycle compressor | 1 | 1 | 1 | 1 | 1 | 1 |
| Auto-thermal reformer | 1 | 1 | 1 | 1 | 1 | 1 |
| H2 expander in FT refinery | 1 | 1 | 1 | 1 | 1 | 1 |
| ***Naptha upgrading to gasoline*** | | | | | | |
| Naphtha reforming | 1 | 1 | 1 | 1 | 1 | 1 |
| C5/C6 isomerization | 1 | 1 | 1 | 1 | 1 | 1 |
| ***Gas Turbine topping cycle*** | | | | | | |
| C-SG expander | 1 | 1 | 1 | 1 | 0 | 0 |
| GT fuel gas compressor | 0 | 0 | 0 | 0 | 0 | 0 |
| Fuel gas saturator | 0 | 0 | 0 | 0 | 0 | 0 |
| Gas turbine (GT) | 0 | 0 | 0 | 0 | 0 | 0 |
| ***Heat recovery and steam cycle*** | | | | | | |
| HRSG + heat exchanger network | 1 | 1 | 1 | 1 | 1 | 1 |
| Steam cycle (ST + condenser) | 1 | 1 | 1 | 1 | 1 | 1 |

Using our base costs, cost scaling factors, and training assumptions, we calculated total plant capital costs as indicated in Table 28 for all of the RC configurations and in Table 29 for all of the OT designs.

For the CTL-RC-V design, with a capacity to produce 50,000 barrels per day (bpd) of gasoline and diesel blendstock, the total plant capital (TPC) is $4.9 billion, or about $98,000 $/bpd. Adding CCS to this design increases the cost only modestly, since the only added cost is for compression of the $CO_2$ that is captured as an intrinsic element of the CTL-RC-V design. The CBTL-RC-V and CBTL-RC-CCS designs produce only about 10,000 bpd and the associated cost penalties arising from the smaller scale relative to the CTL designs are evident in the higher $/bpd costs for the CBTL cases. The BTL systems are of even smaller capacity (about 4,400 bpd). Despite lower oxygen requirements per unit feedstock to the gasifier, lower AGR costs, and no Claus/SCOT plant due to the very small amount of $H_2S$ in the raw syngas, the specific TPC for the BTL cases still fall above those for the CBTL designs.

The systems utilizing OT synthesis (Table 29) are characterized by higher specific TPC values than their corresponding RC counterparts due primarily to the reduced liquids production (for the same feedstock input rates) and added capital investments in the power island.

**Table 27. Number of equipment trains used for capital cost estimate of all OT designs. See text regarding CBTL2-OT-CCS.**

| | *CTL-OT-V* | *CTL-OT-CCS* | *CTL-OTA-CCS* | *CBTL-OT-V* | *CBTL-OT-CCS* | *CBTL2-OT-CCS* | *CBTL-OTS-CCS* | *CBTL-OTA-V* | *CBTL-OTA-CCS* |
|---|---|---|---|---|---|---|---|---|---|
| ***Air separation unit (ASU), and O2 and N2 compression*** | | | | | | | | | |
| ASU (stand alone) | 6 | 6 | 7 | 2 | 2 | 5.8 | 2 | 2 | 3 |
| O2 compressor for C-gasifier | 2 | 2 | 2 | 1 | 1 | 2 | 1 | 1 | 1 |
| O2 comp for B-gasifier or ATR | 0 | 0 | 1 | 1 | 1 | 0.2 | 1 | 1 | 1 |
| N2 compressor to GT | 0 | 2 | 1 | 0 | 1 | 2 | 1 | 0 | 1 |
| N2 expander | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| ***Biomass handling and gasification*** | | | | | | | | | |
| Biomass storage, prep, hand. | 0 | 0 | 0 | 2 | 2 | 1.1 | 2 | 2 | 2 |
| GTI gasifier, gas cooling, cleaning | 0 | 0 | 0 | 2 | 2 | 1.1 | 2 | 2 | 2 |
| ***Coal handling and gasification*** | | | | | | | | | |
| Coal storage, prep, hand. | 10 | 10 | 10 | 2 | 2 | 9.3 | 3 | 3 | 3 |
| GEE quench coal gasifier | 10 | 10 | 10 | 2 | 2 | 9.3 | 3 | 3 | 3 |
| ***Syngas cleanup*** | | | | | | | | | |
| C-WGS reactor (1-stage) and HXs | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| WGS bypass expander | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Rectisol system | 3 | 4 | 4 | 1 | 1 | 3.8 | 1 | 1 | 2 |
| Rectisol recovery compressor | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Downstream WGS (2-stage) & HXs | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 1 | 1 |
| Claus/SCOT | 4 | 4 | 4 | 1 | 1 | 3.8 | 1 | 0 | 1 |
| ***CO2 compression*** | | | | | | | | | |
| CO2 compressor 1, subcritical | 0 | 1 | 1 | 0 | 1 | 1 | 1 | 1 | 1 |
| CO2 compressor 2, subcritical | 0 | 2 | 3 | 1 | 1 | 2 | 1 | 0 | 1 |
| CO2 compressor 3, supercritical (150 bar) | 0 | 1 | 1 | 0 | 1 | 1 | 1 | 0 | 1 |
| ***Fischer-Tropsch synthesis; product distillation and refining*** | | | | | | | | | |
| FT fresh feed compressor | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 3 | 1 |
| FT synthesis; slurry reactor and HXs | 7 | 7 | 7 | 2 | 2 | 7 | 3 | 3 | 3 |
| HC recovery plant | 7 | 7 | 7 | 2 | 2 | 7 | 3 | 2 | 3 |
| H2 production (PSA, etc.) | 5 | 5 | 5 | 2 | 2 | 5 | 2 | 1 | 2 |
| Wax hydrocracking | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Distillate hydrotreating | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Naphtha hydrotreating | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |
| H2-feed, PSA-feed comps. | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Fuel gas comp. inside refinery | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |
| FT recycle compressor | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 | 0 |
| Auto-thermal reformer | 0 | 0 | 1 | 0 | 0 | 0 | 0 | 1 | 1 |
| H2 expander in FT refinery | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |
| ***Naptha upgrading to gasoline*** | | | | | | | | | |
| Naphtha reforming | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| C5/C6 isomerization | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |
| ***Gas Turbine topping cycle*** | | | | | | | | | |
| C-SG expander | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| GT fuel gas compressor | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |
| Fuel gas saturator | 0 | 1 | 1 | 0 | 1 | 0.8 | 1 | 2 | 1 |
| Gas turbine (GT) | 3 | 3 | 3 | 1 | 1 | 3.2 | 2 | 0 | 1 |
| ***Heat recovery and steam cycle*** | | | | | | | | | |
| HRSG + heat exchanger network | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 1 |
| Steam cycle (ST + condenser) | 1 | 1 | 1 | 1 | 1 | 1 | 1 | 0 | 1 |

**Table 28. Total plant capital costs for recycle (RC) configurations.**

| | *CTL-RC-V* | *CTL-RC-CCS* | *CBTL-RC-V* | *CBTL-RC-CCS* | *BTL-RC-V* | *BTL-RC-CCS* |
|---|---|---|---|---|---|---|
| ***Plant capital costs, million 2007$*** | | | | | | |
| Air separation unit (ASU), and O2 and N2 compression | 815 | 815 | 209 | 209 | 94 | 95 |
| Biomass handling, gasification, and gas cleanup | 0 | 0 | 265 | 265 | 266 | 266 |
| Coal handling, gasification, and quench | 1,476 | 1,476 | 237 | 237 | 0 | 0 |
| All water gas shift, acid gas removal, Claus/SCOT | 844 | 844 | 163 | 163 | 58 | 58 |
| CO2 compression | 0 | 67 | 1 | 22 | 1 | 13 |
| Fischer-Tropsch synthesis & refining, recycle compressor, ATRs | 766 | 766 | 229 | 229 | 126 | 127 |
| Naptha upgrading to gasoline (isomerization, catalytic reforming) | 103 | 103 | 41 | 41 | 26 | 26 |
| Power island topping cycle | 35 | 35 | 7 | 7 | 0 | 0 |
| Heat recovery and steam cycle | 840 | 840 | 164 | 164 | 64 | 64 |
| ***Total plant cost ( TPC ), million 2007$*** | 4,878 | 4,945 | 1,316 | 1,336 | 636 | 648 |
| ***Specific TPC , $ per bbl/day*** | 97,568 | 98,908 | 131,478 | 133,530 | 144,135 | 146,714 |

**Table 29. Total plant capital costs for once-through (OT) configurations.**

| | CTL-OT-V | CTL-OT-CCS | CTL-OTA-CCS | CBTL-OT-V | CBTL-OT-CCS | CBTL2-OT-CCS | CBTL-OTS-CCS | CBTL-OTA-V | CBTL-OTA-CCS |
|---|---|---|---|---|---|---|---|---|---|
| ***Plant capital costs, million 2007$*** | | | | | | | | | |
| Air separation unit (ASU), and O2 and N2 compression | 705 | 741 | 836 | 200 | 213 | 722 | 277 | 261 | 350 |
| Biomass handling, gasification, and gas cleanup | 0 | 0 | 0 | 265 | 265 | 199 | 264 | 264 | 264 |
| Coal handling, gasification, and quench | 1,476 | 1,476 | 1,476 | 274 | 274 | 1,372 | 472 | 472 | 472 |
| All water gas shift, acid gas removal, Claus/SCOT | 638 | 730 | 765 | 136 | 154 | 684 | 232 | 210 | 287 |
| CO2 compression | 0 | 59 | 74 | 1 | 24 | 62 | 32 | 1 | 37 |
| Fischer-Tropsch synthesis & refining, recycle compressor, ATRs | 523 | 523 | 799 | 169 | 169 | 523 | 240 | 240 | 382 |
| Naptha upgrading to gasoline (isomerization, catalytic reforming) | 86 | 86 | 86 | 36 | 36 | 86 | 47 | 47 | 47 |
| Power island topping cycle | 278 | 286 | 241 | 81 | 84 | 291 | 129 | 126 | 101 |
| Heat recovery and steam cycle | 701 | 695 | 824 | 161 | 160 | 678 | 250 | 256 | 298 |
| ***Total plant cost (TPC), million 2007$*** | 4,407 | 4,598 | 5,101 | 1,324 | 1,379 | 4,617 | 1,944 | 1,877 | 2,239 |
| ***Specific TPC, $ per bbl/day*** | 120,236 | 125,432 | 139,176 | 163,412 | 170,189 | 125,946 | 149,092 | 143,737 | 171,481 |

# 5 Economic Analysis Assumptions

To provide a consistent framework for our economic analysis, we developed a baseline set of financial parameter and other assumptions.

All costs in this paper are presented in constant (inflation-corrected) 2007 US dollars, and expenditures at different points in time are reconciled assuming a 7% per year real discount rate.

The core of the economic analysis is a comparison of alternative FTL options with respect to levelized production costs, although internal rate of return on equity calculations are also carried out—and the latter may be especially useful in understanding the likely competition among alternative process configurations in a world of sustained high oil prices.

## 5.1 Capital Charge Rates

The key financial parameter underlying a levelized production cost analysis is the levelized capital charge rate (LACCR, the annual percentage of TPI that is charged against the final energy products sold each year). Here the LACCR is calculated using the EPRI-TAG methodology [72], assuming a 7% real discount rate equal to the pre-tax weighted cost of capital. Assuming a 55/45 debt/equity ratio, the plausible combination of real debt and equity costs shown in Table 29 is assumed that is consistent with a 7% real discount rate (*dr*) as defined above.

Under these conditions the real levelized capital charge rate (*LCCR*) is 14.38%/year, assuming that the Owner Cost (OC) as a percentage of TPC is 10%. This OC/TPC value was chosen based on [73].

**Table 30. Financial analysis parameter assumptions.**

| Parameter | Assumed Value |
|---|---|
| ***For calculation of real discount rate, dr*** | |
| Debt fraction of capital investment | 55 |
| Equity fraction of capital investment | 45 |
| Real cost of debt | 4.4% |
| Real cost of equity | 10.2% |
| Corporate income tax rate | 39.2% |
| ***Additional assumptions*** | |
| Owner costs as % of TPC | 10% |
| Construction time | 3 years |
| Property tax and insurance rate | 2% |
| Book life and tax life of investment | 20 years |
| Depreciation schedule | MACRS |
| Capacity factor for FTL plants | 90% |
| Capacity factor for stand-alone power generation | 85% |

To estimate the multiplier of TPC in the annual capital rate that is charged against the final energy products sold each year note that *TPI = IDCF x TPC*, where *IDCF* is the interest during construction factor. To calculate interest during construction, we assume a plant construction time of three years, with equal yearly capital expended at the end of each year during construction, so that $IDCF = [1 + (1 + dr) + (1 + dr)^2]/3 = 1.0716$, so that interest during construction equal to 7.16% of TPC. Thus, the multiplier of TPC to calculate total annual capital charges is *LCCR x IDCF* = 15.41%/year.

## 5.2 Capacity factors

Synfuel plants can be expected to operate with capacity factors of 90% or higher; we assume a 90% capacity factor. For electric power generating plants, we assume a capacity factor of 85%.

## 5.3 Operation and maintenance costs

Detailed estimation of annual operation and maintenance (O&M) costs were not carried out for this study but rather were assumed to be 4% of total plant capital (TPC). This assumption appears to be reasonable based on a comparison to detailed estimates O&M cost estimates made for gasification-based facility by NETL [55]. For example, for the GE Energy IGCC cases in that study the ratio of O&M costs to TPC at 85% capacity factor is 4.6% for the case with $CO_2$ vented and 4.4% for the case with CCS.[13]

## 5.4 Feedstock prices

We assume the prices paid for "as-received" feedstocks (Table 3) delivered to the plant gate are $\$1.71/GJ_{HHV}$ for coal and $\$5/GJ_{HHV}$ for herbaceous biomass.

The assumed coal price is the same as that assumed in [55], which was the average price projected for US coal power generators in 2010 by the Energy Information Administration (EIA) in its *Annual Energy Outlook 2007* report [74].[14]

The assumed biomass price is consistent with delivered biomass price estimates made using a detailed state-by-state biomass logistics model for mixed prairie grasses and corn stover [75]. (For a number of states prices estimated via the detailed logistics modeling are lower than $5/GJ, which can thus be considered to be a price that is not overly optimistic.)

## 5.5 Greenhouse gas emissions valuation

Simulating a carbon policy, we consider a range of prices on GHG emissions, ranging from zero to \$100 per $tCO_{2equiv}$. The GHG emissions price is charged on the total lifecycle emissions, which includes emissions associated with upstream activities (such as coal mining and biomass production and transportation), all emissions at the plant from the conversion process, and all downstream activities (including delivery of FTL fuels to refueling stations and ultimate combustion of FTL fuels in vehicles).

## 5.6 Electricity valuation

An important parameter in the financial analysis, especially for the FTL OT cases, is the price assumed for electricity exported from an FTL facility. For simplicity, we assume that

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[13] For a typical system we have modeled in this study (e.g., CTL-OT-CCS) the overall levelized cost would be ~3% higher if the O&M cost turned out to be 4.6% of TPC instead of 4% of TPC (see Table 32 in Section 6.1) .

[14] If instead the assumed price were taken to be the levelized price projected for US coal power generators according to the EIA's *Annual Energy Outlook 2008* report [76], it would be about the same ($1.66/GJ).

when the GHG emissions value is \$0/t $CO_{2eq}$ the electricity sale price is equal to the average 2007 generation price in the U.S. of \$60 per MWh, as estimated by the Energy Information Administration [76]. For non-zero GHG emission charges, we assume that the electricity sale price is equal to \$60/MWh plus the price of GHG emissions for an emissions rate equal to the U.S. grid-average rate in 2007 (636 kg$CO_2$/MWh).

## 5.7 $CO_2$ transport and storage cost model

For all CCS systems it is assumed that the $CO_2$ (available at 150 bar pressure at the plant gate) is transported 100 km from the conversion facility and stored 2 km underground in a deep saline formation. Costs were estimated using a model for $CO_2$ transmission and storage developed by Ogden [77,78]—a model that takes into account various non-linear variations in costs with scale of transport (tonnes per year) as well as distance.[15]

## 5.8 Breakeven oil price model

An important index for describing synfuel economics is the breakeven crude oil price (BEOP). This is the price of crude oil in \$ per barrel at which wholesale prices of petroleum-derived products would equal (on a \$ per GJ basis) our calculated costs for production of finished FTL fuels.

We estimate the BEOP for a given FTL production cost by subtracting from this cost the difference between the price of crude oil paid by a refiner (the crude oil acquisition cost) and the wholesale price at which the refiner sells the finished petroleum products. This difference is the refiner's margin, typically measured in cents per gallon (Figure 30).[16]

For the period 1990 through 2003, the refiner's margins (in 2007 cents per gallon) were relatively stable and the gasoline margin averaged about 10 cents a gallon more than the diesel margin. Subsequently there were sharp increases for both the gasoline and diesel margins and the diesel margin exceeded the gasoline margin. The data for 2005-2007 appear anomalous and may be a result of a confluence of factors: refinery shutdowns due to hurricane Katrina, various "refinery incidents", and the effects of onetime refinery upgrades to meet stricter low-sulfur diesel requirements and to accommodate heavier crudes.

We use the average margin for the period 1990-2003 to estimate BEOPs. The margin, as estimated by the Energy Information Administration [79], was 32.2 cents per gallon for gasoline (Figure 30). For diesel, we use a margin of 24.9 cents per gallon. This is the 1990-2003 average (from Figure 30) plus 5 cents per gallon. The additional nickel accounts for the estimated additional cost for making low sulfur diesel to comply with the sulfur standards that took effect in 2007 [80]. With these values, the average refinery margin for the diesel and gasoline displaced by FTL from the process designs we simulate (with a 61/39 diesel/gasoline output mix on a lower heating value basis) would be \$10.5 per barrel.[17]

Formally, we calculate the BEOP as follows:

BEOP (\$ per barrel) = $P \cdot 3.784 \cdot 42$, where

[15] For example, the model predicts that for a system with $CO_2$ flow rate of 10 million tonnes per year compared to one with a flow rate of 1 million tonnes per year, the cost of transmission and storage per tonne of $CO_2$ declines 59% for a 100 km transport distance and 69% for a 1000 km transport distance.
[16] The relevant data are the annual average values of the crude oil refiner acquisition cost (Table 5.21) and the refiner margins (Table 5.22) in [79].
[17] If alternatively we were to assume the average margins for 2004-2007, the average margin for the diesel/gasoline displaced would be \$23.2 per barrel.

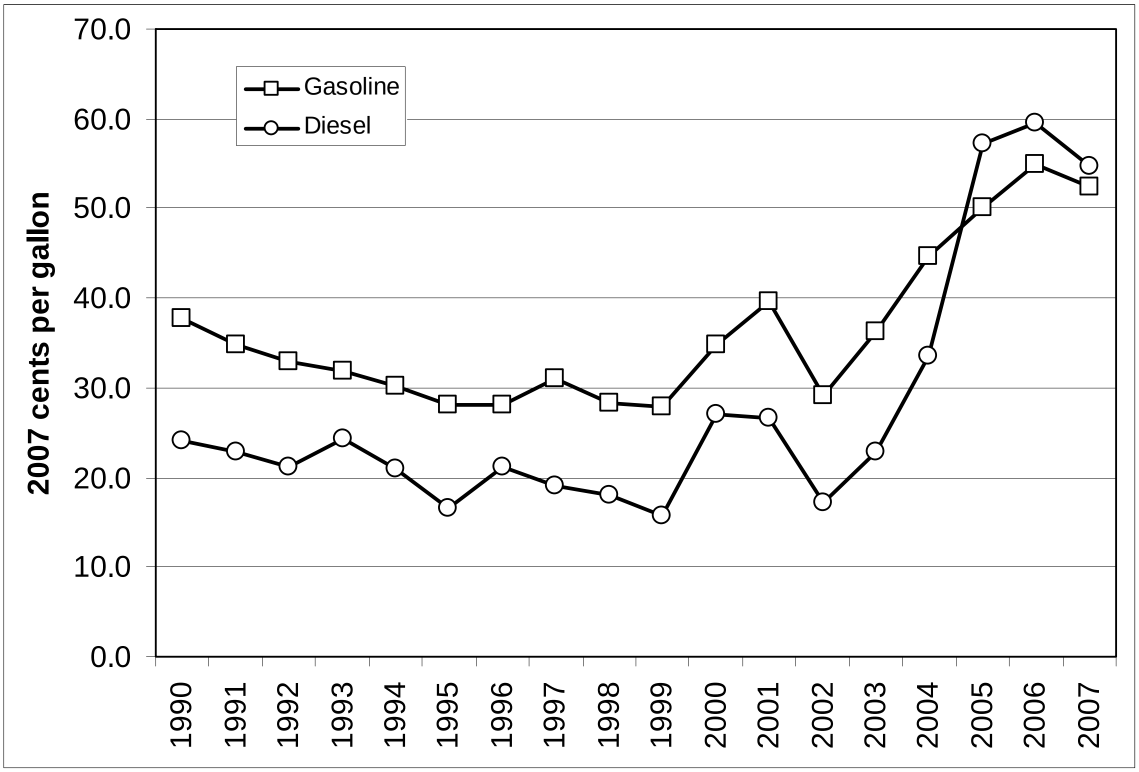


**Figure 30. Historical average annual refiner's margin for gasoline and diesel [81], adjusted to constant 2007 dollars using the GDP deflator [82].**

$P = \{C_{FTL} - f_D \cdot (RM_D / LHV_D + P_{CO2} \cdot EM_D) - f_G \cdot (RM_G / LHV_G + P_{CO2} \cdot EM_G)\} / \{f_D / LHV_D + f_G / LHV_G\}$

$C_{FTL}$ = calculated production cost of FTL in $\$/GJ_{LHV}$

f = fraction of FTL output that is diesel (D) or gasoline (G) on a LHV basis

RM = refinery margin for low-sulfur diesel (D) and reformulated gasoline (G), $ per liter.

LHV = lower heating value of diesel (D) or gasoline (G), in GJ/liter

$P_{CO2}$ = Price of $CO_2$ emissions in $/metric tonne $CO_2$

EM = Lifecycle GHG emissions of petroleum-derived low-sulfur diesel (D) or petroleum-derived reformulated gasoline (G), metric tonnes $CO_{2equiv}$ / $GJ_{LHV}$ (see Section 3.3).

# 6 Economic Results and Discussion

## 6.1 Levelized FTL Production Costs

Using the above economic analysis framework, Table 31 and Table 32 show disaggregated levelized costs of FTL fuels production for RC and OT configurations, respectively, for a GHG emissions price of \$0/t $CO_{2eq}$, so that the electricity sale price is \$60/MWh.

For the RC designs (Table 31) with venting of $CO_2$, the production costs, expressed in terms of breakeven oil prices, range from \$56/bbl for CTL, to \$93/bbl for CBTL, to \$127/bbl for BTL. With CCS added, these figures increase to \$63/bbl, \$103/bbl, and \$139/bbl, respectively.

The OT designs (Table 32), with and without CCS, offer more attractive economics than the corresponding RC designs. With venting of $CO_2$, the breakeven oil prices range from a low of \$40/bbl for the CTL-OT design to \$84/bbl for the CBTL-OT design. With CCS, breakeven oil prices are \$55/bbl to \$73/bbl for CTL designs. In the case of the CBTL-OT designs, the breakeven oil price ranges from \$59/bbl to \$107/bbl, depending on the specific design.

The cost of avoiding GHG emissions (CAE) by utilizing a –CCS rather than the corresponding –V design when the GHG emissions price is \$0/t $CO_{2eq}$ is calculated as:

CAE = {[levelized FTL cost (in \$/GJ) for CCS design] – [levelized FTL cost for V design]}

/[GHG emissions avoided (in tonnes of $CO_{2eq}$/GJ of FTL)].

The CAE is the minimum GHG emissions price needed to induce, via market forces, a shift from the –V option to the –CCS option for a pair of FTL options that differ only with regard to the treatment of the coproduct $CO_2$.

Figure 31 compares CAEs for FTL systems with CAEs for stand-alone power plants that shift from venting $CO_2$ to CCS. Most of the FTL designs have CAEs that are far lower than for power generation—a result of the fact that $CO_2$ capture is intrinsic to the design of any FTL plant, including those that vent the $CO_2$ coproduct. The lowest CAE for FTL plants (~ \$11/t$CO_{2eq}$ for the CTL-RC-CCS design) is only about 1/3 of the lowest CAE for power generation. The highest CAEs for FTL production are comparable to the lowest CAEs for power generation only when autothermal reforming and associated $CO_2$ capture equipment is included in FTL-OT process designs (i.e. FT-OTA designs). As discussed earlier, the reformer enables a larger amount of $CO_2$ to be captured than the corresponding CCS case without a reformer. The significant capital costs for the reformer lead to a relatively high cost of avoided $CO_2$ emissions.

**Table 31. Twenty-year levelized production cost for FTL fuels from RC process designs.**

| | *CTL-RC-V* | *CTL-RC-CCS* | *CBTL-RC-V* | *CBTL-RC-CCS* | *BTL-RC-V* | *BTL-RC-CCS* |
|---|---|---|---|---|---|---|
| ***Levelized FTL cost with 0 \$/tCO2equiv, \$/GJ LHV*** | | | | | | |
| Capital charges | 8.42 | 8.53 | 11.34 | 11.52 | 12.43 | 12.65 |
| O&M charges | 2.18 | 2.21 | 2.94 | 2.99 | 3.23 | 3.28 |
| Coal | 4.14 | 4.14 | 2.34 | 2.34 | 0.00 | 0.00 |
| Biomass | 0.00 | 0.00 | 5.24 | 5.24 | 11.88 | 11.88 |
| CO2 emissions charge | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CO2 disposal charges | 0.00 | 0.49 | 0.00 | 0.89 | 0.00 | 1.29 |
| Coproduct electricity (at 60 \$/MWh) | -2.26 | -1.68 | -2.57 | -1.97 | -2.07 | -1.45 |
| ***Total FTL cost, \$/GJ LHV*** | 12.48 | 13.71 | 19.30 | 21.01 | 25.47 | 27.66 |
| ***Total FTL cost, \$/gallon gasoline equivalent (\$/gge)*** | 1.5 | 1.6 | 2.3 | 2.5 | 3.1 | 3.3 |
| ***Breakeven oil price, \$/bbl*** | 56 | 63 | 93 | 103 | 127 | 139 |

**Table 32. Twenty-year levelized production cost for FTL fuels from OT process designs.**

| | CTL-OT-V | CTL-OT-CCS | CTL-OTA-CCS | CBTL-OT-V | CBTL-OT-CCS | CBTL2-OT-CCS | CBTL-OTS-CCS | CBTL-OTA-V | CBTL-OTA-CCS |
|---|---|---|---|---|---|---|---|---|---|
| ***Levelized FTL cost with 0 $/tCO2equiv, $/GJ LHV*** | | | | | | | | | |
| Capital charges | 10.37 | 10.82 | 12.01 | 14.10 | 14.68 | 10.86 | 12.86 | 12.40 | 14.79 |
| O&M charges | 2.69 | 2.81 | 3.12 | 3.66 | 3.81 | 2.82 | 3.34 | 3.22 | 3.84 |
| Coal | 5.65 | 5.65 | 5.65 | 3.60 | 3.60 | 5.20 | 4.37 | 4.38 | 4.38 |
| Biomass | 0.00 | 0.00 | 0.00 | 6.47 | 6.47 | 1.43 | 4.02 | 4.02 | 4.02 |
| CO2 emissions charge | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 | 0.00 |
| CO2 disposal charges | 0.00 | 0.67 | 0.79 | 0.00 | 1.14 | 0.68 | 0.97 | 0.00 | 1.13 |
| Coproduct electricity (at 60 $/MWh) | -9.24 | -7.76 | -5.91 | -10.29 | -9.01 | -8.04 | -8.24 | -9.88 | -6.33 |
| ***Total FTL cost, $/GJ LHV*** | 9.47 | 12.19 | 15.66 | 17.54 | 20.69 | 12.95 | 17.33 | 14.13 | 21.82 |
| ***Total FTL cost, $/gallon gasoline equivalent ($/gge)*** | 1.1 | 1.5 | 1.9 | 2.1 | 2.5 | 1.6 | 2.1 | 1.7 | 2.6 |
| ***Breakeven oil price, $/bbl*** | 40 | 55 | 73 | 84 | 101 | 59 | 82 | 65 | 107 |

The results in Table 31and Table 32 for FTL production costs (in $/GJ_{LHV}$) are extended to include the impacts of valuing GHG emissions for all 16 process cases in Figure 32, for the 6 RC cases in Figure 33, and for the 10 OT cases in Figure 34. For all these cases the value of the coproduct electricity is the grid average rate shown as a function of the GHG emissions value in Figure 35 [line labeled “Grid (avg)”].

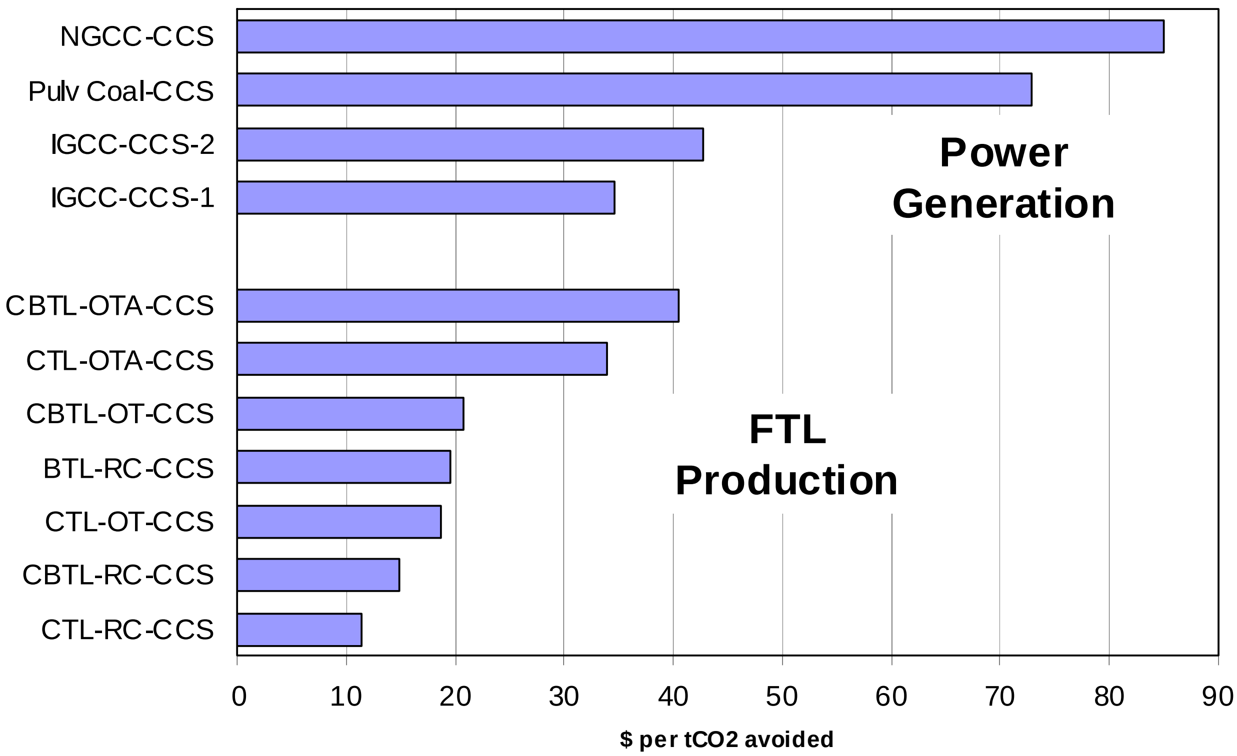


**Figure 31. Avoided cost of $CO_2$ emissions with –CCS process designs compared to corresponding –V designs.**

Section 6.2 discusses in detail, on a case-by case basis, the reasons for the differences in economic performances (FTL production costs, break-even crude oil prices, CAE – see Table 31, Table 32, and Figure 32) and technical performances ($CO_2$ capture rate, 1$^{st}$ Law system efficiencies, etc. – see Table 7, Table 8, Table 9, and Table 10) as a function of GHG emissions price among the 16 cases -- first for RC systems (Section 6.2.1), then for OT cases (Section 6.2.2). We follow with a brief overview analysis (Section 6.2.3) and end with a thought experiment exploring the implications for GHG emissions mitigation and energy security enhancement in the United States of pursuing aggressively the least costly FTL options in the presence of a serious carbon mitigation policy (Section 6.2.4).

## 6.2 Integrated Perspective on Alternative FTL Options

For systems producing both FTL and electricity the GHG emission rate associated with each coproduct is of great interest for the purpose of making comparisons among alternative energy systems that provide the same quantities of liquid fuels and electricity. Several alternative methods for assigning GHG emissions rates to coproducts have been discussed (see Section 3.3), but there is no "correct" methodology. The task of allocating important properties to separate energy coproducts must be carefully considered and reported clearly, with all assumptions transparent. However, the allocation of system emissions between FTL and electricity is arbitrary and has no impact on the economic evaluation of the system (which depends on the total GHG emission rate for the system).

The convention adopted in this Section of the paper is to assign to the power coproduct the GHG emission rate for a stand-alone coal IGCC plant with CCS[18] because of the expectations that: (*i*) *few if any new coal power plants will be built in the US without CCS, and* (*ii*) *if FTL plants were not built, new decarbonized coal power plants would be built either as alternatives to new coal power plants without CCS or as replacements for existing coal power plants.*[19]

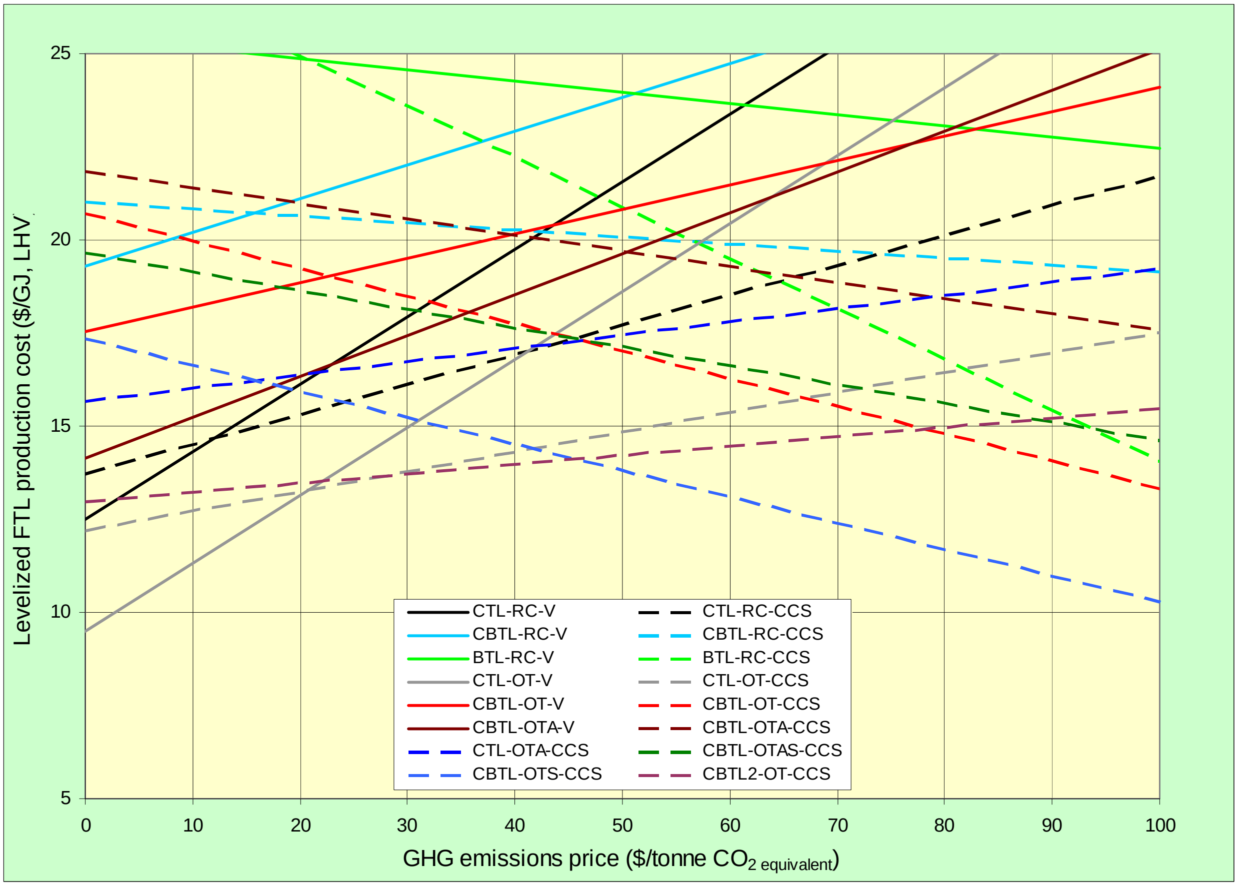


**Figure 32. Levelized FTL production cost for all 16 process designs as a function of GHG emissions price.**

[18] Under this assumption the GHG emission rate allocated to the FTL-OT coproduct electricity would be the rate for coal IGCC-CCS plants, i.e. 138 $kgCO_{2eq}$/MWh. All remaining net lifecycle GHG emissions from the FTL-OT plant would then be allocated to the FTL product.

[19] Alternatively, one could argue that the FTL-OT plant's electricity coproduct will instead "push off the dispatch stack" old, load-following pulverized coal plants whose GHG emission rate is much higher (~1000 $kgCO_{2eq}$/MWh). This assumption would lead to a much lower allocation of the FTL-OT plant's GHG emissions to the synfuels coproduct.

## 6.2.1 Recycle FTL Systems

### 6.2.1.1 CTL-RC systems

*CTL-RC-V*: The CTL-RC systems are designed to produce from bituminous coal 50,000 barrels per day of equivalent crude oil-derived diesel and gasoline. The CTL-RC-V variant generates 874 $MW_e$ of coproduct power (93% steam turbine power, 7% from unfired expanders) of which 447 $MW_e$ is for onsite needs, leaving 427 $MW_e$ net power for export. The overall 1$^{st}$ law efficiency is 50.0%. The overnight construction cost is $4.9 billion ($98,000 per B/D), the levelized plant gate cost of fuel is $1.50 per gallon of gasoline equivalent (gge), and the breakeven crude oil price (BECOP) is $56/barrel (Table 31) with GHG emissions valued at $0/t$CO_{2eq}$. However, the GHG emission rate for the FTL fuels is 2.2 times the rate for the crude oil products displaced (COPD).

*CTL-RC-CCS*: The CTL-RC-V system produces a stream of pure $CO_2$ accounting for 51.5% of the carbon in the coal (77.6% of the coal C not contained in the FTL) that can be captured and stored in appropriate underground media at low incremental cost. Some 40% of the captured $CO_2$ is from the stream of shifted synthesis gas coming directly from the coal gasifier + upstream water-gas-shift reactor and 60% is from the recycle stream downstream of the ATR (the $CO_2$ in the recycle stream is generated mainly via the water-gas-shift activity of the iron catalyst in the FTL synthesis reactor). The FTL GHG emission rate for this CTL-RC-CCS option is 1.03 times the rate for the COPD. The overall 1$^{st}$ law efficiency is 48.6%. The capital cost for the CTL-RC-CCS system is $67 million more than for the CTL-RC-V system (a 1.4% capital cost penalty), and the plant gate-cost of FTL is $1.64/gge (a 10% production cost penalty), and the BECOP is $63/barrel.

An important parameter characterizing the shift from CTL-RC-V to CTL-RC-CCS is the cost of GHG emissions avoided—the minimum GHG emissions value needed to induce CCS by a market mechanism: above this value, $11.4/t $CO_{2eq}$ (see Figure 31), the CTL-RC-CCS option produces FTL at a lower cost than the CTL-RC-V option (see Figure 33).

The economic penalty for CCS is much less for CTL-RC than for power generation. The minimum market price of GHG emissions for inducing a shift from coal IGCC-V to coal IGCC-CCS (the least costly CCS option for bituminous coal power plants) is $34.7/t $CO_{2eq}$ (see Figure 31), and the corresponding increases in capital cost ($/$kW_e$) and generation cost ($/MWh) are 32% and 40%, respectively. The actual cost of GHG emissions avoided is a higher, $44.4/t $CO_{2eq}$, the value needed to induce a shift from a PC-V plant to a coal IGCC-CCS plant, because without CCS, the PC-V option provides power at lower cost than the coal IGCC-V option (Figure 35). For this shift the capture cost is $36/t $CO_{2eq}$, the total cost of capture, $CO_2$ transport, and storage is $43.2/t $CO_{2eq}$, and the capital and generation cost penalties are 52% and 53%, respectively.

Much of the $CO_2$ capture cost for an IGCC-CCS system is associated with creating a pure stream of $CO_2$ from shifted synthesis gas so that the $CO_2$ can be readily captured. In contrast, the much smaller CCS cost for the CTL-RC-CCS system arises because a natural part of the production of FTL is the separation of a stream of pure $CO_2$ accounting for more than half of the C in the coal feedstock; the $CO_2$ capture cost is merely the cost of $CO_2$ compression to 150 bar. Thus the $67 million capital cost penalty for the shift to CCS is the cost of a 110 $MW_e$ $CO_2$ compressor ($609/$kW_e$). Likewise, the $CO_2$ capture cost, some $6.8/t, is entirely for the cost of $CO_2$ compression, most of which ($5.4/t) is for the cost of electricity for the compressor (90.5 kWh per tonne of $CO_2$—see Table 33); the $CO_2$ transport and storage cost is $4.6/tonne—equivalent to 2/3 of the cost of capture (see Table 33).

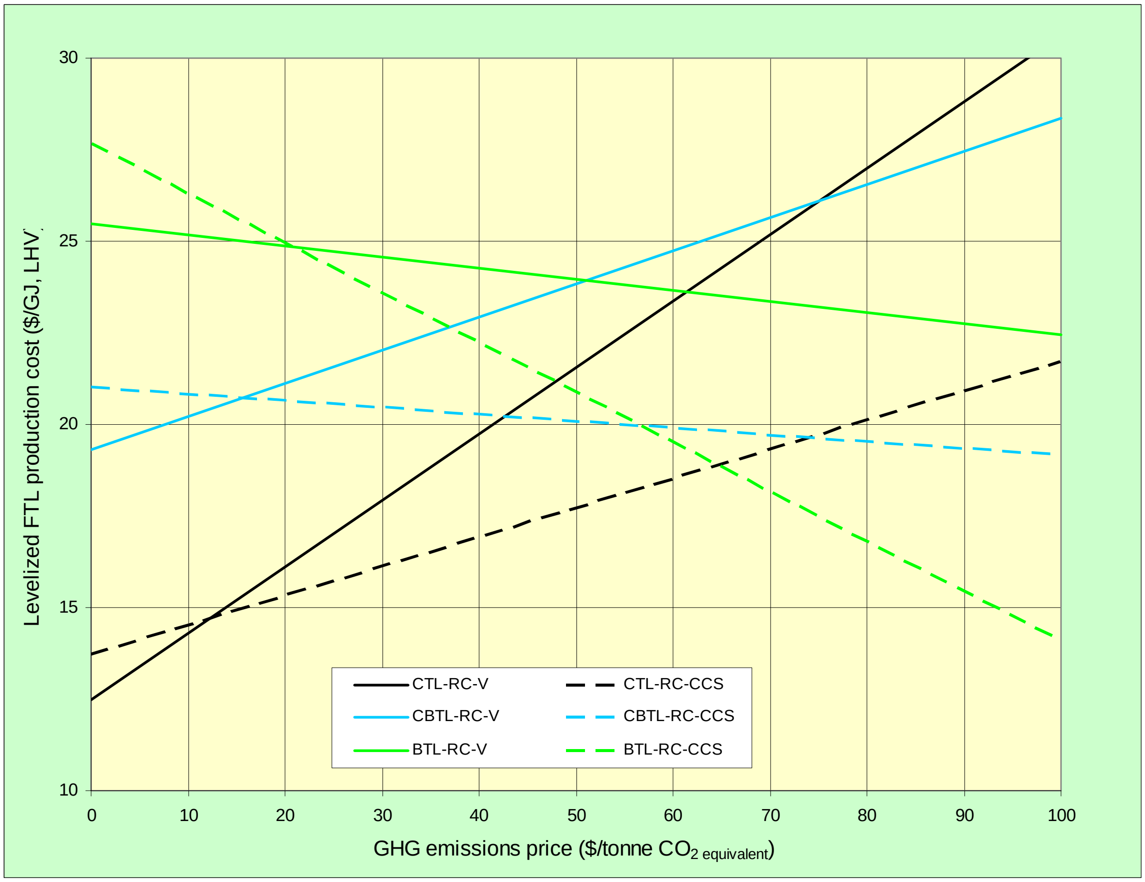


**Figure 33. Levelized FTL production cost for all RC configurations as a function of GHG emissions price.**

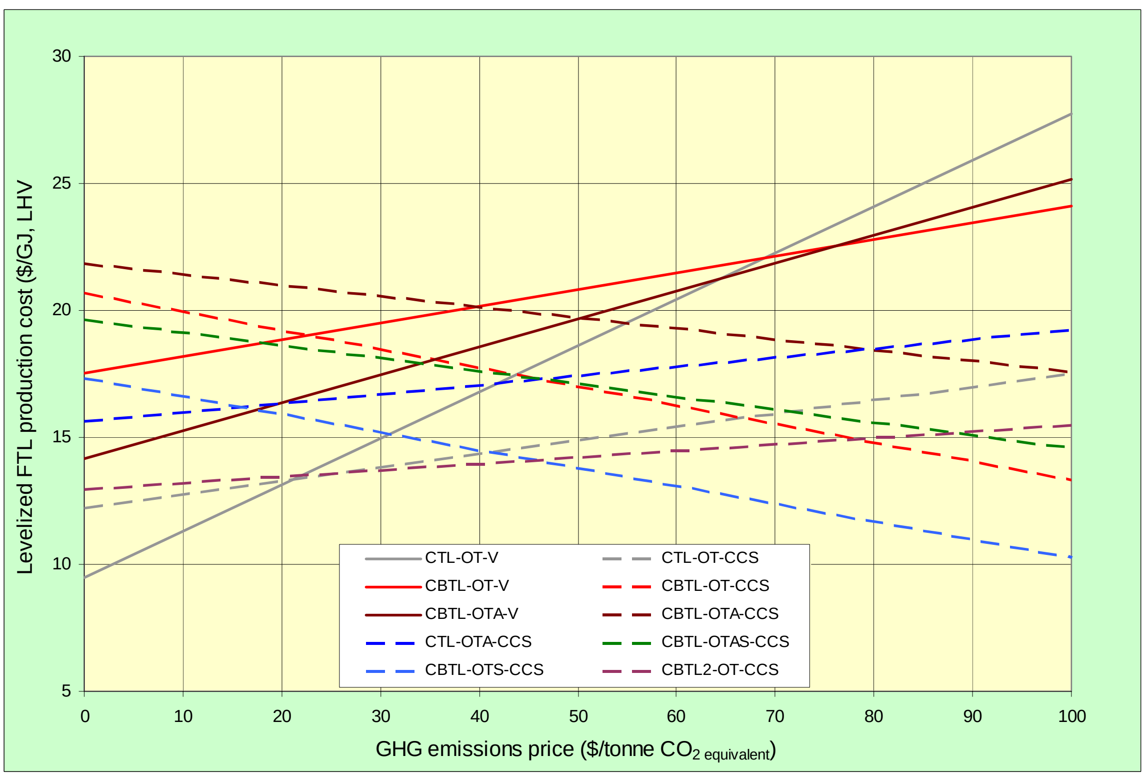


**Figure 34. Levelized FTL production cost for all OT configurations as a function of GHG emissions price.**

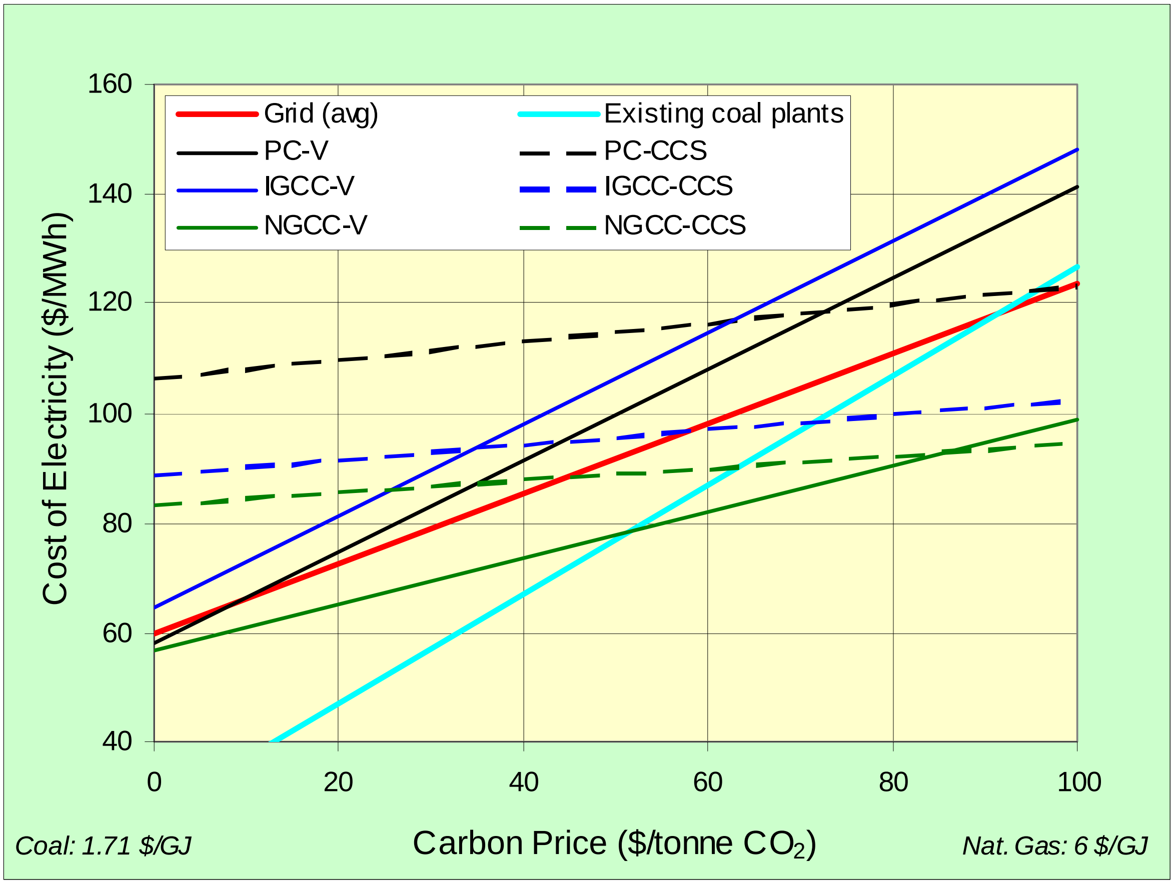


**Figure 35. Levelized power generation costs for alternative technologies as a function of GHG emissions price.**

#### 6.2.1.2 BTL-RC systems

*BTL-RC-V*: For BTL-RC-V systems coproduct power is produced at a rate of 66 $MW_e$ in a steam turbine, of which 32 $MW_e$ is for onsite needs, leaving 34 $MW_e$ available for export. The overall 1st law efficiency is 50.5%. Although the FTL GHG emission rate is a negative 0.14 times the rate for the COPD,[20] costs are relatively high. The overnight construction cost for the plant is $636 million ($144,000 per B/D), the plant gate-cost of fuel is $3.05/gge (1.86 X the cost for FTL via CTL-RC-CCS), and the BECOP is $127/barrel when no value is assigned to the GHG emissions. The high cost of FTL via the BTL-RC-V option is partly due to the assumed high biomass price (~ 3X the coal price) and partly due to the scale sensitivity of the capital cost (when the plant scale is decreased by 50%, the capital cost declines by only 40%). The BTL-RC-V option becomes more and more attractive the higher the GHG emissions value. At $107/$tCO_{2eq}$ the BTL-RC-V option becomes competitive with the CTL-RC-CCS option (Figure 33)—providing FTL at a plant-gate cost of $2.67/gge and a BECOP of $56/barrel.

[20] The GHG emission rate is slightly negative because of two factors: *(i)* about 10% of the carbon in the switchgrass leaves the FTL plant as char in the gasifier ash and is assumed to be sequestered from the atmosphere and *(ii)* the convention assumed here that the electricity coproduct GHG emission rate equals that for a coal IGCC-CCS power plant.

**Table 33. Key characteristics of alternative CTL options with CCS.**

| **CTL option →** | **RC** | **OT** | **OTA** |
|---|---|---|---|
| 1st law efficiency (HHV basis), % | 48.6 | 46.7 | 43.3 |
| Electricity penalty for CCS, kWh per tonne of $CO_2$ | 90.5 | 169.4 | 304.3 |
| GHG emission rate for FTL relative to rate for Crude oil derived products displaced (COPD) | 1.03 | 1.28 | 0.92 |
| Production cost at \$0/t $CO_{2eq}$, \$/gge | 1.64 | 1.46 | 1.88 |
| Breakeven crude oil price, \$/barrel | 62.8 | 54.5 | 73.4 |
| Cost of GHG emissions avoided, \$/t $CO_{2eq}$ | 11.4 | 18.7 | 33.9 |
| % of C not in FTL that is captured and stored | 77.6 | 67.8 | 85.1 |
| Annual $CO_2$ storage rate, $10^6$ tonnes per year | 9.60 | 9.52 | 11.95 |
| **Disaggregated CCS cost, \$ per tonne of $CO_2$ captured/stored** | | | |
| Autothermal reformer for enhanced $CO_2$ capture | | | |
| Capital | - | - | 4.93 |
| O&M | - | - | 1.28 |
| Electricity @ \$60/MWh | - | - | 2.02 |
| **Subtotal** | - | - | **8.23** |
| Reduced gross electricity generation as result of decarbonizing syngas flowing to gas turbine | | | |
| **Reduced revenues from reduced gross electricity generation** | - | **0.44** | **8.09** |
| $CO_2$ compression | | | |
| Capital | 1.08 | 0.95 | 0.95 |
| O&M | 0.28 | 0.25 | 0.25 |
| Electricity @ \$60/MWh | 5.43 | 5.43 | 5.43 |
| **Subtotal** | **6.78** | **6.64** | **6.63** |
| **$CO_2$ transport and storage** | **4.63** | **4.63** | **4.34** |
| | | | |
| $N_2$ Compressor for $NO_x$ control | | | |
| Capital | - | 0.59 | 0.32 |
| O&M | - | 0.15 | 0.08 |
| Electricity @ \$60/MWh | - | 3.58 | 2.03 |
| **Subtotal** | - | **4.33** | **2.44** |
| Additional Rectisol system | | | |
| Capital | - | 1.49 | 1.40 |
| O&M | - | 0.39 | 0.36 |
| Electricity @ \$60/MWh | - | 0.73 | 0.81 |
| **Subtotal** | - | **2.60** | **2.57** |
| Downstream WGS | | | |
| Capital | - | - | 0.24 |
| O&M | - | - | 0.06 |
| **Subtotal** | - | - | **0.30** |
| **Other** | - | **0.03** | **1.28** |
| **Total** | **11.4** | **18.7** | **33.9** |

BTL-RC-CCS: The BTL-RC-V system produces a stream of pure $CO_2$ accounting for 51.4% of the C in the input biomass (75.8% of the biomass C not contained in the FTL). If this $CO_2$ were captured and stored underground the FTL GHG emission rate for the resulting BTL-RC-CCS option would be strongly negative: –1.35 X the rate for the COPD, because storing photosynthetic $CO_2$ underground represents negative $CO_2$ emissions.[21] Even though the capital cost for the BTL-RC-CCS system would be 1.9% higher than for the BTL-RC-V system and the plant gate-cost of FTL at zero GHG emissions value would be 8.6% higher, the GHG emissions value at which this option would become competitive with CTL-RC-CCS is \$65/t $CO_{2eq}$—much

[21] All the biomass-derived $CO_2$ stored underground was originally extracted from the atmosphere during photosynthesis.

lower than in the BTL-RC-V case because of this option's strong negative GHG emission rate. At this GHG emissions price, the FTL could be provided at a plant-gate cost of $2.26/gge and a BECOP of $59/barrel. The $CO_2$ capture cost for the BTL-RC-CC system is $8.0/t $CO_2$ (only slightly higher than for CTL-RC-CCS) but the total incremental cost for $CO_2$ capture, transport to the storage site, and underground storage is a much higher $19.5/t $CO_2$ largely because the $CO_2$ transport cost per tonne of $CO_2$ is so high (small pipe effect) at the low $CO_2$ flow rate for the BTL-RC-CCS option (0.9 million tonnes per year—less than 10% of the rate for the CTL-RC-CCS option).

The cost of GHG emissions avoided (the minimum GHG emissions value needed to induce a shift from BTL-RC-V to BTL-RC-CCS) is $19.5/t $CO_{2eq}$ (see Figure 31), the same as the cost of CCS (measured in $ per tonne of $CO_2$ captured and stored).

#### 6.2.1.3 CBTL-RC Systems

The shortcomings of BTL-RC options are the high feedstock cost and the steep scale economies for both the plant capital cost and for $CO_2$ transport to the storage site. These challenges can be mitigated by coprocessing some biomass with bituminous coal (CBTL-RC options).

*CBTL-RC-CCS*: The CBTL-RC-CCS option is especially interesting because the negative emissions for the biomass component of the system can be used to offset the positive emissions for coal-derived $CO_2$ (from both the energy conversion facility and from the tailpipes of vehicles that eventually burn the fuel).

As noted earlier, our CBTL-RC-CCS plant design uses just enough biomass to reduce to zero the fuel-cycle-wide GHG emission rate for the produced FTL. Under this condition, the scale of the plant (10,000 B/D + 74.6 $MW_e$ of net electricity) is determined by the condition that the biomass delivery rate to the synfuel plant is 1 million dry tonnes per year. Biomass accounts for 43.4% of the feedstock on a HHV basis, and the overall 1$^{st}$ law efficiency is 49.4%.

The amount of zero GHG-emitting FTL produced by this system is 565 liters of gasoline equivalent per dry tonne of biomass (switchgrass)—2.1 X the projected future yield of cellulosic ethanol.[22] This "effective improved yield" is important in light of growing concerns about the "carbon debt" implications of producing biomass for biofuels on croplands [2] —which means that biomass supplies might turn out to be much less than what was thought just a few years ago (it may be necessary to avoid using cropland for growing biomass for biofuels).

The cost of GHG emissions avoided (the minimum GHG emissions value needed to induce a shift from CBTL-RC-V to CBTL-RC-CCS) is $15/t $CO_{2eq}$ (see Figure 33)—less than for the BTL-RC case because of the $CO_2$ pipeline scale economies realized as a consequence of the higher $CO_2$ transport rate (2.1 million tonnes per year).

At a GHG emissions value of $0/t $CO_{2eq}$ the CBTL-RC-CCS option produces FTL at $2.52/gge—24% less than for the BTL-RC-CCS option. However, among the CTL-RC, BTL-RC, and CBTL-RC options with CCS, CBTL-RC-CCS is never the least costly over the GHG value range of $0 to $100 per tonne of $CO_{2eq}$ (Figure 33). Breakeven with CTL-RC-CCS occurs at $74/t $CO_{2eq}$, but BTL-RC-CCS becomes less costly than CTL-RC-CCS at $65/t $CO_{2eq}$. CBTL-RC-CCS cannot make it "into the winners circle" at any GHG emissions price because its cost declines only slowly with GHG emissions price, whereas the cost for BTL-RC-CCS cost falls sharply with emissions price as a result of its strong negative GHG emission rate (Figure 33).

[22] 96 gallons of ethanol per dry short ton or 269 liters of gasoline equivalent per tonne of switchgrass, which Aden et al. [83] project to be possible by 2010.

*CBTL-RC-V*: The CBTL-RC-V option is not interesting in a C-constrained world because the FTL GHG emission rate for a plant with the same liquid fuel production capacity as CBTL-RC-CCS and also fueled with 1 million dry tonnes of biomass per year is 1.22 X the rate for the COPD.

## 6.2.2 Once Through FTL Systems

Here V and CCS variants of OT systems based on CTL, BTL, and CBTL configurations are discussed (see Figure 34).

### 6.2.2.1 CTL-OT systems

*CTL-OT-V*: The CTL-OT systems are designed to consume bituminous coal at the same rate as the CTL-RC systems. This constraint implies that CTL-OT-V produces 36,655 B/D of equivalent crude oil-derived diesel and gasoline. The plant also generates from the syngas unconverted in a single pass through the synthesis reactor a gross amount of electric power in the amount 1673 $MW_e$ via a gas turbine/steam turbine combined cycle power plant. The system requires 393 $MW_e$ of this gross output to meet onsite needs (88% of the needs for a CTL-RC-V plant with the same level of coal input), leaving net power in the amount 1279 $MW_e$ available for export (3 X the rate for a CTL-RC-V plant with the same level of coal input). The overall 1$^{st}$ law efficiency is 49.3%.

The prospective economics are impressive. The overnight construction cost is $4.4 billion (90% of that required for a CTL-RC-V plant consuming coal at the same rate), the plant gate-cost of produced fuel is $1.14/gge (compared to $1.50/gge for CTL-RC-V), and the BECOP is $40/barrel (compared to $56/barrel for CTL-RC-V).

Much of the economic performance gain in shifting from CTL-RC-V to CTL-OT-V is associated with the extraordinarily high marginal efficiency of making extra electricity -- 49.7% on a LHV basis.[23] This is far higher than the expected efficiency (39.3%) of a coal IGCC power plant without CCS, which would require 26% more coal to produce a MWh than does the CTL-OT-V plant (Figure 36).

This high marginal efficiency is a natural consequence of two factors: (*i*) the intense exothermicity of synfuel production, and (*ii*) the fact that the ratio of the FT exotherm energy to the energy in the gases used to fire the power island[24] is much higher in the RC case than the OT case. The exothermicity of synthesis allows for the production of saturated intermediate-pressure steam[25] that can be used for power generation via expansion through a steam turbine. But in the RC case there is less high-temperature heat available (from combustion of the gases that fire the power island) than in the OT case to superheat the intermediate-pressure steam raised by the FT exotherm, so steam turbine power output per unit of FT exotherm is reduced in the RC case relative to the OT case. In short, greater use of the FT exotherm for power generation is feasible

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[23] As shown in Figure 36, the marginal efficiency is calculated for CTL-RC-V and CTL-OT-V plants having the same level of FTL output as the ratio: (extra electricity produced by the CTL-OT-V plant)/(extra coal consumed by that plant).

[24] Purge gas + off-gases from the FT refinery in the RC case and unconverted syngas + off-gases from the FT refinery in the OT case.

[25] The temperature in the FT reactor is maintained nearly constant to maximize conversion. Evaporation of water is required to extract heat from the FT reactor at a sufficiently rapid rate to achieve isothermal conditions. Thus, heat recovery from an FT reactor is accomplished by boiling water. Superheating of the resulting steam is avoided inside the FT reactor to achieve heat transfer rates that are as high as possible.

in the OT case compared to the RC case (the ratio of $MW_e$ of steam turbine output to MW of FTL output is 1.5 X as large for the OT case as for the RC case).[26]

The output of the steam turbine bottoming cycle in this system is 1.22 X the output of the gas turbine. This steam turbine/gas turbine output ratio is far higher than this ratio for a natural gas combined cycle plant (0.5) or for a coal IGCC plant (0.6) because the steam for the steam turbine is produced via heat recovery not only from the hot combustion product exhaust gases of the gas turbine but also from other parts of the system—most notably heat recovered from the highly exothermic synthesis process.

Despite the impressive economics of the CTL-OT-V option, its carbon footprint is extraordinarily high—the FTL GHG emission rate is 2.8 X the rate for the COPD. However, this emission rate can be reduced substantially at relatively low incremental cost, as discussed below.

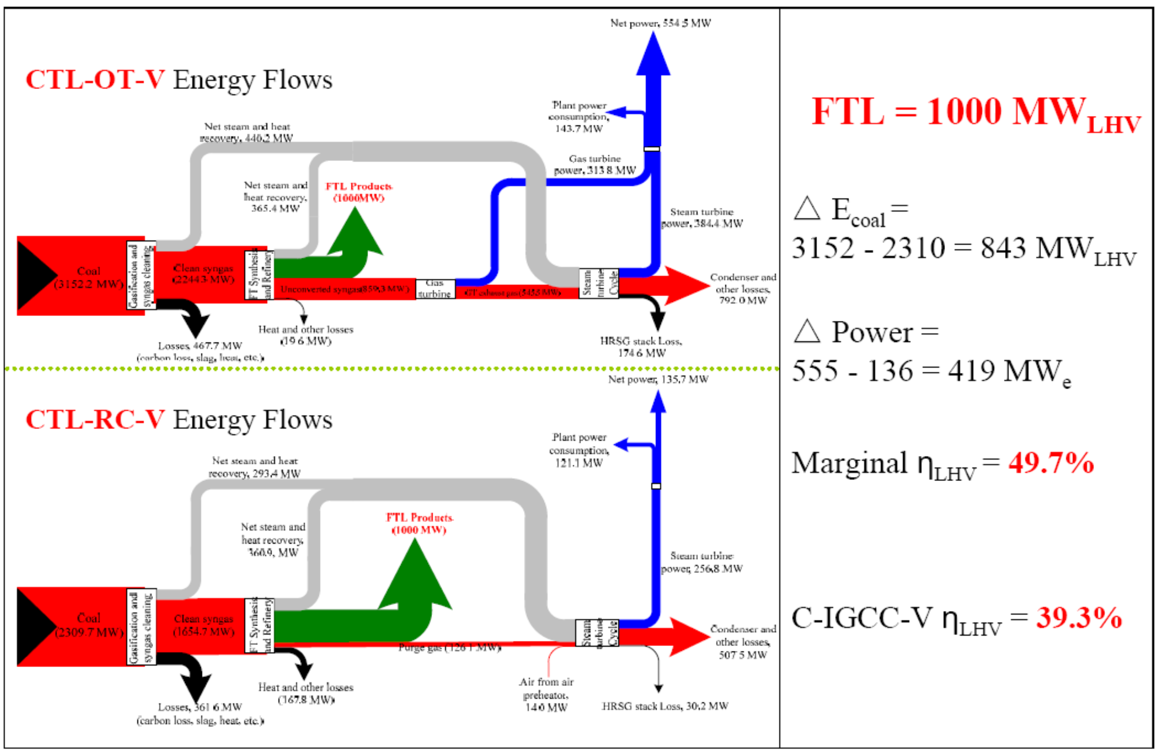


**Figure 36. Energy flows illustrating high marginal electricity efficiency of the CTL-OT-V system.**

*CTL-OT-CCS*: In pursuing CCS for CTL-OT options, $CO_2$ can be captured both upstream of synthesis (from shifted synthesis gas before it enters the synthesis reactor) and downstream of synthesis (extra $CO_2$ is generated via the water gas shift feature of the iron FTL catalyst)—to the extent of 51.1% of the C in the coal (67.8% of the coal C not contained in the FTL). The energy penalty for CCS is manifest as a 16% reduction in the net export power compared to the CTL-OT-V case (to 1075 $MW_e$) and a reduction in the 1st law efficiency from 49.3% to 46.7% (HHV basis).

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[26] This is true regardless of whether only a steam turbine (ST) cycle is used in the RC option (which we and others have chosen) or whether a gas turbine topping cycle is used in a configuration similar to what we have used in our OT cases. (A gas turbine topping cycle would add some efficiency in electricity generation to the power cycle at an RC facility, but the OT configuration would still be advantageous from the standpoint of the steam turbine output.)

As in the CTL-OT-V case, CTL-OT-CCS offers an extraordinarily high marginal efficiency of making extra electricity compared to CTL-RC-CCS: 43.3% on a LHV basis compared to 33.2% for a coal IGCC-CCS plant, which requires 30% more coal to produce a MWh than does the CTL-OT-CCS plant (Figure 37).

But $CO_2$ capture is more energy intensive and more costly for CTL-OT-CCS than for CTL-RC-CCS. The total electricity required for $CO_2$ capture is 169 kWh per tonne of $CO_2$ compared to 91 kWh per tonne for CTL-RC-CCS (see Table 33). The electricity for $CO_2$ compression is the same in both cases but accounts for only 53% of the electricity penalty in the CTL-OT-CCS case. Three factors account for the higher electricity and cost penalties. First, the extra Rectisol® equipment for $CO_2$ capture and solvent regeneration needed for downstream capture and the energy needed to run this equipment must be charged to the GHG emissions mitigation account (rather than the FTL production account), because, absent a carbon policy constraint, this equipment is not needed. However, this cost penalty is relatively modest [equivalent to 39% of the $CO_2$ compression cost (see Table 33)], because the concentration of $CO_2$ in the downstream syngas is much higher than in the upstream syngas (43% vs 17%). Second, a relatively high penalty (65% of the $CO_2$ compression cost—see Table 33) arises because, in shifting from CTL-OT-V to CTL-OT-CCS, there is a relative large penalty associated with adding a $N_2$ compressor that takes $N_2$ from the ASU unit and delivers it to the gas turbine combustor for $NO_x$ control. (No $N_2$ compressor is needed for the CTL-OT-V case because the flame temperature in the gas turbine combustor is sufficiently low to meet $NO_x$ emissions requirements as a result of the high level of $CO_2$ in the syngas flowing to the gas turbine combustor.) Third, removal of $CO_2$ from the syngas flowing to the gas turbine reduces the mass flow to the gas turbine somewhat and thus gross power output from the power island; the effect is relatively modest (accounting for less than 7% of the $CO_2$ compression cost), because most of the mass flow loss from $CO_2$ removal is compensated by the increased mass flow of $N_2$ to the gas turbine required for $NO_x$ emissions control.

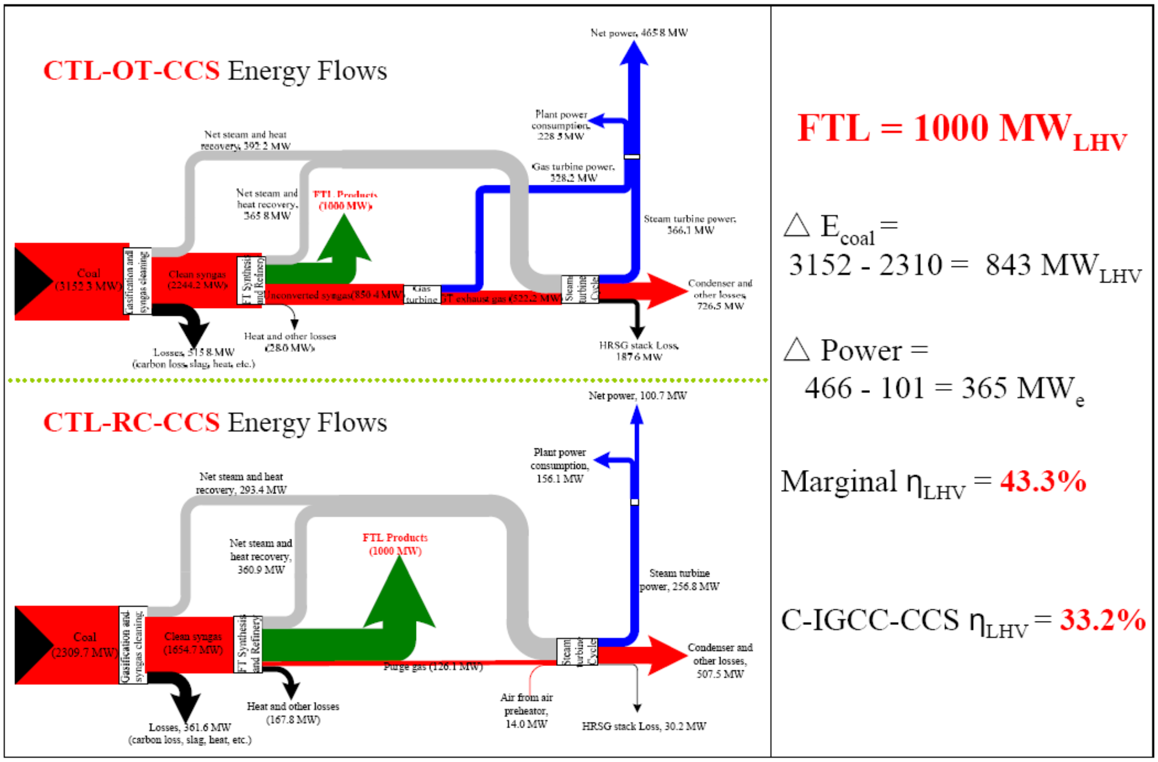


**Figure 37. Energy flows illustrating high marginal electricity efficiency of the CTL-OT-CCS system.**

$CO_2$ transport and storage costs are the same for the CTL-RC-CCS and CTL-OT-CCS because the $CO_2$ storage rates are the same.

The overall CCS cost (at \$18.7/t $CO_2$ of $CO_2$ captured and stored—see Table 33) is 64% higher for than for CTL-RC-CCS but is still much less than for any stand-alone coal power system.

Despite the significantly higher CCS cost for CTL-OT-CCS compared to CTL-RC-CCS, the overall economics are far more favorable for CTL-OT-CCS compared to CTL-RC-CCS: the capital cost for a CTL-OT-CCS system is 93% of the capital cost for a CTL-RC-CCS system; the plant gate-cost of fuel is \$1.46/gge (89% of that for CTL-RC-CCS); and the BECOP is \$55/barrel (compared to \$63/barrel for CTL-RC-CCS). The much better economics are a manifestation of the better thermodynamic performance of OT options compared to RC options.

But the FTL GHG emission rate for CTL-OT-CCS is ~ 1.3 X the rate for the COPD—a rate that is not likely to be acceptable under a serious C-mitigation policy. The high GHG emissions rate for CTL-OT-CCS stems from the fact that the fraction of the coal C not contained in the FTL products that is captured is 68% compared to 78% for the CTL-RC-CCS options.

Even though the FTL GHG emission rate is much higher than that for CTL-RC-CCS, the production cost gap between CTL-RC-CCS and CTL-OT-CCS increases with GHG emissions value because the coproduct credit for the large decarbonized electricity coproduct increases rapidly; as a result there is no GHG emissions value at which CTL-RC-CCS outperforms CTL-OT-CCS (Figure 32).

<u>CTL-OTA-CCS</u>: By introducing an autothermal reformer (ATR) and associated additional $CO_2$ capture equipment downstream of synthesis, the capture fraction in the CTL-OT-CCS case can be increased from 68% to 85% of the coal C that is not contained in FTL products for a CTL system with the same FTL output capacity as the CTL-OT systems described. With this design, the FTL GHG emissions rate would be reduced to 88% of the rate for the COPD.

The ATR converts light ($C_1$- $C_4$) gases to CO and $H_2$. A subsequent water gas shift (WGS) system converts the CO in the syngas to $CO_2$ and $H_2$, and extra $CO_2$ capture equipment is introduced downstream of the WGS reactors. Because all this equipment and the associated electricity penalties have to be charged to the GHG mitigation account, the shift from CTL-OT-CCS to CTL-OTA-CCS entails a substantial energy penalty (the total capture penalty for CTL-OTA-CCS is 304 kWh per tonne of $CO_2$ compared to 169 kWh per tonne of $CO_2$ for CTL-OT-CCS—see Table 33), manifest as a 24% reduction in the net export power compared to the CTL-OT-CCS case, to 818 $MW_e$ and a reduction in the 1$^{st}$ law efficiency from 46.7% to 43.3%.

Moreover, the overall CCS cost is relatively high: \$33.9/t$CO_2$ (81% higher than for CTL-OT-CCS—see Table 29). Most of the increased cost of $CO_2$ capture relative to the CTL-OT-CCS case is accounted for by the addition of the autothermal reformer and associated ASU components (\$8.2/t) and by the loss of revenues (\$8.1/t) from reduced gross power production (see Table 33). In contrast to the shift from CTL-OT-V to CTL-OT-CCS, for which the gross power output was reduced only modestly (from 1673 $MW_e$ to 1664 $MW_e$), the shift from CTL-OT-CCS to CTL-OTA-CCS entails a dramatic loss of 195 $MW_e$ of gross power output. In the CTL-OT-CCS case, the loss of the $CO_2$ mass flow to the gas turbine as a result of $CO_2$ removal was largely compensated for by the addition of $N_2$ mass flow for $NO_x$ control, so that the gross power production fell only modestly. But in the CTL-OTA-CCS case, the $C_1$-$C_4$ gases in the syngas are reformed and shifted and the resulting $CO_2$ is removed, so that the syngas flowing into the gas turbine is mostly $H_2$, and the $N_2$ mass flow added for $NO_x$ control falls far short of compensating for the reduced mass flow associated with decarbonizing the syngas.

The capital cost for CTL-OTA-CCS is 11% higher than for CTL-OT-CCS, the production cost at $0/t $CO_{2eq}$ is $1.88/gge (28% more than for CTL-OT-CCS), and the BECOP is $73/barrel (compared to $55/barrel). Although the cost gap between CTL-OTA-CCS and CTL-OT-CCS narrows as the GHG emissions value increases, the gap still remains substantial even at $100/t $CO_{2eq}$ (see Figure 34).

#### 6.2.2.2 CBTL-OT Systems

*CBTL2-OT-CCS*: This system has the same FTL output capacity as CTL-OT-CCS and CTL-OTA-CCS but part of the feedstock is 1 million dry tonnes of switchgrass per year. This system, providing 36,655 B/D of FTL plus 1113 $MW_e$ of net electricity (compared to 818 $MW_e$ for CTL-OTA-CCS), is fired with 8.6% biomass on a HHV basis and has an FTL GHG emission rate equal to the rate for the COPD (0.81 X the rate for CTL-OT-CCS). The 1$^{st}$ law efficiency is 46.9% (compared to 43.3% for CTL-OTA-CCS). The capital cost is 0.4% higher than for CTL-OT-CCS (compared to 11% higher in the CTL-OTA-CCS case); the production cost at $0/t $CO_{2eq}$ is $1.55/gge (compared to $1.88/gge for CTL-OTA-CCS), which is only 6% higher than for CTL-OT-CCS. The BECOP is $59/barrel compared to $73/barrel for CTL-OTA-CCS.

Moreover, as shown in Figure 32 and Figure 34, CBTL2-OT-CCS becomes the least costly of the 16 FTL options analyzed over the entire GHG emissions price range of $27 to $46 per tonne of $CO_{2eq}$. At $27/t $CO_{2eq}$ the production cost is $1.63/gge and the BECOP is $49/barrel.

CBTL2-OT-CCS is an outstanding candidate option for getting early experience with CCS for power while increasing energy security and providing a synfuel that is no worse than imported oil in terms of GHG emissions. The reason is that the decarbonized coproduct electricity selling price at $27/t $CO_{2eq}$ is $77/MWh—which is only 29% higher than the US average generation cost in 2007 and only 83% and 96%, respectively, of the generation costs for a coal IGCC-CCS plant and a coal PC-V plant at this GHG emissions value.

Also, CBTL2-OT-CCS plants would be outstanding candidates for replacing existing $CO_2$-emissions-intensive coal power with decarbonized power, because these plants would be able to drive out of the market these existing coal power plants in economic dispatch competition if enough CBTL2-OT-CCS plants were built.

The order of plants dispatched on the electric grid is determined competitively by the dispatcher who accepts bids to provide power to the grid. Once built (at which time the capital investment for a power plant becomes a sunk cost) power will be sold into the grid as long as revenues exceed short-run marginal costs. From a power perspective, CBTL2-OT-CCS plants can be considered “must-run” baseload power plants (like nuclear power plants in this respect). Having FTL as a major co-product enables these plants to compete in economic dispatch for power down to very low crude oil prices.

Consider first the situation when the GHG emissions price is zero. In this instance, the average short run marginal cost of existing coal power plants in the US averages $27/MWh (see Figure 38). If there were CBTL2-OT-CCS plants selling power into the grid these polygeneration plants would be able to compete in economic dispatch with these coal plants all the way down to crude oil prices less than $25/barrel (see Figure 38).

A more likely scenario is that CBTL2-OT-CCS plants would not be built without the expectation that the GHG emissions price would average at least $27/t $CO_{2eq}$ over the economic life of the plant. But at this GHG emissions price, CBTO2-OT-CCS plants would be even more competitive in economic dispatch with existing coal power plants—even for oil prices as low as $10/barrel (see Figure 38).

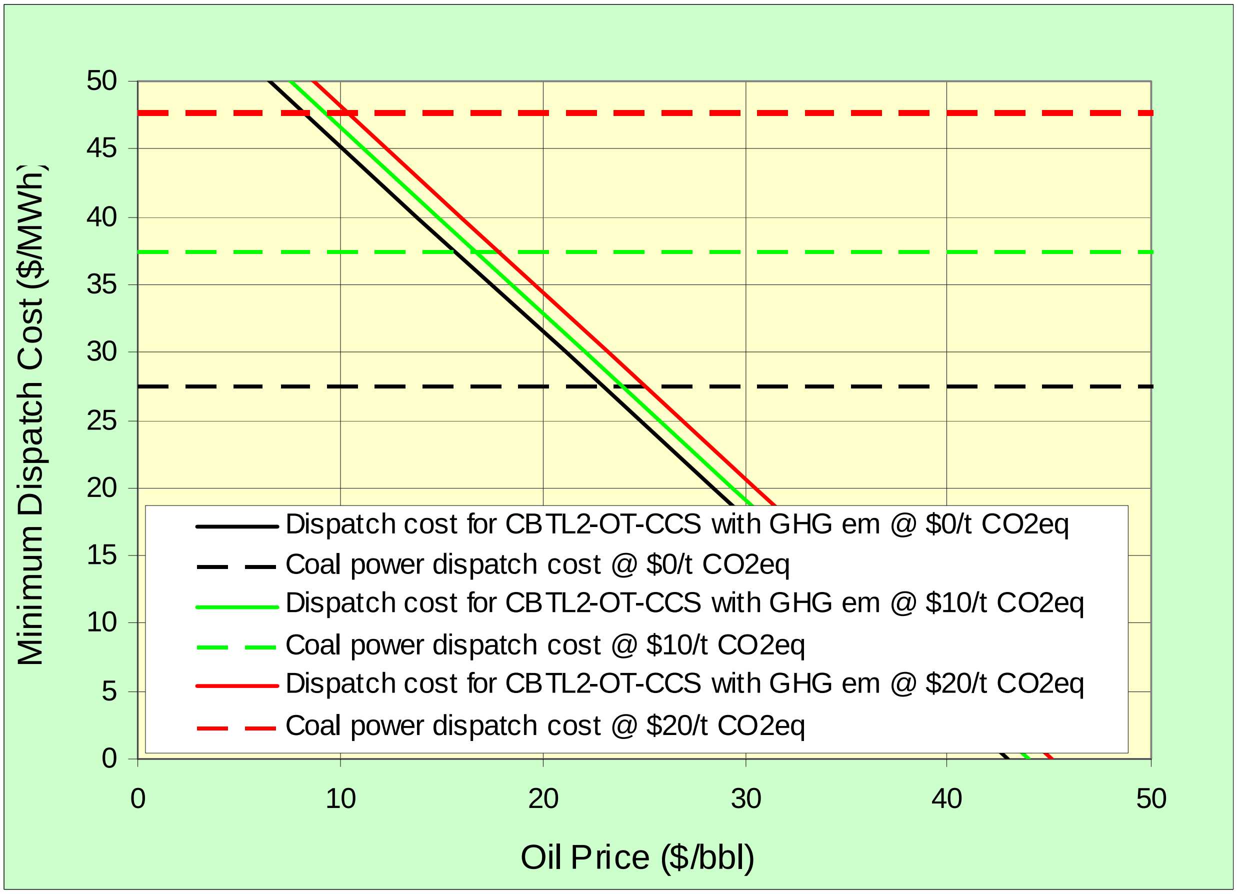


**Figure 38. Dispatch cost analysis.**

*CBTL-OT-CCS*: This system is designed in the same spirit as CBTL-RC-CCS—using just enough biomass in the form of switchgrass to reduce to zero the fuel-cycle-wide GHG emission rate for the produced FTL. This condition and the assumed biomass delivery rate of 1 million dry tonnes per year determine the scale of the plant (8,100 B/D of FTL + 276 $MW_e$ of net electricity). For this plant biomass accounts for 38.1% of the feedstock on a HHV basis (compared to 43.4% in the CBTL-RC-CCS case), and the overall 1st law efficiency is 47.6%.

In contrast to the relatively high costs estimated for CBTL-RC-CCS over all GHG emission values, CBTL-OT-CCS becomes the least costly of all 16 FTL options investigated other than CBTL-OTS-CCS at a GHG emissions value of $78/t $CO_{2eq}$ (see Figure 32 and Figure 34), at which the production cost is $1.79/gge and the BECOP is $30/barrel. As in the case of CBTL-RC-CCS, the CBTL-OT-CCS option offers a zero net GHG-liquid fuel at a much higher biomass yield than any conventional biofuels—specifically, 458 liters of gasoline equivalent per dry tonne of biomass (switchgrass)—1.7 X the projected future yield of cellulosic ethanol.

*CBTL-OTS-CCS*: This CBTL-OTS-CCS (S = soil and root carbon storage) option is like CBTL-OT-CCS except that it involves using mixed prairie grasses (MPGs) instead of switchgrass as the biomass feedstock. Again, this system is designed to use just enough MPGs to reduce to zero the fuel-cycle-wide GHG emission rate for the produced FTL. This condition and the assumed biomass delivery rate of 1 million dry tonnes per year determine the scale of the plant (13,040 B/D of FTL + 406 $MW_e$ of net electricity). For this system the MPGs account for 23.9% of the feedstock on a HHV basis. This percentage is much less than for the switchgrass

(CBTL-OT-CCS) option because of the supplemental carbon storage (beyond the storage realized in the form of supercritical $CO_2$) in soils and roots.

The CBTL-OTS-CCS option offers a zero net GHG-emitting liquid fuel at a much higher biomass yield than the CBTL-OT-CCS option—736 liters of gasoline equivalent per dry tonne of biomass (switchgrass)—2.7 X the projected future yield of cellulosic ethanol.

At GHG emissions values in excess of $46/t $CO_{2eq}$ CBTL-OTS-CCS becomes the least costly of all the 16 FTL systems modeled (see Figure 32), providing FTL at $1.69/gge with a BECOP of $42 a barrel. Across the entire range of GHG emissions values the FTL price is ~ $0.4/gge less than for the CBTL-OT-CCS option (Figure 34). The lower FTL cost reflects both the scale economy of the larger plant and the lower average feedstock cost for CBTL-OTS-CCS compared to CBTL-OT-CCS.

MPGs are likely to be a small fraction of the total biomass supply, a large fraction of which is likely to not offer soil/root C buildup opportunities. Thus CBTL-OTS-CCS and CBTL-OT-CCS should be considered together as complementary zero net-GHG-emitting FTL options.

<u>CBTL-OT-CCS options with autothermal reformers</u>: In light of the scarcity of biomass resources it is worthwhile considering adding autothermal reformers and associated additional $CO_2$ capture equipment downstream of synthesis to increase the percentage of feedstock C not contained in the FTL products that is captured as $CO_2$—from the ~ 67% level for CBTL-OT-CCS and CBTL-OTS-CCS to ~ 85%, thereby making it feasible to realize a zero net GHG emission rate for FTL with a lower percentage of biomass in the feedstock. Keeping the biomass input rate fixed at 1 million dry tonnes per year would also enable larger plants to be built, so that scale economies as well as lower average feedstock cost help offset the increased costs associated with these OTA-CCS options—as discussed above for CTL-OTA-CCS systems. The CBTL-OTA-CCS system so constructed with zero net FTL GHG emissions would have an output of 13,058 B/D of FTL and a biomass input percentage of 23.9% (HHV basis) compared to 8,100 B/D and 38.1% for CBTL-OT-CCS. Similarly, the CBTL-OTAS-CCS system would have an output of 19,107 B/D of FTL and a biomass input percentage of 16.4% compared to 13,039 B/D and 23.9% for CBTL-OTS-CCS.

But the scale economy and lower average feedstock cost benefits turn out to be considerably less than the added costs for the autothermal reformer and associated capture equipment. As in the CTL-OTA-CCS case, these OTA-CCS options never become the least costly FTL options over the entire GHG emissions price range of $0 to $100 per tonne of $CO_{2eq}$ (see Figure 32 and Figure 34).

### 6.2.3 Overview Issues

<u>Focusing on the least costly options</u>: The comprehensive information presented in the prior sections for the 16 FTL options examined in this study may be difficult to grasp easily. However, in the final analysis what is of greatest interest is the set of least costly options without and with a carbon mitigation policy in place.

If attention is restricted to the least costly options for GHG emissions prices in the range $0 to $100 per tonne of $CO_{2eq}$, only the five options (all OT options) shown in Figure 39 warrant close scrutiny. Summarizing:

CTL-OT-V is the clear winner in terms of cost among the 16 FTL options up to a GHG emissions value of $21/t because of the low cost of coal and the powerful thermodynamic advantage of OT options compared to RC options; this option is unattractive though from a carbon policy perspective because the FTL GHG emission rate is 2.8 X that for the COPD.

In the narrow window of GHG emissions prices of \$21/t and \$27/t, CTL-OT-CCS, with an FTL GHG rate that is 1.3 X that for the COPD, is the least costly of the 16 options.

Between \$27/t and \$46/t, CBTL2-OT-CCS is the least costly option. This option offers an early route to CCS for power while enhancing energy security by providing FTL from secure domestic sources at a GHG emission rate that is no greater than for the COPD. At \$27/t$CO_{2eq}$, CBTL2-OT-CCS offers FTL at \$1.6/gge and decarbonized power at \$77/MWh. It would be able to replace power from old coal plants with decarbonized power because it would be highly competitive in economic dispatch (Figure 38).

At a GHG emissions price of \$46/t, CBTL-OTS-CCS fueled with 23.9% biomass in the form of MPGs becomes the least costly option—offering FTL with a zero net GHG emission rate at \$1.7/gge while simultaneously providing decarbonized power at \$89/MWh. A zero FTL GHG emission rate is realized with such a modest biomass input rate because the negative GHG emissions associated with biomass-based CCS are enhanced with the buildup of soil and root carbon in the growing of MPGs on C-depleted soils.

Because MPG supplies that offer soil/root C storage benefits are limited, alternative biomass-based FTL options will also play substantial roles. Excluding CBTL-OTS-CCS results from those shown in Figure 39, CBTL-OT-CCS fueled with 38.1% biomass becomes the least costly option at \$78/t—offering FTL with a zero net GHG emission rate at \$1.8/gge while simultaneously decarbonizing power at \$110/MWh.

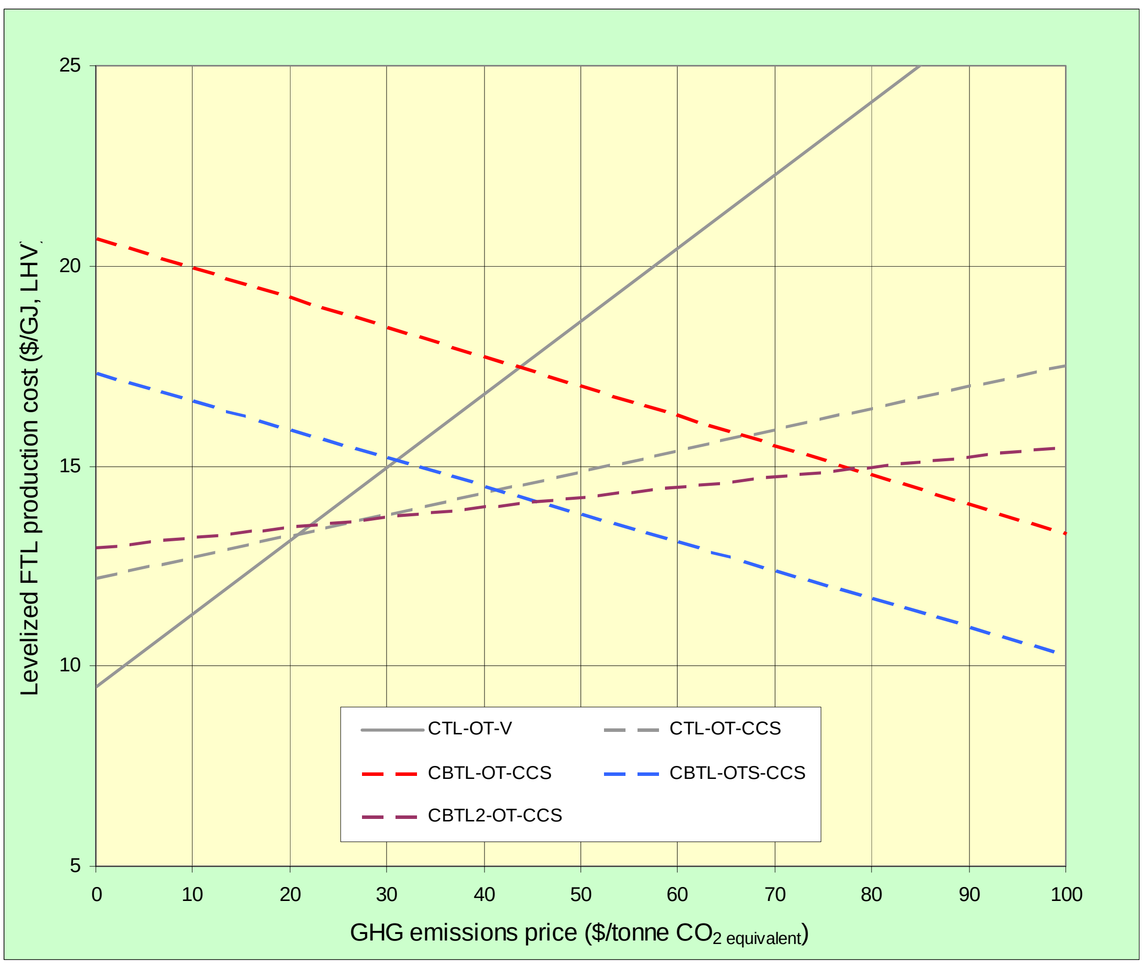


**Figure 39. Levelized FTL production cost for the least costly configuration at each GHG emissions price.**

Protecting synfuel investments against oil price collapse: The present analysis indicates that FTL technologies can be competitive for crude oil prices ~ $60 a barrel or less. But synfuel plants represent multi-billion dollar investments, and prospective investors worry about oil price collapse—since the marginal cost of incremental oil production in the Middle East is generally believed to be much less than $60 a barrel.

Accordingly, many prospective investors in the United States are asking the government to guarantee a floor price on oil to protect these investments. But keeping consumer oil prices high in the event of a collapse in world oil prices may prove to be unpopular.

In the presence of a strong carbon policy and emphasis in synfuel deployment on low-GHG-emitting supply options, there would be no need for a guaranteed floor price on oil. For example, in the presence of an $80/t $CO_{2eq}$ GHG emissions price, the CBTL-OTS-CCS, CBTL-OT-CCS, and CBTL2-OT-CCS options would all be competitive at crude oil prices of $30 a barrel or less and offer FTL at plant-gate costs in the range of $1.4 to $1.8 per gge;[27] in contrast, the CTL-RC-CCS option would not be able to compete if crude oil prices fell much below ~$60 per barrel (Figure 40).

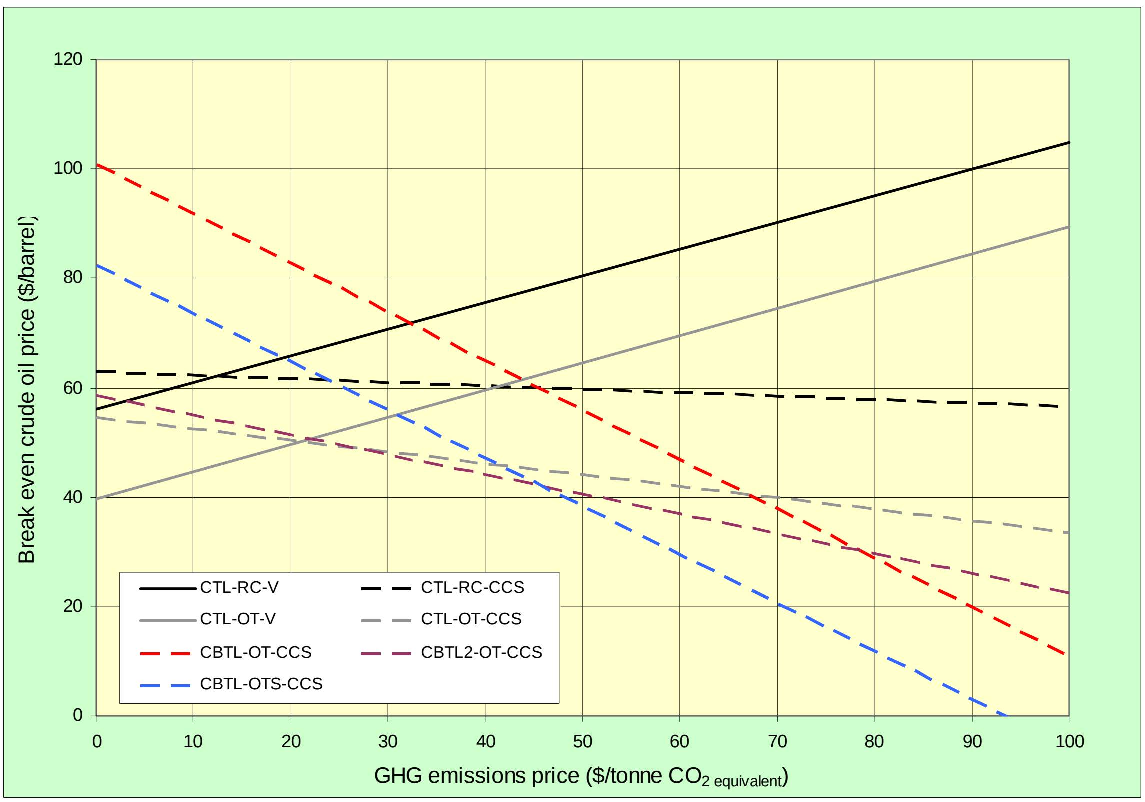


**Figure 40. Breakeven crude oil prices as a function of GHG emissions price.**

Implications of sustained high oil prices: Recent oil prices have been high because demand has been growing faster than supply for liquid fuels. These prices have been far in excess of the marginal cost of new supplies and thus should eventually come down to the marginal cost level – although no one knows when that will be. In recent weeks oil prices have been falling but are still far above marginal costs.

[27] The corresponding "pump prices" seen by the consumer (including distribution and retail taxes) at which these FTL systems could compete would be in the range $2.1 to $2.5 per gge.

A major finding of the present analysis is that OT options will trump RC options in making FTL because they offer much lower production costs. However, if oil prices were to rise once again to very high levels, RC options would tend to be favored over OT options: profits from providing FTL would be so high that project developers would want to squeeze as much FTL out of a tonne of feedstock as possible. That this is so can be gleaned from an internal rate of return (IRR) analysis assuming a high oil price. Figure 41 shows the internal rates of return for the eight most profitable FTL options for a world where the crude oil price is sustained at $150 per barrel.[28] Over almost all of the GHG emissions price range of $0/t to $100/t of $CO_{2eq}$ the most profitable options at this oil price are CTL-RC-V and CTL-RC-CCS. Moreover, a decarbonized FTL option (CBTL-OTS-CCS) does not become the most profitable option until the GHG emissions price reaches ~ $80/t of $CO_{2eq}$ – ~1.8 times the GHG emissions value needed to make CBTL-OTS-CCS the least costly option in a world where the oil price is determined by the marginal production cost (see Figure 32 and Figure 41). And the opportunity to decarbonize a large amount of coproduct power does not become viable until the GHG emissions price reaches ~$70/t, when the CBTL2-OT-CCS option becomes as profitable as the CTL-RC-CCS option.

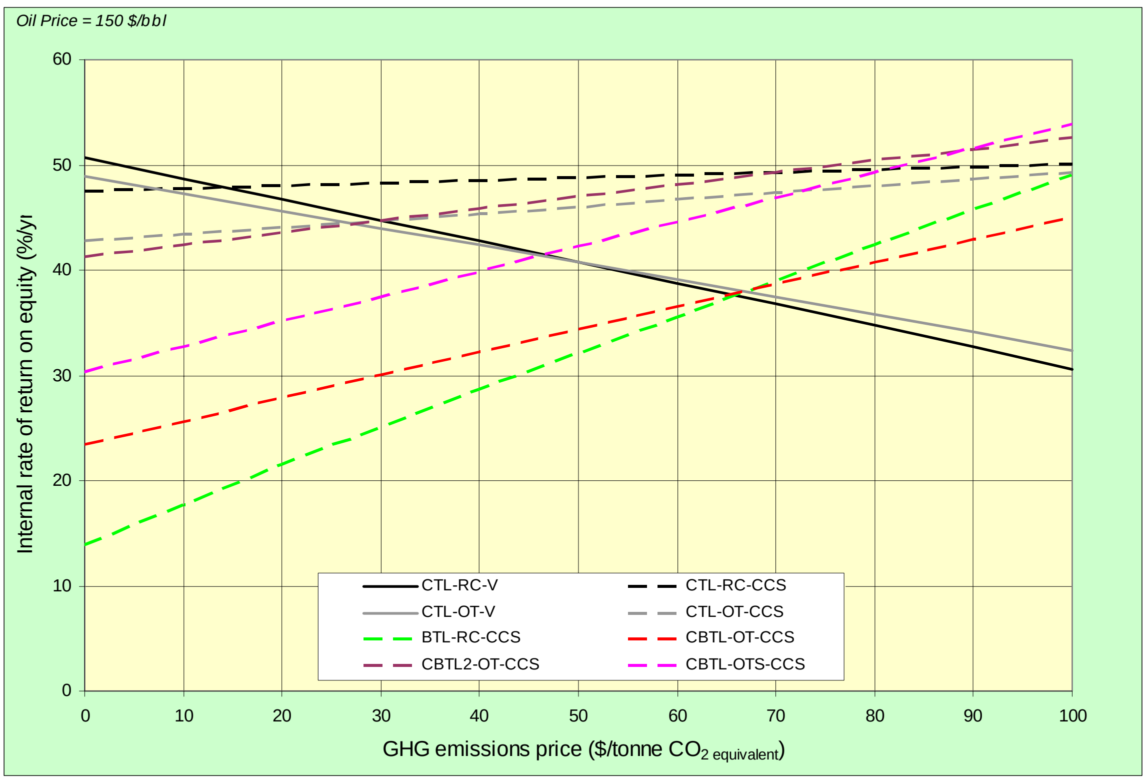


**Figure 41. Internal rates of return as a function of GHG emissions price for the most lucrative FTL production options with oil at $150/bbl.[28]**

This IRR exercise suggests that a promising way to get underway a transition to low carbon liquid fuels at modest GHG emission price levels would be to build up rapidly FTL supplies via CBTL-OT-CCS and CBTL-OTS-CCS systems to levels that would be effective in helping to

[28] Here the internal rates of return (IRR) are real (inflation-corrected) after-tax rates of return on equity for plants that are financed with the same 55/45 debt/equity ratio, the same real debt cost, the same (owner cost)/TPC ratio, the same tax rules, etc., as for the levelized production cost analysis presented in Section 6.1 and parameter values in Table 30—but the real cost of equity is allowed to vary. The IRR is calculated for the assumed crude oil price and at each GHG emissions value by varying the real cost of equity until the levelized FTL production cost equals the cost of displaced electricity and crude oil products.

bring down oil prices to marginal cost levels—a strategy that is explored in the following subsection.

### 6.2.4 A Thought Experiment

There is an emerging consensus that deep reductions in GHG emissions are needed in this half-century in order to avoid dangerous anthropogenic interference with climate from fossil fuel burning. One indicator that such a consensus is emerging is that one of the major pieces of carbon mitigation legislation under consideration in the US Congress, the Lieberman-Warner Climate Security Act of 2008, calls for reducing by 2050 US GHG emissions 70% relative to 2005 levels.

Achieving such a daunting goal will require strong emphasis on reducing emissions from coal power plants and from transportation, which accounted for, respectively, 30% and 33% of US GHG emissions from fossil fuel burning in 2007. Under a serious carbon mitigation policy rapid deployment of CBTL-OT-CCS and CBTL-OTS-CCS polygeneration technologies would be a cost-effective approach to realizing deep reductions in GHG emissions for both of these sectors. For example, at a GHG emissions price ~$80/t of $CO_{2eq}$ these options would offer the least costly ways to provide FTL among the 16 options examined here while decarbonizing both FTL and the coproduct power. Also, rapid deployment of such technologies so as to realize FTL production levels that could have a significant impact in helping bring down the world oil price is key to decarbonizing liquid fuels at moderate GHG emissions price levels (see Section 6.2.3).

Here a thought experiment is presented that would involve deployment by 2035 of 500 such polygeneration plants in the United States—starting with 3 plants on line by 2015 and exponential growth thereafter at an average rate of 29% per year (see Figure 42). Because each plant would consume 1 million tonnes per year, aggregate consumption would reach 0.5 billion tonnes per year by 2035.

Biomass is a scarce resource, and supplies are likely to be less than prospective quantities envisioned several years ago [46] in light of growing concerns about food price impacts [1] and the indirect land use impacts of growing biomass for energy on croplands [2]. However, there are reasonable prospects that US cellulosic biomass supplies that don't require use of cropland will not be less than 0.5 billion tonnes per year.[29]

The only dedicated energy crops assumed for the thought experiment are MPGs grown without inputs after establishment on a degraded land area equivalent to the current amount of land (~ 14 million hectares) in the Conservation Reserve Program (CRP) at an assumed MPG yield of 4.6 dry tonnes per hectare per year [75]. This annual supply of 64 million tonnes of MPGs would be utilized in CBTL-OTS-CCS plants. It is assumed that the remainder of the 0.5 billion tonnes per year of biomass is in the form of crop residues and forest residues that are consumed by CBTL-OT-CCS plants.[30] Thus 12.8% of the plants would be CBTL-OTS-CCS plants and 87.2% would be CBTL-OT-CCS plants, and the outputs of the "average" plant would be 8,700 B/D of FTL + 292 $MW_e$.

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[29] The "billion ton study" [46] estimated potential crop and forest residues to be 425 and 368 million short tons per year, respectively. Converting the latter woody biomass to herbaceous biomass equivalent and both to tonnes leads to a potential US residue supply of 753 million tonnes of herbaceous-biomass-equivalent per year. In addition, some biomass energy crops can be grown on CRP lands and other degraded lands that are not suitable for growing food crops.

[30] Although switchgrass is the feedstock modeled for the CBTL-OT-CCS plant designs, the overall carbon and energy balances are not likely to change much if crop residues (e.g., corn stover) or woody biomass (e.g., forest residues) are used instead. For the thought experiment the simplifying assumption is made that the carbon and energy balances are the same as for switchgrass.

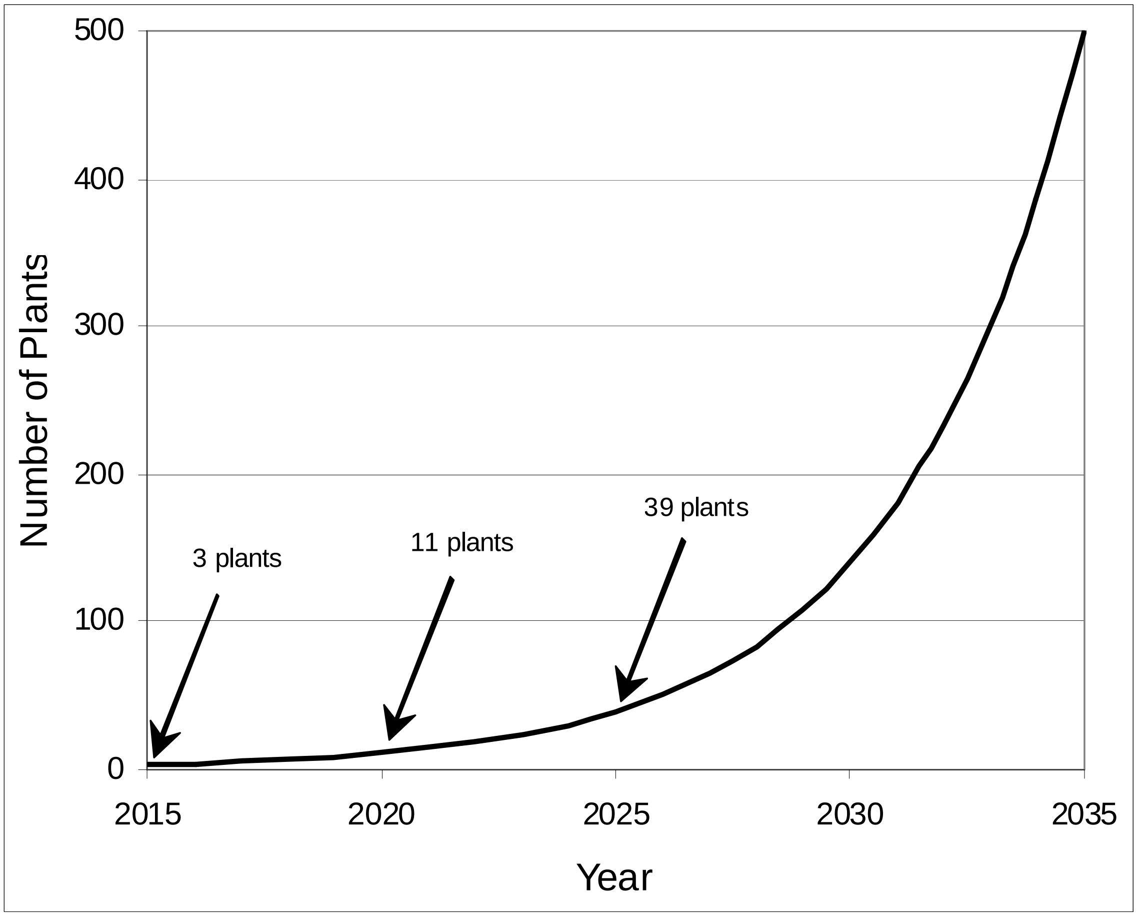


**Figure 42. Thought experiment for deployment of coal/biomass to liquids with CCS.**

The GHG emissions avoided are assumed to be those avoided when: (*i*) the zero GHG-emitting FTL products displace crude-oil-derived products, and (*ii*) the decarbonized electricity coproduct displaces electricity from existing coal power plants. The later assumption is based on: (*i*) the premise of a comprehensive carbon mitigation policy being in place that leads to flat electricity demand during the period out to 2035,[31] so that decarbonized power added during this period would back out existing coal power plants, and (*ii*) the observation that these plants would be highly competitive with existing coal plants in economic dispatch competition (see, for example, Figure 38).

In the thought experiment, productive capacity in 2035 for zero net GHG-emitting FTL would be the energy-equivalent of 4 million barrels per day of crude oil imports, and FTL production would amount to 65 billion gallons of gasoline equivalent per year (equivalent to 46% of US gasoline consumption in 2007). The decarbonized power capacity in 2035 would be 146 $GW_e$, and decarbonized power generation would amount to 1150 TWh per year (equivalent to 58% of coal power generation in 2007). By 2035 the annual rate of $CO_2$ storage would amount to 1.2 Gt $CO_2$ per year, and total GHG emissions avoided via deployment of these

[31] McKinsey & Company [84] projects essentially flat electricity demand through 2030 for its mid-range GHG mitigation scenario for the US—an outcome arising largely from investments in energy efficiency improvement for both new and existing residential and commercial building, which are likely to prominent among the “low-hanging fruit” opportunities that would be exploited under a comprehensive GHG mitigation policy.

technologies would amount to 1.7 Gt $CO_{2eq}$ per year—equivalent to 25% of total US GHG emissions from fossil fuel burning in 2007.

Carbon and energy balances for the thought experiment are summarized in Figure 43. Notably, total coal use for these plants would be only 35% more than the coal consumption rate for the existing coal power plants[32] displaced by these polygeneration plants. If only the coal consumption in excess of that consumed by the existing coal plants is charged to FTL production, the result is that the incremental coal required is just 27% of the coal that would be required if the same amount of FTL were to be provided by CTL-RC-CCS plants (see Figure 43).[33] Moreover, the amount of biomass required to produce this amount of FTL is only 54% as much as the amount of biomass required to produce the energy-equivalent amount of ethanol via future cellulosic ethanol technology[34] (see Figure 43).

The deployment rate envisioned for the thought experiment is very aggressive—but the climate change and energy security risks we face under business-as-usual conditions are formidable and require aggressive action for their mitigation. Moreover, the sustained growth rate of 29% per year during 2015-2035 is not unprecedented. During its heyday, 1957-1977, nuclear power grew worldwide at an average rate of 37% per year (from 0.81 GWe to 98 $GW_e$), and wind power grew worldwide some 28% per year during 1995-2007 (from 4.8 $GW_e$ to 94 $GW_e$). Of course, in both of these historical examples the growth was propelled by promotional public policies. The growth rate envisioned for low GHG-emitting polygeneration plants in the thought experiment could not be realized without an appropriate supporting public policy.

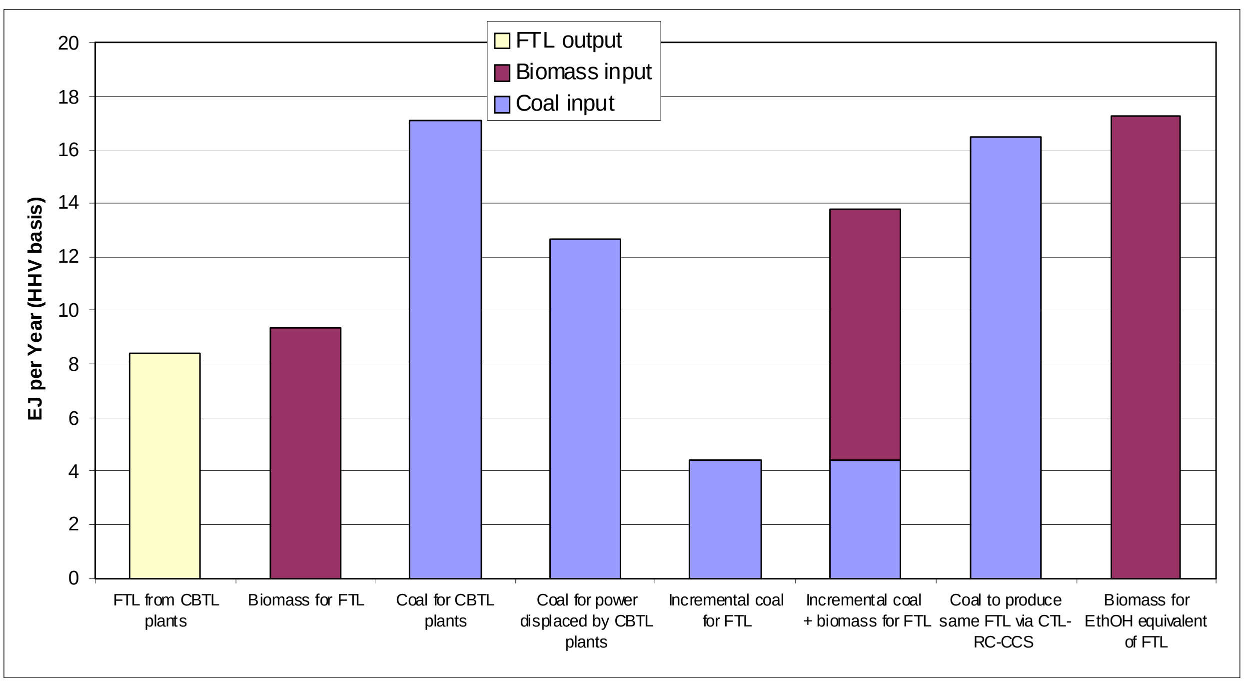


**Figure 43. Thought experiment summary.**

[32] For which the average HHV efficiency in 2007 was 32.8%.

[33] Here the HHV efficiency of making CTL-RC-CCS is assumed to be the "effective efficiency" $\eta_{eff}$ = (FTL output)*100/[total coal input – (power output)/$\eta_{dp}$] = 50.9% (rather than the total First Law efficiency of 48.6% discussed earlier), where $\eta_{dp}$ = the HHV efficiency of a stand-alone coal IGCC-CCS power plant (32.5%).

[34] Here the HHV efficiency of future ethanol production is assumed to be 48.7% (assumed 96 gallons per short dry ton of biomass).

# 7 Summary and Conclusions

In this paper we have developed a detailed, well-grounded, self-consistent framework for process design/simulation and for economic evaluation of alternative FTL systems. And we have applied this framework to describe and compare in quantitative detail 16 alternative CTL, BTL, and CBTL system configurations that could be deployed based on commercial or near-commercial technologies with regard energy balances, GHG-emission characteristics, and economics. In the application of the analytical framework we have given special emphasis to understanding designs that co-process coal and biomass (CBTL) because of the promise that such designs hold for producing at competitive costs under a carbon mitigation policy transportation fuels with no net lifecycle GHG emissions, while doing so using considerably less biomass input per unit of transportation fuel than most other lignocellulosic biomass-based liquid fuels that have been proposed, e.g., cellulosic ethanol.

Some key features of our analytical framework include:

- Simulation assumptions for various unit operations (gasification, water gas shift, acid gas removal, FTL synthesis, gas turbine, etc. – see Table 5) that are, for most operations, consistent with performance that has been demonstrated in existing commercial applications. In a few cases (e.g., biomass gasifier), performance has been demonstrated only at pilot-plant scale, and we assume that such performance can also be realized at commercial scale.
- Greenhouse gas emissions estimates that include net lifecycle emissions for synfuel production and consumption, including emissions associated with activities upstream of the conversion plant, such as coal mining and biomass growing, harvesting, and transportation, as well as emissions downstream of the plant, including transport of the liquid products to refueling stations and combustion of the fuels in vehicles.
- Capital cost estimates in 2007 $ for U.S. construction that take into account the escalations in construction cost that have been observed for several years as a result of increased costs for construction materials, labor, and engineering services.
- Capital cost estimates that assume "$N^{th}$" plant construction, adopting plausible equipment training philosophies and levels of contingencies and indirect costs that are intended to be characteristic of mature process industries.
- Simulation of carbon policies in our financial analysis by assigning a range of different valuations to the net lifecycle GHG emissions associated our process designs.

In applying this analytical framework to a comparison of the 16 FTL configurations that we model, our most important findings were that:

- The $CO_2$ capture costs for the naturally concentrated $CO_2$ streams available in FTL plants are far less than for any stand-alone coal power plant.
- To the extent that oil prices are determined by marginal production costs, OT (once through) options will generally be more cost-effective than RC (recycle) options—a finding that arises because of the high marginal efficiencies of electricity production at OT plants.
- A lower percentage of feedstock carbon not contained in the FTL products is available for capture as $CO_2$ in OT systems than in RC systems…which implies that the FTL GHG emission rate for CTL-OT-CCS is higher than for the corresponding CTL-RC-CCS system and higher also than the rate for the COPD (crude oil products displaced).

- The percentage of feedstock carbon not contained in the FTL products that is available for capture as $CO_2$ can be increased markedly for OT systems by introducing downstream of synthesis an autothermal reformer and associated $CO_2$ capture equipment, thereby substantially reducing the FTL GHG emission rate for such systems (e.g., CTL-OTA-CCS) to a level that is less than for CTL-RC-CCS systems. But FTL costs are considerably higher for FTL-OTA-CCS systems than for FTL-OT-CCS systems and would not be among the least costly FTL options for GHG emission prices up to $100 per tonne of $CO_{2eq}$.
- A more cost-effective way to reduce further the FTL GHG emission rates for FTL-OT-CCS systems would be to coprocess biomass with coal. Such plants cofired with only ~10% biomass could provide decarbonized power and FTL with a GHG emission rate s no greater than that for the COPD at much lower GHG emissions prices than the prices needed to induce CCS for stand-alone coal power plants—making such plants outstanding candidate options for early CCS action. The low cost of GHG emissions avoided makes such synfuel projects especially attractive as $CO_2$ sources for projects to demonstrate the viability of large-scale CCS.
- FTL-OT-CCS plants co-fired with biomass are outstanding candidates for replacing coal power from existing power plants with decarbonized power because these synfuel plants would be highly competitive in economic dispatch down to very low crude oil prices.
- Among the FTL-OT-CCS options that cofire with coal enough biomass to realize a zero net GHG emission rate for FTL, the CBTL-OTS-CCS option stands out. This option uses mixed prairie grasses (MPGs) grown on degraded lands with carbon-depleted soils so that taking into account the buildup of soil and root carbon enables such plants to realize zero net GHG emissions for FTL when fired with only 24% biomass on a HHV basis. These plants would be able to provide both zero GHG-emitting FTL and decarbonized coproduct power at the least cost among all the 16 FTL systems investigated when the GHG emissions price reaches $40/t $CO_{2eq}$. Although such MPG supplies are likely to end up being a relatively modest fraction of total biomass supplies, MPG-fired CBTL-OTS-CCS plants could plausibly make up a large percentage of the early plants that are built offering full decarbonization of energy products.
- At a GHG emissions price ~ $80/t $CO_{2eq}$, FTL-OT-CCS plants fueled with 24% and 38% biomass (CBTL-OTS-CCS and CBTL-OT-CCS plants, respectively) could provide FTL with zero net GHG emissions along with decarbonized coproduct power at lower production costs than for any other of the 16 FTL options analyzed, in quantities sufficient with prospective biomass supplies to displace nearly half of current gasoline supplies and nearly 3/5 of existing coal power.
- Having a carbon policy in place characterized by a GHG emissions price ~ $80/t $CO_{2eq}$ would give CBTL-OT-CCS plant investors powerful insurance against the financial risks of oil price collapse that will threaten investments in capital-intensive synfuel plants—obviating the need for the politically risky alternative of a government guarantee of a floor price on oil.
- Zero GHG-emitting CBTL-OT-CCS plants would require as inputs ~ ½ as much biomass as cellulosic ethanol plants that would produce the same amount of liquid fuel on an energy basis, thereby stretching considerably the amount of scarce biomass resources available for mitigating simultaneously oil insecurity and climate change risks.

- Zero GHG-emitting CBTL-OT-CCS plants would require as inputs ~¼ as much coal as CTL-RC-CCS plants and corresponding less coal mining activity.

Concluding observation: The present analysis suggests that in a world with both a serious carbon policy and a strong synfuels industry in place, an economically attractive way to produce electricity from coal would be via "polygenerating" FTL-OT-CCS plants.

But evolving an industry to produce synfuels and electricity in such a manner would be institutionally challenging. The power industry as presently constituted has indicated no interest in getting involved with synfuels production. And, although the multinational oil companies are beginning to show interest in synfuels in light of their limited global opportunities for new investments in oil exploration and production, they have historically shown little interest in investing in electricity production.

An appropriate new public policy might be needed to overcome the institutional hurdles. In an earlier era, the Public Utility Regulatory Policies Act of 1978 (PURPA) mandated that electric utilities purchase electricity from qualifying industrial cogenerators of heat and power at prices equal to the costs they would avoid by not having to generate this power themselves—a law, the passage of which, created a booming cogeneration industry in the United States. Consideration might be given to figuring out what the appropriate new public policy should be for facilitating the development of a new polygeneration industry suitable for a carbon-constrained and secure-oil-supply-constrained world.

One might be optimistic that such a new public policy will eventually be forthcoming—in light of the substantial economic benefit that could be exploited by deploying OT instead of RC systems for the manufacture of synfuels, and the historical record that in the United States proposed new public policies for overcoming targeted institutional obstacles that have promise in improving economic performance have had relatively good prospects for enactment.

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