Green-House-Gas-Reduced Coal-and-Biomass-to-Liquid-Based Jet Fuel (GHGR-CBTL) Process

FINAL TECHNICAL REPORT

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ABSTRACT

This Final Technical Report describes the work and accomplishments of the project entitled, “Green-House-Gas-Reduced Coal-and-Biomass-to-Liquid-Based Jet Fuel (GHGR-CBTL) Process”. The main objective of the project was to raise the Technology Readiness Level (TRL) of the GHGR-CBTL fuel-production technology from TRL 4 to TRL 5 by producing a drop-in synthetic Jet Propellant 8 (JP-8) with a greenhouse-gas footprint less than or equal to petroleum-based JP-8 by utilizing mixtures of coal and biomass as the feedstock. The system utilizes the patented Altex fuel-production technology, which incorporates advanced catalysts developed by Pennsylvania State University. While the system was not fabricated and tested, major efforts were expended to design the 1-TPD and a full-scale plant. The system was designed, a Block-Flow Diagram (BFD), a Process-Flow Diagram (PFD), and Piping-and-Instrumentation Diagrams (P&IDs) were produced, a Bill of Materials (BOM) and associated spec sheets were produced, commercially available components were selected and procured, custom components were designed and fabricated, catalysts were developed and screened for performance, and permitting activities were conducted. Optimization tests for JP-8 production using C_2 olefin as the feed were performed over a range of temperatures, pressures and WHSVs. Liquid yields of between 63 to 65% with 65% JP-8 fraction (41-42% JP-8 yield) at 50 psig were achieved. Life-Cycle Analysis (LCA) was performed by Argonne National Laboratory (ANL), and a GHGR-CBTL module was added to the Greenhouse gases, Regulated Emissions, and Energy use in Transportation (GREET®) model. Based upon the experimental results, the plant design was reconfigured for zero natural-gas imports and minimal electricity imports. The LCA analysis of the reconfigured process utilizing the GREET model showed that if the char from the process was utilized to produce combined heat and power (CHP) then a feed containing 23 wt% biomass and 77 wt% lignite would be needed for parity with petroleum-based JP-8. If the char is not utilized for CHP, but sequestered in a land fill, 24 wt% biomass and 76 wt% lignite would be required. A TEA was performed on this configuration following DOE guidelines and using the ANL-developed GREET module that showed that the GHGR-CBTL TOC and ECO are 69% and 58% of those for the DOE FT-Liquids Baseline Case, respectively. This analysis shows that the economics of the GHGR-CBTL process are significantly better than a gasification/FT process. No technical barriers were identified. The lower costs and the detailed design that was performed under this project are being used by Altex to attract funding partners to move the GHGR-CBTL development forward.

TABLE OF CONTENTS

LIST OF TABLES

LIST OF FIGURES

EXECUTIVE SUMMARY

This Final Technical Report describes the work and accomplishments of the project entitled, “Green-House-Gas-Reduced Coal-and-Biomass-to-Liquid-Based Jet Fuel (GHGR-CBTL) Process”. The main objective of the project was to raise the Technology Readiness Level (TRL) of the GHGR-CBTL fuel-production technology from TRL 4 to TRL 5 by producing a drop-in synthetic Jet Propellant 8 (JP-8) with a lower greenhouse-gas footprint than petroleum-based JP-8 by utilizing mixtures of coal and biomass as the feedstock. The system utilizes the patented Altex fuel-production technology, which incorporates advanced catalysts developed by Pennsylvania State University. The system was designed, a Block-Flow Diagram (BFD), a Process-Flow Diagram (PFD), and Piping-and-Instrumentation Diagrams (P&IDs) were produced, a Bill of Materials (BOM) and associated spec sheets were produced, commercially available components were selected and procured, custom components were designed and fabricated, catalysts were developed and screened for performance, and permitting activities were conducted. Although cost overruns on equipment and labor resulted in the system not being completely fabricated, enough data from the system-design activities and catalyst testing were available to conduct a Life-Cycle Analysis (LCA) of the Well-to-Wake (WTWa) greenhouse-gas emissions and to develop a Techno-Economic Analysis (TEA) of a full-scale GHGR-CBTL plant for comparison with competing technologies.

GHGR-CBTL is a transformative technology that overcomes the technical and economic barriers that exist in converting coal and biomass to a drop in fuel. The GHGR-CBTL removes the oxygen in its low temperature upstream pyrolysis process, but in contrast to the traditional pyrolysis-bio-oil approaches that require hydrotreating, it does not cool the condensable. But like the proven FT, the process uses gaseous intermediates to build the synthetic fuel, with the required properties to produce the desired synthetic fuel. These attributes increase the GHGR productivity and reduce the capital and operating costs when compared to classical gasification-FT or pyrolysis-bio-oil-hydrotreating.

Under the project a 1 BPD test system and a 40,000 BPD plan were designed. The test system was designed in detail including PI&D, BOM and equipment were selected. Some of the equipment were ordered and received. However, due to higher than planned cost of the equipment and plant engineering the system was not built and was not tested. However, the JP-8 conversion catalyst was identified and was fully tested at laboratory scale and its operating conditions were mapped.

Optimization tests for JP-8 production using C_2 olefin as the feed were performed over a range of temperatures (240 to 450°C), pressures (0 to 180 psig), and WHSVs (1 to 10 hr^{-1}). These tests were conducted with different catalysts; Catalyst P1, PSU-developed catalysts, and commercial catalysts. Detailed test results of C_2-oligomerization optimization are presented Figure ES-1.

As shown in Figure ES-1, both Catalyst P1 and PSU-developed catalysts give between 41 and 42% JP-8 yield with 65% JP-8 fraction and 63 to 65% liquid yield at 50 psig with C_2-olefins. Also, not shown in this figure, both these catalysts perform at 0 psig with lower JP-8 yield up to 5%. In comparison, the commercial catalyst gave similar JP-8 yield of 42% at a higher pressure of 115 psig and had zero liquid yield at 0 psig. Olefin conversion was >95% with Catalyst P1 and PSU-developed catalysts, while it was 83% for the commercial catalyst. The data were used design the oligomerization-reactor design for the TRL-5 system. Availability of the Altex-selected oligomerization catalysts for the TRL-5 system, and JP-8 optimized conditions of low

pressures (0 to 180 psig) versus existing industrial oligomerization at >500 psig makes the Altex-oligomerization for the TRL-5 system both low-risk for JP-8 production and energy-efficient.

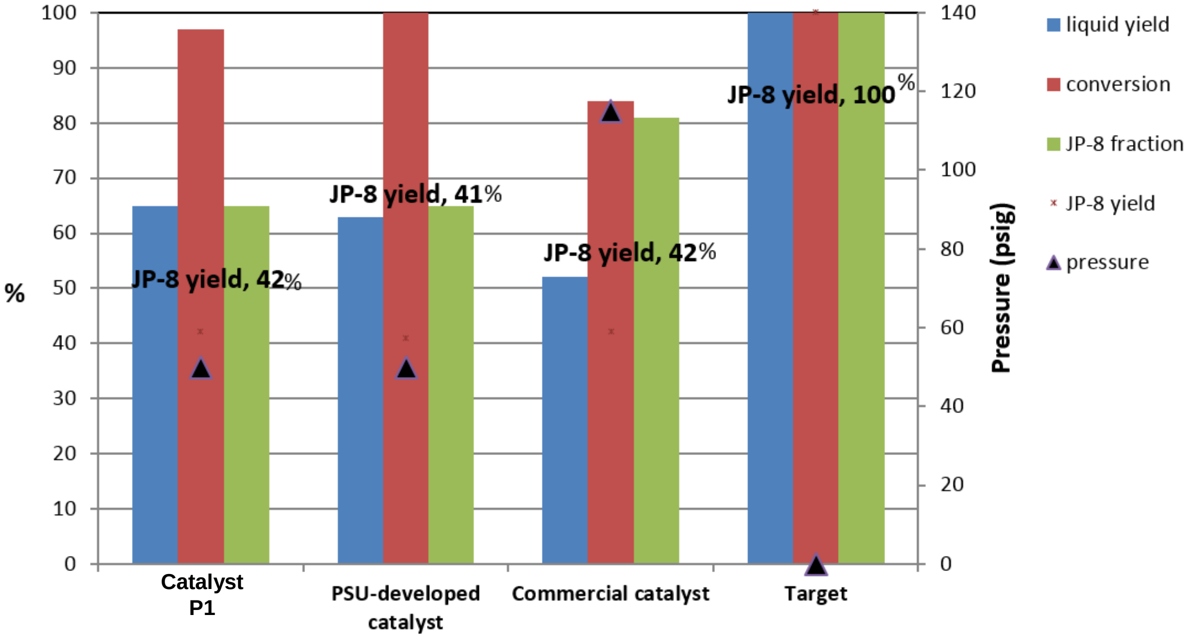


Figure ES-1. Summary of JP-8 yield achieved using Catalyst P1 (50 psi), PSU-developed (50 psi), and Commercial (115 psi) catalysts of ethylene (C_2) oligomerization

Preliminary heat and mass balance data for a 50,000 BPD plant were used in a Life-Cycle Analysis (LCA) to estimate the Well-To-Wake (WTWa) life-cycle Greenhouse-Gas (GHG) emissions. The LCA was performed by Argonne National Laboratory (ANL) and a GHGR-CBTL module was added to the Greenhouse gases, Regulated Emissions, and Energy use in Transportation (GREET®) model. The LCA analysis showed that, based upon the preliminary heat and mass balances, different proportions of biomass would be required for the WTWa GHG emissions to match those of petroleum-based JP-8 (84 gCO_2e/MJ). If the char from the process was utilized to produce combined heat and power (CHP) with the electricity produced offsetting the electricity production elsewhere, then a feed containing 23 wt% biomass and 77 wt% lignite would be needed for parity with petroleum-based JP-8. If the char is not utilized for CHP, but sequestered in a land fill, more biomass is needed since no grid electricity is displaced, and 31 wt% biomass is required. Higher proportions of biomass in either case would result in JP-8 with lower WTWa GHG emissions.

To conduct the TEA, the plant was reconfigured to eliminate natural-gas and electricity imports in order to have a self-sufficient plant. This was accomplished by changing from using the waste-heat-recovery steam from the steam cracker for the production of utility steam to using it for the production of electricity. This led to changing from utilizing the waste-heat-recovery steam and combustion of imported natural gas for the production of utility steam to utilizing the combustion of excess fuel gas and some of the char for the production of utility steam.

The TEA was performed following DOE guidelines listed in the FOA and the referenced documents. A matrix of plant configurations was developed and evaluated. The scenarios included capturing or venting CO_2 in order to estimate the cost of CO_2 capturing. A baseline case with CO_2 capture was included and was compared with a similar FT baseline case that has been studied by DOE with similar assumptions. The results are shown in Figure ES-2 that compares the Total Overnight Cost (TOC) and Equivalent Crude Oil (ECO) price for GHGR-CBTL with those calculated from the results of the DOE/NETL FT-Liquids baseline case[1] (adjustment was made for the difference in capital-charge factor between this project's FOA (0.237) and those used in the FT-Liquids baseline Case (0.218)).

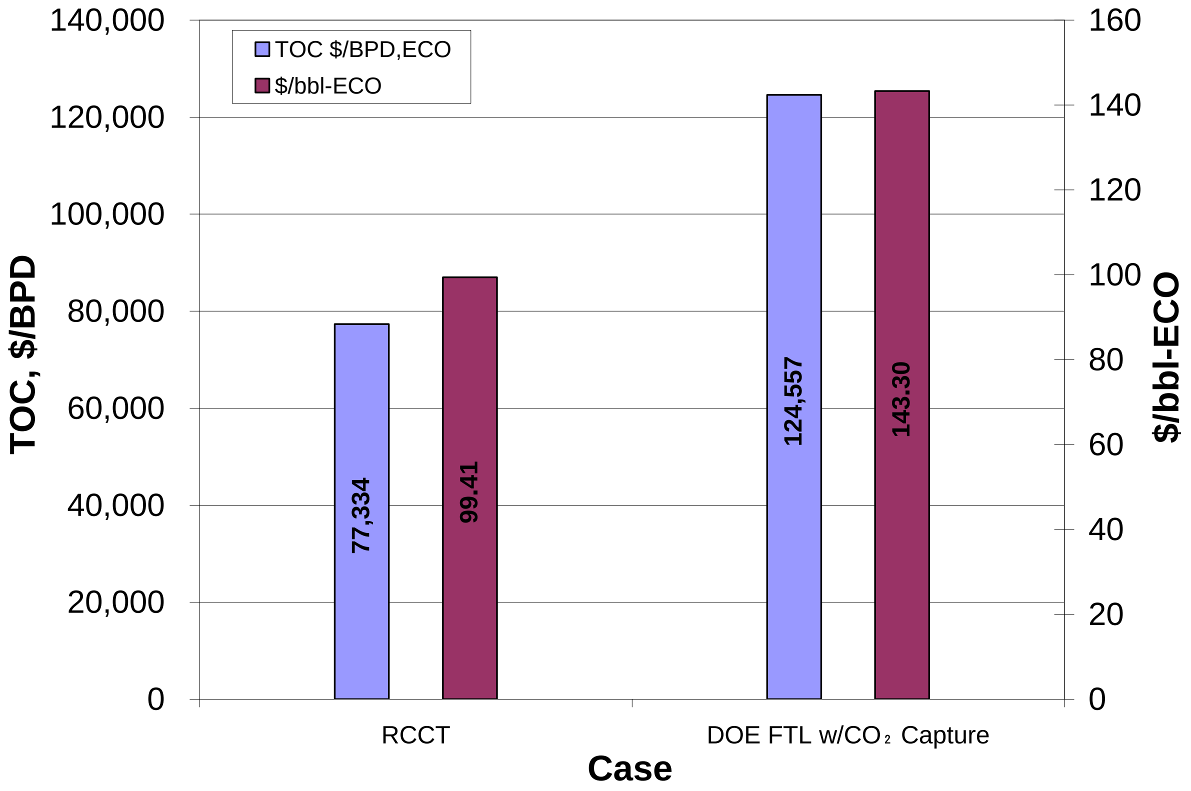


Figure ES-2. Comparison of TOCs and ECO prices for GHGR-CBTL and the DOE/NETL FT-Liquids baseline case.

As shown the GHGR-CBTL TOC and ECO are 62% and 69% of the TOC and ECO for the DOE FT-Liquids Baseline Case, respectively. This shows that the economics of the GHGR-CBTL process are significantly better than a gasification/FT process.

While the 1-TPD system was not fully completed and tested, the previous TRL 4 testing that has been performed on coal and biomass, the detailed design of the 1-TPD system, the tests that were completed, the detailed design of a full-scale plant all support the feasibility of the GHGR-CBTL process.

It is important to note that the GHGR-CBTL process utilizes unit operations that have already been proven and have been put into commercial operation. In particular, ethylene production from the steam cracking of hydrocarbon feedstocks, including heavy feedstocks, is one of the largest chemical industries in the world. This demonstrates that the process and the equipment associated with the process are readily available at the large scales needed for significant fuel production. Although the process needs to be adapted to utilizing pyrogas rather than petroleum-based hydrocarbons, tests to date have demonstrated the production of olefins from pyrogas.

Even more directly related is oligomerization of olefins to liquid fuels. This process is well known and has been commercialized for utilization with both heterogeneous and homogeneous catalysts (*e.g.*, Shell Higher Olefin Process, Dimersol processes), although these operate at higher pressures (>550 psig) than the GHGR-CBTL catalysts. This means that there are drop-in replacements for the GHGR-CBTL oligomerization process that are commercially available.

Utilizing two processes that are so well developed and commercialized greatly reduces the risk of the GHGR-CBTL process.

While the system was not fabricated and tested, major efforts were expended to design the 1-TPD and a full-scale plant. No technical barriers were identified. The TEA was applied to the full scale plant followed DOE guidelines and the lower cost of the GHGR-CBTL than FT is consistent with the GHGR-CBTL process simplicity and its lower temperatures. These lower costs and the detailed design that was performed under this project are being used by Altex to attract funding partners to move the GHGR-CBTL development forward.

1 INTRODUCTION

This Final Technical Report describes the work and accomplishments of the project entitled, "Green-House-Gas-Reduced Coal-and-Biomass-to-Liquid-Based Jet Fuel (GHGR-CBTL) Process".

The main objective of the project was to raise the Technology Readiness Level (TRL) of the GHGR-CBTL fuel-production technology from TRL 4 to TRL 5 by producing a drop-in synthetic Jet Propellant 8 (JP-8) with a lower greenhouse-gas footprint than petroleum-based JP-8 by utilizing mixtures of coal and biomass as the feedstock.

Supporting objectives include: (1) increasing process throughput to >1 bpd, (2) combining previously developed approaches to producing JP-8 from coal, JP-8 from housing waste, gasoline from lignocellulosic biomass, and gasoline from mixtures of coal and biomass to produce JP-8 from mixtures of coal and biomass, (3) assessing the life-cycle Greenhouse Gas (GHG) footprint of synthetic JP-8 produced by the process, and (4) assessing the technical and economic viability of the process.

This report discusses the execution of the project and its results. It includes discussion of the feedstock supply, Full-scale plant process design, Life Cycle Analysis (LCA), test system/component design, fabrication, testing, and assessment.

2 FEEDSTOCK ACQUISITION

This section of this report summarizes activities performed on feedstock acquisition. The goals of these activities were to identify and acquire the needed biomass-feedstock for GHGR-CBTL.

To prepare the coal/biomass mix, and the biomass was supplied by UC Davis and coal was furnished by North American Coal Company.

2.1 Biomass Acquisition

A subcontract was executed with UC Davis for biomass identification, field production and supply under GHGR CBTL project. Under this subcontract, UC Davis has supplied more than 5 tons of switchgrass, wheat straw, and corn stover. In addition, for certain feedstocks, UC Davis provided data to conduct the Life-Cycle Analysis (LCA) of carbon intensity. These data were included in the Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation (GREET) model of carbon intensity developed by Argonne National Laboratory (ANL), another subcontractor under the cost shared GHGR project.

2.1.1 Feed Stock Identification

GHGR-CBTL is feedstock flexible. To show this flexibility three types of biomass were identified to be used for testing under GHGR-CBTL. These are switchgrass, corn stover, and wheat straw. These biomass feedstocks were selected because they are readily or potentially available and they are good candidates in terms of low cost and low maintenance for conversion to biofuels specifically gasoline under GHGR-CBTL. Wheat is the major small grain grown in California and is grown in more than 500,000 acres. Unlike other straws, wheat straw is sturdy and very strong. Hence, it is not used as a bedding material and is generally considered a waste. Disposal of the wheat straw produced in California is a big problem. Farmers are paying for removal of straw from the field. Utilizing this low-value and readily available biomass as a feedstock will eliminate the problems in receiving biomass throughout the year in a commercial scale GHGR-CBTL plant. Hence, wheat straw is considered to be a good candidate for GHGR-CBTL leading to production of biofuel (e.g., gasoline) within California.

Besides producing large quantities of wheat straw, the US is the top corn producer in the world. The cornfield residues in the form of stalks, leaves, and cobs that remain in fields after harvesting are referred as corn stover. According to a CEC 2005 report [2] California produces about 5 million tons per year of field crop residues, principally as cereal straws and corn stover. Because of its high ash content, this material is not used in direct biomass power generation, but it is a good candidate for thermal conversion in biofuels production in GHGR-CBTL.

In the commercial-scale GHGR-CBTL plant, all these three selected biomass materials would be densified in the field and transported to the GHGR-CBTL plant. This would reduce raw-material costs by lowering the feedstock-transportation costs in the GHGR-CBTL plants []. Altex has produced densified logs using over ten tons of these three biomass materials using BBADS equipment as shown in Figure 1. This figure shows the BBADS equipment while being field-tested at UC Davis under support from CEC [3]. Figure 2 shows pictures of densified logs produced from switchgrass, corn stover, and wheat straw. Densifying these residues using BBADS process has potential to reduce the logistics cost involved in transportation of biomass from the field to the GHGR-CBTL plant.

Figure 1. The BBADS equipment that is part of the GHGR-CBTL full scale pant for densifying the feedstocks to minimize the feedstock transportation cost and energy.

Figure 2. Densified feedstock logs; a) switchgrass, b) corn stover, and c) wheat straw.

2.1.2 Biomass Production

Various agricultural crops and crop residues were evaluated at UC Davis as candidates for biomass and liquid fuel production to be tested by Altex. Two biofuels grasses were grown and tested for ease of crop establishment, nitrogen use and requirements, harvest schedule and biomass yield. These included switchgrass (Panicum virgatum), and miscanthus (Miscanthus x giganteous). Other potential biomass sources for fuel production in California include agricultural and urban wastes, rice and wheat straw, corn stover, wood chips from tree pruning and dedicated energy crops.

Switchgrass (Panicum virgatum), and miscanthus (Miscanthus x giganteous) were chosen as dedicated energy crops due to their high yield potential. Switchgrass has some advantages due to its high yield potential combined with excellent conservation attributes and good compatibility with conventional farming practices. For example, switchgrass is established from seeds, and miscanthus is established by the transplanting of rhizomes. Switchgrass (Panicum virgatum) is a perennial warm season C4 grass showing promise as a cellulosic biofuel crop in California. In Davis, 6 lowland and 5 upland ecotypes were evaluated. Lowland ecotypes are taller, coarser, have a more bunch-type growth habit and may be more rapid growing than upland ecotypes. In contrast, upland ecotypes are found in drier upland sites and tend to be finer stemmed and lower yielding. The experiment was set up as a completely randomized block design with six replications. Results from research at UC Davis has shown that with lowland ecotypes, biomass yields up to 17 tons per acre (38 t/ha) could be achieved.

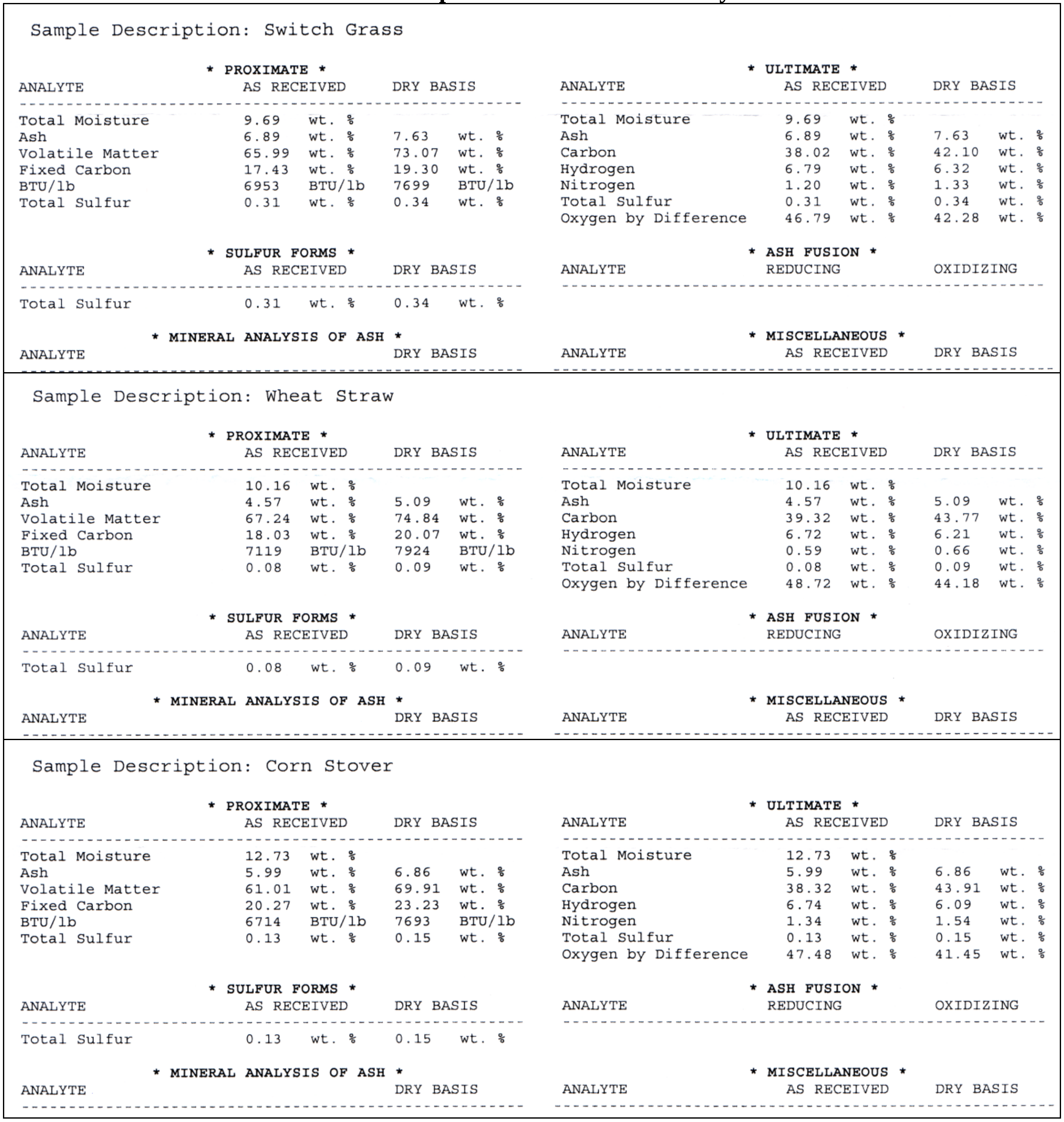
Biofuels grasses (Switchgrass and Miscanthus) were harvested in a two-cut system in 2014-2015 at UC Davis research farm. First harvest took place late-June, second harvest mid-October. The first harvest of switchgrass produced significantly more biomass than subsequent harvests for all varieties. Of the total biomass, approximately 70% was obtained in the first harvest at Davis, and a similar trend was found for Miscanthus as well.

From the harvested test crop, two tons each of Switchgrass and Miscanthus were retained at the UC Davis test site for testing purposes at Altex. In addition to the biofuels grasses, crop residues from corn and wheat (3 ton each) were baled and stored upon request by Altex for initial pilot-scale demonstration of Altex equipment.

2.1.3 Biomass Acquisition

The initial acquisition amounts of the biomass feedstocks were based on one month of testing in the GHGR-CBTL system. This requires 10 tons of biomass. A request was made to UC Davis to harvest, bale, and supply 10 tons of biomass. This amount includes switchgrass, corn stover, and wheat straw. GHGR-CBTL requires particles size of about 1 mm for efficient thermal conversion of the biomass in the pyrolysis process. Altex has contacted several companies to obtain quotes for size reduction of the biomass feed. Based on the quotes received and company capabilities, Andritz was selected to be used for biomass sizing in GHGR-CBTL. Accordingly, ten tons of biomass were transported from UC Davis to Andritz and comminution was carried out. After size reduction of biomass, the same was sent to Hazen research Inc for mixing with ground coal. In addition, switchgrass, wheat straw and corn stover samples were sent to proximate and Ultimate analysis. Table 1 shows the proximate and ultimate analysis of the switchgrass, wheat straw, and corn stover. This data will be used in the design and analysis of the CBDF system.

Table 1 Sample bimoass feedstock analysis



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 Sample Description: Switch Grass

                      * PROXIMATE *                                              * ULTIMATE *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
Total Moisture          9.69   wt. %                      Total Moisture            9.69   wt. %
Ash                     6.89   wt. %     7.63   wt. %     Ash                       6.89   wt. %     7.63   wt. %
Volatile Matter         65.99  wt. %     73.07  wt. %     Carbon                    38.02  wt. %     42.10  wt. %
Fixed Carbon            17.43  wt. %     19.30  wt. %     Hydrogen                  6.79   wt. %     6.32   wt. %
BTU/lb                  6953   BTU/lb    7699   BTU/lb    Nitrogen                  1.20   wt. %     1.33   wt. %
Total Sulfur            0.31   wt. %     0.34   wt. %     Total Sulfur              0.31   wt. %     0.34   wt. %
                                                          Oxygen by Difference      46.79  wt. %     42.28  wt. %

                      * SULFUR FORMS *                                           * ASH FUSION *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                 REDUCING           OXIDIZING
-------------------------------------------------------   ----------------------------------------------------
Total Sulfur            0.31   wt. %     0.34   wt. %

                * MINERAL ANALYSIS OF ASH *                                      * MISCELLANEOUS *
ANALYTE                                  DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
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 Sample Description: Wheat Straw

                      * PROXIMATE *                                              * ULTIMATE *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
Total Moisture          10.16  wt. %                      Total Moisture            10.16  wt. %
Ash                     4.57   wt. %     5.09   wt. %     Ash                       4.57   wt. %     5.09   wt. %
Volatile Matter         67.24  wt. %     74.84  wt. %     Carbon                    39.32  wt. %     43.77  wt. %
Fixed Carbon            18.03  wt. %     20.07  wt. %     Hydrogen                  6.72   wt. %     6.21   wt. %
BTU/lb                  7119   BTU/lb    7924   BTU/lb    Nitrogen                  0.59   wt. %     0.66   wt. %
Total Sulfur            0.08   wt. %     0.09   wt. %     Total Sulfur              0.08   wt. %     0.09   wt. %
                                                          Oxygen by Difference      48.72  wt. %     44.18  wt. %

                      * SULFUR FORMS *                                           * ASH FUSION *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                 REDUCING           OXIDIZING
-------------------------------------------------------   ----------------------------------------------------
Total Sulfur            0.08   wt. %     0.09   wt. %

                * MINERAL ANALYSIS OF ASH *                                      * MISCELLANEOUS *
ANALYTE                                  DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
```

```
 Sample Description: Corn Stover

                      * PROXIMATE *                                              * ULTIMATE *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
Total Moisture          12.73  wt. %                      Total Moisture            12.73  wt. %
Ash                     5.99   wt. %     6.86   wt. %     Ash                       5.99   wt. %     6.86   wt. %
Volatile Matter         61.01  wt. %     69.91  wt. %     Carbon                    38.32  wt. %     43.91  wt. %
Fixed Carbon            20.27  wt. %     23.23  wt. %     Hydrogen                  6.74   wt. %     6.09   wt. %
BTU/lb                  6714   BTU/lb    7693   BTU/lb    Nitrogen                  1.34   wt. %     1.54   wt. %
Total Sulfur            0.13   wt. %     0.15   wt. %     Total Sulfur              0.13   wt. %     0.15   wt. %
                                                          Oxygen by Difference      47.48  wt. %     41.45  wt. %

                      * SULFUR FORMS *                                           * ASH FUSION *
ANALYTE                 AS RECEIVED      DRY BASIS        ANALYTE                 REDUCING           OXIDIZING
-------------------------------------------------------   ----------------------------------------------------
Total Sulfur            0.13   wt. %     0.15   wt. %

                * MINERAL ANALYSIS OF ASH *                                      * MISCELLANEOUS *
ANALYTE                                  DRY BASIS        ANALYTE                   AS RECEIVED      DRY BASIS
-------------------------------------------------------   ----------------------------------------------------
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2.2 Coal Acquisition

As a cost shared item, Coal was supplied by North American Coal Corporation (NACOAL). The relationship between Altex and NACOAL was developed under DARPA supported Clean and Low Cost (CALC) Coal Conversion project. Then, also Low-rank coal (lignite from North Dakota and subbituminous coal from Montana) was procured. One ton of lignite was obtained from the Falkirk Mine. Under this project, the coals was not analyzed but it is similar to what was received and analyzed in the past. The proximate and ultimate analysis of the lignite conducted under the CALC project is presented in Table 2.

Table 2. Analysis of lignite.

Reporting Basis	As Received	Dry	Air Dry
Ultimate (%)			
Moisture	38.19	0.00	5.61
Carbon	35.36	57.21	54.00
Hydrogen	2.34	3.78	3.57
Nitrogen	0.60	0.96	0.91
Sulfur	0.88	1.43	1.35
Ash	14.07	22.77	21.49
Oxygen	8.56	13.85	13.07
Total	100.00	100.00	100.00
Proximate (%)			
Moisture	38.19	0.00	5.61
Ash	14.07	22.77	21.49
Volatile	24.12	39.03	36.84
Fixed C	23.62	38.20	36.06
Total	100.00	100.00	100.00
Sulfur	0.884	1.430	1.350
Btu/lb (HHV)	5771	9338	8814
MMF Btu/lb	6793	12417	
MAF Btu/lb		12091	

NACOAL shipped 20 tons of coal to Hazen Research (Hazen) where it was ground, sieved at 35 and 80 mesh and mixed with the ground biomass that was received from Andritz. A comparison of the ultimate analyses for this lignite and subbituminous coal that was also received under CALC and literature values of subbituminous coal are presented in Table 3.

Table 3. Comparison of Montana subbituminous coal and North Dakota lignite.

Coal	Type	Moisture (As rec'd)	Ash (dry)	C (maf)	H (maf)	N (maf)	S (maf)	Cl (maf)	O (maf)
Montana[(a)]	subB	23.66	5.35	76.00	5.23	0.94	0.53	0.04	16.35
Montana	subB	22.09	5.95	79.30	5.21	1.40	0.49	NM	13.60
North Dakota[(b)]	ligA	39.11	18.96	70.26	4.69	0.69	1.28	NM	23.08

[(a)] Literature data from Penn State Coal Database as reported by Smith, Smoot, Fletcher, and Pugmire, "The Structure and Reaction Processes of Coal," p. 33 (1994).

[(b)] Data obtained from 3rd-party analysis of sized lignite received from Falkirk Mine in North Dakota.

NM = not measured.

2.3 Coal/Biomass Mix Acquisition

Coal requires environmentally controlled storage space to prevent rapid moisture loss leading to production large quantities of fine dust. Large fines in the coal increases the environmental hazards during storage, handling and utilization. So based on the storage facility available and testing needs, it was decided to obtain 10 supersacks of coal and biomass mix from Hazen for utilization under GHGR CBTL testing. Accordingly only 5 tons of coal was ground as per size requirement and blended with ground biomass received from Andrtiz. The blended coal and biomass was received at Altex and will be used for testing in the GHGR CBTL process.

2.4 Feedstock Acquisition Summary

The GHGR-CBTL feedstocks have been identified and acquired. Three biomass feedstocks were selected for testing in the GHGR-CBTL system based on the feedstock availability, cost, and having the potential to increase farmers profit as full scale GHGR CBTL plants are constructed in the future. The feedstocks to be tested in the GHGR-CBTL system are switchgrass, corn stover, and wheat straw. These three feedstocks have also been densified into 40 lb/ft3 logs using Altex BBADS field equipment that would be part of the GHGR-CBTL full-scale plant. By densifying the loose biomass in the field, BBADS reduces biomass transportation costs and energy use. Upon Altex request, UC Davis has produced and shipped 10 tons and the same was sized at Andrtiz using hammer mill for use in GHGR-CBTL. NA Coal has shipping 20 tons of coal to Hazen and sized to required size. Blending of coal and biomass is completed at Hazen and stored in super sacks. The sized biomass and coal will be shipped to Altex in super sacks designed to be used in the bulk-bag unloader of the GHGR-CBTL feed system. With these activities, the biomass identification and acquisition, planned under the GHGR-CBTL project, is completed.

3 LIFE CYCLE ANALYSIS AND LOGISTICS MODELING-

Altex and Argonne National Labs (ANL) worked together to develop the LCA model. Altex performed the H&M balance calculations for LCA and ANL developed a GHGR-CBTL module in GREET. The process steps for carrying out LCA in GREET was based on the well-to-wheel analysis and the process steps are shown in Figure 3.

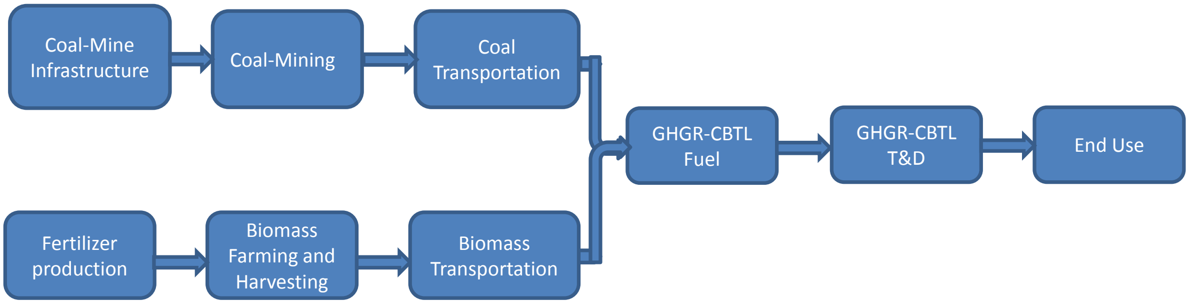


Figure 3 Process steps to carryout LCA under GHGR-CBTL

In order to carry out the LCA a 50,000 GHGR commercial plant was selected as a base. This plant size is reasonable for economical purposes. Also, there is a DOE baseline study on a 50,000-BPD Fischer-Tropsch plant that can be used as a basis for comparison.[4].

The LCA work was carried out with detailed material and energy flow in (i) Production and transportation of coal and biomass to GHGR CBTL plant (ii) production of JP8 within GHGR CBTL plant and (iii) Combustion of biofuel in transportation vehicles and presented below.

Table 4 Summary of LCA data for Biomass in the GHGR-CBTL Process

	Source : GREET Corn (per bushel)	**Source: CA Corn study (San Jose)** Corn (per bushel)	**Source : GREET** Corn Stover (per dry ton collected)	**Source : GREET** Switchgrass (per dry ton)
Farming Energy Use: Btu	9,608	11,133	247,053	177,700
Fertilizer Use				
Grams of Nitrogen	423	567	7,000	7,300
Grams of P_2O_5	146	58	2,000	100
Grams of K_2O	151		12,000	200
Grams of $CaCO_3$	1,150			
Pesticide Use				
Grams of Herbicide	7	0.01		28
Grams of Insecticide	0.06	0.10		
Water Consumption Factor (gallons/bushel)	146	5,576		

3.1 Material and Energy Flow – Production and transportation of coal and biomass

The material flow into the GHGR are biomass and coal. The biomass data were provided to ANL by UC Davis and the coal data were provided by North American Coal. Altex also has provided material and energy flow for biomass densification (BBADS) that is part of the biomass cycle.

3.1.1 Biomass production

Corn, corn stover and switchgrass feedstocks were selected for inclusion in the analysis for GHGR CBTL project. Regarding wheat straw, Argonne has used published results from literature as input. Table 4 summarizes the energy and material flows associated with cultivating and harvesting these biomass types. From Table 4, in addition to reporting the average material and energy flows for corn production in the US per GREET, the regional specific data for corn harvested in San Jose (California) was included by tapping into UC Davis cost studies [5]. This approach was adopted in developing the GHGR-CBTL LCA module in GREET that offered both corn data sets to generate California-specific and national-level results.

3.1.2 Coal production

Table 5 Energy Consumption for Surface Lignite Coal Extraction

Energy consumed: Btu/MMBtu of lignite coal	**Heating-Value Basis**	
	LHV	**HHV**
Diesel fuel	11,671	9,516
Bio-Diesel fuel	897	731
Gasoline	223	182
Electricity	2,495	2,034
Total Energy Consumed	**15,286**	**12,463**

The coal data was developed based on the input from NA Coal Corporation. The data available from NA Coal corporation's southern surface lignite mines that produces 2.5 million tons a year was used for developing metrics specific for GHGR CBTL process. North American Coal's mines use residual oil, natural or coal as energy sources to operate mobile equipment used in coal mining. Assuming a lower heating value (LHV) and higher heating value (HHV) of 5,300 and 6,500 Btu/lb, respectively, for lignite coal, the energy consumption data reported in Table 5. In the GHGR-CBTL LCA module to be developed in GREET, option will provided using energy consumption values based on both LHV and HHV reported in Table 5, the former will however be the default case. Coal mining produces fugitive dust emissions from driving mining vehicles, material handling, and other operations. GREET has default fugitive emissions data from coal mining that will applied in the LCA model.

3.1.3 Biomass Densification

The energy consumption rates for BBADS biomass densification were provided by Altex were included into the GREET model. These rates were based on independently verified experimental measurements by Warren Energy Engineering and tests that were performed at Altex. The electrical energy and thermal energy requirements were converted uniformly to the unit of MMBtu for assembling in the GREET modelling software. The reported values of 0.041 MMBtu/ton and 0.075 MMBtu/ton dry feedstock of electrical and thermal energy, respectively were incorporated into GREET model.

3.2 Material and Energy Flow - GHGR CBTL process

3.2.1 Block Diagram of GHGR CBTL Plant

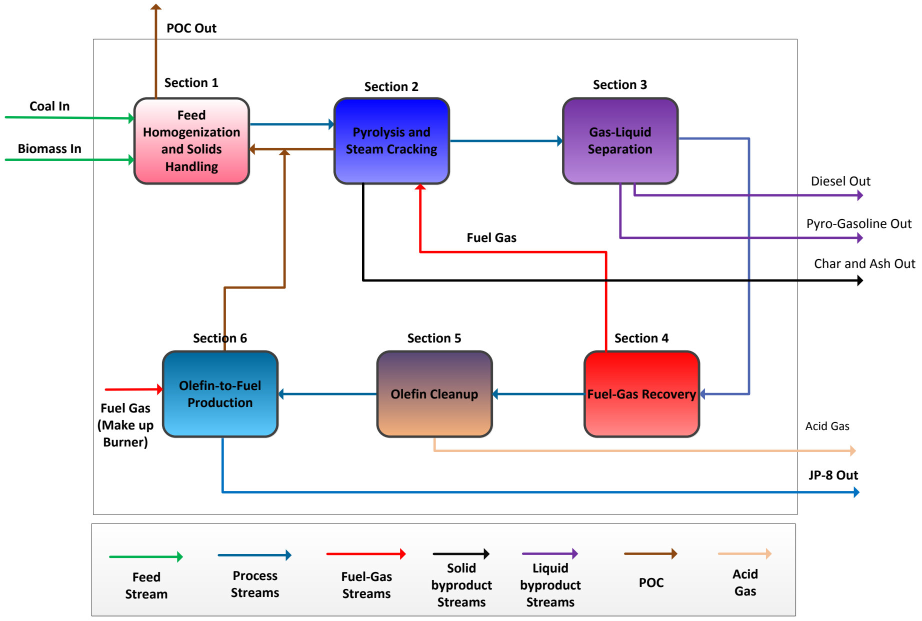


Figure 4 Block Diagram of CBTL GHGR

The GHGR CBTL plant was divided into six different sections, and a block diagram of the plant was prepared (Figure 4). The block diagram was used to prepare Process Flow Diagrams (PFDs) of the different sections using Microsoft Visio. These PFDs were used to develop a flowsheet of the entire plant in CHEMCAD. CHEMCAD was then used to specify the inlet streams, specify the operating conditions of the plant unit operations, and to finalize the heat and mass balance of the entire plant.

CHEMCAD simulates standard unit operations in the chemical process industry including heat exchangers, distillation towers, absorption towers, reactors, pumps, compressors, and many others. Non-standard unit operations can be simulated using combinations of standard unit operations or by using user-developed code for the unit operation. CHEMCAD was used to solve steady-state heat and mass balances around the unit operations and around the entire plant including all recycle streams. This type of information can be used to size the heat exchangers, which can then be used to develop cost estimates for use in the TEA.

After completing the heat and mass balance, the results were linked to the PFDs of the plant sections. The interactive PFD provides a platform for analyzing all the streams and unit operations of the entire GHGR CBTL plant in a user-friendly interface In this way, Altex and ANL engineers can view the PFD and simulation results to further specify, design, and cost equipment without needing to use CHEMCAD.

The GHGR-CBTL process has six main sections and produces jet fuel and No-2 diesel as the main and co-product, respectively.

3.3 Material and Energy Flow – Transportation and Combustion of JP-8

3.3.1 Transportation LCA Assumptions

The LCA of the fuels produced via the GHGR-CBTL process includes transportation and distribution (T&D) steps for both the feedstocks (coal, biomass) and the final fuel products. Important T&D parameters include transportation mode, transportation distances, payload, and feedstock density. Altex has identified three locations for building GHGR CBTL plant, and one site will be finalized and T&D analysis and sensitivity analysis will be carried out. Standard scenarios of payload and the transportation modes were used in the GREET model and updated with the BBADS densification effect. The model is setup such that the user can select transportation mode, distance, biomass type, the payload density and so on. This makes the model general fro adaption to many plant locations.

3.3.2 Fuel Use Stage Parameters

After the fuel (diesel and/or jet) is delivered to the point of use, the fuel will be combusted in an airplane or a marine diesel vehicle. The emission factors for CO_2 emissions in this phase are based on the carbon content of the fuel. SOx emissions are based on the fuel sulfur content. Other emission factors are matched to those of conventional jet (JP-8), which are available in the original GREET model.

3.4 Argonne National Laboratories GHGR GREET Model Development

Collaborating with Altex, Argonne National Laboratory expanded and used the Greenhouse gases, Regulated Emissions, and Energy use in Transportation (GREET®) model [6]to assess life-cycle GHG emissions and water consumption of the GHGR technology. The GREET model includes various feedstocks, conversion technologies, and fuels. In addition, the GREET model includes all transportation modes, including road, rail, marine, and air transportation. For this project, we expanded the GREET model to include wheat straw farming and collection, biomass densification and transportation, the CBTL process (feedstock inputs, energy use, fuel and char production, and gas emissions), char to landfill (Char-LF), and char for combined heat and power (Char-CHP). However, a caveat is that the model for this study was built on GREET 2015, and some minor revisions were included in the GREET 2016 model [7].

Specific issues that Argonne was asked to address included (1) GHG emissions of biomass feedstocks (e.g., corn stover, switchgrass and wheat straw), in g CO2e/dry ton feedstock, from the farming field to the refinery gate; (2) WTWa GHG emissions and water consumption of CBTL fuel composed of 85 wt% coal and 15 wt% biomass; and (3) the break-even point of biomass in the feedstocks and paths forward to achieve 84 gCO_{2e} per megajoule (MJ) (the WTWa GHG emissions level of conventional jet fuel).

3.4.1 Life-Cycle Analysis System Boundary and Functional Unit

The system boundary of this life-cycle analysis (LCA) study encompasses all operations related to coal mining and cleaning, coal transportation, biomass farming and harvesting, biomass densification and transportation, fuel production, fuel transportation and distribution, and jet fuel combustion (Figure 5). These WTWa stages can be divided into Well-to-Pump (WTP) stages and Pump-to-Wake (PTWa) stages. The PTWa stages include the jet fuel combustion, while the WTP stages include the rest of the WTWa stages. In addition, the system boundary includes the indirect GHG emissions and energy and water consumption associated with materials and energy production, such as fertilizer production and application. However, the indirect GHG emissions and energy and water consumption associated with the infrastructure materials and equipment used in coal mining and cleaning, biomass farming, and fuel production are generally much smaller than those associated with fuel production and combustion, and therefore are not included [8]. The functional unit of this LCA is 1 MJ of jet fuel on the basis of the lower heating value (LHV).

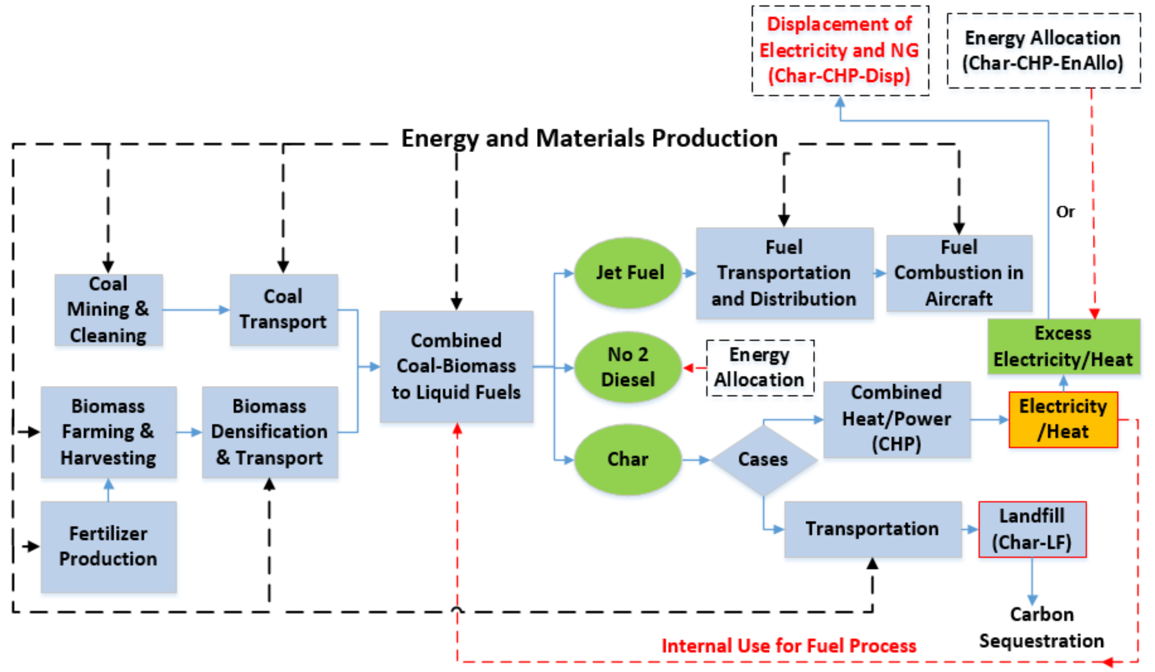


Figure 5. Life-cycle analysis system boundary of combined coal and biomass conversion to liquid fuels (CBTL): coal mining and cleaning, biomass farming and harvesting, fertilizer production, coal transportation, biomass densification and transportation, fuel processing, fuel transportation and distribution, and fuel combustion in aircraft

3.4.2 Coal and Biomass Feedstocks

The coal resource utilized in Altex's CBTL technology is lignite coal. Lignite coal (brown coal) is considered the lowest quality rank of coal because of its relatively low heating content. However, lignite coal has a high content of volatile matter, which makes it easier to convert into gas and liquid fuels than other higher-ranking coals. The coal energy content and carbon ratio are listed in Table 6. In this study, coal is assumed to be from surface mining. Estimates for the energy consumption for lignite coal mining and cleaning are also summarized in Table 6. The process energy sources include diesel, gasoline, and electricity, and the total energy consumption rate is 15,286 Btu/MMBtu of coal mined and cleaned. For surface coal mining and processing, the water consumption factor is 2.6 gal/MMBtu coal [7,9]. During coal mining and cleaning, methane (CH_4) formed during coalification can escape into the atmosphere. Burnham *et al.* [10] reported CH_4 emissions during coal mining and post-mining operations. The average CH_4 emissions were estimated to be 49.5 g CH_4 per MMBtu (or 535 g/ton) of surface-mined coal. Other non-combustion emissions during coal mining and cleaning include volatile organic compounds (VOCs), PM_{10} and $PM_{2.5}$, which are also listed in Table 6.

Biomass feedstocks considered in this study include corn stover, switchgrass, and wheat straw. Table 7 summarizes the energy and material flows associated with cultivating and harvesting these biomass feedstocks, which are based on the latest updated values [6,11,12].

For corn stover, corn grain and stover are harvested separately with two passes (first pass for grain and second pass for corn stover). The second-pass energy consumption is assigned to corn stover. For wheat straw, the collection was modeled in Reference 12.

Table 6. Lignite coal energy content, carbon ratio, energy consumption for surface coal mining and cleaning, and non-combustion emissions during coal mining and cleaning

Lignite coal energy content and carbon ratio	Value	Data source
LHV (MMBtu/ton)	10.6	Altex
C (wt%)	63%	Altex
Energy use in surface lignite coal mining and cleaning[a]	15,286	Altex
Diesel fuel (Btu/MMBtu)	11,671	Altex
Bio-diesel fuel (Btu/MMBtu)	897	Altex
Gasoline (Btu/MMBtu)	223	Altex
Electricity (Btu/MMBtu)	2,495	Altex
Non-combustion emissions during coal mining and cleaning		
VOCs (g/MMBtu)	6.8	GREET
PM_{10} (g/MMBtu)	12	GREET
$PM_{2.5}$ (g/MMBtu)	1.5	GREET
CH_4 (g/MMBtu)	49.5	GREET

[a]The calculated energy consumption values were based on the LHV of lignite coal.

Table 7. Biomass farming energy and fertilizer use, and biomass N content and N_2O emissions.

	Corn stover[a]	Switchgrass[a]	Wheat straw[b]
Farming energy and fertilizer use			
Farming energy use (Btu/dry ton)	192,500	177,700	199,600
N fertilizer (g/dry ton)	7,000	7,300	4,987
P_2O_5 (g/dry ton)	2,000	100	1,269
K_2O (g/dry ton)	12,000	200	6,895
Herbicide (g/dry ton)	0	28	0
Biomass N content and N_2O emission			
N content in biomass (%)	0.77%	0.50%	0.55%
N_2O emission conversion rate from N in biomass (N in N_2O as % of N in biomass)	1.225%	1.225%	1.225%
N_2O emission conversion rate from N in fertilizer (N in N_2O as % of N in N fertilizer)	1.525%	1.525%	1.525%

Data sources: [a]Reference 6 and [b]Reference 11

The supplemental fertilizers (N, P_2O_5 and K_2O) for corn stover and wheat straw are applied to replace the nutrients in the harvested biomass [6,13], because if corn stover and wheat straw remain on the field, these nutrients in biomass could enter the soil and be available for future crops. In this study, we also considered N_2O emissions from below- and above biomass decay and fertilizer application; N_2O emission factors for N in biomass and synthetic fertilizer are assumed to be 1.225% and 1.525% (N in N_2O as % of N in N biomass and fertilizer), respectively, as shown in Table 7 [6,14]. In this analysis, water use is defined as the amount of water withdrawn from freshwater sources. Han *et al.* [15] assumed that corn stover does not consume water, since irrigation is mainly for corn farming and not for corn stover harvesting. We used the same method for corn stover and wheat straw in this study. There is no irrigation water demand for switchgrass farming [6,9].

Unprocessed biomass often has low volumetric energy and bulk density and is aerobically unstable, making handling and transportation inefficient. Densification is one way to increase the volumetric energy density and overcome handling and transportation difficulties. The farming energy use summarized in Table 7 includes the energy demand for baling. For densification cases, this study assumed that the bales are further densified into pellets (small compressed pieces of biomass produced using hammer and pellet mills). The consumption rates of electricity and natural gas (NG) for bale-to-pellet densification were 81,891 and 75,120 Btu/dry ton, respectively, as shown in Table 8. We assumed that the biomass was transported by a heavy heavy-duty 53-ft flatbed trailer (25-ton and 106-m^3 capacity) with fuel economy of 7.4 miles per diesel gallon equivalent (MPDGE), and the one-way travel distance was assumed to be 53 miles. Because pellets are easier to store, handle, and transport than bales, we assumed that the dry matter losses are 2.0% for bales and 0.0% for pellets.

Table 8. Energy consumption for biomass densification, biomass characterization, and transportation of various cropsa

	Corn-stover bales	**Corn-stover pellets**	**Switchgrass bales**	**Switchgrass pellets**	**Wheat-straw bales**	**Wheat-straw pellets**
Energy consumption for biomass densification (bales to pellets)						
Electricity (Btu/dry ton)	NA	81,891	NA	81,891	NA	81,891
NG (Btu/dry ton)	NA	75,120	NA	75,120	NA	75,120
Biomass characterization						
Bulk density (kg/m^3)	137	587	169	641	137	587
LHV (MJ/dry ton)	15,526	15,526	15,242	15,242	11,184	11,184
Energy density (MJ/m^3)	2,352	10,049	2,847	10,770	2,249	9,612
Moisture (%)	12%	6%	12%	6%	12%	6%
Transportation[b]						
Truck payload (dry ton/load)	14	24	17	24	14	24
Energy consumption (Btu/dry ton)	131,288	78,929	106,449	78,929	131,288	78,929

[a] Based upon tests conducted under the BBADS project[3].

[b] Heavy heavy-duty 53-ft flatbed trailer (25-ton and 106-m^3 capacity) with fuel economy of 17,498 Btu/mile, and one-way travel distance of 53 miles.

3.4.3 Coal and Biomass to Liquid Fuels Conversion Process

According to the Altex information, ANL developed a GHG model for the CBTL process including feedstock drying and size reduction, pyrolysis and steam cracking, fractionation and quenching, oligomerization, and fuel fractionation. We assumed that densified biomass would consume the same amount of energy as non-densified biomass, although their moisture contents are different. The CBTL process has a very high char yield: 0.33 ton char/ton feedstock is equivalent to 9.23 MMBtu/ton feedstock. As compared to the yields for jet fuel and diesel (3.61 and 2.20 MMBtu/ton feedstock, respectively), the char accounts for 61% of total energy outputs. Therefore, the utilization of char (namely, disposal to landfill and combustion for energy recovery) is an important issue in this CBTL process.

Char contains a large amount of C. With 15 wt% of biomass share in the total feedstock, the C content of char is 28 kg C/MMBtu of char, and 90% of this C is fossil C while the remaining 10% is biogenic C. Under a carbon neutrality assumption, the 10% C in the char gets biogenic CO_2 credit, since this C is absorbed from the atmosphere via photosynthesis during biomass growth. When the char is buried in landfills, we assume that 80% (by mass) of the C in the char is stable and sequestered, and 20% is labile C and is converted to CO_2 over time [16]. These CO_2 emissions from 20% of the C in the char are taken into account as emissions. In this Char-LF scenario, char is considered as a waste rather than a coproduct displacing conventional products.

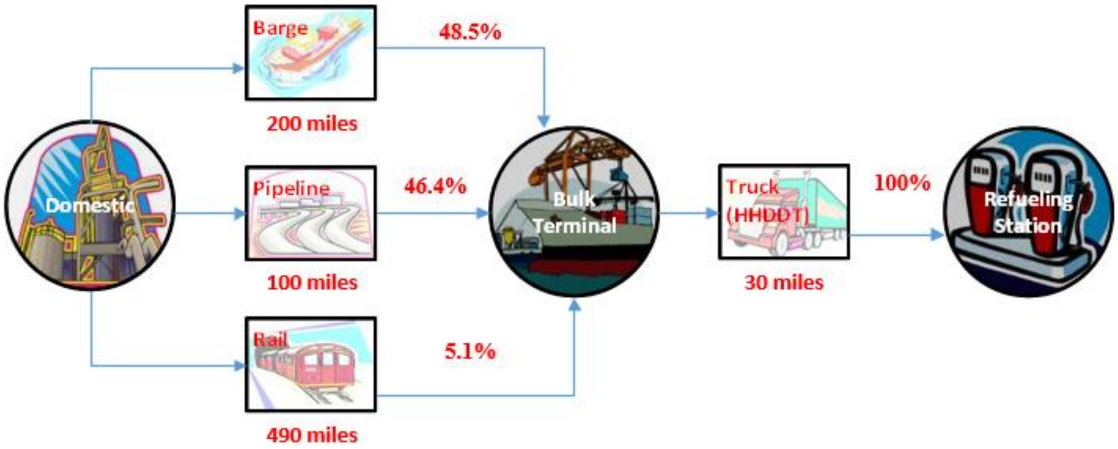
Under the Char-CHP scenario, we assumed that the CHP has 33% electricity efficiency and 64% heat recovery efficiency (the fraction of the useful heat recovered after electricity generation). This study assumes that all exported heat and electricity are used completely, displacing conventional heat (generated from a NG boiler with 80% efficiency) and electricity (U.S. average generation mix). While electricity can be transmitted easily, heat or steam must be consumed by nearby plants. Thus, this heat demand assumption could limit the potential location and size of plants or affect the WTWa results significantly when there is not enough heat demand.

Owing to the large amount of coproduced electricity and heat, the WTWa results of the Char-CHP scenario depend highly on the coproduct handling method [17]. The displacement and allocation methods are widely used to handle coproducts in LCA. The displacement method (Char-CHP-Disp) allocates all energy and emission burdens to the main product, while the energy and emissions related to the displaced products are taken as credits. The energy allocation method (Char-CHP-EnAllo) allocates all energy and emission burdens among all products by their energy, mass, or market value shares. Note that while the Char-CHP scenarios coproduce heat and electricity, the char in the Char-LF scenario is not a coproduct but waste, whose energy and emissions burdens are allocated to jet fuel and diesel. For both scenarios, this study used an energy allocation between jet fuel and diesel. To handle heat and electricity in the Char-CHP scenario, this study examined two coproduct handling methods: Char-CHP-Disp and Char-CHP-EnAllo (Figure 5).

The emissions from char combustion in CHP are taken into account. The CHP emission factors of VOCs, CO, CH_4, and N_2O are 1.50, 12.4, 1.06, and 1.59 g/MMBtu of char burned, respectively, which are based on GREET 2015 [6]. Fossil CO_2 emissions of the CHP process are calculated on the basis of C content in char, fossil C ratio, and C contents of VOCs, CO and CH_4. After counting C contents of VOCs, CO, and CH_4, the fossil CO_2 emission factor is 89,743 g/MMBtu char burned in the CHP process. Biogenic CO_2 emissions from the CHP process are treated as carbon-neutral. In this study, we also considered that the CHP process needs makeup and cooling water. However, since we did not have dedicated CHP process water consumption data, we assumed its water use is the same as for a coal-fired power plant (100.3 gal/MMBtu of electricity available at user sites) [6].

3.4.4 Jet Fuel Transportation and Distribution

Figure 6 shows jet fuel transportation and distribution assumptions used in this study [6]. We assumed that jet fuel is produced domestically, so the share of ocean tank transportation is 0%. First, jet fuel is transported by barge (200 miles and 48.5% share), pipeline (100 miles and 46.4% share), and rail (490 miles and 5.1% share); then jet fuel is distributed from a bulk terminal to refueling stations by heavy heavy-duty trucks (30 miles and 100% share). The fuel types used for barge, pipeline, rail, and truck transportation are different, and include residual oil, electricity, and diesel (Figure 6). The direct and indirect (upstream) GHG emissions, energy consumption, and water use of jet fuel transportation and distribution are calculated using the default values from Reference 6.



	Barge	Pipeline	Rail	Truck
Distance to destination (miles)	200	110	490	30
Energy Intensity				
Origin to destination (Btu/ton-mile)	423	404	274	969
Back-haul (Btu/ton-mile)	312	NA	NA	969
Fuel type used	Residual oil	Electricity	Diesel	Diesel

Figure 6. Jet fuel transportation and distribution assumptions used in this study

3.4.5 Jet Fuel Combustion in Aircraft

The aircraft selected for this study is the single-aisle passenger aircraft. Its emission factors for VOC, CO, CH_4, and N_2O are 1.29 $x10^{-2}$, 9.14 $x10^{-2}$, 1.06$x10^{-4}$, and 2.08 $x10^{-4}$ g/MJ of jet fuel, respectively, which are based on GREET 2015. Combustion fossil CO_2 emissions are calculated on the basis of jet fuel LHV (43.20 MJ/kg), total C ratio (86.2 wt%), fossil C ratio, and C contents of VOC, CO and CH_4. The fossil C ratio in jet fuel is dependent on the shares of biomass and coal in feedstocks. When biomass accounts for 15 wt% of feedstocks, the ratio of fossil C/total jet fuel C is 78 wt%. After counting the C contents of VOCs, CO, and CH_4, the fossil CO_2 emission factor is 57 g/MJ jet fuel combusted in the single-aisle passenger aircraft. Biogenic CO_2 emissions from jet fuel combustion are carbon-neutral.

3.4.6 Feedstock GHG Emissions Results (Field to Refinery)

Figure 7 shows GHG emissions of bio-feedstocks (corn stover, switchgrass, and wheat straw) compared to lignite coal; these include emissions from farming and collection; field treatment, drying, handling and storage; biomass densification; and biomass transportation to the refinery gate. Bio-feedstocks have higher GHG emissions (76,664–142,075 gCO_{2e}/ton feedstock) than lignite coal (52,167 gCO_{2e}/ton feedstock). Supplemental fertilizer and chemical usage contributes 45–75% of the GHG emissions, followed by farming and collection (14–28%). Fertilizer production is very energy-intensive and generates high GHG emissions; in addition, a significant amount of N_2O emission (1.525%) is generated once N fertilizer is applied to the soil [6]. Among the bio-feedstocks examined, switchgrass produces the highest GHG emissions, owing mainly to its high N fertilizer demand and related N_2O emissions. In order to produce uniform and dense feedstocks, we assumed that biomass bales are further densified into pellets for the cases examined in this study. As shown in Table 3, compared to the bulk densities of biomass bales (137–169 kg/m^3), the process of densification increased biomass bulk density by a factor of ~4. In addition, the process of densification decreased biomass moisture by 6%. For biomass bales, owing to their relatively low bulk density, the volume capacity (106 m^3) of the truck is the limiting factor for biomass transportation. On the other hand, for biomass pellets, the weight capability (25 tons) of the truck is the limiting factor for biomass transportation. Compared to the transportation of biomass bales, owing to reduction of transportation energy consumption on a per-ton-of-dry-biomass-delivered basis, the transportation of biomass pellets could reduce GHG emissions by 5,267 gCO_{2e}/ton for corn stover and wheat straw and 2,842 gCO_{2e}/ton for switchgrass (Figure 7). The difference in reduction between switchgrass and the other two feedstocks (corn stover and wheat straw) is due to the difference in the biomass bulk densities of bales and pellets (bulk density ratio of pellets/bales: 641/169 = 3.8 for switchgrass, and 587/137 = 4.3 for corn stover and wheat straw). However, the emission burden of the densification process that results from electricity and NG demands is 20,090 gCO_{2e}/ton. From the viewpoint of overall GHG emissions (bales vs pellets: 95,025 vs 107,419 gCO_{2e}/ton for corn stover, 127,877 vs 142,075 gCO_{2e}/ton for switchgrass, and 76,664 vs 89,421 gCO_{2e}/ton for wheat straw), the densification process (bales to pellets) becomes unfavorable, with additional energy inputs (Figure 4). However, biomass densification results should be interpreted with caution, as the underlying densification energy is process-dependent and not fully optimized here. In addition, the energy requirement for biomass transportation is highly location-dependent. Altex proposed several potential sites: western North Dakota, southeast Texas, and eastern Montana (Table 9). Site-specific transportation information (such as distance and truck vs. rail) could be revisited as joint efforts by Altex and its partners continue. In addition, biomass densification could improve biomass logistics and release stringent constraints of collocation of coal and biomass resources, because biomass densification can reduce the cost of transportation and simplify storage and handling infrastructure.

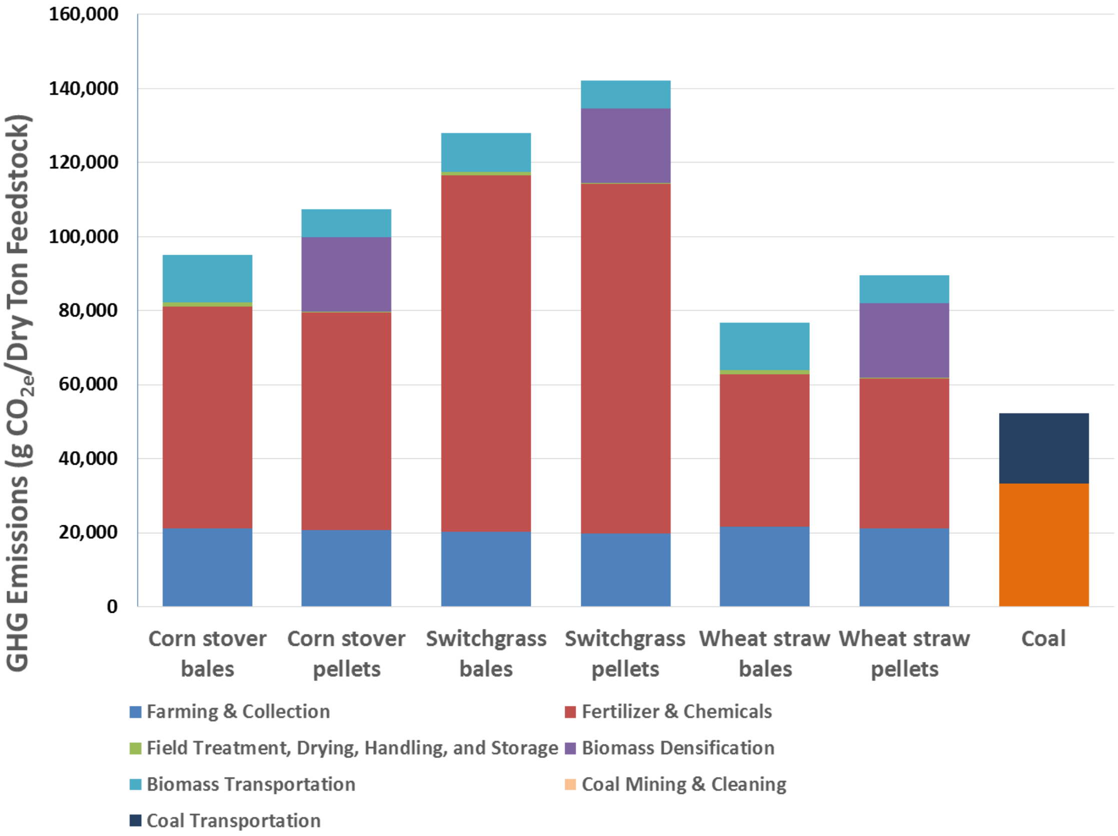


Figure 7. Greenhouse gas (GHG) emissions of feedstock process steps, compared to lignite coal, including farming and collection; field treatment, drying, handling, and storage; biomass densification; and biomass transportation to the refinery gate.

Table 9. Potential sites for the CBTL plant.

Region	Feedstock type	Special characteristics
Western North Dakota	Lignite	Strong State support; biomass is plentiful.
Southeast Texas	Lignite	Forest residues and agriculture residues are plentiful.
Eastern Montana	Sub-bituminous	Huge reserves

3.4.7 Carbon Balance and WTWa GHG Emissions for the Base Case of 85 wt% Coal and 15 wt% Biomass

To simplify the presentation, the following discussion presents results for densified wheat straw only. Similar results for other feedstocks were obtained as well. The share of biomass in feedstocks is an important factor for life-cycle GHG emissions [18]. Altex assumed that 85% of mass was from coal, and 15% of mass was from wheat straw. As shown in Figure 8, through the CBTL process, 13.8% of C is converted into jet fuel, 8.6% of C is converted into No 2 diesel, 30.6% of C is transformed into process CO_2 emissions that are captured by the CO_2 absorber with the efficiency of 95%, and the remaining C (47.0%) is left in the char. The carbon-to-fuel efficiency (including diesel) is 22.4%. Under the scenario of Char-LF, 20% of the C in the char (7.6 g fossil C/MJ and 0.8 g biogenic C/MJ) is emitted as CO_2, and 80% of the C in the char (30.3 g fossil C/MJ and 3.2 g biogenic C/MJ) is sequestrated. The fossil labile C in the char (7.6 g C/MJ) is counted as GHG emissions, and the biogenic stable C in the char (3.2 g C/MJ) takes biogenic CO_2 credit. Thus, the net CO_2 emissions from char disposal are estimated at 15.9 g CO_2e/MJ (or 4.4 g C/MJ). Under the char-CHP scenario (Char-CHP-Disp or Char-CHP-EnAllo), both fossil C (37.9 g C/MJ fuel) and bio-C (4.0 g C/MJ fuel) are combusted. Bio-CO_2 is carbon neutral, and fossil-CO_2 is counted as GHG emissions. Under all scenarios, 95% of the CO_2 emissions from the CBTL process is captured and sequestered. The biogenic CO_2 in the captured CO_2 (1.36 g C/MJ fuel) is counted as carbon credits while 1.29 g C/MJ fuel is taken as GHG emissions. Thus, net GHG emissions from the process emissions after carbon capture is -0.07 g C/MJ. Bio-CO_2 from fuel combustion in aircraft (2.6 gC/MJ fuel) is carbon-neutral (Figure 8).

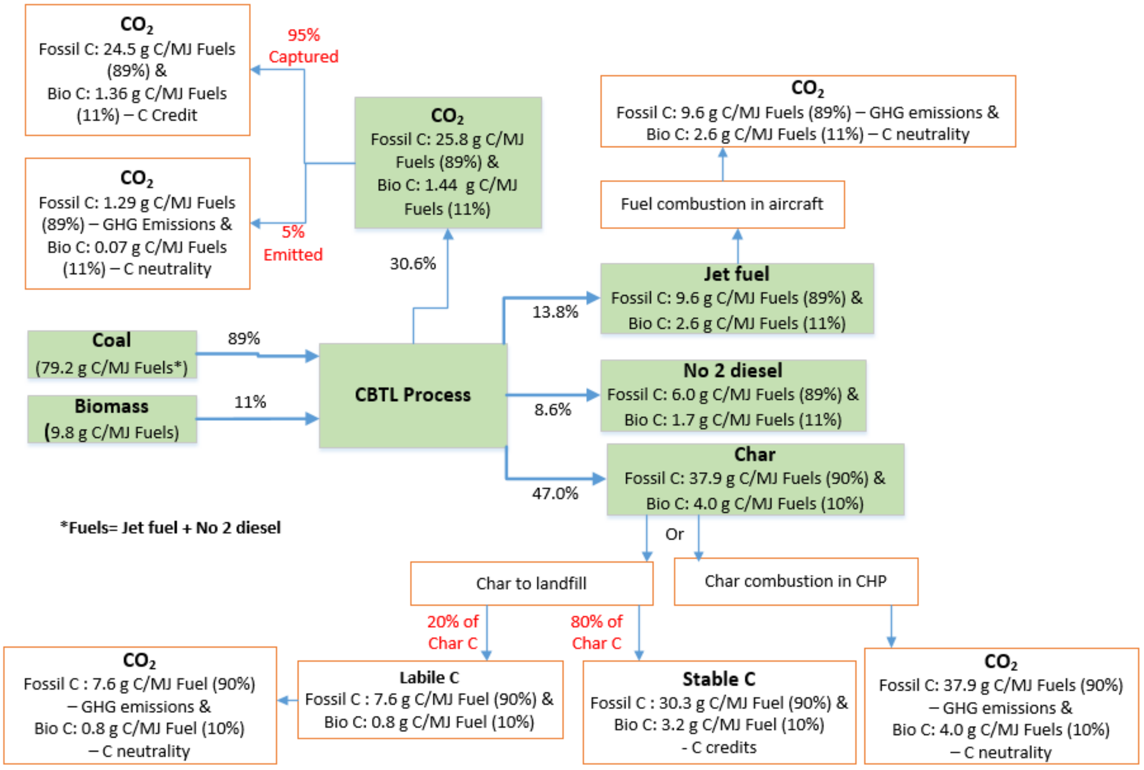


Figure 8. Carbon balance for the case of 85 wt% coal and 15 wt% biomass

The WTWa GHG emissions of jet fuel from the CBTL process, compared to conventional jet fuel, are shown in Figure 9. For conventional jet fuel, GREET evaluates GHG emissions from crude oil recovery and transportation, refining to jet fuel, jet fuel transportation and distribution, and end use [6]. As shown in Figure 9, when biomass accounts for 15% of the total feedstock mass, the jet fuel from the CBTL process has higher WTWa GHG emissions than conventional jet fuel (84 g-CO_2e/MJ). Of the three char application scenarios, Char-CHP-Disp has the lowest GHG emissions (97 g-CO_2e/MJ), and followed by Char-LF. In the Char-CHP-Disp scenario, the combusted fossil C in the CHP contributes 136 g-CO_2e/MJ of GHG emissions, and the displacement credits from excess CHP electricity and heat are 112 g-CO_2e/MJ. When char is sent to a landfill, the processes of char transportation, along with fossil CO_2 from labile carbon, contribute 28 g-CO_2e/MJ of GHG emissions. In the Char-CHP-EnAllo scenario, exported electricity and heat from CHP allocate 50% of the energy and emissions burdens from the processes of feedstocks, fuel process, and CHP. As shown in Figure 9, direct emission from char combustion in the CHP is the largest contributor in the Char-CHP scenario (Char-CHP-Disp or Char-CHP-EnAllo). The net jet fuel combustion (fossil CO_{2e}) contributes 55%, 65% and 46% of WTWa GHG emissions under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. Biomass and coal feedstocks are not major contributors to WTWa GHG emissions, contributing 8.1%, 9.7% and 3.5% of WTWa GHG emissions under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively.

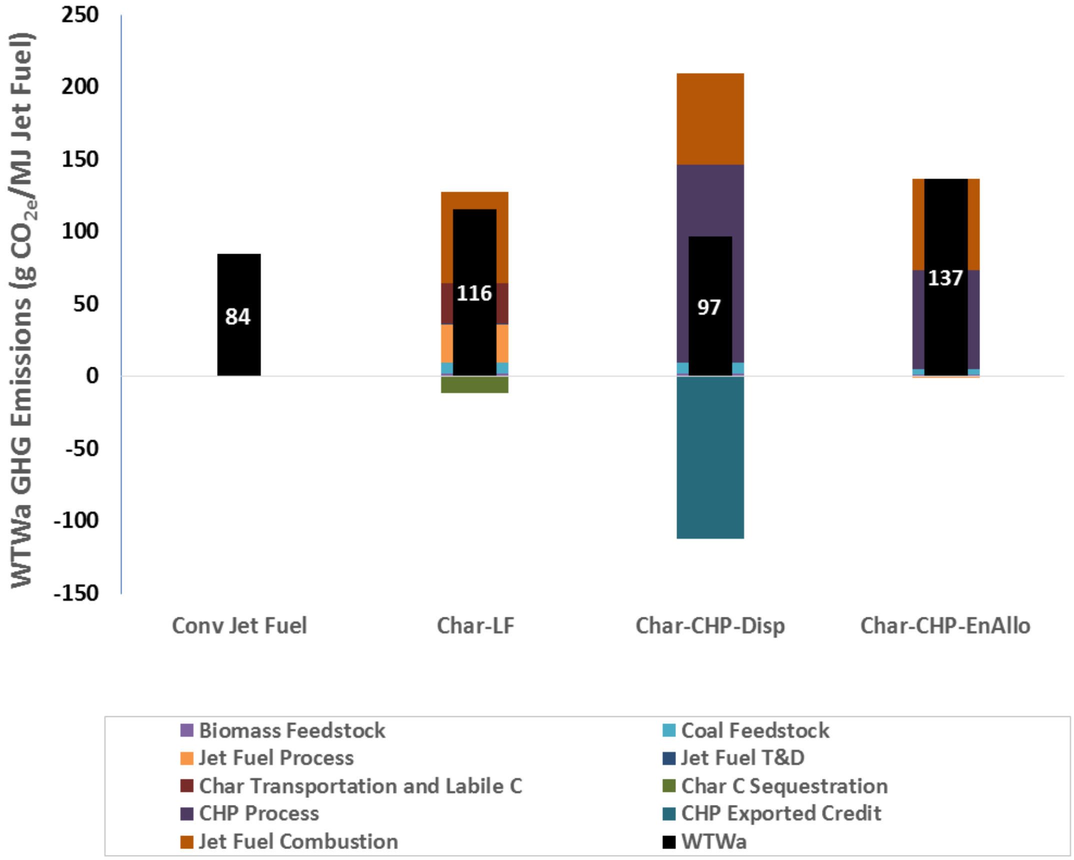


Figure 9. WTWa GHG emissions from 15% wheat straw biomass and 85% lignite coal under three scenarios: char to landfill (Char-LF), char for CHP-displacement (Char-CHP-Disp), and char for CHP-energy allocation (Char-CHP-EnAllo).

3.4.8 WTWa Water Use for the Base Case of 85 wt% Coal and 15 wt% Biomass

As shown in Figure 10, besides impacting GHG emissions, char handling methods also have significant impact on WTWa water use. When char is used for CHP, the CBTL process energy demand is satisfied internally. The net water uses for the CBTL process, which are direct process water demand, are 0.0271 and 0.0137 gal/MJ fuels under the Char-CHP-Disp and Char-CHP-EnAllo scenarios, respectively. On the other hand, under the Char-LF scenario, the net water use for the CBTL process is 0.0623 gal/MJ fuel, which includes both direct process and indirect upstream water demands. Direct water uses in the CHP processes (0.0502 gal/MJ for Char-CHP-Disp and 0.0252 gal/MJ for Char-CHP-EnAllo) are the largest contributors. The water usage of biomass is 0.0020, 0.0020, and 0.0010 gal/MJ under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. The water usage of coal is 0.0074, 0.0074, and 0.0037 gal/MJ under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. The water use credit is -0.133 gal/MJ under the Char-CHP-Disp scenario. Overall, the WTWa water usages are 0.072, -0.046 (water saving), and 0.044 gal/MJ under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. The Char-CHP-Disp and Char-CHP-EnAllo scenarios could achieve 266% and 59% water use reduction compared to conventional jet fuel. The Char-LF scenario increases water use by 161% compared to conventional jet fuel. As shown in Figure 10, under the Char-LF scenario, the fuel process is the major contributor to water use. One good option to minimize freshwater use is to recycle process water.

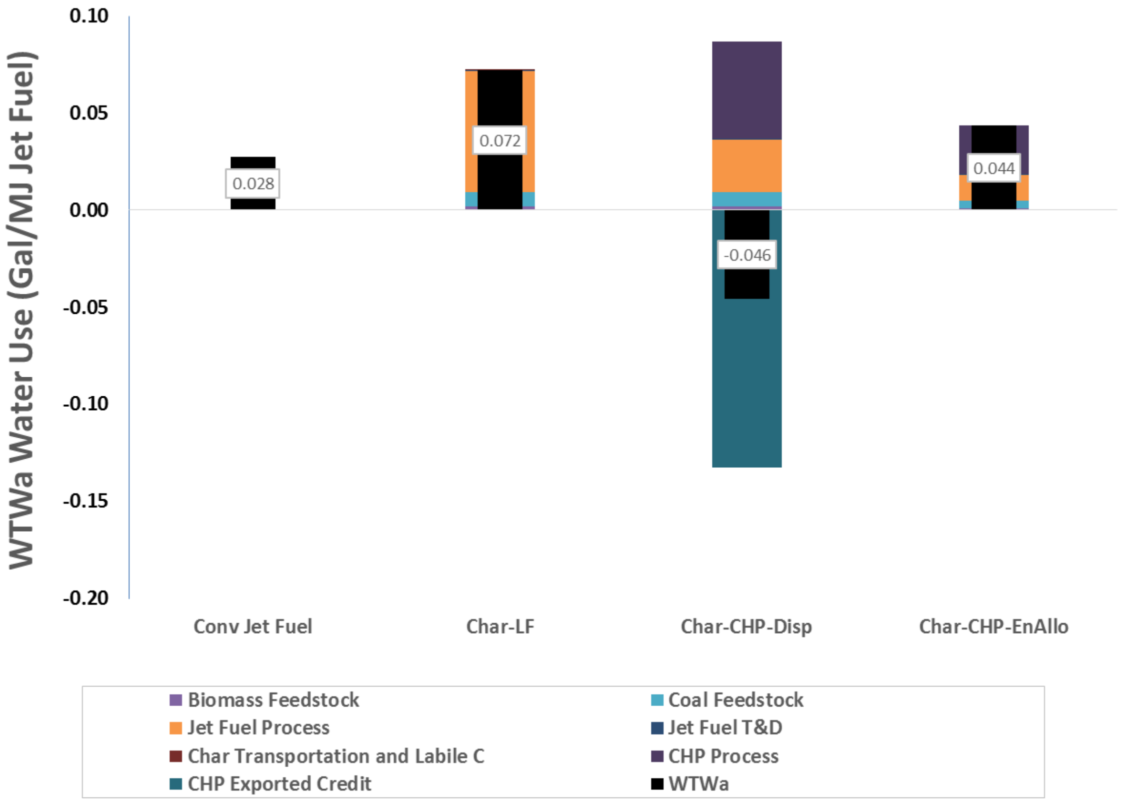


Figure 10. WTWa water use from 15% wheat straw biomass and 85% lignite coal under three scenarios: char to landfill (Char-LF), char for CHP-displacement (Char-CHP-Disp), and char for CHP-energy allocation (Char-CHP-EnAllo).

3.4.9 Paths Forward to 84 g-CO_{2e}/MJ

According to the Energy Independence and Security Act (EISA) of 2007, U.S. Federal agencies cannot enter into contracts for procurement of an alternative fuel that has higher life-cycle GHG emissions than the equivalent conventional fuel [19]. In this study, we did not receive data from Altex related to product yields and optimal process conditions with different biomass feedstock shares. To study biomass share sensitivity and estimate proximate break-even points, we assumed that the yields of jet fuel, diesel, and char are 3.61 MMBtu/ton feedstock, 2.20 MMBtu/ton feedstock, and 0.33 ton/ton feedstock, respectively.

Han *et al.* [16] showed that the product yields (oil, gas, char and water) vary with biomass feedstocks (woody vs. herbaceous), and different feedstocks have different optimal temperatures and pressures for pyrolysis to maximize product yields. Under the assumptions of constant energy demands and fuel yields regardless of the share of biomass and coal feedstocks, the analysis presented below demonstrates a strategy for reducing WTWa GHG emission to 84 gCO_2e/MJ (equivalent to conventional jet fuel) by changing biomass shares (from 0 to 100%) in the feedstock. Figure 11 shows WTWa GHG emissions of 0%, 15%, and 100% biomass under the three scenarios of Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo. Compared to conventional jet fuel (84 gCO_{2e}/MJ), jet fuel from coal alone (0% biomass) increases life-cycle GHG emissions by 70%, 48%, and 84% under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively (Figure 11). Xie *et al.* [18] reported similar results, namely, that FT diesel from coal

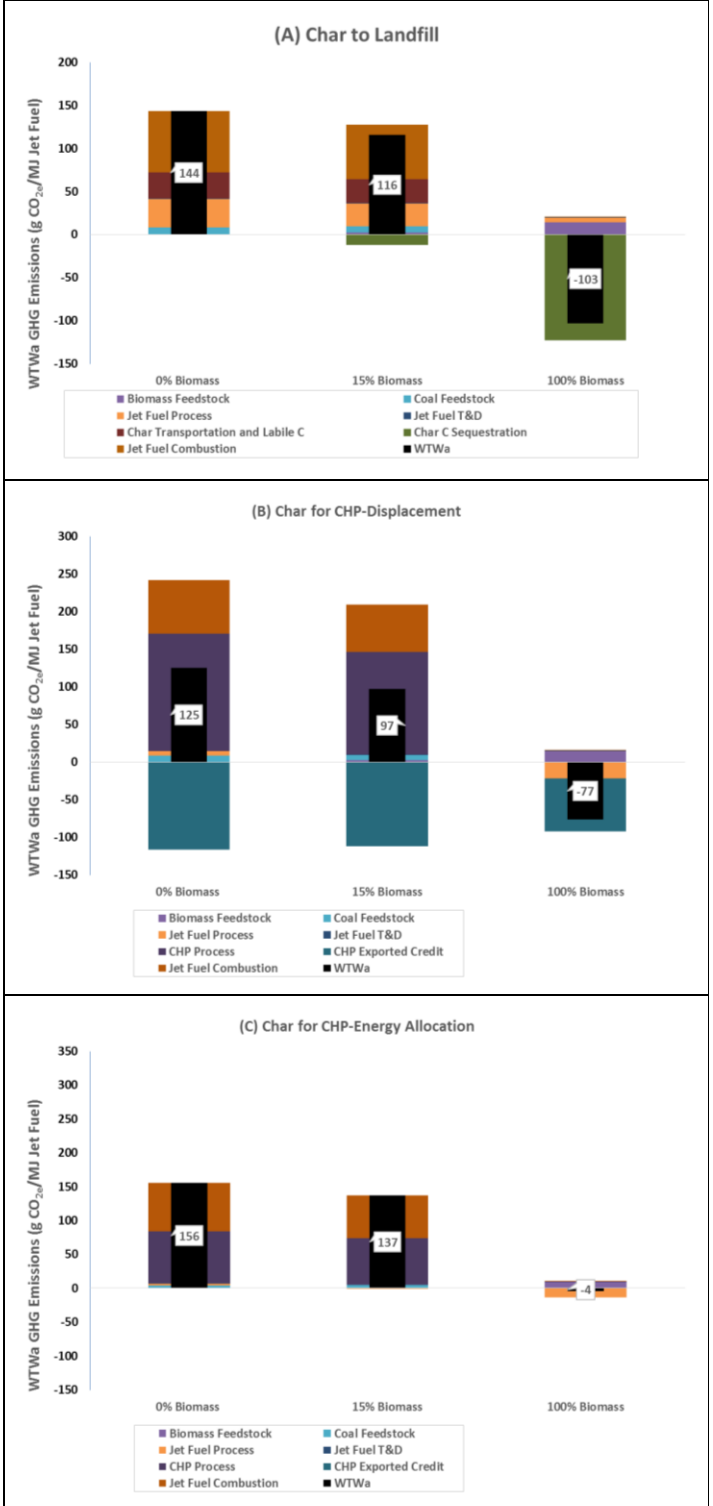


Figure 11. Impact of biomass shares (0%, 15%, and 100%) and char handling on WTWa GHG emissions of jet fuel. (A) char to landfill (Char-LF), (B) char for CHP- displacement (Char-CHP-Disp), and (C) char for CHP-energy allocation (Char-CHP-EnAllo)

increases life-cycle GHG emissions by more than 200% without CCS and 5–29% with CCS, relative to petroleum diesel. When biomass is used as the sole feedstock (100% biomass), jet fuel from the CBTL process decreases life-cycle GHG emissions by 222%, 191%, and 105% under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. In the Char-LF scenario, the process with 100% biomass results in 103 g CO_{2e}/MJ carbon sequestration (negative WTWa emissions). In the Char-CHP-Disp scenario, excess electricity and heat are exported to displace conventional products (U.S. average generation mix and NG), and the displacement credits are over 70 g CO_{2e}/MJ.

Figure 12 shows the break-even point of 84 gCO_{2e}/MJ for the CBTL process under three scenarios.

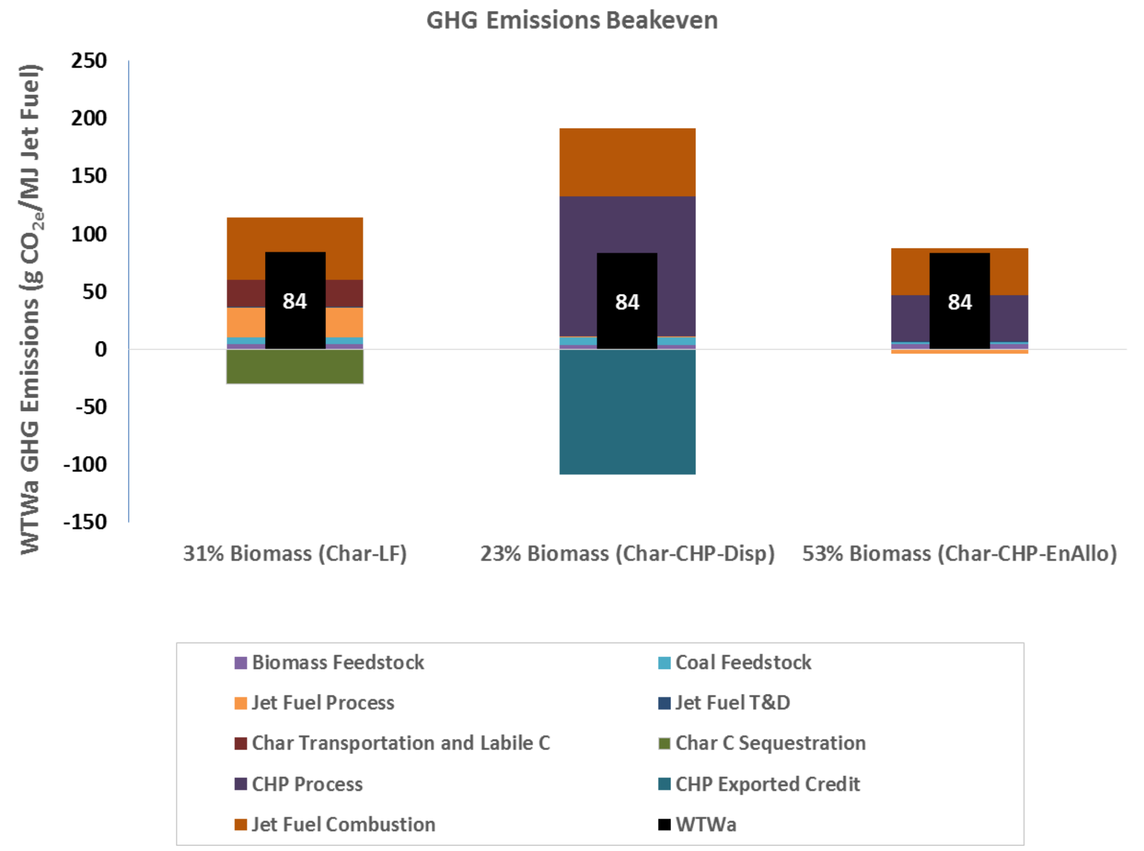


Figure 12. Paths forward to 84 g CO_{2e}/MJ for the combined coal and biomass to liquid fuels (CBTL) pathway under three scenarios: char to landfill (Char-LF), char for CHP-displacement (Char-CHP-Disp), and char for CHP-energy allocation (Char-CHP-EnAllo).

The Char-CHP-Disp scenario needs the lowest biomass supplementation (23%) to achieve WTWa GHG emissions of 84 gCO_{2e}/MJ, and followed by the Char-LF scenario (31%) and the Char-CHP-EnAllo (53%). In the Char-LF scenario, GHG emissions from jet fuel combustion, char transportation-labile C emissions, and fuel processes are 54, 23, and 26 gCO_{2e}/MJ, respectively (Figure 12). In this case, optimizing the fuel process could be a good option to further reduce GHG emissions. In the Char-CHP-Disp scenario, GHG emissions from the CHP process, jet fuel combustion, and CHP exported credits are 122, 59, and -108

gCO_{2e}/MJ, respectively (Figure 12).In the Char-CHP-EnAllo scenario, GHG emissions from CHP process and jet fuel combustion are 41 and 40 gCO_{2e}/MJ, respectively (Figure 12). In both Char-CHP scenarios, a CHP with an integrated CCS could be a good option to further reduce GHG emissions. A caveat is that the displacement approach might be problematic for this study. The large credits and net negative GHG emissions obtained with the displacement method indicate the likelihood of distorted results [17].

3.4.10 LCA Conclusions

The present results evaluate life-cycle GHG emissions and water usage of combined coal and biomass conversion to liquid fuels. Among corn stover, switchgrass and wheat straw, switchgrass has the greatest GHG emissions (per dry ton feedstock), which are mainly due to its high N fertilizer demand and related N_2O emissions. From the viewpoint of overall GHG emissions (per dry ton feedstock), the densification process (bales to pellets) becomes unfavorable, with additional biomass process energy demand. When the feedstock contains 15 wt% densified wheat straw and 85 wt% lignite coal, compared to conventional jet fuel at 84 gCO_{2e}/MJ, WTWa GHG emissions are 116, 97, and 137 gCO_{2e}/MJ under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. WTWa water consumption is 0.072, -0.046, and 0.044 gal/MJ for Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo, respectively (compared to conventional jet fuel at 0.028 gal/MJ). The fuel process is the major contributor to the water usage, so to reduce WTWa water usage, one good option is to recycle the fuel process water and minimize freshwater use. To reach the break-even point of 84 gCO_{2e}/MJ (emissions for petroleum fuels) and comply with the 2007 EISA, Section 526, "*Procurement and Acquisition of Alternative Fuels*," under the assumptions of constant product yields and energy demands regardless of the share of biomass and coal feedstocks, 31 wt%, 23 wt%, and 53 wt% of the feedstock blend needs to be biomass under the Char-LF, Char-CHP-Disp, and Char-CHP-EnAllo scenarios, respectively. A caveat is that, owing to data source limitations, the assumptions of energy usage and product yields for different shares did not represent real conditions, and future improvement to reduce uncertainties is needed. Another issue is that the displacement approach might be problematic for this study. The large credits obtained with the displacement method indicate the likelihood of distorted results. Moving forward, development efforts to further reduce GHG emissions should focus on CCS integration, effective coproduct utilization, fuel process energy reduction and yield improvement, and feedstock production and logistics optimization.

4 GHGR-CBTL SYSTEM DESIGN

The design approach was to produce the test system process flow diagram (PFD) that was based on a full-scale system PFD. To produce the PFD, CHEMCAD was used to define the process flows and conditions. The unit operations in the PFD were added, along with ancillary equipment, to a bill of materials (BOM). The BOM was a living document that lists all the equipment and their specs or part numbers as they are defined. As needed the BOM points to the spec sheets that are used to interact with the suppliers to finalize the equipment selection. The equipment in the BOM were classified as COTS or custom design. As COTS equipment were specified, custom equipment designed, and control-system strategies are developed. This information was combined with the PFD to produce a P&ID and update the BOM.

The specific system design and equipment designs were based on TRL-3 and TRL-4 system data, engineering analysis (including CHEMCAD simulations), and Unitel inputs. Unitel, who has been involved in the design or fabrication of over 1300 pilot plant and demo plant projects, was a Subcontractor to Altex under the project.

5 TRL-5 GHGR-CBTL PROCESS EQUIPMENT BUILDUP

The objective of this activity was to build the TRL-5 system based upon the design developed above.

5.1 Component Procurement

5.1.1 System P&IDs

Following completion of the design based upon the PFDs, the P&IDs for each section were produced. The P&IDs were shared with Unitel for feedback and updates. The sections have been reviewed and updated, based on feedback from the main Unitel Engineer who is working on this project. As discussions with vendors proceeded, the P&ID and BOM were updated to reflect the selection of particular models of equipment and vendor suggestions on operation and control of their equipment. A final review was completed with senior Altex and Unitel staff to cover system operation, safety, what-ifs, risk, and scenarios that may occur during testing. Risk-mitigation strategies were developed either through engineering controls in the P&ID or through procedural controls.

The P&ID is broken into similar sections as discussed above. The P&ID includes the equipment, instruments, and control loops. It also identifies piping materials, sizes, connection types, and services for all of the lines. This information was used to update the BOM and the spec sheets as needed.

The P&ID was linked to the CHEMCAD heat and mass balance. This was then used along with CHEMCAD’s pipe-sizing tools to verify that the selected pipe materials, diameters, and thicknesses were appropriate for each line. This was accomplished by exporting the piping information to a line list in Excel. Then the fluid properties were used to verify that the selected pipe conformed to ASME B31.3 standards. The Excel worksheet verified that the design temperature and pressures were appropriate for the fluid properties and the ASME B31.3 fluid category was identified. The pipe material allowable stresses were interpolated automatically from data from the ASME B31.3 property tables.

5.1.2 Bill of Materials and Spec Sheets

The results from the generation of the PFD and from the subsequent equipment sizing were used to produce a Bill of Materials (BOM) for all of the major equipment needed for the system. After identifying these equipment in the BOM, equipment was divided into three categories: 1) Commercial Off-the-Shelf (COTS) equipment; 2) commercially available equipment, but with long lead times; and 3) equipment that requires custom design and/or fabrication. This allowed for the prioritization of the equipment acquisition by focusing on long-lead items first.

Spec sheets were developed for the equipment in the BOM associated with the latter two categories. This allowed for communication with equipment vendors and reps to obtain quotes for these items.

5.1.3 Overall GHGR Equipment Layout

A 765-ft^{2}room in the Altex facility was dedicated to the GHGR system. This room is approximately19 feet wide, 41 feet long, and 16 feet high. A 3-D CAD model of the layout of the equipment in this space was generated.

The CAD model of the room was then used to produce plan and elevation drawings. These drawings were then sent to the ventilation contractor, electrical contractor, architect, and HazMat engineers so they could conduct their analyses, design and layout the components of their respective systems, and prepare drawings for city permitting. throughout this process, the CAD model and the plan and elevation drawings were updated as feedback was received from the various contractors and consultants.

5.1.4 Fire Safety and Hazmat Safety

In addition to the mechanical and electrical layouts discussed above, fire-safety and hazmat-handling assessments were conducted.

5.1.4.1 Building Occupancies

The first item considered under the fire-safety and hazmat-handling assessments was determining if Altex may exceed the Maximum Allowable Quantities (MAQs) of hazardous materials for the building occupancy codes for which the building has already been permitted. All of the equipment with the capacity to store hazardous materials were identified and the type of fluid (liquid or gas) each piece of equipment held was identified. The fluids were then categorized according to type and degree of hazard (e.g., flammability, toxicity, corrosivity). To determine these classifications, pH, flash points, autoignition temperatures (AITs), and OSHA exposure limits were utilized. AITs were estimated based both on the lowest AIT of the components greater than 5 vol% and on the practices recommended in API Recommended Practice 2216. Flash points were estimated and for gases utilizing CHEMCADs built-in estimator. These were then totaled up to determine if any of the MAQs would be violated. After this analysis it was determined that existing MAQs would not be exceeded.

In addition to fluids, solids are processed by the system. Therefore, the properties of these materials (coal, biomass, and char) were examined to determine if they would be considered unclassified, classified as Class II (flammable solids), or classified as Class III (flammable fibers). It was determined that the coal and char needed to be considered as Class II materials while the biomass needed to be considered as a Class III material. The practical upshots of these classifications were to 1) limit the amount of biomass to be present in the facility to 200 ft^3 at a time (100 ft^3 limit for each of two classified areas in the Altex facility – operation and storage), and 2) classify the areas handling these solids as Class II and Class III.

5.1.4.2 Hazardous-Area Classifications

Because the ventilation system was designed to meet the adequate-ventilation requirements of the California Fire Code and the system and procedures were designed to minimize spills and leaks of gases, liquids, fibers, or dust, the majority of the room in which the GHGR system would be located would be classified as Division 2.

More specifically, the sweep of ventilation from the fluid-processing side of the room to the solids-processing side of the room allows for the areas in the fluid-processing side of the room to remain unclassified or solely Class I. As the ventilation moves from the fluid-processing side of the room to the solids-processing side of the room, areas in the solids-processing side of the room are classified as Class I in addition to being classified as Class II and Class III. Guides for classifying areas are provided for flammable liquids in Figure 13 and for flammable gases in Figure 14.

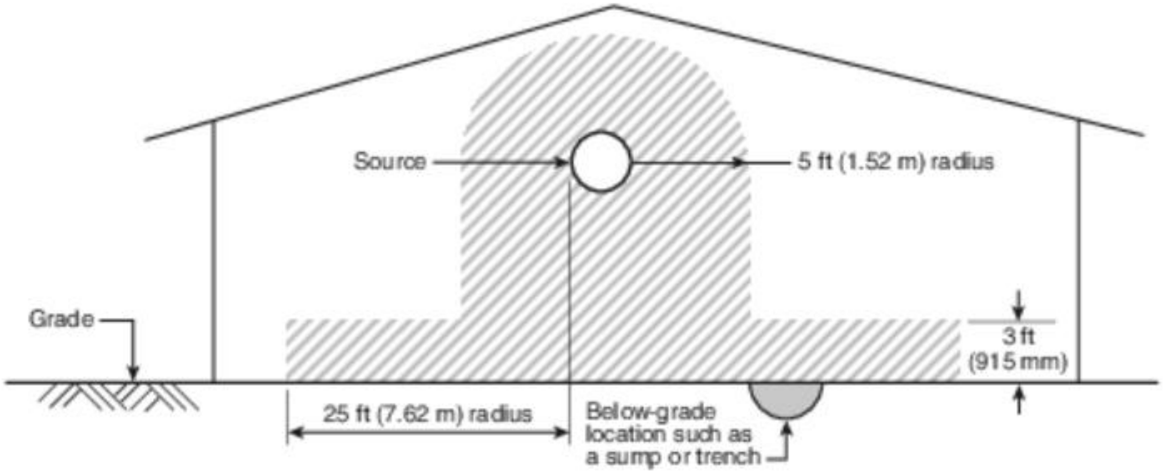
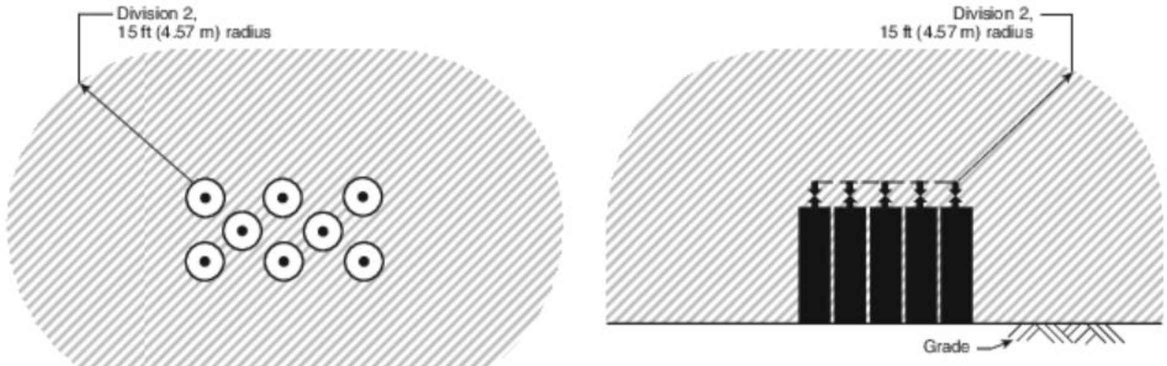


Figure 13. NFPA 497 area classifications for leak of flammable liquids with vapors that are heavier than air indoors with adequate ventilation. Shaded area is Class I, Div. 1, hashed area is Class I, Div. 2.



Condition: Outdoor or indoor well-ventilated location, secured in vertical position, in use in single or manifold configuration.

FIGURE 5.10.15 Compressed Gas Cylinders (lighter than or equal to air, including hydrogen).

Figure 14. NFPA 497 area classifications for leak of lighter-than air or neutrally buoyant gases indoors with adequate ventilation. If the gases are heavier than air, then the entire floor area up to 3 feet high is also Div. 2 as in Figure 13. Hashed area is Class I, Div. 2.

The room in which the system will be located meets the adequate-ventilation requirement, the room contains no sump areas, and the system is designed to have no regular opening or venting of flammable fluids. Therefore, the classifications in Figure 13 and Figure 14 apply. Since there will be flammable liquids and also flammable gases both heavier and lighter than air volumes of radius 5 or 15 feet as appropriate around potential leak points and the volume area up to 3 feet above the floor will be classified as Class I, Division 2. Any remaining areas in the fluid-processing side of the room will be considered unclassified. Based upon the AITs estimated as described above, all streams contain Group C or D mixtures. Therefore all electrical equipment to be placed in the Class I, Div. 2 areas will be rated for Class I, Div.2, Group C. Motors will be sourced as non-sparking (AC induction) TEFC with temperature ratings equivalent to T2.

Because Figure 13 and Figure 14 from NFPA 497 provide recommendations on hazardous-are classification, the independent Chemical Engineering California PE was examining some possible additional area classifications based upon a review of the process. For example, if hydrogen-rich gas had the potential to leak from a specific point in the system during sampling, additional Class I, Div. 1 and/or Class I, Div. 2 areas may be identified (*e.g.*, Class I, Div. 1 in a 3-foot radius from a point source of hydrogen gas). Final area classifications were to be signed off by the independent Chemical Engineering California PE.

5.1.5 Seismic Safety

Because Altex is located in the San Francisco Bay Area, seismic safety is of significant importance. In fact, 3 identified faults capable of producing large earthquakes are within 15 miles of the Altex facility (San Andreas, Hayward, and Calaveras faults). Therefore, in keeping with good engineering practice as well as city permitting requirements, components weighing over 400 lb were designed with seismic-safety constraints. All of these component designs underwent seismic-safety review by a qualified California PE authorized to stamp these drawings. A few items required modifications, but because vendors were notified ahead of time of the seismic requirements, there was no significant changes.

5.1.6 Process-Hazard Assessment

Given the small scale of the process and the types of hazards present, a What-If Analysis was undertaken to identify potential hazards. This was done in a three-step process:

1. Evaluation of options by Altex and Unitel while preparing the P&IDs,
2. A final review of the P&IDs by additional "new eyes" at Unitel and Altex, and
3. An independent review by the California Chemical Engineering PE.

The first two steps were described in Section 5.1.1.

The independent review by the CA PE was nearly, but not completely finished. This review also conducted as a "what-if" analysis. The system was broken down into portions of the system that could be isolated from the other parts of the system by design or by accident or upset condition. These portions of the system were analyzed and then interactions with the other portions of the system were analyzed. A matrix of specific conditions and equipment operational status (*e.g.*, Valve XV-500 is open or blower F-500-B fails) was produced that listed the hazards present in that system under those conditions, what mitigation methods are included in the P&ID, what procedural mitigation methods may be required, and suggestions for modifications to the P&ID to eliminate or further mitigate the identified hazards.

5.2 System Buildup

After preparing the P&IDs, BOM, and spec sheets, equipment was procured. This process included issuing purchase orders (POs), approving equipment drawings (where necessary), and receiving the equipment.

However, because of cost overruns due to higher-than-expected equipment costs and additional labor required for the evaluating the refrigeration and compression options, the DOE instructed Altex to stop system fabrication and cancel all outstanding orders. This was to enable the focus of the remaining project funds on producing the TEA final product. As such, some the key system components were received, some were canceled after issuance of a PO for a fee for work to date, and some were canceled before a PO was issued. The majority of the key components were received with several others canceled before delivery. Charges were incurred for POs for which stop-work orders were issued.

6 PRODUCTION OF CATALYST FOR TESTING

To produce the catalyst needed for the TRL 5 test system two paths are planned to minimize risk. In the first path, Altex-in-house-catalyst from the TRL 3 system is being improved by PSU. This heterogeneous catalyst is improved to provide; high olefin conversion, and liquid yield, higher JP-8 selectivity, easy regeneration process and catalyst recovery, and low operational pressure and temperature. PSU's catalyst improvement is expected to significantly increase the liquid-product yield, especially JP-8 yield, at lower pressures in comparison to commercial catalysts that do not produce liquid-product at lower pressures. The second path involves benchmarking the commercial catalysts. The discussion below summarizes the successful completion of these activities on both fronts and the availability of GHGR-TRL5 system catalysts for oligomerization.

6.1 Benchmarking Commercial Catalysts

Benchmarking commercial catalysts activity at Altex involved testing the performance of three heterogeneous catalysts at comparable operational conditions as those of Altex-in-house catalysts. Ethylene (C_2H_4) and propylene (C_3H_6) were oligomerized in a fixed-bed reactor at temperatures of 240 to 350 °C, pressures of 0 to 175 psi, and WHSV's of 1 to 10 hr^{-1}. All catalysts were tested to determine their selectivity to specific olefins, liquid-fuel yield, carbon-number distribution of the liquid-fuel using the ASTM D2887 Simulated Distillation, and olefin conversion.

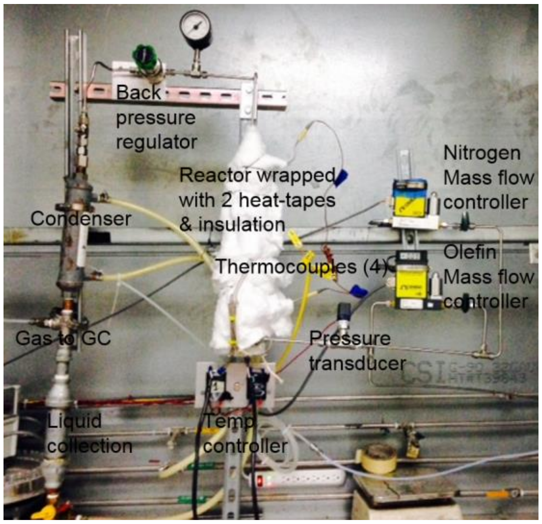


Figure 15. Automated Oligomerization Setup.

The GHGR oligomerization automated test-set-up is shown in Figure 15. It includes; two heat tapes for heating the reactor, two proportional relays to control the firing rate of heat tapes, four thermocouples to measure reactor bottom and top temperatures as well as surface temperature beneath each heat tape, a back pressure regulator to adjust the reactor pressure, and two mass flow controllers to control the flow of olefins and nitrogen gases. Reactor's bottom and top temperatures are used to control the firing rates of the heat tapes on the lower and upper sections of the reactor respectively. A PLC controls the temperature. For user interface (UI) a lab view program is used. The operator sets the desired flow rates and temperature set point for the reactor through UI. The pressure setting is done manually by adjusting the back pressure regulator valve.

Benchmarking oligomerization tests showed that Catalyst 2 was specific to C2 olefin and Catalyst 1 was specific to C3 olefin. Catalyst 3 performed with both C2 and C3 olefins. Table 10 compares the oligomerization test results between the commercial and Altex in-house (PSU) catalysts for C2 and C3 olefins. Catalyst 2 and Catalyst 1 did not perform at 0 psi resulting in no liquid-fuel production. Catalyst 3 gave 10% liquid yield and >17% conversion at 0 psi with 100% C2-olefins, while up to 28% liquid yield and 60% conversion at 170 psi with 20% C3-olefin feed. As shown in Table 10, Altex in-house (PSU) catalysts gave up to 100% conversion of C2 and C3 olefins and liquid yield >20% at 0 psi for both C2 and C3 olefins. With an increase in pressure from 0 to 115 psi, the commercial catalyst, Catalyst 2 gave 52% liquid yield and 84%

C2 conversion. In comparison, liquid yield of 65% and conversion of 100% was obtained with the pressure increase from 0 to 50 psi with PSU catalyst for C2 olefin. Similarly shown in Table 10, the commercial catalyst, Catalyst 1 gave 66% conversion and 50% liquid yield for higher pressures of 120 to 170 psi with C3 olefin. In comparison, PSU catalyst gave 100% conversion and >20% liquid yield at atm pressure with C3 olefin.

Table 10 Comparison between commercial and Altex in-house catalysts.

Catalyst	Olefin Feed	Temperature (°C)	Pressure (psig)	WHSV (h^{-1})	Liquid Yield (%)	Conversion (%)
Catalyst 1	Ethlyene	300-350	0-200	1-10	10-70	17-97
Catalyst 1	Propylene	300-350	0-175	1-10	15-28	30-60
Catalyst 2	Ethlyene	240-290	0-115	1	0-52	84
Catalyst 3	Propylene	240-290	0-175	5	0-50	66
PSU Catalyst	Ethlyene	250-350	0-100	1-24	20-65	50-100
PSU Catalyst	Propylene	250-350	0-50	12	20-45	97-100

C2-olefin was tested on Catalyst 3. WHSV and pressure were varied to see their effect on fuel distribution as shown on Figure 16 and Figure 17. With increasing WHSV from 3.6 to 10 h^{-1} (25 psi and 350 ºC constant-pressure-and-temperature), JP-8 composition in the liquid-product decreased from 57% to 40% and gasoline composition increased from 43% to 60%. Therefore, lower WHSV was favored for JP-8 production with Catalyst 3 as the olefins have higher selectivity to the catalyst at lower WHSV producing heavier hydrocarbon range fuel, JP-8 versus gasoline at higher WHSV.

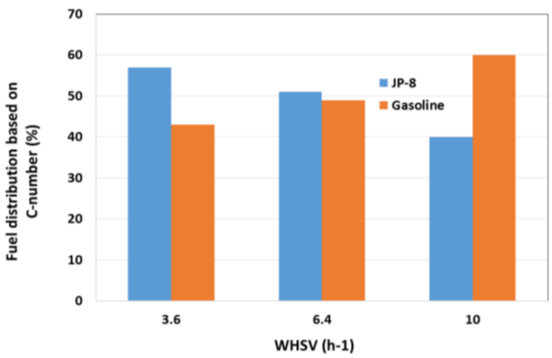


Figure 16. Effect of WHSV on fuel distribution using commercial Catalyst 3.

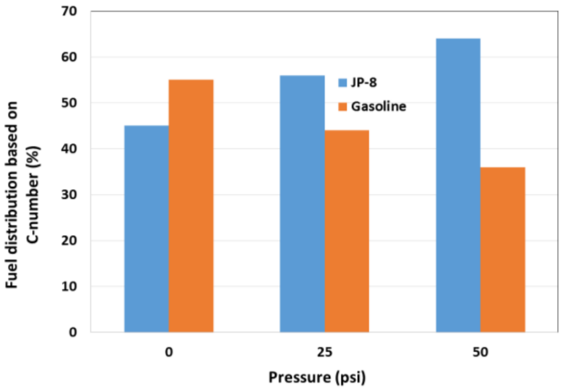


Figure 17. Effect of Pressure on fuel distribution using commercial Catalyst 3.

As shown in Figure 17, with increasing pressure from 0 to 50 psi (3.6 h^{-1} and 350 ºC constant-WHSV-and-temperature), JP-8 composition in the liquid-product increased from 45% to 64% and gasoline composition decreased from 55% to 36%. Therefore, higher pressure was favored for JP-8 production with Catalyst 3 as the olefins have higher selectivity to the catalyst at higher pressure producing heavier hydrocarbon range fuel, JP-8 versus gasoline at lower pressure.

These results show that Catalyst 3 is effective in converting olefins to liquid fuel in JP-8 ranges but require operations at much higher pressure (>50 psi) than PSU-developed catalysts (0 psi).

C2-olefin was also tested on Catalyst 2, which is a C2-olefin-specific non-conventional-oligomerization catalyst. Figure 18 and Figure 19 shows the Catalyst 2 test results. As shown in Figure 18, the catalyst was tested at four different pressures (100, 120, 150, 180 psig) while maintaining the temperature at 350 °C and WHSV at 4.2 hr^{-1} to determine the effect of pressure on liquid yield, JP-8 fraction in the fuel, and the JP-8 yield. With increase in pressure from 100 to 180 psig, liquid yield increased from 91% to 94%, JP-8 fraction in the fuel decreased from 68% to 55%, and combining the two factors the JP-8 yield decreased from 63% to 51%. Therefore, lower pressures of 100 to 120 psig favored JP-8 yield of >60%, and higher pressures of 150 to 180 psig favored JP-8 yield of 52 to 57% as shown in Figure 18. These results have optimized JP-8 fuel production over a low operational-pressure and as a result also reduced the TRL 5 system operational cost.

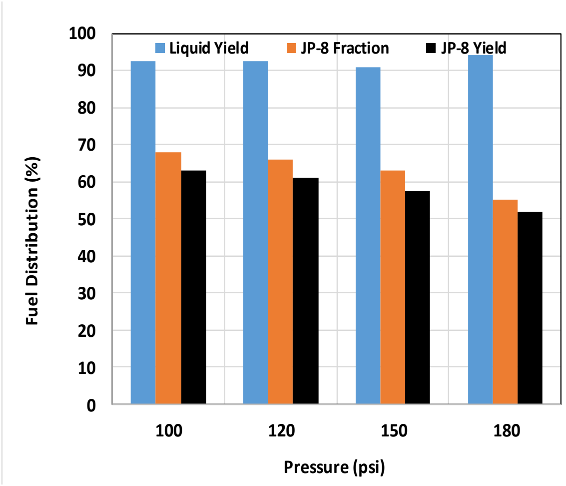


Figure 18. JP-8 production using Catalyst 2 at pressures of 100 through 180 psig, 4.2 h^{-1} WHSV, 350 °C with 12%C2-olefin-feed.

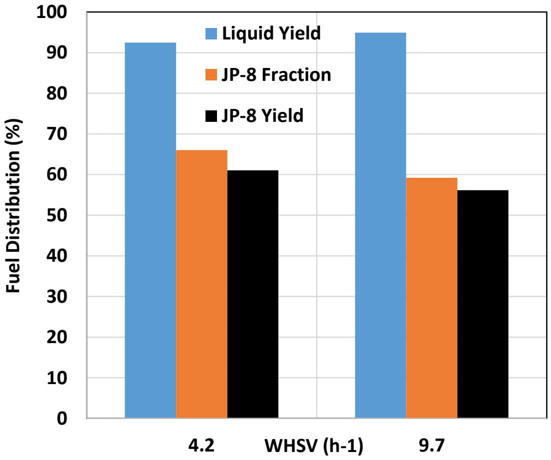


Figure 19. JP-8 production using Catalyst 2 at WHSV of 4.2 and 9.7 h^{-1}, 120 psig, 350 °C with 12%C2-olefin-feed

The tests performed on the Catalyst 2 at WHSV of 4.2 and 9.7 hr^{-1} while maintaining the temperature at 350 °C and pressure at 120 psig to determine the effect of WHSV on liquid yield, JP-8 fraction in the fuel, and the JP-8 yield are shown in Figure 19. With increase in WHSV, liquid yield increased from 92% to 95%, JP-8 fraction in the fuel decreased from 66% to 59%, and combining the two factors the JP-8 yield decreased from 61% to 56% as shown in Figure 19. Both WHSV are within the TRL 5 oligomerization reactor-design basis, and is optimized for JP-8 production. However, lower WHSV of 4.2 hr^{-1} favored JP-8 yield of >60% and higher WHSV of 9.7 hr^{-1} favored JP-8 yield of <60% as shown in Figure 19. These results allow optimized JP-8 production for the TRL 5 system-design.

The Catalyst 2 was also tested at two different temperatures of 350 and 450 °C while maintaining the pressure at 180 psig and WHSV at 7 hr^{-1} to determine the effect of temperature on liquid yield, JP-8 fraction in the fuel, and the JP-8 yield as shown in Figure 20. With increase in temperature from 350 to 450 °C, liquid yield dropped from 90% to 20%, JP-8 fraction in the fuel increased from 62% to 90%, and combining the two factors the JP-8 yield decreased from 57% to 19% as shown in Figure 20. Therefore, low temperature of 350°C favored high JP-8 yield (>55%) over high temperature of 450 °C drastically lowered the JP-8 yield (<20%) as shown in Figure 20. These results indicate that when using Catalyst 2 at 180 psi, the lower 350°C temperature is preferred. Operating the reactor at lower temperature also reduces the energy need.

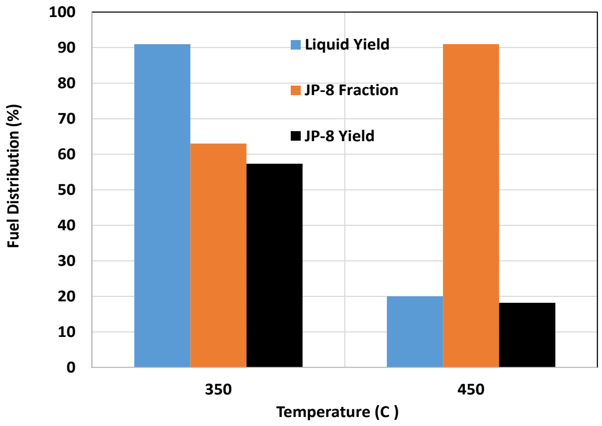


Figure 20 JP-8 production using Catalyst 2 at 350 °C and 450 °C, 180 psig, 7 hr^{-1} WHSV with 12%C2-olefin-feed

Further C3-olefin was tested on Catalyst 3 and Catalyst 1-that is a C3-olefin-specific conventional-oligomerization catalysts. Figure 21 and Figure 22 shows the JP-8 fraction in the liquid fuel from C3 oligomerization, using Catalyst 3 and Catalyst 1, respectively. Both catalysts were tested at three pressures (100, 150, 180 psig) and five WHSV's (1.5 through 10 h^{-1}), while maintaining the temperature at 450 °C. Catalyst 1 (Figure 22) produced higher JP-8 fraction (76% to 87%) than Catalyst 3 (Figure 21) (4 to 72%) at a range of pressures and WHSV's. As shown in Figure 21, using Catalyst 3, 3 different combinations of pressure (150, 100, 180 psig) and respective WHSV (3.6, 5.5, 7 h^{-1}) gave >70% JP-8 fraction while 12 other conditions gave between 4 and 65% JP-8 fractions. On the other hand, as shown in Figure 22, using Catalyst 1, all 15 different conditions produced over 70% JP-8 fraction, giving the flexibility of operating the oligomerization reactor at a wide range of pressures and WHSV. This flexibility makes Catalyst 1 a better choice than Catalyst 3 for JP-8 production in the TRL 5 system.

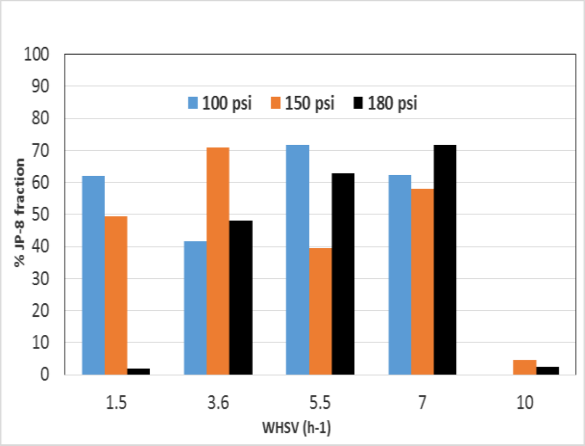


Figure 21. JP-8 fraction using Catalyst 3 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, and T: 450°C.

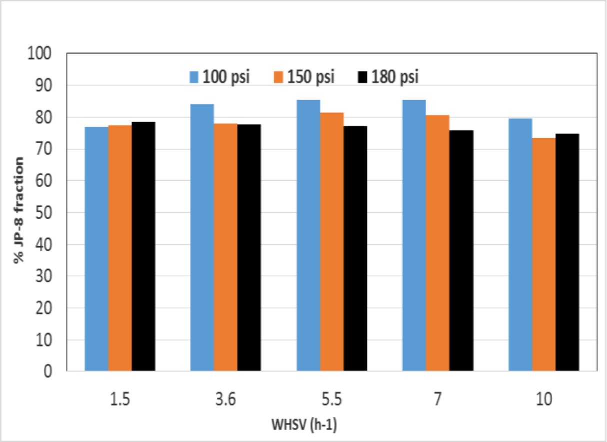


Figure 22. JP-8 fraction using Catalyst 1 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, and T: 450°C.

The corresponding liquid yield for the Figure 21 and Figure 22 data are presented in Figure 23 and Figure 24, respectively. Consistent with the JP-8 fractions, Catalyst 1 produced higher liquid yields (20 to 83%) than Catalyst 3 (5 to 37%) As shown in Figure 23, using Catalyst 3 for oligomerization, 2 different combinations of pressure and WHSV (both 150 and 180 psig at 1.5 h^{-1} WHSV) gave >35% liquid yield while 12 other conditions gave 5-28% liquid. On the other hand, as shown in Figure 24, using Catalyst 1, 2 of the 15 different conditions (both 100 and 150 psig at 1.5 h^{-1} WHSV) gave 70-82% liquid yield and others gave between 19-50% liquid. Therefore, Catalyst 1 performed significantly better than Catalyst 3 for liquid fuel production under the same conditions.

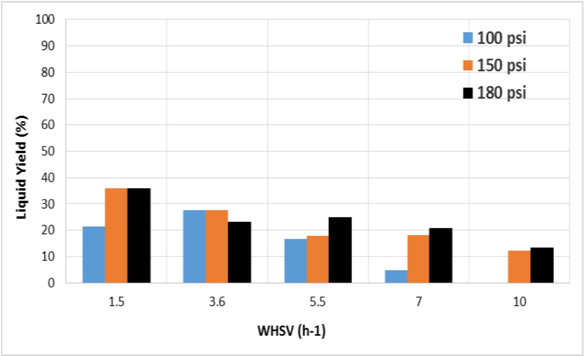


Figure 23. Liquid yield using Catalyst 3 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, and T: 450°C.

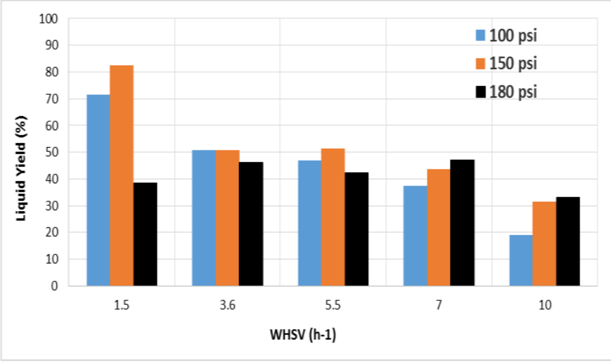


Figure 24 Liquid yield using Catalyst 1 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, T: 450°C.

The JP-8 fractions and liquid yields data presented above have been combined to produce the JP-8 yield and the results are presented in Figure 25 and Figure 26. Again, by comparing these figures it is concluded that Catalyst 1 produces higher JP-8 yield (16% to 65%) than Catalyst 3 (1% to 20%). As shown in Figure 25, using Catalyst 3, 3 different combinations of pressure (150, 150, 180 psig) and WHSV (1.5, 3.6, 5.5 h^{-1}) gave >15% JP-8 yield while 12 other conditions gave between 1 and 14% JP-8 yield. On the other hand, as shown in Figure 26, using Catalyst 1 for oligomerization, 2 of the best conditions gave 55 to 65% JP-8 yield (1.5 h^{-1} at 100 and 150 psig) and other varied between 16 to 43% JP-8 yield. Therefore, Catalyst 1 performed significantly better than Catalyst 3 for JP-8 production under the same conditions.

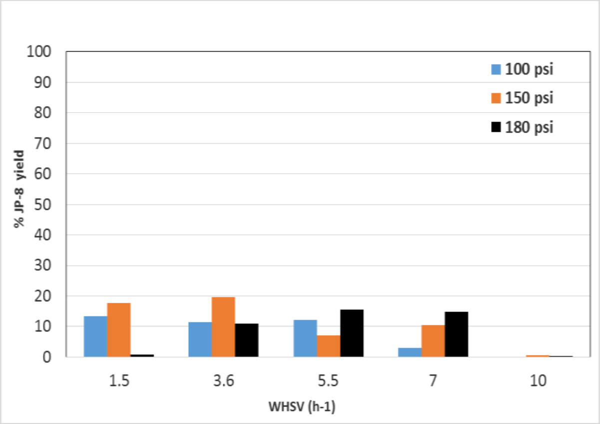


Figure 25. JP-8 yield using Catalyst 3 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, and T: 450 °C

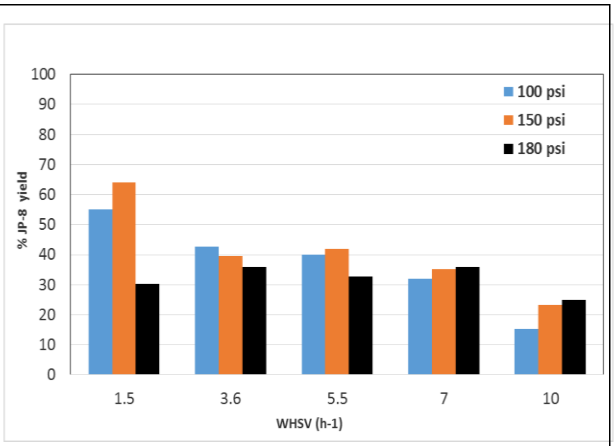


Figure 26. JP-8 yield using Catalyst 1 from C3 oligomerization at pressures of 100 to 180 psig, WHSV: 1.5 to 10 h^{-1}, and T: 450 °C

To explore the effect of temperature, testing was performed at 350 °C while maintaining the pressure at 180 psig and WHSV at 5 h^{-1}. Results are presented in Figure 27 and Figure 28 that show the JP-8 fraction, Liquid yield and JP-8 yield using Catalyst 3 and Catalyst 1, respectively. As shown, Catalyst 1 (Figure 28) produces higher JP-8 yield than Catalyst 3 (Figure 27) at both temperatures. For HZSM, JP-8 yield was 7% to 15% at 350 °C and 16% to 20% at 450 °C, as shown in Figure 27. For Catalyst 1, JP-8 yield was 42% to 65% at 350 °C and 38% to 64% at 450 °C. Therefore, Catalyst 1 performed significantly better than Catalyst 3 for JP-8 production under both temperatures. These results also indicate that when using COD900 catalyst at 180 psi, the lower 350 °C temperature is preferred. Operating the reactor at lower temperature also reduces the energy need.

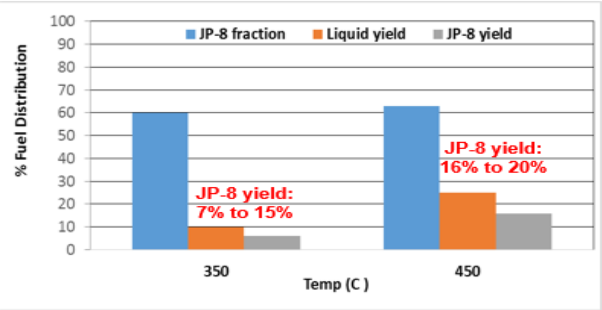


Figure 27. JP-8 production using Catalyst 3 at 350 and 450 °C, 180 psig, 5 h^{-1} WHSV with 20%C3-olefin

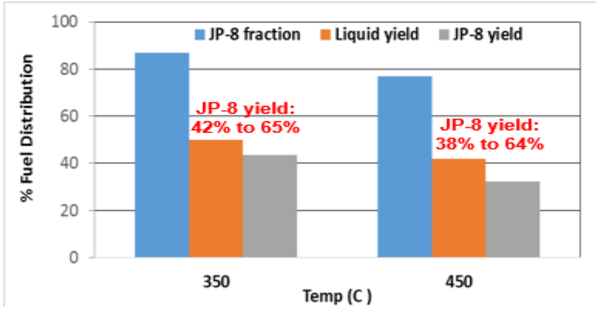


Figure 28 JP-8 production using Catalyst 1 at 350 and 450 °C, 180 psig, 5 h^{-1} WHSV with 20%C3-olefin

In addition, Figure 29 and Figure 30 show comparison of the C-number distribution between the liquid-products from C2 and C3 oligomerization, and available-JP-8-and-diesel samples at Altex facility. The oligomerized product follow similar branch-like distribution as JP-8 and with a carbon range between C8 and C24, peaking at C16 for C2 oligomerization, and a carbon range between C7 and C20, peaking at C12 for C3 oligomerization. In comparison, JP-8 gave a C-number distribution between C7 and C18, peaking at C11 and diesel gave a carbon distribution between C7 and C24, peaking at C12. These C-number distribution co-related to 82% and 87% JP-8 in the liquid-product from C2, and C3 oligomerization respectively. Because these commercial-catalysts are selective to C2 and others to C3, a mix catalyst-bed must be used that is effective on all olefins.

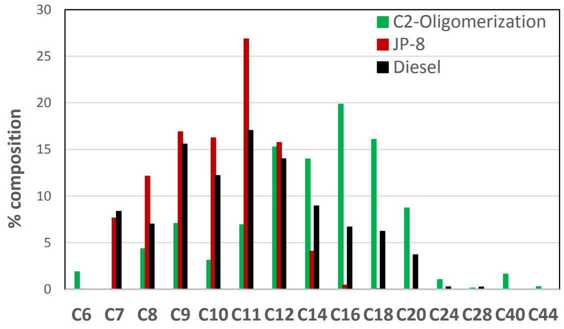


Figure 29. Comparison between C2 oligomerized product, JP-8, and diesel fuel

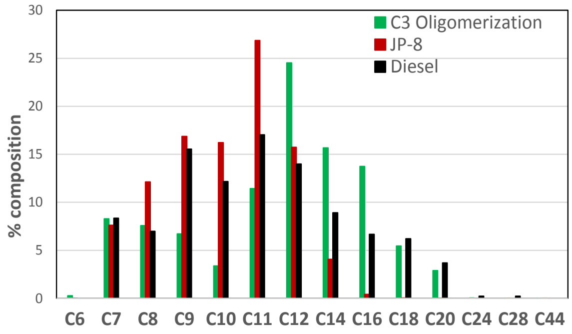


Figure 30. Comparison between C3 oligomerized product, JP-8, and diesel fuel

This comprehensive oligomerization-test matrix conducted for the GHGR-CBTL project show that C2 oligomerization using Catalyst 2 and Catalyst 3, and C3 oligomerization using Catalyst 3 and Catalyst 1 can be achieved at a wide range of pressures (100 to 180 psig), WHSV's (4 to 10 hr^{-1}), and temperatures (350 to 450°C). Oligomerization-optimization tests achieved up to 63% JP-8 yield with Catalyst 2. In general with a diluted olefin-feed stream, with an increase in oligomerization pressure (100 to 180 psig) and WHSV (4 to 10 hr^{-1}), liquid yield increased but JP-8 fraction dropped and as a result JP-8 yield dropped. However, with an increase in temperature (350 to 450°C), the liquid yield dropped, while JP-8 fraction increased and as a result JP-8 yield decreased. JP-8 yield was higher; at pressures of 100 to 120 psig (63%) over 150 to 180 psig (55%), WHSV of 4.2 hr^{-1} (61%) over 9.7 hr^{-1} (56%) and temperatures of 350°C (57%) over 450 °C (19%). Therefore, a combination of temperature, pressure, and WHSV was crucial to the JP-8 production. Further, Altex-designed oligomerization process for the GHGR-CBTL system favors significantly lower pressures of 100 to 120 psig versus traditional oligomerization requiring >500 psig with a flexibility of a range of concentrations from 12% to 100% in the olefin feed.

6.2 Improve Altex-in-house Catalyst Performance by PSU

PSU's catalyst improvement activity involved modification of a baseline catalyst to improve liquid yield, olefin conversion, JP-8 selectivity, coke-formation reduction, and catalyst-durability. For TRL 5 system, the baseline catalyst (Catalyst P1) was modified to produce; Catalyst P2, Catalyst P3, and Catalyst P4. Table 11 compares the % conversion and liquid yield for the PSU-developed catalysts. All PSU's catalysts perform at 0 psi giving 20-40% liquid yield and >80% conversion for C2 oligomerization, and 20-30% liquid yield and >97% conversion for C3 oligomerization.

Table 11 Comparison between PSU-developed catalysts performance for GHGR.

Catalyst	T (°C)	P (psig)	Olefin Feed	WHSV (h^{-1})	Conversion (%)	Liquid Yield (%)
P1	300-500	0-50	C_3H_6	6-24	97-100	20-45
P2	350	0	C_2H_4	12	50-90	0-20
P3	350	0-100	C_2H_4	12	87-100	40-65
	250-350	0	C_3H_6	18	~100	30-45
P4	350	0	C_2H_4	12	86-100	30-36

As shown in Table 11, with increasing pressure, Catalyst P3 had the highest liquid yield up to 65% followed by Catalyst P4, Catalyst P1, and Catalyst P2 giving 36%, 45%, and 20% liquid yield respectively for C2 oligomerization. This trend was similar for %conversion of olefins giving >86% conversion in most cases. For C3 oligomerization, Catalyst P3 had higher liquid yield (30-45%) compared to Catalyst P1 (20-45%).

Figure 31 compares the %conversion of C2-olefin using different PSU-developed catalysts. Catalyst P1 showed an initial conversion close to 100%. With increase in reaction time, the conversion decreased to 90%. As shown in Figure 31a, with increasing amount of the first type of modification (P2a, P2b, and P2c), the initial conversion dropped to 91%, 63%, and 51% respectively. With increase in reaction time, the conversion dropped to 76%, 34%, and 23% for P2a, P2b, and P2c respectively. The results showed that with this first type of modification the catalyst became less active and deactivated faster in terms of ethylene conversion.

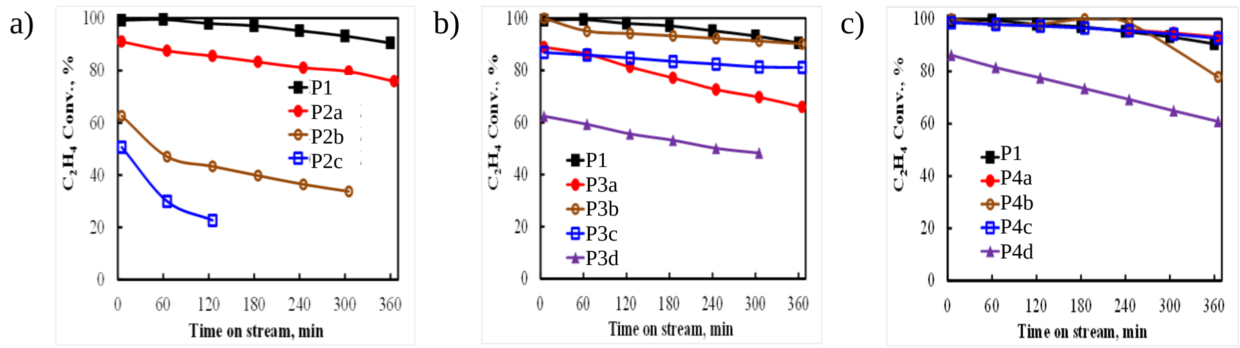


Figure 31. Comparison of ethylene conversion (%) for Catalyst P1, Catalyst P2a-c, Catalyst P3a-d, and Catalyst P4a-d at 350 °C, 12 h^{-1} WHSV, and 0 psi.

As shown in Figure 31b, with the second type of modification, the first amount of modification (P3a), the initial conversion dropped from 100% for Catalyst P1 to 89% for P3a. With increase in reaction time, this conversion dropped to 66%. Upon increasing the second type of modification (P3b), ethylene conversion increased to 100%. With increase in reaction time, this dropped to 90%. Upon further increasing the second type of modification (P3c and P3d), the initial conversion dropped to 87% and 82%, respectively. With increase in reaction time, this dropped to 81% and 48%, respectively.

Further, with the third type of modification, (P4a, P4b, and P4c), ethylene conversion almost overlapped with that of Catalyst P1, as shown in Figure 31c. A similar trend in the ethylene conversion with time was seen. The only exception was Catalyst P4b. It had initial conversion of 100% and dropped to 65% with increase in reaction time. The deactivation rate was much faster based on the decreasing trend in ethylene conversion for Catalyst P4.

Figure 32 compares the average liquid yield of C2-oligomerization using different PSU-developed catalysts. To better assess the effect of the different modification to Catalyst P1 on liquid yield during ethylene oligomerization, the average liquid yield after 6 hours of reaction time was calculated over each sample, and comparison shown in Figure 32. Average liquid yield decreased greatly with the first modification, as shown in Figure 32a. The average liquid yield was 29% for Catalyst P1, followed by 17%, 2%, and 0% over Catalyst P2a, Catalyst P2b, and Catalyst P2c, respectively.

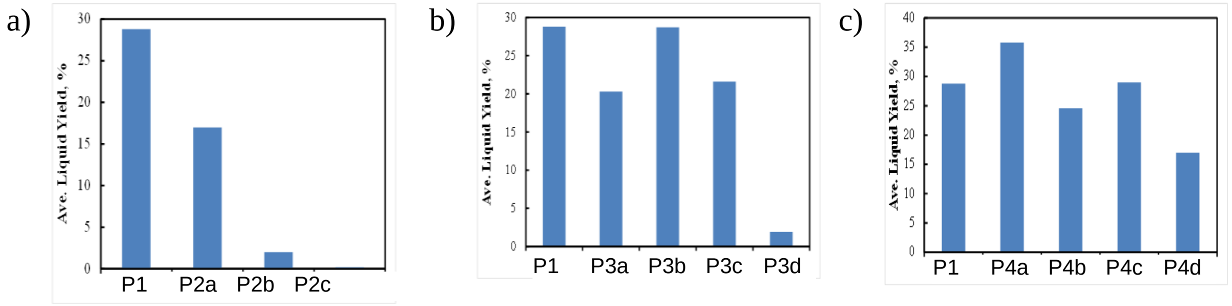


Figure 32. Comparison of liquid yield from ethylene oligomerization (%) of Catalyst P1, Catalyst P2a-c, Catalyst P3a-d, and Catalyst P4a-d at 350 °C, 12 h^{-1} WHSV, and 0 psi.

As shown in Figure 32b, the average liquid yield changed greatly with the second modification. Average liquid yield was 29% for Catalyst P1, followed by 20%, 29%, 22%, and 2% over Catalyst P3a, Catalyst P3b, Catalyst P3c, and Catalyst P3d, respectively.

Further, the average liquid yield also changed greatly with the addition of ZrO_2 as shown in Figure 32c. Average liquid yield was 29% for Catalyst P1, followed by 36%, 25%, 29% and 17% over the Catalyst P4a, Catalyst P4b, Catalyst P4c, and Catalyst P4d, respectively. Therefore, comparing all Catalyst P1-modifications, Catalyst P4a had the highest liquid yield of 36% at 0 psi, 350 °C, and 12 h^{-1} WHSV.

These results show that PSU-modified baseline catalyst performs significantly better than commercially-available catalysts in terms of; performance at lower pressure, liquid yield in JP-8 fuel range, % conversion of olefins, and catalyst-lifetime. To get similar fuel composition in JP-8 ranges as PSU catalysts, commercial catalysts needed 115 psi or higher pressure. PSU-modified baseline catalyst also has an altered acidity and pore structure that reduces coking and increase lifetime of the catalyst.

In addition, PSU's tests on the effects of C2-concentration in the feed over Catalyst P1 at different pressures and WHSV's were studied to optimize liquid yield. As shown in Figure 33a and b, at low ethylene concentration of 20% and extremely low WHSV of 0.75 h^{-1}, no liquid was obtained at pressures of 0 and 100 psig over the 4 hours reaction. Although no liquid was produced, the ethylene conversion was high, close to 100%. It may suggest that the ethylene was oligomerized and transformed into liquids. However, the liquid could not diffuse out and be collected likely due to extremely low WHSV of 0.75 h^{-1}. To prove this hypothesis, WHSV was increased from 0.75 to 3 h^{-1} for the same 20% ethylene feed, and the results are presented in Figure 33c and d.

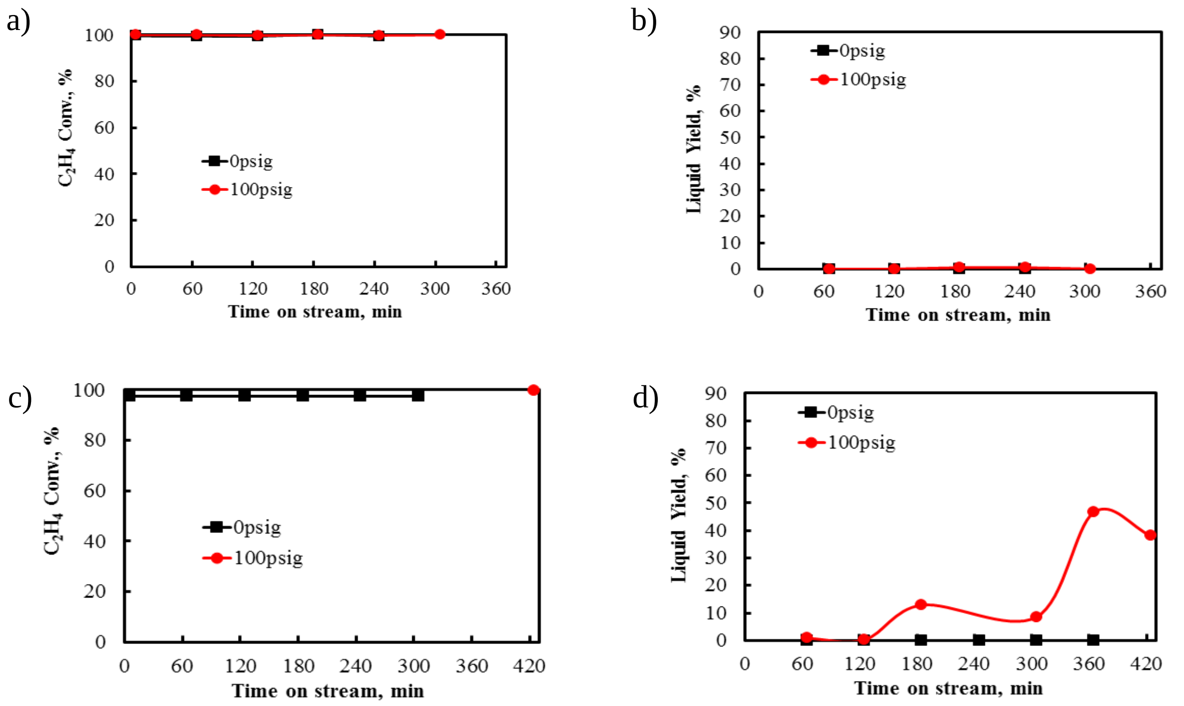


Figure 33 Comparison of ethylene conversion and liquid yield for ethylene oligomerization over Catalyst P1 at 0 and 100 psig, 20 vol% C_2H_4-80 vol% N_2; Temp of 350 °C; WHSV = 0.75 h^{-1} (a, b) and WHSV = 3 h^{-1} (c, d)

At 20% ethylene concentration, at 3 h^{-1} WHSV, no liquid was obtained at pressure of 0 psig over the 4 hours reaction, and conversion was about 98% as shown in Figure 33c and d. However, at 100 psig, little amount of liquid was obtained at the first two hours, and increased to 12.9%, 8.6%, 46.6%, and 38% at the 3rd, 5th, 6th, and 7th hour of oligomerization. The results demonstrate that for low ethylene concentration of 20%, at 100 psig, the liquid fuels can be effectively produced. Higher flow rate benefits the formation and outflow of the liquid products. The results also demonstrate the assumption that no liquid was obtained under the reaction conditions presented in Figure 33b due to the extremely low amount of the formed liquids, which could not flow out of the catalyst. The increase in the flow rate and the pressure could increase the formation of liquid products, and as a result, more liquid was collected. This indicated that the reaction pressure has an impact on the oligomerization of ethylene for low ethylene concentration of 20%. Thus, further studies were done to show the effect of the reaction pressure on the ethylene conversion and liquid yield over Catalyst P1, and the results are shown in Figure 34a and b.

As shown in Figure 34a, ethylene conversion ranged from 97% to 100%. The lowest conversion was obtained at reaction pressure of 0 psig. Unlike ethylene conversion, liquid yield showed greater dependence on the reaction pressure. No liquid was collected at 0 psig. At

50 psig, liquid yield was 0.6, 10.8, 22, 25, 15, and 22.2% at the 1st, 2nd, 3rd, 4th, 5th, and 6th hour of oligomerization respectively. At 100 psig, liquid yield was lower than 50 psig for the first 5 hrs, and increased up to 38 to 48% in the 6th, and 7th hr. At 200 psig, liquid yield varied between 3% and 16%. These results show that increasing the reaction pressure could promote the formation of liquid fuels. However, higher reaction pressure such as 200 psig did not benefit the production of liquid fuels. The liquid yield decreases as the sequence of Y50 > Y100 > Y200 > Y0.

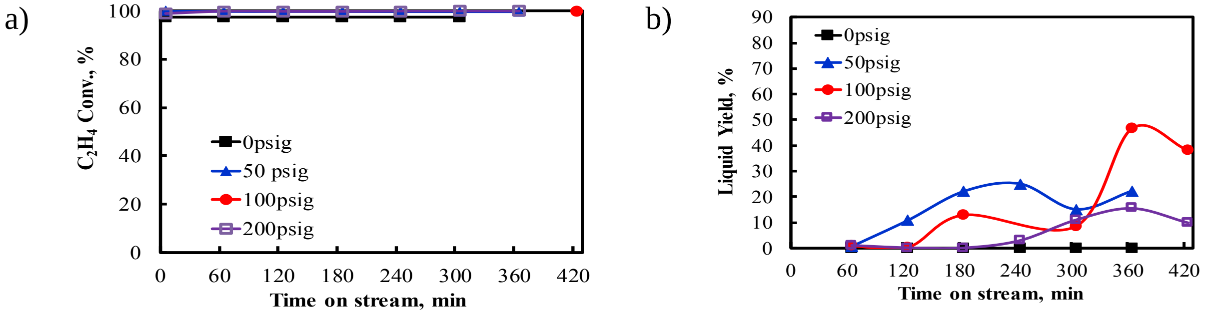


Figure 34. a) Ethylene conversion and b) liquid yield - for 20% ethylene+80% N_2 oligomerization over Catalyst P1 at different pressures, 350 °C, WHSV = 3.0 h^{-1}

Through the comparisons between Figure 33b and d, the results implied that the increase of WHSV had promotional effect on the formation of liquid fuels. Thus, the effect of WHSV on the liquid production was investigated. The results obtained at reaction pressure of 50 psig with different WHSV's are presented in Figure 35a and b.

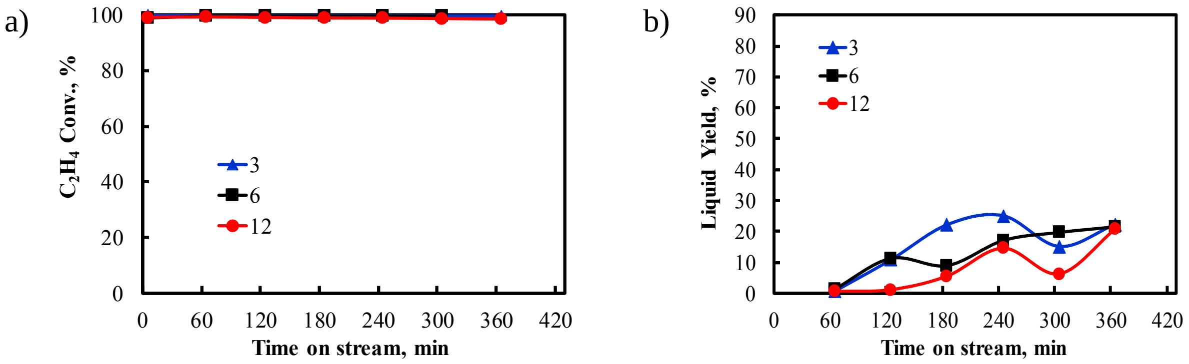
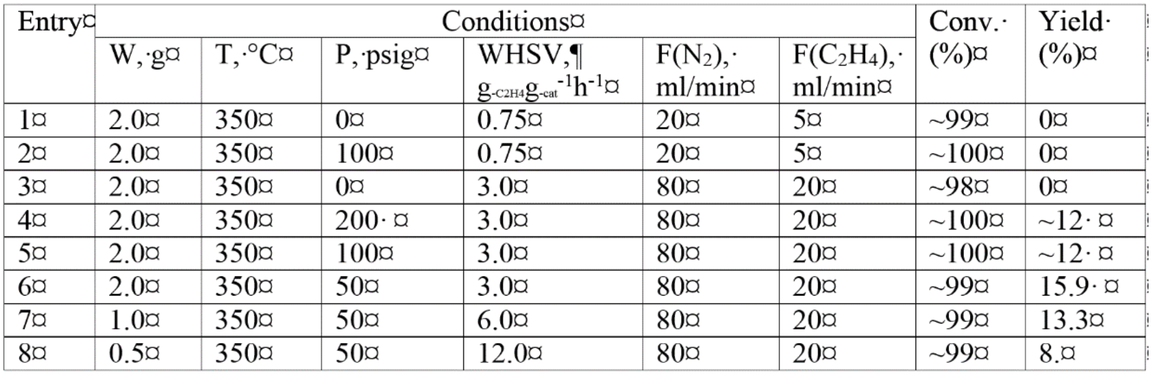


Figure 35. a) Ethylene conversion and b) liquid yield for 20% ethylene+80% N_2 oligomerization over Catalyst P1 at different WHSVs, 350 °C; 50 psig.

For all the different WHSVs, ethylene conversion ranged from 98% to 100% as shown in Figure 35a. The lowest conversion was obtained at WHSV of 12 h^{-1}. Unlike ethylene conversion, liquid yield showed a greater dependence on the WHSV as shown in Figure 35b. At WHSV of 3 h^{-1}, the liquid yield was 0.6, 10.8, 22.2, 25.0, 15.0 and 22.2% at the 1st, 2nd, 3rd, 4th, 5th and 6th hour of reaction, respectively. When WHSV was increased to 6 h^{-1}, the liquid yield became 1.3, 11.3, 8.9, 17.1, 19.8 and 21.6% at the 1st, 2nd, 3rd, 4th, 5th and 6th hour of reaction, respectively. Further increasing the WHSV to 12 h^{-1}, the liquid yield became 0.5, 1.1, 5.5, 14.5, 6.2 and 20.8% at the 1st, 2nd, 3rd, 4th, 5th and 6th hour of reaction, respectively. In general, the liquid yield decreased with the increase in WHSV at the conditions studied this reporting period.

To better assess the effect of both pressure and WHSV on the liquid yield during ethylene oligomerization reaction with low ethylene concentration of 20% over Catalyst P1, the average liquid yield for the 6 hours of reaction time was calculated for each experiment and compared that is presented in Table 12. In summary, the best condition is WHSV of 3.0 h^{-1} with the reaction pressure of 50 psig for the catalytic oligomerization of 20% ethylene in N_2 stream at 350°C.

Table 12. Summary of average ethylene conversion and liquid yield at different conditions obtained in this reported period.



Entry	Conditions						Conv. (%)	Yield (%)
	W, g	T, °C	P, psig	WHSV, $g_{\text{-C2H4}}g_{\text{-cat}}^{-1}h^{-1}$	F(N_2), ml/min	F(C_2H_4), ml/min		
1	2.0	350	0	0.75	20	5	~99	0
2	2.0	350	100	0.75	20	5	~100	0
3	2.0	350	0	3.0	80	20	~98	0
4	2.0	350	200	3.0	80	20	~100	~12
5	2.0	350	100	3.0	80	20	~100	~12
6	2.0	350	50	3.0	80	20	~99	15.9
7	1.0	350	50	6.0	80	20	~99	13.3
8	0.5	350	50	12.0	80	20	~99	8.

6.3 Catalyst for TRL 5 GHGR-CBTL System

The purpose of the current report is to cover the Altex catalyst scale-up activities for the TRL-5 GHGR-CBTL system, which was a catalyst production milestone on the schedule. The scaled up catalyst will be utilized in the TRL-5 test system that requires 7 to 8 kg of catalyst. The needed catalyst is available as described below.

Following a comprehensive set of JP-8-production optimization-tests, Catalyst P1, Catalyst 1, and Catalyst 2 have been selected based on their performance and scale-up availability for the TRL-5 system. PSU's modifications to Catalyst P1 were mainly focused on reducing coking, increasing catalyst life-time, and increasing the JP-8 fraction in the liquid-fuel. Since modifications are not finalized and the baseline Catalyst P1 performance is acceptable, the readily available baseline Catalyst P1 will be used in the TRL-5 test system. Also, the this catalyst was successfully used in the TRL-4 system for oligomerization at 15 psig for liquid-fuel production.

For oligomerization below 100 psig, the reactors will be packed with 100% Catalyst P1, and for oligomerization between 100 and 180 psig, reactors will be packed with a combination of 80-85% Catalyst 2 and 15-20% Catalyst 1. The variation in the oligomerization conditions and ratio of these two catalysts are based on possible variations in the upstream design and the composition of the stream entering the oligomerization system. However, the Altex oligomerization system is flexible and can operate over a range of pressures (0 to 180 psig), temperatures (250 to 450°C), and WHSVs (1 to 12 h^{-1}). Current estimates of the entering composition were used to select a ratio of the Catalyst 2 and Catalyst 1 in order to optimize JP-8 production.

Table 13. Oligomerization catalysts to be used in the TRL-5 system.

Catalyst	Catalyst packing amount (%)	Oligomerization Conditions
Catalyst P1	100% (single catalyst)	<100 psig, 350 to 450 °C
Catalyst 2 (C_2 specific)	80 to 85% (combined with Catalyst 1)	100-180 psig, 350 to 450 °C
Catalyst 1 (C_3 specific)	15 to 20% (combined with Catalyst 2)	100-180 psig, 350 to 450 °C

Further, a brief summary of the performance of each of these catalysts is provided below. These results also include the results for the PSU-modified Catalyst P1 catalysts (P2, P3, and P4).

6.4 Catalyst Performance Summary for TRL-5 GHGR-CBTL system

Optimization tests for JP-8 production using C_2 olefin as the feed were performed over a range of temperatures (240 to 450°C), pressures (0 to 180 psig), and WHSVs (1 to 10 hr^{-1}). These tests were conducted with different catalysts: a baseline catalyst used for the PSU work (Catalyst P1), Catalyst P1 modified by PSU (Catalysts P2, P3 and P4), and commercial catalysts. A summary of the ethylene (C_2) oligomerization optimization performed up to date is presented in Figure 36.

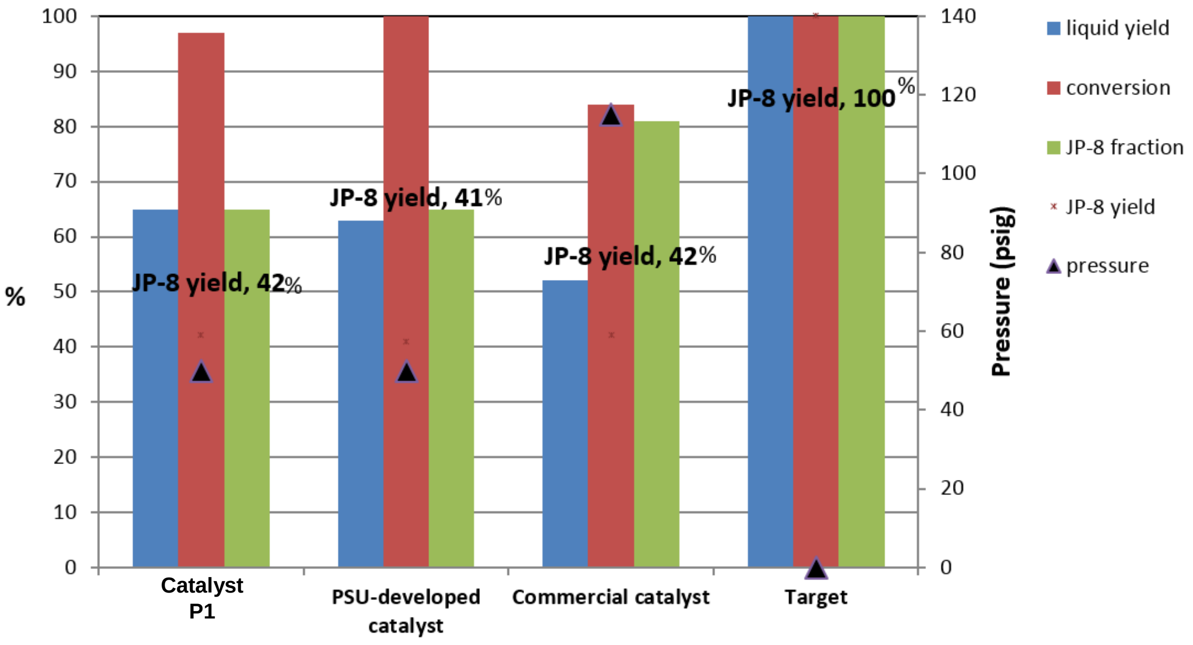


Figure 36. Summary of JP-8 yield achieved using Catalyst P1 (50 psi), PSU-developed catalyst (50 psi), and Commercial (115 psi) catalysts of ethylene (C_2) oligomerization

Figure 36 shows that both Catalyst P1 and PSU-developed catalysts give between 41 and 42% JP-8 yield with 65% JP-8 fraction and 63 to 65% liquid yield at 50 psig with C_2-olefins. Also, not shown in this figure, both these catalysts perform at 0 psig with lower JP-8 yield up to 5%. In comparison, the commercial catalyst gave similar JP-8 yield of 42% at a higher pressure of 115 psig and had zero liquid yield at 0 psig. As shown in Figure 36, olefin conversion was

>95% with Catalyst P1 and the PSU-developed catalysts, while it was 83% for the commercial catalyst. Further, all of the oligomerization-optimization test conditions follow the Altex oligomerization-reactor design for the TRL-5 system. Availability of the Altex-selected oligomerization catalysts for the TRL-5 system, and JP-8 optimized conditions of low pressures (0 to 180 psig) versus existing industrial oligomerization at >500 psig makes the Altex-oligomerization for the TRL-5 system both low-risk for JP-8 production and energy-efficient. Also, the Altex-developed oligomerization system allows options of performing oligomerization reactions at a wide range of pressures (0 to 180 psig) by a combination of multiple catalysts or a single catalyst in the TRL-5 system.

6.5 Conclusion

With a wide-range of oligomerization-optimization-testing, Catalyst P1, Catalyst 1, and Catalyst 2 have been selected based on their performance and scale-up availability for the TRL-5 GHGR-CBTL oligomerization system. For oligomerization below 100 psig, Catalyst P1 will be used, while for 100 to 180 psig, a combination of Catalyst 2 (85%) and Catalyst 1 (15%) will be used. The catalyst-packing options are to optimize JP-8 production that is suitable for the Altex-oligomerization design. With C_2-olefin, Altex has achieved to date a JP-8 yield of 42%. With C_3-olefin, a JP-8 yield of 65% has been achieved. Also, all the oligomerization-optimization test conditions follow the Altex oligomerization-reactor design for the TRL-5 system. Availability of the Altex-selected oligomerization catalysts for the TRL-5 system combined with significantly better JP-8 yield at low-pressures (0 to 180 psig) as compared to existing industrial oligomerization catalysts makes the Altex oligomerization approach for the TRL-5 system both low-risk for JP-8 production and energy-efficient.

7 TECHNO-ECONOMIC ASSESSMENT (TEA)

A Techno-Economic Analysis (TEA) report is being submitted as a separate report that describes the technical and economics potential of the GHGR-CBTL technology in detail. A summary is included herein.

To conduct the TEA, the plant was reconfigured to reduce or eliminate natural-gas and electricity imports in order to have a self-sufficient plant. This was accomplished by changing from using the waste-heat-recovery steam from the steam cracker for the production of utility steam to using it for the production of electricity. This led to changing from utilizing the waste-heat-recovery steam and combustion of imported natural gas for the production of utility steam to utilizing the combustion of excess fuel gas and some of the char for the production of utility steam.

The TEA was performed following DOE guidelines listed in the FOA and the referenced documents. A matrix of plant configurations was developed and evaluated. The scenarios included capturing or venting CO_2 in order to estimate the cost of CO_2 capturing. A baseline case with CO_2 capture was included and was compared with DOE/NETL Fischer-Tropsch Liquids Baseline Case[1].

GHGR-CBTL uses coal and biomass as feed stock. The feedstock blend has been adjusted in each case studied to obtain a Well-to-Wake (WTWa) GHG footprint of 84 g-CO_2e/MJ, which corresponds to petroleum-based JP-8.

7.1 Capital and Operating Costs

7.1.1 Capital Costs

Two separate cost-estimation approaches were used. For the sections of the plant that were designed from the ground up by Altex, all major equipment including pumps, blower, compressors, and vessels were sized and costed individually. For most of these pieces of equipment, CHEMCAD was used to both size and cost the equipment. CHEMCAD utilizes standard methods for sizing most equipment. It also utilizes correlations for equipment cost developed by Walas[20] that was then updated to a June 2011 cost basis by using the appropriate components of the Chemical Engineering Plant Construction Index (e.g., the equipment index for general equipment, the heat-exchanger index for heat exchangers, etc.). For equipment for which CHEMCAD did not have a suitable sizing and/or costing correlation, Excel was used to size the equipment and/or obtain a cost estimate based upon vendor data.

For the sections of the plant for which NETL equivalents were available, the DOE/NETL recommended scaling methods presented in "Quality Guidelines for Energy Systems Studies, Capital Cost Scaling Methodology"[21] were utilized.

Because certain components of the GHGR-CBTL plant have not been demonstrated on a commercial scale, process contingencies were applied to certain plant components based upon the DOE/NETL QGESS presentation, "Introduction to Performing a Techno-Economic Analysis for Power Generation Systems."[22]

The TEA was performed for a matrix of cases and plant configurations. A description of range of plant-configuration variables considered is provided in Table 14. For example, one step in the process can use a baseline process (denoted as R) or an alternative type of process (denoted as C). Process R was used in previous plant designs and has experimental data to back up the process simulations. Process C is an alternative process that process simulations suggested would be a better option than Process R, but does not have as much experimental data to back up the process simulations. Therefore, the use of Process R was considered as the baseline process for comparison with other technologies. The use of Process C was considered as an estimate of near-term improved performance. The cases studied are labeled by a sequence of case designators shown in Table 14 listed in the same order as in Table 14. Therefore, the baseline plant configuration that utilizes the refrigeration route, char sold for firing CHP systems, capturing and sequestering the CO_2, and including the costs of CO_2 T&S costs in the ECO calculation is designated as Case RCCT. A matrix of the cases studied is presented in Table 15.

Table 14. Description of plant-configuration variables.

Plant-Configuration Variable	Case Designator	Description of configuration
Process option		
R	R	Baseline technology: based on past designs and experiments.
C	C	Alternative technology: based upon projected performance (experimental data not yet available)
Fate of char		
Landfill at site	L	Char is sent to on-site landfill (or mine if plant is at the minemouth).
Sell for electricity production	E	Char is sold for electricity production in traditional power plants.
Sell for firing Combined Heat & Power (CHP)	C	Char is sold for firing CHP system.
CO_2 Capture or Venting		
Vented	V	CO_2 in acid gas separated from VOCs and H_2S and vented. CO_2 in flue gas is not captured. VOCs and H_2S sent to sulfur plant.
Captured (e.g., sequestered)	C	Acid gas is combined with flue gas before capture of CO_2, H_2S and VOCs. CO_2 dried and compressed to 2,200 psig. VOCs and H_2S sent to sulfur plant.
CO_2 T&S Costs Included in ECO price?		
Yes	T	CO_2 T&S costs included in ECO calculations
No	[blank]	CO_2 T&S costs not included in ECO calculations

Table 15. Matrix of cases considered.

Case	Increasing Olefin Partial Pressures	Fate of Char	Vent or Capture & Sequester CO_2	CO_2 T&S Costs Included in ECO
RLV	Refrigeration	Landfill	Vent	N/A
RLCT	Refrigeration	Landfill	Capture	Yes
REV	Refrigeration	Electricity	Vent	N/A
RECT	Refrigeration	Electricity	Capture	Yes
RCV	Refrigeration	CHP	Vent	N/A
RCC	Refrigeration	CHP	Capture	No
RCCT[a]	**Refrigeration**	**CHP**	**Capture**	**Yes**
CLV	Compression	Landfill	Vent	N/A
CLCT	Compression	Landfill	Capture	Yes
CEV	Compression	Electricity	Vent	N/A
CECT	Compression	Electricity	Capture	Yes
CCV	Compression	CHP	Vent	N/A
CCC	Compression	CHP	Capture	No
CCCT	Compression	CHP	Capture	Yes

[a] Baseline case

The TOC for the entire facility with CO_2 capture was estimated to be $3,820 million. This cost includes appropriate contingency factors and owners costs. The owner's costs use the same rates and bases as the DOE/NETL FT-Liquids rates and bases.[1] On a per-barrel basis of daily liquid capacity, this cost is $95,778/bpd for the baseline case. Table 17 shows a summary breakdown of the capital costs for the baseline case (Case RCCT). Table 18, Table 19, and Table 20 show similar summaries for 3 other important cases (Cases RCV, CCCT, and CCV). These four cases were selected for inclusion because Case RCV is necessary to calculate the cost of capturing/sequestering CO_2, and Cases CCCT and CCV provide a comparison of the use of Process R and Process C. The TOC is the TPC plus owner's costs.

7.1.2 Operating Costs

7.1.2.1 Fixed Operating Costs

Operating costs were estimated in the standard manner. The bases and rates for estimating operating costs are listed in Table 16 and summarized in Table 21, Table 22, Table 23, and Table 24.

Table 16. Rates and bases for estimating fixed operating costs.

Fixed Operating Cost Component	Rate	Basis
Operating Labor Rate	$39.70	Per hour
Operating Labor Burden	30%	Operating Labor Rate
Labor Overhead Charge	25%	Burdened Labor
Operating Labor (per shift)		
Operators	35	From analysis of unit ops
Skilled Operators	20%	Number of Operators
Foreman	1	1 per plant train
Lab Techs, etc.	3	3 per plant train
Maintenance Labor	1.745%	TEC of Non-ancillary equipment
Administrative & Support Labor	25%	Operating + Maintenance Labor
Property Taxes & Insurance	2%	TPC

7.1.2.2 Variable Operating Costs

The variable operating costs included costs for maintenance materials, raw water, chemicals, and waste disposal. Credits were applied for sales of byproducts and sales of net electricity generation. Feedstock costs are considered separately.

The cost of maintenance materials was estimated as 3.3% of the maintenance-labor basis (TEC - TECX of Accounts N11-N14). The remaining variable operating costs were obtained from flow rates from the heat and mass balances combined with estimated prices of each of the materials required. Prices were estimated from literature resources.

7.1.2.3 Feedstock Costs and Co-Product Credits

The feedstock costs were estimated by utilizing the proximate analysis of the coal received from NA Coal company and the method described in the DOE/NETL fuel-price guidelines.[23] Biomass costs were estimated to be $50 per tonne, as-received, at 10% moisture.

A credit of $60/MWh is included for excess power generated. For the char price, the price was estimated as if it was ND lignite and adjusted based upon its heating value and sulfur content using the method described in the DOE/NETL fuel-price guidelines.[23] this results in a credit of $28.23/ton being included for the cases where the char is sold for producing electricity or for firing CHP systems. It should be noted that for the sensitivity studies below, if the char selling price was set to $0/ton, it was assumed that the char was considered a waste with a negative co-product credit.

7.1.2.4 Operating-Costs Summary

As required by the project, the feedstock blend was adjusted in each case to obtain a Well-to-Wake (WTWa) GHG footprint of 84 g-CO_2e/MJ, which corresponds to petroleum-based JP-8. Table 21, Table 22, Table 23, and Table 24 show the breakdown of the total operating cost for Cases RCCT, RCV, CCCT, and CCV, respectively.

Table 17. Total-plant-cost summary - Case RCCT (Baseline Case).

Acct.	Service	Equipment	Material	Labor		Sales	Bare Erected	Eng. CM	Contingencies		Total Plant Cost	
No.	Process Equipment	Cost	Cost	Direct	Indirect	Tax	Cost	HO & Fee	Process	Project	$	$/BPSD
	TOTAL COST	**$1,465,610**	**$92,407**	**$730,049**	**$0**	**$0**	**$2,288,066**	**$231,047**	**$171,900**	**$440,557**	**$3,131,570**	**$80,522**
	OWNER'S COSTS											
	Pre-production Costs											
	6 months All labor										$29,263	$752
	1 month Maint. Materials										$4,319	$111
	1 month non-fuel consumables										$1,777	$46
	1 month waste disposal										$441	$11
	25% of fuel cost @ 100% CF										$11,313	$291
	2% of TPC										$62,631	$1,610
	total										**$109,745**	**$2,822**
	Inventory Capital											
	60-day supply of fuel and consumables										$3,896	$100
	0.5% of TPC (spare parts)										$15,658	$403
											$19,553	**$503**
	Land										**$900**	**$23**
	Other Owner's Costs (15% of TPC)										**$469,736**	**$12,078**
	Financing Costs (2.7% of TPC)										**$84,552**	**$2,174**
	TOC										**$3,820,420**	**$98,234**
	TASC Multiplier									(Commerical FUELS 35 yr)	**1.1812**	
	TASC										**$4,512,680**	**$116,034**

Table 18. Total-plant-cost summary - Case RCV.

Acct.	Service	Equipment	Material	Labor		Sales	Bare Erected	Eng. CM	Contingencies		Total Plant Cost	
No.	Process Equipment	Cost	Cost	Direct	Indirect	Tax	Cost	HO & Fee	Process	Project	$	$/BPSD
	TOTAL COST	**$1,396,216**	**$92,378**	**$702,095**	**$0**	**$0**	**$2,190,689**	**$222,578**	**$171,900**	**$419,436**	**$3,004,603**	**$77,257**
	OWNER'S COSTS											
	Pre-production Costs											
	6 months All labor										$28,224	$726
	1 month Maint. Materials										$4,107	$106
	1 month non-fuel consumables										$1,777	$46
	1 month waste disposal										$441	$11
	25% of fuel cost @ 100% CF										$14,996	$386
	2% of TPC										$60,092	$1,545
	total										**$109,637**	**$2,819**
	Inventory Capital											
	60-day supply of fuel and consumables										$3,896	$100
	0.5% of TPC (spare parts)										$15,023	$386
											$18,919	**$486**
	Land										**$900**	**$23**
	Other Owner's Costs (15% of TPC)										**$450,690**	**$11,589**
	Financing Costs (2.7% of TPC)										**$81,124**	**$2,086**
	TOC										**$3,670,236**	**$94,372**
	TASC Multiplier									(Commerical FUELS 35 yr)	**1.1812**	
	TASC										**$4,335,283**	**$111,473**

Table 19. Total-plant-cost summary - Case CCCT.

Acct.	Service	Equipment	Material	Labor		Sales	Bare Erected	Eng. CM	Contingencies		Total Plant Cost	
No.	Process Equipment	Cost	Cost	Direct	Indirect	Tax	Cost	HO & Fee	Process	Project	$	$/BPSD
	TOTAL COST	**$1,417,968**	**$92,387**	**$710,254**	**$0**	**$0**	**$2,220,610**	**$225,181**	**$171,900**	**$425,943**	**$3,043,633**	**$78,261**
	OWNER'S COSTS											
	Pre-production Costs											
	6 months All labor										$28,461	$732
	1 month Maint. Materials										$4,174	$107
	1 month non-fuel consumables										$1,777	$46
	1 month waste disposal										$441	$11
	25% of fuel cost @ 100% CF										$11,098	$285
	2% of TPC										$60,873	$1,565
	total										**$106,824**	**$2,747**
	Inventory Capital											
	60-day supply of fuel and consumables										$3,896	$100
	0.5% of TPC (spare parts)										$15,218	$391
											$19,114	**$491**
	Land										**$900**	**$23**
	Other Owner's Costs (15% of TPC)										**$456,545**	**$11,739**
	Financing Costs (2.7% of TPC)										**$82,178**	**$2,113**
	TOC										**$3,713,557**	**$95,486**
	TASC Multiplier									(Commerical FUELS 35 yr)	**1.1812**	
	TASC										**$4,386,454**	**$112,789**

Table 20. Total-plant-cost summary - Case CCV.

Acct.	Service	Equipment	Material	Labor		Sales	Bare Erected	Eng. CM	Contingencies		Total Plant Cost	
No.	Process Equipment	Cost	Cost	Direct	Indirect	Tax	Cost	HO & Fee	Process	Project	$	$/BPSD
	TOTAL COST	**$1,349,511**	**$92,357**	**$682,354**	**$0**	**$0**	**$2,124,222**	**$216,794**	**$171,900**	**$404,983**	**$2,917,899**	**$75,028**
	OWNER'S COSTS											
	Pre-production Costs											
	6 months All labor										$27,149	$698
	1 month Maint. Materials										$3,965	$102
	1 month non-fuel consumables										$1,777	$46
	1 month waste disposal										$441	$11
	25% of fuel cost @ 100% CF										$14,879	$383
	2% of TPC										$58,358	$1,501
	total										**$106,570**	**$2,740**
	Inventory Capital											
	60-day supply of fuel and consumables										$3,896	$100
	0.5% of TPC (spare parts)										$14,589	$375
											$18,485	**$475**
	Land										**$900**	**$23**
	Other Owner's Costs (15% of TPC)										**$437,685**	**$11,254**
	Financing Costs (2.7% of TPC)										**$78,783**	**$2,026**
	TOC										**$3,564,685**	**$91,658**
	TASC Multiplier									(Commerical FUELS 35 yr)	**1.1812**	
	TASC										**$4,210,606**	**$108,267**

Table 21. Total operating and maintenance costs - Case RCCT (Baseline Case).

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
CASE RCCT						
					bbl/day:	38,891
					Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Operating Labor Requirements(O.J.)per Shift:	1 unit/mod.	mods.plant	Plant			
		1				
Skilled Operator	8		8			
Operator	37		37			
Foreman	1		1			
Lab Tech's, etc.	3		3			
TOTAL-O.J.'s	49		49			
					Annual Cost	Annual Unit Cost
					$	$/(bbl/day)
Annual Operating Labor Cost					$ 22,153,076	$ 570
Maintenance Labor Cost	1.74500%	of TEC less 11-14			$ 24,667,303	$ 634
Administrative & Support Labor	25%	of OL+ML			$ 11,705,095	$ 301
Property Taxes and Insurance	2%	of TPC			$ 62,631,408	$ 1,610
TOTAL FIXED OPERATING COSTS					**$ 121,156,883**	**$ 2,110**
VARIABLE OPERATING COSTS						
						$/bbl
Maintenance Material Cost	3.30%	of M&L Basis			**$ 46,648,768**	**$ 3.65137**
Consumables	Consumption		Unit	Initial Fill		$ -
	Initial Fill	/Day	Cost	Cost		$ -
Water (/1000 gallons)	0	1,806	1.67	**$ -**	**$ 990,963**	**$ 0.07757**
Chemicals		6				
Subtotal Chemicals				**$ 4,363,446**	**$ 18,204,530**	**$ 1.42494**
Other						
Butane (tons)	0	0	651.34	$ -	$ -	$ -
Gases, N_2 etc. (/100scf)	0	0	0.00	$ -	$ -	$ -
L.P. Steam (/1000 pounds)	0	0	0.00	$ -	$ -	$ -
Subtotal Other				**$ -**	**$ -**	**$ -**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	22320	0.65	$ -	$ 4,765,878	$ 0.37304
Flyash (ton)	0	0	0.00	$ -	$ -	$ -
Char (ton)	0	628	0.00	$ -	$ -	$ -
Subtotal-Waste Disposal				**$ -**	**$ 4,765,878**	**$ 0.37304**
By-products & Emissions (credit)						
Char (ton)	0	628	28.23	$ -	$ (5,826,162)	$ (0.45604)
Supplemental Electricity (for sale) (MWh)	0	(314)	$ (60.00)	$ -	$ 6,191,006	$ 0.48459
Subtotal By-Products				**$ -**	**$ 364,844**	**$ 0.02856**
TOTAL VARIABLE OPERATING COSTS				**$ 4,363,446**	**$ 70,974,983**	**$ 5.55548**
Less By-product Credis					**$ 70,610,139**	**$ 5.52692**
ND Lignite (ton)	0	50,282	18.19		$ 300,454,135	$ 24
Switchgrass (ton)	0	11,461	50		$ 188,253,540	$ 15
Fuel (ton)		61,743			**$ 488,707,675**	**$ 38.25297**

Table 22. Total operating and maintenance costs - Case RCV.

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
CASE RCV						
					bbl/day:	38,298
					Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Operating Labor Requirements(O.J.)per Shift:	1 unit/mod.	mods.plant	Plant			
		1				
Skilled Operator	8		8			
Operator	36		36			
Foreman	1		1			
Lab Tech's, etc.	3		3			
TOTAL-O.J.'s	48		48			
					Annual Cost	Annual Unit Cost
					$	$/(bbl/day)
Annual Operating Labor Cost					$ 21,700,973	$ 567
Maintenance Labor Cost	1.74500%	of TEC less 11-14			$ 23,456,662	$ 612
Administrative & Support Labor	25%	of OL+ML			$ 11,289,409	$ 295
Property Taxes and Insurance	2%	of TPC			$ 60,092,051	$ 1,569
TOTAL FIXED OPERATING COSTS					**$ 116,539,094**	**$ 2,110**
VARIABLE OPERATING COSTS						
						$/bbl
Maintenance Material Cost	3.30%	of M&L Basis			**$ 44,359,303**	**$ 3.52593**
Consumables	Consumption		Unit	Initial Fill		$ -
	Initial Fill	/Day	Cost	Cost		$ -
Water (/1000 gallons)	0	1,806	1.67	**$ -**	**$ 990,963**	**$ 0.07877**
Chemicals		6				
Subtotal Chemicals				**$ 4,363,446**	**$ 18,204,530**	**$ 1.44700**
Other						
Butane (tons)	0	0	651.34	$ -	$ -	$ -
Gases, N_2 etc. (/100scf)	0	0	0.00	$ -	$ -	$ -
L.P. Steam (/1000 pounds)	0	0	0.00	$ -	$ -	$ -
Subtotal Other				**$ -**	**$ -**	**$ -**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	22320	0.65	$ -	$ 4,765,878	$ 0.37882
Flyash (ton)	0	0	0.00	$ -	$ -	$ -
Char (ton)	0	628	0.00	$ -	$ -	$ -
Subtotal-Waste Disposal				**$ -**	**$ 4,765,878**	**$ 0.37882**
By-products & Emissions (credit)						
Char (ton)	0	628	28.23	$ -	$ (5,826,162)	$ (0.46310)
Supplemental Electricity (for sale) (MWh)	0	3,613	$ (60.00)	$ -	$ (71,207,049)	$ (5.65994)
Subtotal By-Products				**$ -**	**$ (77,033,211)**	**$ (6.12303)**
TOTAL VARIABLE OPERATING COSTS				**$ 4,363,446**	**$ (8,712,536)**	**$ (0.69252)**
Less By-Product Credits					**$ 68,320,674**	**$ 5.43051**
ND Lignite (ton)	0	25,098	18.19		$ 149,968,373	$ 12
Switchgrass (ton)	0	30,310	50		$ 497,840,543	$ 40
Fuel (ton)					**$ 647,808,916**	**$ 51.49149**

Table 23. Total operating and maintenance costs - Case CCCT.

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
CASE CCCT						
					bbl/day:	38,298
					Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Operating Labor Requirements(O.J.)per Shift:	1 unit/mod.	mods.plant	Plant			
		1				
Skilled Operator	8		8			
Operator	36		36			
Foreman	1		1			
Lab Tech's, etc.	3		3			
TOTAL-O.J.'s	48		48			
					Annual Cost	Annual Unit Cost
					$	$/(bbl/day)
Annual Operating Labor Cost					$ 21,700,973	$ 567
Maintenance Labor Cost	1.74500%	of TEC less 11-14			$ 23,836,157	$ 622
Administrative & Support Labor	25%	of OL+ML			$ 11,384,282	$ 297
Property Taxes and Insurance	2%	of TPC			$ 60,872,668	$ 1,589
TOTAL FIXED OPERATING COSTS					**$ 117,794,080**	**$ 2,110**
VARIABLE OPERATING COSTS						
						$/bbl
Maintenance Material Cost	3.30%	of M&L Basis			**$ 45,076,973**	**$ 3.58297**
Consumables	Consumption		Unit	Initial Fill		$ -
	Initial Fill	/Day	Cost	Cost		$ -
Water (/1000 gallons)	0	1,806	1.67	**$ -**	**$ 990,963**	**$ 0.07877**
Chemicals		6				
Subtotal Chemicals				**$ 4,363,446**	**$ 18,204,530**	**$ 1.44700**
Other						
Butane (tons)	0	0	651.34	$ -	$ -	$ -
Gases, N_2 etc. (/100scf)	0	0	0.00	$ -	$ -	$ -
L.P. Steam (/1000 pounds)	0	0	0.00	$ -	$ -	$ -
Subtotal Other				**$ -**	**$ -**	**$ -**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	22320	0.65	$ -	$ 4,765,878	$ 0.37882
Flyash (ton)	0	0	0.00	$ -	$ -	$ -
Char (ton)	0	15,080	0.00	$ -	$ -	$ -
Subtotal-Waste Disposal				**$ -**	**$ 4,765,878**	**$ 0.37882**
By-products & Emissions (credit)						
Char (ton)	0	15080	28.23	$ -	$ (139,827,881)	$ (11.11430)
Supplemental Electricity (for sale) (MWh)	0	1,942	$ (60.00)	$ -	$ (38,284,287)	$ (3.04305)
Subtotal By-Products				**$ -**	**$ (178,112,168)**	**$ (14.15735)**
TOTAL VARIABLE OPERATING COSTS				**$ 4,363,446**	**$ (109,073,824)**	**$ (8.66980)**
Less By-product Credis					**$ 69,038,344**	**$ 5.48756**
ND Lignite (ton)	0	51,752	18.19		$ 309,237,991	$ 25
Switchgrass (ton)	0	10,361	50		$ 170,182,942	$ 14
Fuel (ton)		62,113			**$ 479,420,934**	**$ 38.10707**

Table 24. Total operating and maintenance costs - Case CCV.

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Jun):	2011
CASE CCV						
					bbl/day:	38,298
					Capacity Factor (%):	90
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate(base):	39.70	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Operating Labor Requirements(O.J.)per Shift:	1 unit/mod.	mods.plant	Plant			
		1				
Skilled Operator	7		7			
Operator	35		35			
Foreman	1		1			
Lab Tech's, etc.	3		3			
TOTAL-O.J.'s	46		46			
					Annual Cost	Annual Unit Cost
					$	$/(bbl/day)
Annual Operating Labor Cost					$ 20,796,766	$ 543
Maintenance Labor Cost	1.74500%	of TEC less 11-14			$ 22,641,869	$ 591
Administrative & Support Labor	25%	of OL+ML			$ 10,859,659	$ 284
Property Taxes and Insurance	2%	of TPC			$ 58,357,979	$ 1,524
TOTAL FIXED OPERATING COSTS					**$ 112,656,271**	**$ 2,110**
VARIABLE OPERATING COSTS						
						$/bbl
Maintenance Material Cost	3.30%	of M&L Basis			**$ 42,818,434**	**$ 3.40345**
Consumables	Consumption		Unit	Initial Fill		$ -
	Initial Fill	/Day	Cost	Cost		$ -
Water (/1000 gallons)	0	1,806	1.67	**$ -**	**$ 990,963**	**$ 0.07877**
Chemicals		6				
Subtotal Chemicals				**$ 4,363,446**	**$ 18,204,530**	**$ 1.44700**
Other						
Butane (tons)	0	0	651.34	$ -	$ -	$ -
Gases, N_2 etc. (/100scf)	0	0	0.00	$ -	$ -	$ -
L.P. Steam (/1000 pounds)	0	0	0.00	$ -	$ -	$ -
Subtotal Other				**$ -**	**$ -**	**$ -**
Waste Disposal						
Spent Mercury Catalyst (lb.)	0	22320	0.65	$ -	$ 4,765,878	$ 0.37882
Flyash (ton)	0	0	0.00	$ -	$ -	$ -
Char (ton)	0	15,080	0.00	$ -	$ -	$ -
Subtotal-Waste Disposal				**$ -**	**$ 4,765,878**	**$ 0.37882**
By-products & Emissions (credit)						
Char (ton)	0	15080	28.23	$ -	$ (139,827,881)	$ (11.11430)
Supplemental Electricity (for sale) (MWh)	0	1,942	$ (60.00)	$ -	$ (38,284,287)	$ (3.04305)
Subtotal By-Products				**$ -**	**$ (178,112,168)**	**$ (14.15735)**
TOTAL VARIABLE OPERATING COSTS				**$ 4,363,446**	**$ (111,332,363)**	**$ (8.84932)**
Less By-product Credis					**$ 66,779,805**	**$ 5.30803**
ND Lignite (ton)	0	25,894	18.19		$ 154,726,785	$ 12
Switchgrass (ton)	0	29,714	50		$ 488,051,294	$ 39
Fuel (ton)		55,608			**$ 642,778,080**	**$ 51.09161**

7.2 Cost of Production

In order to estimate the cost of production, as per the FOA, a Capital-Charge Factor (CCF) of 0.237 was used to approximate the financial parameters for a commercial fuels plant. This corresponds to the financial parameters presented in Table 25.

Table 25. Financial parameters for calculating the cost of production.

Parameter	Value
Plant Operational Lifetime	30 years
Debt/Equity Ratio	50/50
Loan Term	15 years
Loan Rate	8%
Financing Fee	2.7% of TPC
Construction Period	5 years
Construction Escalation Rate	3.6%
General Escalation Rate	3%
Depreciation Schedule	20-year declining balance
Effective CCF	0.237

The results of the cost estimations are provided in Table 26. Table 26 is divided into 3 sections based on financial parameters. The first section is for the financial parameters listed in Table 25. The TASCs and ECO prices for the GHGR-CBTL cases are compared to those values from the or the DOE/NETL baseline CTL via Fischer-Tropsch Synthesis case with CO_2 sequestration, but with the TASC and ECO price calculated using the same CCF as shown in Table 25. These values were recalculated because the financial parameters used in that study are significantly more favorable (lower CCF, interest rate, etc.).

The second and third are cost estimations based upon the financial parameters utilized in other published DOE/NETL baseline studies. The second section utilizes the financial parameters from a DOE/NETL CBTL study.[24] The third section utilizes the financial parameters from the DOE/NETL baseline CTL via Fischer-Tropsch Synthesis study.[1] Therefore, this third section is the complement of the first section.

As can be seen in Table 26, the baseline case for the GHGR-CBTL system has a TASC and ECO price that are 58.5% and 69.4% that of the adjusted DOE/NETL baseline CTL via Fischer-Tropsch TASC and ECO price, respectively. If the compression configuration is utilized (Case CCCT), the TASC and ECO price are even more favorable at 58.9% and 65.8% that of the FT-Liquids Baseline Case, respectively.

Figure 37 shows a comparison of TOC per bbl/day for different cases and Figure 38 compares the TOCs and ECO prices for GHGR-CBTL with those of the DOE/NETL FT-Liquids baseline case. As shown both the GHGR TOC and ECO are lower than those for FT. Note that the RCCT, baseline case also captures CO_2 and, therefore, the comparisons in Figure 38 are for similar cases. The 38% lower TOC and 31% lower ECO should overcome the high cost associated with the FT process.

Table 26. GHGR-CBTL Cost of Production of JP-8 and No. 2 diesel.

Case	CO_2 Venting						CO_2 Capture								CO_2 Capture		CO_2 Capture Commercial Fuels	
	RLV	REV	RCV	CLV	CEV	CCV	RLCT	RECT	RCC	RCCT (Baseline)	CLCT	CECT	CCC	CCCT	RCC	RCCT	RCC	RCCT
Total Overnight Cost, $1000	3,670,236	3,670,236	3,670,236	3,564,685	3,564,685	3,564,685	3,820,420	3,820,420	3,820,420	**3,820,420**	3,713,557	3,713,557	3,713,557	3,713,557	3,820,420	4,067,017	3,820,420	3,820,420
Total Overnight Cost, $/bpd	94,372	94,372	94,372	91,658	91,658	91,658	98,234	98,234	98,234	**98,234**	95,486	95,486	95,486	95,486	98,234	104,575	98,234	98,234
Total Overnight Cost, $/bpd$_{ECO}$	74,294	74,294	74,294	72,157	72,157	72,157	77,334	77,334	77,334	**77,334**	75,171	75,171	75,171	75,171	77,334	82,325	77,334	77,334
Total Overnight Cost, $/bpd$_{EPJ}$	92,867	92,867	92,867	90,196	90,196	90,196	96,667	96,667	96,667	**96,667**	93,963	93,963	93,963	93,963	96,667	102,907	96,667	96,667
TASC Multiplier	1.1812	1.1812	1.1812	1.1812	1.1812	1.1812	1.1812	1.1812	1.1812	**1.1812**	1.1812	1.1812	1.1812	1.1812	1.14	1.14	1.1812	1.1812
Total As Spent Cost, $1000	4,335,283	4,335,283	4,335,283	4,210,606	4,210,606	4,210,606	4,512,680	4,512,680	4,512,680	**4,512,680**	4,386,454	4,386,454	4,386,454	4,386,454	4,355,279	4,636,400	4,512,680	4,512,680
Total As Spent Cost, $/bpd	111,473	111,473	111,473	108,267	108,267	108,267	116,034	116,034	116,034	**116,034**	112,789	112,789	112,789	112,789	111,987	119,215	116,034	116,034
Total As Spent Cost, $/bpd$_{ECO}$	87,756	87,756	87,756	85,232	85,232	85,232	91,347	91,347	91,347	**91,347**	88,791	88,791	88,791	88,791	88,160	93,851	91,347	91,347
Total As Spent Cost, $/bpd$_{IEPJ}$	109,695	109,695	109,695	106,540	106,540	106,540	114,183	114,183	114,183	**114,183**	110,989	110,989	110,989	110,989	110,200	117,314	114,183	114,183
Total annual fixed O&M, $1000	116,539	116,539	116,539	112,656	112,656	112,656	121,157	121,157	121,157	**121,157**	117,794	117,794	117,794	117,794	125,925	125,925	121,157	121,157
Total annual variable O&M (90% CF), $1000	68,321	68,321	68,321	66,780	66,780	66,780	70,610	70,610	70,610	**70,610**	69,038	69,038	69,038	69,038	74,177	74,177	70,610	70,610
Total annual feedstock cost (90% CF), $1000	644,485	695,821	647,809	635,132	691,305	642,778	515,507	571,783	488,708	**488,708**	515,507	571,783	479,421	479,421	488,708	488,708	488,708	488,708
Total annual power credit (90% CF), $1000	(71,207)	(71,207)	(71,207)	(38,284)	(38,284)	(38,284)	6,191	6,191	6,191	**6,191**	(38,284)	(38,284)	(38,284)	(38,284)	6,191	6,191	6,191	6,191
Total annual char credit (90% CF), $1000	0	(139,828)	(139,828)	0	(139,828)	(139,828)	0	(139,828)	(139,828)	**(139,828)**	0	(139,828)	(139,828)	(139,828)	(5,826)	(5,826)	(139,828)	(139,828)
Capital Charge Factor	0.237	0.237	0.237	0.237	0.237	0.237	0.237	0.237	0.237	**0.237**	0.237	0.237	0.237	0.237	0.16593	0.16593	0.218	0.218
COP Diesel, $/bbl$_{Diesel}$	128.0	121.0	117.2	127.4	120.8	117.0	139.9	133.3	114.1	**126.8**	134.0	127.5	107.5	120.2	104.0	107.2	108.4	121.1
COP JP-8, $/bblJP-8	126.8	120.0	116.2	126.3	119.8	116.0	138.7	132.2	113.2	**125.7**	132.9	126.4	106.6	119.2	103.1	106.3	107.5	120.1
Equivalent Crude Oil Price, $/bbl$_{ECO}$	**100.3**	**94.9**	**91.9**	**99.9**	**94.7**	**91.7**	**109.7**	**104.5**	**89.5**	**99.4**	**105.1**	**99.9**	**84.3**	**94.2**	**81.5**	**84.1**	**85.0**	**94.9**
Equivalent Petroleum Jet-Fuel Price, $/bbl$_{EPJ}$	125.4	118.6	114.9	124.9	118.4	114.7	137.1	130.7	111.9	**124.3**	131.3	124.9	105.4	117.8	101.9	105.1	106.3	118.7
Cost of CO_2 Capture, $/bbl$_{ECO}$	N/A	N/A	N/A	N/A	N/A	N/A	14.15	14.61	(3.65)	**11.35**	7.83	7.84	(11.24)	3.76	N/A	N/A	N/A	N/A
Reference Case	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**Adj. FTL**	**SRH15**	**SRH15**	**FTL**	**FTL**
Reference Equivalent Crude Oil Price $/bbl$_{ECO}$	131.8	124.7	131.8	131.8	124.7	131.8	143.3	143.3	143.3	**143.3**	143.3	143.3	143.3	143.3	111.2	111.2	133.8	133.8
Reference TASC	7,558,216	7,558,216	7,558,216	7,558,216	7,558,216	7,558,216	7,715,660	7,715,660	7,715,660	**7,715,660**	7,715,660	7,715,660	7,715,660	7,715,660	9,105,180	9,105,180	7,715,660	7,715,660
GHGR $/bbl$_{ECO}$ cost ratio	76.1%	76.1%	69.7%	75.8%	76.0%	69.6%	76.5%	72.9%	62.5%	**69.4%**	73.3%	69.7%	58.8%	65.8%	73.3%	75.6%	63.5%	71.0%
GHGR TASC ratio	57.4%	57.4%	57.4%	55.7%	55.7%	55.7%	58.5%	58.5%	58.5%	**58.5%**	56.9%	56.9%	56.9%	56.9%	47.8%	50.9%	58.5%	58.5%
GREET Model Lignite wt% Required	39.1%	26.7%	38.3%	41.3%	27.8%	39.5%	70.2%	56.6%	76.7%	**76.7%**	74.5%	58.5%	78.9%	78.9%	76.7%	76.7%	76.7%	76.7%
GREET Model Lignite HHV%	45.7%	32.0%	44.8%	48.0%	32.5%	46.1%	75.5%	63.1%	84.1%	**81.2%**	79.3%	64.9%	85.7%	83.1%	84.1%	81.2%	84.1%	81.2%

(a) GHGR-CBTL Results Calculated using economic parameters in the DOE/NETL CBTL report (*Production of Zero Sulfur Diesel Fuel from Domestic Coal: Configurational Options to Reduce Environmental Impact* , DOE/NETL-2012/1542, 12/2011)

(b) GHGR-CBTL Results Calculated using economic parameters in the DOE/NETL FTL report (*Cost and Performance Baseline for Fossil Energy Plants Volume 4: Coal-to-Liquids via Fischer-Tropsch Synthesis* , DOE/NETL-2011/1477, 10/15/2014)

(c) Adj. FTL = FTL study results recalculated with economic parameters from DE-FOA-0000981 - Attachment 1; SRH15 = CBTL study results; FTL = FTL study results

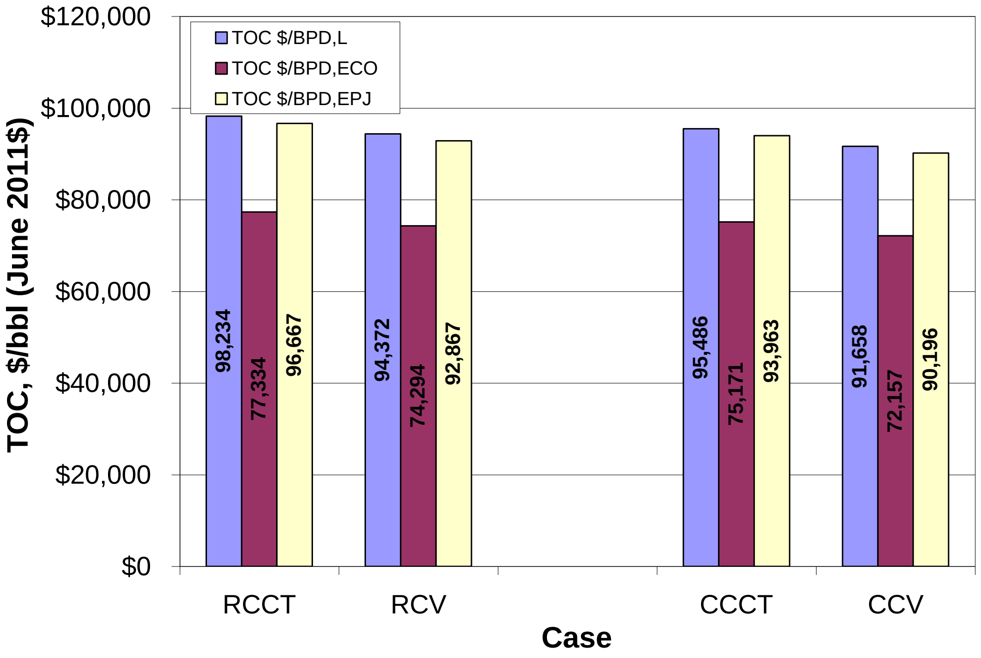


Figure 37. Comparison of TOC per bbl/day for different cases.

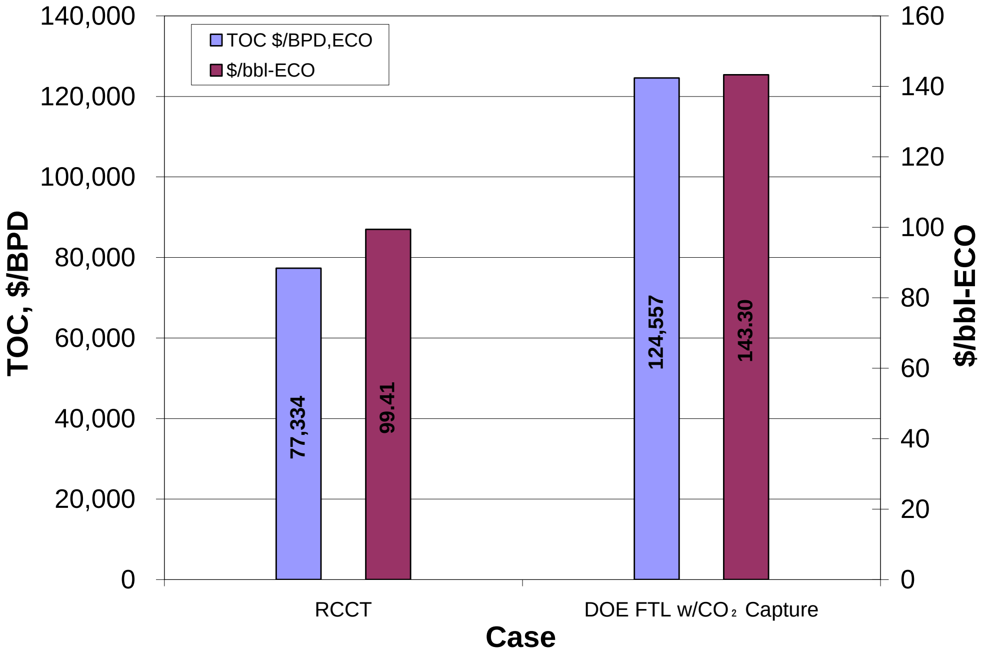


Figure 38. Comparison of TOCs and ECO prices for GHGR-CBTL and the DOE/NETL FT-Liquids baseline case.

7.3 Sensitivity Studies

7.3.1 Critical advanced technology performance parameters

Figure 39 and Figure 40 show summaries of the sensitivity analyses for cases RCCT (baseline) and CCCT, respectively. In both cases, the largest variance in ECO selling price are olefin yield, char yield, and CO_2 gate sales price. It should be noted that the positive impact (lower ECO selling price) of a non-zero CO_2 gate sales price is approximately the same as the potential negative impact (higher ECO selling price) of a 30% reduction in olefin yield from the baseline olefin yield. This holds for both the RCCT and CCT cases.

It should be noted that for the sensitivity analysis with respect to char sales price, the case for the char sales price of zero, a disposal cost of $36/tonne ($32.66/ton) was added to the variable operating costs in lieu of a product credit if the char is sold.

Also, even just considering the CO_2 gate sales price, the best-case scenarios bring the ECO selling price down to less than $60/bbl for both RCCT and CCCT plant configurations. Furthermore, it appears that a combination of loan guarantees and moderate increases in olefin and char yield could lead to an ECO selling price equivalent to current crude spot-market prices of $50/bbl.

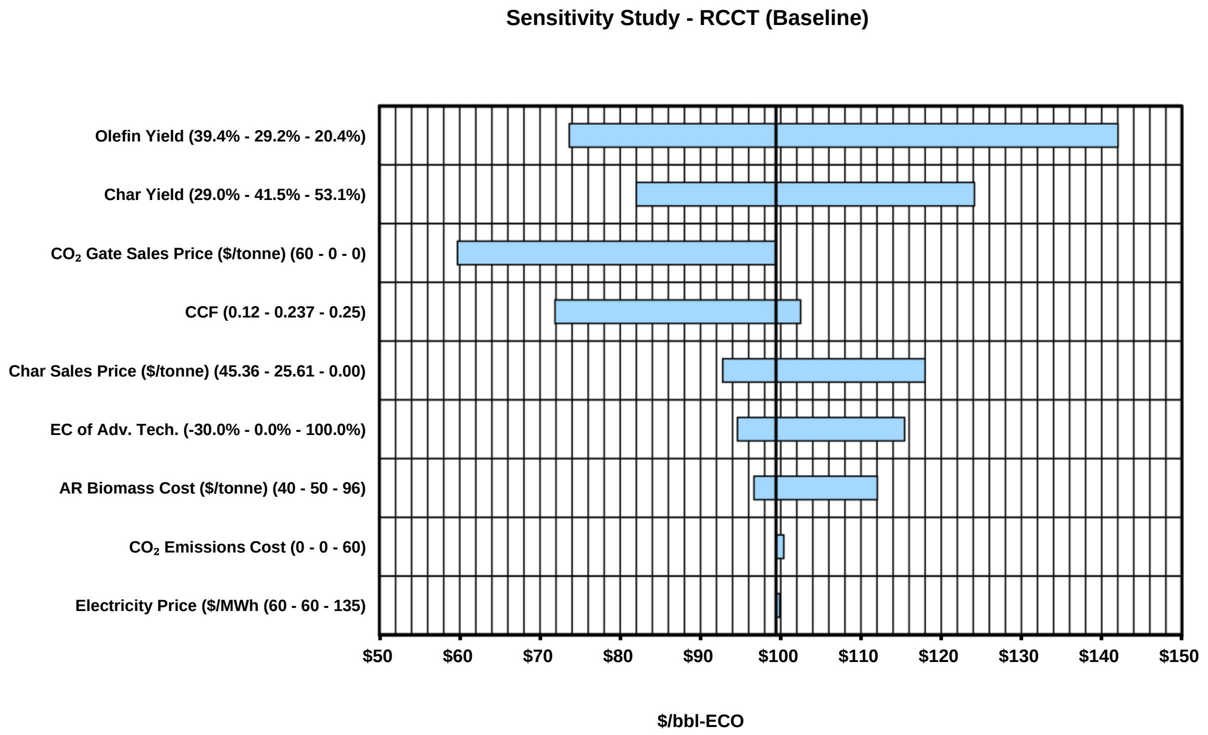


Figure 39. Summary of the sensitivity studies for case RCCT (baseline case).

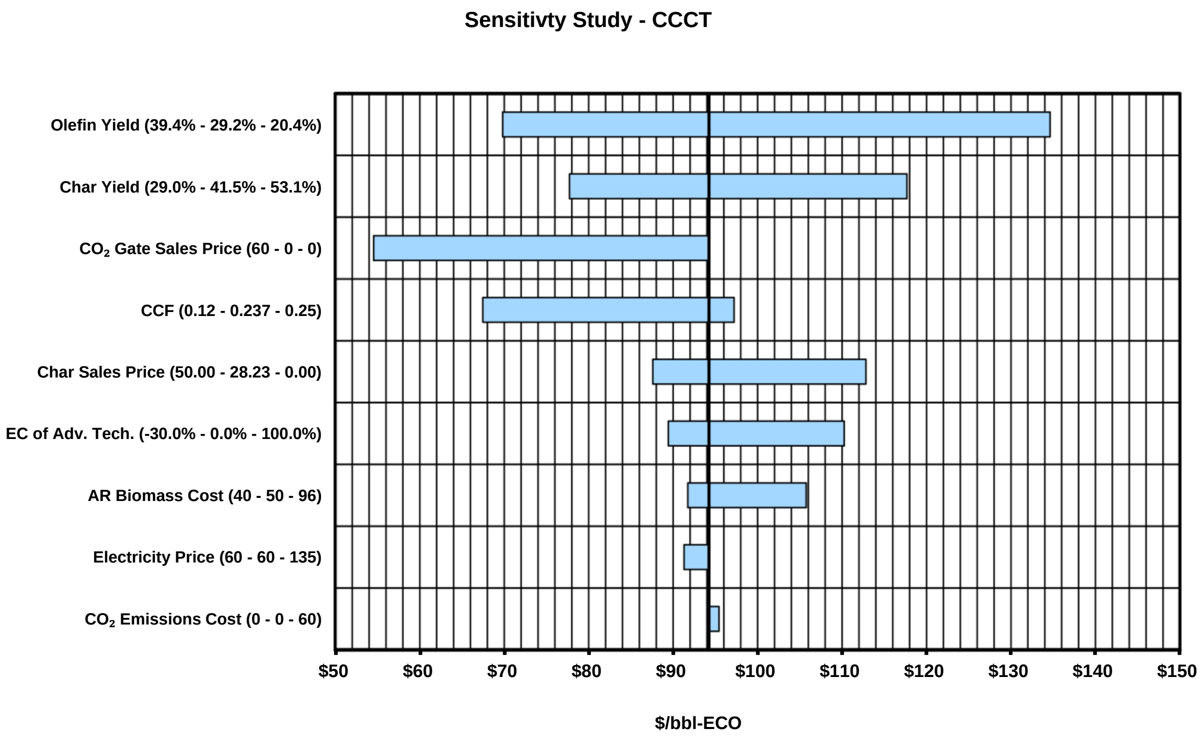


Figure 40. Summary of the sensitivity studies for case CCCT.

7.3.2 Cost of CO_2 Capture

In addition to the sensitivity studies above, the cost of CO_2 capture was also estimated. As with the DOE FT-Liquids Baseline Case, the cost of CO_2 capture was taken to be the difference between the ECO costs in the cases where CO_2 was captured and where it was vented:

$$\text{Cost of } CO_2 \text{ Capture} = ECO_{RCCT} - ECO_{RCV.}$$

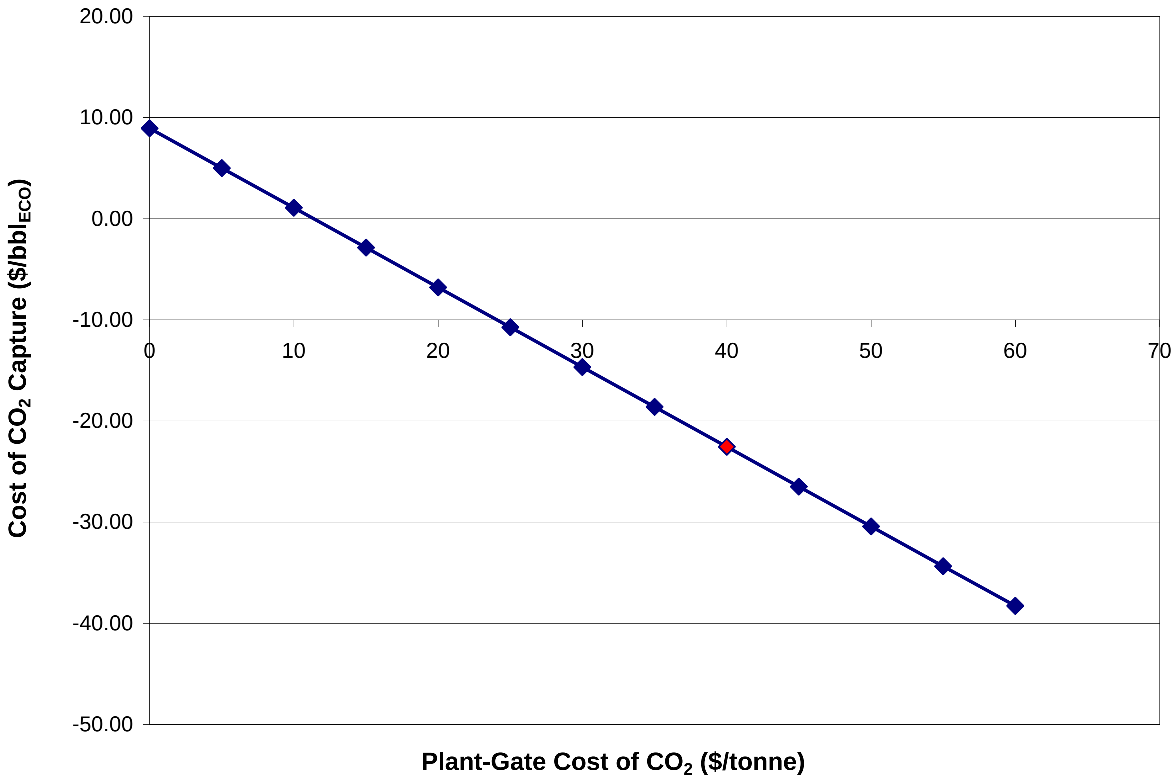


Figure 41. Sensitivity of the cost of CO_2 capture to plant-gate cost of CO_2. Transportation-and-storage costs of $15/tonne included.

It should be noted that since all of the cases where evaluated at a constant WTWa GHG emissions rate of 84 g-CO_2e/MJ, the CO_2-venting configuration (Case RCV) associated with the baseline configuration (Case RCCT) utilized a significantly higher percentage of biomass in the feed than the baseline capture configuration. Because the cost of biomass is significantly higher than that of lignite coal, this has a sizeable impact on CO_2-capture cost. In fact, this drives the CO_2 capture cost negative at plant-gate costs of CO_2 higher than about $25/tonne. Also, these costs include CO_2 Transportation-and-Storage (T&S) costs of $15/tonne. Therefore, without T&S costs, the CO_2-capture costs are negative even at $0/tonne plant-gate cost.

7.4 TEA Conclusions

The TEA demonstrates that the GHGR-CBTL process can significantly reduce the cost of CBTL technologies as compared to approaches utilizing gasification followed by Fischer-Tropsch synthesis. Not only is the ECO selling price 69% that of the DOE FT-Liquids Baseline Case, the TASC is only 59% of the DOE FT-Liquids Baseline Case. So not only can a GHGR-CBTL plant undercut the price of a FT-liquids plant, it also requires significantly less investment than a FT-liquids plant.

It should be noted that the ECO prices above reflect a 20% IRROE. To get an idea of the cost of production without the profit margin included, one can look at the ECO price and the equivalent petroleum jet (EPJ) price for the case where the IRROE is 0%. However, most project-finance terms require a minimum debt-service-coverage ratio (DSCR) of 1.4-1.5. Depending upon the debt:equity ratio and the loan terms, 0% IRROE may result in a minimum DSCR of less than 1.4. Presented in Table 27 is a comparison of 3 cases that were evaluated with the DOE/NETL Power Systems Financial Model (PSFM).

Table 27. Comparison of different scenarios for required IRROE or DSCR.

Scenario	$\$/\text{bbl}_{ECO}$	$\$/\text{bbl}_{EPJ}$	$\$/\text{gal}_{EPJ}$	IRROE (%)	Minimum DSCR
6/2009-11/2003 average wholesale price of jet fuel[a]	88.02[b]	110.02	2.620	-	-
20% IRROE	94.41	124.27	2.959	-	3.93
0% IRROE	52.73	65.91	1.569	-	0.60
Minimum DSCR=1.4	63.95	79.93	1.903	7.62	-
Parity with average wholesale jet-fuel price above	88.02	110.02	2.620	16.52	3.12

[a] Date range utilized in the DOE FT-Liquids Baseline Case for crude and diesel prices

[b] Calculated using the same methodology as ECO prices for the GHGR-CBTL process ($\$/\text{bbl}_{ECO} = 0.8\ \$/\text{bbl}_{EPJ}$).

The case for 20% IRROE was the baseline RCCT case with a CCF of 0.237 and resulted in a JP-8 sales *price* of $\$2.959/\text{gal}_{EPJ}$. The case for 0% IRROE resulted in JP-8 *cost* of $\$1.569/\text{gal}_{EPJ}$, but this resulted in a minimum DSCR much less than 1.4. The case where the minimum DSCR = 1.4 resulted in a JP-8 fuel *cost* of $\$1.903/\text{gal}_{EPJ}$.

It should be noted that the $\$2.959/\text{gal}_{EPJ}$ sales price compares favorably with the average 6/2009-11/2003 average wholesale price of US kerosene-type jet fuel of $2.620/gal[25]. As can be seen in Table 27, if the GHGR-CBTL sales price is set to the same value as the 6/2009-11/2003 wholesale price of jet fuel, the IRROE is 16.52% with a minimum DSCR of 3.12. While this IRROE is lower than the 20% IRROE recommended for a commercial fuels project by the DOE/NETL[26], the minimum DSCR is significantly higher than that typically required for a commercial project (1.4-1.5). This implies that financing terms may be more generous than those associated with a CCF of 0.237. At the average jet-fuel price in Table 27, a 20% IRROE can be obtained if the debt-to-equity ratio is changed from 50/50 to 60/40, the interest rate is reduced from 8% to 7%, and the loan term is increased from 15 to 30 years.

8 CONCLUSIONS, LESSONS LEARNED, AND RECOMMENDATIONS

8.1 Conclusions

This Final Technical Report describes the work and accomplishments of the project entitled, “Green-House-Gas-Reduced Coal-and-Biomass-to-Liquid-Based Jet Fuel (GHGR-CBTL) Process”. The main objective of the project was to raise the Technology Readiness Level (TRL) of the GHGR-CBTL fuel-production technology from TRL 4 to TRL 5 by producing a drop-in synthetic Jet Propellant 8 (JP-8) with a lower greenhouse-gas footprint than petroleum-based JP-8 by utilizing mixtures of coal and biomass as the feedstock. The system utilizes the patented Altex fuel-production technology, which incorporates advanced catalysts developed by Pennsylvania State University. The system was designed, a Block-Flow Diagram (BFD), a Process-Flow Diagram (PFD), and Piping-and-Instrumentation Diagrams (P&IDs) were produced, a Bill of Materials (BOM) and associated spec sheets were produced, commercially available components were selected and procured, custom components were designed and fabricated, catalysts were developed and screened for performance, and permitting activities were conducted. Although cost overruns on equipment and labor resulted in the system not being completely fabricated, enough data from the system-design activities and catalyst testing were available to conduct a Life-Cycle Analysis (LCA) of the Well-to-Wake (WTWa) greenhouse-gas emissions and to develop a Techno-Economic Analysis (TEA) of a full-scale GHGR-CBTL plant for comparison with competing technologies.

GHGR-CBTL is a transformative technology that overcomes the technical and economic barriers that exist in converting coal and biomass to a drop in fuel. The GHGR-CBTL removes the oxygen in its low temperature upstream pyrolysis process, but in contrast to the traditional pyrolysis-bio-oil approaches that require hydrotreating, it does not cool the condensable. But like the proven FT, the process uses gaseous intermediates to build the synthetic fuel, with the required properties to produce the desired synthetic fuel. These attributes increase the GHGR productivity and reduce the capital and operating costs when compared to classical gasification-FT or pyrolysis-bio-oil-hydrotreating.

Under the project a 1 BPD test system and a 40,000 BPD plan were designed. The test system was designed in detail including PI&D, BOM and equipment were selected. Some of the equipment were ordered and received. However, due to higher than planned cost of the equipment and plant engineering the system was not built and was not tested. However, the JP-8 conversion catalyst was identified and was fully tested at laboratory scale and its operating conditions were mapped.

Optimization tests for JP-8 production using C_2 olefin as the feed were performed over a range of temperatures (240 to 450°C), pressures (0 to 180 psig), and WHSVs (1 to 10 hr^{-1}). These tests were conducted with different catalysts; Catalyst P1, PSU-upgraded Catalyst 1 (Catalysts P1, P2, P3, P4), and two commercial catalysts. Detailed test results of C_2-oligomerization optimization are presented Figure ES-1.

Both Catalyst P1 and PSU-upgrdaed-Catalyst P1 catalysts give between 41 and 42% JP-8 yield with 65% JP-8 fraction and 63 to 65% liquid yield at 50 psig with C_2-olefins. Also, not shown in this figure, both these catalysts perform at 0 psig with lower JP-8 yield up to 5%. In comparison, the commercial catalyst gave similar JP-8 yield of 42% at a higher pressure of 115 psig and had zero liquid yield at 0 psig. Olefin conversion was >95% with Catalyst P1 and PSU-developed catalysts, while it was 83% for the commercial catalyst. The data were used design the oligomerization-reactor design for the TRL-5 system. Availability of the Altex-selected oligomerization catalysts for the TRL-5 system, and JP-8 optimized conditions of low pressures

(0 to 180 psig) versus existing industrial oligomerization at >500 psig makes the Altex-oligomerization for the TRL-5 system both low-risk for JP-8 production and energy-efficient.

Preliminary heat and mass balance data for a 50,000 BPD plant were used in a Life-Cycle Analysis (LCA) to estimate the Well-To-Wake (WTWa) life-cycle Greenhouse-Gas (GHG) emissions. The LCA was performed by Argonne National Laboratory (ANL) and a GHGR-CBTL module was added to the Greenhouse gases, Regulated Emissions, and Energy use in Transportation (GREET®) model. The LCA analysis showed that, based upon the preliminary heat and mass balances, different proportions of biomass would be required for the WTWa GHG emissions to match those of petroleum-based JP-8 (84 gCO_2e/MJ). If the char from the process was utilized to produce combined heat and power (CHP) with the electricity produced offsetting the electricity production elsewhere, then a feed containing 23 wt% biomass and 77 wt% lignite would be needed for parity with petroleum-based JP-8. If the char is not utilized for CHP, but sequestered in a land fill, more biomass is needed since no grid electricity is displaced, and 31 wt% biomass is required. Higher proportions of biomass in either case would result in JP-8 with lower WTWa GHG emissions.

To conduct the TEA, the plant was reconfigured to eliminate natural-gas and electricity imports in order to have a self-sufficient plant. This was accomplished by changing from using the waste-heat-recovery steam from the steam cracker for the production of utility steam to using it for the production of electricity. This led to changing from utilizing the waste-heat-recovery steam and combustion of imported natural gas for the production of utility steam to utilizing the combustion of excess fuel gas and some of the char for the production of utility steam.

The TEA was performed following DOE guidelines listed in the FOA and the referenced documents. A matrix of plant configurations was developed and evaluated. The scenarios included capturing or venting CO_2 in order to estimate the cost of CO_2 capturing. A baseline case with CO_2 capture was included and was compared with a similar FT baseline case that has been studied by DOE with similar assumptions.

The GHGR-CBTL TOC and ECO are 62% and 69% of the TOC and ECO for the DOE FT-Liquids Baseline Case, respectively. This shows that the economics of the GHGR-CBTL process are significantly better than a gasification/FT process.

While the 1-TPD system was not fully completed and tested, the previous TRL 4 testing that has been performed on coal and biomass, the detailed design of the 1-TPD system, the tests that were completed, the detailed design of a full-scale plant all support the feasibility of the GHGR-CBTL process.

It is important to note that the GHGR-CBTL process utilizes unit operations that have already been proven and have been put into commercial operation. In particular, ethylene production from the steam cracking of hydrocarbon feedstocks, including heavy feedstocks, is one of the largest chemical industries in the world. This demonstrates that the process and the equipment associated with the process are readily available at the large scales needed for significant fuel production. Although the process needs to be adapted to utilizing pyrogas rather than petroleum-based hydrocarbons, tests to date have demonstrated the production of olefins from pyrogas.

Even more directly related is oligomerization of olefins to liquid fuels. This process is well known and has been commercialized for utilization with both heterogeneous and homogeneous catalysts (*e.g.,* Shell Higher Olefin Process, Dimersol processes), although these operate at higher pressures (>550 psig) than the GHGR-CBTL catalysts. This means that there are drop-in replacements for the GHGR-CBTL oligomerization process that are commercially available.

Utilizing two processes that are so well developed and commercialized greatly reduces the risk of the GHGR-CBTL process.

While the system was not fabricated and tested, major efforts were expended to design the 1-TPD and a full-scale plant. No technical barriers were identified. The TEA was applied to the full scale plant followed DOE guidelines and the lower cost of the GHGR-CBTL than FT is consistent with the GHGR-CBTL process simplicity and its lower temperatures. These lower costs and the detailed design that was performed under this project are being used by Altex to attract funding partners to move the GHGR-CBTL development forward.

8.2 Lessons Learned

The major lesson learned was that at this intermediate scale, the cost of process equipment can be significantly more than one would expect. This is due mainly to the fact that at this scale, equipment is larger than readily available off the shelf, but significantly smaller than typical small petrochemical facilities. This leads to a significant amount of the process equipment needing to be custom fabricated. This involves not only custom materials and fabrication, but also the engineering effort needed to design the custom equipment.

Additionally, because the size of this system is significantly larger than many other systems Altex has produced and tested in the past, the City of Sunnyvale required significant permitting. The time and effort required to learn and to conduct the permitting process required by the city added additional costs. This process required conducting additional analyses of the facilities, process, and system as well as coordinating with subcontractors that are authorized to sign off on component drawings, system drawings, facility drawings, fire-safety plans, hazmat plans. While many of these activities were covered by Altex, formalizing these plans in the format the city required resulted in additional effort. On the positive side, this process has allowed us to better understand the city requirements and processes for future projects.

8.3 Recommendations

The GHGR-CBTL process has the potential to drastically undercut both the capital investment and the fuel price of competing CBTL technologies (*i.e.*, gasification plus Fischer-Tropsch synthesis). This was shown by the TEA, which utilized realistic economic assumptions coupled with estimates of feedstock costs, operating costs, and a Class 4 capital cost estimate.

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