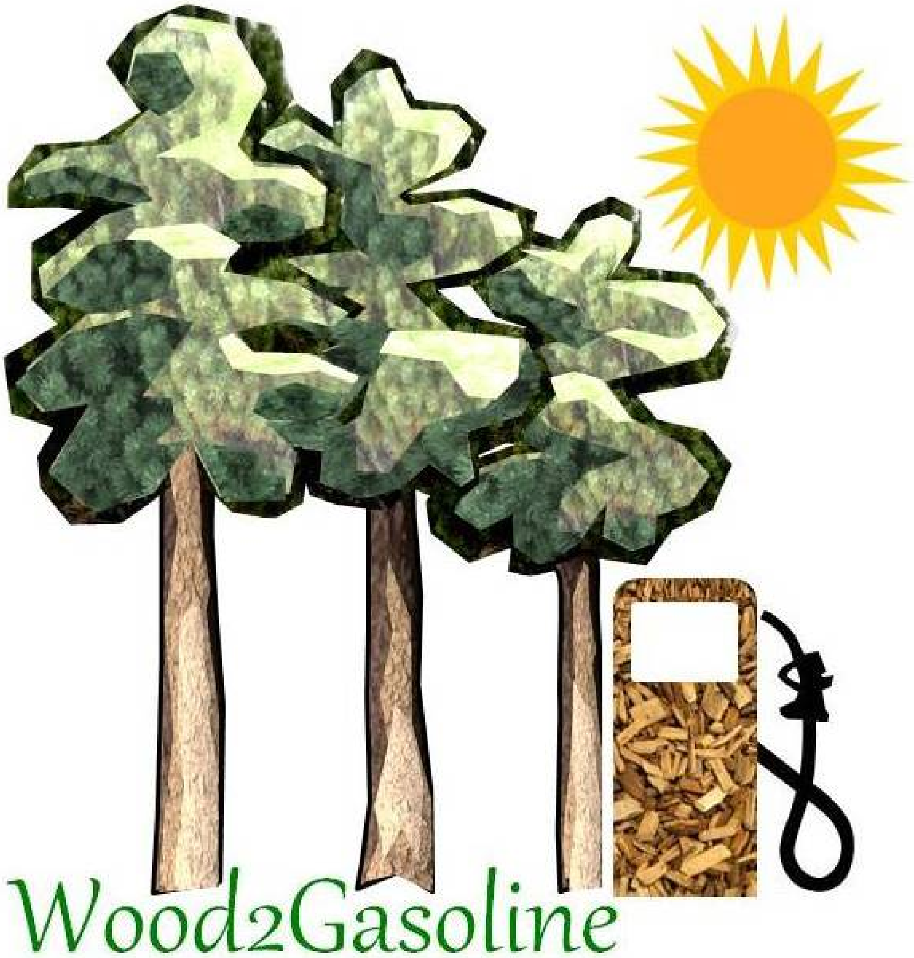
Green Gasoline from Wood using Carbona Gasification and Topsoe TIGAS Process

DE-EE0002874

Final Report

February 19, 2015



UPM PHILLIPS 66 gti ANDRITZ Pulp & Paper HALDOR TOPSØE

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List of abbreviations and definitions:

AC	Andritz-Carbona
AGR	Acid Gas Removal
ARRA	American Recovery and Reinvestment Act
ASU	Air Separation Unit
BD	Bone dry
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BP	Budget Period
BP1	Budget Period 1
BP2	Budget Period 2
BP3	Budget Period 3
CAPEX	Capital expenses
CFR	Code of Federal Regulations
CPCHEM	Chevron Phillips Chemical Company LLC
DME	Dimethyl ether
DMW	Demineralized Water
DOE	Department of Energy
EIA	Energy Information Administration
EPA	Environmental Protection Agency
FAA	Federal Aviation Administration
FIA	Forest Inventory and Analysis
FTP	Federal Test Procedure
GHG	Green House Gas
GTI	Gas Technology Institute
GUU	Gasoline Upgrading Unit
HAZOP	Hazards & Operability
HGF	Hot gas filter
HRSG	Heat Recovery Steam Generation
HTI	Haldor Topsoe, Inc.
IBR	Integrated Bio Refinery
IGCC	Integrated Gasification and Combined Cycle
ISBL	Inside Battery Limit (i.e. one of the licensed process units)
LCA	Life Cycle Analysis
LPG	Liquefied Petroleum Gas
NEPA	National Environmental Policy Act
OPEX	Operating expenses
OSBLPM	Outside Battery Limit (i.e. not one of the licensed process units)
PM	Project Manager
REC	Rapid Energy Center
RVP	Reid Vapor Pressure
SIS	Safety Instrumented System
SwRI	Southwest Research Institute
TIGAS	Topsoe Improved Gasoline Synthesis
TRC	Transportation Research Center, East Liberty, Ohio
UDDS	Urban Dynamometer Driving Schedule
UHGF	Ultra-hot gas filter
WBS	Work Breakdown Structure

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- Block flow diagram for wood to gasoline pilot plant (22.5 bpd)
- Block flow diagram for commercial size wood to gasoline plant (4,106 bpd)

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1. Introduction – Project Participants and Roles

This final report presents the results of a four-year technology demonstration project carried out by a consortium of companies sponsored in part by a $25 million funding by the Department of Energy (DOE) under the American Recovery and Reinvestment Act (ARRA).

The purpose of the project was to demonstrate a new, economical technology for the thermochemical conversion of woody biomass into gasoline and to demonstrate that the gasoline produced in this way is suitable for direct inclusion in the already existing gasoline pool.

The process that was demonstrated uses the Andritz-Carbona fluidized-bed steam-oxygen gasification technology and advanced tar reforming catalytic systems to produce a clean syngas from waste wood, integrated conventional gas cleanup steps, and finally utilizes Haldor Topsoe's (Topsoe) innovative Topsoe Improved Gasoline Synthesis (TIGAS) syngas-to-gasoline process. Gas Technology Institute (GTI) carried out the bulk of the testing work at their Flex Fuel development facility in Des Plaines, Illinois; UPM in Minnesota supplied and prepared the feedstocks, and characterization of liquid products was conducted in Phillips 66 labs in Oklahoma. The produced gasoline was used for a single-engine emission test at Southwest Research Institute (SwRI) in San Antonio, TX, as well as in a fleet test at Transportation Research Center (TRC) in East Liberty, Ohio. The project benefited from the use of existing pilot plant equipment at GTI, including a 21.6 bone dry short ton/day gasifier, tar reformer, Morphysorb® acid gas removal, associated syngas cleanup and gasifier feeding and oxygen systems.

The project team members are among the world leaders in their fields of expertise:

- UPM – one of the world's leading forest industry groups - contributed to the project with expertise in management of forest, from growth through preparation and delivery. Their Minnesota paper mill is a prime candidate to be the site of the first commercial plant;
- GTI and Andritz have developed, licensed, and operated the Andritz-Carbona gasifier, one of the most suitable approaches for the gasification of biomass, also contributing commercial scale expertise in biomass feeding, gasification and gas cleaning systems including acid gas removal;
- Topsoe - one of the world's leading catalyst and technology providers in the manufacture of synthesis gas and synthesis gas products - contributed to the project with the TIGAS process as well as all the catalysts;
- Phillips 66 is a growing energy manufacturing and logistics company. They are the only integrated downstream company to combine leading midstream, chemicals, refining, and marketing and specialties businesses.

The team members were chosen to ensure that the whole concept, from growth of feedstock to sale of gasoline, is covered so that uncertainty risk in technical and strategic considerations is minimized. Collectively the team has extensive experience with operating pilot-scale or demonstration-scale facilities, and track record in successfully scaling-up and commercialization of new technologies. This commercial experience is evidenced by the technology stature, financial standing and growth of the team members.

The following section discusses the partner's organization, how their business fits with the commercialization plan, and how their participation gives the new venture advantages over competition. The section presents core competencies and experience relevant to commercialization.

Haldor Topsoe: Topsoe is a $1 billion global company specializing in catalyst and technology. Topsoe's core businesses are related to the manufacture of commodity chemicals (ammonia, methanol, dimethyl ether), fuels manufacture and refinement (hydrogen, hydrotreating, hydrocracking, substitute natural gas), and the environment (DeNOx for the removal of nitrogen oxides from combustion exhaust, Wet Sulfuric Acid for the removal of sulfur from exhaust gas and the revalorization of sulfur as saleable sulfuric acid, diesel DeNOx for the after treatment of the exhaust from diesel engines). Topsoe's strategy focuses on the development of economically attractive processes encompassing any feed-stock that allows economically self-sustainable processes. The interest of Topsoe in the utilization of biomass as a feed-stock is - in addition to participating and cost sharing in this project - testified in Topsoe's participation in the bioDME pilot project in Piteå, Sweden, in which Topsoe in a consortium of six other companies has demonstrated a breakthrough DME process based on black liquor as a feed-stock, where DME is now used as a fuel for diesel trucks.

Topsoe has established a manufacturing line for the gasoline catalyst in Houston, Texas.

Andritz: Andritz is the leading equipment supplier to the pulp and paper industry. The company has annual revenues in excess of $5.3 billion and devotes 3% of sales to research and development. Andritz has several offices and 1,600 employees in the U.S. with major facilities in Georgia and Pennsylvania. It has a diversity of business interests, including biomass gasification experience since the 1980's as Ahlstrom Machinery Oy. Andritz is a majority owner of Carbona, Inc., a specialist in gasification systems. The company has been developing biomass gasification systems since 1996. Recently it has, in partnership with UPM, invested over $10 million at GTI to develop oxygen gasification of biomass and gas clean-up to produce syngas suitable for further conversion into biofuels. The first biomass gasification plant using Andritz-Carbona technology to produce fuel gas is now in operation in Denmark. Other gasifier applications are ready for market including lime kiln fuel gas production. Continuing gasification development activities include development of pressurized gasifiers for liquid biofuels and upgrading of the steam parameters in recovery boilers and integrated gasification combined-cycle plants (IGCC) for improved electrical efficiency.

UPM: UPM is a global paper and forest products company with annual revenues approaching $14 billion. The company is one of the world's top integrated paper, pulp and forest products manufacturers. UPM has leading market positions in most of its businesses, from forestry products, energy and pulp to printing and label papers.

Fiber- and biomass-based businesses, recyclable raw materials and products as well as resource efficiency are cornerstones of UPM's business. UPM integrates bio and forest industries and builds a sustainable future.

UPM is vertically integrated and controls most of its raw materials. The company is largely self-sufficient in terms of chemical pulp, wood and electrical power. On average, UPM sources about one tenth of the wood used annually from its own forests or from forests where it has felling rights. UPM is the biggest private forest owner in Finland and owns forestland in the US, the UK and Canada. The company produces a significant proportion of its own electricity and the use of biofuels such as bark, forest-based fuels and deinking sludge in the generation of electricity and heat has been steadily increasing.

UPM uses fiber and forest biomass in its current products and the company's aim is to create new growth opportunities based on continuous product development and innovation. Biofuels for use in transportation are a topical example of UPM's innovation work. The new renewable diesel - UPM BioVerno - biorefinery in Lappeenranta, built alongside the UPM Kaukas pulp, paper and saw mill integrate is now in operation. Its annual production capacity is 100,000 tons of fuel produced from tall oil, a wood-based residue of pulp production. UPM BioVerno will significantly decrease the greenhouse gas emission compared to fossil fuels.

UPM's new business group, UPM Biofuels, is investigating the production of liquid biofuels, which complements their experience in using biofuels for power production. UPM has already launched pilot projects in various biofuel technologies. In the U.S., it has concluded piloting of second generation diesel from wood raw material. Another pilot project focuses on catalytic pyrolysis technology, which is being studied in Finland.

Phillips 66: Phillips 66 is a growing energy manufacturing and logistics company. They are the only integrated downstream company to combine leading midstream, chemicals, refining, and marketing and specialties businesses. With this diverse portfolio, Phillips 66 is uniquely positioned to capture opportunities of the changing energy landscape.

The Midstream segment includes Phillips 66's transportation business, including operations of Phillips 66 Partners LP, their master limited partnership which conducted an initial public offering in July 2013; a 50% interest in DCP Midstream, LLC; and natural gas liquids (NGL) operations. DCP Midstream is the largest NGL producer and one of the largest natural gas gatherers and processors in the United States, with 63,000 miles of pipeline, 62 plants and 12 NGL fractionators.

Phillips 66 conducts their Chemicals business through a 50% interest in Chevron Phillips Chemical Company LLC (CPChem), one of the world's top producers of olefins and polyolefins with more than 33 billion pounds of net annual chemicals processing capacity across its product lines.

Phillips 66 Refining operations include 15 refineries with a net crude oil capacity of 2.2 million barrels per day. Their Marketing and Specialties business includes 10,000 owned or supplied outlets, lubricants marketing in more than 70 countries, and other specialty products including petroleum coke, waxes, solvents, polypropylene and pipeline flow improvers.

Phillips 66 has a number of strategic initiatives in biofuels. The company has produced renewable diesel both in the U.S. and in Ireland. They also became a founding member of the Colorado Center for Biorefining and Biofuels and helped establish a biofuels research program at Iowa State University. Phillips 66's emerging technologies program monitors developments and trends in alternative energy sources, and invests in new business development spanning activities from lab-bench work through commercialization. Recent laboratory work investigated the properties of gasoline produced from waste wood feedstock. The company's investments in new technology are strategic, fit with core competencies, and consistent with investment goals.

Gas Technology Institute: GTI is a nonprofit research, development and training organization that serves energy markets by developing technology-based solutions for consumers, industry and government. The organization provides products, services and information that help customers solve problems or capitalize on opportunities related to finding, producing, delivering and using natural gas. More than 800 patents and approximately 400 products have resulted from GTI-led technology developments, including gasification and gas clean-up technologies. Andritz, a GTI research and development partner, has been collaborating with GTI and UPM on the development of waste wood gasification systems. The Flex Fuel development facility, where gasification development work is conducted, and pilot plant equipment from this project will be important assets in the commercialization process as potential customers can perform trials on various feedstock and gain insights into plant costs and operations.

Table 1.1 below shows each participant's role and contribution to the biofuels value chain in the project.

Table 1.1 Project Participants and their Role in the Project

Process Step	**Wood Supply**	**Gasification**	**Gas Cleanup**	**AGR**	**Gasoline Synthesis**	**Product Validation**
Technology		Andritz	Andritz	GTI	Topsoe	
Catalyst			Topsoe		Topsoe	
Demonstration	UPM	GTI	GTI	GTI	GTI	Phillips 66

2. Summary of Project Goals and Actual Accomplishments

The goal of the project was to demonstrate a new, economical technology for the thermochemical conversion of woody biomass into renewable gasoline, applicable to many U.S. sites and which will ultimately create thousands of U.S. jobs. This technology utilizes the ample, domestic renewable biomass resources in the U.S. to create transportation fuels, sufficient in quantity and quality to displace a significant volume of foreign crude oil. Thus, this technology offers a path to genuine energy independence for the U.S., along with the creation of new, large workforces to plant, grow, harvest, and process dedicated biomass crops into fungible fuels. Commercialization of this technology will also reduce U.S. greenhouse gas emissions (GHG) from transportation fuels, since the life cycle analysis (LCA) shows that the renewable gasoline product has 65-74% lower GHG emissions than petroleum-based gasoline, depending on process options selected for power and steam.

The process uses the Andritz-Carbona fluidized-bed steam-oxygen gasification and advanced tar reforming catalytic systems to produce a clean syngas from a wide range of wood resources, integrated conventional gas cleanup steps, and finally utilizes Topsoe’s innovative TIGAS syngas-to-gasoline process. GTI carried out the bulk of the testing work in Des Plaines, Illinois, UPM in Minnesota supplied and prepared the feedstocks, and characterization of liquid products was conducted in Phillips 66 labs in Oklahoma. The produced gasoline was used for single engine emission test at SwRI in San Antonio, TX, as well as in a fleet test at TRC in Ohio. The project benefited from the use of existing pilot plant equipment, including a 21.6 bone dry short ton/day gasifier, tar reformer, Morphysorb® acid gas removal, associated syngas cleanup and gasifier feeding and oxygen systems at GTI.

This project was perfectly suited for funding by ARRA. The total market opportunity for “waste wood to gasoline” technology, assuming all forestry residue, mill residue and wood waste feedstock are processed by this route, is more than 10 billion gallons per year of renewable gasoline. This reduces greenhouse gas emissions by up to 73 million short tons per year, and creates and preserves thousands of new jobs.

The project was originally planned to be performed in three budget periods (BP), but BP2 (Pilot plant installation and testing) and BP3 (Technology evaluation) were from project start combined into a common BP2. During BP1, a detailed design and cost estimate (±10% cost estimate) was made to fully define the project to gasify a nominal 21.6 bone dry short ton/day of biomass, clean the resulting syngas, and convert it to renewable gasoline. Also during BP1, the NEPA documentation was completed, the Risk Management Plan was updated, and the application for BP2 was submitted. Project risks were assigned to four categories: (1) Project Execution; (2) Project Cost; (3) Technical Performance/Operability; and (4) Environment, Health, and Safety. A Risk Matrix, shown in Table 2.1, was constructed to define the impact and probability for each of the categories, and the product of the two constituting a Risk Score. From there, a Risk Register was devised for each category, in which the Risk Score was matched with Mitigation measures, Contingency measures in cases where mitigation might not be sufficient, the owning organization of that risk, and the current status. These Risk Registers were updated regularly through discussions among the team project managers, in order to guide the management of these risks. One example of the final Risk Register is given in Table 2.2. Any Risk Score of 6 or higher required a Contingency measure in case the existing Mitigation was unable to fully mitigate the risk. A status of "OK" meant that the risk was being managed adequately, but not yet "Closed."

Table 2.1 Risk Matrix

	PROBABILITY			
IMPACT	4 (>75%)	3 (51-75%)	2 (26-50%)	1 (<25%)
4 (critical)	16	12	8	4
Project execution: >100% schedule slip, team disintegration				
Project cost: >25% cost escalation				
Technical performance: Major shortfall on 2+ technical targets; testing stops				
Environment, health, & safety: Severe/ multiple moderate injury or health threat; major environmental release				
3 (serious)	12	9	6	3
Project execution: >50% schedule slip, team member(s) quit, chronic late deliverables				
Project cost: 10-25% cost escalation				
Technical performance: Major shortfall on 1/ moderate shortfall on 2+ technical targets; testing interrupted				
Environment, health, & safety: Moderate/ multiple minor injury or health threat; moderate/brief environmental release				
2 (minor)	8	6	4	2
Project execution: >25% schedule slip, team member(s) disengaged, a few late deliverables				
Project cost: 5-10% cost escalation				
Technical performance: Moderate shortfall on 1/ slight shortfall on 2+ technical targets				
Environment, health, & safety: Minor lost-time injury or health threat; minor environmental release				
1 (incidental)	4	3	2	1
Project execution: <10% schedule slip, occasional duplication, rare late deliverables				
Project cost: <5% cost escalation				
Technical performance: Slight shortfall on 1 technical target				
Environment, health, & safety: Minor injury (no lost time); minor irritant; very limited minor environmental release				

Table 2.2 Example Risk Register

Risk name	Impact	Prob	Score	Mitigation	Contingency	Owner	Status	Comments
Labor costs exceed estimates	3	2	6	Implement labor tracking & forecasting	Subsitute lower-cost/ contract labor if possible	All PM's	Closed	Adjust data requirements and test objectives (e.g., fewer samples, terminate test upon meeting key objectives)
Subcontractor/vendor cost growth	4	1	4	Formalize change control & incentives. Conduct bid review meetings.	Maintain alternate subcontractors & vendors	HTAS, GTI	Closed	
Inter-team scope disagreements	3	1	3	Regularly review & update scope definitions		HTI	OK	
Externally imposed scope growth	3	1	3	Review/ clarify scope in regular teleconferences		HTI	OK	
Unanticipated regulatory costs	4	1	4	Maintain up-to-date information from regulatory consultant(s)	Bring in consultant support as needed	HTI, GTI	Closed	
Materials/utilities cost growth	2	2	4	Maintain up-to-date utility rate database	Consider scope revision, value engineering remedy	GTI	Closed	Repair/upgrades of some unit operations (hot filter, flange leaks, etc.) require additional cost
Quality control problems	1	3	3	Conduct regularly scheduled meetings w/ work updates		All PM's	OK	
Late-stage design package changes	4	1	4	Prepare/ conduct preliminary design & HAZOP reviews	Consider value engineering remedy, scope revision	All PM's	Closed	
Duplicative effort	2	3	6	Conduct regularly scheduled meetings w/ work updates	Focused meeting of involved team PMs	All PM's	OK	
TIGAS decommissioning responsibility unclear	2	1	2	TIGAS decommissioning/ removal spelled out contractually		HTI, GTI	OK	

In BP2, the modifications in GTI’s existing Flex-Fuel Testing Facility and gas clean-up equipment were carried out, the TIGAS unit was constructed and installed, liquid product handling facilities were added, and the initial commissioning and subsequent three test campaigns including the Independent Engineer Test were carried out along with all analytical requirements and liquid product sample collection for offsite testing by Phillips 66.

The preparation and delivery of wood for the testing was handled by UPM. The produced gasoline was successfully used for single engine emission test and fleet testing. Finally the detailed technical analysis and process evaluation was conducted.

The project successfully met the six DOE Go/No-Go requirements throughout the project execution:

1.	November 2010	Approval of preliminary design package
2.	April 2011	Approval of detailed design package
3.	May 2011	Approval of BP2 cost estimate
4.	September 2011	Approval of start of procurement
5.	November 2012	Approval of proceeding with shakedown
6.	October 2013	Approval of proceeding with test campaign

More than 10,000 gallons of renewable gasoline were produced. The performance targets for the various process units were met including the predicted efficiency and product quality.

An 80/20 green gasoline/commercial gasoline blend successfully completed a single engine emission test as required for EPA registration. Additionally, there was a comprehensive fleet test with eight vehicles, three pairs of cars and a pair of pick-ups. One vehicle in each pair operated on a 50/50 green gasoline/commercial gasoline blend with 10% ethanol[*] while the other vehicles operated on commercial gasoline with 10% ethanol. Each vehicle successfully completed 75,000 miles of fleet testing. Subsequently an application for EPA fuel registration based on the 80/20 blend has been initiated.

The project was successfully completed within the original budget and schedule. The comparison between budget and actual cost as well as the split between DOE funding and cost share is shown in Table 2.3.

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[*] Both green gasoline and commercial gasoline included 10 % ethanol and a standard additive package.

Table 2.3 Budget Versus Actual Cost

Project Task	Budget	Actual	Deviation
Preparation of Award Activities	$0	$133,555	N/A
Independent Project Review	$0	$68,733	N/A
Preliminary Process Design	$1,205,924	$1,043,401	-13.5%
Engineering Design	$1,742,917	$1,592,300	-8.6%
BP1 Project Management & Reports	$746,912	$928,019	+24.3%
Budget Period 1 – Total	***$3,695,753***	***$3,766,008***	***+1.9%***
Procurement & Installation	$13,686,888	$14,718,877	+7.5%
Commissioning & Shakedown	$2,928,052	$4,116,485	+40.6%
Independent Engineer Test	$497,320	$155,179	-68.8%
Test #1	$1,867,372	$2,471,723	+32.4%
Test #2	$2,483,404	$3,167,864	+27.6%
Test # 3	$2,825,982	$2,069,294	-27.9%
Single Engine Emission Test	$300,000	$563,872	+88%
Fleet Testing	$1,600,000	$$1,174,450	-26.6%
Technology Readiness	$494,811	$48,961	-90.1%
BP2 Project Management & Reports	$4,009,196	$2,390,136	-39.6%
Budget Period 2 – Total	***$30,693,025***	***$30,876,841***	1.0%
Total Project	***$34,388,778***	***$34,642,849***	***0%***
DOE Funding	***$25,000,000***	***$25,000,000 (72.7%)***	***0%***
Project Team Cost Share	***$9,388,778***	***$9,642,849 (27.3%)***	***+2.7%***

The original budget did not include expenses for project activities requested by DOE during the execution of the project such as Independent Project Review questionnaire and seminar as well as extensive post award negotiations with DOE.

During the initial phase of the project it was agreed with DOE to add at no additional project cost the single engine emission test and to submit an application to EPA to register the renewable gasoline blend as well as the fleet test of a blend of the green gasoline and conventional gasoline. Such activities are required for the future registration of the renewable gasoline as transportation fuel. These new activities were funded under the Phillips 66 cost share by shifting Phillips 66 activities foreseen during the design and operation of the pilot plant to the other project team members.

The technology is now ready for a 3200 dry short ton/day commercial facility, offered through Andritz and Topsoe. The demonstrated business, based on the commercialization of the demonstrated “wood waste to gasoline” biorefining technology, has the potential to contribute 1 billion gallons per year of biofuels to the transportation fuel supply by the year 2022 and 2 billion gallons per year by 2028. Twelve plants of this design could contribute enough gasoline in 2022 to satisfy more than 6% of the cellulosic fuel goal in the renewable fuel standard.

3. Schedule Summary

This chapter summarizes the initial and final project schedules.

Figure 3-1 is the Gantt chart for the original schedule. Task 2.4, denoted by the open bars, was not included in the original work plan.

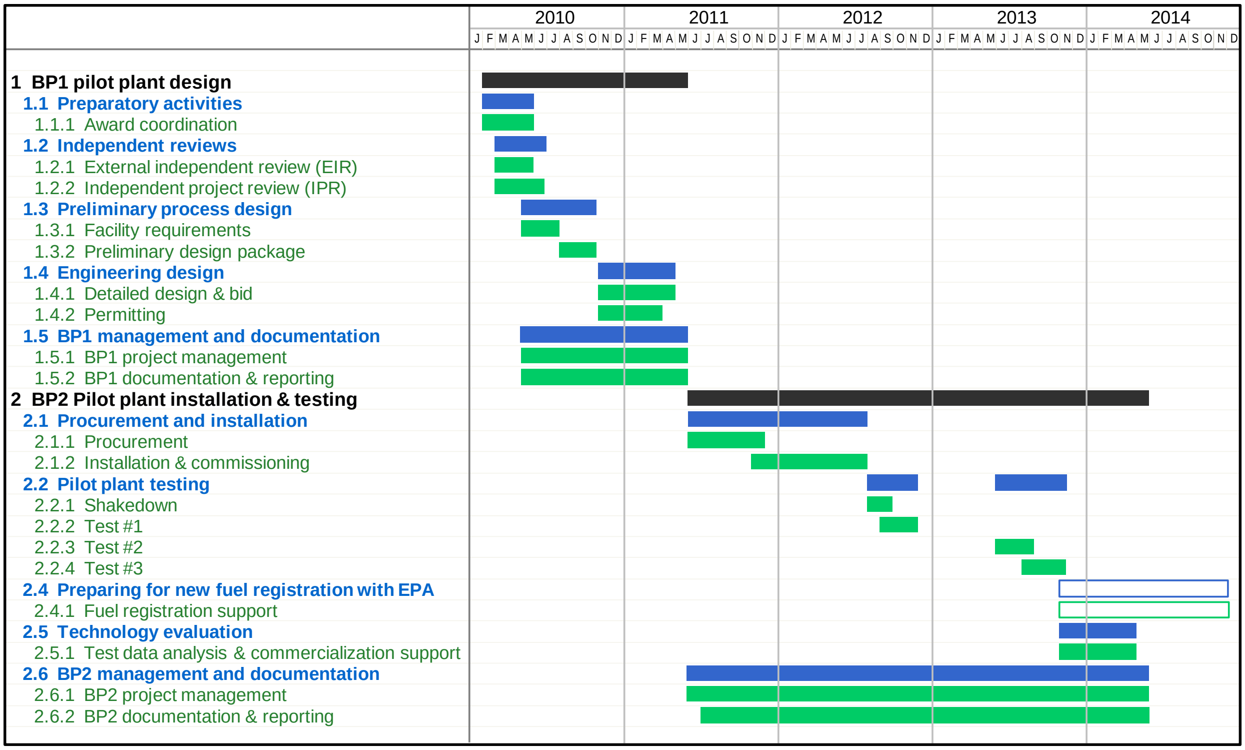

Figure 3-1 Initial WBS Gantt Chart

Figure 3-2 shows the same schedule with the actual performance periods overlaid on the initial. Some Task 2.1 Procurement and Installation slippage in 2012 was associated with a delay in approval of the BP2 application, which was needed to proceed with procurement, followed by a delay in obtaining the FAA Determination of No Hazard (see Chapter 5).

Slippage in Task 2.2 Pilot Plant Testing followed the Task 2.1 schedule, and the project end date was advanced from May 31, 2014 to December 31, 2014 to accommodate these delays, as well as a privately funded moderate fleet test of 50% biogasoline that Phillips 66 was carrying out as a follow-on to Task 2.4 Preparing for New Fuel Registration with EPA.

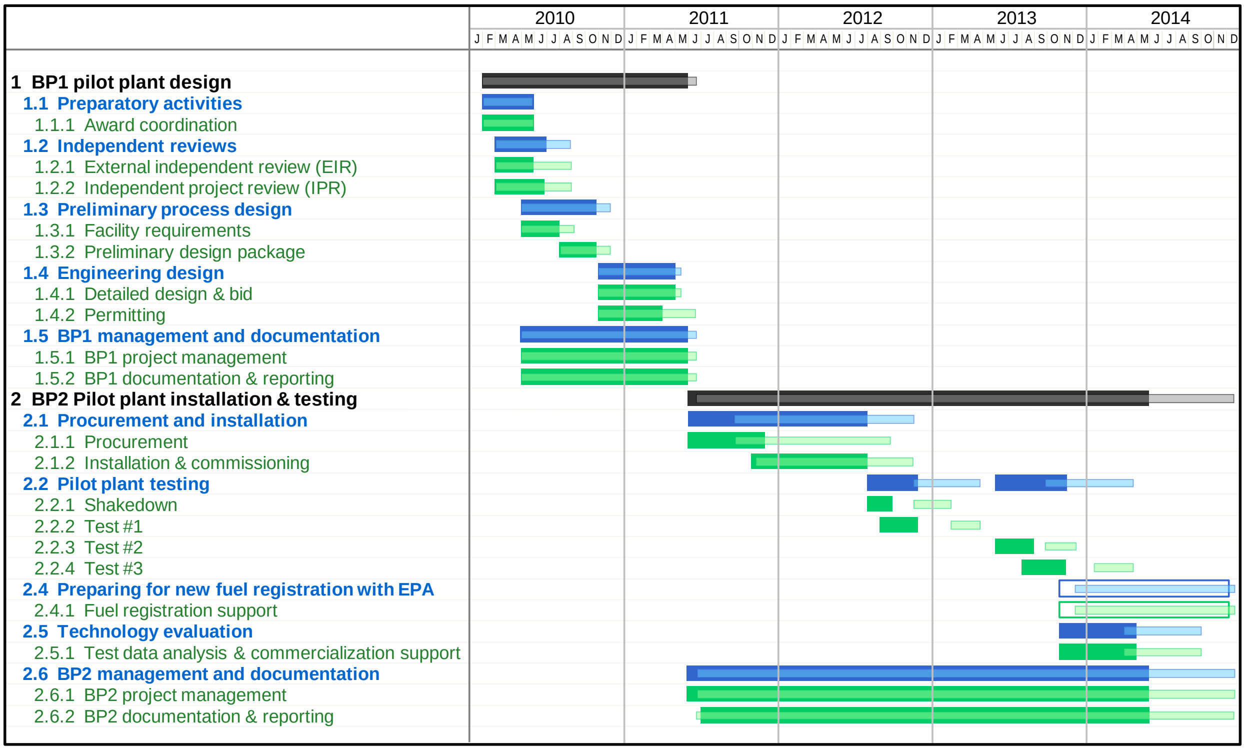

Figure 3-2 Final WBS Gantt Chart Showing Actual Work Periods

4. Pilot Plant Process Description and Block Diagrams

4.1 General Information

The Integrated Bio Refinery (IBR) demonstration unit has a design capacity of 22.5 bbl/d of liquid raw gasoline product from 21.6 bone dry short tons per day of wood feedstock.

The gasoline demonstration unit consists of the following sections:

- Wood Gasification (including tar reforming and gas conditioning)
- Acid gas removal and synthesis gas compression
- DME synthesis and gasoline synthesis

All continuous process streams flowing between the different process units are shown in a block flow diagram attached to this report. The feed storage/handling system, feeding system, gasifier, hot gas filter, tar reformer, direct cooler, and AGR system were pre-existing process components. Pre-existing utilities included cryogenic N_2 and O_2 gas storage, evaporation, and delivery; steam boiler; cooling tower; air cooler for the direct syngas cooler; high-pressure deionized water; and flare. New process units comprised the syngas compressor and TIGAS system. New utilities added for this project included a chiller, temporary storage tanks for product gasoline and process water, and a cryogenic CO_2 storage and delivery system.

4.2 Andritz-Carbona Gasification Process

The Andritz-Carbona gasifier is central to the IBR. The design capacity of the gasifier is 23 short tons/d of as-received feedstock at 6 weight percent moisture. The Gasification section also includes unreacted solids/ash removal, catalytic tar reforming, and syngas scrubbing to condense water and remove most contaminants such as tars, ammonia, and HCl.

4.3 Gasification Process

Biomass or other carbonaceous feedstock is processed in a fluidized bed of inert refractory material at elevated pressure and temperature. Partial oxidation of the feedstock provides heat for simultaneous pyrolysis and reforming reactions. This suite of reactions produces synthesis gas, also known as syngas. Syngas is a mixture of H_2, CO, CO_2, H_2O, CH_4, N_2, and higher hydrocarbons. Depending on the feedstock composition, syngas may also contain trace amounts of H_2S, COS, NH_3, HCN, polyaromatic hydrocarbons, and volatilized inorganics such as KCl. The syngas will also typically include entrained unreacted solids and ash (fly ash) that must be removed before further processing.

Gasification reactions are numerous, but consist chiefly of the following:

$$\text{CHNSO (biomass)} \rightarrow CO, CO_2, H_2, H_2O, CH_4, C_xH_y, NH_3, \text{etc.}$$

$$C + O_2 \rightarrow CO_2$$

$$C + H_2O \rightarrow CO + H_2$$

$$C + 2H_2 \rightarrow CH_4$$

$$C + CO_2 \rightarrow 2CO$$

$$CO + 1/2\ O_2 \rightarrow CO_2$$

$$H_2 + 1/2\ O_2 \rightarrow H_2O$$

$$CO + H_2O \rightarrow CO_2 + H_2$$
$$CH_4 + H_2O \rightarrow CO + 3H_2$$

A block diagram of the Gasification process is shown in Figure 4-1.

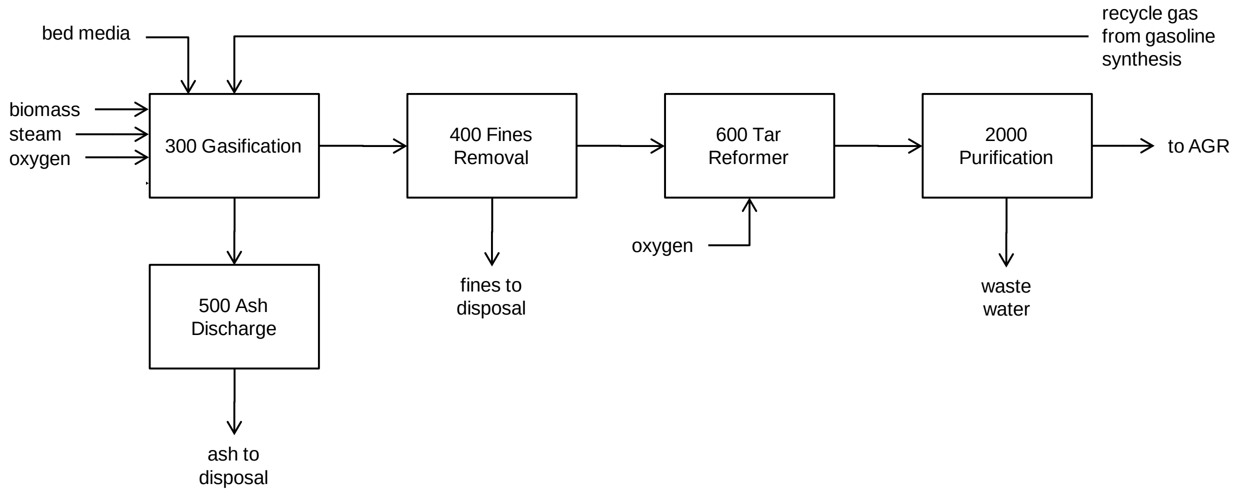

Figure 4-1 Gasification Section Block Flow Diagram

In the following, each of the process units is described sequentially.

4.3.1 Section 300 Gasifier

Solid fuel, fed through a lock hopper system comprising a weigh bin, lock hopper, and constant pressure feed hopper, reacts with steam and air (or oxygen) in the adiabatic refractory-lined pressurized fluidized-bed gasifier. The fluidized bed consists of inorganic granular media such as dolomite, which is metered into the gasifier separately from the biomass. Carbon dioxide can also be fed to the gasifier to assist with fluidization. Hydrogen, methane, carbon oxides and nitrogen constitute the major elements of the gaseous fuel product (syngas). For this project, the Andritz-Carbona gasifier operated at approximately 1560°F and 145-185 psia.

The gasifier reactor consists of three internal sections, namely the lower bed, upper bed and freeboard. The reactor bottom has a coned-shape grid and a central ash discharge. Steam, CO_2, and oxygen are fed proportionally through the grid and the center pipe to the gasifier to fluidize the solids in the bed.

The product gas exits the top of the gasifier. Large particulates are captured in a cyclone and recycled to the gasifier through a dip-leg. This recycling of particulates is important to maximize carbon conversion in the gasifier.

The larger particles of spent ash discharge from the bottom of the gasifier through the ash lock hopper system. Adjusting the gas velocity through the central ash discharge pipe controls the ash discharge rate.

A start-up heater is used to preheat the refractory in the gasifier during start-up. Once the gasifier is sufficiently heated, biomass feeding is started under excess air (combustion) conditions to complete the heat-up process. A second start-up heater provides heat to the downstream units in order to reduce start-up time.

4.3.2 Section 400 Fines Removal

The function of the Fines Removal System is to remove remaining particulates from the raw product gas prior to reforming.

The Flex Fuel Test Facility is equipped with two refractory-lined cyclones in series. The first cyclone separates the solids from the product gas exiting the gasifier freeboard. The separated fines in the cyclone dip-leg are re-injected into the gasifier. A second cyclone is available but was bypassed in this project to improve the performance of the hot gas filter.

The product gas exiting the cyclone is directed to the Hot Gas Filter (HGF), although there is a slipstream available for testing higher-temperature filters. The main line is cooled by direct water spray to a temperature suitable for the HGF. The unit employs multiple candle filter elements to remove the remaining fines. Depending on the filter element material, this temperature is in the range of 800° to 1100°F. The fines removed in the filter are discharged into a water-cooled surge hopper, lock hopper, and subsequently into a collection drum and weighed. The drums are then stored for later disposal. During a portion of the testing under this project, 10 to 15% of the product gas was directed uncooled to an Ultra-Hot Gas Filter (UHGF), followed by cooling for flow measurement and flaring. The UHGF was a refractory-lined vessel equipped with ceramic filter candles and CO_2 back-pulse system, similar to the HGF except for the higher process gas temperature.

4.3.3 Section 500 Gasifier Ash Discharge

The function of the Gasifier Ash Discharge System is to remove and cool the ash discharged from the gasifier in dry solid form.

The Gasifier Ash Discharge System consists of a water-jacketed ash surge hopper and ash lock hopper where the ash is depressurized. The ash is discharged into a drum. Each drum is weighed and collected for later disposal.

4.3.4 Section 600 Tar Reformer

The Tar Reformer is a catalytic reactor that accepts the solids-free syngas from Section 400 and reforms hydrocarbons via steam reforming reactions, using steam already in the syngas to drive the reforming reaction.

$$C_xH_y + xH_2O \rightarrow xCO + (x+y/2)\ H_2$$

The Tar Reformer contains Topsoe Ni-based reforming catalyst. To maintain a temperature sufficient to reduce hydrocarbons to an acceptable level, it is necessary to inject a small amount of oxygen in the tar reformer. The acceptable level of tars in the syngas depends on the downstream processes. In these tests the limit was set by the detection limit for naphthalene – i.e. the process was operated to achieve less than 10 ppm of naphthalene in the syngas. The oxygen increases the syngas temperature to approximately 1635°F through oxidation (combustion) of some syngas components.

Downstream of the tar reformer, the syngas is cooled by deionized water in a second direct spray cooler to a temperature of approximately 700°F.

4.3.5 Section 2000 Purification

The syngas is cooled to condense the in-situ water and also to scrub out impurities remaining in the syngas. The syngas Direct Cooler consists of a spray tower with multiple spray nozzles in countercurrent flow. The scrubber water is recycled from the scrubber sump to an air cooler and two water-cooled heat exchanger to maintain exiting syngas temperature of 104°F. A level controller in the scrubber regulates the amount of excess water that is removed for disposal as waste water. The scrubber recycle circuit also includes pH control by caustic (NaOH) and formic acid (HCOOH) injection systems in order to maintain pH in the range that is permissible for disposal (5 to 10).

The syngas exits the scrubber through a final de-mister at a temperature of 104°F. During start-up and when downstream process steps are not in operation, the syngas is then diverted to another cyclone and then to the flare, where it is combusted with natural gas. When the acid gas removal (AGR) system is in operation, the syngas is directed to the syngas compressor in the AGR section.

4.4 Acid Gas Removal

The purpose of acid gas removal is to reduce the CO_2 and H_2S content in syngas from the gasification section. This section also includes compression of syngas to the pressure required for gasoline synthesis. Waste gases from the AGR section are directed to the flare.

4.4.1 Section 800 Acid Gas Removal Process

After removal of fine solids, tar reforming, and condensation of most of the water vapor in the gasification section, the syngas is composed primarily of CO, H_2, CO_2, and residual water vapor, with lesser amounts of H_2S, CH_4, N_2, and other minor components. The AGR section follows the gasification section. The primary objective of AGR is to selectively remove a major portion of the carbon dioxide (CO_2) from the "raw" syngas. Other acidic components of the raw syngas are also removed in the AGR section, especially hydrogen sulfide (H_2S). The syngas with acid gases removed is termed "sweet" or "sweetened", while the acidic waste gas stream is termed "sour". For this process, the target CO_2 concentration in the sweet syngas (which is routed to the TIGAS section) is 2 mole percent. Because of the low sulfur content of the woody biomass, the concentration of H_2S in the sweetened gas was only a few parts per million even without optimization for H_2S removal.

The AGR section processing is based on absorption of the acidic components from the raw syngas into a physical solvent: a blend of n-formyl morpholine (NFM) and n-acetyl morpholine (NAM) termed Morphysorb®. The Morphysorb® technology was jointly developed by GTI and Uhde GmbH (now Thyssen Krupp) and has many advantages over competing physical solvents.

The absorption of the acidic components into a physical solvent is favored by high pressure and low temperature. Therefore, the initial step in the AGR section is compression of the raw syngas, followed by contacting lean solvent with the syngas in a counter-current flow scheme. After separation from the sweet syngas, the "rich" solvent is regenerated at reduced pressure and moderately-increased temperature, for recycle back to the syngas contactor. With a highly-selective solvent such as Morphysorb®, very little of the desirable components (CO, H_2) of the syngas are lost in the AGR system. Furthermore, solvent losses (by vaporization of the solvent into the syngas) are also minimized by solvent and process-condition optimization.

Figure 4-2 shows the main steps of the AGR section. The syngas (cool, dry, but "sour") is compressed from the gasifier operating pressure (120 to 150 psig) to approximately 1015 psig.

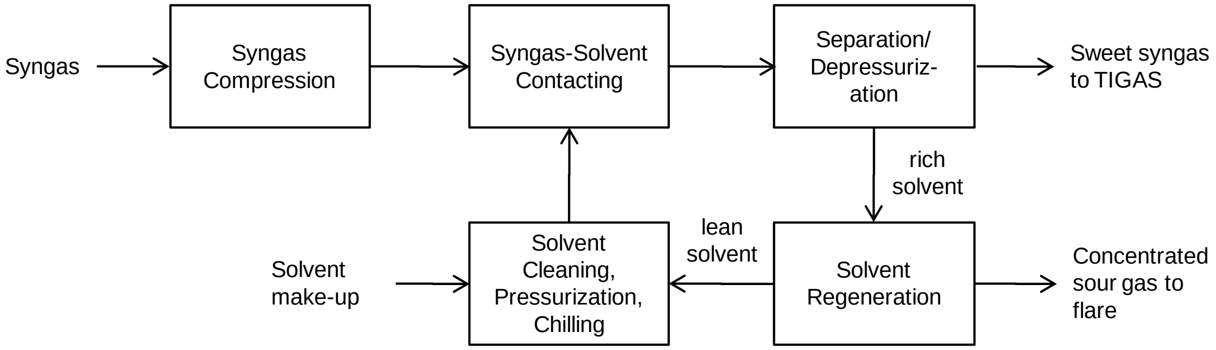

Figure 4-2 Block Flow Diagram of Acid-Gas Removal Section

Next, counter-current contacting of chilled lean solvent with the sour syngas causes the acidic components of the syngas (CO_2 and H_2S) to transfer from the gas phase to the liquid phase.

In the next AGR operation, sweet gas is separated from the rich solvent and, furthermore, a partial de-pressurization (flashing) of the rich solvent produces one stream of concentrated acid gas (CO_2 and H_2S). The sweet syngas is routed to the TIGAS section.

Next, the de-pressurized solvent is heated and stripped/distilled to regenerate “lean” solvent. The second stream of concentrated acid gas from the regenerator is combined with the flashed gas, and this material is routed to the flare.

The lean solvent is then filtered to remove a very small concentration of solid impurities, pressurized, cooled, and chilled for recycle to the syngas/solvent contactor.

4.4.2 Syngas Compressor

Cooled, low-moisture syngas from the Direct Cooler flows to the Syngas Compressor KO Drum, to remove any traces of liquids. The overheads from the KO Drum are routed to the first stage of the Main Syngas Compressor. The partially-compressed syngas is cooled in an interstage cooler, and de-misted in the Inter-Stage KO Drum. The liquid-free vapor is then routed to the second stage of the Main Syngas Compressor.

The fully-pressurized Syngas is cooled in the Syngas Compressor Final Cooler. Since the Main Syngas Compressor is a fixed-volume machine, a portion of the fully-pressurized syngas can be spilled back to the inlet through the Syngas Compressor KO Drum.

4.4.3 Syngas/Solvent Contactor

Pressurized lean solvent is fed onto a distributor in the top section of the High-Pressure Absorber and flows downward through multiple sections of structured metal packing. The bottom section of the Absorber has no internals, and provides a hold-up (or “surge”) volume of liquid, which is maintained by level control on the outlet rich-solvent stream.

The sour syngas is fed into a lower section of the High-Pressure Absorber, above the liquid level but below the structured packing. As the syngas moves up through the dispersed solvent in the structured packing, CO_2 and H_2S are transferred from the syngas to the solvent. The sweet syngas, with small amounts of entrained solvent, exits from the top of the Absorber.

The extent of removal of the acid gases from the syngas is maximized by (1) higher solvent flow rates (up to the point of “flooding” in the packing), (2) lower solvent temperatures (up to the point of excessive solvent viscosity or solvent solidification), and (3) cleaner lean solvent (lower concentrations of CO_2 and H_2S from the regeneration section).

4.4.4 Separation/Solvent Depressurization

The raw sweet syngas from the High-Pressure Absorber is directed to a High-Pressure Separator vessel. Any entrained solvent mist is allowed to settle out in this vessel. The sweet syngas travels upward through a de-mister pad, and exits through a pressure-control valve to the TIGAS section. The small amount of solvent recovered from the sweet syngas is fed through a level-control valve to a Mild-Pressure Separator.

The rich solvent from the bottom of the High-Pressure Absorber is routed through the double-pipe heat exchanger, where the temperature is increased somewhat by exchange with a slip-stream of the lean solvent from the Reboiler/Surge Tank Vessel. The heated rich solvent then flows through the level-control valve into the Mild-Pressure Separator.

A major fraction the absorbed CO_2 and H_2S in the rich solvent is flashed/released with the decrease in pressure in the Mild-Pressure Separator vessel. This sour gas travels upward through a de-mister pad, through a pressure-control valve, to a Reflux Accumulator, and finally to the Flare. The flashed liquid is routed from the bottom of the Mild Pressure Separator, through a level-control valve and heat exchanger, to a Regeneration Stripper.

4.4.5 Solvent Regeneration

The flashed rich solvent out of the Mild Pressure Separator enters a Lean/Rich Solvent Exchanger, a plate-and-frame heat exchanger. By cross-flow through that heat exchanger, the flashed rich solvent is heated from the primary flow of lean solvent from the Reboiler/Surge Tank. As mentioned previously, a slip-stream of the lean solvent from the Reboiler/Surge Tank is routed (by manual control) to pre-heat the rich solvent in the double-pipe heat exchanger before flashing in the Mild-Pressure Separator.

The heated, flashed rich solvent is then fed onto a distributor in the upper section of the Regeneration Stripper, which contains structured metal packing. The dispersed solvent flows downward in the Regenerator Stripper column, in contact with upward-flowing vapor. The vapor in this column consists of the acid gases which out-gas from the solvent assisted by low-pressure nitrogen, which is added at the bottom of the Regeneration Stripper column.

The Regeneration Stripper is mounted directly on the top of the Reboiler/Surge Tank, and the stripped solvent from collects in an internal pocket of the Reboiler/Surge Tank vessel. The stripped liquid flows downward and then through a coiled pipe in the Regeneration Heater, where the heat is supplied by direct firing of natural gas through a burner. The stripped, re-boiled solvent travels back into the primary section of the Reboiler/Surge Tank. The lean (regenerated) solvent exits at the bottom nozzle of the vessel, and is routed to both the heat exchangers, as described above.

4.4.6 Lean Solvent Processing

The cooled lean solvent from the Lean/Rich Solvent Exchanger is routed to the Lean Solvent Circulation Pump, which supplies a slight pressure boost for the solvent cleaning operations. The lean solvent is normally routed through the Lean Solvent Sock Filter, to remove fine particles of debris from the AGR system, but the filter can be bypassed if pressure differential is too high from dirty filter cartridges. A slip-stream of the filtered lean solvent is sent to the Lean Solvent Charcoal Filter to prevent build-up of soluble or extremely fine degradation products in the circulating solvent.

The filtered lean solvent is pressurized through the Lean Solvent Pressurization Pump to about 1000 psig. Since this pump is a constant-volume type, flow rate of the lean solvent is controlled by spilling back some lean solvent through a flow-control valve. The lean solvent is cooled in two stages – through the Lean Solvent Air Cooler and the Lean Solvent Final Cooler. The latter exchanger is supplied with chilled glycol coolant to reduce the lean solvent temperature to 40°F.

4.5 Topsoe Improved Gasoline Synthesis (TIGAS)

The TIGAS process offers an efficient route for the conversion of gasification based synthesis gas into high octane gasoline through methanol (MeOH) and dimethyl ether (DME) as the primary key intermediates. The process that was installed in the IBR is based on a layout where synthesis gas produced from gasification of wood and tar reforming, followed by purification and AGR, is converted first to DME and then to gasoline. A block diagram of the TIGAS unit is shown in Figure 4-3.

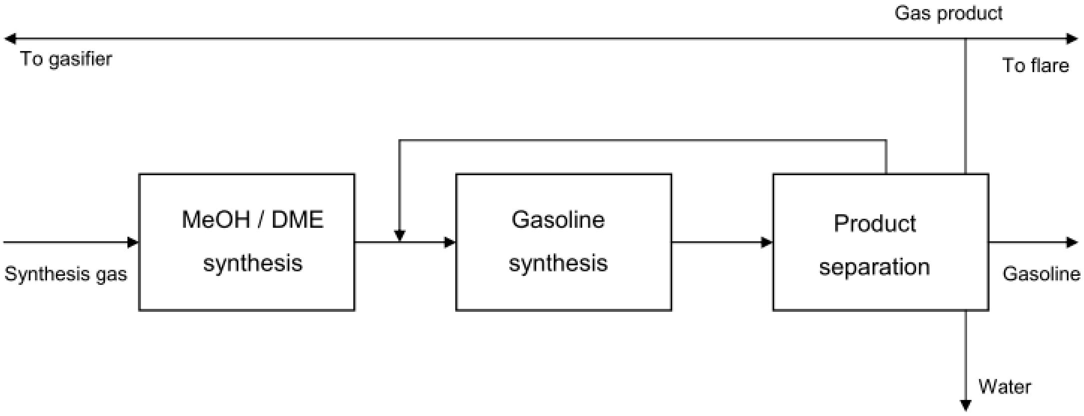

Figure 4-3 TIGAS Block Diagram

The syngas leaving the synthesis gas preparation will normally have a ratio between hydrogen and carbon monoxide of approximately 1.1, which is appropriate when applying a combined MeOH/DME synthesis catalyst. As shown Figure 4-4 below, the theoretically optimum synthesis gas H_2/CO ratio is slightly above 1. When having either a lower or higher ratio, the production of oxygenates will decrease and ultimately so will the gasoline production. The effect of the synthesis gas H_2/CO ratio on the product distribution for the combined MeOH/DME synthesis is shown in Figure 4-4 (equilibrium data). A temperature of 482°F and a pressure of 856 psig were used for the equilibrium calculations used to generate Figure 4-4.

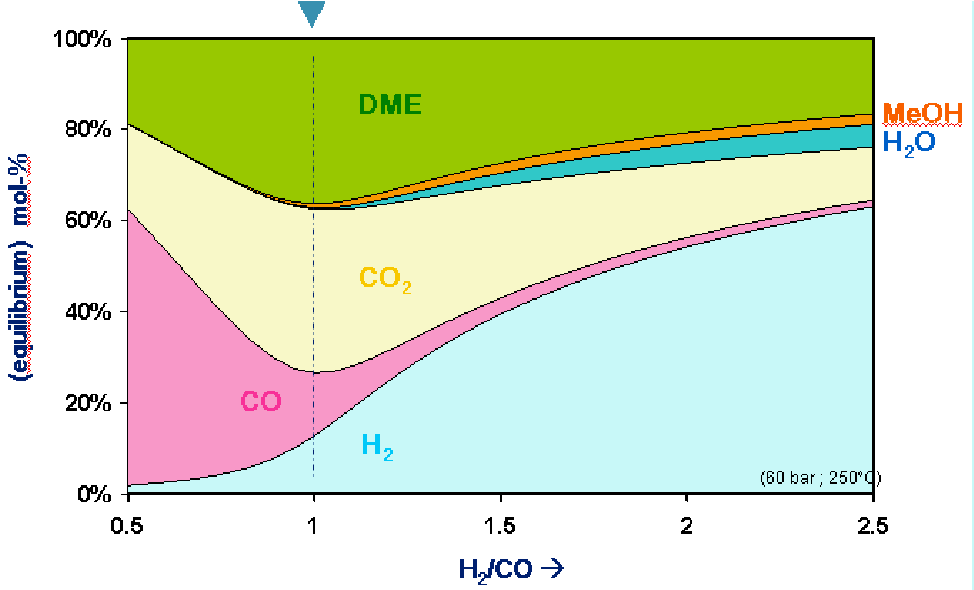

Figure 4-4: Effect of the Synthesis Gas H_2/CO Ratio on the Product Distribution

The Main Synthesis gas compressor is located up-stream the AGR section. The TIGAS section requires high synthesis gas pressure in order to provide a high conversion of synthesis gas to oxygenates (MeOH and DME), as shown in Figure 4-5. The figure shows the equilibrium conversion versus pressure (bar g) of synthesis gas to methanol and dimethyl ether respectively (blue curve). Reference (red curve) indicates conversion to MeOH only. The syngas mole ratio for the calculations used to produce Figure 4-5 is $H_2/CO/CO_2$ = 66/33/1 and the temperature is 465°F.

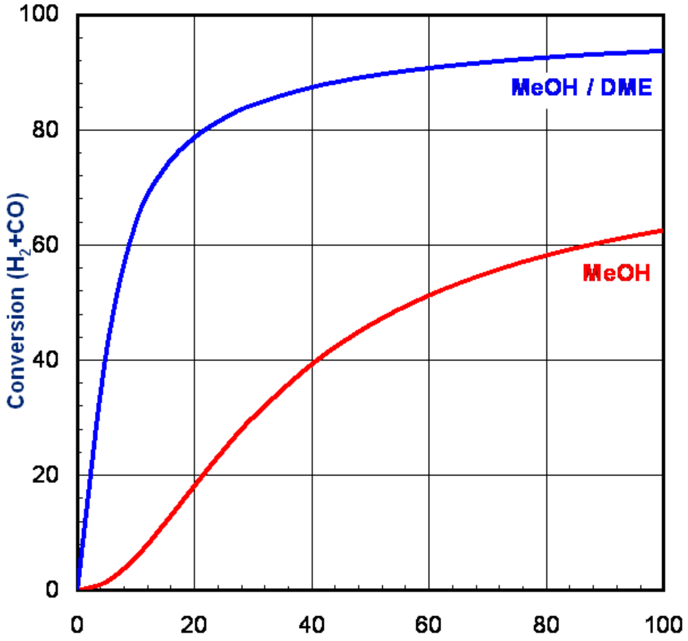

Figure 4-5: Equilibrium Conversion of Synthesis Gas in Percent (y-axis) as Function of Pressure in bar g (x-axis)

The TIGAS synthesis loop feed gas is received at a temperature of about 40°F and a pressure of 1000 psig from the up-stream AGR section. The synthesis gas contains approximately 52.5 mole % H_2, 41.4 mole % CO, 2.3 mole % CO_2 and 3.7 mole % CH_4 + N_2. The CH_4 and N_2 both function as inerts in the synthesis loop.

4.5.1 Oxygenate synthesis

Before the synthesis gas enters the oxygenate synthesis section, it is preheated to 430°F in first the Hot Guard Feed/effluent Exchanger and secondly the Hot Guard Trim Heater before it is passed through the first reactor, the Hot Guard Reactor. The Hot Guard Trim Heater outlet temperature is controlled by adjusting the amount of condensing steam and a process gas by-pass.

The primary function of the Hot Guard Reactor is removal of potential trace amounts of contaminants that would cause catalyst deactivation in the downstream oxygenate synthesis. The reactor contains several layers of catalysts/absorbents that remove a variety of components e.g., COS, NH_3, As, HCN, H_2S and carbonyl compounds. The reactor layer will contain the following catalysts/absorbents:

- Topsoe HTZ: Sulfur absorbent for bulk removal and hydrolysis of COS / HCN.
- Topsoe NT-101: NH_3 absorbent.
- Topsoe CT-101: Carbonyl guard absorbent.
- Topsoe ST-101: Sulfur polisher and As guard.

Besides functioning as a guard reactor for the downstream Oxygenate Reactor hydrogen, carbon monoxide and carbon dioxide are partly converted into methanol in the Hot Guard Reactor, as the ST-101 catalyst is also active for the water gas shift reaction and methanol formation. The following reactions will take place in this reactor:

$$\mathrm{CO + 2H_2 \leftrightarrow CH_3OH} \tag{4.1}$$

$$\mathrm{CO + H_2O \leftrightarrow CO_2 + H_2} \tag{4.2}$$

The partially converted gas from the Hot Guard Reactor contains approximately 3.4 mole % of methanol. Some by-products, primarily ethanol and dimethyl ether, and traces of higher alcohols, ketones and aldehydes, are produced as well, but these are harmless to the downstream synthesis steps. In fact, production of higher oxygenates causes a diversion of reaction heat from the adiabatic gasoline reactor. The diversion has the benefit of reducing the amount of cooling recycle to the gasoline reactor and hereby results in a more energy efficient system.

The partial conversion of synthesis gas through the exothermic methanol reaction causes a temperature increase to approximately 605°F over the catalyst bed in the adiabatic Hot Guard Reactor. The gas must be cooled before it is introduced to the combined MeOH/DME reactor, Oxygenate Reactor, to a temperature of 430°F. The cooling of the gas takes place in Hot Guard Feed/Effluent Exchanger.

In the Oxygenate Reactor the conversion of synthesis gas to both methanol and dimethyl ether takes place according to the following reaction schemes:

(4.1) $CO + 2H_2 \leftrightarrow CH_3OH$

(4.2) $CO + H_2O \leftrightarrow CO_2 + H_2$

(4.3) $2CH_3OH \leftrightarrow CH_3OCH_3 + H_2O$

The combination of the methanol synthesis and the dehydration to dimethyl ether in the same reactor results in high conversion as the methanol product formed accordingly to reaction 4.1 above is continuously being consumed by the dehydration process (reaction 4.3). Especially for carbon monoxide rich conditions, the water gas shift reaction (reaction 4.2) induces a strong enhancement of the conversion because water, formed in the MeOH-to-DME step, is shifted through the reaction with carbon monoxide to form hydrogen and carbon dioxide. The net reaction is then essentially that of hydrogen and carbon monoxide forming DME and carbon dioxide (reaction 4.4) as is seen in Figure 4-4 above.

(4.4) $3H_2 + 3CO \leftrightarrow CH_3OCH_3 + CO_2 \qquad -\Delta H = 246 \text{ kJ/mol}$

For a carbon monoxide rich synthesis gas with a hydrogen to carbon monoxide ratio close to one the equilibrium conversions exceed 90% under typical reaction conditions.

A boiling water reactor design which is used for Oxygenate Reactor ensures efficient heat removal of the reaction heat. The boiling water reactor is constructed as a shell and tube heat exchanger, where the combined methanol and DME synthesis catalyst is loaded in the reactor tubes. The shell side contains water at the boiling point which efficiently removes the reaction heat by evaporation. The shell side is connected to the Steam Drum. The boiling water/steam circulates through the reactor shell driven by the difference in density between the boiling water and boiling water/steam streams entering and exiting the reactor. The boiling temperature in the shell can be controlled by adjusting the pressure of the steam in the steam drum.

The efficient cooling by the design provides for a high conversion and a modest overall temperature increase to approximately 500°F exiting the reactor and with a concentration of methanol and dimethyl ether of 1.7 mole-% and 18.6 mole-%, respectively.

The boiling water reactor is loaded with Topsoe DCK-10 catalyst which is a copper based combined MeOH/DME synthesis catalyst.

4.5.2 Steam System

The steam produced in the boiling water reactor is separated from the boiler water in the steam drum. The produced steam is used for preheating the synthesis gas in Hot Guard Trim Heater. Steam is introduced to the shell side of the exchanger and the condensation of steam provides an efficient heat transfer to the process gas.

All remaining steam produced in the boiling water reactor is condensed in the Air Cooled Steam Condenser and the condensed steam is returned to the Steam Drum which is placed below the two heat exchangers. The steam system is a closed circuit which does not need continuous addition of make-up water. However, it is possible to add make-up boiler feed water to the Steam Drum during operation.

4.5.3 Gasoline Synthesis

In the Gasoline Reactor the oxygenates are converted to water and hydrocarbons by the Topsoe GSK-10 catalyst according to the following reactions:

(4.5) $$n\ CH_3OH \leftrightarrow n\ (CH_2) + n\ H_2O$$

(4.6) $$n\ CH_3OCH_3 \leftrightarrow n\ 2(CH_2) + n\ H_2O$$

where (CH_2) denotes the average composition of the gasoline product. The net reaction may be formulated as:

(4.7) $$\text{Oxygenates (MeOH/DME)} \rightarrow C_{1\text{-}2} + C_{3\text{-}4}\text{(LPG)}, C_{5+}\text{(Gasoline)} + H_2O + \text{Heat}$$

Prior to entering the Gasoline Reactor, the effluent from the Oxygenate Reactor is mixed with recycle gas from the downstream product separation. The mixture is preheated in the Gasoline Feed/Effluent Exchanger and introduced to the Gasoline Reactor. As the inlet temperature of the Gasoline Reactor is essential for the performance, both an electric trim heater and a recycle gas by-pass around Recycle Gas Preheater (which heats up most of the recycle gas) are included in order to maintain efficient control of inlet temperature. The recycle gas by-pass is added up-stream the Electric Trim Heater.

The addition of the recycle stream serves as a heat sink for the exothermic dehydration reaction and reduces the temperature rise in the Gasoline Reactor. The amount of gas which is recycled is determined by the reactor outlet temperature. The reactor outlet temperature is maintained fixed at approximately 745°F; operating temperature shall not exceed 790°F. The reactor recycle rate shall be adjusted to control the temperature increase in a way where the temperature difference is approximately 80-130°F in order to reduce coking rates and irreversible zeolite deactivation caused by the presence of steam at elevated temperatures. The control of inlet temperature and temperature rise reduces the formation of undesired by-products formed by cracking reactions which ultimately would reduce the gasoline product yield.

During the reaction, the zeolite catalyst becomes gradually deactivated by coke lay-down and must be regenerated by controlled coke burn-off. Regeneration will typically be required after 800 hours. The TIGAS demonstration unit provides a single adiabatic reactor and is not equipped with regeneration facilities in consideration of the shorter duration of the test campaigns. As a result it was not necessary to replace the catalyst as the life-time of the catalyst exceeded the run-hours. For an industrial scale plant, the layout will consist of parallel reactors providing intermittent regeneration of individual reactors without affecting the plant productivity.

The effluent from the Gasoline Reactor is further cooled after the Gasoline Feed/Effluent Exchanger in the Recycle Gas Preheater, the Gasoline Cooler and the Gasoline Chiller before separation takes place.

4.5.4 Product Separation

The cooled process stream consists of three phases which are initially separated in the High Pressure Separator. The inlet temperature of the product stream to the separator from the Gasoline Chiller is 41°F. A low temperature decreases the amount of higher hydrocarbons in the gas phase and hence improves the process efficiency. The waste water stream is taken out in the lower part of the separator and depressurized, and dissolved gases are partly removed in the Waste Water Flash Drum. The waste water is sent to the waste water storage tank.

The primary product, the hydrocarbon phase, is taken out in the middle of the High Pressure separation vessel. The product stream is depressurized and the LPG is partially vaporized which causes a significant decrease in temperature. The stream is heated in the Trim Heater to approximately 50°F before it is introduced to the Low Pressure Separator.

In the Low Pressure separator additional gas is removed, and the liquid product is delivered to the storage tank. The product is mainly gasoline but also some LPG and remaining dissolved gases. The gas separated in the Low Pressure separator is sent to flare. The gas phase from the High Pressure separator consists primarily of unconverted synthesis gas, inerts, and light hydrocarbons from the gasoline synthesis.

As previously mentioned, a fraction of the separated gas phase is recycled to the Gasoline Reactor for temperature control. The recycle gas is compressed in the Recycle Gas Compressor before it is preheated in the Recycle Gas Preheater. The remaining fraction is partly recycled to the gasifier section and the remaining fraction of gas is sent to the flare.

5. Project Construction and Commissioning

This chapter will describe activities that were carried out in both Budget Periods, including the following:

- Permitting
- Procurement
- Construction
- Installation
- Commissioning (including Shakedown)

5.1 Permitting

Permitting was first investigated during Budget Period 1 in order to identify required permits and approvals. These are listed and described in Table 5-1 . Permit applications were submitted as the necessary design and operation information became available.

In one case, the final approval to proceed with operations from the City of Des Plaines Fire Department was withheld in order to make changes to a containment wall around the temporary gasoline storage tank.

In another case, the local water quality authority, Municipal Water Reclamation District of Greater Chicago (MWRDGC,) required a sample of wastewater to be analyzed at the start of operations before granting final approval to proceed.

Copies of permits are provided in Appendix A.

Table 5-1 Required Permits

Permit or Approval	**Responsible Agency**	**Description**	**Submittal**	**Approval**
Flammable Material Storage Permit	Illinois State Fire Marshall (ISFM)	Required to install 8,000-gallon aboveground gasoline storage tank. Required to begin CONSTRUCTION.	06/15/2011	06/17/2011
Air Construction Permit	Illinois Environmental Protection Agency (IEPA)	Required for vapor losses from the 8,000-gallon gasoline tank and an 8,000-gallon wastewater tank. Other minor air emission sources may be identified during IEPA permitting process. Required to begin CONSTRUCTION.	08/01/2011	10/28/2011
Federally Enforceable State Operating Permit	Illinois Environmental Protection Agency (IEPA)	Required to expand range of acceptable feedstock types processed in the GTI gasifier. Required to begin OPERATION.	08/01/2011	10/28/2011
Air Pollution Source Permit	Cook County	Installation of air pollution sources require permit. Cook County accepts Air Construction Permit applications that have been submitted to IEPA. Forwarded to County by IEPA. Required to begin OPERATION.	08/15/2011	12/12/2011

Permit or Approval	Responsible Agency	Description	Submittal	Approval
Building Permit (Fire Dept.)	City of Des Plaines	For tank farm and containment. Required to begin CONSTRUCTION, inspection required to begin OPERATION. Note that changes were made to containment wall in Feb 2012.	11/15/2011	01/15/2012 (approved) 06/25/2012 (inspected)
Building Permit (Building Dept.)	City of Des Plaines	Required to begin CONSTRUCTION.	11/15/2011	12/02/2011
Determination of No Hazard to Aviation	Federal Aviation Administration (FAA)	Proximity of site to O'Hare International Airport requires FAA approval and notification of airport authorities when construction crane activity at elevations 150 ft above airport elevation begins and ends. Required to complete CONSTRUCTION.	04/18/2012	07/06/2012
Fire Protection Permit	City of Des Plaines	For sprinkler and fire protection modifications inside process building. Submitted by contractor. Required to begin OPERATION.	08/13/2012	09/21/2012
Wastewater Discharge Permit	Metropolitan Water Reclamation District of Greater Chicago (MWRDGC)	Permit required prior to MWRDGC receiving wastewater (containing traces of organics) from Facility operations. Wastewater to be stored in 8,000-gallon aboveground tank prior to disposal. Required to continue OPERATION subject to sample analysis during first full week of operation.	12/02/2011	03/14/2014

5.2 HAZOP Reviews

The project team conducted Hazard & Operability (HAZOP) reviews at selected times during all phases of the project; whenever major additions, upgrades, or changes were made that could potentially affect safety and/or operability:

- October 2010 (preliminary design)
- March 2011 (AGR modifications)
- April 2011 (TIGAS, syngas compressor, tank farm)
- May 2012 (TIGAS recycle compressor)
- October 2012 (gasifier modifications)
- September 2013 (hot gas filter, gasoline batch tank)

HAZOP reviews were multi-day meetings at GTI facilities. Recommendations for changes were implemented and incorporated into bid packages when needed.

5.3 Construction

All construction-related activities were carried out under Task 2.1 Procurement and Installation.

From the Design Basis developed in BP1, the project team identified prospective bidders, issued Requests for Bid, and held bid meetings at GTI for those who chose to bid. The bid request packages included detailed design information. Bid meetings included all applicants so that everyone would hear the same questions and receive the same answers. This rigorous process ensured that the selection process was fair and efficient.

The major construction and installation packages, number of bidders, and the selected contractors, are given in Table 5-2 .

Table 5-2 Bid Packages

Item/package	Description	# of bidders	Selected vendor(s)
Tank Farm	For gasoline and process water holding tanks; package includes concrete, equipment placement, and piping based on Ridge & Associates design	3	Morrison Construction (incl. subcontract to Hasse Construction for concrete)
Syngas Compressor	Compresses syngas prior to Acid Gas Removal (AGR) and TIGAS	3	AG Equipment
TIGAS Modules	Converts decarbonized (sweet) syngas to gasoline; modular TIGAS pilot plant to be installed in GTI Bay	5	Zeton
Chiller	Supplies chilled fluid for cooling AGR solvent and TIGAS streams	3	AEC Inc.
Concrete	Pads for compressor, AGR access, CO2 storage tank, and TIGAS Bay floor (excludes Tank Farm)	5	Natola Concrete
Building Steel & Equipment Erection	To modify TIGAS Bay steel to accommodate TIGAS modules, lift and place TIGAS modules, lift and place syngas compressor (excludes Tank Farm)	3	MTH Industries
Electrical Installation	All electrical work for new construction including Tank Farm	1	Emcor Hyre Electric
Mechanical Installation & Piping	All new and upgraded interconnecting piping, instrumentation, equipment connections.	5	Amex Nooter Meccon Industries
Insulation	New and upgraded insulation for vessels and piping	2	Luse Thermal Technologies
Fire Protection	Upgrading sprinklers and fire notification in TIGAS Bay	2	F.E. Moran
CO_2 Cryogenic Supply	New storage tank and delivery hardware for cryogenic CO_2	3	Air Liquide
Data & Controls	Process control, data collection, and special instrumentation	Sole source	Novaspect

The Tank Farm was bid as a separate package because of the procurement of a complete design package from Ridge & Associates during the design phase in BP1.

Novaspect was granted a sole source contract for the Distributed Control System (DCS) because they are the official regional distributor for the Emerson DeltaV system already in place and functional at GTI.

Construction began in August 2011 by conducting kick-off meetings with selected major suppliers and contractors (syngas compressor and TIGAS). Work plans were developed while bidding and selection continued for the remaining equipment and services through December 2011.

Excavation and foundations for the Tank Farm began in November 2011, but were delayed in December due to a modification to the containment wall requested by the Des Plaines Fire Department. That was completed in March 2012.

Orders for TIGAS skids and the syngas compressor were placed in February 2012, and detail engineering for structural building modifications to accommodate the TIGAS skids were then carried out to support the final selection of equipment erection and steelwork contractor. Those selections were made in May-June 2012.

AGR Section upgrades included a coolant/solvent heat exchanger, new redundant solvent circulation and pressurization pumps, electric solvent storage tank heater, replacement of pipe unions with flanges, additional heat tracing and insulation, modified solvent/gas mixing head on absorber column, upgraded instruments, anti-foam system, upgraded syngas inlet filter, and safety upgrades. These were started in early 2012 and completed in time for the second shakedown test in December 2012.

Installation jobs completed in May 2012 included a new heat exchanger to increase heat rejection from the syngas cooler water circuit, the AGR pressurization pump described above, a concrete pad for the syngas compressor, and relocation of existing TIGAS Bay steelwork to make room for the TIGAS modules. In June 2012, we completed placement of the syngas compressor and new TIGAS Bay steelwork. Delivery and erection of the TIGAS modules were completed in July 2012. Remaining steelwork and platforms, piping, DCS additions, electric power to major equipment and heat tracing, and instrumentation comprised the construction work in August through October 2012. The final installation tasks in the TIGAS Bay were piping insulation and a recycle compressor, part of the Zeton package.

In the Gasification Section, installation activities included improvements to the bottom ash withdrawal hardware, a new heat exchanger to increase heat removal from syngas, reactant gas preheating, and boiler water preheater to increase steam rate.

Photographs of key installation work are shown in Figure 5-1 through Figure 5-11.

Figure 5-1 Tank Farm Construction and Installation, November 2011-February 2012

Figure 5-2 Syngas Compressor Foundation and Installation, May-June 2012

Figure 5-3 Carbon Dioxide Storage Tank Foundation and Installation, July 2012


2012/05/31 10:28
2012/06/07 09:07
2012/06/08 13:56

Figure 5-4 TIGAS Bay Floor Upgrading, May-June 2012

Figure 5-5 TIGAS Module Fabrication at Zeton, October 2011-June 2012

Figure 5-6 Roof Removal to Accept TIGAS Modules, July 5, 2012

Figure 5-7 Delivery of TIGAS Modules (DME Reactor, Bottom Skid, Top Skid), July 6, 2012

Figure 5-8 Lifting and Installing Bottom TIGAS Module, July 7, 2012

Figure 5-9 Lifting and Installing Top TIGAS Module, July 7, 2012

Figure 5-10 Lifting and Installing DME Reactor, July 7, 2012

Figure 5-11 TIGAS Recycle Compressor & AGR with Upgrades, October 2012

5.4 Commissioning

The project was originally structured so that Commissioning and Shakedown were separate activities, with Shakedown considered part of the Pilot Testing task. However, with the knowledge that Milestone CD-4 (Completion of Commissioning) and issuance of a Commissioning Report were required to proceed with testing, we are including Shakedown for purposes of this report as a part of Commissioning.

Commissioning activities began in September 2011, focusing on existing equipment and instrumentation while bidding and selection for new equipment continued. These activities included checkout of existing instrumentation, stroking valves, loop-testing transmitters, preparation for system cleaning, replacement of existing gas-fired solvent tank heater with electric heater, and updating of interlocks and regulatory logic. AGR commissioning was also supported by a concurrent Illinois State-sponsored program.

A Commissioning Report was submitted in April 2013 to satisfy the CD-4 Milestone, and the text of that report is duplicated here.

> Commissioning teams were set up for the different plant areas and major equipment items in preparation of starting the commissioning work on schedule. Commissioning work including (a) recycle compressor CM-901 commissioning; (b) tank farm instrumentation; (c) AGR commissioning to confirm solvent circulation, gas flows, and instrumentation as well as test with solvent circulation and N_2/CO_2 mixture; and (d) tuning of the chiller controls was completed in November 2013 to allow shakedown to begin on November 26, 2012.

Shakedown testing began with **Test S-1**, which was designed to test and evaluate the gasification section (feed system, gasifier, hot filter, syngas reformer, and scrubber) and begin operation of the new syngas compressor. Three objectives were planned and successfully achieved: (1) stable operation of the gasification section at 1200 lb/h feed rate; (2) stable operation of the gasification section at 1450 lb/h; (3) startup and stable operation at 1450 lb/h with the syngas compressor in operation at full 1000 psig discharge pressure. Pressurized syngas was discharged through the AGR absorber, but without the AGR system in operation except for the use of its exit pressure control valve.
Test S-1 lasted four days. New modifications to the gasifier and reformer performed well, including the new feed injection screw, feed gas preheaters, redesigned oxygen distributor in the reformer, and increased cooling of scrubber recirculation water. Gasifier bed conditions were nominally 1560°F throughout, with gasifier pressure at 95 psig at feed rate #1 (nominally 1200 lb/h) and 120 psig at feed rate #2 (nominally 1450 lb/h). Nominal feed rates based on screw calibration were later checked versus weigh bin averaging and found to be higher than the nominal rates: 1340 lb/h at the lower speed and 1575 lb/h at the higher speed. Duration of set point periods were 15 hr at 1340 lb/h, 16 hr at 1575 lb/h, and 6.5 hr with the syngas compressor in operation.

The estimated carbon conversion was well over 90%. Reforming of hydrocarbons was 99+% for C_6H_6 and higher, and about 94% for CH_4. Scrubber outlet temperature was held within a few degrees of the 104°F design target. The transition from 1340 to 1575 lb/h was made without incident. Bringing the syngas compressor on line required several restarts to fine-tune the procedures without disrupting gasifier stability, and was ultimately successful. The compressor was then operated at 1000 psig discharge pressure with the internal recycle full open and about 50% of syngas exiting through the AGR absorber vessel, with the remainder bypassing to the flare. The gradual process of diverting syngas to the AGR was tested by closing the compressor recycle. During this effort, a sudden increase in suction to the compressor inlet occurred, dropping gasifier pressure excessively and blowing the fluidized bed material out of the gasifier and into the hot filter, terminating the test. This was attributed to the learning process of how to properly tune the compressor controls.

The second shakedown **Test S-2** proceeded on schedule during the week of December 10, 2012. The purpose was to test and evaluate the AGR section operating with pressurized syngas from the gasification section. Five AGR steps were carried out in this test: (1) establish absorption at intermediate pressure, syngas flow, and solvent flow; (2) chill the solvent and add heat to the stripper; (3) increase pressure and syngas flow rate; (4) increase solvent flow rate; and (5) reduce pressures in regeneration vessels. The main target of bringing CO_2 down to the optimal range of 2 to 5 Mole-% in the conditioned gas was met, with CO_2 as low as 1.2 Mole-%. The biomass feed rate was also increased to 90% capacity for a short time. The test also was used to establish procedures for trips and shutdowns of plant sections if necessary during a long test, confirming that the gasifier and cleanup steps can survive a syngas compressor trip without a complete shutdown. This was verified during the test. The AGR was operated for 25 hours over a 35-hour span; with syngas feed up to 2700 lb/h, which is 75% of design capacity.

Problem areas identified included a failure of hot gas filter which allowed particulates to pass through to the tar reformer and scrubber. This required a cleaning of the first stage of reformer catalyst following the test, and replacement of the filter candles with new candles using a silica-treated iron chromium aluminum alloy for better corrosion resistance. Erratic performance of the AGR solvent pump was also observed during the test, which was addressed during turnaround by mechanical modifications to pulsation dampers and pressure relief valves. A third issue was the inability of the chiller to reach the design temperature of 20°F, which was tracked down to a software issue and corrected during turnaround. Two gas flowmeters and a scrubber level control also presented minor problems during the test, and steps were taken to correct or upgrade these instruments.

Following Test S-2, TIGAS catalysts were loaded into all of the TIGAS reactors. A new 456-ton batch of feedstock pellets from bark-on aspen logs was produced at Indeck during the first week of January 2013. A final Test Readiness Review meeting was conducted on January 8-9, 2013.

The shakedown **Test S-3** was conducted on January 14-22, 2013, with the goal of shaking down the TIGAS gasoline synthesis section and the tank farm, and confirming the ability to make gasoline. Prior to startup, catalyst reduction with hydrogen was completed. Test operations began with preheating of the gasifier and tar reformer. Biomass feeding with the new feedstock batch began on the evening of January 16, 2013. Biomass feed rate was brought up to about 1450 lb/h and maintained at that rate throughout the remainder of the test. Tar reformer temperatures were lined out early morning of January 17, 2013, and the syngas compressor and AGR unit were started in late evening. In parallel, the TIGAS reactors were preheated under nitrogen using steam from the onsite boiler. Syngas was introduced to the TIGAS plant on January 18, 2013, and reaction products started showing up in the HP separator around 5 PM. Various interruptions were experienced as staff made adjustments to operating conditions over the next 60 hours, primarily to establish proper temperatures in the TIGAS reactors. Stable gasoline synthesis at about 12 lb/h (~5% of capacity) was established at 5:10 AM of January 21, 2013, and maintained for about 7.5 hours. During the entire synthesis period of 68 hours, about 100 gallons of hydrocarbon product were made. Operational problems with recycle compressor R901 were revealed during the test.

On January 22, 2013 Test S-3 was terminated primarily due to (a) operational problems with the AGR plant because of very cold nighttime temperatures causing AGR fluid freezing in some lines, eventually leading to a pump failure; and (b) pressure buildup across the hot gas filter, which was later attributed to a correctable design flaw with the back-pulsing manifold inside the filter vessel. Overall, the plant logged 134 hours of biomass gasification, 78 hours of AGR operation, and 52 hours of TIGAS operation.

The shakedown tests resulted in the identification of correctable bottlenecks, trips, weak points, and opportunities to improve operating procedures.

A nitrogen leak in the biomass feed unloading and storage system was found and repaired, which subsequently reduced nitrogen consumption. In the hot gas filter, significant bridging of solids was found on one set of candles, which was traced to misalignment of pulse nozzles. The spider that holds the nozzles in position was rebuilt to correct this problem. The silica-treated candles that were installed for Test S-3 were in good condition and were reused in Test #1. The candle supplier Porvair also reported that the problem of the previous set of untreated candles in Test S-2 was caused by oxidative degeneration. The startup procedure was modified accordingly. A number of problems in Test S-3 were caused by extremely cold weather, including failure of a control card in the cryogenic N_2 supply system, freezing of Morphysorb solvent at several points in the AGR system, and damage to an oil cooler in the CM800 syngas compressor. All of these were addressed by either improving heat tracing and insulation or implementing some other form of heating. New quench nozzles and a new level transmitter on the scrubber were determined to have performed well in Test S-3. Numerous improvements to AGR were made, including automating some sequences, increasing heat tracing and insulation, installing a second circulation pump in the system, adding piping to and from solvent storage tank to allow adding more solvent during live testing, repacking pressurization pumps, upgrading PRVs to prevent unwanted response to pressure pulses, and replacing pulsation dampers with a different design to eliminate leakage. There were several improvements to Safety Instrumented System (SIS) logic. One of the oxygen flowmeters to the gasifier was repaired and recalibrated. Instrumentation for the management of chilled coolant between AGR and TIGAS was improved. Startup procedure for TIGAS related to external and internal recycle flows based on lessons learned during Test S-3 was improved. Sampling facilities were enhanced with the addition of methanol, DME, and C_3-C_{5+} hydrocarbon analyses.

6. Woody Biomass Feed Delivery, Handling & Quality Assurance

6.1 Feedstock Characteristics

All biomass feedstock for the gasification process was grown and produced in Minnesota. Economically viable woody biomass can come from several sources including pre-commercial thinning materials, wood waste and wood residues, and/or round-wood that would not otherwise be used for higher-value products (e.g., rotten, malformed, non-marketable species). For this project, non-merchantable round-wood was used. Minnesota forests have been analyzed extensively using FIA (Forest Inventory and Analysis) methods. The annual allowable harvest for the State of Minnesota is 5 million cords per year and it has not been reached. The wood harvested in the early 2000's was 3.8 million cords per year but with the crash of the housing market which closed a number of board plants and several paper machines, the amount of wood harvested averages about 2.5 million cords per year. Because of the lack of markets, there is plenty of wood from round-wood to woody biomass.

All wood for this project was from non-federal lands in Minnesota, harvested according to Minnesota Harvesting Guidelines, and transported by truck. Storage was on site in the forest or in piles at the Blandin paper mill site. Wood grown on public lands is available subject to sustainability criteria for major forest certification standards. Public timber is sold in either oral or sealed bid auctions. Stumpage prices are available online via this website:

http://www.dnr.state.mn.us/forestry/timbersales/stumpage.html

Private stumpage is purchased through negotiated prices. Biomass stumpage data varies by public agency, and is negotiated with private forest owners. Minnesota DNR (Department of Natural Resources) website contains price information:

http://www.dnr.state.mn.us/forestry/um/biomass/index.html

6.2 Qualifications, Experience with Feedstock and Source Management

Blandin Paper Company has purchased its own pulpwood for about 35 years. Forestry wood procurement staff is experienced with stumpage evaluation and contracting with independent professional loggers to harvest and transport pulpwood to the Blandin mill and other mills, currently 650 cords per day (equivalent to about 267,500 bone dry short tons/year). Some logging contractors are making investments in wood grinding and chipping equipment to supply a growing wood-based biomass industry in Minnesota primarily for energy facilities. The Minnesota DNR has developed specific harvesting guidelines for biomass harvesting:

http://www.frc.state.mn.us/Info/MFRCdocs/forest%20biomass%20harvesting.pdf

There are several sources and mechanisms for handling the biomass:

1. UPM Blandin Paper forestry contracts with local loggers to cut and haul wood to a pellet plant.
2. Mill residue is purchased similar to Rapids Energy Center (power plant adjacent to the Blandin paper mill) which currently buys chips and mill residue, half of this will be available.
3. In-woods residue created by loggers with grinders (or contract grinders) would collect tops and defective wood plus low-cost round wood, process in woods, haul to mill.

4. Loggers would haul round wood to the Grand Rapids mill site where an on-site contract grinder would process the material.

6.3 Pilot Project

Because of logistics and convenience, pellets were used instead of chips or ground wood residues. The pellets were produced from aspen (*Populus tremuloides*) round wood with the bark left on to simulate logging residue and mill wastes. The aspen was harvested from various locations in northern Minnesota and trucked to Indeck Energy Ladysmith in northern Wisconsin to be processed into pellets.

Figure 6-1 Local Logger Cutting and Hauling Wood

Figure 6-2 Indeck Energy Ladysmith

The pellets were made softer than the standard fuel pellets with the crush strength being 38% lower and the moisture content at 10% moisture or less by weight. The pellet diameter was approximately ¼ to ½ inch with the length being random from ½ to 2 inches. The pellets were stored in SuperSacks for transport by a common carrier to a warehouse near GTI. From the warehouse, a Bulkmatic truck was used for pneumatic transfer to a silo at GTI. There was 2,245 cords of wood used to produce 2583 BD short tons.

Figure 6-3 Pellets in SuperSacks

Figure 6-4 Bulkmatic Trucks

6.4 Commercial Project

A commercial scale plant would use about 3200 bone dry short tons/day, the wood sourcing and feedstock would be the same as above. The table below summarizes the cost from various sources.

Source Category	Bone Dry Short Tons/day	Cost per Bone Dry Short Tons (2009)	Cost per Bone Dry Short Tons (2014)
Existing REC	250	$48.32	$49.14
In-woods chips, logging residues	260	$57.36	$53.16
Round wood at mill site, contract grinding	2690	$63.32	$64.26
TOTAL	3200	$61.65	$62.17

Table 6-1 Cost of Feed Stock

7. Summary of Operations

This chapter will describe the pilot plant operations carried out during this project, focusing on challenges encountered and solutions implemented. Shakedown testing has already been covered under Chapter 5 Project Construction and Commissioning. This chapter summarizes Tests #1 through #3 and the Independent Engineer Test.

Construction and commissioning of the IBR were completed in February 2013. Extended testing commenced in March 2013 and was completed in March 2014. Detailed Test Reports are attached in Appendices B through D.

7.1 Facility Overview

A detailed process description was presented in Chapter 4. A high-level overview is also given here to help clarify the operations summaries.

Figure 7-1 is a block flow diagram of the pilot IBR (Integrated Bio-Refinery) at GTI. The shaded areas represent the three main process sections: gasification and gas cleanup (tan), compression and acid gas removal (blue), and gasoline synthesis (green). Note that the ultra-hot gas filter (UHGF) and test batch tank shown below were not installed and tested until Test #2.

The nominal pilot IBR capacity is 1930 lb/h (23.2 short tons per day) of as-received (6 wt% moisture) biomass input to produce 246 lb/h (22.5 bbl/day) of gasoline blend stock.

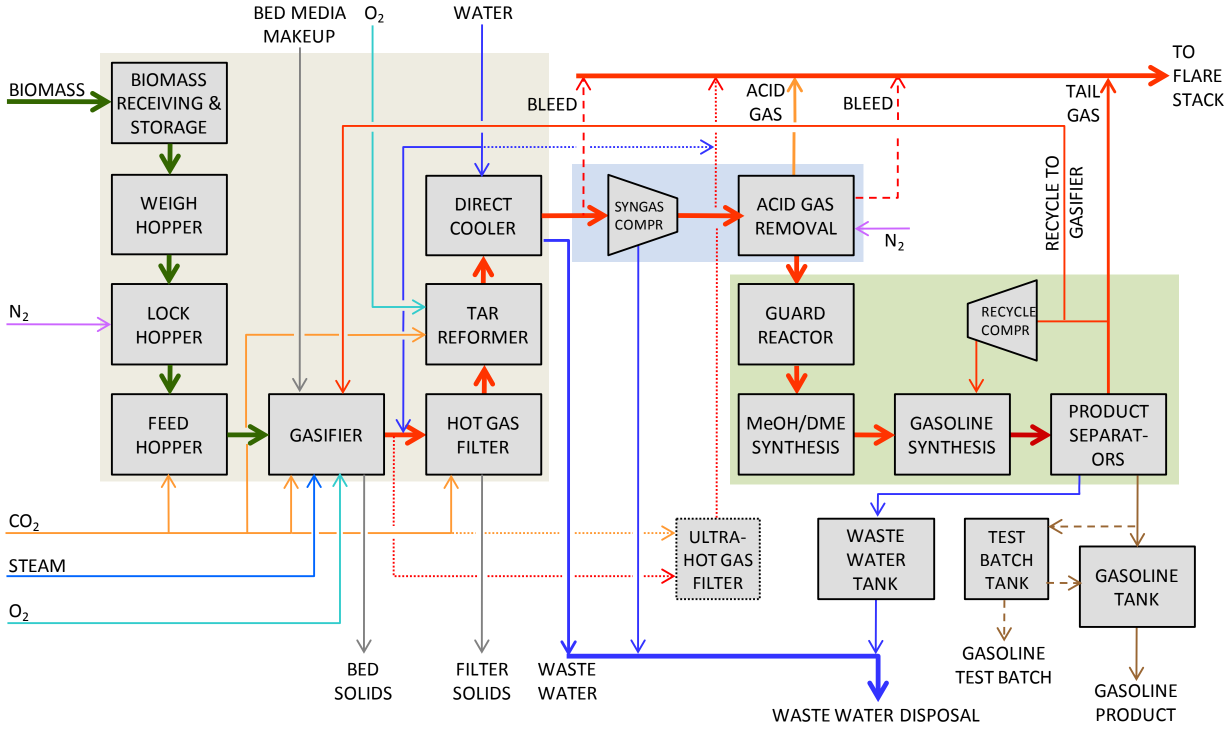

Figure 7-1 Pilot Plant Block Flow Diagram

Utilities which are included in the figure include: biomass storage silo, wastewater storage tank, and gasoline storage tank. Utilities which are not shown include cooling tower, air coolers, cooling fluid chiller, cryogenic gas storage and delivery, steam boiler, deionized water supply, and flare system.

Stream flows are color-coded as follows: green = biomass; red = syngas; violet = N_2; orange = CO_2; med blue = steam; aqua = O_2; dark blue = water; gray = bed media/ash; brown = gasoline. The dashed lines indicate syngas bypasses for startup and control during transients, troubleshooting, or batch collection. The dotted lines denote a non-core component (slip stream filter) that was installed and tested under a separate project.

7.2 Test #1

7.2.1 Operations Summary

The main objective of Test #1 was to operate the IBR at steady state for up to 360 hours, including 72 continuous hours to satisfy the Independent Engineer Test (IE Test) protocol. Secondary objectives were:

1) Validate operational stability and controllability under extended operation.
2) Observe and analyze changes in product quality related to catalyst aging.
3) Produce and analyze process waste water sufficient to finalize local discharge permit.
4) Produce and analyze gasoline product for single engine emissions testing.

The operations crew started up the IBR on March 5, 2013 with the gasification/gas cleanup section producing raw gas that was diverted to the flare through the first bleed. The next step was directing main syngas flow to the syngas compressor and AGR, with sweet syngas from the AGR diverted to the flare through the second bleed. Syngas was then directed to the TIGAS section on March 7. TIGAS tail gas was not recycled to the gasifier during this test.

Various TIGAS recycle compressor trips were experienced through March 11, which were later attributed to condensation of water and/or hydrocarbons that required design and operational changes, such as installation of a control valve on a recycle gas startup line and trimming coolant flow to a syngas heat exchanger upstream of the compressor, were implemented over the following test campaigns. Two kinds of trips were experienced: vibration and rod drop[*]. In this test, safe startup conditions were identified whereby the recycle compressor was started up without vibration trips from March 12 onwards.

On March 12, the IBR entered it first period of full operation and the first bulk quantity of gasoline was produced.

On March 13, the TIGAS section was shut down due to syngas leakage into the test bay. Flanges were re-torqued to eliminate leakage. One flange leak could not be resolved in this way, but a temporary vent was devised. That flange was later modified to limit stress and permanently eliminate the leak.

Also on March 13, the hot gas filter (HGF) showed a sharp upturn in pressure drop. Tar reformer pressure drop also began to rise. Adjustments to the tar reformer O_2 and CO_2 feed rates were implemented to slow down the increase in pressure drop, but these were found to cause fluctuations in the naphthalene levels downstream of the tar reformer. Rising pressure drops were also experienced in the gasifier lower bed.

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[*] The compressor shaft passes through a seal between the drive motor and the engine. A sensor is installed to detect excess wear of that seal, and “rod drop” is the manufacturer terminology for that signal.

Consequently, the gasifier was shut down on March 13 for inspection and cleaning, and was restarted on March 16.

On March 17, during main syngas compressor restart, the compressor tripped due to naphthalene plugging of the inlet strainer. The strainer was cleaned and the syngas compressor successfully brought on line. An additional strainer was later added to allow cleaning in the event of a system upset that resulted in transient high naphthalene levels.

On March 18, the TIGAS section was started and gasoline production resumed. However, TIGAS tripped due to a false high level signal from a relief knockout pot. The trip was bypassed for further operation and the sensor was replaced. Also on March 18, rod drop trips previously mentioned with the recycle compressor were experienced, and this situation was rectified by the manufacturer on March 19. The recycle compressor vendor Knox Western, which was selected by Zeton (the TIGAS system fabricator) did not have a proven design for this size and application. A more thorough vetting of bidders for this recycle compressor may have helped avoid this. Before TIGAS could be restarted, the main syngas compressor tripped on March 19 and 20, again due to naphthalene residues. Continued buildup in tar reformer second stage pressure drop was also observed. On March 21, adjustments to O_2 and CO_2 distribution in the tar reformer were successful in reducing naphthalene content in the syngas. The syngas compressor and TIGAS were successfully started, and gasoline production resumed for the following three days.

On March 22, the test was ended due to high HGF pressure drop. Total run time for Test #1 was 299 hours of Gasification, 215 hours Gasification with AGR, and 47 hours with the full IBR.

Test #1 satisfied secondary objectives 1 and 2. The IE Test objective, as well as secondary objectives 3 and 4, were deferred to Test #2.

Total biomass fed during Test #1 was 212 short tons. During gasoline production, total biomass fed was 34 short tons. Total raw gasoline produced during Test #1 was 1.8 tons, or approximately 573 gallons.

7.2.2 Post-Test Inspection and Improvements

The HGF was later inspected, revealing that one candle was broken at the weld and the metallic filter media itself had deteriorated at some locations. The cause of this was later identified as corrosive breakdown of the filter material. The process team decided to abandon metallic HGF elements previously supplied by Porvair, and work with Pall Corporation to install ceramic filter elements commercially referenced for biomass-based syngas

The tar reformer was also inspected. Dust buildup from the failed filter and erosion to one of the oxygen distributors was found. A new oxygen distributor and an improved syngas distribution system were designed and implemented for Test #2.

Gasoline yield did not reach the expected value for the syngas rate and composition that was measured at the inlet to the TIGAS section. The project team concluded that the main reason for this was that the gasoline catalyst volume was over-dimensioned, causing partial cracking of the gasoline before it leaves the reactor. Therefore, part of the gasoline catalyst was removed for Test #2. Another cause was high N_2 content in the syngas due to higher-than-expected N_2 back-pulsing of the HGF. To correct this, the back-pulsing gas was changed from N_2 to CO_2 for Test #2. A final step to raise gasoline yield was to increase the TIGAS operating pressure and reduce the syngas bleed stream upstream of the syngas compressor to convert more syngas.

Changes that were made prior to the subsequent test were as follows:

- Syngas filter: changed from metallic to Pall ceramic candles, new back-pulse system using CO_2 instead of N_2.
- Tar reformer: repaired oxygen lance, redesigned syngas distributor.
- Ultra-hot gas filter (UHGF): installed Pall high-temperature ceramic candles in existing refractory-lined filter vessel, sized for slipstream at 10-15% of syngas flow.
- Acid gas removal: improved instrumentation, upgraded piping to reduce leakage and increase solvent circulation rate, added solvent strainer, reduced carryover, installed continuous CO_2 monitor on sweet gas for improved real-time control.
- Gasoline synthesis: removed half of the catalyst bed to adjust the kinetic balance between conversion and cracking.
- TIGAS recycle compressor: upgraded shaft seal and installed bracing to reduce vibration.
- Strainer: installed a naphthalene guard strainer in the cool dry syngas exiting the direct cooler upstream of the main syngas compressor.
- Leak avoidance: installed “Superbolt” tensioners to facilitate hot torqueing of large vessel flanges in the TIGAS section in the event of leak development; added an expansion loop and reoriented the heads on one of the TIGAS heat exchangers to correct a persistent leak found during Test #1.
- Control: installed a new pressure control valve upstream of the gasoline reactor to open and depressurize the system in the case of a safety system trip; reoriented coriolis-type flowmeter measuring high-pressure gasoline flow, to allow more accurate reading with two-phase flow.
- TIGAS recycle compressor: replaced intake caps and valve cages to reduce the compressor amp draw and installed a new, redesigned packing set to correct a persistent leak during Test #1.
- Gasoline collection: installed a temporary 550-gallon sample batch tank to isolate 500-gallon cuts of gasoline product, in order to capture an optimal gasoline batch for engine emission testing.

7.3 Test #2

7.3.1 Operations Summary

The main objective of Test #2 was to operate the IBR at steady state for up to 480 hours, including 72 continuous hours to satisfy the Independent Engineer Test protocol. Secondary objectives were:

1) Maximize throughput and identify the limits of operation, primarily pressure and temperature limits.
2) Maximize gasoline production at maximum achievable biomass feed rate.
3) Demonstrate steady operating conditions at higher severity to gain knowledge and experience with respect to establishing design parameters for industrial plant layout.
4) Observe and analyze changes in product quality related to catalyst aging.
5) Collect and analyze process waste water samples to complete requirements for local water discharge permit.
6) Produce 500 gallons of gasoline product for single engine emissions testing.

Operators started the IBR on October 19. Gasification was delayed to clean up a water leak into the biomass feeding system. AGR and TIGAS systems were brought into ready/idle state during the cleanup. Once the cleanup was complete, operators started feeding wood in the evening of October 21, and made the switch from combustion to gasification in the morning of October 22. The slipstream UHGF was on line from the beginning at 10% of the total syngas flow.

The main syngas compressor was started up on October 22, followed by an AGR stabilization period. One compressor trip was encountered because of an incorrectly positioned cooling water valve, but that was corrected and the compressor ran without incident from that point on. TIGAS was brought on line on October 22 and gasoline production began on October 23.

Logging of IE Test conditions started on October 23 at 18:30. However, the IE Test was interrupted 56 hours into the IE Test when a cracked pipe weld required the TIGAS section to be taken offline for repairs. The Gasification and AGR sections of the plant continued operating without interruption during the TIGAS outage, and the TIGAS section was brought back online at midday on October 27. During this time, a 457-gallon batch of gasoline was collected for possible use in engine emissions testing, and gasoline collection in the batch tank resumed in the afternoon on October 27. Gasoline aromatic content was already below the upper target (<35 vol%) for engine emission testing and was gradually decreasing. Also during this initial period, the UHGF was taken out of service on October 25 because of solids appearing in the exit line. This did not adversely affect the main syngas stream through the HGF, which thereafter received 100% of the raw syngas flow without incident.

Starting on October 28, a series of steps were taken to maximize syngas conversion: reducing pre-compression syngas bleed, introducing recycle gas to the gasifier, and increasing biomass feed rate. The feed rate was lowered back to the earlier rate after 14 hours because of some irregularities observed in the gasifier bottom discharge, and recycle gas was discontinued after 49 hours to ensure uninterrupted collection of two 500-gallon batches of gasoline for engine emissions testing.

On November 1, prior to planned shutdown of the IBR, the operators changed conditions in the Guard Reactor to define the boundary conditions for temperature in that vessel. As expected, the temperature limit was reached and the test was terminated.

About 680 gallons of gasoline was shipped to Southwest Research Institute (SwRI) in San Antonio, Texas, for single-engine emissions testing, which is discussed in detail in Chapter 9.

Following the test and submission of a draft IE Test Report, the Independent Engineer and the project team agreed that because of the interruption in the original IE Test attempt, a later uninterrupted 72-hour period would be designated as the official IE Test Period. We selected the 72-hour period of 10/29/13 through 10/31/13. All of the IE Test protocol requirements were met for that duration. A copy of the IE Test Report is provided as Appendix E.

Test #2 satisfied the IE Test objective and all of the secondary test objectives.

Total biomass fed during the test was 182 short tons, of which 150 short tons was fed during gasoline production. Total raw gasoline produced was 11.9 short tons, or approximately 3954 gallons. Total run time for Test #2 was 248 hours of Gasification, 235 hours of Gasification with AGR, and 194 hours with the full IBR.

7.3.2 Post-Test Inspection and Improvements

No anomalies were found in the gasifier, but inspection revealed that the gasifier bottom ash screw seal needed replacement. HGF components, including the new Pall ceramic candles were in good shape. The candles were removed and cleaned with a 20% citric acid solution per Pall's recommendations. The tar reformer catalysts were not removed or replaced. No adverse conditions were found during inspection of the direct syngas cooler or syngas compressor. Normal maintenance was performed in accordance with manufacturer's instructions.

Inspection of the UHGF revealed that the candles were unbroken, but one of the ceramic seals had failed, allowing solids to break through. A gasket design modification was devised in consultation with Pall Corporation.

In the AGR skid, one of the solvent-solvent heat exchangers was found to be partially plugged on the low-pressure side, but opening a bypass valve was sufficient to continue operation and with the planned heat load and weather conditions, it was not deemed necessary to dismantle and clean the heat exchanger for the remainder of testing. Solvent from Test #1 was drained and stored for reuse in Test #2. No significant carryover loss of Morphysorb® was observed in this test.

The TIGAS subsystem was blocked in under nitrogen to prevent air infiltration that would compromise TIGAS catalysts. Although test data indicated some loss of activity, as expected, of the methanol-DME catalyst, Topsoe decided not to replace any of the catalysts for Test #3. The gasoline synthesis catalyst also showed the expected signs of deactivation, validating Topsoe's kinetic model.

A series of changes were made to the IBR during the December 2013-March 2014 downtime. These included:

- Recycle gas: added provisions for recycle gas to the biomass injection screw purge, which can be switched between CO_2 and recycle gas, in addition to the previously used recycle gas inlet port in the gasifier bottom section to test the feasibility and effect of bringing recycle gas to the gasifier at alternate locations. The operations team also installed two new flow controllers for these recycle gas inputs.
- Syngas bleed streams: installed bypass piping and valving for the pre-compression syngas bleed to obtain direct measurement of the bleed flow.
- Controls: relocated transmitters for TIGAS tail gas flowmeters, removed an unneeded check valve, and installed flex hoses in and out of the recycle compressor to minimize possible vibration effects on flow instruments and piping; installed a new flowmeter to measure CO_2 purges in the gasification section between the first quench and the tar reformer inlet; upgraded various other controls to reduce operator intervention requirements and minimize nuisance trips that were observed in previous tests.
- TIGAS recycle compressor: replaced shaft packing with redesigned seal hardware to allow the seal to run cooler and with reduced leakage.
- Filters: modified the HGF candle seal fixture to reduce gasket compression, replaced pulse valve sealing surfaces, and replaced and redesigned the UHGF candle seal that had failed in Test #2 after about 70 hours. The seal was redesigned after consultation with Pall Filter, who advised that seal compression was too low. An electric heater (to compensate for heat loss because of small size of filter vessel) that failed in Test #2 was also replaced, and heat tracing was added to the solids collector to avoid condensation.

- AGR: cleaned up Morphysorb® solvent by circulation through a carbon bed to remove hydrocarbons.

7.4 Test #3

7.4.1 Operations Summary

The main objective of Test #3 was to conduct a final demonstration run at a fixed set of optimized operating conditions as identified based on the operational experience gained through Tests #1 and #2, and supported by data analyses and models derived through previous tests. Secondary objectives were:

1) Validate the efficacy of modifications made after Test #2.
2) Maximize gasoline production at maximum achievable biomass feed rate.
3) Produce sufficient raw gasoline (up to 10,000 gallons) to support moderate fleet testing.
4) Observe and analyze changes in product yield and quality as a result of catalyst aging.

Test #3 startup activities commenced on March 2, 2014 with AGR and TIGAS brought to idle conditions under nitrogen, but difficulties in obtaining full flow of AGR solvent because of partially frozen solvent delayed the startup of biomass feeding until March 4. Operators started feeding wood to the gasifier around noon on March 4, and syngas flow to the UHGF was also started at that time. The biomass feed rate was raised for transition to gasification late on March 5, with maximum feed rate of about 1600 lb/h achieved in early afternoon of March 6.

The main syngas compressor was then started and AGR conditions were then stabilized to hold CO_2 between 2 and 3 vol% at the absorber exit. Flow of sweet gas to TIGAS was started on March 6. Gasoline flow from the TIGAS separators began about 4 hours later on March 6, adding to approximately 2810 gallons that were left over from Test #2.

On March 7, flow to the UHGF was increased from 11% of raw syngas to 15% in order to sustain the filter at the desired minimum temperature of 1300°F. Bleed to flare was adjusted to maintain constant flow to AGR and TIGAS. That same day, the first tanker pickup of gasoline took place, which was also an opportunity to commission the gasoline transfer hardware and procedures. A load of 2865 gallons was removed from the T-717 tank and shipped to a blending facility to hold while awaiting the remainder of the product later in the test.

Recycle of TIGAS tail gas to the gasifier was started on March 7, and recycle to the gasifier biomass injection screw was added 24 hours later. Some temperature fluctuations in the bed solids discharge section of the gasifier indicated possible disruption in solids removal, but operators continued both recycles until March 9, when another syngas leak in the TIGAS bay necessitated shutdown of the TIGAS section. Gasification and AGR operation continued through the TIGAS downtime, with product gas diverted to the flare.

The source of the syngas leakage was found to be cracks in the flexible hoses installed around the recycle compressor to isolate vibration. Those lines were replaced with hard piping, and TIGAS was brought back into service on March 10. Gasoline production then continued for the remaining six days of the test.

The UHGF was taken out of service on March 11 due to plugging of a sample line, but this did not disrupt main IBR operations. The extra syngas was easily accommodated by the AGR and TIGAS sections.

To continue evaluation of tail gas recycle, operators restarted recycle gas flow to the injection screw on March 11. Recycle was not restored to the gasifier bottom because of concerns that bed discharge might be compromised. The recycle was maintained for 47 hours and then discontinued on March 13 to facilitate a comparison with and without recycle. Recycle gas was replaced with CO_2 in the screw purge, and flow to TIGAS was also reduced by 6% in response to lower syngas throughput. This required adjustments in the AGR operation to restore CO_2 content in the sweet gas to a constant 2 vol%. At that time, conditions were held steady to obtain mass balance and performance data without tail gas recycle.

Flow to TIGAS had to be reduced by 15% on March 15 because the steam condenser which controls temperature in the methanol-DME reactor could not reject enough heat. This was traced to an installation error which had not previously been detected

Full IBR operations continued at steady conditions until a voluntary shutdown on March 16. Gasoline production during the second TIGAS operational period of 135 hours totaled 3742 gallons. A final tanker pickup on March 19 totaled 4881 gallons, so the total of the two shipments to the blending facility for preparation of finished gasoline for the fleet test was 7746 gallons. Smaller batches were also shipped to Phillips 66 and/or Haldor Topsoe for further analysis and testing.

All of the Test #3 test objectives were met.

Total biomass fed during Test #3 was 188 short tons. During gasoline production, total biomass fed was 160 short tons. Total raw gasoline produced during Test #3, based on the averages of flow sensors and tank levels, was 16.0 tons, or approximately 5243 gallons. Total run time for Test #3 was 257 with Gasification, 242 hours with Gasification and AGR, and 196 hours with the full IBR.

7.5 Challenges and Solutions

Pilot testing of the IBR yielded commercially-relevant information on equipment, operating procedures, startup/shutdown procedures, safety systems, upset conditions, redundancy, and a host of other key issues. Below is a partial list of challenges encountered, how they were met by the project team, and how they relate to scale-up and future commercial-scale projects based on this technology. Lessons learned in connection with the project execution are described in Chapter 11.

7.5.1 Biomass Feeding

CHALLENGE: Feeding of wood in the form of chips and/or pellets had previously been demonstrated in the GTI pilot plant using existing equipment, but not at design capacity for extended periods in excess of about five days. In prior testing, wood was supplied in 1800-lb sacks, moved by fork truck from storage, and manually unloaded into a mechanical transport system to the weigh bin at the top of the gasification building. This was considered too labor-intensive and risky for reliable 24/7 feeding for up to 30 days.

> SOLUTION: We commissioned an existing pneumatic truck unloading system and 6500 ft^3 storage silo, previously used for coal, for biomass. The storage silo contained 3 to 5 days of feedstock, and bags were also held in reserve as a hedge against delivery interruptions. This proved effective, as it allowed the system to remain in operation through winter weather which could have impeded operation with the old method, and also in the instance when a delivery truck had the wrong pneumatic connections and could not deliver its load. In all 804 hours of gasifier operation in the program, not a single interruption was experienced due to biomass feed unloading, handling, or feeding to the gasifier.

7.5.2 Gasification

CHALLENGE: The Andritz-Carbona gasifier was not initially outfitted for extended operation beyond five continuous days in pilot operation. Bottom ash withdrawal, wall effects at high feed rate contributing to high levels of heat loss, and gradual residue buildup on injection screw were identified as key impediments to long-term operation at high capacity.

> SOLUTION: We built and installed a new inclined water-cooled bottom ash transport screw, implemented input gas preheating with steam, and modified the feed injection screw design. These improvements were commissioned in the first shakedown test (S-1) and associated procedures were updated. These steps resulted in enhanced reliability of gasifier operation. By the end of the testing campaign, all parties were satisfied that the gasifier could operate indefinitely at more than 80% design capacity.

7.5.3 Tar Reforming

CHALLENGE: The tar reformer must hold tar concentrations, particularly naphthalene, at very low ppm levels for an extended time to avoid fouling downstream equipment, contaminating the AGR solvent, and possibly poisoning TIGAS catalysts. In previous pilot test programs with biomass, the tar reformer was operated in “dirty” mode (upstream of the hot gas filter) to minimize heat losses, as is done in the Andritz-Carbona commercial gasifier at a combined heat and power plant in Skive, Denmark.

> SOLUTION: Our project team determined from prior experience that the “dirty” mode of tar reforming would not be reliable enough at pilot scale to meet the more stringent requirements of a gasoline synthesis system. Therefore, the “clean” reformer mode in which the tar reformer is located downstream of the hot gas filter was employed.

CHALLENGE: In the “clean” reformer mode, the catalyst is supported on pellets rather than formed bodies (monoliths), requiring particulate removal to below 10 ppmw for extended operation. During operation, some carbon deposition was observed which was attributed to poor distribution of syngas and oxygen. This was also thought to be responsible for some irregularities in the reforming performance, including some short episodes of naphthalene breakthrough.

> SOLUTION: To correct this, we designed and installed a modified syngas distributor plate and improved oxygen distributors after Test #2. These changes resulted in more stable performance in Test #3.

7.5.4 Hot Gas Filter

CHALLENGE: The hot gas filter (HGF) must remove entrained solids down to single-digit ppmw level to protect tar reformer performance. Initially, metallic filter candles that were used in previous tests were used for the IBR tests, but some media degradation was observed in shakedown testing, requiring tar reformer catalyst removal for cleaning. A new filter candle with silica coating was then installed, but that element also failed in two ways: (a) breakdown of the filter media allowing solids leakage, and (b) failure of a seam weld, allowing bulk passage of solids. Metal dusting was suspected but not confirmed.

> SOLUTION: The project team switched to an advanced ceramic-based filter candle from Pall Corporation. This required a change in the filter vessel head and the back-pulse system as well. Pall provided both the candles and the pulse valves, and the IBR project provided a new filter vessel head. Pall also supported a separate effort to test a higher-performance “ultra-hot gas filter” (UHGF) that was tested on a 10-15% slipstream of the dirty gas exiting the gasifier. Both of the ceramic elements performed well and established ceramic elements as the preferred solution for commercial application.

7.5.5 Gas Cooling

CHALLENGE: The direct syngas cooler, which was sufficient for prior testing, did not have enough heat removal for the subsequent syngas compression. Also, level control was not found during shakedown testing to be robust enough for integrated operation.

> SOLUTION: We installed an additional heat exchanger using cooling tower water to supplement the existing air cooler, and upgraded the level control to a delta-P cell for more reliable cooler tank level control.

7.5.6 Syngas Compression

CHALLENGE: The newly installed syngas compressor required cool, clean syngas at a steady flow to supply high-pressure syngas to the AGR. Compressor trips had the potential to send pressure pulses both upstream and downstream, disrupting gasification and/or AGR performance and possibly resulting in a complete shutdown. Another challenge was heat and condensate management in the compressor. In the shakedown testing and continuing into Test #1, compressor trips occurred which delayed fully integrated operation. Instances of naphthalene breakthrough also occurred, resulting in loss of syngas flow due to plugging of the inlet strainer.

> SOLUTION: Extensive training and staff experience, augmented by vendor training, was key to achieving trouble-free operation. Setting operating margins between normal operation and trip points was part of that learning curve. Startup and shutdown procedures were developed and refined. The use of a pre-compression syngas bleed stream to the flare was helpful in quickly switching syngas from the AGR back to the flare when trips occurred. Management of internal recycle was also developed and refined over the course of shakedown and Test #1. An existing vessel that was not originally part of the IBR design was pressed into service as a pre-compression surge tank, helping to cushion the impact of pressure pulses during startup. Drainage of condensate from KO pots was automated. Finally, a pre-compression "naphthalene trap" was added to the system inside the process building to prevent compressor outage in the instance of a momentary breakthrough at the tar reformer. All of these steps resulted in very reliable compressor startup, operation, and shutdown by the time Test #3 was conducted.

CHALLENGE: Excessive pre-compression bleed to flare, which wasted part of syngas that should be used to increase gasoline yield. The existing control loop for diverting syngas between the compressor and the bleed did not afford low-level flow monitoring and/or fine control over flow.

> SOLUTION: We added a low-level flowmeter and bypass loop to allow gradual reduction of the bleed stream, and also tuned the controls to permit a rapid but controlled switch to bleed in the event of a compressor outage. This was fully implemented in Test #3 and worked well, allowing operators to reduce bleed to less than 5% of syngas flow.

7.5.7 Acid Gas Removal

CHALLENGE: The AGR system was a refurbished unit previously operated on natural gas at lower pressures and flows. It was upgraded for syngas, but had not yet been commissioned for that service at the pressure and capacity needed for the IBR. In addition, the Morphysorb® solvent needed to be chilled, requiring a new heat exchanger to cool the solvent with a coolant solution provided by a new chiller. The AGR system needed to be brought to the same level of reliability and resistance to upsets as the rest of the IBR. We began commissioning the AGR using nitrogen. Following that, shakedown Test S-2 was conducted to commission the AGR on syngas at meaningful conditions, which required gasifier and syngas compressor operation. Among the troubleshooting needs encountered starting with Test S-2 and continuing through Tests #1 and 2 were: (a) leaking pipe unions on high-pressure solvent side; (b) erratic flow from pressurization pump; (c) solvent carryover from absorber; (d) poor temperature distribution in absorber indicating inadequate gas-liquid contacting; (e) solvent freezing in cold (<15°F) weather; (f) and inability to supplement solvent during live operation.

> SOLUTION: Solutions to the individual problems included: (a) replacement of pipe unions with standard pipe flanges; (b) installation of a backup pressurization pump, pulse bottle upgrades, pressure relief valve upgrades; (c) design modification of absorber top head for better gas-liquid contacting; (d) augmented heat tracing and insulation; and (e) design and implementation of a subsystem to add solvent during operation. A variety of instrumentation and control improvements were also made to achieve better automation and data collection.

7.5.8 Gasoline Synthesis

CHALLENGE: Syngas leakage from TIGAS reactors and some heat exchangers occurred at operational temperature and pressure, even though the flanges were torqued to spec as instructed by the fabricators. This was attributed to thermal expansion which reduced tension on the bolts, or in one instance (heat exchanger) from distortion of a flange seal on a close-coupled connection.

> SOLUTION: In the first instance where this leakage necessitated a TIGAS shutdown during Test #1, we mitigated the problem by flooding the TIGAS section with nitrogen and then hot-torqueing the bolts. This remedied the problem temporarily, but for subsequent testing we purchased and installed “Superbolt” bolt tensioners which could be tightened with hand tools. This approach was successful. In the case of the leaking heat exchanger, the close-coupled connection was modified to include an expansion loop. This was an important improvement for pilot testing because of the high cost of outages.

CHALLENGE: The TIGAS recycle compressor experienced vibration trips rod drop trips during startup and syngas leakage continuously, starting with Shakedown Test S-3. Gasoline synthesis cannot continue without the recycle compressor, so it was critical to arrive at a long-term solution.

> SOLUTION: The recycle compressor was a first-of-a-kind from the vendor, who concluded that the shaft seal and packing design was not adequately wear-resistant, and also led to syngas leakage. In addition, we uncovered an error in the compressor body fabrication. Both of these were corrected by the manufacturer, which took care of the leakage and wear problems. Vibration trips; however, were caused by a combination of factors, chiefly (a) transient condensation in the recycle gas during startup, and (b) excessive vibration due to inadequate mounting and stiffening of connecting lines. These were corrected by upgrading the mounting method and adding bracing to the compressor piping, and also by developing a startup procedure that kept the vibration levels below the trip point during startup.

CHALLENGE: Syngas leaks developed from cracked welds in the TIGAS section, specifically from an instrument mounting on a pipe tee, and from flexible hoses connected to the recycle compressor in an effort to reduce transmitting vibration to some key flow instruments.

> SOLUTION: After weld repair, the vulnerable instrument mounting was modified to reduce stress on the weld joint. The flex hoses were replaced with rigid piping. The lesson learned is to examine the piping and mounting of system components with more attention paid to potential stress points when vibration is encountered. Additional bracing at key points to reduce displacement due to vibration is also recommended.

CHALLENGE: The steam condenser for the methanol-DME reactor was key to reaching TIGAS capacity because it was the method by which heat is removed from the exothermic reactions in the reactor. During Test #3 operation to maximize TIGAS throughput, heat rejection from the condenser was found to be at its maximum capacity. The air flow through the air-cooled steam condenser was found to be restricted by the installation using dampers and ducts to select between inside and outside air, in response to ambient conditions.

> SOLUTION: During testing, we partially mitigated this restriction by spraying water into the air cooler inlet in order to increase heat capacity. However, we also had to reduce TIGAS input rate to avoid overheating the methanol-DME reactor. For future testing, the air condenser housing and duct dampers would need to be modified to allow capacity air flow through the condenser. It would also be advisable to increase the fan speed and motor size.

8. Catalyst Performance

The processes required for production of gasoline from raw synthesis gas derived from gasification of wood rely on a number of chemical reactions taking place on heterogeneous catalysts manufactured by Topsoe. The following catalysts were supplied to the project by Topsoe:

- Tar reforming catalyst, installed in Tar Reformer
- Hot guard catalyst installed in Hot Guard Reactor
- Methanol/DME catalyst installed in Oxygenate Reactor
- Gasoline catalyst installed in Gasoline Reactor

Each of these catalysts has a specific purpose, and the reactions taking place on each of the catalysts are briefly described in the following:

8.1 Catalyst Purpose and Reactions

The tar reforming catalyst has a primary purpose of converting tar (a carbon rich mix of partly poly-aromatic hydrocarbons) to synthesis gas and a secondary objective to convert methane in the raw synthesis gas to carbon monoxide and hydrogen. The main reactions taking place on the tar reforming catalyst are well described by the reactions:

$$C_nH_m + n\ H_2O \rightarrow n\ CO + (n+m/2)\ H_2 \tag{8.1}$$
$$CO + H_2O \leftrightarrow CO_2 + H_2 \tag{8.2}$$
$$CH_4 + H_2O \leftrightarrow 3\ H_2 + CO \tag{8.3}$$

The hot guard catalyst is primarily an adsorption mass. It comprises various types of active materials (ZnO, metals and zeolites) and is designed to adsorb all the most common types of catalyst poisons such as S, Fe, Ni, etc. Moreover, it removes activated nitrogen compounds such as ammonia and cyanides from the synthesis gas. Examples of the adsorption reactions taking place are:

$$Fe(CO)_5 \rightarrow Fe(ads) + 5\ CO \tag{8.4}$$
$$H_2S + ZnO \rightarrow H_2 + ZnS \tag{8.5}$$

Common for the reactions taking place over this catalyst is that they leave a “fingerprint” so that it subsequently is possible to determine which kind of catalyst poisons there were in the raw synthesis gas.

The Methanol/DME catalyst really catalyzes two reaction steps simultaneously – namely the methanol synthesis and the methanol dehydration to DME. Moreover, the catalyst is shift active meaning that the three main reactions taking place on this catalyst can be described through:

$$CO + 2\ H_2 \leftrightarrow CH_3OH \tag{8.6}$$
$$CO + H_2O \leftrightarrow CO_2 + H_2 \tag{8.7}$$
$$2\ CH_3OH \leftrightarrow CH_3OCH_3 + H_2O \tag{8.8}$$

Finally, the gasoline catalyst catalyzes a whole range of reactions that are well described through:

(8.8) $2\ CH_3OH \leftrightarrow CH_3OCH_3 + H_2O$

(8.9) $n/2\ CH_3OCH_3 \rightarrow C_nH_{2n} + n/2\ H_2O$

(8.10) $z\ C_nH_{2n} \leftrightarrow C_xH_y + C_pH_q$

Where reactions (8.10) represent many different reactions, the end result is that light olefins combine to a mixture of paraffins, naphthenes and aromatics with an overall stoichiometry of $-(CH_2)-$.

Due to the shape selectivity of the gasoline catalyst, the majority of components resulting from reaction (8.10) fall in the range from C_5 to C_{10} with a very sharply declining tail consisting of higher hydrocarbons. This mix of hydrocarbons constitutes a nearly ideal gasoline and therefore one of the main performance parameters for the gasoline catalyst is the so-called C_5^+ yield.

The following sections describe the performance observed for each of the catalysts installed.

8.2 Performance of Tar Reforming Catalyst

The overall performance of the tar reforming catalyst was very satisfying. Except for very brief periods where the system tripped, the catalyst resulted in a tar conversion that was far better than required for the down-stream process. Catalyst volume sizing and staging of the tar reformer was based on a rough tar reforming kinetics originally developed for sulfur passivated reforming of methane. From the test campaigns, useful information was obtained regarding kinetics adjustment. This includes the sensitivity to sulfur amount in the process gas, reaction orders for different hydrocarbons, and the site blocking effects of the different poly-aromatic hydrocarbons (tar). On the basis of the original design calculations, it was assumed that almost complete methane conversion could be achieved. However, the methane conversion obtained was limited to about 85%. Several reasons contributed to this. Firstly, as a consequence of a very poor gas distribution in the top of the reformer, a significant temperature gradient radially across the catalyst bed was observed. This inhomogeneity also made it difficult to identify the ideal operating conditions. The inlet temperature to the four catalyst beds were furthermore chosen conservatively in order to ensure a sufficient conversion during the entire operation period. Secondly, the tar reformer served as a water gas shift section ensuring the desired H_2/CO ratio (module) to the TIGAS section. This limited the ability to increase temperature and hence methane conversion in the last reforming stage. In general the operation condition was primarily selected in order to ensure an optimal syngas module to the TIGAS section as well as complete tar removal and secondarily on achieving a completely equilibrated exit gas with respect to methane steam reforming.

8.3 Performance of the Hot Guard Catalyst

The catalysts installed in the hot guard reactor were after the tests unloaded and analyzed for impurities. The analysis confirms that none of the guard beds had been saturated by poisons. We had for the pilot plant – as intended – been conservative in our design, however the analysis of the data still provided excellent information about the types and amounts of catalyst poisons that must be expected in a syngas generated by wood gasification. This knowledge will be used to optimize the design of the commercial plant.

8.4 Performance of the Methanol/DME Catalyst

The oxygenate compounds produced over this catalyst are DME (CH_3OCH_3) and methanol (CH_3OH). The selectivity between the two is given primarily by the feed gas composition; the higher the H_2/CO ratio, the higher the selectivity is towards methanol. The feed gas composition was reasonably stable throughout the tests. A typical feed gas composition (after acid gas removal) in dry mole percent is $H_2/CO/CO_2/N_2/CH_4$ = 49/42/2.4/3/3.6. With this gas composition the predominant oxygenate species at equilibrium is DME, and a typical measured exit composition included 23-26 mole percent of DME and 2-4 mole percent of methanol.

Even though the catalyst was subjected to catalyst deactivation during the time of operation the exit composition stayed close to equilibrium throughout the entire run. However, it was observed that over time, the hot spot temperature in the reactor decreased from about 720°F initially to around 630°F in the final test. Moreover, the location of the hot spot moved downwards in the bed. These observations cannot be explained by catalyst deactivation alone, but also changing process conditions such as increasing flow rate which will affect the hot spot. These observations were taken into account when developing the commercial plant lay-out described in Chapter 13. The catalyst system will be slightly modified in a commercial plant in order to minimize the hot spot while maintaining a high H_2+CO conversion. This will be obtained by modifying the ratio between methanol and DME catalyst and increase the reactor temperature slightly.

8.5 Performance of the Gasoline Catalyst

The produced hydrocarbon product contains both lighter hydrocarbon (LPG) and heavier hydrocarbons (C_{5+} or gasoline):

$$2CH_3OH \overset{-H_2O}{\underset{H_2O}{\Leftrightarrow}} CH_3OCH_3 \overset{-H_2O}{\rightarrow} \begin{pmatrix} \mathrm{C5+} \\ \mathrm{LPG} \end{pmatrix}$$

The selectivity of the desired product, C_{5+}, is normally defined as the ratio between carbon atoms ending up as C_{5+} components and carbon atoms entering the gasoline reactor loop as either methanol or DME.

The gasoline selectivity decreases at high reactor temperature. This is due to enhanced rates of cracking reactions that decompose C_{5+} components into lighter hydrocarbons. Moreover, in this context high reaction temperature is equivalent to high catalyst activity, which means that for a fixed reactor temperature the gasoline selectivity will increase over time (simultaneously as the aromatic decreases) due to the reversible catalyst deactivation by coking. The highest gasoline yield is therefore always expected late in a given cycle.

In a commercial plant a multiple, parallel reactor system is required for continuous operation, and the system of reactors operating displaced in time would in that case provide a nearly constant average product distribution.

The gasoline reactor in the pilot plant was originally designed to operate at an inlet and outlet temperature of 660°F and 745°F, respectively, and a weight hourly space velocity of 1.6 h^{-1}. After Test #1 it became clear that the design flow rate for the unit would never be reached and it was therefore decided to remove half of the catalyst (the top part) due to the reduced run-time and throughput. Even at this lower catalyst volume we never reached the end of cycle conditions and design value for the space velocity where the maximum gasoline productivity would be anticipated. By lowering the reactor temperature below the design value, we managed to reach a C_5^+ selectivity of 84%.

The amount of coke deposit on the catalyst was measured to 8.7 weight percent C after the tests. From experience, methanol breakthrough normally occurs when the coke content reaches 25-30 weight percent C, i.e., there was room for longer operation. Methanol breakthrough was actually measured late in Test #3, but this “breakthrough” was related to the low reactor temperatures applied to compensate for the lower feed rate, i.e., the reactor temperatures were decreased to limit the cracking rates, and thereby increase the gasoline yield. Analysis of the catalyst revealed no abnormalities, so all in all the catalyst performance that we observed matched our expectations quite well, and there is therefore no need to change the design of the gasoline unit in future commercial units.

9. Woody Biomass Gasoline EPA Registration Testing

The original project award for demonstrating the woody biomass to gasoline technology did not include work for registering the resulting renewable gasoline. However, before a transportation fuel can be introduced into commerce it must first be registered with the Environmental Protection Agency (EPA). In early 2010, DOE requested that the project scope be modified to include work that would show that the renewable gasoline could be registered. As a result, Phillips 66 agreed to develop a registration plan that would include the Tier 1 engine emissions testing results and the toxicology literature search results. It is anticipated that Tier 2 will not be required because the renewable gasoline is substantially similar to existing conventional gasoline; however, this determination rests with EPA.

An outside lab, Southwest Research Institute (SwRI), was hired to perform the engine emissions testing. SwRI® has extensive experience in performing the analysis. All test results are reported by SwRI® in a confidential report made directly to Phillips 66, and is included in Appendix F.

A registration request submission was also written by Phillips 66 summarizing the engine test results and data supporting the Phillips 66 claim that the woody biomass gasoline is substantially similar to known gasoline fuel within the United States. Included in that application is a toxicology literature search, which was done by a Phillips 66 toxicologist (Appendix H).

Credentials for SwRI are quoted directly from them below:

SWRI® QUALITY CONTROL AND QUALITY ASSURANCE

To demonstrate SwRI's constant goal to provide quality emissions data in our project efforts, the Engine, Emissions, and Vehicle Research Division (EEVRD) maintains certification to International Organization for Standardization (ISO) 9001:2000 and accreditation to ISO/IEC 17025:2005 standards. Standard operating procedures and routine instrument calibration and calibration records are included in these standards. Based on the successful completion of third party audits, the EEVRD is able to maintain registration under ISO 9001:2000, "Quality Management System," and accreditation by ISO/IEC 17025:2005, General Requirements for the Competence of Testing and Calibration Laboratories." The SwRI Office of Automotive Engineering (OAE) Quality Policy Statement:

SwRI OAE Quality Policy Statement

> *"The Office of Automotive Engineering provides unequaled capabilities for the research, development, evaluation, and qualification of transportation systems, vehicles, vehicles, fuels, lubricants, and emissions-related products. Quality excellence is the foundation for the management of our business and the keystone to customer satisfaction. It is our objective to ensure that our final products are internationally recognized with unquestioned quality and are delivered to our clients in a professional, cost effective, and timely manner.*
>
> *We are committed to comply with ISO 17025, ISO 9001:2000, and all customer-required standards of excellence. Continual improvement of this policy occurs through regular review of the quality system's suitability to meet customer, employee, and supplier needs."*

> *Throughout this project, SwRI implemented our QA/QC plan in a manner consistent with the program objectives, including spot-checking of records, accuracy/precision charts, notebooks, calibration tags, and other quality control elements including chain of custody of samples. Listed below are a few of the key process that ensure the quality standards are implemented.*

This work was performed to address EPA requirements for registration of designated fuel and fuel additive (F/FA) as stipulated by sections 211(b) and 211(e) of the Clean Air Act (CAA). In general, standard mandatory requirements for F/FA registrations are contained in a three tiered structure. Definitions of baseline, non-baseline, and “atypical” F/FA are discussed in detail below. Other special provisions include experimental F/FA, relabeled products, and products exclusively for off-road use.

Each F/FA is sorted into one of two broad "fuel families": conventional or alternative. The conventional fuel families are diesel and gasoline, and the alternative fuel families include methanol, ethanol, methane, and propane. Each fuel family is then subdivided into three "F/FA categories": baseline, non-baseline, and “atypical.” Candidate fuel Woody Biomass Gasoline (WBG) is a non-baseline gasoline fuel because it comes from a non-conventional renewable source.

Regulated exhaust emissions for total hydrocarbons (THC), non-methane hydrocarbons (NMHC), carbon monoxide (CO), oxides of nitrogen (NO_x), total particulate matter (PM) and carbon dioxide (CO_2) were evaluated for all tests conducted. In addition, hydrocarbon speciation was performed to determine volatile-phase exhaust hydrocarbons, aldehydes, and ketones. Samples were also collected for volatile- and particulate-phase polycyclic aromatic hydrocarbons (PAH) and nitrated polycyclic aromatic hydrocarbons (NPAH).

In addition to the combustion emissions, Tier 1 requires testing of the evaporative emissions for the fuels and fuel additives which have a Reid Vapor Pressure (RVP), pursuant to Code of Federal Regulations (CFR) Title 40, part 80, Appendix “E” of 2.0 pounds per square inch or greater. A method of determining the evaporative emissions from gasoline was developed by Southwest Research Institute® (SwRI®) and approved by EPA. This method uses an Evaporative Emission Generator (EEG) to determine the head space composition of fuels rather than using a method that employs an evaporative emissions enclosure and an entire vehicle. The fuels were introduced to the EEG, and hydrocarbon speciation was performed on the head space composition to determine volatile-phase hydrocarbons.

This analysis includes the combustion emission measurements and evaporative emission measurements that were conducted as part of the requirements for the registration of an additive or fuel as stipulated by sections 211(b) and 211(e) of the CAA.

Combustion emissions for evaluating light-duty vehicle exhaust emissions were generated according to the exhaust emission portions of the Federal Test Procedure (FTP) for certification of new motor vehicles found in 40 CFR part 86, Subpart B. The Urban Dynamometer Driving Schedule (UDDS), pursuant to 40 CFR part 86, Appendix I (a) was applied for the testing. The UDDS is the result of more than ten years of effort by various groups to translate the Los Angeles smog-producing driving conditions to dynamometer operations, and is a non-repetitive driving cycle covering 7.5 miles in 1372 seconds with an average speed of 19.7 mph. The maximum speed is 56.7 mph.

9.1 Summary of a Single UDDS

Segment	Duration, sec.	Distance, miles	Average Speed, mph

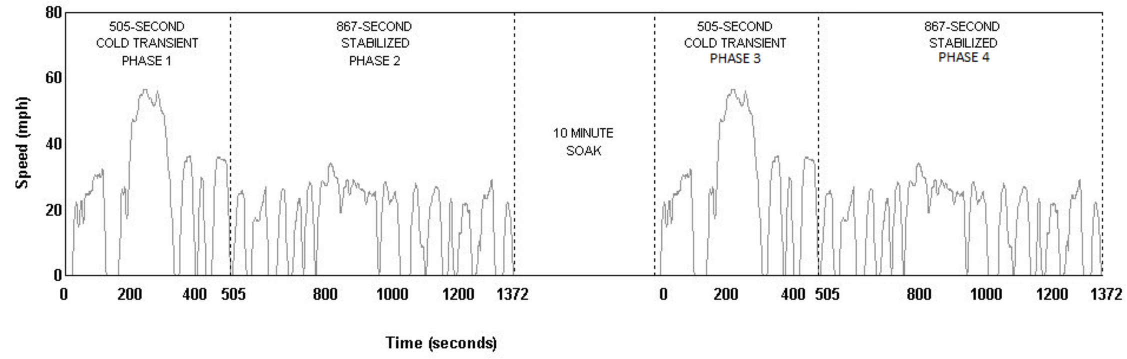

Figure 9-1 4-bag FTP driving schedule

Figure 9-2 Vehicle during FTP driving cycle (front view)

Figure 9-3 Vehicle during FTP driving cycle (side view)

Emission testing for characterization of combustion emissions was conducted both with the after-treatment device intact and with the after-treatment device rendered non-functional (as per EPA requirements for Tier 1 testing). After vehicle selection was made, the entire vehicle after treatment system was removed and configured with a blank spool piece (hollow tube) in accordance with EPA 211(b) methodology for testing the vehicle without a catalyst. The purpose of this step was to simulate a "worst case" condition, where the catalyst has melted and has been blown out of the exhaust system.

9.2 Test Plan for Light-Duty Vehicle Testing

STEP	DESCRIPTION
1	When received at SwRI, the vehicle was checked for an intact after treatment device, exhaust leaks, acceptable tires, proper oil level, proper transmission fluid level, and proper vehicle operation on the mileage accumulation dynamometer (MAD). The vehicle fuel was drained and refueled with the base fuel. Mileage accumulation for 4000 miles was performed using a MAD.
2	For emission testing after the first 4000 miles, the vehicle's exhaust system was prepared for connection to the Constant Volume Sampler (CVS); all necessary calibrations of the testing equipment were performed.
3	The vehicle was run over two UDDS sequences to prepare it for testing the following day.
4	The vehicle was soaked overnight (12 to 36 hours).
5	A 4-bag FTP cycle was run. Gaseous and particulate emissions were collected and analyzed for the regulated and unregulated emissions including hydrocarbon speciation, aldehydes, ketones, PAH, and NPAH.
6	Steps 3 through 5 were repeated two additional times on different days.
7	The catalytic converter was removed and replaced with a blank spool piece and Steps 3 through 6 were repeated.
8	The catalytic converter was re-installed. The base fuel was drained and refueled with the WBG fuel. Mileage accumulation for 4000 miles was performed using a MAD. (**Note:** No other fuel was used until mileage accumulations and entire emission test sequences with WBG were completed.)
9	The vehicle was prepared with two UDDS sequences, and Steps 4 through 7 were repeated.
10	All samples collected were analyzed and final report was prepared.

9.3 Vehicle Selection per EPA Protocol

For the purpose of testing, a vehicle was selected and purchased. The vehicle was selected to meet the following criteria:

- Less than 500 miles of prior operation (actual mileage was 39 miles).
- Same type, class, and subclass which consumed the most gallons of fuel in the fuel family over the past three years.
- Represent the most common fuel metering system and the most common of the most important emission control system devices or characteristics with respect to the emission reduction performance for the model year in which testing began.
- One of the five highest selling models from the current model year.
- Unaltered from the specification of the original equipment manufacturer and to remain under the control of SwRI throughout the testing.

9.4 Woody Biomass Gasoline Blendstock Properties

Additional analyses for the candidate fuel were performed to confirm if this fuel complies with EPA definition as “substantially similar” to unleaded gasoline. Sections 211(f)(1)(A) and (B) of the Clean Air Act prohibit the introduction into commerce, or increases in the concentration in use of fuels and fuel additives that are not “substantially similar” to fuels or fuel additives used in the certification of 1975 or subsequent model-year vehicles. Section 211(f)(1)(A) established this prohibition effective March 31, 1977, for fuels used in light-duty motor vehicles; Section 221(f)(1)(B) established this prohibition for motor vehicles effective November 15, 1990.

At present, EPA defines fuels and fuel additives that meet criteria should include the following as “substantially similar” with respect to Section 211(f)(1)(A) of the CAA:

1. The fuel must contain carbon, hydrogen, and oxygen, nitrogen, and/or sulfur, exclusively, in the form of some combination of the following:

   a. hydrocarbons;
   b. aliphatic ethers;
   c. aliphatic alcohols other than methanol;
   d. (i) up to 0.3 percent methanol by volume;
      (ii) up to 2.75 percent methanol by volume with an equal volume of butanol, or higher molecular weight alcohol.

2. The fuel must contain no more than 2.0 percent oxygen by weight, except fuels containing aliphatic ethers and/or alcohols (excluding methanol) must contain no more than 2.7 percent oxygen by weight.

3. The fuel must possess, at the time of manufacture, all of the physical and chemical characteristics of an unleaded gasoline as specified in American Society for Testing Materials (ASTM) Standard D4814-88 for at least one of the Seasonal and Geographical Volatility Classes specified in the standard.

4. The fuel additive must contain only carbon, hydrogen, and any one or more of the following elements: oxygen, nitrogen, and/or sulfur.

9.5 Evaporative Emission Composition Testing

An EEG is a fuel tank or vessel which is heated to cause the volatile portion of the fuel or fuel additive to evaporate at a desired rate. The SwRI EEG vessel was designed according to the requirements of the Code of Federal Regulation (CFR) Title 40 – Protection of Environment, Part 79 – Registration of Fuels and Fuel Additives, Subpart F – Testing Requirements for Registration, Section 79.57 – Emission Generation. For these experiments, the EEG was a stainless steel cylinder with a flange on the top for the introduction of the fuel sample. A bleed valve, closed-tip thermocouple which extends to the liquid volume, pressure gauge, and septum-type sampling port were mounted on the top flange, and this top flange is bolted onto the main portion of the vessel with six bolts. A Teflon® ring was used to provide a seal between the bottom and top of the vessel. The assembled vessel was then wrapped with a custom-made thermal blanket which included two parts: one for covering a cylindrical vessel bottom and the other for covering the top. Both parts of the thermal blanket were connected to a temperature controller to maintain the desired temperature.

9.6 Summary

Tests were performed on a light-duty vehicle to provide data to address EPA requirements for registration of a designated F/FA as stipulated by section 211(b) and 211(e) of the CAA. A vehicle was tested according to procedures established in Title 40 CFR Part 79, Subpart F, Section 79.57 and Title 40 CFR Part 86, Subpart D. Emissions characterization was performed with two fuels: with the base gasoline and with WBG. All test results are reported to Phillips 66 in a confidential report. This report contains data and analysis to support the Phillips 66 claim that the woody biomass gasoline is substantially similar to known gasoline fuel within the United States.

10. Fleet Test Procedure and Test Results

This chapter reproduces the results of the report “Comparison of Candidate and Commercial Fuel in an Eight Vehicle Fleet” conducted by Transportation Research Center Inc. Please note that the Transportation Research Center Inc. does not endorse or certify products of manufacturers. The manufacturer’s name appears solely to identify the test article. Transportation Research Center Inc. assumes no liability for the report or use thereof. It is responsible for the facts and the accuracy of the data presented herein. This report does not constitute a standard, specification or regulation.

10.1 Program Design Summary

Transportation Research Center Inc. (TRC Inc.) performed vehicle testing for Phillips 66 to compare the effects of a candidate fuel and a commercial gasoline fuel on the vehicle, engine components, and exhaust after-treatment systems. The test utilized four pairs of vehicles. In each pair, one vehicle ran on commercial gasoline and the other ran on the candidate 50% bio-gasoline blend. Each vehicle accumulated 75,000 miles of on-road vehicle operation. A summary of the steps completed during the program are listed below:

1) Vehicle selection by Phillips 66
2) TRC Inc. vehicle procurement.
3) TRC Inc. baseline engine visual inspection.
4) Vehicle break-in to 4,000 odometer miles using the Standard Road Cycle (SRC) and specified candidate or commercial fuel.
5) Baseline Emissions and Fuel Economy Testing using FTP/HWFETx3 test cycles. All vehicles are tested using EPA Federal Certification Test Fuel.
6) Mileage Accumulation to 75,000 odometer miles using the Standard Road Cycle (SRC) and specified candidate or commercial fuel. Scheduled maintenance performed following manufacturer’s schedule.
7) Final Emissions and Fuel Economy Testing using FTP/HWFETx3 test cycles. All vehicles are tested using EPA Federal Certification Test Fuel.
8) Final engine visual inspection.
9) TRC Inc. engine disassembly inspection and measurement.

10.2 Test Vehicle Fleet Specifications and Procurement

TRC Inc. was provided a list of four vehicle manufacturer and model configurations by Phillips 66. TRC Inc. purchased eight vehicles total, two new vehicles of each manufacturer and model combination. All vehicles were purchased new from manufacturer lots in the Central Ohio area and care was taken to ensure the vehicles were identically equipped within each pair. The details for each vehicle are included in Table 10-1. All vehicles were delivered to TRC and received initial inspections to verify they were damage free.

Table 10-1 Vehicle Fleet Specifications

Model Year	# Cylinders	Engine	Equipment Requested
2014	4	1.8 L	Gasoline Port Fuel Injection
2014	4	1.5 L	Port Fuel Injection
2014	4	1.5 L	Direct Injection Turbocharged
2014	6	3.5 L	Direct Injection Turbocharged

Upon receipt at TRC Inc. each vehicle was inspected, photographed, and assigned a unique test ID number to be used for the duration of the program and data reporting. In each vehicle pair one vehicle was randomly assigned to operate on commercial fuel and one on the candidate fuel.

10.3 Baseline Engine Inspection

Prior to starting any testing activities or mileage accumulation, each vehicle engine was inspected and photographed using an Olympus IPLEX video scope with light source. The video scope is a non-invasive photography tool which uses a 6mm lens tip on an articulated flexible fiber optic cable to take engine component photos without fully disassembling the engine.

The inspection was completed following the Coordinating Research Council (CRC) Intake Valve Deposit (IVD) rating method. Engine disassembly was limited to the removal of selected external engine components to allow fuel injector removal. The engine was not disassembled past this point. The IVD was rated on all intake valves by a CRC qualified rater.

TRC Inc.'s goal was to document the initial engine condition without introducing any possible changes due to disassembly prior to program start. At the completion of the program each engine was inspected a second time using the video scope to record the same images as pre-program for a direct comparison of engine component wear.

Photos were taken to be documented in the confidential report. Video scope photographs included valves, piston tops, cylinder walls, and intake runners when appropriate. Visual deposit ratings were completed following CRC methods for the exhaust valves, intake ports, head combustion chamber, intake runners, and piston tops.

10.4 Track Mileage Accumulation Cycle

The eight-vehicle fleet was driven on TRC Inc.'s 7.5-mile test track to complete mileage accumulation. TRC Inc. used the EPA Standard Road Cycle (SRC) for this mileage accumulation.

TRC Inc. maintained the eight-vehicle test fleet to the specifications stated by each manufacturer in the owner's manual for normal duty operation. This maintenance included changing oil, oil filters, and air filters. Consumable items such as tires and headlights were changed as needed.

10.5 Mileage Accumulation Fuel

One vehicle from each model pair was randomly selected to operate on either the candidate fuel or commercial gasoline. The candidate fuel was provided by Phillips 66 while the commercial gasoline for the program was drawn from TRC Inc.'s on-site fuel supply.

The candidate fuel was delivered to TRC Inc. in two bulk deliveries directly from Phillips 66's supplier. Upon delivery a one-gallon sample from each delivery was collected and the fuel was transferred to TRC storage/dispensing tanks which had been previously emptied and steam cleaned. Following delivery approximately 30 gallons of fuel was dispensed to flush the pump and lines and the pump fuel filter was replaced with a new filter.

TRC Inc. maintains a constant supply of unleaded gasoline for track mileage accumulation programs. TRC Inc.'s fuel for this program was regular unleaded, additized to a Top-Tier fuel.

10.6 Compression and Leak-Down Measurements

At five mileage points each vehicle's compression and leak-down were measured for comparison. Measurements were recorded at approximately 4,000, 20,000, 40,000, 60,000 and 75,000 odometer mileages.

10.7 Emissions and Fuel Economy Testing

TRC Inc. completed a chassis dynamometer-based Triple FTP/Triple HWFET Exhaust Emissions Test at 4,000 and again at 75,000 odometer miles for each test vehicle. TRC Inc. set the first test point at 4,000 miles to allow the catalyst to be adequately broken in and stabilized. TRC Inc. completed all emissions testing using federal certification test fuel.

Following the fuel change procedure the vehicle performed two Highway Fuel Economy Driving Schedules (HWFET) and completed the Society of Automotive Engineers (SAE) J2264 Chassis Dynamometer Simulation of Road Load Using Coast down Techniques procedure. The SAE J2264 procedure is used to determine the chassis dynamometer loading coefficients which will produce the correct road load to simulate vehicle operation on a road. The SAE J2264 procedure was completed for each vehicle prior to the 4,000 and 75,000 miles emissions test points.

At the completion of the SAE J2264 procedure the vehicle completed a single Urban Dynamometer Driving Schedule (UDDS) preconditioning. This procedure is followed by a soak period of at least 12 hours, but not exceeding 36 hours. All preconditioning procedures are performed at an ambient temperature range of 68 - 86°F. Upon completion of the soak, a single FTP Exhaust Emissions Test was performed on the vehicle immediately followed by a Highway Fuel Economy Test (HWFET). Following the completion of the FTP/HWFET the vehicle immediately started a second soak period of at least 12 hours. Upon completion of the second soak, a second FTP/HWFET Test was performed on the vehicle. Following the completion of the second test the vehicle started a third soak period of at least 12 hours. The vehicle was not preconditioned prior to the second and third FTP/HWFET Tests unless the soak period exceeded 72 hours. At the completion of the third soak period the final FTP/HWFET Tests was performed.

Gaseous emissions were measured using a CFV-CVS sampling technique and an AVL Analytical Sampling Train. All tests were completed and processed using actual Federal Certification Fuel properties. Both the chassis dynamometer and the exhaust gas analytical sampling train received weekly quality control inspections prior to the inception of each testing. Routine monthly verifications are also completed on this equipment.

The emissions for all phases were collected on a dilute bag basis for analysis. The results are provided in grams per mile (g/mi) for total hydrocarbons (THC), non-methane hydrocarbons (NMHC), carbon monoxide (CO), carbon dioxide (CO_2) and oxides of nitrogen (NOx). Fuel Economy results were calculated for each test phase.

At the completion of mileage accumulation and the 75,000 emissions test points, each vehicle engine was inspected and photographed a second time using an Olympus IPLEX video scope with light source. This inspection was identical in procedure to the initial inspection completed at vehicle delivery as described previously.

The photos are organized for a direct comparison of the pre- and post-test condition of each part between the candidate and commercial fuel vehicles. Visual deposit ratings were completed following CRC methods for the exhaust valves, intake ports, head combustion chamber, intake runners, and piston tops.

10.8 Post Mileage Engine Measurement

At the completion of the post mileage engine inspection each vehicle was disassembled and the following measurements were recorded: valve stems (intake and exhaust; top and bottom), valve seat width (intake and exhaust), cam lobes (intake and exhaust), cam journals (intake and exhaust), and cylinder bore (top, middle, bottom).

Additional bench photography was taken of combustion chambers, intake ports, intake runners, intake valves, and piston tops.

10.9 Valve Weights

At the completion of post-test measurements the exhaust valves were removed and weighed. The valves were then buffed to remove all deposits and weighed a second time to determine total deposit accumulation.

10.10 Equipment Dispersal

At the completion of the program all test vehicles, parts, and candidate fuel remaining were dispersed per Phillips 66 instructions. No physical property was retained by TRC Inc. at the completion of the program.

10.11 Candidate Fuel Disposition

The remaining Candidate Fuel received by Phillips 66 was consumed internally by TRC at their test facility. No fuel was sold/released/consumed outside of the facility.

10.12 Program Observations

TRC Inc. reviewed the emissions results, measurements, photographs, and vehicle operation reports of each vehicle to identify any notable differences between the matched vehicle pairs operating on the candidate and commercial fuels. The observations are based on only the information collected during this program. Mathematical data were not analyzed for statistical significance due to the limited sample size.

The observations were made to identify any operational or mechanical changes which would be noticed by the typical consumer when operating the program vehicles for an extended period of time (75,000 odometer miles) on the candidate fuel as opposed to commercial fuel operation.

Data reviewed included:

- Pre- and Post-Program Video Scope Photography
- Post-Program Engine Disassembly Photography
- Intake Valve Weights and CRC Ratings
- Engine Component CRC Ratings
- Post-Program Engine Measurements
- Observed Calculated Fuel Economy
- Pre- and Post-Program FTP and Highway Emissions Test Results
- Lone Term Fuel Trim Data
- Compression and Leak Down Measurements
- Vehicle Maintenance
- Vehicle Oil Consumption

All results from this fleet test have been reported to Phillips 66 in a comprehensive, confidential report.

10.13 Overall Fleet Observations

There were only minor insignificant differences in the candidate versus commercial fuel fleet vehicle operation over the course of the program. There were no differences observed which would have had any impact on the consumer in terms of vehicle operation or perception.

11. Lessons Learned From Project Execution

The purpose of this chapter is to review the experiences gained from the execution of the project. The technical lessons learned are described elsewhere in this report.

From an overall perspective it must be concluded that the project was well managed and that to a large extent it met the targets set forth initially. However, there is of course always room for improvement and some of the observations made in this chapter may be of interest when executing future projects of this type.

11.1 Budget and Schedule

The start-up activities of the project included some tasks that were not originally foreseen to take place such as the extensive Independent Project Review. The project team had to provide extensive information in form of detailed questionnaires, etc. and participate in workshops conducted by an outside consulting firm. The impact on project cost and schedule had to be incorporated within the original project schedule and budget without expanding these. Better understanding of DOE requirements and communication with DOE prior to finalizing budget and schedule would have minimized this issue.

An unexpected delay of DOE approval to move into Budget Period 2 due to political discussions in Congress caused the work on the project to slow down substantially for several weeks. After restart we did, however, manage to complete the project within schedule. The delay was of course completely out of DOE's and our control.

All the main activities of the project stayed reasonably within the original budget except for the higher cost of commissioning. However, this was compensated by the savings of project management and reporting cost, and by reducing the test durations in Tests #2 and #3.

It was also possible within the budget to allocate, at the recommendations of DOE, the Phillips 66 funds to a new activity including the engine emission test at Southwest Research Institute in San Antonio, TX, as well as the fleet testing of a renewable gasoline blend. It was very beneficial for the project that DOE was flexible enough to modify the original scope to include such new tasks that enhanced the value of the project.

11.2 Engineering Design

The Integrated Bio-Refinery (IBR) consists of several technologies provided by different companies of the project team. The integration of these technologies requires several calculation iterations which are time consuming. A well-defined procedure for this activity is required for achieving optimal execution of this task.

Active participation by the Independent Engineer in the engineering phase assured that the final design would meet the expectations of DOE.

One component of the project that we encourage for future projects is to conduct a preliminary HAZOP study well before final engineering design and bid solicitation. Although this HAZOP was not comprehensive because final P&IDs were not yet developed, there were many good ideas generated during those sessions which contributed to safe and successful operations later in the process, and also helped the final HAZOP studies to proceed more smoothly.

11.3 Procurement and Installation

Clear definition of responsibilities and the use of conventional procurement activities including preparation of detailed bid documents, conducting open bidders' meetings, etc., as well as evaluation and selection of suppliers resulted in very few change orders and execution of this phase within budget.

Assignment of a dedicated Construction Manager (CM) by GTI was critical to maintaining subcontractor discipline and avoiding cost overruns. The CM enforced strict management-of-change rules, whereby any subcontractor scope change that was not already captured within a current task description required the subcontractor to submit a form detailing the task description, cost estimate, cost estimate breakdown (labor hours, supplies, rentals, etc.), and schedule. By spelling this out prior to subcontract execution and making it routine, the inevitable changes required for successful installation and commissioning were manageable and their cost impacts were fully captured in budget updates. A weekly construction meeting with all active subcontractors present was also a key component of this process.

11.4 Commissioning

Commissioning of an installation comprising new, untested technologies, broadly defined to include hot shakedown testing, typically presents the greatest potential for unforeseeable challenges. Although the project work plan included a very detailed, staged commissioning of the integrated pilot unit, this task incurred a significant budget overrun.

This overrun was caused by the discovery and necessity to correct operational challenges in the course of shakedown testing, which required full 24/7 staffing and facility support. Examples included:

- Hot gas filter (HGF) candle degradation – metallic HGF candles, which had performed adequately in prior coal gasification tests, experienced degradation under exposure to the biomass-derived product gas. This degradation caused a premature shutdown and necessitated an extra dismantling and cleanup of the downstream tar reformer.
- Syngas compressor startup and stable operation during transitions – the syngas compressor is equipped with a large number of sensors and safety-related trip points. There is a significant learning curve to operate within the equipment boundary conditions during startup and during transitions such as switching the gas discharge from flare to AGR, accommodating changes in feed rate and upstream pressure, transitioning from internal gas recycle to full discharge, managing output pressure changes, and operating at changing ambient conditions. These learnings also had to be communicated to a number of operating personnel.
- AGR operation – the AGR system, originally designed for purification of natural gas, was refurbished and upgraded for syngas service and relocated from Texas to GTI's Des Plaines facility in 2007, but was never fully commissioned until this project due to lack of funding. A variety of design shortcomings were identified during the second and third shakedown tests (S-2 and S-3), including erratic performance of the AGR solvent pump, malfunctioning instrumentation, and leaky pipe connections.
- Purchased support equipment commissioning – some of the critical packaged equipment items added for this project required significantly more troubleshooting than allowed for in the budget. One example was the inability of the purchased chiller to reach the design temperature of 20°F, which was tracked down to a software issue and corrected during turnaround. Another was a rental air compressor for startup, which was improperly wired. Manufacturers and vendors were typically responsive in fixing problems that arose with their equipment, but costs were still incurred for the

project due to testing interruptions, downtime, and project personnel participation in the troubleshooting activities.

While these challenges added to the cost of commissioning, the extensive troubleshooting reduced interruptions to the subsequent pilot plant testing which would most likely otherwise have caused even more substantial overruns and schedule slips. We recommend that the budget and schedule for the commissioning task be structured to accommodate contingency for unforeseeable costs and delays in future similar demonstration projects. The current budgeted DOE budgeting procedure requiring firm quotes for each equipment item and installation task makes it challenging to anticipate these costs. Because many of the unit operations comprising the biorefinery are either uniquely designed or sized (*e.g.*, for pilot plant scale), or used in a new type of service (*e.g.*, for biomass-derived gas), traditional quoting methods often undershoot the actual time and manpower requirements to get the system actually up and running properly and safely.

11.5 Testing

The testing portion of this project generated knowledge and experience essential to de-risking of the technology for commercial scale-up, such as:

- Integrated operation of the entire biorefinery for extended periods.
- Validation of biorefinery integrated control strategy.
- Confirmation of gasification stability and resistance to upsets in other parts of the system.
- Confirmation that the gas cleanup steps (filter, tar reformer, direct cooler, and AGR) can deliver syngas of acceptable quality and purity for the TIGAS plant.
- Confirmation that Morphysorb AGR solvent can deliver predicted CO_2 and H_2S removal with very low losses of CO and H_2, and acquisition of process correlations.
- Confirmation that TIGAS tail gas can be recycled to the gasifier without causing any gasification upsets.
- Achievement of biomass-to-gasoline conversion targets (gasoline yield).
- Validation of catalyst performance and confirmation of catalyst space velocity decisions to optimize product yield and quality.
- Validation of predicted gasoline product quality suitable as a drop-in motor fuel blendstock.

The testing campaign also resulted in milestones which were not anticipated in the original proposal:

- Testing and validation of an ultra-hot gas filter (UHGF) technology that can increase energy efficiency and/or simplify the commercial process.
- Production of enough gasoline blendstock for a moderate fleet test to ensure compatibility with commercially available automobile engines.

Based on these achievements, the key lesson learned is that a pilot test campaign with sufficient data acquisition and interpretation can substantially reduce the risk associated with deployment of a commercial biorefinery to produce gasoline blendstock from woody biomass. This is confirmed by the fact that all of the technology licensors involved in this project have stated that the pilot results are sufficient to support commercial offering of their technologies.

For future projects with similar testing objectives, additional lessons were learned which could help maximize value from available funds.

- As expected, the measurements, analysis and treatment of the operating data from the operation of the IBR demonstration plant was a very complex but important task. It is recommended that the procedures, data collection schedule, etc. are discussed in detail and clearly defined at an early stage of the project to ensure that it matches the requirements and expectations of all project team members. In this project, sampling and analyses to adequately characterize performance of individual system components required more manpower than in the original budget.
- In the Gasification Section, the upgraded metallic hot gas filter candles mentioned in the previous section on Commissioning failed during Test #1, requiring unplanned post-test dismantling, removal, inspection, and analysis. Consultation with experts on metallurgy in biomass gasification might have helped to anticipate this obstacle and expedite the change to a ceramic-based candle, which turned out to be the eventual solution. However, such consultations would have to be included in the original project budget.
- The recycle compressor in the TIGAS unit created significant costs due to lengthy startup troubleshooting. The recycle compressor was procured by the fabricator of the TIGAS skid-mounted unit, and the support from the manufacturer of the compressor was unsatisfactory. A more detailed review of the critical equipment procured and supplied by others is recommended for future projects.
- Flange leaks from major high-pressure vessels and heat exchangers were a source of significant downtime and expense. Pilot testing requires far more extensive use of flanged connections than a commercial installation, because of the need to open vessels for inspection and modifications, and also to accommodate extensive instrumentation. We found that flange tightening procedures recommended by vendors and fabricators were not sufficient to avoid leaks at process temperatures and pressures. Projects that employ this type of equipment should include provisions for hot torqueing of flanges with minimal disruption of testing operations. In this project, we eventually employed bolt tensioners (Superbolts) that could be re-torqued with hand tools under hot conditions when needed. This solution incurs substantial additional capital expense for equipment, but greatly reduces downtime and associated operating expenses.
- In general, the operating expenses during testing were higher than originally budgeted, which resulted in fewer operating hours in order to stay within budget. For reasons alluded to above, these operating expenses are difficult to estimate accurately without a contingency element that is typically included in industrial R&D, so it is recommended to find ways to estimate these costs more conservatively in future similar projects.

12. Commercialization Plan

The strategy for the successful introduction and ongoing sales of waste wood to gasoline technologies encompasses major activities to achieve a long term goal of creating two billion gallons per year capacity. The team members, who will be offering the commercialized technology, are confident that the pilot scale plant can be scaled up to 4,000 short tons per day (inlet the gasification unit) to meet the needs of the market. It is anticipated that the project participants, UPM, Topsoe, Andritz, and GTI, will play active roles in the commercial venture. Topsoe and Andritz will license the TIGAS and gasification technologies and provide the process performance guarantees for the commercial plant. An EPC contractor will complete the design of the plant including the necessary off-sites and complete the procurement, construction and commissioning of the plant as well as assure the environmental and other regulatory approvals needed to build and operate a full scale plant. Phillips 66 intends to apply for certification for the use of the green gasoline as a transportation fuel. UPM has expressed interest in being the first adopter of this new process.

The team will prepare press releases and other marketing communication materials to announce the availability of the new technology. Results will be published in papers at selected gasification, biomass and transportation fuel conferences. Other marketing information for this technology will be placed on team members' websites. Sales strategy will focus in the south (Louisiana-East Texas), northwest (Washington-Oregon), north central (Minnesota-Wisconsin), and northeast (Maine) with abundant waste wood resources.

Potential sites for the first commercial plant will be identified. UPM's locations are primary candidates for the first site. The leading candidate is UPM's Blandin Paper Mill in Grand Rapids, Minnesota. Available feedstock within a 50 mile radius has already been characterized and can be acquired at an average price of about $62 per dry short ton.

New product development and continuous improvement are a necessity to maintain competitiveness of the technologies during the growth phase. This is mandatory for all members of the commercialization team whose core businesses are technology development and licensing and equipment manufacturing. For example, Topsoe invests a significant fraction of its revenues in research to maintain the competitiveness of TIGAS and all its licensed technologies.

The target will be to sell and build 3 to 4 plants per year to meet two billion gallons per year target. Once the plant construction business reaches maturity in the late 2020s, the options for technology suppliers to continue their growth include service business to maintain and improve plant performance and marketing and sales of the next generation of waste wood to gasoline technology.

Critical success factors include:

- Economics of potential commercial plant meet or exceed return on investment criteria.
- Fuel meets motor vehicle specifications.
- Regulatory and licensing approvals obtained.
- First commercial sale completed.

13. Commercial Plant Design

The commercial plant comprises a wood yard to receive and handle the feedstock, a woody biomass drying unit followed by a gasification unit, an AGR unit, the synthesis unit, and the gasoline upgrading unit. Each of these units is described in the following including a summary of the production and consumption figures and the total installed cost. Reference is made to the attached block flow diagram (BFD).

13.1 Wood Yard and Wood Biomass Drying Unit

The capacity of the wood yard and biomass drying unit is 6400 short tons per day of 50% wet woody biomass feed. The biomass feed - in the form of chips of approximate size of 2-inch x -1 inch x 0.5-inch – is delivered to the plant site by trucks.

The wood yard consists of a biomass feed truck unloading station, several conveyers to transport feed from receiving to storage area, from storage area to fine screening and silos and then to transport prepared feed to dryers, open air storages (donut shape) with pile marking and discharge equipment, fine screening of chips and oversize crushing, and two wet feed storage silos for the prepared wood chips.

Woody biomass drying is handled by six (6) parallel closed loop type belt dryers. Sized and contaminant free woody biomass is conveyed from the wet biomass storage area with belt conveyors to the biomass distributing screws and conveyors which feed the dryers in an optimized manner to achieve high availability. The arrangement of screw conveyors allows flexible feed supply to the gasification units. The screws discharge into the woody biomass feeding bins of the dryers, from where the biomass is fed to the feed module of the dryers containing two parallel screw conveyors which distribute the biomass at the inlet section of the dryer belt. These screws across the dryer inlet ensure that biomass is distributed evenly on the dryer belt.

Drying of the woody biomass bed is done by hot air, heated by water coil heat exchangers placed along the dryer tunnel in the roof section. The hot water can be supplied from the heat recovery water system of the plant or according to the optimized entire plant integration from other sources. The heat consumption of the dryers is different in winter and summer operation, which has to be considered at the design of the dryer heating system and the overall plant heat recovery system. Ventilation of the dryer tunnel is controlled by a number of exhaust fans located outside the dryer. Ventilation air is ambient air. The drying air is sucked through the fuel layer and the belt then it is exhausted through silencers. Due to low air velocity and the screening effect of the fuel layer and the belt, the dust emission to atmosphere is low.

The dryness (moisture content) of the woody biomass discharged from the dryer is measured at the discharge chute. The dryness of the discharged biomass controls the drying air flow through the biomass bed travelling on the dryer belt, by changing the speed of the air fans. Dried woody biomass is discharged from the dryers' outlet by screw conveyor onto a collecting belt conveyor which collects the dried biomass from all 6 dryers. The dried biomass is transported with two (2) belt conveyors to two (2) dried woody biomass storage silos.

13.2 Gasification Unit

The gasification unit consists of four trains each having a capacity of 1000 short tons per day (20% wet) woody biomass input. The gasifiers are oxygen blown, pressurized system generating syngas from woody biomass. Each train of the gasification unit includes the following sub-processes:

- Solids feeding systems (pressurizing, feeding of fuel, bed material and bed additive)
- Gasification system including the gasifier with the cyclone
- Bottom ash discharge system
- Gas cooler (water injection)
- High temperature syngas filter system including filter ash discharge system
- Syngas tar reforming reactor
- Syngas cooler (superheating)
- Syngas scrubber system with cooling system
- Syngas final heater system
- Syngas tar adsorption
- Gasification unit specific auxiliary systems

The gasification process converts woody biomass to clean synthesis gas. The biomass is gasified in oxygen – steam atmosphere. The raw gas is slightly cooled and the fly ash is removed in a hot gas filter. The filtered gas is led to a catalytic tar reformer where hydrocarbons (methane and tars) are reformed to H_2 and CO. After reforming, the syngas is cooled in a gas cooler generating superheated steam. The syngas is cooled further in a water scrubber. The scrubber also removes the remaining contaminants and most of the ballast water vapor from the syngas. The syngas is slightly heated up after scrubbing in order to reduce relative humidity before compression.

13.2.1 Process Description

13.2.1.1 Solid Feeding Systems

Woody biomass from the drying unit is fed by one inclined belt conveyor to two gasification units from the dry woody biomass storage silo (i.e., two silos and two conveyors for four gasifiers). From the belt, conveyor biomass is distributed onto two chain conveyors on top of two gasification units through one mixing bin and two screw conveyors.

The biomass feeding system of one gasification unit includes three parallel lock-hopper based feeding lines. Woody biomass is fed into the intermediate storage silos at the top of the feeding lines. The storage silo serves also as buffer storage. The storage silo is purged with nitrogen to avoid dust explosion or self-ignition of the dried biomass. The storage silo is equipped with weighing systems. From the storage silo the biomass is moved through screw discharger, distribution screw and feed in valve into two lock-hoppers, where the biomass is pressurized from ambient to system pressure by using carbon dioxide. Each feeding line contains two lock hoppers which are alternately pressurized and depressurized. From the lock-hoppers the pressurized biomass is fed through the discharge valve and distributor screw to the surge hopper. The surge hopper is operated at the same pressure as the gasifier and allows continuous biomass feed to gasifier. The biomass is fed from the surge hopper through surge hopper discharger, metering screw, safety valve and the feeding screw into the gasifier. The safety valve separate the feeding line from the gasifier in case the gasifier pressure is higher than the surge hopper pressure (gas back flow) or when the feeding screw temperature exceeds the safety limit. In addition a positive flow of CO2 into the gasifier over the feed screw is maintained to prevent back flow. The biomass feeding screw is water cooled.

The bed material storages are located next to the gasification units. Two gasification units have common bed material storage silo. The bed material (dolomite) is supplied by trucks to the plant site and pneumatically conveyed into the atmospheric bed material storage silos. The bed additive (kaolin) is supplied also by trucks to the plant site and pneumatically conveyed into the atmospheric bed additive storage silos (one for two gasification units). From these storage silos the bed material and additive is dosed into mixing/conveying hoppers, through weigh hoppers and mixing screw. From the mixing/conveying hopper the mix is conveyed pneumatically to the bed material feeding system of the gasification units.

The bed material feeding system is a kind of lock hopper system including a bed material weigh hopper at the top from where the mix of bed material and bed additive is fed into the lock/surge-hopper for bed material pressurization to system pressure. The metering screw at the bottom of the surge hopper moves bed material to the fuel feeding screw of one (or two) of the biomass feeding lines for feeding it in the gasifier bed. The bed material feeding system is operated periodically. During lock hopper refill the metering screw stops and the bed material feeding system is isolated from the biomass feeding line with valves.

13.2.1.2 Gasifier

Woody biomass is gasified in the gasifier reactor by using oxygen and steam as gasification agent. The gasifier is an Andritz-Carbona pressurized bubbling fluidized bed type gasifier operated at 1560 ºF and 130 psig. The gasifier is a refractory lined pressure vessel. Gasification agents (oxygen & MP super-heated steam) are introduced through a valve system and the gas distributor grid to the bottom of the gasifier. The gasification oxygen is supplied from an external air separation unit (ASU) at a purity level of 99.5 mole percent.- Oxygen is distributed inside the gasification unit from a distribution header. Oxygen is preheated with MP steam to avoid condensation of gasification steam when mixed with oxygen. Oxygen preheating occurs before it enters the valve skid. MP steam is provided from outside the gasification unit. Recycle gas from the TIGAS process is also fed to the distributor grid of the gasifier. Recycle gas is preheated also with MP steam. The MP steam condensate is returned to external condensate recovery system. The biomass derived raw gas exits the gasifier at the top of the reactor. The entrained dust is partly separated from the hot gas in a cyclone and returned to the fluidized bed via the return pipe, called dipleg.

Bed material and fuel ash (together called gasifier bottom ash) are removed through the bottom of the gasifier by using water cooled cooling screw and a lock-hopper. The lock hopper is pressurized with carbon dioxide. The bottom ash is conveyed pneumatically through a conveyor hopper to the ash storage silo by using nitrogen. Two gasification units have a common bottom ash storage silo.

During startup the gasifier reactor is heated by the startup burner which is located in a refractory lined pressure vessel connected to the bottom part of the gasifier reactor. The startup burner is fired with natural gas. The steps of startup procedure are heat up with startup burner, combustion of biomass in air atmosphere and air blown gasification of biomass. For these steps air is supplied by the startup air system. The startup air system includes a screw compressor and air receiver tank. Startup air system is connected to gasification agent valve system of the gasifier.

13.2.1.3 Gas Cleanup

The gas cleanup and treatment system includes gas filtering, tar reforming, cooling and scrubbing.

The raw gas exiting the cyclone of the gasifier is cooled with water quench to the temperature of the hot gas filter unit. The water quench is a set of nozzles in the gas duct. The quench water is filtered scrubber condensate. During the startup and shutdown of the gasification unit, when no scrubber condensate is available, the gas is cooled by mixing steam or nitrogen to the main gas flow. The hot gas filter system is a candle filter unit. The filter elements are ceramic candles capable of operating at ultra-high temperature similar to those tested in the pilot plant, installed into a tube sheet. The filter candles are arranged in clusters. The pressure vessel of the filter is refractory lined. The filter candles are cleaned by carbon dioxide back pulsing. The filter unit is operated at system pressure, the pulsing gas (CO_2) is at ambient temperature and pressurized outside the gasification unit. The fly ash is removed from the filter bottom. In the fly ash removal and storage system the fly ash is cooled in water cooled ash screw, depressurized in a lock hopper system including a buffer hopper, a lock hopper and a conveyor hopper and transferred to the atmospheric fly ash silo. The lock hopper is pressurized with carbon dioxide while the fly ash transportation gas is nitrogen. Two gasification units have a common fly ash storage silo. The disposal cost of the fly ash is insignificant.

The filtered gas enters the tar reformer for the reformation of the hydrocarbons by using a catalyst system. The reformer reactor is a fixed bed reactor containing pelletized catalyst. The reformer temperature is increased from filter exit temperature to the required catalyst temperature by injecting oxygen and carbon dioxide mixture through burners at the inlet of the reactor stages. The CO_2 is recycled from the AGR. Both oxygen and carbon dioxide are heated by MP steam before entering the valve skid. The steam condensate of the preheating system is returned to the condensate recovery system. The tar reformer is a refractory lined pressure vessel accommodating the internals holding the catalyst bed(s) and providing even gas distribution. The reformer system is equipped with an external startup heater system. The startup heater system consists of a burner and a combustion chamber connected to a gas/gas heat exchanger where high purity nitrogen is heated up by burning natural gas. The combustor system has a separate air supply fan. Reformer heat up occurs simultaneously with gasifier heat up. During heat up the two flue-gas flows are mixed and directed to the flare. The flue-gas of the burner system is also exhausted through the flare stack.

The reformed gas of about 1650ºF is cooled in a water tube type gas cooler. The gas cooler generates HP super-heated steam which is utilized outside the gasification unit. The gas cooler includes evaporator section, superheater and steam drum. The hot gas enters the gas cooler from the top, through the superheater section and exits at the bottom. The superheater section is equipped with water injection for steam temperature control. The boiler feed water of the gas cooler is supplied from outside of the gasification unit. The preheated feed water is fed to the steam drum of the gas cooler. The gas cooler is equipped with soot blowing to remove depositions from the evaporator tubes if any. Water spray system is installed at the gas cooler gas exit to adjust the gas temperature in case of capacity decrease of the gas cooler. This quench will use scrubber water.

The cooled gas is further cooled in a two stage water scrubber to about 100ºF. The scrubber removes most of the water vapor and remaining contaminants from the gas and provides protection of the syngas compressor in case of reformer or filter malfunction. The scrubber has an inlet quench system where the water is pumped by cooling pumps through nozzles into syngas flow. The scrubber 1st stage provides low temperature heat (hot process water) which can be utilized for example in fuel drying. In the second stage of the scrubber gas is cooled also through a filler bed by recirculated water (pump and heat exchanger). If necessary, chemicals (sodium hydroxide or formic acid) can be dosed to the water circuits of the scrubber for pH adjustment or ammonia or chloride removal from the syngas. The separated scrubber condensate is partly used in the quench systems (upstream filter, downstream gas cooler). The remaining water is transferred to the waste water treatment plant.

The cooled, saturated syngas from the scrubber is heated up to 120ºF by heat exchange with hot water from the heat recovery water system (refer to scrubber heat recovery) in order to reduce gas relative humidity before the syngas is passed through activated carbon containing tar adsorbers, where the traces of tars are removed. The tar adsorber includes two vessels from which one is in operation while the other one is regenerated by steam. The clean gas is then sent to the Syngas Compressor

13.2.1.4 Auxiliary Systems

During start-up and shutdown of the gasification unit, in an emergency situation and when the downstream AGR or TIGAS process cannot accept the syngas, the flare system is used to burn gas safely and with low emissions. The flare is a natural draft systems utilizing natural gas as support fuel. The flare system contains burner system with pilot burner and ignition system, knockout drum, flare stack and gas valves. Gas can be flared at different temperature levels: from hot gas filter, before and after the scrubber and downstream the tar adsorber. In case of flaring for the hot gas filter the gas is cooled with water quench. During heat up the flare serves as a stack where the flue gas of the startup burners and the flue gas of the biomass combustion in the gasifier are released. All four gasification units are equipped with its own separate flare system allowing flaring up to 50% load of the gasification unit. This provides significant flexibility of plant operation.

The internal auxiliary systems in the gasification units are supplied from external sources. The gasification units utilize N_2 and CO_2 as inert gas. Nitrogen is produced in the ASU and used for purging, inerting , conveying and pressurization in all system, where this gas is not mixing into the syngas flow. Nitrogen is also used for emergency purposes and replacing unavailable carbon dioxide in case of malfunction of the carbon dioxide system. Carbon dioxide is recovered in the AGR and used for purging, inerting , back pulse cleaning and conveying in all systems where this gas is mixing into the syngas flow.

The gasification units have nitrogen distribution systems including one HP header supplying the HP consumers and one MP header from where the LP nitrogen line and LP header is branching for the low pressure consumers. Nitrogen is supplied externally from the ASU of the plant.

The gasification units have carbon dioxide distribution system including an HP carbon dioxide tank and a carbon dioxide buffer tank. The HP tank supplies continuous consumers through an HP header and periodic consumers. The buffer tank supplies consumers in the gasification units which have periodically larger carbon dioxide consumption (lock hoppers).

The gasification units are equipped with high pressure (HP) cooling water system. This is a closed loop cooling system of elevated water pressure to avoid cooling water boiling. This loop cools the bottom ash, filters ash cooling screws and the three fuel feeding screws. The high pressure water is circulated by two circulating pumps and cooled in a heat exchanger, which is connected to the heat recovery system.

The gasification units are equipped with high pressure sealing water systems. The sealing water system supplies water to the mechanical seals of the fuel feeding screw shafts. The system includes two pumps, cooling heat exchanger which is connected to the LP cooling water system.

The gasification units have heat recovery water systems providing hot process water. The heat is recovered from the scrubber 1st stage and from the HP cooling water system. The recovered heat is used in the gasification units for syngas heating and may be used for fuel dryer heating or other purposes externally of the gasification units.

The LP cooling water distribution systems of the gasification units provide cooling for the HP cooling and sealing water system (heat exchanger) for the start-up air compressor cooling and the 2nd stage of the syngas scrubber.

13.3 Acid Gas Removal Unit

Acid gas removal (AGR) is required to eliminate most of the sulfur gases (H_2S and COS) and CO_2 from the syngas prior to gasoline synthesis. Morphysorb®, a physical solvent developed jointly by GTI and Uhde GmbH (now Thyssen Krupp), is specified for this process. The physical solvent selectively absorbs acid gases from the “sour gas” at high pressure and low temperature, yielding a “sweet gas” for further processing and a “rich” solvent which can then be regenerated by depressurization to release the dissolved gases. Morphysorb® is a mixture of N-formyl and N-acetyl morpholine, selected for high selectivity, low vapor pressure, low cost, high stability, and low toxicity.

In the biorefinery, cleaned, dried, pressurized “sour” syngas is filtered and then routed to the Absorber column where it encounters pressurized, cooled “lean” solvent in countercurrent fashion. Pressure at the Absorber inlet is 1175 psig. Temperature of the cooled solvent is 55°F. The Absorber contains a packing material to maximize gas-liquid contact.

The “sweet” syngas exits the absorber, passes through a knockout drum to remove any solvent droplets, and continues to the TIGAS section of the plant.

The “rich” solvent is then depressurized and heated via heat exchange in three Flash stages operating at successively lower pressures. A portion of the CO_2-rich off-gas from the first flash stage is recycled to the Tar Reformer as a diluent, while the remaining off-gases are combined into an acid gas stream as described below.

Following the third flash stage, the solvent is further depressurized in the Stripper Column to about 10 psig and stripped with plant N_2 to maximize acid gas release. The solvent then proceeds to the Reboiler, where it is indirectly heated with steam to drive off any remaining acid gases. The stripped solvent temperature is 150°F. All of the acid gas streams from the Flash stages, Stripper, and Reboiler are then combined to form a single acid gas stream that exits the process for further treatment and disposal.

A portion of the acid gas stream is diverted to a separation step (membrane, PSA, or other) to separate high-purity CO_2 for recompression and use in the gasification plant. This CO_2 is used mainly for biomass feeding lock hopper pressurization, but also as a supplemental gas to the gasifier to provide a portion of the required fluidization. The remainder of this slipstream containing unrecovered CO_2, N_2, CO, H_2, H_2S, and COS, are recombined with the main acid gas stream for cleanup to remove sulfur gases before final discharge.

The “lean” AGR solvent is then pumped back up to system pressure and cooled for recycle to the Absorber. Solvent cooling is accomplished by indirect heat exchange with a chilled fluid on the way back to the Absorber. A slipstream of solvent is also taken through a carbon filter to remove dissolved hydrocarbons. Makeup solvent can also be added to the slipstream to replace solvent lost through evaporation, mainly in the acid gas.

13.4 DME Synthesis

Before the compressed synthesis gas from the acid gas removal unit enters the TIGAS synthesis unit, it is preheated to 430°F in the Hot Guard Feed/Effluent Exchanger and passed through the first reactor, the Hot Guard Reactor.

The primary function of the hot guard reactor is removal of potential trace amounts of contaminants that would cause catalyst deactivation in the downstream oxygenate synthesis. The reactor contains several layers of catalysts/absorbents that remove a variety of components e.g., COS, NH_3, As, HCN, H_2S and carbonyl compounds. The reactor layer will contain the following catalysts/absorbents:

Topsoe HTZ:	Sulfur absorbent for bulk removal and hydrolysis of COS and HCN.
Topsoe NT-101:	NH_3 absorbent.
Topsoe CT-101:	Carbonyl guard absorbent.
Topsoe ST-101:	Sulfur polisher and As guard.

The process gas leaving Hot Guard Reactor is cooled further heated to about 430°F in the MeOH/DME Preheater and passed to the MeOH/DME Reactor where the conversion of synthesis gas to methanol and dimethyl ether takes place according to the following reaction schemes:

(13.1) $$CO + 2H_2 \leftrightarrow CH_3OH$$

(13.2) $$CO + H_2O \leftrightarrow CO_2 + H_2$$

(13.3) $$2CH_3OH \leftrightarrow CH_3OCH_3 + H_2O$$

MeOH/DME Reactor is a boiling water reactor which ensures efficient heat removal of reaction heat. The boiling water reactor is constructed as a shell and tube heat exchanger, where the combination of methanol and DME synthesis catalysts is loaded in the reactor tubes. The shell side contains water at the boiling point which efficiently removes the reaction heat by evaporation. The shell side is connected to the Steam Drum. The hot spot temperature in the catalyst bed is controlled by adjusting the pressure of the steam in the steam drum.

The efficient cooling by the design provides for a high conversion and a modest temperature rise, to approximately 540°F exit the reactor and the concentration of methanol and dimethyl ether is 1.4 mole percent and 23 mole percent, respectively.

Some of the produced steam is used for preheating the synthesis gas in heat exchangers Hot Guard Trim Heater and MeOH/DME Preheater.

13.5 Gasoline Synthesis

In the Gasoline Reactors (three in parallel of which two are in operation), the oxygenates are converted to water and hydrocarbons by the Topsoe GSK-10 zeolite catalyst:

(13.5) $$n\ CH_3OH \leftrightarrow n\ (CH_2) + n\ H_2O$$

(13.6) $$n\ CH_3OCH_3 \leftrightarrow n\ 2(CH_2) + n\ H_2O$$

where (CH_2) denotes the hydrocarbon product. The net reaction may be formulated as:

(13.7) $$\text{Oxygenates (MeOH/DME)} \rightarrow C_{1\text{-}2} + C_{3\text{-}4}\text{(LPG)}, C_{5+}\text{(Gasoline)} + H_2O + \text{Heat}$$

Prior to entering the Gasoline Reactors the effluent from MeOH/DME Reactor is mixed with recycle gas from the downstream product separation. The mixture is preheated in the Gasoline Feed/effluent Exchanger and introduced to the Gasoline Reactors. As the inlet temperature of the Gasoline Reactors is essential for the performance, both an Electric Start-up Heater and a gas by-pass around the feed/effluent exchanger are included in order to maintain efficient control of inlet temperature with sufficient controllability margin.

The addition of the recycle stream serves as a heat sink for the exothermic dehydration reaction and reduces the temperature increase in Gasoline Reactors. The amount of gas recycled is determined by the reactor outlet temperature. The reactor outlet temperature is maintained fixed at approximately 750°F. The reactor recycle rate shall be adjusted to control the temperature rise to approximately 80-130°F in order to reduce coking rates and irreversible zeolite deactivation caused by the presence of steam at elevated temperatures. The control of inlet temperature and temperature increase reduces the formation of undesired by-products formed by cracking reactions which ultimately reduces the gasoline product yield.

During the operation, the zeolite catalyst GSK-10 becomes gradually deactivated by coke lay-down and must be regenerated by controlled burn-off. The need for regeneration will be observed by the methanol slip in the reactor effluent. After regeneration, the catalyst is ready to be taken into operation again. Such a period between two regenerations is defined as a cycle.

The gasoline synthesis unit consists of three parallel adiabatic Gasoline Reactors providing intermittent regeneration of individual reactors without affecting the plant productivity with two reactors in operation and one in regeneration or stand-by mode. The cycle life of the individual reactors is approximately 30 days, and gasoline properties are expected to vary over the life of the cycle, and from cycle to cycle. The three reactor system will ensure consistent gasoline quality during the operation of the plant. After approximately 20 number of cycles, the catalyst must be replaced.

The regeneration involves addition of air in controlled concentration at increasing temperatures. The control philosophy of the regeneration system is based on monitoring the oxygen concentration in and out of the reactor. When the oxygen slip occurs, the inlet temperature is increased to the next level.

During the initial regeneration sequence, the hydrocarbon in the catalyst is burned and water is formed and separated from the regeneration gas. The carbon on the catalyst will react with the oxygen in the air at different rates depending on the chemical structure.

For controlling of the oxygen concentration, the flow of filtered ambient air from the Air Compressor is adjusted in proportion to the regeneration loop recycle flow. The recycle flow is assured by the Regeneration Compressor. The required concentration of oxygen in the recycle to the reactor will be within the range of 0.3 to 20 mole percent. The oxygen level is increased keeping a maximum temperature rise over the reactor of 90°F.

The effluent gas from the reactor will contain water, carbon dioxide, and carbon monoxide. The liquid is separated in the Regeneration Separator.

A purge stream from the loop is kept to have a CO_2 concentration at about 21 mole percent and no more than 25 mole percent.

13.5.1 Product Separation

The process gas leaving the Gasoline Reactors is cooled and chilled to about 40°F. The cold process stream now consists of three phases which are separated in the High Pressure Separator. The process water is taken out in the lower part of the separator and depressurized, and dissolved gases are partly removed in the Process Water Flash Drum. The process water is finally sent to on-site water treatment facilities before reuse and/or disposal.

The primary product, the raw gasoline phase, is taken out in the middle of the High Pressure Separator. The product stream is depressurized and the LPG is partially vaporized which causes a significant decrease in temperature. The stream is heated in the Effluent Cooler to approximately 50°F before it is sent to fractionation section.

The gas phase from the High Pressure Separator consists primarily of unconverted synthesis gas, inerts, and light hydrocarbons from the gasoline synthesis. Most of this is recycled to the Gasoline Reactors for temperature control. The recycle gas is compressed in the Recycle Gas Compressor before it is preheated in the Recycle Gas Preheater. The remaining fraction is partly recycled to the gasification unit and the remaining fraction of gas is used as fuel gas.

13.6 Gasoline Fractionation

The hydrocarbon phase is sent to the De-ethanizer, where the light gas components (H_2, CO_2, CH_4, C_2H_6 and C_2H_4) are separated from the raw gasoline at the top of the column. The light gas is used as fuel gas within battery limit.

The De-ethanizer uses the De-ethanizer Steam Reboiler to strip the gases off the gasoline.

Water entering the De-ethanizer will leave with the fuel gas. In case too much water reaches the column, a separate water phase will occur on the top tray. This is automatically removed by De-ethanizer Water Draw-off Pot.

The bottom product consists of LPG, gasoline, and a small fraction of higher boiling material called heavy fuel oil (HFO), which is used as a sulfur-free fuel in the auxiliary boiler/HRSG unit. The product stream is sent to the LPG Splitter where the main part of the LPG is removed. By controlling the amount of LPG that remains in the bottom product, the gasoline Reid vapor pressure (RVP) of the final gasoline product is adjusted.

The LPG Splitter uses the LPG Splitter Steam Reboiler to supply the heat required for the fractionation. The overhead product stream is cooled in the LPG Splitter OVH Condenser and then sent to the LPG Splitter OVH Drum. The liquid is pressurized in the LPG Splitter Reflux Pump before most of it is returned to the column as reflux and the remaining exported as LPG product. The LPG product stream is primarily butane and propane in a 3/1 ratio (weight). The liquid from the LPG Condenser is minimum 10°F sub-cooled to minimize flashing in the LPG storage.

The bottom product from the LPG Splitter is sent to the Gasoline Splitter where the gasoline is split into three fractions: Light gasoline, heavy gasoline, and a minor sulfur free fuel HFO stream. The overhead product stream is cooled in the Gasoline Splitter OVH Condenser and sent to the Gasoline Splitter OVH Drum. The liquid is pressurized in the Gasoline Splitter Reflux Pump before some is returned to the column as reflux and the remaining exported as light gasoline product. The light gasoline constitutes approximately 75% of the total product gasoline and is after cooling in the Light Gasoline Product Cooler sent directly to the Light Gasoline Shift Tank.

The remaining approximately 25% has a relatively high durene content and is therefore sent to the gasoline upgrading unit ("GUU") where the durene content is reduced and the octane number increased. The lower part of the Gasoline Splitter separates the heavy gasoline from the sulfur free fuel stream. The fraction of gasoline which is lost here is very small but it contains the heaviest hydrocarbons and hence the adjustment of the flow rate of this fuel stream is used to control the final boiling point of the gasoline to meet the gasoline product specifications.

13.7 Gasoline Upgrade Unit

One of the components that greatly differs between the synthetic gasoline produced from synthesis gas and a conventional gasoline product is the durene (1, 2, 4, 5-tetramethylbenzene). Durene has a high melting point and with a high content of durene, the gasoline can plug the injection system or freeze in the tank.

The GUU consists of an Isomerization Reactor system where durene is isomerized resulting in a lower melting point to meet final gasoline product specification. Furthermore, the upgrade unit increases the octane number of the final product.

Before being preheated, the gasoline stream which is sent to the GUU is mixed with a small amount of imported hydrogen. After preheating in Feed-Effluent Exchanger followed by a trim heater the stream of gasoline and hydrogen is fed to the two Isomerization Reactors. The two reactors in series allow replacement of the catalyst in one of the reactors during operation of the plant.

The gasoline product from the GUU is cooled in the Feed-Effluent Exchanger and Air Cooler and separated from off gases in the HP Cold Separator. The liquid gasoline is cooled in the Heavy Gasoline Product Cooler. The gasoline is depressurized and the off-gas released is separated in the Low Pressure Cold Separator. The product is sent to Heavy Gasoline Shift Tank from where it is mixed with the light gasoline and exported as gasoline product.

Key specifications for the gasoline produced in the unit are summarized in Table 13-1. The RVP-specification is the one that has been used to calculate the production yield. The RVP (and therefore also yield and to a slight degree compositions) may vary and will depend on the location of the plant, the time of the year, and the specific product gasoline requirements. The production of gasoline with a different RVP will impact the yields of gasoline and LPG, since higher RVPs allow for more LPG components to be included in the gasoline product.

Table 13-1 Gasoline Specifications

Flow	lb/h	44,068
Pressure	psig	45
Temperature	deg. F	114
RVP	psia	9.5
n-Parafins	wt %	~5
i-Parafins	wt %	45-55
Olefins	wt %	5-15
Naphthenes	wt %	5-10
Aromatics	wt %	25-35
Benzene	wt%	<0.5
Sulfur	ppmw	< 5
Octane	(RON+MON)/2	88

13.8 Production and Consumption Figures

All continuous process streams flowing between the different units are shown in the block flow diagram (BFD) attached to this report. As it is shown on the diagram, on a daily basis 6400 short tons wood with 50% moisture is converted into 4106 barrels of refined gasoline and 554 barrels of LPG. This constitutes 1.283 bbl gasoline blendstock and 0.173 bbl LPG per dry ton of wood.[*]

The higher gasoline yield for the commercial plant compared to the pilot demonstration plant is partly due to the higher degree of integration - including higher level of recycle flow - in the commercial plant. Due to the existing gasifier in the pilot plant, it was not possible to increase recycle to the optimal degree, but it was always the intention to do so in the commercial plant. Another important factor is the nature of the Topsoe GSK-10 gasoline catalyst. The gasoline yield will increase during the cycle life and peak just before the catalyst is getting ready for regeneration. The operation with several gasoline reactors in parallel in the commercial plant results in a relatively steady, average yield that is higher than the catalyst

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[*] Specification feedstock contains 1.25 wt% ash on a dry basis. Product densities are 6.13 lb/gal (gasoline) and 4.66 lb/gal (LPG).

yield experienced in the single gasoline reactor in the pilot plant, where the tests were carried out over fresh catalyst.

A minor natural gas import is required to close the power and steam balance. The overall water balance for the plant results in a water export of about 1,600 STPD which is treated before leaving the site. Finally the minerals balance shows an import of about 50 short tons per day of gasifier bed material balanced by an ash export of 153 short tons per day. The power balance is shown in Table 13-2.

Table 13-2 Overall Power Balance

Unit	Description	kW
50	ASU	7,450
80	Aux. boiler/HRSG	75
100	Wood dryer	3,950
300	Gasifier	750
700	Syngas compressor excl. turbine	75
800	AGR	4,100
900	TIGAS	1,050
1000	Fractionation	3,150
	Miscellaneous	2,230
	Total consumption	22,830
80	Power generation	(22,830)

13.9 Estimated CAPEX Contributions

- The CAPEX contributions listed below include all costs associated with the construction of each of the units of the IBR:
    - Equipment and instrumentation costs
    - Process piping and electrical
    - Foundation, piperacks and structures
    - Installation including construction management
    - Engineering, licenses, contractors fees and contingency
    - Catalysts (1st charge)

- Costs related to utilities and infrastructure as well as their connections to the ISBL process units has been collected under the heading: OSBL costs. These are the facilities for preparation/handling of:
    - Waste water treatment/plant water
    - Demineralized water (DMW) and boiler feed water (BFW) preparation
    - Cooling water
    - Instrument air and plant air
    - Tankage for product gasoline and LPG
    - Chiller for cooling AGR solvent and TIGAS products

- The following costs are excluded from the cost estimate:
    - Costs of buildings
    - Cost of land
    - Costs related to possible excessive site preparation and site development
    - Capital spares
    - Forward escalation
    - Financial charges
    - Taxes, duties and levies
    - Owner’s cost

All of the costs below are November 2014 costs:

Table 13-3 Survey of CAPEX Contributors

CAPEX Contributions		
Unit number	Description	Cost
ISBL Units		
100	Wood yard and dryer	$46 MM
200	Air separation unit	$48 MM
300	Gasification and gas cleanup unit	$238 MM
700	Syngas compression unit	$9 MM
800	AGR	$52 MM
900 + 1000	TIGAS unit incl. prod. upgrade	$90 MM
80	Aux Boiler/HRSG unit	$3 MM
	Power generation unit	$14 MM
	ISBL contingency (25%)	$125 MM
Total ISBL		$625 MM
OSBL Units		
	Waste Water treatment	$19 MM
	Tank farm for gasoline and LPG	$11 MM
	Cooling Water tower	$14 MM
	DMW unit	$2 MM
	Chiller unit	$8 MM
	Additional units & Infrastructure (assumed as 75% of OSBL costs)	$40 MM
Total OSBL		$94 MM
Total		$719 MM

The costs above have been calculated an exchange rate of 1€=$1.12, which is the exchange rate that results from the “broad method” as defined by Bank of International Settlements when basing the calculations on the exchange ratios on December 19, 2014. This CAPEX also corresponds to a capital efficiency $175,050 per bbl/d gasoline output (2014$).

14. Commercial Plant Scale–up Considerations

14.1 Overall Plant Size

The overall size of the commercial plant is dictated by two main criteria – economics of scale and availability of woody biomass residue at economical price. The first leads to a plant size as large as possible to minimize the production cost of gasoline. However, there are limits to the availability of biomass residue at attractive price at a given location. Large quantities of biomass residue available from logging operations for lumber and pulp mills are concentrated in certain parts of U.S. – Southeast, Northwest, Northeast and upper Midwest. The most important factor in the cost of wood residue feedstock is transportation and hauling and studies have shown that collection from beyond a radius of 50 to 75 miles from plant site is not economical. Based on these considerations and several studies by others, at a plant site in any of the above geographical areas, about a maximum of 6000 to 7000 short tons per day of forestry residue can be available at economical price. Therefore, we have selected the commercial plant size to be 6400 short tons per day of greenwood forest residue (~50% wet) or dried biomass feed to gasifiers of 4000 short tons per day (20% wet) corresponding to 3200 short tons per day bone dry wood. This size corresponds to an ASU producing approximately 2,000 short tons per day of oxygen and this is well within the range that is industrially proven.

14.2 Scale-up

The commercial plant has four main subsystems – gasification, acid gas removal, TIGAS and oxygen plant. Of these, except for gasification, all other systems can be scaled up so that a single train or unit is sufficient to handle the entire plant production. Acid gas removal units like Morphysorb®, Rectisol and Selexol are today in commercial operation in a size range that includes the size of the proposed integrated biorefinery (IBR). The TIGAS gasoline unit comprises equipment items like fixed bed catalytic reactors, heat exchangers, compressors and separators that are similar in design to those commercially demonstrated in numerous methanol units of similar or larger scale. A single train 15,000 bpd TIGAS gasoline plant was recently awarded to Topsoe confirming the successful commercialization of such larger gasoline plants.

A single gasifier is not able to handle 4000 short tons per day of biomass feed. Based on our experience from several different pilot and commercial plants, we have selected a single gasifier of 1000 short tons per day capacity so that the commercial plant has 4 trains of gasification plant. The basis for selecting this size for a single gasifier are primarily the testing done at GTI for producing clean synthesis gas from different woody biomass feedstock and the commercial plant in Skive, Denmark (Figure 14-1).

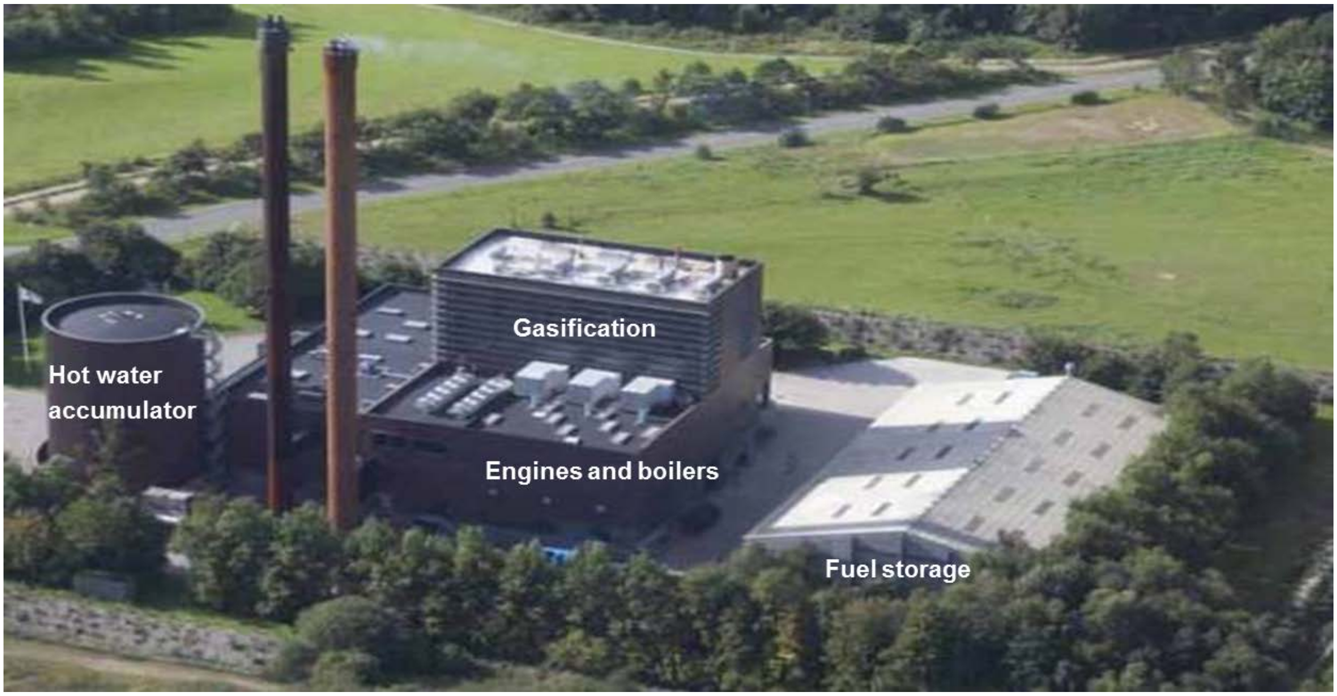

Figure 14-1 Skive Gasification Plant

The gasifier at GTI has operated under oxygen gasification and pressure of 10 barg. In several tests totaling more than 4000 hours all design parameters have been established – temperature, fluidization velocity, oxygen feeding, lock hopper feeding, carbon conversion, residence time, steam/carbon ratio, tar reforming, gas recycle, etc. The Skive plant is a commercial delivery and has been operating for several years. This plant has a gasifier of about 12 feet diameter, operating pressure of 3 barg and feed capacity of about 150 short tons per day of biomass. Operating experience from this plant has proved scaling up the gasifier dimensions and hydrodynamics of BFB. The commercial gasifier capacity of 1000 short tons per day is therefore obtained by keeping the gasifier dimensions approximately the same as the Skive plant but increasing the operating pressure from 3 barg to 10 barg and also taking into account a higher gasification reaction rate due to increase in pressure.

With 4 gasifier trains the availability of the commercial plant also increases as there is some spare capacity in each gasifier so that about 90% output can be maintained with one gasifier down.

15. Greenhouse Gas/Lifecycle Analysis

The LCA submitted with the original application stated 91.7% GHG reduction for the IBR gasoline product compared to conventional gasoline. New information based on pilot testing, product evaluation, and further detailing of the commercial plant layout required a recalculation of that analysis, resulting in a revision of the GHG reduction to 65-74%, depending on process options discussed below.

The basis for the change comes from one item that was not included in the original analysis and two changes in the commercial plant design:

- Biogenic CH_4 and N_2O emissions from gasoline combustion, totaling 0.49 g CO_2-eq/MJ.[*]
- Use of a steam drive for syngas compression, which requires an auxiliary boiler fueled partly by process gases (tail gases) and byproduct heavy fuel oil (HFO).
- Additional on-site processing of the raw gasoline via distillation and isomerization of distillation bottoms to maximize product value. This results in lower gasoline yield per ton of biomass, which raises the relative impact of feedstock production and transportation.
- Evaluation of several options to meet steam and power requirements, which in turn affect both the LCA GHG reduction and plant economics.

The LCA is expressed in terms of the product energy (MJ) per unit of GHG emissions (g CO_2-eq), we evaluated five options (detailed later in Chapter 16) to provide the needed steam and power for pumps, blowers, compressors, and other powered equipment. These options and the resulting LCA data are shown in Table 15-1 below.

Table 15-1 Power and Steam Options for Commercial Plant

IBR Case	Electric power source	On-site power plant fuel (if used)	Auxiliary steam fuel/energy[†]	LCA GHG reduction
1	Grid	--	Natural gas	65.3%
2	Grid	--	LPG	66.9%
3	On-site	LPG	LPG + HRSG	69.4%
4	On-site	LPG	Natural gas + HRSG	71.0%
5	On-site	Natural gas	HRSG	73.7%

In cases 3, 4, and 5, an on-site power plant and heat recovery steam generator (HRSG) were included to supply power as well as waste heat to offset steam requirements in the auxiliary boiler. LPG that is not burned on site is available for sale, increasing both revenue and plant output of bio-derived product.

LCA results for petroleum gasoline and the five IBR cases are detailed below in Table 15-2.

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[*] "GREET Model, The Greenhouse Gases, Regulated Emissions, and Energy Use in Transportation Model," Argonne National Laboratory, Chicago, Illinois (2013).

[†] In addition to process gases (tail gases) and byproduct HFO.

Table 15-2 LCA Calculations

Case	Petroleum gasoline	IBR Case 1	IBR Case 2	IBR Case 3	IBR Case 4	IBR Case 5
			g CO_2eq/ MJ product			
Feedstock chemicals	0.00	0.00	0.00	0.00	0.00	0.00
Feedstock Production	6.92	6.88	6.88	6.88	6.88	6.88
Feedstock Transportation	1.33	4.03	4.03	4.03	4.03	4.03
Fuel Production	9.29	19.20	17.14	14.33	12.89	11.58
Fuel Transportation	1.04	1.04	1.04	1.04	1.04	1.04
Fuel Combustion	72.60	0.49	0.49	0.49	0.49	0.49
Total	91.18	31.64	29.58	26.77	25.33	24.02
LCA GHG reduction	--	65.3%	66.9%	69.4%	71.0%	73.7%

16. Commercial Plant Proforma

The project team developed preliminary (±30%) techno-economic analyses of five cases referenced in Section 15. The cases are described again below.

1. Facility power purchased from the grid; facility steam from auxiliary boiler fueled by process gases, byproduct fuel oil, and natural gas; 100 percent of produced LPG sold for revenue.
2. Facility power purchased from the grid; facility steam from auxiliary boiler fueled by process gases, byproduct fuel oil, and LPG; the portion of LPG not consumed in the auxiliary boiler is sold for revenue.
3. Facility power provided by an LPG-fired on-site gas turbine (Solar Centaur 40) followed by a heat recovery steam generator (HRSG); facility steam provided by HRSG and auxiliary boiler fueled by process gases, byproduct fuel oil, and remaining LPG.
4. Facility power provided by an LPG-fired on-site gas turbine (GE 10-1) followed by HRSG; facility steam provided by HRSG and auxiliary boiler fueled by process gases, byproduct fuel oil, and natural gas; all LPG is consumed on site.
5. Facility power provided by a natural gas-fired gas turbine (P&W SwiftPac 25) followed by HRSG; facility steam provided by HRSG with supplemental firing of process gases and byproduct fuel oil; 100 percent of produced LPG sold for revenue. This is the base case.

16.1 Proforma Methodology and Assumptions

The analyses were documented in pro forma spreadsheets that were adapted from a 2004 DOE-provided template, updated to reflect technology-specific factors *(e.g.*, product and byproduct streams) and recommendations by the Independent Engineer. Three sets of input data provided inputs to the calculation table: (a) technical and financial assumptions; (b) capital expense estimates; and (c) commodity price data forecasts from EIA databases.

The Technical and Financial Assumptions common to all IBR cases are given in Table 16-1.

- Availability ramp-up was based on typical gasification plant ramp-up.
- Mature availability, *i.e.*, after the ramp-up period was based on natural gas providing start-up power and steam, with tail gases and HFO flared until system is ready for supplemental firing fuel switchover.
- Heating values were obtained from pilot data or published typical values.
- Year 1 of financing was assumed to start January 1 in year of construction.
- Process unit installations assumed to commence one year before startup.
- Overtime hours were not considered.
- Inflation from 2014 to plant construction was 2.00%, subsequent year inflation was linked to EIA price forecasts for commercial gasoline.
- Labor rates were obtained from U.S. Department of Labor based on plant siting in the State of Minnesota.

Table 16-1 Technical and Financial Assumptions (All Cases)

Operational		
Availability, % (year 1)	51.0	
Availability, % (year 2)	87.1	
Availability, % (year 3+)	91.0	
Fuel Properties		
Biomass moisture (as rec'd), wt%	50.0	
Biomass moisture, (as fed), wt%	20.0	
Gasoline density, lb/gal	6.13	
Heating values, Btu/lb	HHV	LHV
Biomass (dry basis)	8,283	7,721
Biomass (as rec'd)	4,119	3,336
Gasoline	19,776	19,262
LPG	21,561	19,679
Heavy fuel oil	21,874	19,965
Natural gas	24,822	22,233
Hydrogen	61,127	51,682
Financial		
Dollar basis year	2014	
Startup year	2019	
Term of loan, years	25	
Lifetime of plant, years	30	
Salvage value, % of CAPEX	10.0	
Interest rate, % annual	8.0	
Equity investment, % of CAPEX	50.0	
Insurance, % of CAPEX	1.0	
Property tax, % of depreciated value	1.0	
License fee, $/bbl gasoline	1.0	
Management fee, % of revenue	1.0	
Maintenance, % of CAPEX	2.5	
Contingency, % of OPEX	16.0	
Labor benefits, % of wages	35.0	
Annual labor hours per worker	2088	
Labor Category wage, $/h		
Operator	$25.80	
Operator Supervisor	$27.07	
Site Manager	$49.57	
EH&S Manager	$42.15	
Process Engineer	$43.99	
Lab Supervisor	$43.16	
Lab Technician	$23.84	
Production Manager	$46.98	
HR Manager	$52.05	
Purchasing & Receiving	$29.37	
Logistics Coordinator	$36.52	
Accounting Manager	$40.17	
IT Technician	$29.49	
General Administration	$16.26	

16.2 Capital Expenses (CAPEX) Estimates

CAPEX estimates were previously shown for the base Case 5 in Table 13-2, but are repeated in Table 16-2 with additional detail for the four alternate IBR cases. The only variations involve (a) LPG storage tanks in cases where LPG is consumed in the boiler or on-site power plant, and (b) cases that include an on-site power unit. All estimates were obtained in 2014$ except for on-site power plants which were in 2013$ and were therefore adjusted to 2014$. For the Proforma calculations, CAPEX estimates were escalated to 2018$ based on assumed 2.00% general inflation from 2014 to 2018.

Table 16-2 CAPEX Estimates (2014$MM)

Unit/Area	Case 1	Case 2	Case 3	Case 4	Case 5
ISBL					
Wood yard and dryer	$45.6	$45.6	$45.6	$45.6	$45.6
Air separation unit	$48.0	$48.0	$48.0	$48.0	$48.0
Gasification plant	$238.4	$238.4	$238.4	$238.4	$238.4
Syngas compression	$8.8	$8.8	$8.8	$8.8	$8.8
AGR plant	$52.0	$52.0	$52.0	$52.0	$52.0
TIGAS including product upgrading	$89.6	$89.6	$89.6	$89.6	$89.6
Auxiliary boiler	$3.2	$3.2	$3.2	$3.2	$3.2
On-site power plant (when equipped)	$0.0	$0.0	$4.2	$7.9	$14.3
Contingency 25% of ISBL units	$121.4	$121.4	$122.4	$123.4	$125.0
OSBL					
Water treatment	$18.6	$18.6	$18.6	$18.6	$18.6
Gasoline storage	$6.0	$6.0	$6.0	$6.0	$6.0
LPG storage	$5.3	$2.2	$0.0	$0.0	$5.2
Cooling tower	$13.7	$13.7	$13.7	$13.7	$13.7
Demineralizer	$2.2	$2.2	$2.2	$2.2	$2.2
Chiller	$8.1	$8.1	$8.1	$8.1	$8.1
Infrastructure 50% of OSBL units	$27.0	$25.5	$24.3	$24.3	$26.9
Contingency 25% of OSBL units	$13.4	$12.7	$12.2	$12.2	$13.5
TOTAL	$701.3	$696.0	$697.3	$702.0	$719.1

16.3 Pricing

Price forecasts for key products and utilities (gasoline blendstock, LPG, natural gas, electricity) were obtained from U.S. Department of Energy, Energy Information Administration, 2014 Annual Energy Outlook (AEO) tables, Reference Case.[*] State and Federal taxes were subtracted from products, and wholesale price of blendstock (RBOB) was adjusted for by subtracting ethanol from commercial E10, with an effective ethanol price derived from the E85 price forecast. An estimated transportation and marketing cost of $0.12/gal was also subtracted from retail E10 and E85 products. A summary of key EIA price forecasts for years 2018-2048 is shown in Figure 16-1.

Pricing for the feedstock and other utilities and supplies (bed media, hydrogen, AGR solvent, catalysts, SOx/NOx permits) were obtained from published or private sources and escalated per general inflation.

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[*] Accessed 11/11/2014.

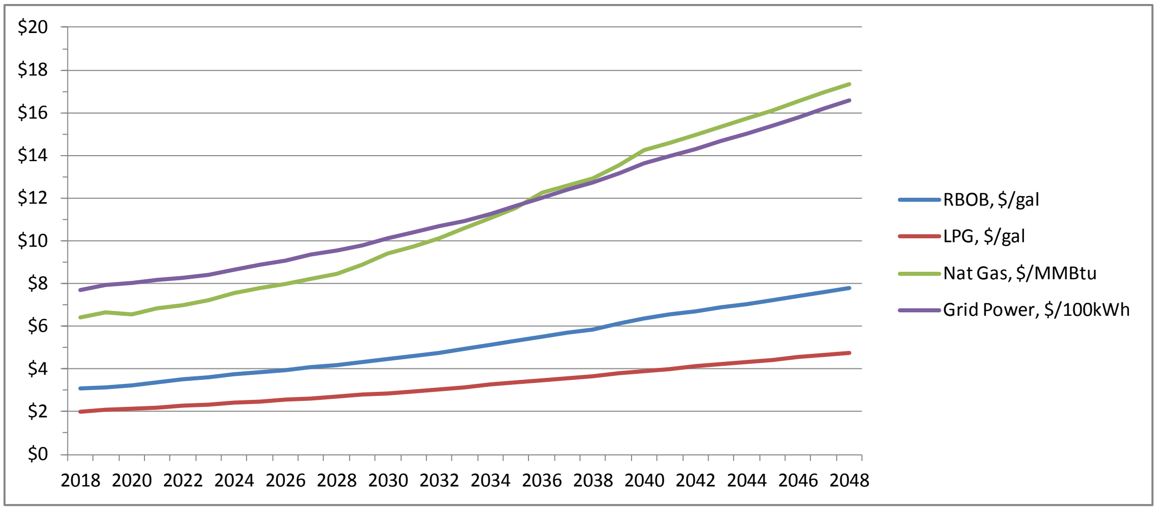

Figure 16-1 EIA Price Forecasts for Wholesale Gasoline (RBOB), LPG, Natural Gas, and Electricity

16.4 Process Inputs

Input and output data specific to the five IBR cases are shown in Table 16-3, along with summaries of steam and power options.

In Cases 3,4, and 5, on-site power was added to the plant design and included in the CAPEX. The power units in all cases were simple -cycle gas turbines that were available off the shelf and of suitable size for the available fuel and power needs. Costs and specifications for the power plants were obtained from *Gas Turbine World 2013*.[*] In all cases, an assumption was made that the power packages could be equipped for firing either natural gas or LPG without added cost, and that supplemental firing and HRSG packages could be integrated with the ISBL Auxiliary Boiler without significant added CAPEX. Power plant costs were adjusted from 2013$.

Table 16-3 Input/Output Data with Steam and Power Summaries

	Case 1	Case 2	Case 3	Case 4	Case 5
Inputs					
Biomass fuel, as rec'd, lb/h	533,216	533,216	533,216	533,216	533,216
Biomass fuel, as fed, lb/h	333,260	333,260	333,260	333,260	333,260
Biomass fuel, bone dry, lb/h	266,608	266,608	266,608	266,608	266,608
Dryer air, lb/h	3,560,979	3,560,979	3,560,979	3,560,979	3,560,979
Fluid bed media, lb/h	4,224	4,224	4,224	4,224	4,224
Hydrogen, lb/h	169	169	169	169	169
Boiler feed water, lb/h	108,786	108,786	108,786	108,786	108,786
AGR Solvent, lb/h	74	74	74	74	74
Natural gas, lb/h	2,422	0	0	1,050	9,200
Electricity (from grid), MW	22.83	22.83	18.34	14.65	0.00

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[*] *Gas Turbine World 2013 GTW Handbook*. Vol. 30. June 2013.

	Case 1	Case 2	Case 3	Case 4	Case 5
Outputs					
Gasoline, lb/h	44,068	44,068	44,068	44,068	44,068
LPG, lb/h	4,520	1,783	0	0	4,520
Ash, lb/h	12,738	12,738	12,738	12,738	12,738
Dryer air exhaust, lb/h	3,760,935	3,760,935	3,760,935	3,760,935	3,760,935
Process water, lb/h	201,928	201,928	201,928	201,928	201,928
Blowdown, lb/h	2,112	2,112	2,112	2,112	2,112
Acid gas, lb/h	269,167	269,167	269,167	269,167	269,167
N_2, lb/h	530,291	530,291	530,291	530,291	530,291
Steam summary, MMBtu/h					
From process gases	66.79	66.79	66.79	66.79	66.79
From HFO	18.05	18.05	18.05	18.05	18.05
From natural gas	53.86	0.00	0.00	23.35	0.00
From LPG	0.00	53.86	34.08	0.00	0.00
From onsite power plant waste heat	0.00	0.00	19.78	30.51	53.86
Power Summary, MW					
From grid	22.83	22.83	18.34	14.65	0.00
Onsite power from natural gas	0.00	0.00	0.00	0.00	22.83
Onsite power from LPG	0.00	0.00	4.49	8.18	0.00

16.5 Proforma Results

Based on these inputs and assumptions, operating revenues, operating expenses, net revenues, debt service, net cash flow, gasoline production cost, and internal rate of return (IRR) were calculated for the five cases. Capital efficiency in terms of $/bbl/day and $/annual gallon were also determined.

IRR was calculated for two scenarios: (1) where no biofuel support policy was considered, and (2) where the Cellulosic Biofuel Waiver Credit stipulated in 40CFR80 section 80.1456(d) creates a higher market price in competition with conventional gasoline where the Obligated Parties (refiners) can opt to either purchase petroleum-based RBOB and pay the waiver or purchase bio-derived gasoline from the IBR. This applies to the IBR product because the LCA GHG reduction is above 60% and therefore the gasoline qualifies as a Cellulosic Biofuel (CB).

Key results of this analysis are shown in Table 16-.4. The tradeoffs between LPG sale, LPG consumption on site, and source of electricity involve a combination of economic and environmental policy drivers. The IRR in the absence of a policy driver ranges from 9.2 to 11.8%. With the current driver of CB Waiver Credits projected forward, the IRR ranges from 13.1 to 15.9%.

It is evident that the lowest gasoline production cost, highest IRR, and highest GHG reduction are all achieved in Case 5 with on-site power provided by a natural gas-fired turbine. The next highest IRR comes from Case 1 with 100% grid power, but that one has the lowest GHG reduction. All cases where LPG is consumed on site are intermediate in GHG reduction but significantly less profitable than either Case 1 or Case 5 because of the loss of byproduct revenue.

Detailed Proforma spreadsheets are provided in Appendix J.

Table 16-.4 Results of Pro Forma Evaluation for Five IBR Cases

IBR Case	CAPEX, $MM	Capital Efficiency (2014$)		LCA GHG reduction	Gasoline production cost, $/gal	IRR, %	
		$/bbl/d	$/annual gal			No policy	Cellulosic biofuel support
1	$701.3	$170,708	$12.78	65.3%	$2.62	11.3%	15.4%
2	$696.0	$169,437	$12.69	66.9%	$2.75	10.0%	13.9%
3	$697.4	$169,756	$12.71	69.4%	$2.84	9.1%	12.9%
4	$702.0	$170,890	$12.80	71.0%	$2.83	9.2%	13.0%
5	$719.1	$175,057	$13.11	73.7%	$2.56	11.8%	15.9%

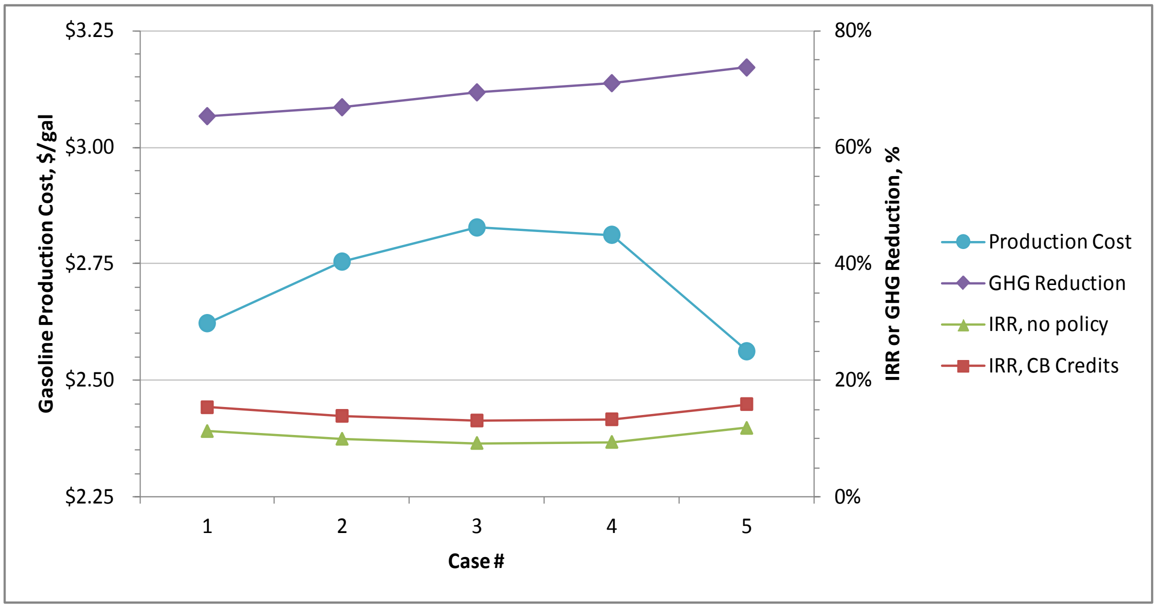
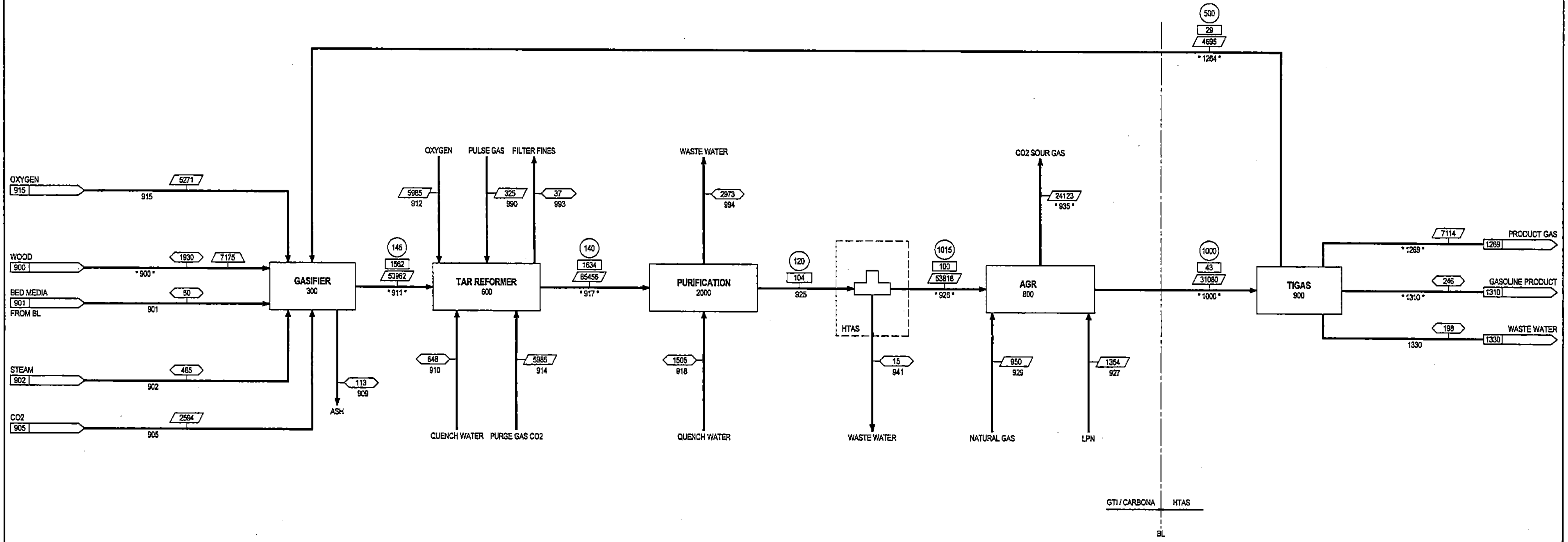


Figure 16-2. Gasoline Production Cost, IRR, and GHG Reduction for Five IBR Cases

Attachments:

- Block flow diagram for wood to gasoline pilot plant (22.5 bpd)
- Block flow diagram for commercial size wood to gasoline plant (4,106 bpd)



STREAMS	" 900 "		" 911 "		" 917 "		" 926 "		" 935 "		" 1000 "		" 1264 "		" 1269 "		" 1310 "	
COMP.	lb/h	wt %	SCFH	mole %	SCFH	mole %	SCFH	mole %	SCFH	mole %	SCFH	mole %	SCFH	mole %	SCFH	mole %	lb/h	wt %
ASH	45	2.31																
C2			327	0.61	21.8	0.03	22	0.04	12	0.05	9.57	0.03	5.59	0.12	10	0.14		0.01
C3													8.63	0.18	23.8	0.33	1	0.25
C4													26.8	0.57	103	1.45	18	7.37
C5+			207	0.38	45.5	0.05	35	0.07	35	0.15	0.466		14.8	0.31	56.2	0.79	222	90.53
CH_4			4376	8.11	95.7	0.11	86.1	0.16	14	0.06	74.6	0.24	75.2	1.60	110	1.54		
CO			9756	18.08	15174	17.76	14810	27.52	1943	8.06	12874	41.42	723	15.40	1008	14.17		0.01
CO_2			14008	25.96	21144	24.74	20999	39.01	20289	84.10	724	2.33	1535	32.70	2640	37.11	5	1.83
COS			0.594															
H_2			11807	21.88	17552	20.54	17280	32.11	967	4.01	16321	52.52	1855	39.53	2538	35.68		
H_2S			4.64	0.01	4.61	0.01	4.47	0.01	4.4	0.02								
HCN			0.54				0.484											
N_2			480	0.89	550	0.64	533	0.99	809	3.35	1075	3.46	450	9.59	625	8.79		
NH_3			44.6	0.08	1.88		0.161		0.16									
WOOD	1769	91.69																
H_2O	116	6.00	12950	24.00	30867	36.12	48.4	0.09	49	0.20	0.373							
TOTAL	1930	100.00	53962	100.00	85456	100.00	53818	100.00	24123	100.00	31080	100.00	4695	100.00	7114	100.00	246	100.00
MOLE WEIGHT	8.09		23.34		23.04		25.93		40.54		14.71		23.12		25.32		87.65	

○ Pressure psig
⬡ Flow lb/h
⏢ Lower Heating Value BTU/lb
▱ Flow SCFH
▭ Temperature °F

Calc No's: 1297812
Layout: L1
PFF: F1
Case: C2
Name: PFD1

ORIGINAL IF RED

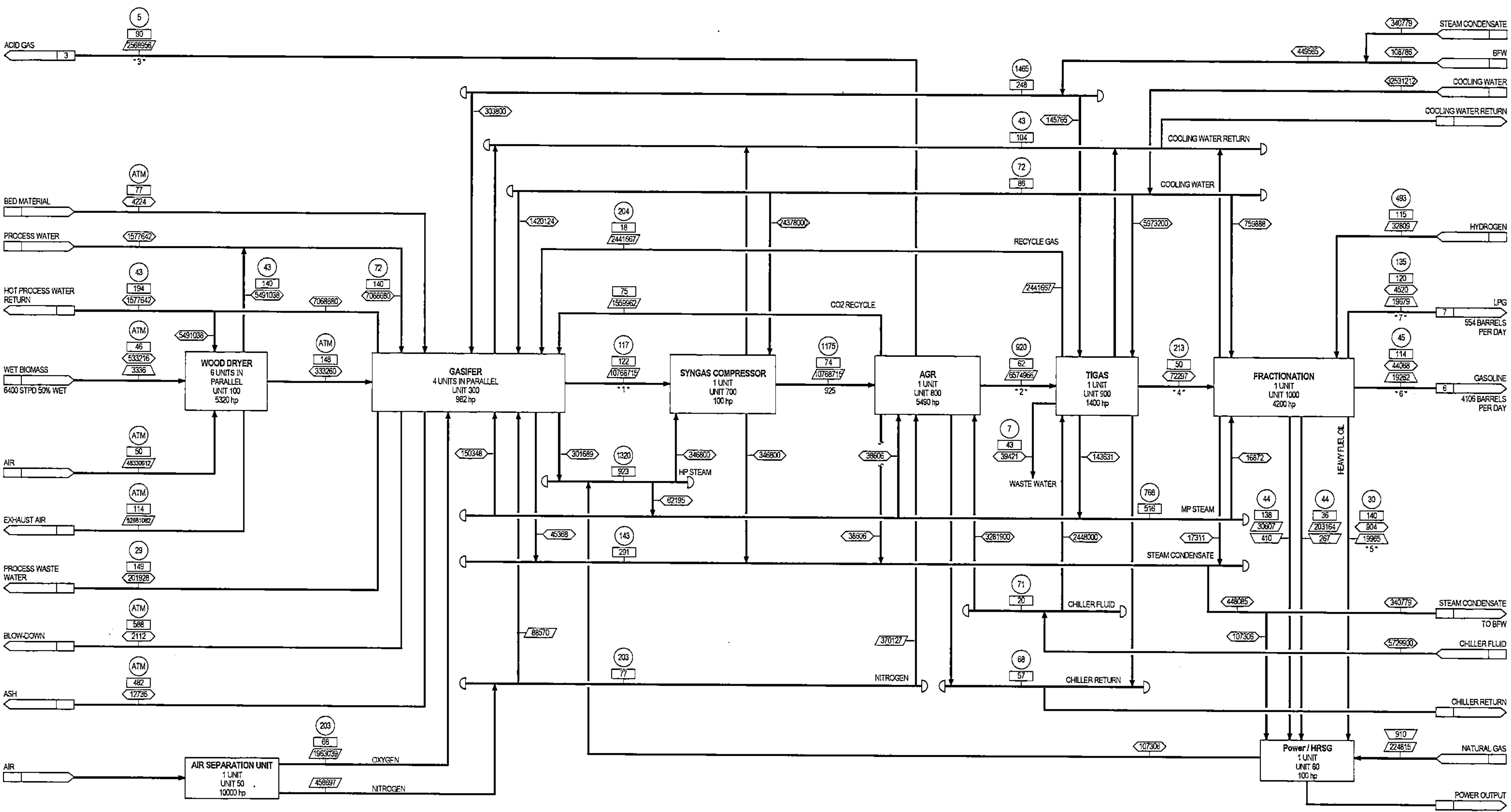
Rev	Description	Date	Made By	Checker	Approver
2	General Revision	19-Nov-2010	MAES		WA
1	General Revision	8-Oct-2010	MAES	ANK	JTJ
0	First Issue	12-Aug-2010	MAES	ANK	JTJ

HALDOR TOPSØE A/S
DK-2800 LYNGBY, DENMARK

Green Gasoline from Wood
DOE - Carbona Gasifier and Topsoe TIGAS
22.5 barrels/day TIGAS Plant

GTI/Carbona

Doc Grp: 41
Job: S-02460
Doc No: 1346681
Rev: 2



Calc No's: 1776446
Layout: L4
PFF: BLOK

Case: C1
Doc: BLOK

STREAMS	"1"		"2"		"3"		"4"		"5"		"6"		"7"	
COMP.	SCFH	mole %	SCFH	mole %	SCFH	mole %	lb/h	wt %	lb/h	wt %	lb/h	wt %	lb/h	wt %
C2H6							70	0.10			14	0.03		1 ppm
C5+							45335	62.69	904	100.00	43316	98.30	90	2.00
CH_3OH														
CH_4	46305	0.43	44052	0.67	599	0.02	107	0.15				11 ppm		
CO	3332917	30.95	3080371	46.85	95476	3.72	887	1.23						
CO_2	3790588	35.20	131499	2.00	2183460	85.00	19156	26.50						
H_2	3468603	32.21	3276963	49.84	53275	2.07	28	0.04				7 ppm		
H_2S					236	92 ppm						5 ppm		
LPG							6667	9.22			738	1.67	4430	98.00
N_2	44152	0.41	42080	0.64	233216	9.08	47	0.07						
H_2O	86150	0.80			2694	0.10								
TOTAL	10768715	100.00	6574966	100.00	2568956	100.00	72297	100.00	904	100.00	44068	100.00	4520	100.00
MOLE WEIGHT	25.14		15.29		41.06		66.05		152.98		94.43		55.13	

○	Pressure psig
⬡	Flow lb/h
⏢	Lower Heating Value BTU/lb
▱	Flow SCFH
⏢	Lower Heating Value BTU/SCF
▭	Temperature °F

HALDOR TOPSØE

REV.	DESCRIPTION	DATE	DRAWN	CHECK	APPR.

Green Gasoline from Wood
DOE - Carbona Gasifier and Topsoe TIGAS
4106 barrels/day TIGAS Plant
Commercial case
Design

Process Flow Diagram

SHEET SIZE
UNIT
NAME C1 BLOK

DOCUMENT ID
S-02460
JOB NUMBER DOCUMENT NUMBER REV

TABC 2015-01-06 09:09:30