Conversion of Forest Residues to a Methane-Rich Gas in a High-Throughput Casifier

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ABSTRACT

Research was conducted in a process research unit to develop an entrained bed gasifier which is supplied heat by recirculating a stream of sand between a separate combustion vessel and the gasifier. The char remaining after gasification of the wood provides the fuel for the combustor.

The research program was conducted in two phases. In the first phase, a 6 in. I.D. gasifier was used to establish the feasibility of the concept for a wide variety of biomass feeds. The second phase of the program was conducted with a 10 in. I.D. gasifier, and a fully automated feeder system, to evaluate gasifier performance at very high feed rates.

During the course of the program the following experimental results were obtained.

- It was demonstrated that the gasification system could handle any reasonable **form** of woody biomass with minimal preparation. For example, both hard and soft wood chips, bark, sawdust, coarsely shredded stump material, and whole tree chips (including leaves and stems) were successfully gasified in the system.
- The heating value (hhv) of the product gas was between 450 and 500 Btu/SCF and was independent of the moisture level of the feed. [Obviously the yield of product gas decreases as the moisture level increases because more heat (char) is required to evaporate the moisture leaving less heat (char) for gasification.]
- Woody biomass throughputs of 3400 lbs/ft²-hr were **achieved while** maintaining carbon conversions sufficient for thermally balanced operation.
- Long term operational stability was demonstrated in a 5 day test.
- The parameter having by far the greatest effect on carbon conversion was gasifier temperature. Other parameters such as wood throughput, gas velocity, even gas type did not have effects that were greater than the statistical scatter of the data. Wood type was found to have some influence especially in tests conducted in the 6 in. I.D. gasifier.
- Tar yields in the product gas were between 0.5 and 1.5 weight percent of the dry wood.

The experimental results were used to develop design parameters and detailed energy and material balances for a conceptual plant. A preliminary cost analysis is presented in the report based on the conceptual design.

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FINAL REPORT

on

CONVERSION OF FOREST RESIDUES TO A METHANE-RICH
GAS IN A HIGH-THROUGHPUT GASIFIER

to

PACIFIC NORTHWEST LABORATORY

March 2, 1988

INTRODUCTION

This report describes the status and presents performance data of a biomass gasification process developed by Battelle's Columbus Division with the support of the U.S. Department of Energy's Biomass Energy Technology Division. The process development effort was initiated in 1980 with the start-up of a process research unit (PRU) to determine the feasibility of producing medium-Btu gas from biomass without the use of oxygen.

There is a need for such a system because low-Btu gas requires derating of combustion equipment and is expensive to distribute even short distances. Most biomass plants will certainly be under 1000 tons/day (dry basis) which makes the unit cost of oxygen prohibitively expensive. Therefore the production of a medium-Btu gas based on steam/oxygen gasification, while technically feasible, was projected to be prohibitively expensive.

State-of-art biomass gasification systems based on coal gasification technology are designed to provide the substantially longer residence times required to convert relatively unreactive coal and to handle a much denser feed material than biomass. The approach taken in this program was to develop a reactor system to exploit the unique physical properties and high reactivity of biomass-type materials.

EXECUTIVE SUMMARY

The Battelle process employs an entrained gasifier which is supplied heat by recirculating a stream of sand between a separate combustion vessel and the gasifier. The char remaining after gasification of the wood provides the fuel for the combustor. In the commercial concept, both gasifier and combustor would operate in the entrained mode. However, the PRU utilized a conventional fluid-bed combustor. This was a sensible expedient for the PRU because it allowed the combustor to also serve as a surge vessel providing hot sand inventory to allow smooth circulation of the sand between the combustor and gasifier.

The PRU program was conducted in two phases. In the first phase, a 6 in. I.D. gasifier was employed. This phase established the feasibility of the concept and demonstrated the ability to handle a surprisingly wide variety of biomass feeds as received without preparation. For example, coarsely shredded bark was successfully gasified in spite of its fibrous stringy nature with some strips of bark greatly exceeding the diameter of the 6 in. I.D. gasifier. We were pleasantly surprised that this material could be gasified without any reactor bridging or plugging occurring. The tests in the 6 in. gasifier also established that sufficiently high conversion levels of biomass could be achieved in a simple entrained gasification reactor to allow the system to operate thermally balanced with no production of excess char. Elimination of excess char is easier with separate gasification and combustion reactors because, in a two reactor process, the combustor temperature can be maintained at a much higher level than is reasonable for an average gasifier temperature in a single reactor gasifier.

The second phase of the program was conducted with a 10 in. I.D. gasifier installed in place of the 6 in. unit and with the biomass feed system automated to allow operation at the very high rates demonstrated as being feasible during the first program phase.

During the course of the program the following experimental results were obtained.

- It was demonstrated that the gasification system could handle any reasonable form of woody biomass with minimal preparation.

For example, the following feeds were successfully gasified over the course of the program.

- Wood chips (both hard and soft wood)
 - Bark
 - Sawdust
 - Coarsely shredded stump material (residue from the solvent extraction of naval stores)
 - Whole tree chips including branch fragments and leaves
- The heating value (hhv) of the product gas was between 450 and 500 Btu/SCF and was independent of the moisture level of the feed. [Obviously the yield of product gas decreases as the moisture level increases because more heat (char) is required to evaporate the moisture leaving less heat (char) for gasification.]
 - Woody biomass throughputs of 3400 lbs/ft²-hr were achieved while maintaining carbon conversions sufficient for thermally balanced operation (62 percent).
 - Long term operational stability was demonstrated in a test of 5 days duration.
 - The parameter having by far the greatest effect on carbon conversion was gasifier temperature. Other parameters such as wood throughput, gas velocity, even gas type did not have effects that were greater than the statistical scatter of the data. Wood type was found to have some influence especially in tests conducted in the 6 in. I.D. gasifier.
 - Tar yields were between 0.5 and 1.5 weight percent of the dry wood. (In a commercial process this tar would be burned in the combustor to provide supplementary heat for gasification.)

In addition to the gasification tests, separate combustion tests were conducted in a circulating fluid-bed to determine emissions and design parameters for a circulating fluid-bed combustion system.

Over the course of the program, several cost assessments were conducted to assist in guiding development efforts as well as to provide economic data to potential users of the technology. The most significant of the cost assessments was described in a joint Battelle/Weyerhaeuser paper⁽¹⁾ evaluating the Battelle gasification process for cogeneration of electricity and steam in a typical pulp and paper plant. Part of this evaluation compared Battelle gasification with direct combustion and indicated a very significant cost advantage over direct combustion. For example, the total capital

requirements for installing a 570 TPD gasifier together with a natural gas boiler were estimated at \$15 million for the Battelle gasification system compared to \$27 million for a new combustion boiler. In addition to significantly reduced costs, the production of medium-Btu gas offers the following advantages compared to steam.

The ability to distribute the gas to various site locations to replace oil or gas.

- Utilization of combined cycle power generation which greatly increases the electricity generated per unit of fuel heating value.

PROCESS DESCRIPTION

This section briefly describes the commercial concept to provide a reference point for the experimental and process modeling efforts described in subsequent sections of this report. The integrated commercial concept also establishes the basic unit operations required to convert the incoming raw biomass into a cooled, clean, medium-Btu gas thereby defining the data required for the design of supporting unit operations.

The integrated commercial process will consist of the following systems and associated subsystems.

- Gasification System
 - Gasifier
 - Combustor
 - Solids circulation
 - Solids separation
 - Ash removal
- Biomass Handling/Feeding System
 - Wood drying
 - Wood feeding to gasifier
- Environmental Systems
 - Ash disposal

- Waste water treatment
- Waste Heat Recovery
 - Flue gas
 - Product gas
- Gas Clean-Up
 - Product gas scrubbing/**tar** removal
 - Flue gas

While it was not feasible in this program to generate data for any of the above subsystems except gasification, the data collected provides the designer information for the design of supporting subsystems. For example, data generated using physically diverse woody biomass and with widely varying moisture content provides a basis for the design of the wood handling and feeding system. Data on the tar and aqueous condensate allows the specification of environmental systems and gas clean-up systems. Heat and material balances over the integrated system based upon **PRU** data allow the effect of various heat recovery options on the system's **thermal** efficiency to be evaluated.

Conceptual System Description

The **Battelle** biomass gasification process **allows** the production of a medium-Btu product gas without the need for an oxygen plant. The process uses two physically separate reactors, a gasification reactor in which the biomass is converted into a medium-Btu gas and residual char which is then burned in a separate combustion reactor to provide the heat required for gasification. Heat transfer between zones is accomplished by circulating sand between the gasification and combustion reactors. Figure 1 is a simplified sketch illustrating the basic concept.

The concept itself is not new. For example, this basic approach has been applied to coal gasification by various organizations, including Battelle who circulated hot ash agglomerates between combustion and

FIGURE 1. SIMPLIFIED SCHEMATIC OF BATTELLE'S BIOMASS GASIFICATION SYSTEM

gasification zones. Consolidated Coal Company, who circulated calcined dolomite for heat transfer between combustion and gasification zones; and OGAS, who employed hot char transfer between combustion and gasification zones. (2) The Japanese have discussed large-scale demonstration of a two-zone process utilizing conventional fluid-beds to gasify municipal solid waste derived fuels (RDF) also employing circulating sand as the heat transfer media. (3)

EXPERIMENTAL PROGRAM

The major program effort was devoted to conducting the experiments necessary to establish the feasibility of converting biomass to a medium-Btu gas in an **integrated** Battelle process and generating the data required for scale-up and economic evaluation of the process. As will be described, the PRU employs a "bubbling" fluid-bed rather than a circulating fluid-bed combustor. Since the commercial concept will employ a circulating bed rather than a bubbling fluid-bed, separate combustion tests were conducted with wood char in a circulating-bed combustor to establish the design parameters for the commercial plant. Separate combustion tests also provided more information than could have been gotten if a circulating fluid-bed combustor were employed in the PRU because the PRU combustor required the use of natural gas as a supplementary fuel to compensate for the large heat losses inherent in the PRU.

Gasification

The gasification program was conducted in two phases. The first phase of the program utilized a 6 in. I.D. gasifier with essentially manual loading of the wood feed hopper. For the second phase of the program a new 10 in. I.D. gasifier was installed and the feed system was modified to allow completely automated feeding with a maximum capacity (based on wood chips) of approximately 20 to 25 TPD. The installation of the larger gasifier was necessary to provide more credible data for **scaling-up** to larger systems. Other modifications were made during the second program phase to reduce system heat losses such as relining of the combustor.

Gasification PRU Description

A schematic of the gasification PRU as modified to allow automated wood feeding is shown in Figure 2. The base of the gasifier vessel in the PRU is shown in Figure 3. The modified gasifier is constructed from a 10 in. Schedule 40 pipe and is constructed with multiple temperature and pressure/sampling ports through its length. A total of 12 temperature locations and 13 pressure taps are located in the gasifier vessel. The gasifier is surrounded by electric heating elements over the entire length to balance heat losses and then insulated on the outside of the heaters with ceramic fiber insulation. The gasifier has an active length of 22.7 ft, and is constructed with a "top hat" to minimize erosion at the exit of the reactor.

The PRU combustor is a 40 in. I.D. fluid bed reactor with an active height of $11\frac{1}{2}$ ft. The combustor is a refractory lined vessel with a total refractory thickness of 7 in. The refractory lining consists of a 4 in. cast refractory inner surface surrounded by 3 in. of refractory board insulation. This lining was designed to give a metal shell temperature of about 170 F to minimize heat losses from the vessel.

Sand enters the combustor through a 10 in. diameter line connected to the disengager. This line enters at a level of about 12 in. above the air distributor. Sand enters the L-valve at a level of about 24 in. above the distributor. A seal between the gasifier and combustor environments is provided by the fluidized sand bed in the combustor. The combustor cyclone solids discharge also is fed directly back into the fluid bed to reduce sand losses from the PRU system. A photograph of the modified PRU gasification system is shown in Figure 4.

The design philosophy applied to the PRU was to allow rapid heat-up and maximize the system's responsiveness to changes in operating parameters (especially temperature) so as to allow the maximum amount of data to be collected over any period of time. For example, the gasifier and all connecting piping were constructed without refractory linings to reduce heatup and cool down time as well as the time required to reach steady state. Obviously, this is a completely different approach than would be taken for the design of a commercial unit.

FIGURE 2. BATTELLE'S BIOMASS GASIFICATION PRU

FIGURE 3. BASE OF BATTELLE GASIFIER

FIGURE 4. BATTELLE'S BIOMASS GASIFICATION PRU

In the design of the PRU, consideration was given to the installation of a circulating fluid-bed as a sand heater. However, because of the low solids inventory in a circulating fluid-bed, a surge vessel is required to provide a hot sand inventory for stable flow through the L valve. Because of the high heat losses inherent in the PRU system, this surge vessel would have to be fired to insure sand hot enough to gasify the wood. Therefore, it was more expedient to use a conventional bubbling fluid-bed combustor in the PRU because it could function as a surge vessel as well as a combustor. The combustor diameter was selected to allow access for installing refractory. This resulted in a combustor I.D. approximately 40 in. which could supply heat in excess of that required for the 6 in. gasifier.

Hot sand circulation between gasifier and combustor is controlled through a conventional L-valve⁽⁴⁾ which utilizes a low flow of gas to aerate solids thus allowing flow. Solids circulation rates are controlled by varying the flow of L-valve gas. Sand was obtained in 100 lb bags and for most of the tests had a particle size range of 20 to 50 mesh.

The dimensions of the L-valve circulation system are shown in Figure 5. The L-valve piping was increased from 3 to 4 in. when the larger gasifier was installed. There is no way to directly measure solids circulation rates through the L-valve and the solids flow through the valve is adjusted to provide the desired temperature difference between the gasifier and combustor. Thus, the gasifier temperature as well as the temperature differential between gasifier and combustor is the parameter of concern and not the circulation rate of sand.

The wood feed system is shown schematically in Figure 6. It consists of a 15 ft diameter storage silo with provision for blowing ambient or heated air for in-situ wood drying. Wood is screw fed from the center of the floor of the silo into the inlet of a 10 HP blower which pneumatically transports the wood into the lock hopper assembly. The lock hopper assembly is installed on a load cell which provides a constant recording of the weight of the hopper assembly plus the contained wood. The transport of wood from the silo to the hopper assembly and the charging of the pressurized lock hopper is controlled electronically by sensing the wood level in the hopper. When operating at maximum wood rates, the lock hopper cycle times are on the order of 2 minutes. Depending on the bulk density of the feed, maximum wood

FIGURE 5. L-VALVE SYSTEM

FIGURE 6. WOOD FEED SYSTEM

rates are on the order of 1 ton/hour. The wood is fed into the bottom of the gasifier where it contacts the incoming hot sand and the wood/sand suspension is transported up through the gasifier. The rate of heat transfer between wood and sand is sufficiently high to gasify most of the wood by the time the char/sand suspension exits the gasifier.

The product gas enters a disengagement vessel after exiting the gasifier where char and sand are separated from the product gas. The char and sand flow from the disengager by gravity to the combustor. After separation of the sand and char, the product gas passes through an additional cyclone and is then cooled in a spray tower and scrubbed in a venturi. A side stream of the cooled clean gas is sampled for analyses and the remainder is burned in a flare. Samples of the condensate are collected periodically for analyses and the tar is separated from the aqueous phase to measure tar yields over the entire run period.

For the reasons described above, the combustor is a conventional bubbling fluid-bed. In addition to the char from the gasifier, natural gas is used as fuel. This is necessary because the PRU was designed to allow rapid attainment of steady-state and therefore has much higher heat losses than a unit designed to operate extended periods at as close to adiabatic conditions as possible. Utilization of an auxiliary fuel also provides maximum flexibility in terms of varying gasifier and combustor temperatures. The flue gas from the combustor passes through a cyclone where entrained sand is removed and returned to the combustor. After fine particulates are removed and the gas is cooled by scrubbing, the primary combustion products are monitored to guide control of the combustor.

To achieve the highest wood throughputs the system is capable of and to maximize wood conversion to gas, the limiting factor turned out to be the ability of the combustor to supply sufficient heat. As fuel to the combustor increases so does the combustion air requirement and resulting flue gas. The capacity limit of the fluid-bed combustor is determined by flue gas entraining excessive bed material. To gain additional combustor capacity, pure oxygen was blended with the combustion air to reduce the flue gas rate at any given combustor fuel input. By utilizing oxygen enrichment of the combustor air, it was possible to increase wood throughput to where it was limited by the wood feed system.

Gasifier Controls and Instrumentation

The gasification PRU is equipped with the following primary control systems which are listed below together with a brief description of their function.

- Wood Feed Rate. Obviously controlling the wood feed rate is of critical importance in generating reliable data. The automated system has already been described. The weight of the lock hopper assembly is constantly monitored by a load cell which records the weight on a strip chart recorder. This provides a continuous direct reading of the wood feed rate.
- Gasifier Inlet Gas Rate. Steam, together with nitrogen in various proportions, is used to provide mixing of the sand and wood and transport of the wood/sand suspension through the gasifier. The nitrogen is metered by conventional orifice plates and the orifice AP continually recorded. The nitrogen is preheated typically to 400 F before entering the gasifier. The steam rate is also measured by an orifice in the steam line downstream of the steam superheater. Both nitrogen and steam feed systems have the separate capacity to provide sufficient gas for entrainment of the wood/sand suspension.
- L Valve Sand Flow Rate. Varying sand flow through the L valve is the primary means of controlling the gasifier temperature. There is no means of directly measuring the actual sand flow rate. Instead, the gasifier temperature is used as the control point for sand flow. In the PRU, nitrogen is used as the L valve feed gas for expediency. Typically, a nitrogen rate of approximately 1 to 1.5 SCF/min is sufficient to create enough sand flow to provide a gasifier temperature of 1600 F when operating at a wood feed rate of 1000 lbs/hr (2000 lbs/hr-ft²). This flow is negligible compared to total product gas flow which typically is on the order of 235 SCF/min.
- Fuel and Air Rate to Combustor. Both char and natural gas are the fuels for the PRU fluid-bed combustor. The natural gas is used to compensate for the high heat losses inherent in the PRU and to provide maximum flexibility in allowing a wide range of carbon conversions to be achieved. The natural gas flow is measured by an orifice plate measurement. Combustion air is controlled by monitoring the oxygen in the flue gas. The inlet air is controlled to provide flue gas oxygen levels of about 1 to 2 volume percent.

In addition to the control systems listed above, the following measurements are made continuously over the course of an experiment.

- Product gas composition
- System temperatures
- System pressures.

The product gas rate is calculated based upon a nitrogen balance. The basic assumption is that the product gas composition is unaffected by any gas leaks. Periodically, organic tar condensate is collected, weighed, and stored for subsequent analyses. The water separated from the tar phase has been analyzed for selected runs to determine the waste water treatment required to make it suitable for discharge.

Pressure is measured at approximately 30 locations in the PRU and temperature is approximately 80 locations. The temperatures and pressures necessary for interpretation of the data are continuously recorded on strip charts in the control room. Near the end of the test program a computer data acquisition system was installed that is capable of collecting all of the process data and storing it on floppy disk. Other process measurements are made by digital indicators in the control panel.

The sample streams pass through a water cooled condenser to remove water, followed by a series of impingers placed in an ice bath. Gases then are filtered prior to analysis.

Product and flue gases are analyzed by a Carle Model 111 H Gas Chromatograph (GC) which provides a gas analysis every 20 minutes during tests. The GC is backed up by infrared analysis for methane and carbon monoxide in the product gas stream. Infrared instruments are Beckman Model 864 analyzers. Oxygen level in the flue gas is verified by a Beckman Model 755 paramagnetic oxygen analyzer.

Tar is separated from the quench water periodically and its weight is recorded. The tar samples always contain significant amounts of emulsified water (typically 50 percent or more) and therefore require distillation and ultimate analyses after the experiment before their contribution to the overall carbon balance can be made. Char cannot be sampled because it flows at high temperature directly from the disengager into the combustor.

Calculation of the char from an overall carbon balance around the combustor is not as accurate as the product gas carbon balance since, in most cases, natural gas is the major fuel input to the combustor. However, the most important performance parameter is the gas production which was most accurately determined. Details of the calculational procedure as well as a precision analyses of the data are discussed in Appendix B.

Typically, an experiment was completed in a single 10 hour shift with data collection periods ranging from 4 to 7 hours. However, two test periods were extended to five 24 hour days to evaluate long-term process stability.

Objectives of the Experimental Program

The objectives of the experimental program are the following:

- Demonstrate basic process operability and the stability of the system at steady-state conditions.
- Determine gas yields and composition.
- Identify any reaction byproducts particularly those that may influence the environmental impact of the process.
- Identify the parameters controlling gasification and quantify their effect.
- Determine gasifier performance, especially throughput.
- Determine the effect of various forms of woody biomass and confirm their ability to be used without special preparation.

The above objectives have been successfully achieved in the program.

Experimental Results and Analyses

The experimental program was conducted in two phases. The first phase was conducted with the 6 in. I.D. gasifier in place. The second program phase utilized a 10 in. I.D. gasifier and automated feed system.

The first program phase started with the gasifier using a proprietary Battelle development, the so-called multi-solid fluid-bed (MSFB).⁽⁵⁾ The

principle of the MSB is that very coarse particles that are ordinarily not able to be fluidized by gas can be fluidized in a flowing suspension of gas and finer solids. The coarse fluidized particles, result in a greatly increased residence time for the entrained solids. In addition, the entrained solids seem to penetrate bubbles thereby, creating smoother fluidization because "slugging" is minimized.

Until this program, experience with the MSB concept has been limited to combustion applications.. For combustion the MSB concept has been demonstrated to have many advantages⁽⁶⁾ and is the basis for a commercially available circulating fluid-bed combustor. However, during gasification rapid elutriation of the dense phase particles was caused by the high net generation of gas during gasification. (During combustion, oxygen is consumed and H₂O and CO₂ produced with no net generation of gas. In gasification, about three moles of gas are produced for each mole of conveying gas fed.) Because of the elutriation of dense phase particles, attempts to operate in the MSB mode were abandoned.

The high reactivity of biomass coupled with the high heat transfer rates existing in the gasifier, by virtue of the high interfacial area of the sand particles, resulted in suitable carbon conversions without operating in dense phase MSB mode. However, some tests were conducted with internals in the gasifier to determine whether these internals would influence gasifier performance. In the 6 in. gasifier the internals were found to have a significant effect on carbon conversion. However, in the second phase of the program conducted with a 10 in. I.D. gasifier, no detectable effect of the internals was found.

Except for a few tests in the second phase using internals, the second phase of the program was conducted with the gasifier operating in the entrained mode without internals or a dense phase fluid-bed.

As the program progressed, the available data was analyzed and correlated. These intermittent results are available in the form of reports presented at annual DOE Contractors' Meetings.⁽⁷⁻¹³⁾ This final report summarizes and correlates all available data.

Reactants and Reaction Products

Before discussing the detailed results of the experimental program, the distribution of reaction products generated by the rapid rate thermal decomposition of woody biomass as well as the woody biomass used in these tests are described.

The conversion of biomass can be represented by the simplified equation:

As will be seen, the relative yields of char and product gas depend almost entirely on temperature with other parameters having, at most, a secondary effect on the distribution between char and product gas.

Product Gas

The product gas composition remains constant regardless of the values selected for the operating parameters or feed gas. For example, when pure nitrogen was used as the feed gas rather than steam or **steam/nitrogen** mixtures, no significant changes in the heating value of the product gas (N_2 free) from the typical 450 Btu/SCF to over 500 Btu/SCF or in carbon conversion were measured. Thus, it appears that thermal decomposition governs the conversion of woody biomass with steam-carbon gasification reactions playing a negligible role. (These results suggest that recycled product gas could be considered as an alternative to steam for a commercial plant.) Product gas analyses for all the tests conducted are reported with the test results shown in Appendix A. A typical gas analyses is shown in Table 1. The high heating value is due to the relatively high concentrations of methane and ethane formed by the decomposition of wood. Unlike conventional single reactor gasification systems, the composition and heating value of the product gas is independent of feed moisture level. This is illustrated in Figure 7 where the dependency of heating value of the product gas on the moisture content of the wood feed is shown. Increases in moisture level increase the heat requirements for gasification

FIGURE 7. PRODUCT GAS HEATING VALUE DEPENDENCE ON WOOD MOISTURE

TABLE 1. TYPICAL PRODUCT GAS ANALYSIS - DRY BASIS

	Dry Gas Component (N ₂ Free) % Volume	Contributing to Heating Value
H ₂	17.5	56.9
CO	50.4	161.9
CO ₂	9.4	--
CH ₄	15.5	156.9
C ₂ H ₄	6.1	97.8
C ₂ H ₆	1.1	19.5
Calculated Heating Value		
By ASIM Method D3588		
493 Btu/SCF		

both in systems employing single gasifiers and the Battelle process. However, in the **Battelle** process the heat for gasification is generated in a separate reactor so the increased flue gas does not dilute the gasifier product gas whereas, in a single reactor gasifier, the increased combustion products from the increased char required to evaporate the additional moisture becomes part of the product gas thereby reducing its heating value.

Char

The residual char is an intermediate product that is completely consumed in the combustor to provide the gasification heat. Because of the integration of the gasifier and combustor and the high temperatures, it has been impossible to collect char samples. The char analyses are, therefore, estimated by calculating the carbon, hydrogen, and oxygen balances over the gasifier.

Organic Liquids

The organic liquid phase is a tar consisting of highly condensed aromatics containing virtually no functional groups such as OH groups⁽¹⁴⁾ and

is, thereby, essentially insoluble in water. Within our ability to measure them, the yield of condensed species did not seem to be affected by any of the controllable system parameters. Typically, the yield of organic liquids formed is approximately $\frac{1}{2}$ to 1 weight percent of the incoming dry wood. A surprising result shown in Figure 8 is that the yield of tar appeared independent of temperature over the temperature range covered in this program. Typical physical properties of the tar are summarized in Table 2 and a tar distillation curve is shown in Figure 9.

TABLE 2. TAR PHYSICAL PROPERTIES

Tar Viscosity	9800 CP 851 C
Density	1.14 g/ml 822 C
HHV (wet)	10,000 Btu/lb
Tar Yield	0.5 to 1% by weight of dry wood

The tar is very insoluble in water so that the waste water from the scrubber should be treatable in a conventional waste water treatment facility. The analyses of a typical sample of the waste water from the PRU scrubber is summarized below.

Waste Water:

pH:	6.0
BOD:	800 to 2000 mg/l
COD:	1500 to 5500 mg/l
Total Organic Carbon:	100 to 200 ppm

Woody Biomass Feedstocks

The wood types used over the course of the PRU program are summarized below:

- Mixed hardwood chips
- Cherry chips
- Mixed oak, birch, maple chips
- Pine chips

FIGURE 8. DEPENDENCE OF TAR YIELD ON GASIFIER TEMPERATURE

FIGURE 9. TAR DISTILLATION CURVE

- Hardwood bark
- Pine sawdust
- Red oak chips
- Birch and maple chips
- Swamp oak chips
- Gum chips
- Whole tree chips (hog fuel)
- Shredded stump (after resin extraction).

The material varied widely in physical form as the photographs in Figure 10, a-d indicate. All of the materials could be fed without preparation and without reactor plugging or bridging. In fact, the highly fibrous bark shown in Figure 10-d was gasified in the 6-in. I.D. gasifier in spite of some of the bark strands being almost 12-in. long. This is of prime economic importance because it minimizes front-end preparation costs and allows the user maximum flexibility in purchasing feed materials.

Approach to Data Correlation and Analyses

The purpose of the experimental program is to establish the relationship between the system design and operating parameters and system performance. The system operating parameters that can be independently controlled are listed in Table 3.

TABLE 3. PRU INDEPENDENT OPERATING PARAMETERS

<ul style="list-style-type: none"> • Wood Feed Rate • Wood Type and Moisture Level • Feed Gas Flow Rate and Composition • Sand Circulation Rate • Combustor Temperature = f (fuel rate)
--

For example, any (within reason) desired gasifier temperature can be achieved by varying the sand circulation rate and, within narrower bounds, the combustor temperature.

Hog Fuel

100 mm

(b) Hog Fuel

1 2 3 4 5

SAWDUST

(a) Sawdust

FIGURE 10. PHOTOGRAPHS OF WOODY BIOMASS USED IN THE TEST PROGRAM

HARDWOOD

(c) Hardwood

BARK

(d) Bark

In addition to the operating parameters the gasifier was also changed during the program. As indicated, the first phase of the program was conducted with a 6 in. I.D. gasifier with a 10 in. I.D. gasifier installed for the second phase of the program. Reactor internals included the dense phase particles evaluated in the initial testing of the MSB and baffles tested in both gasifiers.

The parameters determining the system performance are listed in Table 4.

TABLE 4. PRU PERFORMANCE PARAMETERS

<ul style="list-style-type: none"> • Carbon Conversion • Gas Yield and Composition • Gasifier Throughput • Organic Condensate Yield and Composition

In addition to the above measurable performance parameters, other more subjective measures of performance are equally important. These parameters include:

- System mechanical operability.
- System stability which is defined as the ability to reach and maintain a steady state without requiring constant adjustments of the operating parameters.

Presentation and analyses of the performance data will be done separately for the two program phases with the conclusions based upon consideration of all available data.

Phase 1 - Six In. I.D. Gasifier Tests

Complete run data for all gasification tests are tabulated in Appendix A. This data was reviewed to determine the operating parameters having a statistically significant effect on the above defined performance parameters. Appendix B contains the details of the computational methods and

how the various variables were statistically evaluated as to their influence on the system performance. In general, the data scatter in the 6 in. unit is considerably greater than in the 10 in. gasifier. It appears that the most likely cause for this scatter is in the measurement of the wood feed rate, which was continually monitored by a load cell in all tests in the 10 in. unit, and which was therefore more accurate.

Gasifier Temperature. Results of this analyses indicate that, with a given wood feed, gasifier temperature is the only parameter that significantly influences system performance and that the effect, if any, of other operating parameters is not detectable with the data scatter in the 6 in. unit.

Gasifier temperatures were ordinarily within a 20 F range from the bottom to the top of the gasifier. The temperature at the gasifier top was reduced due to losses from the unit. The gasifier temperature reported was an average of the temperatures over the relatively isothermal lower- to mid-zone of the gasifier.

All the carbon conversion to gas data collected in the 6 in. unit is plotted against gasifier temperature in Figure 11. This figure shows the data both with and without internal baffles. These data indicate that a family of curves may be drawn depending on whether baffles are used as well as on the type of wood used. The family of curves shows that baffles result in increased conversion and, that for any given gasifier temperature, conversion is higher for bark and sawdust than for wood chips. There does not seem to be any affect due to wood type, i.e., hardwood or softwood.

Gasifier Throughput. In the 6 in. unit, carbon conversion as well as other performance parameters were independent of gasifier throughput. This independence is shown in Figure 12 which plots carbon conversion and gasifier temperature against gasifier throughput. The fact that the conversion and temperature curves are reflections of each other indicates that any changes in conversion are due to temperature changes and not to throughput. The maximum throughput in the 6 in. unit was limited by the fact that wood was manually loaded into the hopper which required hoisting drums by crane. As Figure 12

FIGURE 11. SUMMARY OF CARBON CONVERSION DEPENDENCE ON TEMPERATURE - 6 IN. GASIFIER

FIGURE 12. CARBON CONVERSION IS INDEPENDENT OF THROUGHPUT - 6 IN. GASIFIER

indicates the highest feed rate (dry) was about 350 lbs/hr which results in a throughput of approximately 1800 lbs/hr-ft² in the 6 in. gasifier.

Feed Gas Flow Rate and Composition. Another variable that conceivably could influence performance was the gas inlet velocity. The reasoning behind this supposition is that particle residence time would be inversely proportional to the inlet gas velocity. To test this supposition the gasifier temperature and carbon conversion to gas are plotted against inlet velocity in Figure 13. As this data indicates, there is no apparent influence of inlet gas velocity on carbon conversion.

The role of gas composition as it affects either reaction with the carbonaceous char or vapor phase reactions was evaluated by utilizing nitrogen rather than steam in selected experiments. Obviously, if the reactions of steam with the carbonaceous solid is a factor or if vapor phase reactions such as hydrocarbon reforming, tar cracking, or water gas shift ($\text{CO} + \text{H}_2\text{O} = \text{CO}_2 + \text{H}_2$) are important, eliminating steam as a feed gas should have a significant influence, e.g., lower carbon conversion, higher tar yields, higher concentrations of hydrocarbon gases, or a higher CO/H_2 ratio. However, tests with pure nitrogen rather than steam indicated no changes in any of the aforementioned performance parameters.

Therefore, the conversion of woody biomass, at least over the range of carbon conversion attained, appears to occur completely through pyrolyses with no measurable contribution due to steam reactions with either the solid phase or vapor phase.

Phase 2 - Ten In. I.D. Gasifier Tests

The test program was continued with the installation of a larger gasifier to determine gasifier scaleability. Wood throughput in the 6 in. unit was limited by the ability to supply wood which was transported to the lock hopper by crane. Therefore together with the installation of the larger gasifier, the entire feed system was automated to allow wood feed rates as high as 20 TPD (dry) or as low as 2 TPD (dry) to be achieved.

Another modification made during the course of the Phase 2 program was to increase the heat generation capability of the integrated fluid-bed

FIGURE 13. CARBON CONVERSION IS INDEPENDENT OF INLET GAS VELOCITY - 6 IN. GASIFIER

combustor. This was necessary because it was found that as the throughput increased, the available heat supply could no longer maintain the desired gasifier temperature. The factor limiting the heat generation rate in the combustor was excessive elutriation of sand as the **fuel/air** input was increased. The modification included the addition of a cyclone return leg to the combustor to reduce the make-up sand and installing an oxygen feed system to provide oxygen enriched air thereby reducing the gas velocity required to burn a given amount of fuel. This reduced elutriation as **well** as the sensible heat removed in the flue gas.

Effects of Process Variables

The process variables evaluated during the Phase 2 program were the same as for the Phase 1 study. However, because **much** greater amounts of feed material were required, not as **many** different feeds were evaluated in the Phase 2 tests as in Phase 1. Also, with the exception of tests with baffles, results in the 6 and 10 in. gasifiers were in good agreement. Therefore, it was not felt necessary to completely repeat all of the feed materials tested in the 6 in. unit. The effects of these process variables in the 10 in. unit are summarized **below**.

Internal Baffles

Because reactor internals had a very significant effect in the 6 in. unit, they were tested in the 10 in. unit. However, in the 10 in. unit the baffles did not measurably increase carbon conversion compared to the empty gasifier. This is illustrated in Figure 14 where the points generated with baffles for both the 6 and 10 in. gasifier are compared with the least squares fit for both units.

These results were explained by speculating that geometry is important in determining the effectiveness of these internals. For example, the baffles create a turbulent zone of increased residence time in their vicinity. The temperature at this zone is probably the key factor in determining the effect on conversion. Because the placement and geometry effects of the baffles appeared to be reactor size related, it was decided

FIGURE 14. BAFFLES IN THE 10 IN. GASIFIER HAD NO INFLUENCE ON CONVERSION

that trying to scale-up to larger commercial sizes would require an experimental and modeling effort beyond the scope of the program. In addition, there would remain the risk that whatever scale-up criteria were developed between the 6 and 10 in. unit may not work for larger units. Therefore, the experiments with baffles were terminated.

Gasifier Temperature

The only gasifier parameter having any measurable effect on carbon conversion was the gasifier temperature. There was less scatter in the 10 in. data than in data from the 6 in. unit.

The dependence of carbon conversion on gasifier temperature is shown in Figure 15. By utilizing oxygen enriched air for certain tests in the 10 in. unit the attainable gasifier temperature range was extended to almost 1900 F. As Figure 15 indicates, the relationship between carbon conversion and gasifier temperature is linear over the entire temperature range.

Gasifier Throughput

Confirming the results in the 6 in. gasifier, gasifier throughput did not affect any of the gasifier performance measures. This is shown in Figure 16 which shows the same data as plotted in Figure 12 for the 6 in. unit. The fact that both temperature and conversion are essentially reflections of each other when plotted against wood feed rate indicates that conversion is a function of temperature and not throughput.

In the 6 in. gasifier, the throughput was limited by the ability to manually transport wood to the feed hopper. In the 10 in. unit the throughput is limited by the automated feed system which, in turn, is limited by the size of metering screws. At the higher feed rates, the system also becomes limited by heat supply in that, even with oxygen enrichment, it becomes impossible to generate and transfer sufficient heat to maintain gasifier temperatures above the desired 1500 F.

FIGURE 15. DEPENDENCY OF CARBON CONVERSION ON GASIFIER TEMPERATURE - 10 IN. GASIFIER

CARBON CONVERSION DEPENDS ON TEMPERATURE

FIGURE 16. CARBON CONVERSION IS INDEPENDENT OF THROUGHPUT - 10 IN. GASIFIER

In spite of the physical limitations, throughputs of over **3800 lbs/ft²-hr** based on dry wood were achieved in tests in the 10 in. unit while maintaining acceptable carbon conversion to gas.

System Mechanical Operability and Stability

As previously discussed, two of the most important performance parameters are the mechanical operability of the system and its stability. Stability is defined as the system's ability to produce a uniform product in spite of normal fluctuations in operating parameters and over extended periods of time. In order to quantify these somewhat subjective performance measures, extended operations were conducted. In addition to providing information on the above performance measures, the extended tests also allowed more accurate collection of the tars which, because of the relatively low yield, could not be accurately measured during the typical **4** hour data collection periods.

Though the PRU was not designed to demonstrate long term operability and the program was not funded to conduct round the clock operation, limited extended tests were conducted. The longest of these tests was conducted over a 5 day period operating **24** hours per day with wood feeding for over **80** hours, **72** hours of which were uninterrupted. (The interruption was required to repair a weld in the L-valve system.) This extended test demonstrated not only the system stability but also established that the system was easy to control and that flow through the L-valve remained steady (as evidenced by constant gasifier temperature) once set.

EXPERIMENTAL PROGRAM - COMBUSTION

Although the gasifier is the prime reactor in the Battelle gasification system, the char combustor is also critical to operation of the process. The combustor must be capable of (1) completely burning the char produced during gasification, (2) reheating the circulating sand phase, and (3) providing for smooth recirculation of sand between the gasifier and combustor. A fluid bed combustor seems, at first glance, to provide these capabilities. The high gasifier throughputs allow very compact gasifiers,

but if a conventional fluid-bed combustor were to be used, there would be a large difference in the relative sizes of the gasifier and combustor.

For example, the relative diameters of the Battelle gasifier and a conventional fluid bed combustor are shown in Figure 17. The figure also shows a Battelle gasifier coupled to a high throughput circulating type fluid-bed, which is capable of operating at the high throughputs needed to minimize the difference in vessel diameter between the gasifier and combustor. Because the closer matching of vessel diameters should simplify solids distribution in the combustor by greatly reducing the number of sand inlets required, a separate test series was conducted burning char in a circulating fluid-bed.

Circulating Fluid-Bed Combustor

A circulating fluid bed combustion PRU was used for the combustion tests. A schematic of the combustion PRU is shown in Figure 18. It consists of a 6 in. I.D. combustor with the circulating solids separated from the flue gas by a conventional cyclone. From the cyclone a portion of the solids can be directly returned to the gasifier through a "hot" solids recycle loop. Combustion temperature is controlled by a fluid-bed external heat exchanger which cools the circulating solids prior to returning the cooled solids to the combustor through a "cold" solids recycle loop. The relative solids flow through "hot" and "cold" loops are regulated by separate L valves.

In the case of wood char feed, no hot recycle was required to achieve complete carbon utilization. The flue gas is monitored for CO and NO, in addition to the other combustion products. Additional cyclones remove finer particulates. In the case of wood char combustion there was almost complete separation of sand and wood ash due to the very fine size of the fly ash compared to the circulating sand.

Objectives of the Experimental Program

The specific objectives of the circulating combustion tests were the following.

FIGURE 17. RELATIVE SIZES OF CONVENTIONAL AND CIRCULATING FLUID-BED COMBUSTORS TO INTEGRATE WITH GASIFIER

FIGURE 18. CIRCULATING FLUID-BED PILOT COMBUSTOR SCHEMATIC

- Demonstrate the combustion efficiency of wood char in a high throughput combustor.
Determine the effect (if any) of wood ash on combustion temperature limits.
- Provide data for design of commercial scale equipment.
- Determine expected emissions from the combustor.
- Determine if condensed organic material from the product gas stream can be efficiently and cleanly burned.

Experimental Results and Analyses

A wood derived charcoal was selected as the available fuel most closely resembling the char produced during combustion. An analyses of the char is shown in Table 5 below.

TABLE 5. COMBUSTION FEEDSTOCK ANALYSIS

Composition (%Dry Basis)	Char
Carbon	76.99
Hydrogen	2.76
Nitrogen	0.50
Chlorine	0.01
Sulfur	0.06
Ash	7.61
Oxygen (difference)	12.07
Moisture, %	5.33
HHV, Btu/lb (dry)	12,252

Results from the circulating fluid-bed combustion tests are summarized in Tables 6 and 7. These results are anticipated to be conservative compared to a commercial system because the char in a commercial system will enter the combustor preheated to gasifier temperature while, in the PRU tests reported here, the char was fed at ambient temperature. In addition, to the tests with char, a single test was conducted with bark. Considering that

TABLE 6. MSFB COMBUSTION TEST SUMMARY

Run No.	Feed Type	Moisture (%)	Throughput (lb/hr-ft ²)	Combustor Temp. (F)	Excess Air (%)	Carbon Utilization (%)	Nitrogen Oxides ppm	lb/MM (Btu)	Carbon Monoxide (ppm)	lb/MM (Btu)
3A	Char	5.3	228	1800	48.2	94.6	106	0.19	80	0.09
3B	Char	5.3	230	1850	11.5	94.9	175	0.24	73	0.06
4	Char	5.3	262	1900	19.6	98.8	160	0.22	60	0.05
5	Char	5.3	230	1825	10.7	102.0	208	0.28	61	0.05
6	Bark	13.2	370	1600	13.4	80.3	153	0.20	97	0.08
7	Ash Doped Char	4.4	287	1750	2.8	86.4	242	0.29	231	0.17

TABLE 7. COMBUSTION TESTING PHASE TWO

TEST NUMBER	2-3	2-4	2-8	2-8*	2-9A*	2-9B*	2-10	2-11	2-12	2-13	2-14*
FEED MATERIAL	CHAR	CHAR	CHA	CHAR	CHAR	CHAR	CHAR	CH+TAR	CH+TAR	CH+TAR	CHAR
CHAR RATE LB/HR	30	55	57.5	55.8	56.4	57.2	55	47	47	46.5	52
TAR RATE LB/HR	0	0	0	0	0	0	0	8	5.7	6	0
TEMPERATURE - F	1876	1597	1820	1817	1700	1770	1715	1900	1890	1850	1700
CHAR THROUGHPUT	150	275	287.5	.279	282	286	275	235	235	232.5	260
EXCESS OXYGEN %	1	3.2	.4	.48	1.5	1.1	2.45	1.9	3.1	2.2	2.7
FLUE GAS CO2 %	16	14.6	14.6	14.5	14.5	15	15.8	16	15.3	14.9	15
CO EMISSION, PPM	150	150	75	54	110	65	95	55	20	50	70
CO EMISSION LB/MMBTU	.13	.14	.07	.05	.10	.06	.08	.05	.02	.05	.06
NOX EMISSION, PPM	450	370	N/A	N/A	N/A	N/A	N/A	N/A	N/A	550	620
NOX EMISSION LB/MMBT	.21	.23	NONE	NONE	NONE	NONE	NONE	NONE	NONE	.06	.11
BED TYPE	DEN.BED	BAFF.	NONE	NONE	NONE	NONE	NONE	NONE	NONE	NONE	NONE

* SAND CIRCULATION TESTS - SHORT DATA PERIOD

this test was conducted at a temperature significantly lower than the char tests, it seems to fit the trend of carbon utilization as a function of temperature exhibited by the char. The influence of combustor temperature on carbon utilization is shown in Figure 19. These data indicate that 1800 F or higher combustor temperatures should provide complete carbon utilization.

When combusting low volatile or high moisture content fuels, one method of achieving higher conversion levels is to increase the amount of excess air. This approach is typically used in conventional wood fired boilers to help compensate for the high wood moisture content. It was initially felt that the low volatile content of the gasified char might require high excess air levels in the combustor. Data generated during the combustion tests showed that excess air levels as low as 2 percent were sufficient to achieve the high levels of carbon utilization measured. As a result, a design level of 10 percent excess air is felt to be adequate for complete combustion of the char produced during gasification.

In one test, additional wood ash was added to increase the ash level to 20 percent of the char. This test determined whether increased ash would create any slagging problems. By doping the char with wood ash, it was felt that the build-up of wood ash could be simulated.

The test with the ash doped char indicated that ash agglomeration should not be a problem. Not only was no evidence of slagging observed, but the fine particle size of the ash compared to the sand caused rapid removal of ash from the system by elutriation. Therefore, the conceptual commercial design has two cyclones downstream of the combustor. The first cyclone will separate primarily sand for recirculation to the gasifier while the second more efficient cyclone will remove primarily wood ash as well as attrited sand.

In addition to the char, the combustor will also be used to burn the tar byproduct produced by gasification. Therefore tests were conducted with char + tar to determine whether the tar could be completely burned. In addition, to the combustion with char, separate combustion tests were conducted with the tar to measure polyaromatic hydrocarbons (PAH) to allow comparison, for example, with typical PAH emissions from the combustion of natural gas. Using the CO concentration as an indication of completeness of combustion, the tar/char blends were found to burn as efficiently as the tar alone.

FIGURE 19. EFFECT OF COMBUSTOR TEMPERATURE ON CARBON UTILIZATION

Therefore, injection of the tar into the combustor is an effective and efficient means of tar disposal.

Combustion Emissions

During circulating fluid-bed combustion tests, CO and NO_x were monitored to determine whether emissions would conform to new source performance standards (NSPS) for stationary sources. In addition, some separate burner tests were made with the byproduct tar to determine PAH formation.

The available NO_x emission data is summarized in Figure 20 which also shows the NSPS for NO_x. As Figure 20 indicates the combustor operating with char and byproduct tar should be well within NSPS limits.

The levels of CO in the gas ranged from 0.02 to 0.2 lbs/MM Btu with no apparent dependence on temperature or excess oxygen level. Therefore, no problems are anticipated in achieving acceptable levels of CO at the combustor conditions projected for commercial plants.

The separate burner tests conducted on the tar to determine PAH formation also indicated that the tar should burn as cleanly as natural gas with respect to PAH formation. This test was conducted passing the combustion products through a modified EPA Method 5 train and analyzing the extract for various PAH constituents using high resolution gas chromatography and mass spectrometer techniques. Figure 21 is a schematic of the sampling system. Results of this test are summarized in Table 8 where the results from burning natural gas in the same apparatus are also shown. This result indicates that the tar should burn cleanly and without PAH emissions.

Combustion Test Summary

The combustion tests indicated that char and the resulting tar can be burned completely in a circulating fluid-bed with emissions well within NSPS. The combustor throughput used for the conceptual design of a commercial forest residue gasifier (FRG) plant 150 lbs/ft²-hr is very conservative compared to the throughputs used in the experimental study which approached 400 lbs/ft²-hr. Excess oxygen levels could be maintained low 2 percent while still

6 INCH COMBUSTOR

FIGURE 20. NO_x EMISSIONS FROM COMBUSTION TESTING

FIGURE 21. MODIFIED METHOD 5 SAMPLING SYSTEM

TABLE 8. HRGC/MS DATA FOR MODIFIED METHOD 5 SAMPLES

Compound	Total Amount Found in the Extract, μg	
	Char + Tar Combustion	Natural Gas Combustion
Naphthalene	6.4	10
Phenanthracene	5.8	19
Anthracene	0.2	1
Fluoranthene	1.1	26
Pyrene	0.5	31
Benzo[a]anthracene	0.03	4
Chrysene	0.1	5
Cyclopenta[a,d]pyrene	ND ⁽¹⁾	
Benzo[k]fluoranthene	0.05	1
Benzo[e]pyrene	ND	0.6
Benzo[a]pyrene	ND	0.7
Benzo[g,h,i]perylene	ND	
Indeno[1,2,3-c,d]pyrene	ND	
Dibenzo[a,h]anthracene	ND	
Coronene	ND	

(1) Not detected.

achieving complete combustion and acceptable CO and NO_x concentrations in the product gas.

CONCEPTUAL SYSTEM DESIGN/ECONOMICS

An important part of the program is to utilize the experimental data for the conceptual design of integrated gasification plants. The resulting conceptual designs provide the basis for the specification of supporting unit operations which then allow the determination of process economics. At this stage of development, the primary purpose of the process economics is to determine whether the project can be commercially competitive with other approaches to generating energy from biomass. In addition, the conceptual design exercise helps to guide the experimental program and to establish the uncertainties in scaling-up the process. The first step in this exercise was the development of a flowsheet for the integrated process.

Process Flowsheet

A simplified schematic flowsheet for the integrated process is shown in Figure 22. The flowsheet shows a plant that will produce a cooled clean medium-Btu (400–500 Btu/SCF) gas from woody biomass. It therefore not only includes the gasification system but also the following supporting unit operations.

- Wood receiving and handling
Wood drying
- Wood feeding
- Waste heat recovery from the product gas and the flue gas
- Product gas quenching and scrubbing.

Wood is received in the form of whole tree chips, shredded bark, sawdust, or whatever form of wood is available. The wood is screened to remove tramp material and then dried using waste heat in the flue gas from the process combustor. From the drier, wood goes to a storage silo from which it is fed

FIGURE 22. SCHEMATIC OF BIOMASS GASIFICATION SYSTEM

into the gasification system either via a lock hopper and metering bin containing a screw feeder or with a screw extruder. (Typical system pressures will only be a few psig.) From the metering bin, the wood is conveyed to the base of the gasifier where it is mixed with incoming hot sand from the combustor. The mixture of wood and sand is entrained through the gasifier by steam injected at the base of the gasifier.

In the gasifier, the wood is converted into primarily a medium-Btu product gas, which is separated from the circulating sand/char suspension in a cyclone. The hot product gas will typically be used to preheat combustion air. Because of the potential for tar condensation, heat is recovered only to a product gas temperature of about 750 F. The product gas is then scrubbed to remove any remaining particulate material as well as the small amount of tar formed in the gasifier. Scrubber water is recycled with excess water going to a waste water treatment facility. Tar separated from the scrubber is injected into the combustor for disposal.

The sand/char suspension is fed from the cyclone into the base of a high-throughput circulating bed combustor where the char is burned together with the separated tar to provide the heat for gasification. The hot sand and wood ash are removed from the combustor by entrainment and passed through a first cyclone where primarily the relatively coarse sand is removed. The hot sand is recirculated to the gasifier through a non-mechanical "L-valve" which controls the circulation rate of the hot sand. The circulation rate is adjusted to provide the desired gasifier temperature. After separation of the coarser sand, the much finer wood ash is removed from the system in a second cyclone.

The hot flue gas first preheats combustion air and then dries incoming wood before final particulate removal by a cyclone and baghouse and venting to the atmosphere.

Typically, temperatures in the gasifier will be on the order of 1600 F and the combustor will operate at temperatures in the range of 1900 F. In many cases, there is sufficient waste heat to generate some export steam in addition to the steam required by the gasifier. The overall system thermal efficiency depends on such factors as the raw wood moisture level, limitations on air preheat temperatures, which is based on metal-wall temperature considerations, plant size which determines heat losses, gasifier throughput

and other factors. The interrelationships between these parameters was determined by a heat and material balance model developed for the integrated system. As much as possible, the model is based upon experimental data from the PRU.

Heat and Material Balance Model

The computer model to simulate performance of a commercial Battelle gasification system consists of a complete mass and energy balance including calculated system heat losses for each major process vessel as well as process piping. Flexibility is built into the model to allow the effect of operating and design parameters on system performance and operation to be estimated. In addition to helping with process conceptual designs, the model has been used to determine the critical experimental parameters for PRU tests as well as to help interpret the results of PRU tests. The model allows the calculation of thermal performance in terms of controllable process and design parameters, as well as process parameters such as temperatures and flow rates.

The model predicts performance as a function of the following variables:

- Plant size
- Wood type
- Wood analysis
- Wood heating value
- Feed moisture
- Gasifier temperature, or
- Carbon conversion to gas
- Gas composition
- Carbon conversion to liquid products
- Heat loss from plant
- Turndown
- Steam generation efficiency.

In calculating a system mass balance, experimental data generated in the PRU is used to predict the elemental balance around the gasifier and

generate values for the gasifier exit streams such as the product gas, char and condensible organics. In combination with the energy balance, the relative mass flows are balanced so that the char sent to the combustor is enough to satisfy gasification heat requirements with no excess char or heat production. The gasifier is supplied with heat from the combustor which generates the heat from burning the char from the gasified biomass. Therefore the factors that influence the heat balance between the gasifier and the combustor are critical to the process. The primary factors that influence this thermal balance are the reactivity of the wood and the heat requirements for gasification.

The reactivity of the wood determines the gasifier temperatures required to achieve various conversion levels. For example, if the reactivity of the wood were zero, no wood would be converted in the gasifier and therefore the only heat absorbed would be sensible energy. This would result in temperatures in the combustor approaching adiabatic flame temperatures and the gasifier would function primarily as a preheater for the combustor. On the other hand, with high reactivities, the conversion level in the gasifier is limited by the available heat. Since the only heat available is from the char, the system will be in balance at the gasifier temperature at which the char remaining after gasification is sufficient to supply the heat to maintain the gasifier temperature.

For example, if the gasifier temperature drops, more char is produced which means additional heat is generated which will tend to increase the temperature back to where the system is in balance. If the gasifier temperature increases, more biomass is converted which results in less char and less heat which reduces the temperature to where the char is once again sufficient to provide the required heat. Thus, the system is self stabilizing which has been demonstrated during PRU operations.

Of course, if additional fuel is burned in the combustor, higher gasifier conversions can be achieved. For example, in the PRU tests where methane was used as an auxiliary fuel, carbon conversions in excess of 90 percent were achieved.

The heat requirements for gasification determine the distribution between char and product gas that can be produced. These heat requirements depend on the product distribution from gasification. Fortunately, the

relatively high concentration of gaseous hydrocarbons and the low CO/H_2 ratio reduce gasifier heat requirements allowing higher conversions to gas to be achieved than if a gas having a more favorable composition for synthesis would be produced.

With mass flows determined, projected vessel sizes are calculated using experimental data to estimate gasifier and combustor throughputs and conventional design methods for other process vessels such as cyclones. These vessel sizes are then used as inputs for determining heat losses from the system.

In predicting "balanced operation", the model adjusts the mass balance so as to achieve **thermally** balanced operation. That is, each section of the plant is in thermal equilibrium, where heat inputs equal outputs (including heat loss) at steady state. Additionally, the drying of the wood is adjusted so that the flue gas dries the wood as close to 10 percent moisture as possible.

The model output provides a process summary as well as details of each of the process streams. The output includes:

- Production rates of product gas and steam,
- Thermal and conversion efficiencies,
- Detailed heat losses,
- A total mass and energy balance,
- Detailed mass energy data for each stream, and
- Combustor conditions.

Table 9 presents a typical output. The case listed is for a 200 TPD (based on dry wood) plant using softwood as the feedstock and steam as the conveying medium in the gasifier.

Since heat losses are expected to increase as plant size decreases, heat loss sensitivity to plant size was examined. Figure 23 is a graph of heat loss versus plant size for a wood throughput of 2000 lbs/hr-ft^2 . This graph illustrates that heat losses rapidly increase as the size of plant decreases.

The model was also used to evaluate the effect of wood throughput on cold gas efficiency for a gasifier and combustor of fixed dimensions. For

TABLE 9. MASS AND ENERGY BALANCE SUMMARY FOR FOREST RESIDUE GASIFICATION MODEL

CASE • 1			
PLANT SIZE (TPD)		200.0	SOFTWOOD
DRY FEED RATE (LB/HR)		16666.67	
FEED MOISTURE CONTENT (%)		10.0	
STEAM RATE (LB/HR)		5186.23	
PURGE GAS FEED (LB/HR)		85.9	
CARBON CONVERSION TO GAS (%)		66.6	
PROD GAS RATE (SCF/HR)		211894.5	
ENERGY PRODUCT (MM-BTU/DAY)		2381.3	
PRODUCT HHV (BTU/SCF)		468.3	
COLD GAS EFFICIENCY (%)		68.9	
EXPORT ENERGY (MM-BTU/HR)		6.81	
(LB-STEAM/HR 85% EFF)		4927.3	
TOTAL HEAT LOSS (MM-BTU/HR)		9.36	
PLANT/HT EXC (% OF FEED HHV)		4.8 / 1.7	
GASIFIER INLET VELOCITY (FT/SEC)		10.30	
COMBUSTOR GAS VELOCITY (FT/SEC)		20.92	
GASIFIER TEMPERATURE (F)		1510.0	
COMBUSTOR TEMPERATURE (F)		1810.0	
COMBUSTION DATA			
AIR REQ (LB/HR)		43255.4	
INLET AIR TEMP (F)		77.0	
FLUE GAS RATE (SCF/HR)		588024.8	
SAND RATE (LB/HR-FT ²)		32900.2	
(GASIFIER)			
OVERALL ENERGY BALANCE (MMBTU/HR)			
IN		OUT	
WOOD	144.84	PROD GAS	121.19
STEAM	7.23	FLUE GAS	24.39
AIR	0.01	CONDENSIBLE	1.41
		DIRECT LOAD	4.77
TOTAL	152.13		151.76
HEAT LOSS SUMMARY			
		MMBTU/HR	% OF WHV
GASIFIER		0.441	0.31
COMBUSTOR		0.720	0.50
GASIFIER CYCLONES		0.134	0.09
COMBUSTOR CYCLONES		0.848	0.59
SURGE POTS		0.554	0.38
L VALVES		0.141	0.10
PIPING		4.018	2.79
SUM		6.855	4.76
(HEAT EXCHANGE)		2.500	1.74
TOTAL		9.356	6.49
ELEMENT BAL-GASIFIER (LBS/HR)			
IN		OUT	
C	8511.67	8511.67	
H	1823.68	1823.68	
N	110.93	110.93	
ASH	13799.41	13799.41	
	121.67	131.67	
	13.33	13.33	
TOTAL	23790.58	23790.68	
VESSEL SIZES (FT)			
		ID	OD HEIGHT
GASIFIER		3.3	4.3 45.9
COMBUSTOR		5.6	6.7 45.3
GASIFIER CYCLONES		2.4	3.0 9.3
COMBUSTOR CYCLONES		4.4	4.9 17.5
SURGE POTS		6.7	7.8 13.3
L VALVES		0.6	1.3 15.0
DETAILED MASS AND ENERGY BALANCE			
=====			
NET PRODUCT GAS			
IN		LB/HR	MM-BTU/HR
WOOD	16666.7	144.8	
MOISTURE	1851.3	0.2	
STEAM	5186.2	7.3	
NITROGEN	85.9	0.0	
TOTAL	23790.7	152.1	
GAS		LBMOLE/HR	VOLE % MM-BTU/HR
H ₂	115.9	11.6	15.4
CO	257.2	25.7	34.1
CO ₂	61.3	6.2	1.0
CH ₄	87.2	8.7	35.1
C ₂ H ₄	29.1	2.9	
C ₂ H ₆	3.3	0.4	18.4
H ₂ O	443.2	44.2	14.4
N ₂	4.0	0.4	0.0
TOTAL	1002.3	100.0	121.2
CONDENSIBLES			
OUT		LB/HR	MM-BTU/HR
C	2731.79	40.3	
H	269.00	12.3	
ASH	145.00	0.1	
TOTAL	3136.00	53.9	
AIR REQUIREMENT			
O ₂		LBMOLE/HR	MM-BTU/HR
	313.5	0.00	
N ₂	1186.5	0.00	
TOTAL	1500.0	0.01	
FLUE GAS			
O ₂		LBMOLE/HR	MOLE % MM-BTU/HR
	231.8	14.9	4.8
H ₂ O	104.5	6.7	3.7
O ₂	28.6	1.8	0.4
N ₂	1186.5	76.5	15.4
TOTAL	1551.5	100.0	24.3
ASH (LB/HR)	171.7		0.1
SUM (LB/HR)	46391.4		24.4

FIGURE 23. HEAT LOSS INCREASES SIGNIFICANTLY IN SMALLER PLANTS

this exercise, a nominal 10 ft gasifier was selected as the largest practical size in order to minimize the influence of heat losses. Two modes of operation were examined for their effect on the system cold gas efficiency. The first mode involved entrained operation of gasifier and combustor with gas velocities in both vessels at least 10 ft/sec. The second mode of operation involved operation of both gasifier and combustor at inlet velocities of about 2 ft/sec, more typical of conventional fluid-beds.

Figure 24 thus compares the influence of throughput (at constant reactor size) on cold gas efficiency for two-reactor gasification systems operating in the conventional fluid-bed mode as well as the entrained mode. As Figure 24 indicates, throughput has a significant impact on cold gas efficiency and that, below throughputs of about 1500 lbs/ft²-hr, the decrease in cold gas efficiency becomes very significant. The two-reactor fluid-bed and entrained modes are discontinuous because more sensible heat is removed in both product and flue gas phases in the entrained mode than in the fluid-bed mode. However, this distinction is relatively academic because atmospheric pressure fluid-bed gasification systems typically operate in the neighborhood of 200 lbs/ft²-hr. Therefore, throughput affects process economics by its influence on overall thermal efficiency as well as its influence on capital investment, and the ability to produce modular systems.

Process Economics

The most recent and probably the most accurate published cost estimate of producing medium-Btu gas using the Battelle process was described in a joint study by Weyerhaeuser and Battelle presented at an AIChE meeting. (1) Capital and operating costs for this study were estimated by Weyerhaeuser using process flowsheets and heat and material balances for the FRG process provided by Battelle.

The setting for this study was a "typical" pulp and paper mill producing 800 TPD pulp. This study examined the use of the FRG process to produce a medium-Btu gas to replace oil and natural gas and to generate electricity with a gas combustion turbine. A wood gasification plant with a capacity of approximately 1000 dry TPD was required to generate all of the plant's steam demand (225,000 lbs/hr at 850 psig and 825 F) from the gas

FIGURE 24. COLD GAS EFFICIENCY VERSUS THROUGHPUT FOR FLUID AND ENTRAINED GASIFIERS

combustion turbine exhaust. The plant energy utilization before and after installation of the gasifier is summarized in Table 10.

The installed equipment costs for this plant are summarized in Table 11.

The installed equipment cost for this 1000 TPD gasification facility is approximately \$13.4 million.

The components of the operating cost and the estimated requirements for each of these operating cost components are summarized in Table 12. Using this table enables one to estimate annual operating costs for any given sized plant by selecting labor rates, wood prices, utility prices, and the prices of sand and water.

The largest single operating expense is the cost of wood. These estimates allow one to determine the price of natural gas or distillate oil at which one would invest in a FRG gasification system. However, the investment decision will also be affected by special tax credits and possibly local PURPA regulations determining the price of power generated from renewable resources. A discussion of the influence of these various incentives for using renewables is beyond the scope of this report.

Cost Comparison with Direct Combustion

Perhaps a more meaningful evaluation of FRG economics is to compare it with direct combustion. Significant investments in the direct combustion of biomass are currently being made. Therefore, an important question is, assuming FRG technology were commercially available, could it compete with combustion? This comparison was made in Reference 1 with the results summarized in Table 13.

The first comparison in Table 13 is of a FRG system producing a cooled clean MBG to generate the pulp mills steam needs. To provide these steam requirements (225,000 lbs/hr) would require a FRG plant with a wood capacity of approximately 570 TPD (dry).

In the second case, the FRG facility is expanded to provide enough additional capacity to provide MBG for the lime kilns in addition to providing the plant's steam needs. In the first case, the FRG plant including a natural gas boiler to generate the steam would cost less than 60 percent of the cost

**TABLE 10. ENERGY UTILIZATION -- BATTELLE GASIFIER/GAS
TURBINE RETROFIT 800 TPD KRAFT MILL**

	Average Mill Conditions	
	Before Retrofit	After Retrofit
Steam Generation - lbs/hr		
Recovery Boiler	450,000	450,000
Combination Boiler	225,000	--
Waste Heat Boiler - 850 psig	--	225,000
Waste Heat Boiler - 150 psig	--	99,000(a)
Power Generation - MW		
Steam Turbine	23	23
Gas Turbine	0	39
Condensing Turbine	--	7
Fossil Fuel Use - MMBtu/hr		
Lime Kiln	96	--
Combination Boiler	130	--
Power Consumption - MW	26	32
Wood Requirements - ton/hr	14.4	42.9
Net Power Production Increase*		
		37 MW

(a) **Used** in new condensing turbine to generate additional 7 MW of electricity.

TABLE 11. INSTALLED EQUIPMENT COST FOR A 1000 TPD (DRY)
FRG FACILITY

	<u>\$ X 1000</u>
SITE	160
(Road, drainage, lighting)	
WOOD HANDLING / RETREIVAL	2880
(Truck dump, screen, wet storage)	
WOOD DRYING / RETREIVAL	2591
(Dryer, ducts, storage, conveying)	
GASIFICATION	4274
Surge bin	60
Feeder System	944
Gasifier	224
Gasifier Cyclone	30
Surge	88
Combustor	252
Combustor Cyclones	100
Surge	88
Startup Burner	50
Sand Hopper	15
Heat Recovery	648
Ash Removal / Recovery	90
Air Blower	30
Refractory Lined Pipe	163
Scrubber System	720
Foundations & Structures	650
Gas Oucting	72
Gas Booster Blower	50
ELECTRICAL & INSTRUMENTS	3173
PIPING	125
(Including fire protection)	
OTHER	225
(Including waste water treatment, painting)	
TOTAL INSTALLED EQUIPMENT COST	13428

TABLE 12. ELEMENTS OF OPERATING COST
(Battelle Wood Gasification Process)

Element	Basis
Wood	Approximately 100 MM Btu/hr MBG Output from 200 TPD (Dry)
Water	1000 gal/hr
Sand Makeup	0.015 lb/lb Wood (0.1% Loss)
Power	3.5 kW/daily ton Wood
Operating Labor	1.5 man/shift
Maintenance Labor	1 man/shift
Operating Supplies	30% of Operating Labor
Administration & Management	2.4% of Inst. Equipment
Taxes & Insurance	

of a conventional combination boiler. Even with an expansion in capacity to the 740 TPD required to provide gas for the lime kiln in addition to the steam, the total capital required for a FRG facility is less than 65 percent of that required for a combination boiler sufficient to provide only the plant's steam needs.

PROCESS SCALE-UP CONSIDERATIONS

Experimentation in the PRU facility determined the conditions necessary to produce a medium-Btu gas at carbon conversions sufficiently high to eliminate the production of excess char. In addition, the PRU has been operated for extended periods of time (days), demonstrating smooth stable operation with no variation in gasifier performance. The remaining parameters that must be resolved by scale-up are discussed below. They include the determination of: thermal performance, solids mixing, and system availability/reliability. In addition, other performance parameters such as biomass throughput should be confirmed in a larger-scale system having the same design and components as a commercial system.

TABLE 13. WHY NOT BURN WOOD DIRECTLY?

	Battelle Gasifier Gas Boiler	New Combination Boiler
Steam Output, lb/hr	225,000	225,000
Wood Input, TPD	570	500*
Installed Equipment Cost, \$ x 10 ⁶	9.6	20.6
Total Capital Required, \$ x 10 ⁶	15.2	27.0
<u>Lime Kiln Fuel Replacement</u>		
Oil/Gas Usage, Btu/hr	0	96
Wood Input, TPD	740	500*
Installed Equipment Cost, \$ x 10 ⁶	12.5	20.6
Total Capital Required, \$ x 10 ⁶	17.0	27.0

*90% of energy supplied by wood.

Thermal Performance is defined as the ability of the gasification system to produce a given yield of product gas of a given composition per unit of dry wood fed. Thermal performance is, therefore, the single most important factor in determining the economics of the gasification system. The thermal performance upon which existing cost feasibility evaluations were based was calculated based upon normalized experimental material balances, published and estimated thermochemical data, and calculated heat losses. Heat losses in the PRU are unavoidably high because of its relatively small scale and use of metal wall rather than refractory lined gasifier and solids transfer loops. These heat losses are compensated for by burning a supplementary fuel (natural gas) in the fluid-bed combustor. Therefore, the design of the PRU does not allow thermal performance to be measured so the thermal performance of large systems must be calculated. The calculations are affected by the following factors:

- Experimental material balance closures and identification of all reaction products,
- System heat losses, both in the PRU and those calculated to exist in larger units.
- The accuracy of thermochemical data, which are either gathered from the literature, measured (e.g., heat of combustion) or estimated (e.g., the composition of gasified char transferred to the combustor).

Since all the above elements are subject to error and/or uncertainty, it is possible for the actual thermal performance of the unit to deviate from the calculated performance, based on data obtained in units where heat losses are appreciable. Therefore, in order to provide the confidence needed to provide performance guarantees for commercial plants thermal performance must be experimentally determined at a scale large enough to minimize system heat losses.

To estimate how large a FRG facility would be required to reduce heat losses to a negligible fraction of the heating value of the incoming wood, refer to Figure 23 which shows that, at a throughput of 2000 lbs/ft²-hr, heat losses (reported as a percent of the wood heating value) increase rapidly below 200 TPD.

Solids Mixing between the recirculating sand and the incoming biomass is essential for insuring that the necessary heat transfer occurs in the **relatively** short residence time available in the gasifier. The critical zone in this mixing occurs is at the gasifier inlet where hot recirculating sand must mix with the incoming wood feed. In relatively small-diameter reactors, achieving inlet mixing between the feed and hot recirculating sand is simple because the **maximum** possible separation between sand and feed inlets is small. However, in **much** larger commercial-sized gasifiers, the inlet configuration will be critical in insuring thorough initial mixing of sand and feed. To minimize this problem, a novel gasifier concept was demonstrated under this program and a patent application filed to protect this concept. However, **larger** scale tests will be required to prove the concept works regardless of gasifier size.

Biomass Feed Rates of over 22 TPD have been achieved in the present PRU facility. The factor that limits the throughput in the PRU is the ability to provide heat from the existing conventional fluid-bed combustor.

Biomass throughput is important in determining system economics. It determines not only the capital investment required for a particular energy output but also the **maximum** capacity for a shop-fabricated gasification system. In addition, as shown in Figure 24, throughput influences cold gas efficiency because of its effect on heat losses. In the present PRU facility heat losses and combustor capacity limit throughput. Therefore it is important that the proposed scale-up facility proposed be of sufficient size to allow the factors affecting **maximum** throughput to be determined.

A Gas Clean-Up System is required to produce the cooled clean gas that will be required for most end use applications **e.g.**, being compressed to use in gas combustion turbines or distributed throughout the plant to replace natural gas and petroleum based fuels. The basic requirements of the gas clean-up system are the following:

- (1) Secondary particulate removal of the fine wood ash from the flue gas.

- (2) Heat recovery from the product gas to cool it from the gasifier exit temperature to as low a temperature as possible without initiating condensation.
- (3) Quenching and scrubbing of the product gas to remove tarry condensates and any remaining particulates.
- (4) Removal of mist from the cooled clean product gas.
- (5) Separation of the water/tar phases to allow water recirculation and tar combustion.

Discussions with vendors of gas clean-up systems, as well as with a few users of such systems, indicate that modifications in both design and operation of such systems will probably be necessary to have them work well with a particular gas. Our limited experience with a recently installed commercially available scrubber confirms the need to develop a "learning curve" with regard to gas clean-up systems for biomass gasification. However, since the gas clean-up system used in the PRU could be identical in design to a commercial system, and since the gas inlet conditions to the scrubber could also be identical to a commercial system, this important unit operation can be investigated more thoroughly in the existing PRU.

System Availability/Reliability are two parameters that are critical in determining the system's economics. System availability is defined as simply the time that the system can operate between planned maintenance. Thus, system availability influences the economics because it determines the total energy production per unit of capital investment, influences maintenance costs, and requires the user to purchase supplemental energy during downtimes or, to install spares.

System reliability is a measure of the entire system's reliability from wood feed to final end use of the cooled cleaned product gas, to function in an integrated manner. Thus, it is anticipated that reliability will be improved by progressing up the learning curve by operating of a fully integrated facility.

Obviously, to maximize reliability and determine the system's potential availability it will be necessary to operate an integrated system that utilizes commercial design philosophy.

Thus, the existing PRU could be effectively utilized to generate more scale-up information on the important supporting unit operation of gas clean-up. However, the critical issues of

- Thermal performance,
- System reliability and availability, and
- Biomass throughput.

As well as the overall validity of scale-up criteria must be demonstrated in a scaled-up facility.

PROGRAM SUMMARY

Experimental Gasification Results

- **It** was demonstrated that the gasification system could handle any reasonable form of woody biomass with minimal preparation. For example, the following feeds were successfully gasified over the course of the program.
 - Wood chips (both hard and soft wood)
 - Bark
 - Sawdust
 - Coarsely shredded stump material (residue from the solvent extraction of naval stores)
 - Whole tree chips including branch fragments and leaves
- The heating value (HHV) of the product gas was between **450** and **500 Btu/SCF** and was independent of the moisture level of the feed. [Obviously the yield of product gas decreases as the moisture level increases because more heat (char) is required to evaporate the **moisture** leaving less heat (char) for gasification.]
- Woody biomass throughputs of **3400 lbs/ft²-hr** were achieved while maintaining carbon conversions sufficient for thermally balanced operation.
- Long term operational stability was demonstrated in a test of 5 days duration.
- The only parameter significantly affecting carbon conversion was gasifier temperature. Other parameters such as wood throughput, gas velocity, even gas type did not have effects that were greater than the statistical scatter of the data.

Wood type was found to have some influence especially in tests conducted in the 6 in. I.D. gasifier.

- Tar yields were between 0.5 and 1.0 weight percent of the dry wood. (In a commercial process this tar would be burned in the combustor to provide supplementary heat for gasification.)

These results indicate that biomass is nature's way of storing a 450 Btu/SCF gas in solid form requiring only heating the biomass to 1500 to 1600 F to release the gas. Reaction rates for the conversion of biomass to gas appear to be limited only by particle heat-up rates. Thus the conversion of biomass to a medium-Btu gas can be carried out at very high throughputs in very compact reactor systems.

The two-reactor concept with heat generated in one reactor and gasification in the other utilizing circulating hot solids to transfer the heat between reactors was demonstrated to be especially well suited to the conversion of high-reactivity biomass. Data collected during the program will allow the design of commercial systems.

Experimental Combustion Results

Results indicated that wood char can be completely combusted in a circulating fluid-bed without recycle at combustor temperatures approximately 1800 F. Emissions of both CO and NO_x will fall well within NSPS. In separate tests the tar byproduct was burned to measure PAH emissions. These tests indicated that tar combustion would not result in significant PAH formation. The char throughputs obtained (nearly 400 lbs/ft²-hr) were well above those used in the conceptual design (150 lbs/ft²-hr). Therefore the circulating fluid-bed combustor provides an efficient low emission means of generating heat for the gasifier.

Process Engineering Heat and Material Balance Models

Conceptual designs for commercial plants allowed the development of a detailed heat and material balance model for the integrated gasification system proceeding from the wood dryer to the production of cooled clean product

gas. Utilization of the detailed heat and material balances allowed the influence of various waste heat recovery schemes as well as other system design parameters on overall plant thermal efficiency to be determined. These simulations indicated cold gas thermal efficiencies exceeding 70 percent could be achieved with a 40 percent moisture wood.

Economic Feasibility

The conceptual design also allowed estimations to be made of the capital costs required for an integrated FRG gasification facility. These cost estimates indicated that a FRG gasification system including a gas-fired boiler would require a significantly lower capital investment than a conventional wood/fossil fuel combination boiler. Therefore, this technology has, at least in larger plants, the economic potential to replace combustion as a means of generating energy from woody biomass

Process Scale-Up

Because of the size, design philosophy, and purpose of the PRU, certain performance parameters could not be determined by PRU testing and will have to be determined at the next stage of scale-up. These performance parameters include:

- System availability and reliability
- Thermal efficiency
- Solids mixing.

The process heat and material balance model allowed the scale-up required to reduce heat losses to a minimum to be determined. At a nominal throughput of 2000 lbs/ft²-hr, the projected plant size, at which heat losses approach a minimum, is approximately 200 TPD. A plant this size should therefore allow the direct determination of thermal efficiency as well as the other performance parameters.

CONCLUSIONS AND RECOMMENDATIONS

Based on the results of this program, the FRG process appears to have considerable commercial **potential**. The next step towards realizing this commercial potential is to scale the process up to a wood capacity of approximately 200 TPD utilizing **all** commercial components. Since the gas produced will be approximately 100 MM **Btu/hr**, **it** should be installed at a site where the value of the gas produced economically justifies the cost of building and operating the test facility. Efforts are under way to proceed with such a scale-up.

Another important spin-off from this program would be the application of FRG technology to other similar feed materials. The feed material of the greatest immediate commercial importance is RDF. **It** is anticipated that the same advantages that FRG technology has over direct combustion for woody biomass would extend directly to RDF conversion. Therefore, initiating a PRU program on the gasification of RDF offers a major opportunity to provide a cleaner more economic conversion to **MSW** utilization than combustion.

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APPENDIX A

TEST DATA

FOREST RESIDUE GASIFICATION - 1ST PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	1	2	3	4	5	6	7	8	9	10
LAB BOOK RUN NUMBER	6A	6B	7A	7B	7C	9	11A	11B	13A	13B
RUN CONDITION	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0	1-0-0
WOOD TYPE	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	CHERRY	CHERRY
WOOD SIZE, INCHES	1/4*3/2	1/4*3/2	1/4*3/2	1/4*3/2	1/4*3/2	1/4	1/4	1/4	1/2	1/2
MOISTURE, %	48.3	48.3	30	30	28.3	17.2	25.2	25.2	72	26
WOOD FEED RATE, LB/HR	238	238	176	176	172	147	326	326	274	165
DRY FEED RATE, LB/HR	123	123	123	123	123	122	244	244	203	122
GASIFIER TEMPERATURE, F	1325	1325	1400	1420	1450	1440	1410	1480	1430	1460
GASIFIER PRESSURE, PSIG	3.8	3.8	4	4	4	3.2	3.8	3.8	4	4
NITROGEN FEED, LB/HR	0	0	0	0	0	0	0	0	0	0
STEAM RATE, LB/HR	160	181	158	180	180	327	208	264	189	189
GASIFIER GAS VELOCITY, FT/SEC	13.09	14.81	13.33	15.35	15.59	29.44	17.66	23.48	16.20	16.46
COMBUSTOR TEMPERATURE, F	1480	1550	1632	1600	1650	1643	1590	1610	1650	1640
COMBUSTOR PRESSURE, PSIG	3.8	3.8	4	4	4	3.2	4.2	4.2	4.4	4.4
COMBUSTOR GAS VELOCITY, FT/SEC	1.3	1.3	1.4	1.4	1.4	1.8	1.7	1.7	1.5	1.5
PRODUCT GAS RATE, SCF/LB-MAF	6.00	4.80	8.40	10.20	12.50	8.10	9.00	9.60	8.00	8.50
CARBON CONVERSION TO GAS, %	37.70	29.90	49.29	60.21	63.10	49.90	54.70	58.70	50.40	54.10
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	11.05	11.17	11.40	12.09	13.26	14.80	12.39	14.86	11.85	11.85
CARBON DIOXIDE	19.49	18.58	17.26	16.40	15.74	15.30	16.28	16.27	13.91	13.56
ETHYLENE	4.88	4.92	5.01	5.26	5.38	5.00	4.71	5.26	5.01	5.12
ETHANE	1.00	1.58	1.00	.94	.82	.90	1.10	.89	1.34	1.28
METHANE	12.86	13.28	15.38	15.60	15.64	15.70	15.96	14.51	17.82	19.03
CARBON MONOXIDE	50.78	50.51	49.95	49.81	49.18	48.30	49.56	49.21	50.09	49.16
H2/CO RATIO	.22	.22	.23	.24	.27	.31	.25	.30	.24	.24
PRODUCT GAS HHV, BTU/SCF	425	440	451	457	460	458	456	453	484	484
BTU YIELD, BTU/LB-MAF	2550	2111	3789	4662	5751	3711	4104	4351	3670	4197
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	82.84	82.84	83.83	83.83	83.53	83.88	83.83	83.83	81.65	81.65
FIXED CARBON	15.72	15.72	15.63	15.63	16.04	15.36	14.92	14.92	17.00	17.00
ASH	1.44	1.44	.54	.54	.43	.76	1.25	1.25	1.35	1.35
CARBON	49.31	49.31	49.83	49.83	49.99	48.97	49.49	49.49	49.82	49.82
HYDROGEN	6.03	6.03	5.76	5.76	5.77	5.84	5.82	5.82	5.81	5.81
NITROGEN	1.8	.18	.22	.22	.22	.22	.19	.19	.31	.31
CHLORINE	.04	.04	.03	.03	.03	.02	.02	.02	.02	.02
SULFUR	.02	.02	.05	.05	.04	.04	.05	.05	.02	.02
OXYGEN (by difference)	42.98	42.98	43.57	43.57	43.54	44.15	43.18	43.18	42.97	42.97
BTU/LB (DRY)	8366	8366	8340	8340	8339	8422	8373	8373	8408	8408

* A	-	B	-	C	-
1 - STEAM		0 - NO NITROGEN		0 - NO DENSE BED	
		1 - 90% NITROGEN		1 - 1/4" MASTER BEADS	
		2 - 80% NITROGEN		2 - 112" FISH SPINE INSULATORS	
		3 - 70% NITROGEN		3 - 3/8" SS TUBING	

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FOREST RESIDUE GASIFICATION - 1ST PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	11	12	13	14	15	16	17	18	19	20
LAB BOOK RUN NUMBER	14A	14B	15A	15B	16A	16B	16C	17A	17B	18A
RUN CONDITION	1-0-0	1-0-0	1-0-0	1-0-1	1-0-0	1-0-0	1-0-1	1-0-0	1-0-0	1-0-0
WOOD TYPE	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	MIXED	OAK, B.M	OAK, R.M
WOOD SIZE, INCHES	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4
MOISTURE, %	36	36	29.9	29.9	28.9	28.9	28.9	20.2	20.2	4 5
WOOD FEED RATE, LB/HR	317	216	197	197	69	69	89	79	307	64
DRY FEED RATE, LB/HR	203	138	138	138	49	49	49	63	245	81
GASIFIER TEMPERATURE, F	1460	1525	1500	1520	1560	1608	1640	1530	1500	1540
GASIFIER PRESSURE, PSIG	4	4	4.8	4.8	5.4	5.4	5.4	4	4	4
NITROGEN FEED, LB/HR	0	0	0	0	0	0	0	0	0	0
STEAM RATE, LB/HR	179	268	206	183	178	178	178	181	181	174
GASIFIER GAS VELOCITY, FT/SEC	15.59	24.13	17.56	15.76	15.17	15.52	15.77	16.34	16.09	15.78
COMBUSTOR TEMPERATURE, F	1640	1743	1700	1790	1810	1850	1860	1720	1740	1720
COMBUSTOR PRESSURE, PSIG	4.4	4.4	5.2	5.2	5	5	5	4.2	4.2	4.4
COMBUSTOR GAS VELOCITY, FT/SEC	1.8	2	2	2	1.9	1.9	1.9	1.6	1.6	1.4
PRODUCT GAS RATE, SCF/LB-MAF	9.40	9.80	10.10	11.00	10.50	10.80	12.20	11.70	9.70	10.70
CARBON CONVERSION TO GAS, %	55.60	62.80	60.84	64.72	61.90	64.10	69.40	69.10	58.10	61.70
PRODUCT GAS COMPOSITION, % N2 FRLE										
HYDROGEN	13.60	13.30	11.96	14.82	19.29	18.07	22.86	17.56	14.87	17.26
CARBON DIOXIDE	15.40	14.20	13.76	13.75	13.71	14.17	13.68	14.27	14.65	13.73
ETHYLENE	4.80	5.30	5.29	5.23	5.06	4.93	4.40	5.03	4.87	5.38
ETHANE	1.40	1.30	1.38	1.16	1.39	1.32	1.28	1.23	1.30	1.27
METHANE	17.50	18.20	18.61	17.99	17.90	18.03	17.58	17.07	17.80	17.82
CARBON-MONOXIDE	47.30	47.70	49.00	47.03	42.84	43.48	40.21	44.84	48.51	44.78
H2/CO RATIO	.29	.28	.24	.32	.45	.42	.57	.39	.32	.38
PRODUCT GAS HHV, BTU/SCF	475	489	494	488	487	483	475	478	478	487
BTU YIELD, BTU/LB-MAF	4465	4788	4988	5342	5108	5219	5789	5572	4846	5207
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	92.76	92.76	82.15	82.15	82.59	82.59	82.59	80.15	80.15	80.84
FIXED CARBON	16.33	16.33	17.32	17.32	16.57	16.57	16.57	17.76	17.76	18.91
ASH	.91	.91	.53	.53	.84	.84	.84	2.09	2.09	2.25
CARBON	49.54	49.54	49.20	49.20	49.72	49.72	49.72	49.83	49.83	48.85
HYDROGEN	5.88	5.88	5.88	5.88	5.99	5.99	5.99	5.87	5.87	5.92
NITROGEN	.26	.26	.17	.17	.17	.17	.17	.32	.32	.38
CHLORINE	.03	.03	.02	.02	.02	.02	.02	.03	.03	.02
SULFUR	.04	.04	.12	.12	.05	.05	.05	.04	.04	.04
OXYGEN(by difference)	43.34	43.34	44.08	44.08	43.21	43.21	43.21	41.82	41.82	42.44
BTU/LB (DRY)	8386	8386	8332	8332	8117	8117	8117	8373	8373	8120

* A	-	B	-	C
1 - STEAM	0 - NO NITROGEN	0 - NO DENSE BED		
	1 - 90% NITROGEN	1 - 114" MASTER BEADS		
	2 - 80% NITROGEN	2 - 1/2" FISH SPINE INSULATORS		
	3 - 70% NITROGEN	3 - 3/8" SS TUBING		

FOREST RESIDUE GASIFICATION - 1ST PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	21	22	23	24	25	26	27	28	29	30
LAB BOOK RUN NUMBER	188	19A	19B	20A	20B	20D	21B	21C	22A	22B
RUN CONDITION	1-0-2	1-0-0	1-0-0	1-0-0	1-0-0	1-2-0	1-3-0	1-3-2	1-0-0	1-0-2
WOOD TYPE	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M
WOOD SIZE, INCHES	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4
MOISTURE, %	4.5	12.2	12.2	11.8	11.8	11.8	11.9	11.9	10.8	10.8
WOOD FEED RATE, LB/HR	64	69	164	83	150	150	151	151	60	60
DRY FEED RATE, LB/HR	61	61	111	56	132	132	133	133	54	54
GASIFIER TEMPERATURE, F	1570	1440	1470	1550	1560	1560	1520	1460	1500	1560
GASIFIER PRESSURE, PSIG	4	3.8	3.8	4.2	4.2	4.8	4.6	4.6	4	4
NITROGEN FEED, LB/HR	0	0	0	0	0	260	196	196	0	0
STEAM RATE, LB/HR	136	226	210	202	169	42	54	54	108	108
GASIFIER GAS VELOCITY, FT/SEC	12.52	19.69	18.58	18.22	15.32	18.38	15.66	15.19	9.60	9.90
COMBUSTOR TEMPERATURE, F	1720	1640	1700	1760	1780	1820	1720	1650	1700	1780
COMBUSTOR PRESSURE, PSIG	4.4	4	4	4.6	4.4	5.2	5	5	4.4	4.4
COMBUSTOR GAS VELOCITY, FT/SEC	1.4	1.3	1.3	1.6	1.6	1.6	1.5	1.2	1.3	1.3

PRODUCT GAS RATE, SCF/LB-MAF	10.60	9.30	8.90	13.10	11.90	10.30	10.60	9.60	11.40	12.10
CARBON CONVERSION TO GAS, %	59.60	59.20	54.10	73.84	66.89	58.99	60.54	55.30	63.87	64.58
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	19.47	12.99	14.99	17.13	17.23	15.89	14.89	13.60	17.09	20.25
CARBON DIOXIDE	14.05	12.33	13.78	16.44	14.84	16.48	19.52	19.53	15.89	15.85
ETHYLENE	5.00	5.45	5.04	5.04	5.10	5.07	5.10	5.10	5.20	5.07
ETHANE	1.06	1.46	1.34	.63	.68	.56	.74	.95	.68	.36
METHANE	17.02	18.11	17.82	16.48	16.65	18.87	18.38	16.58	16.19	16.58
CARBON MONOXIDE	43.41	49.65	47.03	44.29	45.51	45.13	43.37	44.24	44.97	42.88
H2/CO RATIO	.45	.26	.32	.39	.38	.35	.34	.31	.38	.47
PRODUCT GAS HHV, BTU/SCF	474	498	485	457	464	458	448	453	459	449
BTU YIELD, BTU/LB-MAF	5023	4633	4313	5982	5527	4722	4752	4300	5231	5431
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	80.84	81.36	81.36	81.49	81.49	81.49	80.10	80.10	80.84	80.84
FIXED CARBON	16.91	17.08	17.08	17.13	17.13	17.13	18.58	18.58	18.88	18.88
ASH	2.25	1.56	1.56	1.38	1.38	1.38	3.31	3.31	2.5	2.5
CARBON	48.95	49.00	49.00	49.10	49.10	49.10	48.88	48.88	49.04	49.04
HYDROGEN	5.92	5.94	5.94	5.86	5.86	5.86	5.84	5.84	5.98	5.98
NITROGEN	.38	.29	.29	.32	.32	.32	.41	.41	.32	.32
CHLORINE	.02	.02	.02	.01	.01	.01	.02	.02	.02	.02
SULFUR	.04	.04	.04	.04	.04	.04	.07	.07	.04	.04
OXYGEN (by difference)	42.44	43.15	43.15	43.29	43.29	43.29	41.47	41.47	42.12	42.12
BTU/LB (DRY)	8120	8295	8295	8343	8343	8343	8266	8266	8318	8318

* A	-	B	-	C
1 - STEAM	0 - NO NITROGEN	0 - NO DENSE BED		
	1 - 90% NITROGEN	1 - 114" MASTER BEADS		
	2 - 80% NITROGEN	2 - 1/2" FISH SPINE INSULATORS		
	3 - 70% NITROGEN	3 - 3/8" SS TUBING		

FOREST RESIDUE GASIFICATION - 1ST PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	31	32	33	34	35	36
LAB BOOK RUN NUMBER	22C	22D	23A	23B	24A	24B
RUN CONDITION	1-3-0	1-3-2	1-0-0	1-3-0	1-3-0	1-3-3
WOOD TYPE	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M	OAK, B, M
WOOD SIZE, INCHES	+1/4	+1/4	+1/4	+1/4	+1/4	+1/4
MOISTURE, %	10.6	10.6	4.3	4.3	4.3	4.3
WOOD FEED RATE, LB/HR	60	60	117	148	148	148
DRY FEED RATE, LB/HR	54	54	112	142	142	142
GASIFIER TEMPERATURE, F	1540	1580	1510	1540	1500	1530
GASIFIER PRESSURE, PSIG	4	4	4	4	4	4
NITROGEN FEED, LB/HR	130	130	0	232	258	258
STEAM RATE, LB/HR	36	36	242	64	71	71
GASIFIER GAS VELOCITY, FT/SEC	10.85	11.06	21.62	19.34	21.06	21.38
COMBUSTOR TEMPERATURE, F	1740	1760	1720	1720	1680	1700
COMBUSTOR PRESSURE, PSIG	4.4	4.4	4.2	4.2	4.2	4.2
COMBUSTOR GAS VELOCITY, FT/SEC	1.3	1.3	1.4	1.4	1.4	1.4
PRODUCT GAS RATE, SCF/LB-MAF	13.10	12.60	9.80	10.90	11.60	11.40
CARBON CONVERSION TO GAS, %	73.10	69.65	54.12	62.08	65.40	63.70
PRODUCT GAS COMPOSITION, % N2 FREE						
HYDROGEN	16.66	18.71	18.44	15.23	14.33	15.23
CARBON DIOXIDE	16.11	15.72	16.37	15.81	15.59	15.40
ETHYLENE	5.32	5.20	5.00	5.35	5.34	5.42
ETHANE	.58	.48	.65	.69	.82	.69
METHANE	16.09	15.90	16.45	17.25	17.25	17.36
CARBON MONOXIDE	45.23	44.02	43.09	45.67	46.68	45.90
H2/CO RATIO	.37	.43	.43	.33	.31	.33

PRODUCT GAS HHV, BTU/SCF	458	455	456	469	471	472
BTU YIELD, BTU/LB-MAF	5997	5823	4473	5109	5466	5377
WOOD ANALYSIS, % DRY BASIS						
VOLATILE MATTER	80.84	80.84	82.00	82.00	82.00	82.00
FIXED CARBON	16.66	16.66	16.41	16.41	16.41	16.41
ASH	2.5	2.5	1.59	1.59	1.59	1.59
CARBON	49.04	49.04	49.53	49.53	49.53	49.53
HYDROGEN	5.96	5.96	6.05	6.05	6.05	6.05
NITROGEN	.32	.32	.28	.28	.28	.28
CHLORINE	.02	.02	.02	.02	.02	.02
SULFUR	.04	.04	.07	.07	.07	.07
OXYGEN (by difference)	42.12	42.12	42.46	42.46	42.46	42.46
BTU/LB (DRY)	8316	8316	8368	8368	8368	8368

* A	-	B	-	C	
1 - STEAM		0 - NO NITROGEN		0 - NO DENSE BED	
		1 - 90% NITROGEN		1 - 1/4" MASTER BEADS	
		2 - 80% NITROGEN		2 - 1/2" FISH SPINE INSULATORS	
		3 - 70% NITROGEN		3 - 3/8" SS TUBING	

FOREST RESIDUE GASIFICATION - 2ND PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	37	38	39	40	41	42	43	44	45	46
LAB BOOK RUN NUMBER	213	2.4	2.5A	2.5B	2.5C	2.50	2.6	3.7	3.8A	3.8R
RUN CONDITION	B-S	8-S	B-S	8-S	B-N-S	8-S	S	S	N-S	N
WOOD TYPE	HARD	HARD	HARD	HARD	HARD	HARD	SOFT	SOFT	SOFT	SOFT
MOISTURE, %	5.31	5.31	8.15	8.15	8.15	8.15	5.53	8.23	6.19	6.19
WOOD FEED RATE, LB/HR	196	150	118	137	135	299	279	195	134	134
DRY FEED RATE, LB/HR	186	142	108	126	124	275	264	183	126	126
GASIFIER TEMPERATURE, F	1550	1489	1453	1510	1574	1455	1425	1470	1466	1526
GASIFIER PRESSURE, PSIG	2.6	4	2	2	2	2	1.8	3.6	2.3	4
NITROGEN FEED, LB/HR	0	0	0	0	230	0	0	0	40	335
STEAM RATE, LB/HR	214	363	270	202	45	275	307	341	328	0
GASIFIER GAS VELOCITY, FT/SEC	21.09	32.09	26.23	20.21	19.92	26.75	30.11	30.50	31.10	20.00
COMBUSTOR TEMPERATURE, F	1687	1636	1630	1630	1630	1630	1533	1640	1572	1660
COMBUSTOR PRESSURE, PSIG	2.8	5.3	1.9	1.9	1.9	1.9	2	3.2	2.6	4
COMBUSTOR GAS VELOCITY, FT/SEC	1.80	1.50	1.70	1.70	1.70	1.70	1.80	1.70	1.60	1.60
PRODUCT GAS RATE, SCF/LB-MAF	14.30	13.40	11.95	11.12	13.15	11.87	10.20	11.40	12.40	13.10
CARBON CONVERSION TO GAS, %	86.00	79.90	73.00	73.20	79.00	71.70	58.80	84.70	64.50	66.10
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	19.93	16.31	14.67	15.39	16.61	15.55	16.63	18.65	17.20	17.09
CARBON DIOXIDE	13.65	13.37	13.89	13.99	13.42	13.89	9.44	9.62	14.43	14.97
ETHYLENE	5.01	5.33	5.25	5.39	5.61	5.28	5.73	5.81	5.83	6.85
ETHANE	.63	.63	.90	.76	.53	.78	.89	.81	.84	1.76
METHANE	16.53	16.85	15.87	16.00	16.81	16.49	15.21	15.39	15.09	15.67
CARBON MONOXIDE	44.26	47.50	49.49	48.47	47.02	48.01	52.06	49.72	51.61	48.67
H2/CO RATIO	.45	.34	.30	.32	.35	.32	.32	.38	.33	.36
PRODUCT GAS HHV, BTU/SCF	466	473	467	467	474	470	483	484	483	511
BTU YIELD, BTU/LB-MAF	6659	6334	5584	5198	6239	5579	4924	5512	5985	6700
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	81.9	82.29	81.47	81.47	81.47	81.47	82.51	82.54	83.1	63.1
FIXED CARBON	17.15	16.62	17.06	17.06	17.06	17.06	16.71	16.82	16.23	16.23
ASH	.95	1.09	1.47	1.47	1.47	1.47	.78	.84	.67	.67
CARBON	49.73	49.77	50.56	50.56	50.56	50.56	50.8	51.4	50.77	50.77
HYDROGEN	6.06	6.07	5.8	5.8	5.8	5.8	6.25	6.29	8.3	6.3
NITROGEN	.24	.26	.32	.32	.32	.32	.15	.12	.16	.15
CHLORINE	.11	.0	.01	.01	.01	.01	.03	.03	.15	.15
SULFUR	.04	.05	.03	.03	.03	.03	.14	.14	.01	.01
OXYGEN (by difference)	42.87	42.76	41.81	41.81	41.81	41.81	41.75	41.34	41.95	41.96
BTU/LB (DRY)	8430	8290	8479	8479	8479	8479	8771	8752	8887	8887

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN (FLOWS APPROXIMATE)
 S - STEAM

FOREST RESIDUE GASIFICATION - 2ND PHASE RUN RESULTS - 6 INCH GASIFIER

TEST NUMBER	47	48	49	50	51	52	53	54	55	56
LAB BOOK RUN NUMBER	2.9C	2.10	2.12A	2.12B	2.13	2.14	2.15A	2.15B	2.16A	2.17A
RUN CONDITION	N	S	N-S	N-S	S	B-C-S	B-C-S	B-C-S	S	S
WOOD TYPE	SOFT	SOFT	SOFT	SOFT	BARK	SOFT	SOFT	SOFT	BARK	BARK
MOISTURE, %	6.19	4.83	7.57	7.57	20.4	9.78	11.96	11.96	16.92	20.46
WOOD FEED RATE, LB/HR	356	136	134	200	107	154	179	179	101	71
DRY FEED RATE, LB/HR	334	129	124	185	85	139	158	158	84	66
GASIFIER TEMPERATURE, F	1448	1527	1450	1454	1472	1620	1468	1490	1488	1535
GASIFIER PRESSURE, PSIG	6	3.6	4.1	4.8	2.1	2.4	3.2	3.3	2.7	2.6

NITROGEN FEED, LB/HR	300	0	230	230	0	0	0	0	0	0
STEAM RATE, LB/HR	0	370	60	60	315	306	377	385	361	371
GASIFIER GAS VELOCITY, FT/SEC	17.60	34.08	23.00	22.30	30.73	30.05	34.26	35.38	33.88	38.28
COMBUSTOR TEMPERATURE, F	1655	1685	1544	1616	1553	1762	1627	1703	1710	1730
COMBUSTOR PRESSURE, PSIG	5.6	3.8	4.4	5.1	2.2	2.4	3	3.1	2.7	2.8
COMBUSTOR GAS VELOCITY, FT/SEC	1.40	1.60	1.50	1.60	1.50	1.60	1.60	1.50	1.60	1.60
PRODUCT GAS RATE, SCF/LB-MAF	10.10	12.10	10.70	10.80	10.10	13.30	11.80	12.80	11.70	13.30
CARBON CONVERSION TO GAS, %	60.00	69.60	65.70	64.30	58.80	76.20	68.80	72.80	64.50	72.80
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	15.91	18.11	14.10	14.48	22.77	20.55	17.66	19.58	22.56	21.70
CARBON DIOXIDE	9.57	9.58	11.88	11.24	20.01	9.95	17.2	9.84	18.02	18.53
ETHYLENE	8.05	5.85	6.96	6.48	6.27	5.75	5.91	5.89	6.41	6.43
ETHANE	1.09	.62	2.42	1.97	.90	.62	.83	.80	1.34	.87
METHANE	15.89	15.74	15.89	15.41	14.18	14.90	16.63	15.34	14.86	16.13
CARBON MONOXIDE	51.49	50.30	46.71	50.44	35.85	48.22	50.24	48.45	38.81	38.34
H2/CO RATIO	.31	.36	.29	.29	.84	.43	.36	.40	.61	.60
PRODUCT GAS HHV, BTU/SCF	494	481	518	503	448	478	488	483	488	489
BTU YIELD, BTU/LB-MAF	4991	5823	5538	5337	4538	8325	5788	6231	5481	6237
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	83.1	84.37	81.83	81.83	74.52	83.18	83.16	83.15	76.86	73.88
FIXED CARBON	16.23	15.23	17.14	17.14	16.94	16.35	16.28	16.28	14.88	16.81
ASH	.67	.4	1.03	1.03	8.54	.47	.69	.59	9.37	10.83
CARBON	50.77	51.06	50.87	50.87	47.06	51.57	51.5	51.5	47.27	48.27
HYDROGEN	6.3	6.32	6.27	6.27	5.74	6.12	6.12	6.12	5.2	6.07
NITROGEN	.15	.21	.18	.18	.39	.18	.12	.12	.4	.47
CHLORINE	.15	.1	.03	.03	.01	.02	.02	.02	.03	.02
SULFUR	.01	.04	.04	.04	.11	.08	.04	.04	.05	.08
OXYGEN(by difference)	41.95	41.9	41.58	41.58	38.15	41.58	41.81	41.61	37.68	37.46
BTU/LB (DRY)	8687	8551	8795	8795	7798	8868	8865	8865	7721	7815

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN (FLOWS APPROXIMATE)
 S - STEAM

FOREST RESIDUE GASIFICATION - 2ND PHASE UN RESULTS - 6 INCH GASIFIER

TEST NUMBER	57	58	59	60	61	62	83	64	65	66
LAB BOOK RUN NUMBER	2-178	2-179	2-180	2-181	2-182	2-183	2-208	2-209	2-210	2-211
WOOD CONDITION	BARK	BARK	BARK	BARK	BARK	BARK	SOFT	SOFT	HARD	HARD
MOISTURE, %	20.46	25.11	25.11	25.11	25.11	25.11	8.77	8.77	10.75	10.75
WOOD FEED RATE, LB/HR	58	64	64	64	64	64	98	98	161	161
DRY FEED RATE, LB/HR	46	48	48	48	48	48	89	89	135	135
DRY FEED TEMPERATURE, F	1513	1625	1559	1559	1559	1559	1470	1470	1514	1514
GASIFIER PRESSURE, PSIG	2.6	2.4	2.7	2.7	2.7	2.7	1.2	1.2	2.4	2.4
NITROGEN FEED, LB/HR	0	0	0	0	0	0	0	0	0	0
STEAM RATE, LB/HR	371	355	381	381	381	381	414	414	402	402
GASIFIER GAS VELOCITY, FT/SEC	35.89	36.72	37.50	37.50	37.50	37.50	42.63	42.63	39.36	39.36
COMBUSTOR TEMPERATURE, F	1727	1765	1760	1760	1760	1760	1826	1826	1624	1624
COMBUSTOR PRESSURE, PSIG	12.6	12.4	12.7	12.7	12.7	12.7	3.8	3.8	3.8	3.8
COMBUSTOR GAS VELOCITY, FT/SEC	1.60	1.70	1.70	1.70	1.70	1.70	1.60	1.60	1.60	1.60
PRODUCT GAS RATE, SCF/LB-MAF	12.60	16.60	16.00	16.00	16.00	16.00	8.20	8.20	8.00	8.00
CARBON CONVERSION TO GAS, %	69.30	84.80	78.90	78.90	78.90	78.90	53.40	53.40	51.00	51.00
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	23.02	26.16	27.89	27.89	27.89	27.89	17.58	17.58	15.94	15.94
CARBON DIOXIDE	18.35	18.60	18.89	18.89	18.89	18.89	12.20	12.20	12.46	12.46
ETHYLENE	6.02	5.88	5.65	5.65	5.65	5.65	5.51	5.51	5.93	5.93
ETHANE	.89	.87	.73	.73	.73	.73	1.13	1.13	1.43	1.43
METHANE	14.62	14.25	13.42	13.42	13.42	13.42	15.86	15.39	16.71	16.71
CARBON MONOXIDE	37.09	34.44	33.43	33.43	33.43	33.43	40.74	40.48	48.15	48.15
H2/CO RATIO	.62	.76	.83	.83	.83	.83	.37	.41	.33	.33
PRODUCT GAS HHV, BTU/SCF	454	448	437	437	437	437	479	482	485	485
BTU YIELD, BTU/LB-MAF	5723	7409	6997	6997	6997	6997	4409	4337	4172	4256
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	73.56	73.34	73.34	73.34	73.34	73.34	85.81	85.81	81.18	81.18
FIXED CARBON	15.81	15.46	15.46	15.46	15.46	15.46	13.77	13.77	17.1	17.1
ASH	10.63	11.2	11.2	11.2	11.2	11.2	4.2	4.2	1.72	1.72
CARBON	46.27	46.3	46.3	46.3	46.3	46.3	52.66	52.66	50.28	50.28
HYDROGEN	5.07	5.4	5.4	5.4	5.4	5.4	7.03	7.03	5.82	5.82
NITROGEN	.47	.47	.47	.47	.47	.47	1.2	1.2	.21	.21
CHLORINE	.02	.02	.02	.02	.02	.02	.02	.02	.03	.03
SULFUR	.08	.08	.08	.08	.08	.08	.08	.08	.08	.08
OXYGEN (by difference)	37.46	36.59	36.59	36.59	36.59	36.59	39.69	39.69	41.87	41.87
BTU/LB (DRY)	7615	7727	7727	7727	7727	7727	8810	8810	8478	8478

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN (FLOWS APPROXIMATE)
 S - STEAM

FOREST RESIDUE GASIFICATION - 2ND PH				
TEST NUMBER	67	68	69	70
LAB BOOK RUN NUMBER	2.22A	2.228	2.22C	2.22D
RUN CONDITION	S	S	S	N-S
WOOD TYPE	SAWDUST	SAWDUST	SAWDUST	SAWDUST
MOISTURE, %	8.18	8.18	8.18	8.18
WOOD FEED RATE, LB/HR	111	104	91	103
DRY FEED RATE, LB/HR	102	95	84	95
GASIFIER TEMPERATURE, F	1495	1611	1623	1465
GASIFIER PRESSURE, PSIG	2.4	2.4	2.4	4.2
NITROGEN FEED, LB/HR				200
STEAM RATE, LB/HR	344	360	386	69
GASIFIER GAS VELOCITY, FT/SEC	33.36	35.20	37.97	24.25
COMBUSTOR TEMPERATURE, F	1675	1696	1737	1780
COMBUSTOR PRESSURE, PSIG	2.6	2.6	2.6	4.2
COMBUSTOR GAS VELOCITY, FT/SEC	1.50	1.60	1.60	1.50
PRODUCT GAS RATE, SCF/LB-MAF	12.70	13.30	13.30	10.90
CARBON CONVERSION TO GAS, %	73.10	76.60	76.60	64.70
PRODUCT GAS COMPOSITION, % N2 FREE				
HYDROGEN	18.76	19.28	19.77	17.73
CARBON DIOXIDE	7.14	7.33	7.46	7.00
ETHYLENE	6.48	6.14	6.34	6.69
ETHANE	.55	.49	.43	2.72
METHANE	15.62	15.64	15.67	15.52
CARBON MONOXIDE	51.38	50.82	50.44	50.35
H2/CO RATIO	.37	.38	.37	.36
PRODUCT GAS HHV, BTU/SCF	198	191	193	532
BTU YIELD, BTU/LB-MAF	6320	6534	6558	5797
WOOD ANALYSIS, % DRY BASIS				
VOLATILE MATTER	84.44	84.44	84.44	84.44
FIXED CARBON	15.2	15.2	15.2	15.2
ASH	.36	.36	.36	.36
CARBON	51.33	51.33	51.33	51.33
HYDROGEN	6.13	6.13	6.13	6.13
NITROGEN	.12	.12	.12	.12
CHLORINE	.07	.07	.07	.07
SULFUR	.02	.02	.02	.02
OXYGEN (by difference)	41.97	41.97	41.97	41.97
BTU/LB (DRY)	8802	8802	8802	8802
B - TWI C - INTERNAL CATALYST N - NITROGEN (FLOWS APPROXIMATE) S - STEAM				

A-9

TEST NUMBER	71	72	73	74	75	76	77	78	79	80
LAB BOOK RUN NUMBER	3.1	3.2A	3.2B	3.4	3.5	3.6A	3.6B	3.7	3.A	3.B
RUN CONDITION	N-S	N	N	N-S	N-S	N-S	N-S	N-S	B-N-S	B-N-S
WOOD TYPE	DFO OAK	RFD OAK	RFD OAK	RFD OAK	RFD OAK	RFD OAK	RFD OAK	RFD OAK	RIR-MAP	RIR-MAP
MOISTURE, %	14.42	14.76	14.76	15.69	6.88	9	9	10.79	10	10
WOOD FEED RATE, LB/HR	486	508	875	588	700	816	777	1045	7AA	AAA
DRY FEED RATE, LB/HR	416	433	746	496	652	833	707	932	707	522
GASIFIER TEMPERATURE, F	1280	1425	1167	1504	1562	1359	1466	1382	1445	1470
GASIFIER PRESSURE, PSIG	3.8	3.2	4.8	6.2	8.2	8.4	7.8	7	5.8	5.2
NITROGEN FEED, LB/HR	139	156	156	173	315	169	204	218	160	160
STEAM RATE, LB/HR	389	0	0	404	401	481	415	371	406	412
GASIFIER GAS VELOCITY, FT/SEC	13.67	3.21	2.53	14.71	16.18	13.62	14.19	8.00	14.3A	15.1A
COMBUSTOR TEMPERATURE, F	1525	1626	1576	1866	1897	1806	1845	1670	1734	1680
COMBUSTOR PRESSURE, PSIG	4.2	3.4	5	6.6	9	8.7	8.4	7.7	6.A	5.7
COMBUSTOR GAS VELOCITY, FT/SEC	1.64	1.26	1.43	1.74	1.68	1.68	1.68	1.72	1.79	1.77
PRODUCT GAS RATE, SCF/LB-MAF	7.66	11.70	5.16	10.74	13.94	8.50	10.84	9.94	10.25	10.5A
CARBON CONVERSION TO GAS, %	45.62	66.15	31.88	59.00	76.36	50.54	61.82	59.40	58.04	57.76
PRODUCT GAS COMPOSITION, % NZ FR _{SS}										
HYDROGEN	8.25	18.35	8.62	20.59	20.51	14.66	18.15	14.94	17.20	20.81
CARBON DIOXIDE	6.32	14.36	21.58	14.24	17.08	14.39	13.93	14.21	12.75	14.78
ETHYLENE	3.81	4.75	2.80	4.80	3.94	4.18	4.63	4.30	4.78	4.28
ETHANE	1.01	.69	1.22	.48	.25	1.02	.73	.AA	.74	.AO
ACETYLENE										
METHANE	16.10	15.85	14.48	15.90	14.41	16.59	16.30	16.47	16.3A	15.3A
CARBON MONOXIDE	49.52	46.00	51.30	43.98	43.80	49.18	46.27	49.13	48.15	44.17
H2/CO RATIO	.27	.40	.17	.47	.47	.30	.39	.30	.3A	.47
PRODUCT GAS HHV, BTU/SCF	444	456	406	455	421	459	480	459	4RA	444
BTU YIELD, BTU/LB-MAF	3401	5339	2095	4882	5863	3897	4981	4563	4778	4704
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	81.07	81.28	81.28	70.42	82.82	83.38	83.38	80.47	83.99	84.48
FIXED CARBON	16.17	17.69	17.69	28.37	16.55	15.48	15.48	17.50	15.AA	14.AA
ASH	2.76	1.03	1.03	1.21	.63	1.14	1.14	2.03	.38	.87
CARBON	49.34	50.03	50.03	57.75	50.08	49.77	49.77	48.AO	48.AA	50.14
HYDROGEN	5.93	6.37	6.37	5.58	0.05	5.99	5.99	5.88	6.12	6.10
NITROGEN	.07	.21	.21	.44	.21	.20	.20	.25	.10	.03
CHLORINE	.03	.00	.00	.03	.08	.02	.02	.03	.03	.03
SULFUR	.13	.03	.03	.87	.00	.00	.00	.01	.AA	.AA
OXYGEN (by difference)	41.74	42.30	42.30	34.12	42.92	42.88	42.88	42.32	43.45	42.80
BTU/LB (dry)	8220	8412	8412	9789	8398	8524	8524	8431	84.AA	84.A1
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B C N S	TWI INTERNAL CATALYST NITROGEN STEAM									

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	81	82	83	84	85	86	87	88	88	80
LAB BOOK RUN NUMBER	3.10A	3.10B	3.14	3.16	3.16	3.17	3.1A	3.1A	3.1A	3.31
RUN CONDITION	B-N-S	B-N-S	N-S	N-S	N-S	N-S	N-S	N-S	N-S	N-S
WOOD TYPE	BIR+MAP	BIR+MAP	BIR+MAP	BIR+MAP	BIR+MAP	HARD	HARD	HARD	S OAK	S OAK
MOISTURE, %	8.12	8.12	12.37	3.66	10	3	4.7	3.7	16	6.6
WOOD FEED RATE, LB/HR	408	782	457	418	397	426	426	41.1	34A	490
DRY FEED RATE, LB/HR	375	719	400	403	357	412	405	388	2A8	397
GASIFIER TEMPERATURE, F	1548	1429	1465	1386	1422	1461	1408	143A	1371	143A
GASIFIER PRESSURE, PSIG	5	6.8	4.4	4.1	3.4	2.6	2.6	2.4	2.8	2.7
NITROGEN FEED, LB/HR	160	152	118	106	225	231	333	2A	101	124
STEAM RATE, LB/HR	413	437	430	432	487	447	422	426	438	419
GASIFIER GAS VELOCITY, FT/SEC	15.95	14.27	15.48	14.96	19.31	18.76	18.63	18.21	15.81	1A.8A
COMBUSTOR TEMPERATURE, F	1804	1675	1722	1707	1623	1672	1628	1675	1630	1630
COMBUSTOR PRESSURE, PSIG	5	6.8	3.8	3.8	3.3	2.4	2.4	3.3	3.5	3.8
COMBUSTOR GAS VELOCITY, FT/SEC	2.04	1.67	2.50	2.60	2.17	2.12	2.20	2.13	2.16	2.28
PRODUCT GAS RATE, SCF/LB-MAF	11.08	9.82	8.97	10.98	8.24	10.72	10.30	10.2A	in 21	10.04
CARBON CONVERSION TO GAS, %	60.48	55.04	50.21	63.17	49.63	62.27	62.28	68.36	68.68	67.02
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	24.18	19.36	20.92	19.81	19.19	18.41	16.85	18.64	18.21	20.47
CARBON-DIOXIDE	15.42	14.57	15.30	18.11	14.82	12.66	11.2A	13.83	15.04	14.97
ETHYLENE	4.15	4.31	4.88	4.83	4.48	4.79	4.89	4.83	4.63	4.8A
ETHANE	.31	.71	.42	1.27	.77	.64	.A2	.73	.91	.72
ACETYLENE								.28	.73	.37
METHANE	14.42	15.83	16.81	15.75	17.13	17.06	17.01	1A.8A	1A.81	in 12
CARBON MONOXIDE	41.91	45.23	42.57	42.23	43.61	46.46	47.38	43.37	44.61	43.51
H2/CO RATIO	58	.43	.49	.47	.44	.40	.35	.43	.41	.47
PRODUCT GAS HHV, BTU/SCF	131	450	151	459	461	170	471	4a3	4A2	4A1
BTU YIELD, BTU/LB-MAF	4777	4417	4051	5038	3798	6037	4866	4967	4721	4626
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	83.63	83.63	84.42	84.50	84.09	83.41	84.18	86.66	83.47	84.33
FIXED CARBON	15.93	15.93	14.96	15.00	15.45	16.66	16.26	13.74	15.72	14.8A
ASH	.44	.44	.63	.5	.46	.93	.61	.61	.61	.89
CARBON	51.13	51.13	60.96	50.37	50.41	48.63	50.12	49.83	4A.84	49.8A
HYDROGEN	5.73	5.73	5.70	5.88	6.13	6.00	8.23	6.42	6.30	6.33
NITROGEN	.14	.14	.13	.1A	.3A	.31	.13	.1A	.1a	.12
CHLORINE	.02	.02	.02	.03	.03	.03	.03	.06	.02	.02
SULFUR	.03	.03	.02	.00	.00	.00	.03	.04	.00	.00
OXYGEN (by difference)	42.51	42.51	42.56	43.03	42.58	43.10	42.86	43.08	42.77	42.84
BTU/LB (DRY)	8513	8513	8300	8394	8464	8280	8402	838A	8A00	84A0

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	91	92	93	94	95	96				
LAB BOOK RUN NUMBER	3.22A	3.22B	3.23	3.24	3.25A	3.25B				
RUN CONDITION	N-S	N-S	N-S	N-S	N-S	N-S				
WOOD TYPE	S OAK	S OAK	SAWDUST	SAWDUST	S OAK	S OAK				
MOISTURE, %	10	10	40	35	5	5				
WOOD FEED RATE, LB/HR	275	275	340	618	200	1008				
DRY FEED RATE, LB/HR	248	248	204	337	190	958				
GASIFIER TEMPERATURE, F	1374	1388	1290	1326	1628	1313.5				
GASIFIER PRESSURE, PSIG	3.6	4.1	3.4	3.2	5	5				
NITROGEN FEED, LB/HR	124	124	128	62	115	0				
STEAM RATE, LB/HR	429	434	430	417	411	357				
GASIFIER GAS VELOCITY, FT/SEC	15.49	15.34	15.05	13.83	15.61	9.77				
COMBUSTOR TEMPERATURE, F	1555	1605	1620	1768	1766	1675				
COMBUSTOR PRESSURE, PSIG	3.3	3.8	3.4	3	4.8	5.4				
COMBUSTOR GAS VELOCITY, FT/SEC	2.19	2.23	1.39	1.24	1.79	1.96				
PRODUCT GAS RATE, SCF/LB-MAF	9.98	9.75	8.82	6.31	14.86	9.34				
CARBON CONVERSION TO GAS, %	57.53	57.25	47.52	40.57	77.76	55.90	.00	.00	.00	.00
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	19.38	20.04	14.16	16.81	26.41	15.92				
CARBON DIOXIDE	14.95	14.92	12.02	10.53	14.75	15.05				
ETHYLENE	4.41	4.51	4.74	4.96	4.02	3.66				
ETHANE	.82	.78	1.06	.93	.00	1.24				
ACETYLENE	.24	.24	.00	.25	.50	.53				
METHANE	16.42	16.58	16.26	15.61	14.75	16.48				
CARBON MONOXIDE	43.80	43.01	51.75	50.92	40.06	47.11				
H2/CO RATIO	.44	.47	.27	.33	.68	.34				
PRODUCT GAS HHV, BTU/SCF	459	461	472	476	436	458	0	0	0	0
BTU YIELD, BTU/LB-MAF	4578	4493	4160	3001	6472	4279				
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	85.07	85.07	81.92	80.65						
FIXED CARBON	14.26	14.26	17.43	18.78						
ASH	.67	.67	.65	.67						
CARBON	49.81	49.81	53.92	53.11	50.00	50.00				
HYDROGEN	6.05	6.05	6.53	6.50						
NITROGEN	.11	.11	.21	.18						
CHLORINE	.02	.02	.04	.04						
SULFUR	.01	.01	.04	.03						
OXYGEN(by difference)	43.33	43.33	38.61	39.57						
BTU/LB (DRY)	8448	8448	8011	8085						

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

FOREST RESIDUE GASIFICATION RUM RESULTS - 10 INCH GASIFIER

	97	98	99	100	101	102	103	104	105	106
TEST NUMBER	97	98	99	100	101	102	103	104	105	106
LAB BOOK RUN NUMBER	4.2	4.4A	4.4B	4.4C	4.5	4.6	4.7	4.8A	4.8B	4.9
RUN CONDITION	N-S	N-S	N-S	S	N-S	N-S	N-S	N-S	S	N-S
WOOD TYPE	GUM	GUM	GUM	GUM	HARD	HARD	HARD	HARD	HARD	HARD
MOISTURE, %	7.17	4.18	4.18	4.18	5.5	4.42	5.74	4.69	4.59	4
WOOD FEED RATE, LB/HR	358.5	387.7	505	1122	460	3PO	270	420	424	306
DRY FEED RATE, LB/HR	333	371	484	1075	425	373	349	401	405	294
GASIFIER TEMPERATURE, F	1542	1558	1580	1431	1623	1612	1507	1651	1627	1544
GASIFIER PRESSURE, PSIG	6.8	6.4	7	7.4	7	6.6	5.8	3.4	6	3.8
NITROGEN FEED, LBIHR	119.03	107.11	102.78	105.32	107.63	133.60	107.36	09.27	.00	104.72
STEAM RATE, LB/HR	459	439	453	391	468	432	407	447	372	449
GASIFIER GAS VELOCITY, FT/SEC	15.14	14.75	14.61	11.91	15.87	15.27	13.89	18.08	11.3A	1A.97
COMBUSTOR TEMPERATURE, F	1738	1788	1768	1765	1833	1754	1647	1893	1836	1694
COMBUSTOR PRESSURE, PSIG	6.8	6	7.2	7.2	7.4	5.7	6.4	3.8	6.8	3.8
COMBUSTOR GAS VELOCITY, FT/SEC	2.88	3.48	3.64	3.63	3.67	3.56	3.58	3.39	3.73	3.18
PRODUCT GAS RATE, SCF/LB-MF	11.11	12.72	11.87	11.38	14.45	12.02	11.78	13.83	14.57	12.99
CARBON CONVERSION TO GAS, %	65.18	67.11	63.66	66.12	77.16	86.39	68.98	73.14	77.63	68.88
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	24.99	26.08	25.4	19.1	24.58	21.9	20.87	24.87	25.48	22.47
CARBON DIOXIDE	15.025	14.44	14.33	15.1	13.21	13.99	14.02	13.7	13.41	12.54
ETHYLENE	4.45	4.22	4.1	4.27	4.06	4.6	4.71	4.09	3.84	4.47
ETHANE	.40	.35	.33	.94	.23	.52	.47	.11	.21	.36
ACETYLENE	.29	.30	.30	.07	.37	.29	.35	.67	.40	.34
METHANE	14.80	15.50	15.98	18.00	18.76	16.74	16.88	16.81	16.45	16.45
CARBON MONOXIDE	40.05	39.13	39.5	42.57	40.75	41.98	42.72	41.05	40.21	43.17
H2/CO RATIO	.62	.67	.64	.45	.60	.52	.49	.61	.63	.62
PRODUCT GAS HHV, BTU/SCF	442	446	447	467	455	463	465	447	450	463
BTU YIELD, BTU/LB-MAF	4915	5668	5306	5315	6575	5559	5478	6178	6552	6020
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	83.15	82.92	82.92	82.92	83.80	85.20	84.20	85.60	85.60	85.2A
FIXED CARBON	15.57	14.84	14.84	14.84	14.99	14.21	14.54	13.95	13.95	14.07
ASH	1.28	2.24	2.24	2.24	1.21	.59	1.28	.45	.45	.67
CARBON	50.88	49.98	49.98	49.98	50.22	50.50	50.37	51.30	51.30	50.82
HYDROGEN	6.06	5.95	5.95	5.95	5.90	5.93	6.00	6.12	6.12	6.08
NITROGEN	.15	.18	.18	.18	.06	.18	.19	.11	.11	.14
CHLORINE	.02	.02	.02	.02	.02	.01	.01	.03	.03	.03
SULFUR	.04	.12	.12	.12	.09	.01	.02	.10	.10	.20
OXYGEN(by difference)	41.57	41.51	41.51	41.51	42.50	42.78	42.15	41.89	41.89	42.08
BTU/LB (DRY)	8475	8380	8380	8380	8417	8383	8191	8488	8488	8488

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	107	108	109	110	111	112	113	114	115
LAB BOOK RUN NUMBER	4.11	4.12	4.14	4.15	4.17A	4.17B	4.19A	4.19B	4.19C
RUN CONDITION	N-S	N-S	N-S	N-S	N-S	N-S	N	N	N
WOOD TYPE	HARD	PINE	HOG	PINE	PINE	PINE	PINE	PINE	PINE
MOISTURE, %	5	5	3	9.7	5	5	5.79	5.79	5.79
WOOD FEED RATE, LB/HR	357	494	312.5	407.5	1207	1405	400	400	1179
DRY FEED RATE, LB/HR	339	469	303	368	1147	1335	380	300	1120
GASIFIER TEMPERATURE, F	1650	1615	1500	1579	1350	1313	1707	1764	1600
GASIFIER PRESSURE, PSIG	4.8	5.2	3.6	3.8	6.2	6.2	8	8.6	9.6
NITROGEN FEED, LB/HR	92.76	77.96	113.15	101.19	108.63	108.63	88.70	88.70	88.70
STEAM RATE, LB/HR	439	440	345	331	346	346	0	0	0
GASIFIER GAS VELOCITY, FT/SEC	16.38	15.52	13.44	13.41	10.03	10.70	1.65	1.65	1.47
COMBUSTOR TEMPERATURE, F	1920	1992	1630	1743	1640	1590	1900	1900	1900
COMBUSTOR PRESSURE, PSIG	4.8	4.8	4	4	6.4	6.4	8.4	8.6	8.6
COMBUSTOR GAS VELOCITY, FT/SEC	4.43	3.48	2.94	2.24	2.61	2.55	2.84	2.81	2.70
PRODUCT GAS RATE, SCF/LB-MF	14.93	11.42	9.96	13.03	9.65	9.35	16.06	15.47	13.08
CARBON CONVERSION TO GAS, %	76.62	71.54	59.73	70.93	67.37	54.48	81.43	80.99	74.23
PRODUCT GAS COMPOSITION, % N2 FREE									
HYDROGEN	26.39	26.09	21.54	22.99	16.18	15.2	24.81	23.19	18.18
CARBON DIOXIDE	12.55	10.52	9.07	9.62	12.43	12.3	8.91	7.89	8.25
ETHYLENE	3.69	4.21	4.96	5.22	4.36	4.22	3.71	3.67	5.57
ETHANE	.14	.29	.59	.34	1.35	1.42	.25	.30	.75
ACETYLENE	.49	.42	.27	.41	.17	.18	.24	.20	.16
METHANE	15.65	14.19	15.19	14.99	16.05	15.98	15.98	16.77	17.04
CARBON MONOXIDE	41.09	44.28	48.38	46.43	49.46	50.72	46.18	47.98	49.05
H2/CO RATIO	.64	.59	.45	.60	.33	.30	.64	.48	.37
PRODUCT GAS HHV, BTU/SCF	445	450	473	471	470	469	457	468	494
BTU YIELD, BTU/LB-MAF	6644	5135	4714	6144	4538	4388	7343	7213	6462
WOOD ANALYSIS, % DRY BASIS									
VOLATILE MATTER	85.37		70.29	85.04	84.34	84.34	84.34	84.34	84.34
FIXED CARBON	14.11		12.82	14.83	15.22	15.22	15.22	16.22	15.32
ASH	.52		16.89	.13	.44	.44	.44	.44	.44
CARBON	50.84		45.36	51.27	52.44	52.44	62.44	83.44	63.44
HYDROGEN	6.07		5.83	6.19	6.36	8.36	6.36	6.38	8.36
NITROGEN	.22		.18	.13	.14	.14	.14	.14	.14
CHLORINE	.03		.03	.02	.01	.01	.01	.01	.01
SULFUR	.44		.02	.13	.27	.27	.31	.27	.37
OXYGEN (by difference)	41.88		31.89	42.13	40.34	40.34	40.34	40.34	40.34
BTU/LB (DRY)	8469		7681	8748	8901	8901	8901	8901	8901

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	117	118	119	120	121	122	123	124	125	126
LAB BOOK RUN NUMBER	51A	51B	52A	52B	52C	54A	54B	55A	55A	55A
RUN CONDITION	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE
WOOD TYPE	N-S	N-S	N-S	N-S	N-S	N-S	N-S	N-S	N-S	N-S
WOOD MOISTURE, %	7.8	7.8	5.82	5.82	5.82	4.97	4.4	4.4	4.4	4.4
WOOD FEED RATE, LB/H ¹⁰	270.3	1239	434.8	1362.3	1986.4	497.9	1897	504.7	1224	1449
DRY FEED RATE, LB/HR	249	1142	409	1233	1873	498	1814	485	1175	1775
GASIFIER TEMPERATURE, F	1725	1609	1684	1610	1553	1765	1512	1750	1663	1618
GASIFIER PRESSURE, PSIG	5.22	6.2	5.2	6.8	14.4	6.6	10.4	8	9.72	12.05
NITROGEN FEED, LB/HR	96.28	00	101.70	00	00	00	00	92.2A	7A.7A	12.7A
STEAM RATE, LB/HR	374	374	367	378	423	110	34	175	134	0
GASIFIER GAS VELOCITY, FT/SEC	17.40	11.24	14.15	11.04	8.88	3.59	82	175	134	0
COMBUSTOR TEMPERATURE, F	1870	1780	1872	1820	1760	1985	1959	1981	1980	1910
COMBUSTOR PRESSURE, PSIG	5.2	6.8	5.2	7.6	15.2	7.6	11.2	8	8	12
COMBUSTOR GAS VELOCITY, FT/SEC	3.17	2.82	2.88	2.61	1.90	3.28	2.79	2.19	2.08	2.08
PRODUCT GAS RATE, SCF/LB-MF	16.14	12.99	14.43	14.53	11.88	15.25	10.95	15.11	12.7A	11.4A
CARBON CONVERSION TO GAS, %	72.06	65.77	72.00	74.17	61.82	75.04	62.38	78.12	71.34	65.87
PRODUCT GAS COMPOSITION, % N ₂ FUEL										
HYDROGEN	27.97	20.96	24.66	20.23	18.16	26.98	17.1	23.4	19.59	17.1
CARBON DIOXIDE	10.46	9.3	9.03	9.18	9.58	6.5	8.65	7.13	8.83	10.6A
ETHYLENE	3.66	5.43	4.18	5.36	5.3	3.03	4.84	3.74	5.18	4.83
ETHANE	.00	.44	.15	.64	.85	.46	1.16	.16	.66	1.24
ACETYLENE	.50	.26	.45	.23	.65	.37	.14	.37	.17	.13
METHANE	14.12	15.82	14.85	15.78	16.24	14.37	16.48	15.45	1A.21	1A.33
CARBON MONOXIDE	43.29	47.69	46.67	48.71	49.41	48.34	50.63	49.74	49.25	50.3
H ₂ /CO RATIO	.65	.44	.53	.42	.37	.5A	.34	.47	.40	.34
PRODUCT GAS HHV, BTU/SCF	439	481	457	481	486	444	485	460	4A3	4A4
BTU YIELD, BTU/LB-MAF	7083	6249	6589	6896	6721	6768	5314	6956	6177	5638
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER ¹⁰	84.79	84.79	85.35	85.35	85.35	84.27	84.27	84.66	84.66	84.66
FIXED CARBON	14.89	14.89	14.58	14.58	14.58	15.29	15.29	15.21	15.21	15.21
ASH	32.32	32.32	32.32	32.32	32.32	32.32	32.32	32.32	32.32	32.32
CARBON	52.55	52.55	52.88	52.88	52.88	52.02	52.02	51.89	51.89	51.89
HYDROGEN	6.42	6.42	6.29	6.29	6.29	6.13	6.13	6.16	6.16	6.16
NITROGEN	1.14	1.14	1.18	1.18	1.18	1.23	1.23	1.2	1.2	1.2
CHLORINE	.01	.01	.03	.03	.03	.01	.01	.02	.02	.02
SULFUR	.24	.24	.02	.02	.02	.01	.01	.03	.03	.03
OXYGEN (by difference)	40.32	40.32	40.62	40.62	40.62	41.16	41.16	41.65	41.65	41.65
BTU/LB (DRY)	8992	8992	8843	8843	8843	8652	8652	8652	8652	8652

B - TWI
C - INTERNAL CATALYST
N - NITROGEN
S - STEAM

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	127	128	129	130	131	132	133	134	135	136
LAB BOOK RUN NUMBER	5.6A	5.6B	5.6C	5.8	5.9	5.11A	5.11B	5.12A	5.12B	5.17
RUN CONDITION	N	N	N	N	N	N	N	N	N	N-S
WOOD TYPE	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE	PINE
WOOD MOISTURE, %	12.41	12.41	12.41	10.81	7.07	6.61	6.61	6.28	6.28	5.92
WOOD FEED RATE, LB/HR	540	382	128	1970	578	636	1013	613	1888	391
DRY FEED RATE, LB/HR	473	335	112	1757	537	594	946	574	1864	368
GASIFIER TEMPERATURE, F	1756	1771	1797	1423	1760	1680	1644	1776	1423	1857
GASIFIER PRESSURE, PSIG	7.4	7.2	7.2	10.8	8.1	7.38	7.74	9.9	11.7	6.3
NITROGEN FEED, LB/HR	98.67	73.35	277.09	78.27	116.97	94.88	A6.43	152.88	152.88	A7.84
STEAM RATE, LB/HR	0	0	0	0	0	0	0	0	0	365
GASIFIER GAS VELOCITY, FT/SEC	1.93	1.46	5.57	1.13	2.20	1.79	1.75	2.71	3.17	14.17
COMBUSTOR TEMPERATURE, F	1950	1950	1960	1950	1910	1995	1970	1900	1980	2159
COMBUSTOR PRESSURE, PSIG	7.2	7	7	9.8	8	5.7	5.8	8.4	R.8	8
COMBUSTOR GAS VELOCITY, FT/SEC	2.57	2.59	2.55	1.83	1.84	2.35	2.24	2.13	2.00	7.34
PRODUCT GAS RATE, SCF/LB-MF	17.75	17.24	18.55	10.68	17.87	14.50	13.15	18.58	A.87	20.02
CARBON CONVERSION TO GAS, %	85.85	86.98	88.56	61.32	86.62	77.32	72.89	84.54	56.67	85.67
PRODUCT GAS COMPOSITION, % N2 FREE										
HYDROGEN	23.29	23.81	28.39	14.46	25.85	21.24	18.24	23.88	15.81	35.57
CARBON DIOXIDE	7.46	7.51	7.09	11.27	7.24	7.74	8.54	7.6	11.87	10.43
ETHYLENE	3.92	3.86	2.35	4.08	2.89	4.8	5.7	3.46	4.39	1.33
ETHANE	.23	.26	.00	1.60	.00	.38	.59	.20	1.55	.00
ACETYLENE	.28	.30	.63	.30	.32	.23	.20	.28	.48	.39
METHANE	15.75	15.45	13.56	15.82	15.17	18.41	16.40	18.18	18.12	17.84
CARBON MONOXIDE	49.07	48.81	47.47	52.45	48.52	49.22	49.33	48.38	49.77	39.74
H2/CO RATIO	.47	.49	.60	.28	.53	.43	.38	.50	.32	.an
PRODUCT GAS HHV, BTU/SCF	464	461	429	474	444	480	482	460	478	3A7
BTU YIELD, BTU/LB-MAF	8230	7954	7958	5052	7943	6957	6485	7633	4707	7953
WOOD ANALYSIS, % DRY BASIS										
VOLATILE MATTER	83.89	83.89	83.89	83.83	83.98	84.52	84.52	83.93	83.93	84.39
FIXED CARBON	15.78	15.78	15.78	15.82	15.65	15.1	15.1	15.82	15.82	16.28
ASH	.33	.33	.33	.35	.37	.38	.38	.25	.25	.35
CARBON	52.37	52.37	52.37	52.96	52.42	52.28	52.28	52.54	52.54	52.48
HYDROGEN	6.04	6.04	6.04	6.16	6.02	6.09	6.09	6.04	6.04	5.99
NITROGEN	.02	.02	.02	.13	.1	.14	.14	.04	.04	.05
CHLORINE	.02	.02	.02	.02	.03	.03	.03	.01	.01	.01
SULFUR	.25	.25	.25	.24	.09	.25	.25	.57	.57	.48
OXYGEN (by difference)	40.97	40.97	40.97	40.12	40.97	40.84	40.84	40.55	40.55	40.63
BTU/LB (DRY)	8739	8739	8739	8900	8816	8735	8735	8885	8885	8701

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

FOREST RESIDUE GASIFICATION RUN RESULTS - 10 INCH GASIFIER

TEST NUMBER	137	138	139	140	141	142
LAB BOOK RUN NUMBER	5.14	5.15	5.16	5.17A	5.17B	5.18
RUN CONDITION	N-S	N-S	N-S	N-S	N-S	N-S
WOOD TYPE	PINE	PINE	SAWDUST	BARK	HOOFUCEL	PINE
WOOD MOISTURE, %	7.71	8.01	13.67	12.3	33.85	10
WOOD FEED RATE, LB/HR	429	336	2116	300	551	447
DRY FEED RATE, LB/HR	396	309	1827	263	364	402
GASIFIER TEMPERATURE, F	1839	1851	1316	1786	1687	1749
GASIFIER PRESSURE, PSIG	7.56	6.66	13.1	3.6	5.2	3.8
NITROGEN FEED, LB/HR	79.63	85.96	86.18	86.66	86.02	84.48
STEAM RATE, LB/HR	368	355	377	321	335	333
GASIFIER GAS VELOCITY, FT/SEC	13.14	13.47	a.38	14.02	12.77	14.04
COMBUSTOR TEMPERATURE, F	2040	2070	1860	1900	1030	1960
COMBUSTOR PRESSURE, PSIG	6.6	7.4	11.6	4.6	4.6	4.8
COMBUSTOR GAS VELOCITY, FT/SEC	2.35	2.09	1.85	2.25	2.24	2.21
PRODUCT GAS RATE, SCF/LB-MF	21.08	21.43	0.85	10.60	14.84	15.97
CARBON CONVERSION TO GAS, %	04.42	95.09	52.51	95.67	78.44	77.35
PRODUCT GAS COMPOSITION, % N2 FREE						
HYDROGEN	33.31	33.89	13.03	30.48	32.25	27.86
CARBON DIOXIDE	9.65	8.95	12.13	10.89	14.75	9.19
ETHYLENE	2	1.59	3.63	3.26	4.17	3.39
ETHANE	.00	.00	1.68	.48	.00	.00
ACETYLENE	.47	.35	.00	1.08	.48	.44
METHANE	12.93	12.80	14.74	13.43	12.33	13.83
CARBON MONOXIDE	41.64	42.42	54.79	40.41	38.03	45.3
H2/CO RATIO	80	80	.24	.75	.90	.62
PRODUCT GAS HHV, BTU/SCF	112	407	455	441	419	437
BTU YIELD, BTU/LB-MAF	8683	8714	4484	8648	6221	6977
WOOD ANALYSIS, % DRY BASIS						
VOLATILE MATTER	87.07	85.70	80.45	80.23	81.02	85.70
FIXED CARBON	12.47	14.07	18.32	16.7R	16.83	14.07
ASH	.46	.23	1.33	3.98	2.15	.23
CARBON	51.55	51.58	53.10	50.36	60.73	51.58
HYDROGEN	5.98	5.78	6.04	5.83	5.71	5.78
NITROGEN	1.1	.06	.17	.11	.33	.08
CHLORINE	.02	.02	.03	.03	.03	.02
SULFUR	.02	.01	.02	.07	.05	.01
OXYGEN (by difference)	41.86	42.32	39.31	38.62	41	42.32
BTU/LB (DRY)	8815	8834	8925	8404	8462	8834

B - TWI
 C - INTERNAL CATALYST
 N - NITROGEN
 S - STEAM

APPENDIX B

SAMPLE CALCULATION AND STATISTICAL ANALYSIS

APPENDIX B

SAMPLE CALCULATIONS AND STATISTICAL ANALYSIS

PRU Calculations

Calculation of the major flows and carbon conversion are accomplished as follows:

The major flows of steam, nitrogen, air, and natural gas are all based on pressure drop across an orifice. This gives an equation of the type

$$\text{Flow} = K \sqrt{\rho \Delta P}$$

where K is a constant calculated from the orifice and pipe sizes, ρ is density calculated from the material and its temperature and pressure. P is the pressure drop across the orifice.

Product gas flow is calculated from the nitrogen flow and product gas composition using a ratio of nitrogen content of the product gas to nitrogen flow into the gasifier.

$$\text{Product Gas Flow} = \text{Nitrogen Flow} / \text{Fraction Nitrogen in Product Gas}$$

The most important calculation is carbon conversion.

$$\text{Carbon Conversion} = \left(\frac{\text{Moles } N_2 \text{ Into Gasifier}}{N_2 \text{ Content in Product}} \times \frac{\text{Carbon Content in Product} \cdot 12}{\text{Wood Feed Rate} \times \text{Wood Carbon Content}} \right)$$

Carbon conversion is calculated by taking the amount of carbon in the product gas divided by the feed carbon. The values are calculated from the product gas composition, flows of nitrogen and wood feed, and feed analysis for carbon. Much effort is given to the accuracy of this calculation since carbon conversion is directly related to gas production and efficiency. This concern led us to conduct a statistical analysis of the data.

Statistical Analysis of Experimental Data

Examination of all the data generated during this program indicates that carbon conversion is primarily a function of gasifier temperature. However, there is a considerable amount of scatter in this data as shown in Figure 8-1. Therefore, the following questions were posed.

- (1) Is the scatter observed statistically consistent with the methods of measurement, analytical techniques, and calculational methods employed?
- (2) What is the best fit to the data and what probability is there that the actual (as opposed to the calculated) carbon conversion will fall within certain values?
- (3) How can the methods of raw data measurement and analytical techniques be improved to reduce the scatter in calculated carbon conversion?
- (4) Are there other parameters involved that perhaps have a second order effect on carbon conversion?

The first step in the statistical analysis was to eliminate data that, for known reasons, was faulty. Then, carbon conversion, the prime measure of performance, was plotted versus temperature, the main independent variable. Figure B-1 shows the linear regression and that the actual data are linear with gasifier temperature. A standard deviation of 5.8 percent is calculated for carbon conversion with respect to the regression. This is seen graphically by the band bounded by parallel lines in Figure B-1. This shows that another single observation (experimental test run) has a 68 percent probability of falling within this band. A precision analysis was conducted, calculating the variance in carbon conversion as a function of the variance in the operating parameters. This analysis indicates that the scatter observed is easily due to normal experimental errors.

What might be of more critical importance to the design and scale-up efforts is the accuracy of the regression line. That is, if the test set

FIGURE B-1

(INCLUDES BOTH 6 AND 10 IN. GASIFIER DATA)

were to be replicated, where would the new regression line be? This would indicate how close the given experimentally determined line is to the theoretical one. Figure 8-2 presents the regression line with 95 percent confidence limits (i.e., there is a 95 percent probability that the theoretical line or replicate will fall within these bounds). As seen there is excellent agreement, especially in the range where operation is expected. This is almost an expected result in a statistical analysis, that means or averages behave much better with less scatter than single observations, almost by a factor of three in the present case.

It should be noted that performance, whether it be a pilot scale or commercial plant, will follow a curve not a single observation. Nothing can be said of the variance of a single observation in a commercial plant since the measurement techniques may vary from what is presently used. However, the average or theoretical curve should be the same. The only assumption implicit here is that the variances seen are random, which has been verified to some extent.

Contributions to Error

In order to understand and possibly decrease the experimental errors occurring, the relative contribution of individual measurement error to the overall variance in carbon conversion was studied. The total deviation in a dependent variable is the square root of the sum of the relative variance in each independent variable used to calculate it. This analysis also serves as a check on the calculation procedure. Since carbon conversion is the prime measurement of performance it was studied to determine contributions to the variance in the calculated carbon conversions. The measurements used in this calculation are: GC measurements of gas carbon and nitrogen content, nitrogen feed to the gasifier, feed rate of wood, and carbon content of the wood. Each of these component's contribution to the variance was determined. This is seen on a relative basis in Figure B-3. The error in the carbon conversion calculation is most sensitive to the measurements of the nitrogen in the product gas and nitrogen flow. It is somewhat less sensitive to carbon content of the product gas and the feed rate and relatively insensitive to measurement variances in the feed carbon content.

REGRESSION ANALYSIS

FIGURE B-2

SIGMA-2 CONTRIBUTION

FIGURE B-3

Second Order Effects

Although temperature appears to have the strongest effect of any independent variables, the data was analyzed for second order effects. To do this, a so called "statistical tree"^(a) was constructed. Carbon conversion was again chosen as the performance predictor of importance. All independent variables were individually and equally divided into high and low values with a mean carbon conversion calculated in both ranges. Variables used included: temperature, feed rate, steam rate, feed moisture, gasifier pressure, inlet velocity, and air to the combustor, the last serving as a check for bias since air rate should have no direct effect on conversion. The variable, temperature, which resulted in the greatest difference (most effect on carbon conversion) between these means became the top level of the "tree". Then the process was repeated separately for only high temperature points then low temperature points. The portrayal in Figure B-4 shows the dominant variable affecting carbon conversion at each level.

A surprising observation is the complete dominance of temperature in high temperature regions masking effects from the other variables studied. However, at low temperatures feed moisture has a significant effect on conversion. This may be due to the reduced rate of moisture vaporization at lower temperatures. Feed rate appears as a minor effect in that lower rates favor marginally higher conversions. At low temperature and high moisture conditions inlet velocity had a slightly greater effect than temperature. That high inlet velocities favored conversion may be due in part to the increased heat input supplied by the steam which helps vaporize the moisture at the lower temperatures. However, these effects are obscured somewhat since inlet velocities varied only over a narrow range at the lower temperatures.

The main conclusion here is that at the 65 to 75 percent control conversion levels of interest for commercial designs temperature has by far the dominant effect suggesting heat transfer in the gasifier will be the primary contributor to conversion, while lower wood feed rates marginally enhanced conversion.

(a) This approach was suggested by Dr. Ralph E. Thomas, retired Battelle staff member.

[illegible]

FIGURE B-4. STATISTICAL TREE ANALYSIS

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