The Development of a 20-Inch Indirect Fired Fluidized Bed Gasifier

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March 1988

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SUMMARY

This report discusses the design, fabrication and operation of a 20" I.D. fluidized bed gasifier producing medium Btu gas. The reactor is indirectly heated using 30 x 1-inch U-tubes inserted in the inert bed. The U-tubes are heated using flue gases produced from a propane burner system located at the bottom of the reactor. The feed material was dry wood chips fed into the bed with a 6" auger. The experiments included two different feed points. The reactor was fed both into the bed and at the top of the bed. The fluidizing medium was superheated steam. The steam was superheated to 1000°F using an electric preheater and a regenerative system included in the exhaust system. The gas produced from the 20" reactor was passed through a cyclone for char removal and routed to the flare for combustion and disposal.

The parameters measured during the experimental runs were:

1. wood feed rate,
2. steam flow rate,
3. steam temperatures,
4. bed temperatures,
5. free board temperatures,
6. product gas temperatures,
7. bed differential pressures,
8. char production,

9. gas production,
10. gas analyses, and
11. tar production.

The parameters measured in the laboratory were:

1. moisture contents (wood and char),
2. ash contents (wood and char), and
3. tar content.

All these measurements were only taken after steady state performance was assured. The minimum time for steady state was four hours. Several runs were made in the extended format including continuous operations up to 120 hours. The system operating performance was excellent with system control very easy to maintain. The important operating characteristics were the bed temperature, the steam-to-wood ratio and the bed differential pressure.

The data analysis included the system performance showing the relationship between the important output parameters and the bed temperature as well as the steam-to-wood ratio. Figures are included showing the gas heating value, H_2/CO ratio, tar production, gas production and gasification efficiency versus the bed temperature at different steam-to-wood ratios when feeding into the bed. The data also discusses the difference in system performance as a function of feed point with improvements demonstrated when feeding to the top of the bed.

This data was used to conceptualize a commercial scale system. The preliminary process schematics and design parameters are discussed and shown in the table and figure.

In conclusion, the system design and the operation appear satisfactory for a scale-up and commercialization. Additional work reported in the appendix suggests even better performance using a recycle concept. The recycle is accomplished by using injectors driven by the fluidizing steam. The addition of recycle suggests the system ability to adjust the H_2/CO ratio and reduction in the tar and char levels by the increased residence times in the reactor. The reduction of tar and char levels will increase the gas production thus improve the gasification efficiency. Additional work needs to be undertaken to demonstrate the flexibility of the recycle system.

TABLE OF CONTENT

	PAGE
SUMMARY.....	iii
LIST OF ILLUSTRATIONS.....	ix
LIST OF TABLES.....	x
INTRODUCTION.....	1
SYSTEM DESCRIPTION.....	1
Feed System.....	3
Gasifier.....	7
Instrumentation and Control.....	10
SYSTEM OPERATION.....	13
EXPERIMENTAL RESULTS AND ANALYSIS.....	15
Analysis of the Performance of the Gasifier System....	15
Quality of Data.....	16
Physical Effects.....	18
Chemical Effects.....	19
Sensitivity of Operational Parameters.....	21
Comparison of Top Feed and Bottom Feed.....	27
COMMERCIAL SYSTEMS.....	30
CONCLUSION.....	36
REFERENCES.....	38

TABLE OF CONTENT (cont.)

	PAGE
APPENDICES	
A. TABLES.....	41
B. DATA REDUCTION PROGRAM.....	99
C. RECYCLE AND ALTERNATE FEED.....	110
Process Description and Operation of	
Recycle System.....	111
Results and Discussion.....	116
Alternate Feed Material.....	123
Summary.....	126

LIST OF ILLUSTRATIONS

FIGURE	PAGE
1. INDIRECT FIRED FLUID BED WOOD GASIFIER.....	2
2. FIRED TUBES.....	4
3. LOWER AND UPPER FEED SYSTEMS.....	5
4. UPPER FEEDER DESIGN.....	6
5. GASIFIER SYSTEM FLOW DIAGRAM.....	9
6. INSTRUMENTATION DIAGRAM.....	11
7. HEATING VALUE VS. TEMPERATURE.....	22
8. H ₂ /CO RATIO VS. TEMPERATURE.....	23
9. TAR PRODUCTION VS. TEMPERATURE.....	24
10. GAS PRODUCTION VS. TEMPERATURE.....	25
11. GASIFICATION EFFICIENCY VS. TEMPERATURE.....	28
12. COMPARISON OF HEATING VALUE VS. TEMPERATURE.....	29
13. COMPARISON OF H ₂ /CO RATIO VS. TEMPERATURE.....	31
14. COMPARISON OF EFFICIENCY VS. TEMPERATURE.....	32
15. COMMERCIAL SYSTEM FLOW DIAGRAM.....	33
C.1. STEAM INJECTOR DETAIL DIMENSIONS.....	113
C.2. DIAGRAM FOR RECYCLE FLOW MEASUREMENT.....	115
C.3. H ₂ /CO RATIO VS. TEMPERATURE FOR TOP FEED RUNS....	119
C.4. COMPARISONS OF GAS COMPOSITION BETWEEN WOOD AND RICE HULLS.....	125

LIST OF TABLES

TABLE	PAGE
A.I. UNIT SPECIFICATIONS.....	42
A.II. TEST RESULTS.....	43
A.III. CARBON BALANCE.....	51
A.IV. HYDROGEN BALANCE.....	57
A.V. OXYGEN BALANCE.....	63
A.VI. SENSIBLE HEAT.....	69
A.VII. HEAT OF COMBUSTION.....	75
A.VIII. HEAT TRANSFER DATA.....	81
A.IX. HEAT BALANCE.....	87
A.X. SCALE-UP SPECIFICATIONS.....	93
C.I. RECYCLE TEST RESULTS.....	117

INTRODUCTION

The production of medium BTU gas from biomass material has received considerable attention during the last several years. The procedures normally used in these production processes have included either indirect heating of the biomass material (1) or the use of pure oxygen with steam in the gasifier (2, 3). The steam is used for temperature control, reaction with char to produce H_2 , and possible shifting of the gas produced. Depending upon the end use of the gas (either as a fuel gas or as synthesis gas), the constituents can be controlled by using bed catalyst (4) or other techniques, including increased pressures, reduced residence times, and variations in other important reactor variables. The University of Missouri-Rolla Gasification Research on Wood (GROW) project has designed, built, and operated a system using indirect heating to produce medium BTU gas.

SYSTEM DESCRIPTION

The present indirect fired pilot gasifier is shown in Figure 1. This unit has been run over 4000 hours during the past four years. It has been proven to be very rugged and easy to operate. The system consists of several components which will be described in detail.

FIGURE 1. INDIRECT FIRED FLUID BED WOOD GASIFIER

Feed System

During the experimentation two feed locations have been investigated resulting in a final reactor configuration containing two separate feeder designs. The feeder used for the significant part of the data is a simple 6-in auger feeding into the reactor at the bottom of the tubes. Because of the close spacing of the 1-in fire tubes (see Figure 2), the feed material was injected tangentially to the reactor shell into the largest void volume. The auger is a 8-in pitch stainless steel screw, driven by a variable speed Reeves drive. The screw is located at the bottom of a large oval-shaped container. Figure 3 shows the lower feeder system with the weighed input going into the feed hopper. The weighing system is a suspended 55-gallon barrel with an adjustable trigger bottom release, which permits accurate determination of the feed rate and the total feed supplied to the reactor. With this feeder design the system could be operated with feed rates varying from 50 to 750 lb/hr.

The second feeder, used in later tests, was easier to use but more complex in design. This feeder design is shown in Figure 4. The unit includes a live bottom bin driven by a chain drive from a variable speed motor. The adjustable speed live bottom permits control of the wood feed to the reactor by feeding the injector auger. The injector runs at a constant speed, transporting the metered product to the

FIGURE 2. FIRED TUBES

FIGURE 3. LOWER AND UPPER FEED SYSTEM

FIGURE 4. UPPER FEEDER DESIGN

reactor. The wood material is fed to the reactor just at the top of the tubes, eliminating most of the problems with tube spacing and bed mixing. The design also provides the capability of a deeper wood storage, helping to eliminate gas feedback from the reactor to the feed bin. At the present time, the reactor can be run in either configuration or with both feed systems operating simultaneously.

Gasifier

The reactor part of the gasifier system, as shown in Figure 1, consists of three major components: a fluidized bed reactor, a fire tube bundle, and two propane burners mounted in a combustion chamber. The reactor is a stainless steel cylinder measuring 20-in I.D. and 17-ft high, with the tube bundle inserted and several access ports.

The tube design, as shown in Figure 2, was formed from 30 U-tubes, using 0.824-in I.D. x 1.05-in O.D. rolled alloy tubes (RA-330). All the tubes were made to pass through the distributor plate (plate A) and were circumferentially welded on the tube sheet (plate B). Originally ceramic tubes were used as an inlet section to protect the metal tubes from the high-temperature flue gas. The ceramic tubes were inserted into the larger stainless steel sleeves on plate B. The larger sleeves were packed to seal the tubes from leaking. The sleeves on plate C were packed for the same reason and also to allow for expansion. After several

runs, all of the ceramic tubes cracked and were replaced by the 1-in RA-330 tubes welded to plate B. The inlet section was formed by pounding in a wet-type refractory around PVC pipe. The initial start-up eliminated the PVC pipe, resulting in a refractory lined inlet.

The assembled bundle was slipped into the opening of the tee section of the Van Packer combustion chamber, a large section of prefabricated chimney, and set on a compressible gasket. The two propane burners, as shown in Figure 1, were mounted on the ends of the tee section. The burners were capable of producing 2 M BTU/hr at maximum firing. The burner control initially sensed the bed temperature and the over-temperature control sensed the combustion chamber temperature. Later the system was modified to control the burner chamber temperature. The unit specifications including the inert bed material are shown in Table A.I.

The entire gasifier system is shown in Figure 5. The steam is supplied by a 45-hp low pressure boiler. The boiler output is superheated to 1000°F using an electrical preheater with the capacity of 0.6 M BTU/hr. Later flow arrangements permitted the steam to pass through the superheater via a pipe inserted in the flue gas stack and into the reactor. This simple energy recovery system permitted an additional 200°F superheat to the steam. The wood is stored in a large bin which is used to supply a

FIGURE 5. GASIFIER SYSTEM FLOW DIAGRAM

weigh barrel. The weighed feed is dumped into the feed hopper, which augers the wood into the bed of the reactor.

The produced gas passes through the cyclone for char removal and enters the venturi scrubber, where much of the tar produced in the reaction is condensed by the scrubber water. Water in the scrubber system is pumped in a closed loop from the scrubber through a settling tank, where the tar is allowed to settle out, and to the fin-fan cooler piped directly to the scrubber, completing the closed loop system.

The cleaned gas from the scrubber passes through the tar drop-out drum and the demister for further cleaning. The cleaned product gas is then routed to the flare for disposal with part of the gas being continually sampled. Basically, the downstream system is the same as that used in the original 40-in gasifier (5). In later experimental runs the produced gas bypassed the scrubber, going directly to the flare, thus eliminating the problems of scrubber water disposal.

Instrumentation and Control

The instrumentation includes the temperature, pressure, feed, and flow measurements required to provide the data necessary to determine system performance. These necessary measurements are shown in Figure 6.

FIGURE 6. INSTRUMENTATION DIAGRAM

Wood feed rate was determined by timing the batches of the weighed material dumped into the feed hopper. The moisture and ash contents were recorded as well. Steam flow was measured by use of an orifice and checked with metering the water into the boiler. In later work, a turbine flow meter was installed in the steam line. The meter monitored both flow and pressure then reported the corrected instantaneous flow rate and also integrated the rate with time into the totalizer. The cleaned product gas was measured using an orifice plate, which has been calibrated with compressed air and corrected for the product gas. Product gas was monitored with a Beckman continuous gas analyzer for carbon monoxide and oxygen content. A Carle gas chromatograph was used to check the composition of the gas. Later, a turbine flow meter was installed in this line and used to measure gas flow. The meter monitored the pressure and temperature of the gas and used these readings to calculate the mass flow rate.

Char separated from the cyclone was weighed, and the sample analyzed to determine the ash content. It was assumed that all the ash in the wood remains as ash in the char at steady-state. The tar content was estimated using total carbon method in water, which was collected by bubbling product gas through a series of water-filled glass cylinders and metered with a wet test meter. Typical samples of wood, char, and tar were periodically sent for carbon, hydrogen, and oxygen analysis. Water content of the

product gas was estimated by measuring the dew point. This measurement was checked by the increased amount of condensed water during tar sample collection. The scrubber water flow was also measured along with its temperature. The propane consumption and the combustion air flow permitted the amount of heat exchanged to be calculated.

Reactor pressures were measured at the top of the reactor, at the top of the fixed bed, and at the steam inlet. From the differential pressures between these three points, the bed expansion can be monitored easily. This accompanied with the temperature measurements along the reactor provided evidence whether nonuniformities or agglomeration were occurring in the reactor. Samples of the bed were obtained from a sampling tube. The bed-char level was determined by ashing the drawn sample. Such information was useful in establishing the rate of solids conversion in the bed. Steady-state conditions were checked with periodical recording of temperatures, pressures, flow rates, and the gas analyses.

SYSTEM OPERATION

The operation of the reactor requires an initial start-up of the compressor, boiler, preheater, and propane burners. The compressor supplies air to the reactor to fluidize the bed and to transfer the heat from the tubes to the bed during start-up. The propane burners and the

electrical preheater are used to elevate the bed temperature. On reaching 900-1000°F, the air supply is shut off and steam is used to fluidize the bed. The reactor temperature begins to drop, but with the careful adjustment of the burner temperature control the temperature can be stabilized at approximately 1100-1400°F depending on the desired operating temperature. During the start-up operation, wood is supplied at a slow rate to prevent backflow of hot gas through the feed auger. Then the wood feed rate and steam rate are adjusted to the desired level, while maintaining the desired reactor temperature. As steady-state is approached, the product gas is sampled and analyzed periodically.

It is important to prevent melting, fluxing or cementing of the bed particles during the run. For this reason, care should be taken to prevent any of the following:

1. Temperature in the bed above 1600°F.
2. Accumulation of excessive char or carbon in the bed.
3. Accumulation of excessive wood ash in the bed. As a precaution to prevent high temperature in the bed, the bed temperatures were closely monitored. In addition, there is always steam flow through the bed during operation helping to absorb heat, diluting the bleeding air, and preventing combustion hot spots.

Accumulation of char may occur during some runs, and should be burned out slowly with air at the end of an

operational period. Normally, before starting the next run, air was used as a fluidizing medium to burn out the char and to help bring up the temperature. A build up of ash can melt and/or cement the bed particles together. The bed was occasionally inspected and sampled for any signs of fusing or agglomeration. If these were noted, the bed was replaced.

In summary, the unit has been operated for over 4000 hours, with continuous runs of up to 5 days. The unit is easy to operate and flexible in its functions.

EXPERIMENTAL RESULTS AND ANALYSIS

Table A.II presents the operating data and the results for all of the test runs. The reactor bed temperature is the average temperature of T5, T6, and T7 (Fig. 6). The computer program used for data reduction is included in Appendix B.

Analysis of the Performance of the Gasifier System

The experimental data provide considerable insight into the important gasification parameters. The parameters that can be varied include: bed temperature from 1051°F to the maximum of 1466°F, wood feed rate from 100 to 400 lb/hr, and steam rate from 171 to 590 lb/hr. The effects of the parameters studied will be discussed later.

Quality of Data. Using the installed instrumentation, data were obtained during each run to develop the overall material and energy balances. The calculations were performed using the results of the ultimate analyses of wood, tar, and char, shown as followings:

	WOOD	TAR	CHAR
WEIGHT % (DRY BASIS):			
C	48.19	69.15	87.40
H	6.60	6.68	5.64
O	43.63	22.32	0.26
ASH	1.02	1.84	6.70

HEATING VALUE:

BTU/LB (DRY BASIS)	8319	10066	15798
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The heating value of the char is calculated from the heating values of C and H contained in the char. The steady state heat balance around the reactor can be found as:

$$HTIN = HTOUT$$

$$HTIN = HVW + HVSI + HFT$$

$$HTOUT = HVG + HVT + HVC + HVSO$$

$$GSEFF = HVGC / HVW \times 100\%$$

$$THEFF = HVGC / HTIN \times 100\%$$

where

HTIN = Total heat entering reactor

HTOUT = Total heat exiting reactor

HVW = Total heating value of wood

HVSI = Total heat of vaporization and sensible heat of steam entering reactor

HFT = Total heat transferred from the fired tubes

HVG = Total heating value and sensible heat of dry gas

HVT = Total heating value and sensible heat of tar

HVC = Total heating value and sensible heat of char

HVSO = Total heat of vaporization and sensible heat of steam exiting reactor

GSEFF = Cold gasification efficiency of the reactor (Table A.VII)

HVGC = Total heating value (heat of combustion) of cold dry product gas

THEFF = Thermal efficiency of the reactor (Table A.IX)

The results of the reduced data are given in Tables A.II through A.IX. The accuracy can be improved by installing a pressure regulator at the boiler outlet. The "ON-OFF" boiler control produces a cyclic flow variation in the entire system. The only flow measurement not responding to this variation is that of combustion-air flow when operated under the "excess air" mode. The temperature profiles were consistent except in those cases where the bed was only partially fluidized. Tar production, especially at low operating temperature, caused potential plugging in the piping and in the tar sample line, resulting in less accurate measurements of the steam outlet. These factors, in addition to the feed rate, are the principal ones influencing the material and energy balances.

The deviations between input and output of carbon, hydrogen and oxygen at steady state conditions, as listed in Tables A.III-V, are within 10%, while the higher deviation of energy (see Table A.IX) caused by heat losses from the reactor to the atmosphere.

Physical Effects. Chemical and physical processes in which fluidized beds are involved often require heat transfer. Heat can be introduced to the bed by combustion via fluidizing air or by heat transfer via a hot internal surface. This research was concerned with the use of internal heat transfer tubes. As most fluidized beds are vertical cylinders, a vertical tube bundle is easier to be mounted in such a vessel than a horizontal one. However, the heat transfer coefficient is lower for vertical tube bundle.

Genetti and Knudsen (6) postulated that a particle at the bed temperature moves from the bulk of the bed toward the heat transfer surface at a higher temperature. The particle gains energy from the fluid flowing around it while it is in the vicinity of the surface. After a period of time the particle returns to the bulk of the bed, where it releases its energy. The heat transfer coefficient depends on the duration of the particle approaching hot surface resulting from the degree of fluidization. The tube-side, bed-side, and overall heat transfer coefficients are shown in Table A.VIII.

During experimental operation, the propane burners were operated to provide a high temperature gas to the fired tubes thus maximized the rate of heat transfer. However, limitations are imposed by the maximum allowable temperature for the alloy used for the fired tubes, which is 2250°F (7). The apparent overall heat transfer coefficient (U_o) was calculated without considering external heat losses. As shown in Table A.VIII, the overall heat transfer coefficient of the first nine runs ranged from 5.4-8.3 BTU/hr-ft²-°F. During these runs the burners were operated in the "on ratio" mode, which supplied a flue gas flow of 850 to 1200 lb/hr. In an attempt to increase the overall heat transfer coefficient, the burners were operated in the "excess air" mode, which supplied a less variable flue gas flow up to 2000 lb/hr. This increased the velocity of the flue gas inside the tubes thereby increased the heat transfer process. The results are demonstrated in runs 10 and 11. The tube-side heat transfer coefficient was greatly increased but this increase has little effect on the overall coefficient.

Chemical Effects. The gasification mechanism involves a number of reactions, which occur simultaneously in the reactor. These include the thermal decomposition of wood into CO, CO₂, H₂, and many organic oxygen-containing compounds, which are subsequently decomposed. The lower molecular weight species present in the product gas mixture

are H_2 , CO , CO_2 , and hydrocarbons, including CH_4 , C_2H_4 , C_2H_6 , C_3^+ . The following reactions take place between these species:

1. Carbon-Steam Reaction.

2. Carbon-Carbon Dioxide Reaction.

3. Hydrogenation.

4. Water-Gas Shift Reaction.

5. Methanation.

6. Pyrolysis.

The key reactions are those between carbon and steam or carbon dioxide. These two reactions are endothermic, resulting in the need of energy supply through indirect heating in order to drive the desired gasification reaction. The water-gas shift reaction increases the proportion of hydrogen in the synthesis gas. Part of hydrogen then combines with carbon to produce methane via hydrogenation.

The last reaction is pyrolysis, the thermal decomposition of wood. Thermal decomposition (8), however, is also a feature of partial combustion, hydrogenation, and reaction with steam, as well as reaction with other materials wherein conversion takes place as the temperature

is elevated. This produces a range of products depending on the temperature and the rate of heating.

In order to determine the gasifier performance and the effects of the independent gasification parameters to an extent that would allow designing of a gasification system, it is necessary to be able to estimate the heating value of the gas on a dry basis, and the production rates of produced gas, tar, and char. All of these should be estimated from independent design variables, such as the reactor bed temperature and the steam-to-wood ratio.

Sensitivity of Operational Parameters

The reactor bed temperature has primary influence on the gasification process and was chosen as the main variable for comparison. The steam-to-wood ratio has a secondary effect and is used as a parameter on each figure for the bottom feed runs. Three levels of steam-to-wood ratio were used and designated by low, medium, and high corresponding to <1.2 , >1.2 but <2.0 , and >2.0 , respectively.

The interdependence of these variables are shown in Figures 7 through 10. Figure 7 shows that the higher heating value (HHV) of the product gas decreases with increasing bed temperature at any steam-to-wood ratio. This is caused primarily by an increase in the H_2/CO ratio at the higher temperature as seen in Figure 8. This results from the impact of the water-gas shift reaction and secondary

FIGURE 7. HEATING VALUE VS. TEMPERATURE

FIGURE 8. H₂/CO RATIO VS. TEMPERATURE

FIGURE 9. TAR PRODUCTION VS. TEMPERATURE

FIGURE 10. GAS PRODUCTION VS. TEMPERATURE

decomposition of higher hydrocarbons at the higher temperature. Operation at the medium steam-to-wood ratio produced the highest heating values. Low ratios resulted in minimal fluidization while high ratios produced a lean fluidized bed. Figure 9 shows the amount of tar produced per unit mass of dry wood at various bed temperatures. The plot indicates that tar production is lower at higher temperatures. As wood enters the reactor, it partially decomposes at low temperature, producing tar. Some of the tar undergoes further decomposition at the higher temperatures as it rises in the reactor. Char production is about the same at all temperatures with limited fluctuations from the cyclone efficiency caused by different steam rates.

The amounts of char and tar produced affect the total energy content of the product gas. Conservation of energy requires that the decrease in tar production at essentially constant char rates must mean an increase in product gas energy per unit feed. Because the actual heating value of the gas decreased with temperature, as seen in Figure 7, the amount of gas produced per unit feed must increase. This is shown in Figure 10. The increase in total energy of product gas per unit feed is directly related to the efficiency of the gasification process which is expressed in terms of the thermal conversion. The gasification efficiency is the fraction of wood's heating value appearing in the product gas, which is defined as the total heating value of the product gas divided by the total heating value of the wood.

As seen in Figure 11, the gasification efficiency increases with the bed temperature, ranging from 58 to 81 percent.

Steam consumption is a parameter of economic and operational significance. The minimum steam rate of 170 lb/hr was fixed by the need to maintain the fluidization of the bed. The steam-to-wood ratio was varied from 0.7 to 2.8. Best operation for a high heating value appear to be at a medium steam-to-wood ratio of 1.5. This condition is sufficient to keep the bed well fluidized. Lower steam rates severely limited fluidization. Higher rates favored slightly higher H_2/CO ratio thereby lowering the heating value of the product gas.

Comparison of Top Feed and Bottom Feed

Figure 12 is a graph of the HHV of the gas produced versus the bed temperature. The figure shows the results of top feed runs (no run of high steam-to-wood ratio was operated under this condition) and bottom feed runs (only runs of low and medium steam-to-wood ratios are included). It is as expected that the reactor system produces different types of gas in the two configurations. Gas from top feed runs is higher in HHV and shows an increase in HHV with increasing temperature. Gas from bottom feed runs shows the opposite effect. Gas from the upper feeder is higher in methane and lower in hydrogen than gas from the lower feeder.

FIGURE 11. GASIFICATION EFFICIENCY VS. TEMPERATURE

FIGURE 12. COMPARISON OF HEATING VALUE VS. TEMPERATURE

Another important parameter in the steam driven gasification process is shown in Figure 13. The figure shows graphs of the H_2/CO ratio of the product gas versus the bed temperature. Again the comparison shows the different behavior between the two feed points. As stated earlier, the H_2 level is reduced in the top feed system and the slopes, as a function of the bed temperature, are totally different. This is primarily due to the lack of secondary reactions in the upper feed process. Figure 14 shows the gasification efficiency of the two feed systems, and the top feed runs show higher efficiency with the same dependence on temperature. The gasification efficiencies are listed in Table A.VII.

COMMERCIAL SYSTEMS

The reactor design for the commercial adaptation of the indirect heated gasifier technology is shown in Figure 15. The design has been carefully evaluated to provide excellent flexibility, low cost, and high conversion efficiencies. The reactor system becomes an integrated power plant facility with capability of producing medium BTU gas, steam, and low BTU gas. The unit is constructed by placing the medium BTU gasifier on top of a low BTU gasifier. The low BTU gasifier is the same technology developed in the original gasification work at the GROW facility (9) and is discussed in great detail in the Phase I report. The

FIGURE 13. COMPARISON OF H₂/CO RATIO VS. TEMPERATURE

FIGURE 14. COMPARISON OF EFFICIENCY VS. TEMPERATURE

FIGURE 15. COMMERCIAL SYSTEM FLOW DIAGRAM

addition of combustion air to a combustor at the bottom of the medium BTU gasifier permits the controlled flue gas temperature necessary for the heat transfer through the tubes in the medium BTU system. The radiation shield will protect the tube section entrance from the radiant energy but will permit its transfer to the top of the bed of the low BTU gasifier. This radiant component will increase the gas quality in the gasifier.

The quantity of combustion air added will be that amount required to obtain the necessary flue gas temperature but not enough to produce an oxidizing atmosphere. The lack of oxygen in the flue gas stream should help to protect the heat transfer tubes from burnout. At the same time the air addition will be sufficient to burn the tar and char in the gas stream, providing a clean gas for the heat transfer process. The wood to the low BTU gasifier will be the as-received material at the commercial site and the gasifier will also convert the more reactive char from the steam driven system, preventing any disposal problems and reducing wood needs. The unit will be refractory lined and insulated to be as energy and cost efficient as possible.

The design of the medium BTU gasifier has been changed from the U-tube concept to the more simple straight tubes. The reactor should be relatively short in height to permit the majority portions of the tubes to be submerged in the bed at the full load operating condition. The suggested tube spacing and other important design features are

calculated and listed in Table A.X. These factors listed are for a 10 or 40 M BTU/hr plant. The distributor will be constructed of 3/4-in ceramic balls. The inert material should be matched with the feed material specification. The tubes should use turbulators to improve the gas-side heat transfer coefficient. The feed point should be in the middle of the expanded bed. The produced gas passes through a two-stage cyclone with the first stage returning the active steam produced char to the low BTU gasifier, as shown in Figure 15. The second stage will remove the ash from the produced gas before the gas is cleaned of tar. The cleaning process may not be necessary in the commercial system. The tubes will be RA-330 alloy and must be permitted to expand in the upper tubesheet, as is typical in the design of a fire tube boiler.

The exhaust flue gas leaves the medium BTU gasifier and enters the steam production section. At this point additional combustion air is added, raising the flue gas temperature and totally oxidizing the low BTU gas. The high temperature flue gas passes through a convective superheater bank upon entering the convective boiler bank. The waste energy of the flue gas is sufficient to produce the steam required to run the process. The plant is sufficiently flexible to produce additional steam using the concept of a reducing atmosphere through the tubes and over-designing the low BTU gasifier. The flue gas passes through the steam generation section into the economizer. Table A.X shows the

temperature and the important variables in this process. The remaining energy in the flue gas stream appears sufficient to dry the wood required for the medium BTU gasifier.

In summary, the medium BTU gasifier design is a complex system totally based on proven technologies. The design is flexible and energy efficient and has the potential for an excellent operating characteristic while supplying a commercial facility with medium BTU gas.

CONCLUSION

In conclusion, the indirect fired gasification system has been proven to be a flexible one. The operation has demonstrated a large variation in the operating parameters while maintaining a consistent gas production and wood conversion capability.

The data has demonstrated that the most important system parameter is the bed temperature. The system design, including the fluidization features, in conjunction with the indirect firing, permit easy and consistent control of the bed temperature. The heating value of product gas, the gasification efficiency and gas yield, are all directly related to this bed temperature. In the range of temperatures between 1161 and 1450°F, the heating value varied from 550 to 350 BTU/SCF for bottom feed runs. This variation is a direct result of the higher ratio of H_2/CO at

the higher temperatures when feeding into the bed. This result is somewhat different in the top feed situation with H_2/CO remaining relatively constant and the heating value increasing slightly with temperature. The H_2/CO ratio increases primarily due to the secondary reactions and water-gas reaction at higher temperature. In all cases, the gas yield and efficiencies increase with increasing bed temperature.

These results were evidenced for all steam-to-wood ratios. However the steam-to-wood ratio has a secondary effect related to bed fluidization. The steam-to-wood ratio was varied from 0.7 to 2.8 in the studies. The operation at high steam rate is not economical, while a minimum steam rate is required for an adequate degree of fluidization. The best operation for high heating value is near 1.5 lb of steam per lb of dry wood. Future work with recycle of product gas will lower the steam demands for fluidization as well as the energy input and further improve the system efficiency and economics.

These data were used to design the commercial scale system for the output of 10 or 40 M BTU/hr. The integration of a low BTU gasifier with the indirect fired system resulted in a compact and efficient power plant package using a single biomass fuel. Another advantage in this approach is the fact that the low BTU gasifier will convert the char produced in the medium BTU system. This reduces the environmental problems with the non-integrated approach.

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TABLE OF CONTENT OF APPENDICES

	PAGE
APPENDICES	
A. TABLES OF SPECIFICATIONS AND REDUCED DATA.....	41
B. DATA REDUCTION PROGRAM (FORTRAN IV).....	99
C. ADDITIONAL WORK ON RECYCLE AND ALTERNATE FEED.....	110
Process Description and Operation of	
Recycle System.....	111
Results and Discussion of recycle runs.....	116
Alternate Feed Material.....	123
Summary.....	126

APPENDIX A
TABLES

TABLE A.I. UNIT SPECIFICATIONS

Reactor:

Type: Fluidized bed

Size: 20 in I.D. x 17 ft height

Material: 1 section of 4 ft carbon steel

2 sections of 6 ft stainless steel (SS)

6 in steam inlet section (SS)

4 in flue gas outlet section (SS)

Bed: 6 in height of gravel,

22 in height of 0.0115 in mean diameter sand

Feed: 10 - 60 mesh oak sawdust with 7% moisture

Fired Tubes:

Type: 30 vertical U-tubes

Size: 0.824 in I.D. x 1.05 in O.D. x 7 ft long

Material: Rolled alloy (RA-330)

Burners:

Type: Eclipse 104 MVTA nozzle mixing medium velocity
burners, 1 M Btu/hr each

Fuel: Propane

Chamber: Van Packer Tee with 22 in I.D. projection,
30 in I.D. for burners

TABLE A.II. TEST RESULTS

RUN No.	MASS IN		OPERATING DATA				MASS OUT					GAS ANALYSIS								
	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO	C ₃ +	O ₂	N ₂
1	155	370	1054	574	40.0	885	98.4	365.7	24.4	19.9	429	.1050	.2309	.0299	.0094	.0969	.4532	.0275	.0102	.0370
2	120	370	1140	627	32.7	828	79.9	372.9	11.9	19.1	425	.1606	.2025	.0327	.0101	.0631	.4616	.0291	.0108	.0297
3	210	460	1051	584	43.3	913	124.7	459.9	46.0	20.9	466	.1251	.2149	.0321	.0097	.1284	.4205	.0289	.0105	.0299
4	210	525	1184	1073	54.2	1038	140.5	549.6	23.1	22.5	521	.2064	.1889	.0370	.0102	.1544	.3620	.0327	.0006	.0078
5	240	560	1105	1122	46.2	828	147.1	553.8	35.0	24.6	537	.1738	.1989	.0346	.0109	.1655	.3720	.0387	.0001	.0055
6	315	590	1112	1121	54.2	990	231.2	604.9	38.9	32.3	512	.1532	.2387	.0286	.0105	.1441	.3740	.0451	.0001	.0058
7	400	590	1084	1119	54.2	990	282.3	605.9	60.6	42.5	492	.1291	.2711	.0260	.0103	.1393	.3740	.0443	.0001	.0058
8	187	375	1237	1069	53.0	1107	138.1	400.3	17.2	18.8	538	.2257	.2165	.0348	.0111	.1717	.2828	.0415	.0028	.0132
9	210	405	1200	981	36.6	767	145.5	441.3	23.5	25.4	538	.1912	.2269	.0374	.0114	.2019	.2797	.0315	.0039	.0162
10	323	430	1359	1261	92.5	1917	248.8	435.3	33.5	37.0	442	.2645	.1588	.0228	.0102	.0930	.4240	.0205	.0006	.0057
11	323	430	1365	1213	92.5	1980	261.3	463.5	34.3	28.5	440	.2735	.1499	.0239	.0098	.0867	.4289	.0201	.0007	.0065
21	300	250	1259	830	80.1	1642	210.9	275.6	23.9	22.6	418	.2580	.1522	.0241	.0094	.0819	.4434	.0138	.0000	.0172
23	300	261	1300	782	83.4	1642	226.9	284.7	21.8	22.3	435	.2242	.1580	.0279	.0097	.0864	.4700	.0167	.0000	.0069
24	300	259	1259	817	80.1	1642	217.3	288.8	25.0	22.6	403	.2749	.1515	.0229	.0085	.0753	.4200	.0129	.0000	.0341
25	320	261	1211	769	73.1	1642	234.8	288.6	26.0	19.3	430	.2193	.1722	.0217	.0099	.0837	.4486	.0237	.0002	.0208

TABLE A.II. TEST RESULTS (CONT.)

RUN No.	MASS IN		OPERATING DATA				MASS OUT				GAS ANALYSIS									
	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆ (Mole Fraction)	CH ₄	CO	C ₃ +	O ₂	N ₂
26	301	193	1204	862	73.1	1657	230.6	227.1	24.5	13.9	444	.1817	.1798	.0223	.0117	.0938	.4789	.0240	.0001	.0078
27	218	206	1309	851	73.1	1657	180.9	204.3	18.8	14.0	391	.3054	.1656	.0221	.0074	.0637	.3992	.0129	.0008	.0229
28	216	195	1343	880	73.1	1657	186.9	181.8	20.9	8.1	371	.3671	.1648	.0202	.0059	.0560	.3501	.0089	.0003	.0268
29	216	253	1328	845	73.1	1657	169.6	258.5	17.0	12.0	404	.2721	.1641	.0258	.0086	.0640	.4389	.0139	.0011	.0102
37	212	263	1289	1062	73.1	1715	163.0	270.7	16.2	12.0	442	.1976	.1259	.0310	.0107	.0828	.5338	.0132	.0004	.0045
45	200	238	1324	1093	89.5	1715	156.7	227.8	16.7	11.3	436	.3528	.1822	.0260	.0085	.1120	.2843	.0160	.0002	.0180
48	204	253	1193	1037	75.4	1852	159.4	250.1	19.3	15.4	433	.3092	.1969	.0191	.0093	.0999	.3113	.0264	.0008	.0271
49	173	217	1207	1020	86.5	1852	130.1	221.8	15.0	9.8	452	.2428	.1830	.0261	.0103	.1231	.3560	.0217	.0012	.0358
50	200	217	1197	1039	88.0	1852	160.6	210.0	18.9	13.2	433	.2697	.1849	.0217	.0101	.1109	.3383	.0209	.0038	.0397
52	187	184	1450	1098	65.4	1917	170.6	158.7	15.9	11.3	377	.3540	.2347	.0234	.0054	.1001	.2255	.0092	.0007	.0469
53	200	184	1420	1057	65.4	1967	176.2	165.3	16.5	12.1	358	.3720	.2510	.0193	.0051	.0964	.1873	.0088	.0009	.0592
54	218	274	1195	1107	91.0	1917	170.1	278.4	18.8	8.2	442	.2234	.2137	.0228	.0105	.1284	.3641	.0190	.0031	.0178
56	202	274	1415	1118	86.5	1980	194.1	242.7	19.4	6.9	374	.4165	.2396	.0178	.0054	.0960	.1948	.0091	.0003	.0205
58	206	264	1377	1052	91.0	1980	174.6	267.4	13.4	11.7	427	.1790	.2517	.0225	.0132	.1289	.3567	.0182	.0007	.0291
64	290	191	1161	1060	92.5	1980	222.9	211.6	25.4	13.7	440	.1752	.2456	.0217	.0108	.1175	.4020	.0248	.0000	.0025

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT					GAS ANALYSIS								
RUN No.	FEED lb/hr	STEAM lb/hr	T_B F	T_S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO	C ₃ +	O ₂	N ₂
																	(Mole Fraction)			
65	253	171	1246	930	83.4	1715	196.2	191.7	19.2	8.6	472	.1766	.2086	.0292	.0109	.1286	.4164	.0258	.0001	.0039
123	100	237	1384	877	78.4	1485	83.9	241.1	8.0	7.0	398	.3525	.1690	.0260	.0078	.0700	.3438	.0110	.0021	.0178
124	100	237	1389	876	78.4	1485	81.8	243.3	8.1	7.0	383	.3863	.1783	.0233	.0068	.0670	.3032	.0098	.0025	.0228
126	100	214	1405	865	76.7	1485	82.0	199.9	8.0	6.6	398	.3653	.1715	.0260	.0058	.0753	.3307	.0104	.0006	.0144
127	100	214	1403	853	76.7	1485	81.3	200.8	8.0	6.6	392	.3774	.1736	.0251	.0054	.0723	.3195	.0099	.0008	.0161
128	100	193	1401	843	76.7	1485	84.4	186.3	8.3	5.3	399	.3684	.1717	.0256	.0058	.0778	.3279	.0099	.0003	.0125
129	100	193	1399	846	76.7	1485	84.2	187.3	8.3	5.3	391	.3785	.1757	.0243	.0055	.0753	.3131	.0097	.0008	.0173
131	100	254	1310	827	73.1	1485	76.7	256.0	7.5	6.0	408	.3006	.1798	.0265	.0079	.0703	.3809	.0167	.0013	.0159
132	100	254	1302	830	73.1	1485	76.8	255.2	7.5	6.0	411	.2966	.1771	.0254	.0095	.0688	.3834	.0184	.0020	.0189
134	100	254	1306	824	73.1	1485	78.1	258.3	7.6	6.0	407	.2932	.1782	.0270	.0081	.0665	.3927	.0169	.0016	.0159
135	100	224	1304	818	73.1	1485	82.1	219.5	8.1	6.5	415	.3016	.1762	.0261	.0095	.0691	.3815	.0190	.0014	.0157
136	100	224	1297	815	73.1	1485	82.0	214.4	7.9	6.5	422	.2830	.1741	.0272	.0084	.0666	.4046	.0223	.0009	.0130
137	100	224	1297	821	73.1	1485	81.2	221.0	8.1	6.5	406	.3181	.1805	.0246	.0077	.0695	.3651	.0174	.0014	.0158
139	100	189	1301	819	73.1	1485	80.0	195.2	7.8	4.0	414	.2978	.1788	.0258	.0101	.0701	.3869	.0176	.0006	.0124
140	100	189	1300	821	73.1	1485	79.2	195.3	7.8	4.0	414	.3104	.1830	.0249	.0080	.0714	.3693	.0198	.0003	.0130

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT				GAS ANALYSIS									
RUN No.	FEED lb/hr	STEAM lb/hr	T_B F	T_S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO	C ₃ +	O ₂	N ₂
141	100	264	1216	808	69.4	1485	77.2	256.9	8.7	5.0	419	.2519	.1882	.0235	.0112	.0651	.4179	.0245	.0012	.0166
142	100	264	1216	807	69.4	1485	76.0	258.3	8.7	5.0	420	.2731	.1913	.0220	.0104	.0678	.3865	.0271	.0022	.0196
143	100	264	1221	812	69.4	1485	78.6	257.9	8.7	5.0	439	.2427	.1838	.0245	.0110	.0615	.4272	.0338	.0008	.0147
144	100	226	1232	782	69.4	1485	81.8	219.5	9.1	4.0	409	.2438	.1755	.0231	.0093	.0600	.3954	.0287	.0135	.0507
145	100	226	1229	738	69.4	1485	81.2	219.7	9.1	4.0	423	.2546	.1854	.0242	.0092	.0623	.4176	.0283	.0017	.0168
146	100	226	1227	739	69.4	1485	80.2	219.9	9.1	4.0	424	.2699	.1933	.0236	.0095	.0660	.3919	.0288	.0007	.0163
147	100	198	1237	740	69.4	1485	85.2	197.1	9.7	3.0	424	.2658	.1932	.0235	.0098	.0696	.3906	.0275	.0021	.0180
148	100	198	1228	740	69.4	1485	85.1	199.6	9.8	3.0	419	.2769	.1993	.0231	.0096	.0740	.3815	.0239	.0001	.0115
149	100	198	1229	736	69.4	1485	85.4	201.4	9.9	3.0	419	.2831	.2040	.0230	.0093	.0769	.3751	.0228	.0000	.0059
150	336	244	1272	901	81.7	1485	243.3	294.7	22.6	18.8	458	.1989	.1680	.0281	.0104	.0939	.4714	.0255	.0002	.0035
151	285	244	1274	913	81.7	1485	218.1	264.2	20.3	16.0	458	.1989	.1680	.0281	.0104	.0939	.4714	.0255	.0002	.0035
152	283	244	1284	929	81.7	1485	217.2	263.1	20.2	15.9	458	.1989	.1680	.0281	.0104	.0939	.4714	.0255	.0002	.0035
153	310	227	1289	925	81.7	1485	225.7	274.0	21.0	15.1	468	.1930	.1576	.0291	.0108	.0994	.4794	.0260	.0005	.0044
154	283	227	1284	929	81.7	1485	209.9	254.9	19.5	13.7	468	.1930	.1576	.0291	.0108	.0994	.4794	.0260	.0005	.0044
155	277	227	1288	927	81.7	1485	209.4	254.3	19.5	13.5	468	.1930	.1576	.0291	.0108	.0994	.4794	.0260	.0005	.0044

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT					GAS ANALYSIS								
RUN No.	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO (Mole Fraction)	C ₃ +	O ₂	N ₂
156	287	200	1291	928	81.7	1485	214.7	225.1	19.8	13.7	452	.2244	.1540	.0297	.0096	.0909	.4671	.0209	.0002	.0032
157	293	200	1291	930	81.7	1485	218.7	229.2	20.2	14.0	452	.2244	.1540	.0297	.0096	.0909	.4671	.0209	.0002	.0032
158	305	200	1306	931	81.7	1485	226.0	236.9	20.8	14.5	452	.2244	.1540	.0297	.0096	.0909	.4671	.0209	.0002	.0032
161	300	242	1328	1042	76.7	1852	243.2	253.7	21.3	16.8	492	.1848	.1260	.0375	.0104	.1325	.4477	.0212	.0061	.0337
162	300	242	1291	1040	76.7	1852	237.7	249.6	22.0	16.8	474	.2054	.1453	.0331	.0098	.1275	.4134	.0216	.0074	.0366
163	300	242	1272	1037	76.7	1852	242.7	247.2	21.8	16.8	482	.1786	.1469	.0329	.0107	.1280	.4469	.0231	.0056	.0274
164	110	256	1399	1011	73.1	1852	86.3	248.9	8.3	6.2	450	.3227	.1509	.0324	.0067	.1187	.3382	.0124	.0023	.0157
165	110	256	1403	1021	73.1	1852	85.3	246.3	8.2	6.2	449	.3268	.1543	.0319	.0067	.1169	.3338	.0133	.0016	.0147
166	200	238	1379	1040	78.4	1852	150.7	249.8	14.3	13.3	455	.3139	.1564	.0310	.0072	.1201	.3462	.0148	.0010	.0093
167	200	238	1375	1045	78.4	1852	151.8	246.5	14.1	13.3	463	.2948	.1537	.0324	.0076	.1228	.3633	.0157	.0011	.0087
168	200	238	1381	1051	78.4	1852	152.3	244.1	13.9	13.3	472	.2782	.1480	.0345	.0081	.1273	.3790	.0156	.0009	.0084
169	200	214	1371	1050	78.4	1852	147.4	229.5	13.1	13.3	487	.2497	.1432	.0364	.0088	.1328	.4030	.0183	.0008	.0071
170	200	214	1361	1049	78.4	1852	146.9	228.4	13.1	13.3	489	.2490	.1435	.0366	.0089	.1333	.4011	.0189	.0009	.0077
171	200	214	1368	1046	78.4	1852	146.3	228.7	13.1	13.3	486	.2549	.1455	.0360	.0086	.1329	.3967	.0185	.0007	.0062
172	200	250	1329	1024	78.4	1852	145.5	258.0	13.5	13.1	482	.2541	.1499	.0339	.0088	.1280	.3963	.0203	.0008	.0079

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT					GAS ANALYSIS								
RUN No.	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆ (Mole Fraction)	CH ₄	CO	C ₃ +	O ₂	N ₂
173	200	250	1324	1021	78.4	1852	144.9	257.4	13.5	13.1	479	.2606	.1541	.0327	.0087	.1246	.3884	.0218	.0008	.0084
174	200	250	1320	1020	78.4	1852	146.9	255.6	13.4	13.1	489	.2383	.1492	.0339	.0093	.1285	.4094	.0231	.0009	.0074
175	200	250	1325	1022	78.4	1852	145.5	256.2	13.4	13.1	489	.2488	.1487	.0340	.0089	.1287	.3973	.0234	.0011	.0091
176	200	244	1326	1022	78.4	1852	156.6	246.8	14.8	10.5	473	.2732	.1528	.0309	.0085	.1212	.3810	.0213	.0011	.0100
177	200	244	1316	1020	78.4	1852	157.7	246.5	14.8	10.5	483	.2648	.1554	.0324	.0088	.1278	.3848	.0220	.0002	.0038
178	200	244	1321	1019	78.4	1852	159.1	246.3	14.8	10.5	488	.2556	.1531	.0329	.0092	.1281	.3946	.0233	.0001	.0032
179	200	277	1214	979	78.4	1852	147.5	278.4	16.7	11.8	486	.2178	.1621	.0285	.0101	.1256	.4177	.0278	.0007	.0097
180	200	277	1218	978	78.4	1852	146.3	279.8	16.8	11.8	481	.2357	.1696	.0279	.0097	.1226	.4001	.0279	.0001	.0064
181	200	277	1217	973	78.4	1852	151.3	277.8	16.7	11.8	502	.1930	.1609	.0302	.0106	.1263	.4390	.0328	.0004	.0062
182	200	240	1209	973	78.4	1852	149.4	249.0	16.8	9.9	492	.2186	.1691	.0272	.0103	.1249	.4103	.0323	.0003	.0070
183	200	240	1211	972	78.4	1852	152.4	248.5	16.8	9.9	501	.1958	.1614	.0281	.0105	.1229	.4403	.0351	.0002	.0058
184	200	240	1209	971	78.4	1852	151.2	249.5	16.8	9.9	485	.2043	.1590	.0279	.0100	.1218	.4203	.0312	.0026	.0229
185	200	204	1332	1034	76.7	1852	167.0	211.0	14.8	5.5	477	.1953	.1271	.0384	.0088	.1259	.4383	.0187	.0052	.0424
186	200	204	1333	1037	76.7	1852	163.7	209.5	14.7	5.5	488	.2087	.1287	.0382	.0089	.1272	.4471	.0195	.0031	.0195
187	200	229	1314	1006	81.7	1852	164.2	241.0	15.1	5.6	495	.1912	.1210	.0376	.0090	.1251	.4606	.0241	.0050	.0265

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT					GAS ANALYSIS								
RUN No.	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄ (Mole Fraction)	CO	C ₃ +	O ₂	N ₂
188	200	229	1308	1005	81.7	1852	163.0	239.1	15.0	5.6	483	.1940	.1269	.0364	.0088	.1225	.4538	.0220	.0052	.0304
189	200	229	1308	1007	81.7	1852	162.8	238.6	15.0	5.6	490	.1962	.1298	.0368	.0089	.1229	.4534	.0240	.0045	.0236
190	200	225	1288	1011	81.7	1852	159.0	233.2	14.7	5.6	488	.1956	.1310	.0359	.0092	.1253	.4537	.0226	.0047	.0221
191	200	225	1289	1009	81.7	1852	158.8	232.7	14.6	5.6	489	.1968	.1313	.0351	.0091	.1238	.4523	.0243	.0048	.0225
192	300	280	1206	1015	81.7	1852	241.8	287.1	25.8	10.6	490	.1597	.1508	.0308	.0098	.1194	.4888	.0292	.0017	.0099
193	300	280	1212	1014	81.7	1852	240.2	286.3	25.8	10.6	502	.1605	.1460	.0315	.0100	.1222	.4922	.0314	.0007	.0057
194	300	280	1208	1016	81.7	1852	238.3	286.1	25.8	10.6	501	.1688	.1471	.0312	.0098	.1210	.4852	.0319	.0006	.0046
195	300	223	1241	1045	81.7	1852	237.6	237.2	24.7	10.6	512	.1556	.1337	.0348	.0101	.1257	.5035	.0309	.0006	.0051
196	300	223	1232	1040	81.7	1852	237.4	236.1	24.5	10.6	507	.1581	.1408	.0331	.0100	.1228	.4984	.0318	.0005	.0046
197	300	223	1231	1038	81.7	1852	238.0	235.7	24.5	10.6	508	.1539	.1407	.0331	.0100	.1234	.5012	.0321	.0007	.0049
198	300	223	1229	1021	81.7	1852	237.3	234.7	24.4	10.6	510	.1580	.1458	.0322	.0099	.1211	.4942	.0348	.0004	.0036
199	300	217	1219	1017	81.7	1852	236.2	232.6	25.1	10.6	510	.1595	.1521	.0299	.0101	.1202	.4865	.0376	.0004	.0038
200	300	217	1218	1014	81.7	1852	236.4	232.5	25.1	10.6	512	.1594	.1533	.0299	.0102	.1207	.4851	.0382	.0002	.0030
201	300	217	1228	1023	81.7	1852	233.4	233.6	25.2	10.6	516	.1762	.1500	.0301	.0100	.1193	.4704	.0402	.0004	.0036
202	200	256	1293	1036	81.7	1852	155.3	265.4	14.6	8.4	514	.2229	.1381	.0351	.0089	.1240	.4341	.0332	.0003	.0034

TABLE A.II. TEST RESULTS (CONT.)

MASS IN		OPERATING DATA					MASS OUT					GAS ANALYSIS								
RUN No.	FEED lb/hr	STEAM lb/hr	T _B F	T _S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO (Mole Fraction)	C ₃ +	O ₂	N ₂
203	200	256	1296	1039	81.7	1852	156.4	264.1	14.5	8.4	521	.2104	.1365	.0359	.0092	.1274	.4438	.0344	.0001	.0023
204	200	256	1220	1001	71.3	1852	148.5	268.5	16.1	8.2	514	.1744	.1458	.0323	.0101	.1215	.4727	.0369	.0003	.0061
205	200	256	1214	1000	71.3	1852	149.0	268.1	16.1	8.2	511	.1714	.1481	.0315	.0099	.1205	.4777	.0365	.0002	.0042
206	200	244	1218	998	71.3	1852	153.7	245.8	16.6	8.0	514	.1758	.1516	.0309	.0098	.1183	.4708	.0393	.0002	.0033
207	200	244	1209	990	71.3	1852	154.7	245.8	16.6	8.0	514	.1712	.1522	.0301	.0097	.1165	.4730	.0410	.0003	.0060
208	200	220	1211	987	71.3	1852	146.6	233.0	15.7	8.0	521	.1736	.1527	.0295	.0097	.1171	.4674	.0445	.0002	.0052
209	200	220	1208	984	71.3	1852	146.9	233.6	15.8	8.0	526	.1748	.1528	.0296	.0098	.1176	.4633	.0468	.0003	.0050
210	200	250	1400	1053	78.4	1852	155.1	265.5	13.3	7.4	522	.2256	.1299	.0398	.0085	.1295	.4434	.0296	.0003	.0035
211	200	250	1382	1054	78.4	1852	155.0	265.0	13.2	7.4	530	.2134	.1281	.0403	.0088	.1323	.4409	.0330	.0001	.0033
212	200	204	1448	1064	78.4	1852	150.8	224.5	12.8	7.4	538	.2019	.1208	.0429	.0086	.1374	.4525	.0325	.0002	.0031
213	200	204	1427	1064	78.4	1852	151.1	224.0	12.8	7.4	538	.1984	.1220	.0424	.0089	.1380	.4563	.0322	.0001	.0018
214	200	195	1466	1069	74.9	1852	145.7	215.1	12.3	7.4	544	.1918	.1198	.0439	.0090	.1416	.4604	.0324	.0001	.0011
215	200	195	1440	1066	74.9	1852	146.4	215.0	12.3	7.4	548	.1893	.1199	.0424	.0089	.1401	.4605	.0361	.0001	.0029

TABLE A.III.
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
1	69.5	16.9	17.4	39.4	0.0588
2	53.8	8.3	16.7	32.2	0.0620
3	94.2	31.8	18.3	51.6	0.0787
4	94.2	16.0	19.7	62.3	0.0399
5	107.7	24.2	21.5	66.0	0.0370
6	141.3	26.9	28.2	100.6	0.1017
7	179.5	41.9	37.1	120.0	0.1086
8	84.5	11.9	16.4	60.9	0.0547
9	94.9	16.3	22.2	63.6	0.0754
10	146.0	23.2	32.3	104.9	0.0986
11	146.0	23.7	24.9	110.3	0.0883
21	136.3	16.5	19.8	86.6	-0.0983
23	136.3	15.1	19.5	95.6	-0.0445
24	136.3	17.3	19.8	86.4	-0.0938
25	145.4	18.0	16.9	96.8	-0.0945
26	136.8	17.0	12.2	97.3	-0.0759
27	99.1	13.0	12.2	71.2	-0.0267
28	98.2	14.5	7.1	70.9	-0.0581
29	98.2	11.7	10.5	68.9	-0.0711
37	96.3	11.2	10.5	70.3	-0.0452

Note: Deviation = (out - in) / in

TABLE A.III. (CONT.)
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
45	90.9	11.6	9.9	63.4	-0.0669
48	92.7	13.3	13.5	63.8	-0.0227
49	78.6	10.4	8.6	53.2	-0.0824
50	90.9	13.1	11.5	63.9	-0.0264
52	85.0	11.0	9.9	61.5	-0.0313
53	90.9	11.4	10.5	60.5	-0.0928
54	99.1	13.0	7.2	69.3	-0.0968
56	91.8	13.4	6.0	70.4	-0.0218
58	93.6	9.2	10.2	69.3	-0.0525
64	131.8	17.6	12.0	91.9	-0.0791
65	115.0	13.3	7.5	83.9	-0.0900
123	45.0	5.6	6.1	33.0	-0.0084
124	45.0	5.6	6.1	31.1	-0.0495
126	45.0	5.5	5.8	32.3	-0.0316
127	45.0	5.6	5.8	31.6	-0.0455
128	45.0	5.7	4.7	33.3	-0.0301
129	45.0	5.8	4.7	32.6	-0.0445
131	45.0	5.2	5.2	30.7	-0.0861
132	45.0	5.2	5.2	30.8	-0.0846
134	45.0	5.2	5.2	31.3	-0.0713

TABLE A.III. (CONT.)
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
135	45.0	5.6	5.7	33.2	-0.0128
136	45.0	5.4	5.7	33.6	-0.0061
137	45.0	5.6	5.7	32.4	-0.0298
139	45.0	5.4	3.5	32.4	-0.0825
140	45.0	5.4	3.5	31.9	-0.0930
141	45.0	6.0	4.4	31.4	-0.0720
142	45.0	6.0	4.4	30.7	-0.0867
143	45.0	6.0	4.4	32.7	-0.0423
144	45.0	6.3	3.5	31.7	-0.0782
145	45.0	6.3	3.5	33.1	-0.0456
146	45.0	6.3	3.5	32.6	-0.0568
147	45.0	6.7	2.6	34.5	-0.0265
148	45.0	6.8	2.6	34.5	-0.0254
149	45.0	6.8	2.6	34.7	-0.0198
150	151.2	15.6	16.5	104.5	-0.0967
151	128.3	14.0	14.0	93.7	-0.0514
152	127.4	13.9	13.9	93.3	-0.0490
153	139.5	14.5	13.2	98.0	-0.0991
154	127.4	13.5	12.0	91.2	-0.0839
155	124.7	13.5	11.8	91.0	-0.0680

TABLE A.III. (CONT.)
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
156	129.2	13.7	11.9	92.2	-0.0878
157	131.9	14.0	12.2	93.9	-0.0897
158	137.3	14.4	12.7	97.0	-0.0957
161	135.3	14.7	14.7	105.4	-0.0032
162	135.3	15.2	14.7	100.3	-0.0379
163	135.3	15.0	14.7	104.2	-0.0098
164	49.6	5.7	5.4	36.0	-0.0500
165	49.6	5.7	5.4	35.6	-0.0600
166	90.2	9.9	11.6	63.7	-0.0561
167	90.2	9.7	11.6	64.8	-0.0449
168	90.2	9.6	11.6	65.8	-0.0349
169	90.2	9.1	11.6	64.9	-0.0519
170	90.2	9.0	11.6	64.6	-0.0546
171	90.2	9.0	11.6	64.3	-0.0583
172	90.2	9.4	11.5	63.5	-0.0654
173	90.2	9.3	11.5	63.0	-0.0715
174	90.2	9.3	11.5	64.6	-0.0538
175	90.2	9.3	11.5	63.8	-0.0623
176	90.2	10.2	9.2	67.5	-0.0367
177	90.2	10.2	9.2	69.0	-0.0202

TABLE A.III. (CONT.)
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
178	90.2	10.2	9.2	70.1	-0.0084
179	90.2	11.5	10.3	64.3	-0.0451
180	90.2	11.6	10.3	63.5	-0.0530
181	90.2	11.5	10.3	67.1	-0.0138
182	90.2	11.6	8.7	65.4	-0.0496
183	90.2	11.6	8.7	67.6	-0.0256
184	90.2	11.6	8.7	65.1	-0.0527
185	88.1	10.2	4.8	71.0	-0.0232
186	88.1	10.1	4.8	71.7	-0.0163
187	89.7	10.5	4.9	72.1	-0.0257
188	89.7	10.4	4.9	70.5	-0.0443
189	89.7	10.4	4.9	71.1	-0.0373
190	89.7	10.1	4.9	69.4	-0.0590
191	89.7	10.1	4.9	69.3	-0.0603
192	134.6	17.9	9.3	106.6	-0.0066
193	134.6	17.8	9.3	107.4	-0.0009
194	134.6	17.8	9.3	106.6	-0.0071
195	134.6	17.1	9.3	107.7	-0.0045
196	134.6	17.0	9.3	106.9	-0.0106
197	134.6	17.0	9.3	107.2	-0.0084

TABLE A.III. (CONT.)
CARBON BALANCE (LB/HR)

RUN	FEED	TAR	CHAR	DRY GAS	DEVIATION
198	134.6	16.9	9.3	107.0	-0.0108
199	134.6	17.4	9.3	106.2	-0.0128
200	134.6	17.4	9.3	106.5	-0.0111
201	134.6	17.4	9.3	105.5	-0.0178
202	89.7	10.1	7.3	70.4	-0.0217
203	89.7	10.0	7.3	71.4	-0.0107
204	89.7	11.1	7.2	67.0	-0.0496
205	89.7	11.1	7.2	67.1	-0.0482
206	89.7	11.5	7.0	69.3	-0.0216
207	89.7	11.5	7.0	69.6	-0.0181
208	89.7	10.9	7.0	66.4	-0.0613
209	89.7	10.9	7.0	66.8	-0.0564
210	89.7	9.2	6.5	70.7	-0.0368
211	89.7	9.2	6.5	71.4	-0.0294
212	89.7	8.9	6.5	70.1	-0.0469
213	89.7	8.9	6.5	70.3	-0.0451
214	89.7	8.5	6.5	68.2	-0.0726
215	89.7	8.5	6.5	68.6	-0.0678

TABLE A.IV.
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
1	9.5	41.1	1.6	1.12	3.3	40.6	-0.0999
2	7.4	41.1	0.8	1.08	2.8	41.4	-0.0666
3	12.9	51.1	3.1	1.18	5.2	51.1	-0.0778
4	12.9	58.3	1.5	1.27	8.3	61.1	-0.0091
5	14.7	62.2	2.3	1.39	8.4	61.5	-0.0651
6	19.4	65.6	2.6	1.82	11.5	67.2	-0.0481
7	24.6	65.6	4.0	2.39	12.6	67.3	-0.0734
8	11.6	41.7	1.1	1.06	9.1	44.5	0.0231
9	13.0	45.0	1.6	1.43	9.4	49.0	0.0336
10	20.0	47.8	2.2	2.08	12.8	48.4	-0.0643
11	20.0	47.8	2.3	1.61	13.5	51.5	-0.0151
21	18.7	27.8	1.6	1.28	9.9	30.6	-0.1022
23	18.7	29.0	1.5	1.26	10.1	31.6	-0.1033
24	18.7	28.8	1.7	1.28	10.3	32.1	-0.0811
25	19.9	29.0	1.7	1.09	10.2	32.1	-0.1145
26	18.7	21.4	1.6	0.78	9.5	25.2	-0.1180
27	13.6	22.9	1.3	0.79	8.8	22.7	-0.1137
28	13.4	21.7	1.4	0.46	10.3	20.2	-0.1143
29	13.4	28.1	1.1	0.68	7.7	28.7	-0.1098
37	13.2	29.2	1.1	0.68	6.7	30.1	-0.1189

Note: Deviation = (out - in) / in

TABLE A. IV. (CONT.)
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
45	12.4	26.4	1.1	0.64	10.7	25.3	-0.0606
48	12.7	28.1	1.3	0.87	9.4	27.8	-0.0666
49	10.8	24.1	1.0	0.55	7.1	24.6	--0.0754
50	12.4	24.1	1.3	0.74	8.8	23.3	-0.0976
52	11.6	20.4	1.1	0.64	10.2	17.6	-0.1111
53	12.4	20.4	1.1	0.68	10.7	18.4	-0.0981
54	13.6	30.4	1.3	0.46	8.6	30.9	-0.0924
56	12.6	30.4	1.3	0.39	13.4	27.0	-0.0518
58	12.8	29.3	0.9	0.66	7.7	29.7	-0.1033
64	18.0	21.2	1.7	0.77	9.5	23.5	-0.1374
65	15.7	19.0	1.3	0.48	9.2	21.3	-0.1112
123	6.2	26.3	0.5	0.40	4.8	26.8	-0.0206
124	6.2	26.3	0.5	0.40	5.0	27.0	-0.0078
126	6.2	23.8	0.5	0.37	4.9	22.2	-0.0854
127	6.2	23.8	0.5	0.37	5.0	22.3	-0.0813
128	6.2	21.4	0.6	0.30	5.2	20.7	-0.0578
129	6.2	21.4	0.6	0.30	5.2	20.8	-0.0526
131	6.2	28.2	0.5	0.34	3.9	28.4	-0.0557
132	6.2	28.2	0.5	0.34	3.9	28.4	-0.0585
134	6.2	28.2	0.5	0.34	3.8	28.7	-0.0500

TABLE A.IV. (CONT.)
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
135	6.2	24.9	0.5	0.37	4.2	24.4	-0.0711
136	6.2	24.9	0.5	0.37	4.0	23.8	-0.0961
137	6.2	24.9	0.5	0.37	4.3	24.6	-0.0641
139	6.2	21.0	0.5	0.23	4.1	21.7	-0.0500
140	6.2	21.0	0.5	0.23	4.2	21.7	-0.0460
141	6.2	29.3	0.6	0.28	3.4	28.5	-0.0934
142	6.2	29.3	0.6	0.28	3.6	28.7	-0.0833
143	6.2	29.3	0.6	0.28	3.5	28.7	-0.0876
144	6.2	25.1	0.6	0.23	3.5	24.4	-0.1016
145	6.2	25.1	0.6	0.23	3.7	24.4	-0.0972
146	6.2	25.1	0.6	0.23	3.8	24.4	-0.0915
147	6.2	22.0	0.6	0.17	4.0	21.9	-0.0740
148	6.2	22.0	0.7	0.17	4.1	22.2	-0.0607
149	6.2	22.0	0.7	0.17	4.2	22.4	-0.0501
150	20.7	27.1	1.5	1.06	10.8	32.7	-0.0826
151	17.6	27.1	1.4	0.90	9.7	29.4	-0.1168
152	17.4	27.1	1.3	0.89	9.7	29.2	-0.1179
153	19.1	25.2	1.4	0.85	10.2	30.4	-0.0787
154	17.4	25.2	1.3	0.78	9.5	28.3	-0.1077
155	17.1	25.2	1.3	0.76	9.5	28.3	-0.1018

TABLE A.IV. (CONT.)
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
156	17.7	22.2	1.3	0.77	10.0	25.0	-0.1163
157	18.1	22.2	1.3	0.79	10.2	25.5	-0.1086
158	18.8	22.2	1.4	0.82	10.6	26.3	-0.0964
161	18.5	26.9	1.4	0.95	12.6	28.2	-0.0928
162	18.5	26.9	1.5	0.95	12.4	27.7	-0.1052
163	18.5	26.9	1.5	0.95	11.9	27.5	-0.1216
164	6.8	28.4	0.6	0.35	5.7	27.7	-0.0502
165	6.8	28.4	0.5	0.35	5.6	27.4	-0.0595
166	12.4	26.4	1.0	0.75	9.7	27.8	-0.0257
167	12.4	26.4	0.9	0.75	9.5	27.4	-0.0408
168	12.4	26.4	0.9	0.75	9.4	27.1	-0.0506
169	12.4	23.8	0.9	0.75	8.8	25.5	-0.0441
170	12.4	23.8	0.9	0.75	8.8	25.4	-0.0475
171	12.4	23.8	0.9	0.75	8.8	25.4	-0.0455
172	12.4	27.8	0.9	0.74	8.6	28.7	-0.0643
173	12.4	27.8	0.9	0.74	8.6	28.6	-0.0657
174	12.4	27.8	0.9	0.74	8.4	28.4	-0.0742
175	12.4	27.8	0.9	0.74	8.6	28.5	-0.0688
176	12.4	27.1	1.0	0.59	9.4	27.4	-0.0607
177	12.4	27.1	1.0	0.59	9.5	27.4	-0.0580

TABLE A.IV. (CONT.)
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
178	12.4	27.1	1.0	0.59	9.5	27.4	-0.0598
179	12.4	30.8	1.1	0.67	7.9	30.9	-0.0882
180	12.4	30.8	1.1	0.67	8.1	31.1	-0.0809
181	12.4	30.8	1.1	0.67	7.9	30.9	-0.0910
182	12.4	26.7	1.1	0.56	8.1	27.7	-0.0742
183	12.4	26.7	1.1	0.56	7.9	27.6	-0.0807
184	12.4	26.7	1.1	0.56	7.8	27.7	-0.0796
185	12.1	22.7	1.0	0.31	8.5	23.4	-0.0921
186	12.1	22.7	1.0	0.31	8.7	23.3	-0.0913
187	12.3	25.4	1.0	0.32	8.5	26.8	-0.0683
188	12.3	25.4	1.0	0.32	8.3	26.6	-0.0795
189	12.3	25.4	1.0	0.32	8.4	26.5	-0.0780
190	12.3	25.0	1.0	0.32	8.2	25.9	-0.0885
191	12.3	25.0	1.0	0.32	8.2	25.9	-0.0900
192	18.4	31.1	1.7	0.60	11.0	31.9	-0.1267
193	18.4	31.1	1.7	0.60	11.3	31.8	-0.1236
194	18.4	31.1	1.7	0.60	11.4	31.8	-0.1219
195	18.4	24.8	1.6	0.60	11.4	26.4	-0.1215
196	18.4	24.8	1.6	0.60	11.2	26.2	-0.1275
197	18.4	24.8	1.6	0.60	11.2	26.2	-0.1296

TABLE A.IV. (CONT.)
HYDROGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
198	18.4	24.8	1.6	0.60	11.2	26.1	-0.1311
199	18.4	24.1	1.7	0.60	11.2	25.8	-0.1229
200	18.4	24.1	1.7	0.60	11.3	25.8	-0.1219
201	18.4	24.1	1.7	0.60	11.7	26.0	-0.1094
202	12.3	28.4	1.0	0.47	8.9	29.5	-0.0586
203	12.3	28.4	1.0	0.47	8.8	29.3	-0.0629
204	12.3	28.4	1.1	0.46	7.4	29.8	-0.0819
205	12.3	28.4	1.1	0.46	7.3	29.8	-0.0855
206	12.3	27.1	1.1	0.45	7.6	27.3	-0.1080
207	12.3	27.1	1.1	0.45	7.6	27.3	-0.1095
208	12.3	24.4	1.1	0.45	7.3	25.9	-0.0925
209	12.3	24.4	1.1	0.45	7.5	26.0	-0.0875
210	12.3	27.8	0.9	0.42	9.0	29.5	-0.0430
211	12.3	27.8	0.9	0.42	9.1	29.4	-0.0431
212	12.3	22.7	0.9	0.42	8.8	24.9	-0.0402
213	12.3	22.7	0.9	0.42	8.7	24.9	-0.0435
214	12.3	21.7	0.8	0.42	8.4	23.9	-0.0536
215	12.3	21.7	0.8	0.42	8.5	23.9	-0.0534

TABLE A.V.
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
1	63.0	328.9	5.4	0.05	51.0	325.0	-0.0496
2	48.7	328.9	2.7	0.05	41.4	331.5	-0.0246
3	85.3	408.9	10.3	0.05	62.6	408.8	-0.0502
4	85.3	466.7	5.2	0.06	68.5	488.5	-0.0047
5	97.5	497.8	7.8	0.06	71.8	492.2	-0.0626
6	128.0	524.4	8.7	0.08	117.7	537.6	-0.0114
7	162.5	524.4	13.5	0.11	148.0	538.6	-0.0159
8	76.5	333.3	3.8	0.05	65.5	355.8	0.0118
9	85.9	360.0	5.3	0.07	69.1	392.3	0.0199
10	132.2	382.2	7.5	0.10	129.1	387.0	-0.0162
11	132.2	382.2	7.7	0.07	135.1	412.0	0.0423
21	123.4	222.2	5.3	0.06	109.9	245.0	-0.0018
23	123.4	232.0	4.9	0.06	119.4	253.1	0.0180
24	123.4	230.2	5.6	0.06	111.4	256.7	0.0131
25	131.7	232.0	5.8	0.05	122.1	256.5	0.0119
26	123.8	171.6	5.5	0.04	121.9	201.9	0.0597
27	89.7	183.1	4.2	0.04	95.5	181.6	-0.0090
28	88.9	173.3	4.7	0.02	98.9	161.6	-0.0294
29	88.9	224.9	3.8	0.03	90.7	229.8	-0.0013
37	87.2	233.8	3.6	0.03	85.0	240.6	-0.0075

Note: Deviation = (out - in) / in

TABLE A.V. (CONT.)
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
45	82.3	211.6	3.7	0.03	78.8	202.5	-0.0624
48	83.9	224.9	4.3	0.04	80.6	222.3	-0.0374
49	71.2	192.9	3.3	0.03	64.1	197.1	-0.0304
50	82.3	192.9	4.2	0.03	79.3	186.7	-0.0529
52	76.9	163.6	3.5	0.03	88.2	141.1	-0.0683
53	82.3	163.6	3.7	0.03	91.1	146.9	-0.0557
54	89.7	243.6	4.2	0.02	88.1	247.4	-0.0132
56	83.1	243.6	4.3	0.02	104.7	215.7	-0.0360
58	84.8	234.7	3.0	0.03	92.0	237.7	0.0087
64	119.3	169.8	5.7	0.04	120.9	188.1	0.0359
65	104.1	152.0	4.3	0.02	102.3	170.4	0.0300
123	40.8	210.7	1.8	0.02	43.8	214.3	0.0102
124	40.8	210.7	1.8	0.02	42.8	216.3	0.0141
126	40.8	190.2	1.8	0.02	43.1	177.7	-0.0600
127	40.8	190.2	1.8	0.02	42.8	178.4	-0.0581
128	40.8	171.6	1.8	0.01	44.5	165.6	-0.0288
129	40.8	171.6	1.9	0.01	44.2	166.5	-0.0257
131	40.8	225.8	1.7	0.02	40.4	227.6	-0.0099
132	40.8	225.8	1.7	0.02	40.2	226.9	-0.0136
134	40.8	225.8	1.7	0.02	41.3	229.6	0.0005

TABLE A.V. (CONT.)
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
135	40.8	199.1	1.8	0.02	43.0	195.2	-0.0236
136	40.8	199.1	1.8	0.02	43.0	190.5	-0.0423
137	40.8	199.1	1.8	0.02	42.8	196.4	-0.0190
139	40.8	168.0	1.7	0.01	42.3	173.5	0.0136
140	40.8	168.0	1.7	0.01	41.8	173.6	0.0117
141	40.8	234.7	1.9	0.01	40.8	228.3	-0.0363
142	40.8	234.7	1.9	0.01	39.7	229.6	-0.0357
143	40.8	234.7	1.9	0.01	41.0	229.2	-0.0325
144	40.8	200.9	2.0	0.01	40.3	195.1	-0.0407
145	40.8	200.9	2.0	0.01	42.6	195.3	-0.0305
146	40.8	200.9	2.0	0.01	42.1	195.4	-0.0319
147	40.8	176.0	2.2	0.01	44.6	175.2	-0.0031
148	40.8	176.0	2.2	0.01	45.3	177.4	0.0101
149	40.8	176.0	2.2	0.01	45.9	179.0	0.0202
150	136.9	216.9	5.0	0.05	126.9	262.0	0.0545
151	116.1	216.9	4.5	0.04	113.8	234.8	0.0097
152	115.3	216.9	4.5	0.04	113.3	233.9	0.0081
153	126.3	201.8	4.7	0.04	116.1	243.6	0.0523
154	115.3	201.8	4.4	0.04	108.0	226.6	0.0157
155	112.9	201.8	4.4	0.03	107.8	226.0	0.0219

TABLE A.V. (CONT.)
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
156	117.0	177.8	4.4	0.04	111.7	200.1	0.0147
157	119.4	177.8	4.5	0.04	113.7	203.8	0.0243
158	124.3	177.8	4.7	0.04	117.5	210.6	0.0399
161	122.5	215.1	4.8	0.04	113.6	225.5	-0.0304
162	122.5	215.1	4.9	0.04	112.5	221.9	-0.0435
163	122.5	215.1	4.9	0.04	117.2	219.7	-0.0364
164	44.9	227.6	1.9	0.02	42.5	221.2	-0.0473
165	44.9	227.6	1.8	0.02	42.2	218.9	-0.0567
166	81.7	211.6	3.2	0.03	75.2	222.0	-0.0137
167	81.7	211.6	3.1	0.03	75.6	219.1	-0.0224
168	81.7	211.6	3.1	0.03	75.3	217.0	-0.0303
169	81.7	190.2	2.9	0.03	72.3	204.0	-0.0143
170	81.7	190.2	2.9	0.03	71.8	203.0	-0.0195
171	81.7	190.2	2.9	0.03	71.9	203.3	-0.0181
172	81.7	222.2	3.0	0.03	71.8	229.3	-0.0353
173	81.7	222.2	3.0	0.03	71.7	228.8	-0.0372
174	81.7	222.2	3.0	0.03	72.4	227.2	-0.0404
175	81.7	222.2	3.0	0.03	71.2	227.7	-0.0425
176	81.7	216.9	3.3	0.03	77.5	219.4	-0.0316
177	81.7	216.9	3.3	0.03	78.4	219.1	-0.0294

TABLE A.V. (CONT.)
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
178	81.7	216.9	3.3	0.03	78.9	218.9	-0.0285
179	81.7	246.2	3.7	0.03	73.4	247.4	-0.0432
180	81.7	246.2	3.7	0.03	73.5	248.7	-0.0391
181	81.7	246.2	3.7	0.03	75.2	246.9	-0.0397
182	81.7	213.3	3.8	0.03	74.5	221.3	-0.0222
183	81.7	213.3	3.7	0.03	75.9	220.9	-0.0192
184	81.7	213.3	3.8	0.03	73.7	221.7	-0.0236
185	79.8	181.3	3.3	0.01	77.9	187.5	-0.0275
186	79.8	181.3	3.3	0.01	78.7	186.3	-0.0293
187	81.2	203.6	3.4	0.01	77.4	214.2	-0.0069
188	81.2	203.6	3.4	0.01	77.4	212.6	-0.0127
189	81.2	203.6	3.3	0.01	77.8	212.1	-0.0129
190	81.2	200.0	3.3	0.01	76.3	207.3	-0.0225
191	81.2	200.0	3.3	0.01	76.1	206.9	-0.0248
192	121.9	248.9	5.8	0.03	121.0	255.2	-0.0183
193	121.9	248.9	5.8	0.03	119.8	254.5	-0.0235
194	121.9	248.9	5.7	0.03	118.9	254.3	-0.0260
195	121.9	198.2	5.5	0.03	117.0	210.9	-0.0150
196	121.9	198.2	5.5	0.03	117.8	209.8	-0.0158
197	121.9	198.2	5.5	0.03	118.1	209.5	-0.0161

TABLE A.V. (CONT.)
OXYGEN BALANCE (LB/HR)

RUN	FEED	STEAM	TAR	CHAR	DRY GAS	STEAM OUT	DEVIATION
198	121.9	198.2	5.4	0.03	118.0	208.6	-0.0189
199	121.9	192.9	5.6	0.03	117.7	206.7	-0.0094
200	121.9	192.9	5.6	0.03	117.8	206.7	-0.0092
201	121.9	192.9	5.6	0.03	115.2	207.7	-0.0139
202	81.2	227.6	3.3	0.02	75.4	235.9	-0.0203
203	81.2	227.6	3.2	0.02	75.7	234.8	-0.0228
204	81.2	227.6	3.6	0.02	73.0	238.7	-0.0182
205	81.2	227.6	3.6	0.02	73.8	238.4	-0.0166
206	81.2	216.9	3.7	0.02	76.1	218.5	-0.0389
207	81.2	216.9	3.7	0.02	76.4	218.5	-0.0378
208	81.2	195.6	3.5	0.02	72.0	207.1	-0.0220
209	81.2	195.6	3.5	0.02	71.8	207.6	-0.0211
210	81.2	222.2	3.0	0.02	74.7	236.0	-0.0067
211	81.2	222.2	3.0	0.02	73.9	235.5	-0.0106
212	81.2	181.3	2.9	0.02	71.2	199.6	-0.0042
213	81.2	181.3	2.9	0.02	71.7	199.1	-0.0041
214	81.2	173.3	2.7	0.02	68.8	191.2	-0.0151
215	81.2	173.3	2.7	0.02	68.8	191.1	-0.0157

TABLE A.VI.
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
1	472349.	13388.	8673.	28464.	532902.
2	481370.	6173.	7842.	22322.	533969.
3	589361.	26493.	9550.	40237.	680246.
4	790734.	14279.	10983.	55520.	831315.
5	856072.	20550.	11441.	54031.	824277.
6	901661.	22648.	14888.	80331.	897815.
7	901119.	33935.	18849.	90276.	888349.
8	564120.	10497.	9093.	56359.	603781.
9	592855.	13485.	11536.	53745.	651674.
10	684835.	21243.	18537.	98708.	664458.
11	675341.	21824.	14339.	104735.	708421.
21	348595.	16425.	12300.	89022.	432967.
23	358170.	15278.	12359.	95524.	450227.
24	359596.	17197.	12289.	92284.	453591.
25	356609.	17462.	10265.	93994.	449728.
26	271956.	15967.	7163.	86808.	349857.
27	289232.	13396.	7870.	81244.	325246.
28	276389.	14983.	4611.	88894.	290033.
29	354524.	12165.	6796.	74444.	412332.
37	394789.	11209.	6543.	65812.	425799.

TABLE A.VI. (CONT.)
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
45	360656.	12240.	6546.	81669.	366466.
48	376869.	12462.	7864.	68002.	384253.
49	321546.	9679.	5007.	53148.	340864.
50	323443.	11772.	6509.	63489.	318668.
52	279249.	9520.	5361.	67557.	237859.
53	275779.	10028.	5800.	71159.	248595.
54	416973.	11667.	4042.	64267.	421943.
56	418360.	13590.	3792.	97361.	383787.
58	395076.	7524.	5198.	56101.	392672.
64	286534.	15032.	6413.	75471.	315707.
65	246305.	11732.	4161.	71040.	289201.
123	335592.	5113.	3527.	35341.	368472.
124	335483.	5171.	3534.	35524.	372149.
126	301843.	4935.	3239.	33963.	302446.
127	300661.	4945.	3233.	33890.	303454.
128	270269.	5103.	2598.	35104.	281674.
129	270536.	5050.	2558.	34610.	281689.
131	353822.	4701.	2962.	30118.	389144.
132	354172.	4615.	2917.	29623.	385900.
134	353471.	4617.	2884.	29455.	389035.

TABLE A.VI. (CONT.)
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
135	311105.	4900.	3151.	31468.	330593.
136	310796.	4749.	3128.	30649.	321995.
137	311414.	4855.	3102.	30917.	331045.
139	262581.	4588.	1870.	29315.	290569.
140	262755.	4610.	1878.	29569.	291127.
141	365445.	5150.	2373.	27340.	383699.
142	365323.	5194.	2381.	27737.	386177.
143	365930.	5153.	2365.	27870.	384786.
144	310140.	5397.	1884.	28623.	327460.
145	305566.	5392.	1880.	28723.	327599.
146	305669.	5413.	1886.	29009.	328187.
147	267890.	5543.	1357.	29568.	291044.
148	267890.	5518.	1334.	29236.	293109.
149	267526.	5594.	1340.	29664.	296234.
150	348198.	14782.	9744.	94589.	454800.
151	349545.	13044.	8137.	83294.	405490.
152	351340.	12925.	8041.	82478.	403099.
153	326444.	13476.	7641.	86171.	420014.
154	326862.	12396.	6900.	79148.	389170.
155	326653.	12418.	6780.	79328.	388810.

TABLE A.VI. (CONT.)
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
156	287892.	12501.	6829.	81701.	343058.
157	288076.	12733.	6971.	83218.	349425.
158	288168.	13157.	7257.	85990.	361068.
161	361040.	14602.	9123.	103358.	398090.
162	360817.	14754.	8946.	99416.	388851.
163	360483.	14527.	8896.	98445.	384282.
164	378276.	5449.	3209.	38911.	384622.
165	379453.	5508.	3277.	39516.	383479.
166	354853.	9796.	7210.	70958.	392129.
167	355401.	9694.	7230.	70631.	387333.
168	356057.	9609.	7237.	70259.	383725.
169	320054.	9076.	7270.	67190.	361369.
170	319956.	9000.	7244.	66729.	359183.
171	319660.	9031.	7257.	66870.	359986.
172	370905.	9261.	7083.	65318.	404438.
173	370560.	9215.	7063.	65097.	403176.
174	370445.	9124.	7043.	64819.	399914.
175	370675.	9153.	7050.	65018.	400941.
176	361779.	9993.	5593.	70049.	385099.
177	361554.	9896.	5546.	69883.	383468.

TABLE A.VI. (CONT.)
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
178	361442.	9877.	5540.	70006.	382972.
179	405229.	10585.	5914.	58553.	424895.
180	405101.	11221.	6233.	62534.	435088.
181	404464.	11141.	6233.	62995.	432005.
182	350438.	10456.	4886.	58371.	377627.
183	350328.	10394.	4867.	58241.	376437.
184	350218.	10455.	4876.	57918.	378109.
185	303597.	9379.	2756.	64962.	322264.
186	303878.	9456.	2797.	65860.	321718.
187	337853.	9590.	2799.	63993.	367705.
188	337747.	9478.	2787.	62856.	364377.
189	337958.	9428.	2779.	62916.	363185.
190	332469.	9272.	2796.	61809.	355757.
191	332262.	9262.	2799.	61889.	355121.
192	414254.	16161.	5254.	89171.	436288.
193	414126.	16248.	5296.	90259.	436269.
194	414383.	16154.	5270.	89684.	435205.
195	333001.	16078.	5471.	93117.	365603.
196	332488.	15715.	5376.	90920.	361541.
197	332283.	15645.	5360.	90567.	360638.

TABLE A.VI. (CONT.)
SENSIBLE HEAT (BTU/HR)

RUN	STEAM	TAR	CHAR	DRY GAS	STEAM OUTLET
198	330539.	15499.	5333.	90186.	358459.
199	321247.	15611.	5222.	87770.	352633.
200	320947.	15592.	5216.	87836.	352438.
201	321846.	15667.	5216.	88444.	354139.
202	381220.	9822.	4458.	67716.	413674.
203	381573.	9794.	4466.	67919.	411978.
204	377098.	10230.	4117.	57602.	410086.
205	376980.	10165.	4096.	57108.	408851.
206	359085.	10441.	3987.	58964.	374302.
207	358187.	10441.	3987.	59103.	374288.
208	322652.	9676.	3899.	55011.	352126.
209	322348.	9708.	3903.	55514.	353040.
210	374240.	9513.	4216.	73044.	423528.
211	374355.	9444.	4193.	72650.	421869.
212	306412.	8947.	4104.	68670.	354551.
213	306412.	8902.	4093.	68319.	353400.
214	293342.	8446.	4044.	64911.	337856.
215	293073.	8376.	4014.	64710.	336687.

TABLE A.VII.
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
1	1200473	245376	314597	536349	44.7
2	929399	120129	301801	455041	49.0
3	1626448	462927	330481	769630	47.3
4	1626448	232933	355190	1081010	66.5
5	1858797	352050	388282	1122940	60.4
6	2439672	391401	509621	1589184	65.1
7	3097996	609891	670670	1784289	57.6
8	1459203	172680	296498	1105214	75.7
9	1638677	236940	401426	1136623	69.4
10	2520441	337417	583928	1726202	68.5
11	2520441	345248	449912	1830166	72.6
21	2353445	240417	357534	1380133	58.6
23	2353445	219725	353065	1481699	63.0
24	2353445	251955	357534	1394112	59.2
25	2510341	261468	305095	1485887	59.2
26	2361290	246871	219719	1446279	61.2

Note: Gasification efficiency (GAS.EFF.) (%)

$$= \frac{\text{Heat of combustion of cold dry gas}}{\text{Heat of combustion of feed}} \times 100\%$$

TABLE A.VII. (CONT.)
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
27	1710170	189185	220837	1146981	67.1
28	1694480	210450	128712	1211887	71.5
29	1694480	171024	189850	1072262	63.3
37	1663101	163463	189493	1073768	64.6
45	1568963	168137	178767	1188639	75.8
48	1600343	194209	243123	1110588	69.4
49	1357153	150692	154633	900028	66.3
50	1568963	190262	208561	1087126	69.3
52	1466981	159785	178290	1073894	73.2
53	1568963	166357	190685	1061057	67.6
54	1710170	189139	129904	1104400	64.6
56	1584653	195444	108333	1302775	82.2
58	1616032	134604	184130	1025962	63.5
64	2274997	255566	216010	1336126	58.7
65	1984739	192992	135684	1300369	65.5
123	776995	80889	110663	573669	73.8
124	776995	81640	110663	557465	71.7
126	776995	80508	104761	571505	73.6
127	776995	80832	104761	564645	72.7
128	776995	83338	84104	592892	76.3
129	776995	83779	84104	584855	75.3

TABLE A.VII. (CONT.)
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
131	776995	75594	94432	499900	64.3
132	776995	75356	94432	502234	64.6
134	776995	76260	94432	503217	64.8
135	776995	81033	103285	545393	70.2
136	776995	79118	103285	541152	69.6
137	776995	81569	103285	536666	69.1
139	776995	78598	63447	527424	67.9
140	776995	78639	63447	527333	67.9
141	776995	87267	79677	482502	62.1
142	776995	87738	79677	487283	62.7
143	776995	87606	79677	508007	65.4
144	776995	91759	63447	495567	63.8
145	776995	91862	63447	513559	66.1
146	776995	91930	63447	515461	66.3
147	776995	97402	47216	545089	70.2
148	776995	98613	47216	545974	70.3
149	776995	99524	47216	550498	70.8
150	2610702	227424	297462	1615306	61.9
151	2214435	203888	252312	1448143	65.4
152	2198895	203037	250541	1442096	65.6
153	2408683	211484	237852	1535646	63.8

TABLE A.VII. (CONT.)
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
154	2198895	196700	217135	1428299	65.0
155	2152275	196247	212532	1425004	66.2
156	2229975	199371	215970	1460830	65.5
157	2276594	203071	220485	1487945	65.4
158	2369834	209837	229515	1537522	64.9
161	2335975	214542	266160	1791376	76.7
162	2335975	221127	266160	1699753	72.8
163	2335975	218979	266160	1710787	73.2
164	856524	83507	97592	669775	78.2
165	856524	82631	97592	661819	77.3
166	1557317	143663	209971	1166818	74.9
167	1557317	141761	209971	1170254	75.1
168	1557317	140392	209971	1180685	75.8
169	1557317	131987	209971	1146308	73.6
170	1557317	131368	209971	1144806	73.5
171	1557317	131572	209971	1140813	73.3
172	1557317	136320	207013	1115569	71.6
173	1557317	136034	207013	1107232	71.1
174	1557317	135072	207013	1122348	72.1
175	1557317	135373	207013	1124539	72.2
176	1557317	149058	165611	1196851	76.9

TABLE A.VII. (CONT.)
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
177	1557317	148886	165611	1221454	78.4
178	1557317	148744	165611	1233661	79.2
179	1557317	168129	186312	1079150	69.3
180	1557317	168985	186312	1074481	69.0
181	1557317	167788	186312	1111577	71.4
182	1557317	169157	156739	1099808	70.6
183	1557317	168858	156739	1118015	71.8
184	1557317	169490	156739	1086517	69.8
185	1520713	148668	86635	1205319	79.3
186	1520713	147648	86635	1224090	80.5
187	1548998	152481	88246	1223655	79.0
188	1548998	151310	88246	1185552	76.5
189	1548998	150972	88246	1199648	77.4
190	1548998	147577	88246	1167986	75.4
191	1548998	147262	88246	1167857	75.4
192	2323497	260126	167667	1684679	72.5
193	2323497	259394	167667	1718995	74.0
194	2323497	259210	167667	1716207	73.9
195	2323497	248342	167667	1744557	75.1
196	2323497	247114	167667	1719630	74.0
197	2323497	246753	167667	1720134	74.0

TABLE A.VII. (CONT.)
HEAT OF COMBUSTION (BTU/HR)

RUN	FEED	TAR	CHAR	DRY GAS	GAS. EFF. (%)
198	2323497	245687	167667	1717852	73.9
199	2323497	252825	167667	1703404	73.3
200	2323497	252773	167667	1708700	73.5
201	2323497	253993	167667	1730949	74.5
202	1548998	146926	132369	1223696	79.0
203	1548998	146223	132369	1235643	79.8
204	1548998	162157	129427	1101402	71.1
205	1548998	161949	129427	1093333	70.6
206	1548998	167028	126486	1133192	73.2
207	1548998	167022	126486	1132939	73.1
208	1548998	158338	126486	1089056	70.3
209	1548998	158693	126486	1103207	71.2
210	1548998	133615	117661	1243813	80.3
211	1548998	133363	117661	1259906	81.3
212	1548998	129144	117661	1238990	80.0
213	1548998	128857	117661	1235557	79.8
214	1548998	123737	117661	1200118	77.5
215	1548998	123647	117661	1208022	78.0

TABLE A.VIII.
HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
1	630.4	13.5	18.7	6.8	422867.
2	548.3	12.7	12.7	5.4	295127.
3	707.3	13.8	16.4	6.4	451504.
4	723.1	15.4	21.1	7.7	547632.
5	797.1	12.9	16.9	6.3	497702.
6	818.5	14.8	18.4	7.1	571680.
7	821.1	14.8	18.8	7.1	579293.
8	616.6	16.1	24.2	8.3	509275.
9	482.0	12.0	33.4	7.9	376964.
10	692.9	25.0	26.7	10.8	738999.
11	649.5	25.6	29.3	11.4	733446.
21	475.5	22.1	16.8	7.9	338390.
23	411.2	22.1	22.0	9.2	341767.
24	475.5	22.1	16.8	7.9	338390.
25	437.3	22.0	16.4	7.8	307122.
26	461.0	22.2	14.4	7.2	298817.

Note : Heat transfer coefficients are in BTU/hr-ft²-°F

Heat transfer is in BTU/hr

LMTD = Log Mean Temperature Difference

TABLE A.VIII. (CONT.)

HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
27	350.9	22.2	17.4	8.1	255502.
28	361.7	22.2	16.4	7.8	254406.
29	367.9	22.2	16.4	7.8	258792.
37	551.3	22.8	8.9	5.2	258990.
45	459.6	23.0	8.3	4.9	204178.
48	502.0	24.2	8.2	4.9	223619.
49	536.1	24.3	7.3	4.5	218153.
50	604.6	24.3	6.4	4.1	222015.
52	254.6	24.8	8.3	5.0	115002.
53	231.9	25.3	10.3	5.9	123729.
54	709.0	25.0	6.8	4.3	273081.
56	415.9	25.6	8.8	5.3	197826.
58	492.4	25.6	10.4	6.0	266532.
64	747.5	25.6	6.8	4.3	289056.
65	503.3	22.9	11.6	6.3	284983.
123	405.5	20.5	17.7	7.9	289912.
124	401.3	20.5	18.0	8.0	288921.
126	359.0	20.5	19.1	8.3	267811.
127	359.6	20.5	19.2	8.3	268801.
128	362.0	20.5	19.0	8.2	268801.
129	366.2	20.5	18.5	8.1	267811.

TABLE A.VIII. (CONT.)

HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
131	329.3	20.4	18.4	8.1	240035.
132	333.6	20.4	18.4	8.1	242998.
134	336.6	20.4	17.5	7.9	238553.
135	337.9	20.4	17.0	7.7	235096.
136	345.4	20.4	17.0	7.8	241022.
137	347.4	20.4	16.8	7.7	240528.
139	355.1	20.4	15.9	7.4	238059.
140	354.9	20.4	15.7	7.4	236083.
141	325.7	20.4	16.8	7.7	225168.
142	333.1	20.4	16.2	7.5	225661.
143	339.2	20.4	15.4	7.3	222704.
144	326.4	20.4	16.4	7.6	222704.
145	329.1	20.4	16.2	7.5	223197.
146	328.1	20.4	16.1	7.5	221719.
147	327.6	20.4	15.5	7.3	215806.
148	331.5	20.4	15.1	7.2	214821.
149	330.2	20.4	15.3	7.3	216299.
150	526.9	20.5	15.2	7.3	344664.
151	556.3	20.5	16.5	7.6	381414.
152	550.5	20.5	16.5	7.6	377937.
153	538.9	20.5	16.3	7.6	367509.

TABLE A.VIII. (CONT.)

HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
154	545.5	20.5	16.0	7.5	367012.
155	540.9	20.5	16.2	7.5	367012.
156	549.7	20.5	15.2	7.3	359563.
157	550.6	20.5	15.2	7.2	359066.
158	539.0	20.5	15.5	7.3	355590.
161	449.4	24.2	25.1	10.3	415752.
162	479.7	24.2	23.2	9.9	425534.
163	495.3	24.2	22.3	9.7	430426.
164	328.2	24.2	34.8	12.0	355176.
165	329.9	24.2	33.5	11.8	350904.
166	407.0	24.2	28.6	11.0	402661.
167	413.8	24.2	27.5	10.8	401437.
168	410.0	24.2	27.7	10.8	398989.
169	417.7	24.2	27.1	10.7	402049.
170	429.0	24.2	25.8	10.4	402661.
171	419.6	24.2	27.0	10.7	403273.
172	394.8	24.2	25.6	10.4	369616.
173	409.1	24.2	25.4	10.3	380631.
174	414.8	24.2	24.6	10.2	380019.
175	413.2	24.2	24.4	10.1	376959.
176	412.0	24.2	24.6	10.2	376959.

TABLE A.VIII. (CONT.)

HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
177	419.6	24.2	24.1	10.1	380019.
178	413.6	24.2	24.8	10.2	380019.
179	366.5	24.2	22.3	9.6	318211.
180	365.2	24.2	22.1	9.6	315764.
181	364.6	24.2	22.3	9.7	316988.
182	377.4	24.2	20.7	9.3	314540.
183	374.2	24.2	21.1	9.4	315152.
184	373.2	24.2	21.5	9.5	317599.
185	399.2	24.2	24.4	10.1	363783.
186	399.7	24.2	24.2	10.1	362560.
187	432.3	24.3	20.3	9.2	356155.
188	436.0	24.3	20.2	9.1	358607.
189	436.0	24.3	20.2	9.1	358607.
190	453.3	24.3	19.3	8.9	362898.
191	452.2	24.3	19.4	8.9	362898.
192	575.4	24.3	17.1	8.3	428490.
193	568.6	24.3	17.4	8.4	428490.
194	573.9	24.3	17.1	8.3	427877.
195	558.6	24.3	16.8	8.2	410713.
196	561.7	24.3	17.0	8.2	416230.
197	560.5	24.3	17.2	8.3	418069.

TABLE A.VIII. (CONT.)

HEAT TRANSFER DATA

RUN NO.	LMTD F	INSIDE TUBE COEFFICIENT	BED SIDE COEFFICIENT	OVERALL COEFFICIENT	HEAT TRANSFER
198	559.6	24.3	17.4	8.3	420521.
199	567.8	24.3	17.1	8.3	422973.
200	569.7	24.3	17.0	8.2	422360.
201	561.5	24.3	17.2	8.3	419908.
202	495.9	24.3	20.7	9.3	413165.
203	492.4	24.3	20.9	9.3	413165.
204	457.3	24.2	17.7	8.4	347518.
205	460.9	24.2	17.7	8.4	349957.
206	455.0	24.2	19.0	8.8	359712.
207	460.4	24.2	18.9	8.8	363370.
208	462.2	24.2	18.5	8.7	360322.
209	467.2	24.2	18.1	8.5	359102.
210	420.0	24.2	23.6	10.0	376347.
211	433.6	24.2	22.9	9.8	381855.
212	374.3	24.2	28.2	10.9	367780.
213	393.7	24.2	26.1	10.5	372064.
214	355.7	24.2	30.9	11.4	364672.
215	381.4	24.2	27.4	10.7	368948.

TABLE A.IX.
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
1	2095688.	1679744.	-0.1985	25.6
2	1705894.	1447276.	-0.1516	26.7
3	2667311.	2319558.	-0.1304	28.9
4	2964812.	2581223.	-0.1294	36.5
5	3212569.	2773565.	-0.1367	35.0
6	3913010.	3505881.	-0.1040	40.6
7	4578404.	4096248.	-0.1053	39.0
8	2532596.	2254117.	-0.1100	43.6
9	2608495.	2505425.	-0.0395	43.6
10	3944273.	3450480.	-0.1252	43.8
11	3929225.	3474628.	-0.1157	46.6
21	3040428.	2528784.	-0.1683	45.4
23	3053381.	2627865.	-0.1394	48.5
24	3051429.	2578949.	-0.1548	45.7
25	3174070.	2623887.	-0.1733	46.8
26	2932061.	2372653.	-0.1908	49.3

Note: Deviation = (out - in) / in

Note: Thermal efficiency (%)

$$= \frac{\text{Heat of combustion of cold dry gas}}{\text{Total heat in}} \times 100\%$$

TABLE A.IX. (CONT.)
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
27	2254904.	1984748.	-0.1198	50.9
28	2225274.	1949555.	-0.1239	54.5
29	2307795.	1938863.	-0.1599	46.5
37	2316880.	1936075.	-0.1644	46.3
45	2133795.	2002453.	-0.0616	55.7
48	2200829.	2020491.	-0.0819	50.5
49	1896852.	1614044.	-0.1491	47.4
50	2114420.	1886377.	-0.1079	51.4
52	1861231.	1732255.	-0.0693	57.7
53	1968470.	1753670.	-0.1091	53.9
54	2400224.	1925351.	-0.1978	46.0
56	2200837.	2105069.	-0.0435	59.2
58	2277639.	1806183.	-0.2070	45.0
64	2850586.	2220316.	-0.2211	46.9
65	2516025.	2005170.	-0.2030	51.7
123	1402497.	1177669.	-0.1603	40.9
124	1401397.	1166140.	-0.1679	39.8
126	1346648.	1101353.	-0.1822	42.4
127	1346456.	1095755.	-0.1862	41.9
128	1316065.	1084807.	-0.1757	45.0
129	1315341.	1076640.	-0.1815	44.5

TABLE A.IX. (CONT.)
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
131	1370850.	1096846.	-0.1999	36.5
132	1374163.	1095072.	-0.2031	36.5
134	1369017.	1099895.	-0.1966	36.8
135	1323194.	1099818.	-0.1688	41.2
136	1328812.	1084071.	-0.1842	40.7
137	1328936.	1091435.	-0.1787	40.4
139	1277634.	995806.	-0.2206	41.3
140	1275833.	996599.	-0.2189	41.3
141	1367606.	1068003.	-0.2191	35.3
142	1367977.	1076183.	-0.2133	35.6
143	1365628.	1095459.	-0.1978	37.2
144	1309838.	1014132.	-0.2258	37.8
145	1305757.	1032456.	-0.2093	39.3
146	1304381.	1035328.	-0.2063	39.5
147	1260690.	1017216.	-0.1931	43.2
148	1259705.	1020996.	-0.1895	43.3
149	1260819.	1030066.	-0.1830	43.7
150	3303561.	2714093.	-0.1784	48.9
151	2945391.	2414296.	-0.1803	49.2
152	2928169.	2402204.	-0.1796	49.2
153	3102633.	2512272.	-0.1903	49.5

TABLE A.IX. (CONT.)
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
154	2892765.	2329738.	-0.1946	49.4
155	2845937.	2321106.	-0.1844	50.1
156	2877426.	2320247.	-0.1936	50.8
157	2923734.	2363836.	-0.1915	50.9
158	3013588.	2444334.	-0.1889	51.0
161	3112763.	2797239.	-0.1014	57.5
162	3122324.	2698993.	-0.1356	54.4
163	3126881.	2702065.	-0.1359	54.7
164	1589975.	1283060.	-0.1930	42.1
165	1586881.	1273817.	-0.1973	41.7
166	2314830.	2000535.	-0.1358	50.4
167	2314153.	1996863.	-0.1371	50.6
168	2312362.	2001869.	-0.1343	51.1
169	2279419.	1933163.	-0.1519	50.3
170	2279932.	1928291.	-0.1542	50.2
171	2280248.	1925490.	-0.1556	50.0
172	2297835.	1944994.	-0.1536	48.5
173	2308505.	1934823.	-0.1619	48.0
174	2307778.	1945327.	-0.1571	48.6
175	2304949.	1949080.	-0.1544	48.8
176	2296053.	1982243.	-0.1367	52.1

TABLE A.IX. (CONT.)
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
177	2298888.	2004733.	-0.1280	53.1
178	2298776.	2016401.	-0.1228	53.7
179	2280755.	1933529.	-0.1522	47.3
180	2278180.	1944844.	-0.1463	47.2
181	2278767.	1978042.	-0.1320	48.8
182	2222293.	1877035.	-0.1554	49.5
183	2222794.	1893543.	-0.1481	50.3
184	2225132.	1864097.	-0.1623	48.8
185	2188091.	1839971.	-0.1591	55.1
186	2187151.	1858195.	-0.1504	56.0
187	2243004.	1908458.	-0.1492	54.6
188	2245351.	1864598.	-0.1696	52.8
189	2245562.	1877163.	-0.1641	53.4
190	2244363.	1833433.	-0.1831	52.0
191	2244156.	1832426.	-0.1835	52.0
192	3166239.	2659334.	-0.1601	53.2
193	3166110.	2694117.	-0.1491	54.3
194	3165755.	2689384.	-0.1505	54.2
195	3067208.	2640825.	-0.1390	56.9
196	3072213.	2607954.	-0.1511	56.0
197	3073847.	2606753.	-0.1520	56.0

TABLE A.IX. (CONT.)
HEAT BALANCE (BTU/HR)

RUN	HEAT IN	HEAT OUT	DEVIATION	THERMAL EFF. (%)
198	3074555.	2600671.	-0.1541	55.9
199	3067714.	2585120.	-0.1573	55.5
200	3066802.	2590210.	-0.1554	55.7
201	3065248.	2616064.	-0.1465	56.5
202	2343380.	1998651.	-0.1471	52.2
203	2343734.	2008383.	-0.1431	52.7
204	2273613.	1875011.	-0.1753	48.4
205	2275933.	1864920.	-0.1806	48.0
206	2267793.	1874390.	-0.1735	50.0
207	2270554.	1874255.	-0.1745	49.9
208	2231969.	1794583.	-0.1960	48.8
209	2230447.	1810541.	-0.1883	49.5
210	2299583.	2005381.	-0.1279	54.1
211	2305205.	2019078.	-0.1241	54.7
212	2223189.	1922058.	-0.1355	55.7
213	2227472.	1916781.	-0.1395	55.5
214	2207011.	1856766.	-0.1587	54.4
215	2211018.	1863109.	-0.1574	54.6

TABLE A.X. SCALE-UP SPECIFICATIONS

	20-in	A	B
Output(M BTU/hr)	1.35	10	40
Feedrate(lb/hr)(10% mois.)	300	2222	8889
Dry feedrate(lb/hr)	270	2000	8000
Steam flowrate(lb/hr)	405	3000	12000
Superficial velocity(ft/s)	3	3	3
Gas prod.(SCF/hr)	3000	22222	88889
Gas per dry feed(SCF/lb)	11.11	11.11	11.11
HHV* (Btu/SCF)	450	450	450
MW (lb/lb-mole)	25	25	25
Steam/Wood (Dry) ratio	1.5	1.5	1.5
Tar prod.(lb/hr)	21.6	160	640
Char prod.(lb/hr)	16.2	120	480
Max ash prod.(lb/hr)	1.9	14	56

* : Gas composition (Medium BTU gas)

H ₂	CO	CH ₄	C ₂ +	C ₃ +	CO ₂
20.0%	40.0%	13.0%	4.5%	2.5%	20.0%

TABLE A.X. SCALE-UP SPECIFICATIONS(CONT.)

	20-in	A	B
Output(M BTU/hr)	1.35	10	40
Prop. const.(lambda)	1.0	7.41	29.63
Feed flux(lb/hr-ft ²)	139	140	140
Heat input(Btu/lb)(dry)	1000	1000	1000
Reactor diam.(ft)	1.79	4.70	9.40
Diam. at top*(ft)	1.79	6.65	13.29
Tube O.D.(in)	1.05	1.9	1.9
No. tubes required**	30 U-Tubes	76	300
Dist. plate area(ft ²)	2.16	15.88	63.49
Tube spacing(in)	2.0	5.0	5.25
Fluidizing ht.(ft)	6	20	20
Bed ht.(ft)	3	10	10
Reactor ht.(ft)	15	34.2	36.7

* : Reactor diameter of commercial scale is increased
by 1.414 to reduce superficial velocity to 2 ft/s.

** : Heat transfer coeff. assumed const. = 6 Btu/hr-ft²-°F

Log mean temp diff. = 450°F

Flue inlet temp. = 2200°F

Flue outlet temp. = 1750°F

Reactor temp. = 1300°F

TABLE A.X. SCALE-UP SPECIFICATIONS(CONT.)

	A	B
Output(M BTU/hr)	10	40
Feed bin capacity*(gal)	462	1847
Diameter(ft)	3.84	6.10
Height(ft)	4.95	7.86
Cyclone dimensions:		
Flow rate(ft ³ /s)	46.33	185.31
Inlet width,w(ft)	0.621	1.243
Inlet height,h(ft)	1.243	2.485
Body diam.,Db(ft)	2.219	4.438
Outlet diam.,Do(ft)	0.976	1.953
Length,L(ft)	3.884	7.768
Inlet velocity(ft/s)	60	60

* : Feedbins for both reactors hold a sufficient weight of wood for one hour continuous operation at full load.

TABLE A.X. SCALE-UP SPECIFICATIONS(CONT.)

	A	B
Output(M BTU/hr)	10	40
Low BTU Reactor Specifications:		
Ht.(ft)	22.5	25.0
Fluidizing ht.(ft)	10	10
Feedrate(lb/hr)(40% mois.)	3300	13200
Fluidization air*(lb/hr)	3000	12000
Air for combustion(lb/hr)	7590	30360
Mass flow flue gas(lb/hr)	13890	55560
Max. ash prod.(lb/hr)	13.9	55.4
Feed flux(lb/hr-ft ²)	180	180
Reactor diam.(ft)	4.831	9.663
Diam. at top**(ft)	6.83	13.67
Low BTU gas prod.(lb/hr, dry basis)	3960	15840
Feed bin capacity*** (gal)	673	2692
Diam.(ft)	4.36	6.92
Ht.(ft)	5.62	8.91

* : 25% of stoichiometric air required for wood combustion

** : Reactor diameter of commercial scale is increased
by 1.414 to reduce superficial velocity to 2 ft/s.

***: Feedbins for both reactors hold a sufficient weight of
wood for one hour continuous operation at full load.

TABLE A.X. SCALE-UP SPECIFICATIONS(CONT.)

	A	B
Output(M BTU/hr)	10	40
Steam Generator Specifications:		
Inlet flue gas flowrate(lb/hr)	13400	53600
Inlet temp.(F)	1650	1650
Inlet Dimensions(in)	57.0x36	71.3x72
Outlet dimensions(in)	59.6x36	39.5x72
Overall ht.(ft)	8.77	11.60
Outlet temp(F)	860	860
Boiler:		
L.M.T.D.(F)	990	990
Surface area(ft ²)	192.1	582.0
No. tubes	103	156
Tube spacing(in)	4.75	4.75
Tube O.D.(in)	2.375	2.375
Tube length(ft)	3	6

TABLE A.X. SCALE-UP SPECIFICATIONS(CONT.)

	A	B
Output(M BTU/hr)	10	40
Steam Generator Specifications:		
Superheater:		
L.M.T.D.(F)	910	910
Surface area(ft ²)	67.2	223.8
No. tubes	36	60
Tube spacing(in)	7.125	7.125
Tube O.D.(in)	2.375	2.375
Tube length(ft)	3	6
Economizer:		
L.M.T.D.(F)	748	748
Surface area(ft ²)	21.7	70.2
No. tubes	21	34
Tube spacing(in)	3.29	3.29
Tube O.D.(in)	1.315	1.315
Tube length(ft)	3	6

APPENDIX B
DATA REDUCTION PROGRAM

C
 C VARIABLE DICTIONARY
 C
 C JUN : RUN NUMBER
 C MO : MONTH
 C DAY : DAY
 C YR : YEAR
 C WF : WOOD FEED RATE (WET BASIS)
 C MOIST : MOISTURE CONTENT OF WOOD
 C WS : STEAM RATE
 C DPAN : PRESSURE DROP MEASURED IN PROPANE LINE
 C AIR : AIR RATE IF UTILIZED
 C TS : STEAM INLET TEMPERATURE
 C TB : BED TEMPERATURE
 C TO : OVERHEAD TEMPERATURE
 C T1 : FLUE GAS INLET TEMPERATURE
 C TL : FLUE GAS OUTLET TEMPERATURE
 C P : PRESSURE IN THE REACTOR
 C MTAR : POUND TAR PER POUND MOLE OF DRY GAS
 C MCHAR : POUND CHAR PER POUND OF DRY WOOD
 C H2 : MOLE FRACTION OF HYDROGEN IN PRODUCT GAS
 C CO2 : MOLE FRACTION OF CARBON DIOXIDE IN PRODUCT GAS
 C C2H4 : MOLE FRACTION OF ETHYLENE IN PRODUCT GAS
 C C2H6 : MOLE FRACTION OF ETHANE IN PRODUCT GAS
 C O2 : MOLE FRACTION OF OXYGEN IN PRODUCT GAS
 C N2 : MOLE FRACTION OF NITROGEN IN PRODUCT GAS
 C CH4 : MOLE FRACTION OF METHANE IN PRODUCT GAS
 C CO : MOLE FRACTION OF CARBON MONOXIDE IN PRODUCT GAS
 C C3H8 : MOLE FRACTION OF PROPANE IN PRODUCT GAS
 C FCO2 : MOLE FRACTION OF CARBON DIOXIDE IN FLUE GAS
 C FN2 : MOLE FRACTION OF NITROGEN IN FLUE GAS
 C FO2 : MOLE FRACTION OF OXYGEN IN FLUE GAS
 C FAP : PRESSURE DROP MEASURED IN COMBUSTION AIR LINE
 C WGR : POUND STEAM PER POUND MOLE OF DRY GAS
 C DGFP : PRESSURE DROP MEASURED IN PRODUCT GAS LINE
 C T8 : PRODUCT GAS TEMPERATURE NEAR ORIFICE
 C CBED : CARBON IN BED
 C VGP : SPECIFIC VOLUME OF STEAM AT INLET TEMPERATURE

IMPLICIT REAL*4(A-H,L-Z)

INTEGER*4 IR(198)

REAL*4 C1(198),C2(198),C4(198),C5(198),C6(198),C7(198),
 &C8(198),C9(198),C10(198),C11(198),C12(198),C13(198),
 &CA(198),CB(198),CC(198),CD(198),CE(198),CF(198),
 &D1(198),D2(198),D3(198),D4(198),D5(198),D6(198),
 &D7(198),D8(198),D9(198),D10(198),D11(198),D12(198),
 &D13(198),D14(198),D15(198),D16(198),D17(198),D18(198),
 &D19(198),D20(198),D21(198),D22(198),D23(198),
 &F1(198),F2(198),F3(198),F4(198),F5(198),F6(198),F7(198),F8(198),
 &H0(198),H1(198),H3(198),H4(198),H5(198),H6(198),H7(198),
 &H8(198),H9(198),H10(198),H11(198),H12(198),H13(198),H14(198),
 &H15(198),H16(198),
 &HT1(198),HT2(198),HT3(198),HT4(198),HT5(198),

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&M1(198),M2(198),M3(198),M4(198),M5(198),M6(198),M7(198),
&M8(198),M9(198),M10(198),M11(198),M12(198),M13(198),M14(198),
&M15(198),M16(198),M17(198),M18(198),M19(198),
&PL1(198),PL2(198),PL3(198),PL4(198),PL5(198),PL6(198),
&PL7(198),PL8(198),PL9(198),PL0(198)
J=0
DO 100 I=1,198
  READ(5,*,END=10)JUN,MO,DAY,YR,WF,MOIST,WS,DPAN,AIR,TS,TB,TO,T1,TL,
  &P,MTAR,MCHAR,H2,CO2,C2H4,C2H6,O2,N2,CH4,CO,C3H8,FCO2,FN2,F02,
  &FAP,WGR,DGFP,T8,CBED,VG
  WRITE(6,*)JUN,MO,DAY,YR,WF,MOIST,WS,DPAN,AIR,TS,TB,TO,T1,TL,
  &P,MTAR,MCHAR,H2,CO2,C2H4,C2H6,O2,N2,CH4,CO,C3H8,FCO2,FN2,F02,
  &FAP,WGR,DGFP,T8,CBED,VG
C
C WOOD FRACTION (DRY BASIS)
C
  C=.4819
  H=.0660
  O=.4363
  ASH=.0102
C
C WOOD FRACTION (WET BASIS)
C
  F=1.-MOIST
  MC=C*F
  MH=H*F
  MO=O*F
C
C FLOWRATE OF WOOD COMPOSITIONS
C
  WC=MC*WF
  WH=MH*WF
  WO=MO*WF
  WH2O=MOIST*WF
  DWF=WF*F
C
C FLOWRATE OF STEAM COMPOSITIONS AND MOISTURE CONTENT
C
  MOIS=WF-DWF
  HM=2*MOIS/18
  HO=16*MOIS/18
  SH=2*WS/18
  SO=16*WS/18
  SIN=MOIS+WS
C
C CHAR RATE
C
  CHAR=MCHAR*DWF
450  MWG=CO2*44+CO*28+H2*2+C2H4*28+C2H6*30+C3H8*42+O2*32+
  &N2*28+CH4*16

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C
C   DRY GAS FLOW AND STEAM OUTPUT
C
      IF(JUN.GT.20)GO TO 500
      DGF=27.*SQRT(MWG*520.*DGFP/(460.+T8))
      MDGF=DGF/MWG
      GH20=WGR*MDGF
      GO TO 550
500   WG=WGR/18.
      MWG=WG/(1.+WG)*18.+1./(1.+WG)*MWG
      WGF=27.*SQRT(MWG*520.*DGFP/(460.+T8))
      MDGF=WGF/MWG/(1.+WG)
      DGF=MWG*MDGF
      GH20=WGR*MDGF
C
C   TAR RATE
C
550   TAR=MTAR*MDGF
C
C   CARBON BALANCE
C
      MCPG=CO2+CO+2*C2H4+2*C2H6+3*C3H8+CH4
      CG=12*MCPG*DGF/MWG
      CTAR=.6915*TAR
      CCHAR=.874*CHAR
      COUT=CG+CTAR+CCHAR+CBED
      DELC=(COUT-WC)/WC
C
C   H2 BALANCE
C
      MHPG=H2+2*C2H4+3*C2H6+3*C3H8+2*CH4
      H2G=2*MHPG*DGF/MWG
      H2TAR=.0668*TAR
      H2CHA=.0564*CHAR
      HGH20=2*GH20/18
      H2IN=WH+SH+HM
      H2OUT=H2G+H2TAR+HGH20+H2CHA
      DELH2=(H2OUT-H2IN)/H2IN
C
C   O2 BALANCE
C
      MOPG=CO2+.5*CO
      O2G=32*MOPG*DGF/MWG
      O2TAR=.2232*TAR
      O2CHA=.0026*CHAR
      OGH20=16*GH20/18
      O2IN=W0+S0+H0
      O2OUT=O2G+O2TAR+OGH20
      DELO2=(O2OUT-O2IN)/O2IN
      CHOG=CG+H2G+O2G
      CFRA=CG/CHOG
      HFRA=H2G/CHOG
      OFRA=O2G/CHOG

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MSIN=SIN+DWF
MSOUT=CHAR+TAR+DGF+GH2O
DELS=(MSOUT-MSIN)/MSIN
C
C HEAT BALANCE
C
C A. SENSIBLE HEAT OF PRODUCT
C
1401 T2=TO+460.
      TI=540.
      DELT=TO-77
      CALL HEAT(T2,HCO2,HCO,HCH4,HH2O,HN2,HH2,HC2H4,HC2H6,HC3H8,H02,TI)
      HPDG=C02*HCO2+C0*HCO+CH4*HCH4+N2*HN2+H2*HH2+C2H4*HC2H4+C2H6*HC2H6
      & +C3H8*HC3H8+O2*H02
C
C BTU/HR OF DRY GAS
C
      BPHG=HPDG*DGF/MWG
      BH2O=HH2O*GH2O/18+(GH2O*1048)
      BTAR=.5*CTAR*DELT+H02*O2TAR/32+HH2*H2TAR/2
      BCHAR=.5*(CHAR+CBED)*DELT
C
C TOTAL SENSIBLE HEAT OF PRODUCT (BTU/HR)
C
      SHP=BPHG+BH2O+BTAR+BCHAR
C
C B. HEAT OF COMBUSTION OF PRODUCT
C
      HVPG=C0*121750+H2*122970+CH4*383040+C2H4*607010+C2H6*671090
      & +C3H8*885580
      HHV=HVPG/359
      HVDG=HVPG*DGF/MWG
      HVTAR=10066*TAR
      HVCHAR=169290*CCHAR/12+122970*H2CHA/2
      HVP=HVDG+HVTAR+HVCHAR
      EOUT=SHP+HVP
C
C C. HEAT OF COMBUSTION OF DRY WOOD
C
      HVW=8319*DWF
C
C D. SENSIBLE HEAT OF STEAM
C
      SST=WS*(.46*(TS-77)+1048.)
C
C E. HEAT TRANSFER FROM TUBES
C
      FAIR=700.*SQRT(FAP)
      IF(FN2.EQ.0.)GO TO 2000
      FMG=44.*FCO2+28.*FN2+32.*FO2
      FN2I=.77*FAIR
      FDG=FMG*FN2I/(28.*FN2)
      FO2O=32*FO2*FN2I/(28*FN2)

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FCO20=44*FCO2*FN2I/(28*FN2)
FC3H8=FCO20/3.
FWG=FC3H8*4.
FGOUT1=FWG+FDG1
2000 PANE=51.7*SQRT(DPAN)
FGOUT=PANE+FAIR
HFIN=FGOUT*.317*(T1-77.)
HFOUT=FGOUT*.317*(TL-77.)
HFINP=PANE*20127.
IF(JUN.GT.20)GO TO 2010
HEX=HFINP-HFOUT
GO TO 2020
2010 HEX=HFIN-HFOUT
2020 IF(FN2.EQ.0.)GO TO 2100
ASTOI=15.54*FC3H8
EXAIR=(FAIR-ASTOI)*100/ASTOI
C WRITE(6,*)PANE,FC3H8
C WRITE(6,*)FGOUT,FGOUT1
C HLOSS=.10*(HVW+SST+HEXI)
C HEX=HEXI+HLOSS
C
C F. FIND THE OVERALL HEAT COEFF. U; Q=UA(LMTD)
C
2100 LMTD=(T1-TL)/ALOG((T1-TB)/(TL-TB))
IF(JUN.GT.20)GO TO 2110
OVERU=HEX/(LMTD*99.)
GO TO 2120
2110 OVERU=HEX/(LMTD*90.)
C
C FIND HEAT COEFF. INSIDE THE TUBES
C
2120 VELOC=FGOUT/8.392
RENOL=VELOC*50.795
HCI=.0135*(RENOL**0.8)
C
C FIND HEAT COEFF. OUTSIDE THE TUBES
C
OVERUI=1./OVERU
HCII=1./HCI
HCOE=1.05/ (.824*(OVERUI-HCII-.0052))
C
C TOTAL HEAT IN
C
EIN1=HVW+SST+HFINP
EIN=HVW+SST+HEX
C WRITE(6,*)EIN,EIN1
DELE=(EOUT-EIN)/EIN
THER=HVDG*100./EIN
C
C H2/CO AND STEAM/DRY WOOD
C
HCRAT=H2/CO
SWRAT=WS/DWF

```

```

C
C   LB C IN GAS/ LB C IN WOOD & COLD GAS EFFICIENTCY
C
      CRAT=CG/WC
      CGEF=HVDG/HVW
C
C   RESIDENCE TIME
C
      R=1545/MWG
      V=DGF*R*(TB+460.)/(P*144.*3600.)
      VS=VG*GH2O/3600.
      RT=28./(V+VS)
C
C   STORE FOR MAKING TABLE
C
      IR(I)=JUN
      C11(I)=CFRA
      C12(I)=HFRA
      C13(I)=OFRA
      C1(I)=H2
      C2(I)=CO2
      C10(I)=C2H4
      C4(I)=C2H6
      C5(I)=O2
      C6(I)=N2
      C7(I)=CH4
      C8(I)=CO
      C9(I)=C3H8
      CA(I)=DGF
      CB(I)=GH2O
      CC(I)=TAR
      CD(I)=CHAR
      CE(I)=HEX
      CF(I)=HHV
      D1(I)=WF
      D2(I)=WS
      D3(I)=PANE
      D4(I)=T1
      D5(I)=TL
      D6(I)=TB
      D7(I)=TO
      D8(I)=TS
      M1(I)=WC
      M2(I)=CTAR
      M3(I)=CCHAR
      M4(I)=CG
      M5(I)=DELC
      M6(I)=WH
      M7(I)=SH
      M8(I)=H2TAR
      M9(I)=H2G
      M10(I)=DEIH2

```

```

M11(I)=WO
M12(I)=SO
M13(I)=O2TAR
M14(I)=O2G
M15(I)=DELO2
M16(I)=H2CHA
M17(I)=O2CHA
M18(I)=HGH2O
M19(I)=OGH2O
H1(I)=SST
H0(I)=BTAR
H3(I)=BCHAR
H4(I)=BPHG
H5(I)=BH2O
H6(I)=BPHG+BH2O
H7(I)=HVV
H8(I)=HVTAR
H9(I)=HVCHAR
H10(I)=HVDG
H11(I)=BH2O
H12(I)=E IN
H13(I)=EOUT
H14(I)=DELE
H15(I)=THER
H16(I)=HPDG
F1(I)=FCO2
F2(I)=FN2
F3(I)=FO2
F4(I)=FAIR
F5(I)=FC3H8
F6(I)=EXAIR
F7(I)=FGOUT
F8(I)=WGR
HT1(I)=LMTD
HT2(I)=HCI
HT3(I)=HCOE
HT4(I)=OVERU
HT5(I)=FAIR
PL1(I)=HCRAT
PL2(I)=SWRAT
PL3(I)=CRAT
PL4(I)=CGEF
PL5(I)=MDGF
PL6(I)=DWF
PL7(I)=MWG
PL8(I)=RT
PL9(I)=S IN
PL0(I)=DELMs
J=J+1
100 CONTINUE

```

```

10  WRITE(6,169)
169  FORMAT(1H1,'RUN',3X,'SWR',3X,'H2/CO',3X,'C-CONV',
& 3X,'GS-EFF',2X,'MDGF',4X,'DWF',4X,'SIN',5X,'RT',4X,'BED-F',
& 2X,'BTU/MOLE',2X,'MWG',3X,'WGR',3X,'MS-DV'/)
DO 171 IQ=1,J
WRITE(6,170)IR(IQ),PL2(IQ),PL1(IQ),PL3(IQ),PL4(IQ),PL5(IQ),
& PL6(IQ),PL9(IQ),PL8(IQ),D6(IQ),H16(IQ),PL7(IQ),F8(IQ),PL0(IQ)
170  FORMAT(1X,I3,2X,F5.3,2X,F5.3,3X,F6.4,3X,F6.4,F6.2,3X,F5.1,
& 2X,F5.1,3X,F4.2,3X,F5.0,3X,F6.0,F7.1,F6.1,F7.3/)
171  CONTINUE
WRITE(6,11)
11  FORMAT(1H1,19X,'GAS COMPOSITIONS'/)
WRITE(6,12)
12  FORMAT(1X,'RUN',3X,'H2',3X,'CO2',3X,'C2H4',2X,'C2H6',2X,
& 'CH4',4X,'CO',4X,'C3+',3X,'O2',4X,'N2'/)
DO 101 II=1,J
WRITE(6,13)IR(II),C1(II),C2(II),C10(II),C4(II),C7(II),C8(II),
& C9(II),C5(II),C6(II)
13  FORMAT(1X,I3,1X,9(F5.4,1X)/)
101  CONTINUE
WRITE(6,14)
14  FORMAT(1H1,20X,'OPERATING DATA'/)
WRITE(6,15)
15  FORMAT(1X,'RUN',3X,'FEED',2X,'STEAM',2X,'PROPANE',2X,'BED TEMP.',
& 3X,'STEAM',3X,'BURNER AIR')
WRITE(6,16)
16  FORMAT(1X,'NO.',2X,'LB/HR',2X,'LB/HR',3X,'LB/HR',7X,'F',
& 9X,'F',7X,'LB/HR'/)
DO 102 IJ=1,J
WRITE(6,17)IR(IJ),D1(IJ),D2(IJ),D3(IJ),D6(IJ),D8(IJ),F4(IJ)
17  FORMAT(1X,I3,3X,F4.0,2X,F5.0,2X,F6.1,5X,F5.0,5X,F5.0,1X,F9.0/)
102  CONTINUE
WRITE(6,18)
18  FORMAT(1H1,20X,'PRODUCTION DATA'/)
WRITE(6,19)
19  FORMAT(1X,'RUN',3X,'DRY GAS',4X,'STEAM OUT',4X,'TAR',5X,'CHAR',
& 3X,'HEAT VALUE')
WRITE(6,20)
20  FORMAT(1X,'NO.',4X,'LB/HR',7X,'LB/HR',5X,'LB/HR',3X,'LB/HR',
& 4X,'BTU/SCF'/)
DO 103 IA=1,J
WRITE(6,21)IR(IA),CA(IA),CB(IA),CC(IA),CD(IA),CF(IA)
21  FORMAT(1X,I3,3X,F6.1,5X,F7.1,5X,F5.1,3X,F5.1,F10.0/)
103  CONTINUE
WRITE(6,22)
22  FORMAT(1H1,14X,'CARBON BALANCE (LB/HR)'/)
WRITE(6,23)
23  FORMAT(1X,'RUN',5X,'FEED',4X,'TAR',4X,'CHAR',4X,
& 'DRY GAS',4X,'ERROR'/)

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DO 104 IB=1,J
WRITE(6,24)IR(IB),M1(IB),M2(IB),M3(IB),M4(IB),M5(IB)
24 FORMAT(1X,I3,4X,F5.1,2X,F5.1,3X,F5.1,3X,F7.1,4X,F7.4/)
104 CONTINUE
WRITE(6,25)
25 FORMAT(1H1,16X,'HYDROGEN BALANCE (LB/HR)'/)
WRITE(6,26)
26 FORMAT(1X,'RUN',2X,'FEED',2X,'STEAM',3X,'TAR',3X,'CHAR',2X,
& 'DRY GAS',3X,'STEAM OUT',2X,'ERROR'/)
DO 105 IC=1,J
WRITE(6,27)IR(IC),M6(IC),M7(IC),M8(IC),M16(IC),M9(IC),M18(IC),
& M10(IC)
27 FORMAT(1X,I3,1X,F5.1,2X,F5.1,1X,F5.1,2X,F5.2,1X,F6.1,6X,F5.1,
& 4X,F7.4/)
105 CONTINUE
WRITE(6,28)
28 FORMAT(1H1,20X,'OXYGEN BALANCE (LB/HR)'/)
WRITE(6,29)
29 FORMAT(1X,'RUN',2X,'FEED',2X,'STEAM',3X,'TAR',3X,'CHAR',2X,
& 'DRY GAS',2X,'STEAM OUT',3X,'ERROR'/)
DO 106 ID=1,J
WRITE(6,30)IR(ID),M11(ID),M12(ID),M13(ID),M17(ID),M14(ID),M19(ID),
& M15(ID)
30 FORMAT(1X,I3,1X,F5.1,2X,F5.1,1X,F5.1,2X,F5.2,3X,F5.1,5X,F5.1,
& 4X,F7.4/)
106 CONTINUE
WRITE(6,31)
31 FORMAT(1H1,15X,'SENSIBLE HEAT (BTU/HR)'/)
WRITE(6,32)
32 FORMAT(1X,'RUN',4X,'STEAM',5X,'TAR',4X,'CHAR',3X,
& 'DRY GAS',3X,'STEAM OUT'/)
DO 107 IE=1,J
WRITE(6,33)IR(IE),H1(IE),H0(IE),H3(IE),H4(IE),H5(IE)
33 FORMAT(1X,I3,F10.0,2X,F6.0,2X,F6.0,2X,F7.0,4X,F7.0/)
107 CONTINUE
WRITE(6,34)
34 FORMAT(1H1,12X,'HEAT OF COMBUSTION (BTU/HR)'/)
WRITE(6,35)
35 FORMAT(1X,'RUN',4X,'FEED',8X,'TAR',7X,'CHAR',6X,'DRY GAS'/)
DO 108 IG=1,J
WRITE(6,36)IR(IG),H7(IG),H8(IG),H9(IG),H10(IG)
36 FORMAT(1X,I3,2X,F8.0,3X,F8.0,3X,F8.0,3X,F9.0/)
108 CONTINUE
WRITE(6,37)
37 FORMAT(1H1,15X,'HEAT BALANCE (BTU/HR)'/)
WRITE(6,38)
38 FORMAT(1X,'RUN',5X,'HEAT IN',7X,'HEAT OUT',5X,'ERROR',3X,
& 'THERMAL EFF.(%)'/)

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DO 109 IH=1,J
WRITE(6,39)IR(IH),H12(IH),H13(IH),H14(IH),H15(IH)
39 FORMAT(1X,I3,5X,F8.0,6X,F8.0,3X,F8.4,8X,F4.1/)
109 CONTINUE
WRITE(6,40)
40 FORMAT(1H1,20X,'HEAT TRANSFER DATA'/)
WRITE(6,41)
41 FORMAT(1X,'RUN',2X,'LMTD',2X,'INSIDE TUBE',4X,'BED SIDE',6X,
& 'OVERALL',7X,'HEAT'/,
& 1X,'NO.',4X,'F',3X,'COEFFICIENT',3X,'COEFFICIENT',3X,
& 'COEFFICIENT',2X,'TRANSFER'/)
DO 110 IP=1,J
WRITE(6,42)IR(IP),HT1(IP),HT2(IP),HT3(IP),HT4(IP),CE(IP)
42 FORMAT(1X,I3,1X,F6.1,1X,F8.1,4X,F9.1,5X,F9.1,5X,F9.0/)
110 CONTINUE
9999 STOP
END

C
SUBROUTINE HEAT(T2,HCO2,HCO,HCH4,HH2O,HN2,HH2,HC2H4,HC2H6,HC3H8,
& HO2,TI)
HCO2=16.2*(T2-TI)-6.53E+3*ALOG(T2/TI)-1.41E+6*(1./T2-1./TI)
HCO=9.46*(T2-TI)-3.29E+3*ALOG(T2/TI)-1.07E+6*(1./T2-1./TI)
HCH4=4.52*(T2-TI)+(.00737/2.)*(T2*T2-TI*TI)
HH2O=19.86*(T2-TI)-(2.*597.)*(SQRT(T2)-SQRT(TI))+7500.*ALOG(T2/TI)
HN2=9.47*(T2-TI)-3.47E+3*ALOG(T2/TI)-1.16E+6*(1./T2-1./TI)
HH2=5.76*(T2-TI)+(.578/2000)*(T2*T2-TI*TI)+40*(SQRT(T2)-SQRT(TI))
HO2=11.515*(T2-TI)-344.*(SQRT(T2)-SQRT(TI))+1530.*ALOG(T2/TI)
HC2H4=4.23*(T2-TI)+.00589*(T2*T2-TI*TI)
HC2H6=4.01*(T2-TI)+.00818*(T2*T2-TI*TI)
HC3H8=2.258*(T2-TI)+.016*(T2*T2-TI*TI)-1.81E-6*(T2*T2*T2-TI*TI
& *TI)
RETURN
END

```

APPENDIX C
RECYCLE AND ALTERNATE FEED

Process Description and Operation of Recycle System

The major component of the recycle system is the venturi injector. In the injector the recycled gas from the product gas line flows due to the lower pressure created by the throttled fluidization steam. The injector was installed downstream of the pneumatic control valve through which the desired steam flow was controlled before entering the electric preheater. The injector recycle inlet was connected to a surplus turbine meter and a check valve using 2-inch piping. The piping connected the inlet of the check valve to the main gas line below the cyclone. After a test run, it was discovered that no recycle flow could be detected. The suspected problem was related to flow restrictions in the recycle line. Several attempts to eliminate these possible restrictions were made including moving the inlet of the recycle line to the christmas tree (5, 9). This move was tested and proved to be insufficient to provide the desired performance.

Consequently the injector was moved behind the electric preheater to raise the steam temperature while leaving the recycle connection at the christmas tree. This move greatly reduced the length of recycle line improving the recycle efficiency. The increased steam temperature also decreased the steam density thus increasing the steam volume flow rate, thereby lowering the throttled pressure. After another unsuccessful test, a new injector (as shown in

Figure C.1) with a smaller throat was designed and fabricated. A test run showed as much as 10-in W.C. pressure difference across the recycle line. This pressure difference was thought to be enough to drive the recycle flow. However, no reading was obtained from the turbine meter. Later, it was found that the char/tar mixture in the gas had stopped the turbine while almost totally blocking the line. The turbine meter was removed and pressure drop measurements versus air flow across a section of the pipe were made to produce a calibrated flow measuring section. The system was run using this recycle system. The gasifier performance was significantly different but the planned flow measurement did not work.

The flow measurement of the recycle gas was not only difficult to accomplish but also difficult to maintain. This results from the nature of the recycle gas. The product gas stream was unclean. The recycle gas was full of char particles and vapor phase tar. On contacting the cooler surface of recycle pipe, the tar starts to condense and the sticky tar traps additional char particles. This produced a layer of tar/char mixture growing thicker with time. Both the decreasing cross-sectional area and the changing surface condition influenced the differential pressure measurement resulting in a deviation from the original calibration of the flow.

Another possible solution for the flow measurement considered was to inject a known rate of neon into the

FIGURE C.1 STEAM INJECTOR DETAIL DIMENSIONS

recycle line. As shown in Figure C.2, the concentrations of Ne in the samples obtained before and after the Ne injection port, along with the known flow rate of Ne stream are sufficient to calculate the volume flow rate of the recycle stream. The procedure is shown as following:

Ne balance:

$$F_{Ne} \times Y_{Ne,0} + F_{Rec} \times Y_{Ne,1} = (F_{Rec} + F_{Ne}) \times Y_{Ne,2}$$

Assume $F_{Ne} \ll F_{Rec}$, then $F_{Rec} + F_{Ne} = F_{Rec}$

$$F_{Ne} \times Y_{Ne,0} + F_{Rec} \times Y_{Ne,1} = F_{Rec} \times Y_{Ne,2}$$

$$F_{Ne} \times Y_{Ne,0} = F_{Rec} \times (Y_{Ne,2} - Y_{Ne,1})$$

$$\text{so, } F_{Rec} = \frac{Y_{Ne,0}}{Y_{Ne,2} - Y_{Ne,1}} \times F_{Ne}$$

There are drawbacks to this method. It requires two additional gas samples for the recycle flow determination, and it takes about 35 minutes for each gas sample to be analyzed by the gas chromatograph. The other problem is the accuracy of the Carle chromatograph and the necessity for big neon rates. Besides the extra time and effort, the gas sampling ports and the recycle line would still be easily plugged and need periodical cleaning. Other alternative procedures are still under consideration.

The next step was to purchase a commercial venturi/nozzel injector system from the Burner Design Corporation in St. Louis. It was installed in parallel to the fabricated injector. The piping arrangement and the reactor minimum steam requirement necessitated the fabricated injector to be used during operation. The piping arrangement allowed the

FIGURE C.2 DIAGRAM FOR RECYCLE FLOW MEASUREMENT

recycle flow to have the following options: both injectors, the fabricated injector only, the commercial injector only, or none (recycle flow been totally shut-off).

Results and Discussion

The results of the data for the recycle runs are given in Table C.I. In the table, runs no. 231, 236 and 255 were operated without recycle, runs no. 231-260 (except 3 runs mentioned above) were operated with the fabricated injector, runs no. 264-267 were operated with the commercial injector, and run no. 268 was operated using both injectors.

The major effect of recycling is shown in the H_2/CO ratio of the product gas seen in Figure C.3. The H_2/CO ratio varied between 0.3 to 0.6 for the top feed runs without recycle with an increasing H_2/CO trend with temperature. For the top feed runs with recycle (using the fabricated injector), the H_2/CO ratio varied between 0.4 to 0.9 with a greater slope with temperature. The comparisons of the results of the top feed runs with and without recycle (using the fabricated injector) show that the H_2/CO ratio of the product gas is larger with recycle at any temperature (above 1200°F). The difference between the H_2/CO ratio at any temperature increases with increasing temperature. It is believed that the recycled gas upon entering the donut section is shifted almost entirely toward H_2 . This donut section is the highest temperature point in the reactor and

TABLE C.I. RECYCLE TEST RESULTS

RUN No.	MASS IN		OPERATING DATA				MASS OUT				GAS ANALYSIS									
	FEED lb/hr	STEAM lb/hr	T_B F	T_S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HIV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO (Mole Fraction)	C ₃ +	O ₂	N ₂
231	200	273	1328	975	86.5	1642	179.2	276.4	13.1	3.7	476	.1809	.1492	.0420	.0084	.1160	.4401	.0219	.0053	.0361
232	200	243	1348	987	85.0	1642	179.6	247.4	13.5	3.7	477	.2146	.1610	.0406	.0081	.1158	.4113	.0229	.0025	.0233
233	200	243	1348	990	85.0	1642	177.6	247.0	13.5	3.7	483	.2256	.1647	.0405	.0081	.1170	.4014	.0246	.0014	.0167
236	200	265	1399	1047	98.1	1852	179.4	268.7	12.3	3.7	492	.2200	.1467	.0448	.0083	.1280	.4245	.0181	.0018	.0077
237	200	265	1385	1050	85.0	1852	175.0	266.1	12.2	3.7	485	.2341	.1464	.0428	.0083	.1253	.4146	.0172	.0020	.0092
238	200	265	1365	1038	85.0	1852	172.8	263.6	12.8	3.7	486	.2446	.1519	.0411	.0081	.1206	.4009	.0213	.0020	.0094
241	200	215	1258	1041	98.1	1852	174.2	220.2	14.7	2.5	493	.2084	.1369	.0388	.0093	.1160	.4584	.0240	.0004	.0078
242	200	215	1320	1048	92.5	1852	171.9	218.5	13.6	2.5	497	.2439	.1345	.0385	.0089	.1154	.4229	.0265	.0004	.0090
246	200	238	1429	1051	76.7	1565	183.3	235.9	13.5	3.7	459	.2809	.1492	.0317	.0078	.1263	.3870	.0117	.0006	.0048
247	200	251	1455	1057	80.1	1565	183.5	241.9	13.8	3.7	449	.3042	.1530	.0306	.0072	.1220	.3681	.0100	.0006	.0045
249	200	212	1409	1039	78.4	1642	177.9	204.6	13.1	3.0	450	.2873	.1575	.0306	.0074	.1236	.3745	.0110	.0009	.0072
250	200	250	1428	1042	81.7	1642	179.2	242.2	13.4	3.0	448	.3030	.1628	.0296	.0073	.1215	.3549	.0122	.0009	.0077
251	200	250	1421	1044	80.9	1642	177.7	240.4	13.3	3.0	453	.3035	.1609	.0294	.0073	.1220	.3546	.0143	.0018	.0062
252	200	228	1310	1008	78.9	1715	178.1	227.2	13.6	3.7	475	.2155	.1492	.0302	.0088	.1302	.4349	.0192	.0026	.0095
253	200	228	1315	1016	78.4	1715	176.6	224.5	13.9	3.7	471	.2486	.1537	.0291	.0085	.1285	.4069	.0186	.0015	.0074

TABLE C.I. RECYCLE TEST RESULTS (CONT.)

RUN No.	MASS IN		OPERATING DATA				MASS OUT				GAS ANALYSIS									
	FEED lb/hr	STEAM lb/hr	T_B F	T_S F	FUEL lb/hr	AIR lb/hr	GAS lb/hr	STEAM lb/hr	TAR lb/hr	CHAR lb/hr	HHV Btu/SCF	H ₂	CO ₂	C ₂ H ₄	C ₂ H ₆	CH ₄	CO	C ₃ +	O ₂	N ₂
255	200	228	1403	998	80.1	1642	185.4	229.3	13.1	2.2	458	.2538	.1547	.0304	.0081	.1240	.3964	.0155	.0060	.0111
256	200	228	1407	993	81.7	1642	180.0	226.4	12.9	4.8	455	.2691	.1584	.0296	.0081	.1239	.3889	.0135	.0011	.0074
257	200	228	1402	982	80.1	1642	179.7	226.6	13.0	4.8	442	.2768	.1622	.0284	.0077	.1183	.3724	.0130	.0063	.0148
258	200	228	1417	1010	80.1	1642	179.2	227.7	13.0	6.0	443	.2848	.1678	.0288	.0077	.1199	.3671	.0121	.0040	.0078
260	200	228	1425	1034	81.7	1642	176.4	224.3	12.8	6.0	440	.2829	.1644	.0285	.0076	.1206	.3671	.0112	.0065	.0112
264	200	206	1394	1039	80.1	1642	189.1	196.7	14.3	3.2	423	.3285	.1828	.0273	.0065	.1148	.3187	.0088	.0030	.0096
265	200	206	1402	1052	80.1	1642	194.5	191.3	15.3	3.2	404	.3721	.1952	.0223	.0063	.1045	.2810	.0081	.0028	.0077
266	200	206	1417	1061	78.4	1642	191.8	191.0	15.3	3.2	400	.3811	.1918	.0219	.0063	.1003	.2822	.0073	.0022	.0069
267	200	201	1420	1071	78.4	1642	190.4	183.4	15.4	3.7	389	.4016	.2008	.0189	.0060	.0964	.2592	.0069	.0027	.0074
268	200	201	1422	1068	78.4	1642	192.7	182.5	15.4	3.0	396	.3854	.1982	.0198	.0063	.1011	.2727	.0076	.0025	.0064
269	100	208	1506	1061	81.7	1642	--	--	--	--	293	.5884	.2301	.0060	.0011	.0332	.1306	.0000	.0024	.0082
270	100	208	1507	1051	81.7	1642	--	--	--	--	276	.6155	.2463	.0033	.0005	.0228	.1008	.0000	.0029	.0079
271	100	208	1458	1022	81.7	1642	--	--	--	--	304	.5429	.2428	.0092	.0016	.0465	.1462	.0000	.0025	.0082

Note: Runs no. 269-271 were rice hulls feed with limited data collected.

H₂ / CO RATIO VS. TEMP.

FIGURE C.3 H₂/CO RATIO VS. TEMPERATURE FOR TOP FEED RUNS

the partial pressure of the steam is very high in this section. The combination of the high temperature and the high steam partial pressure shifts the CO and H₂O to H₂ and CO₂. Secondary reactions also add to the variation including breaking down the heavier recycled tar, reaction of char with steam, and etc. The water gas shift reaction not only changes the compositions of the product gas but also releases some extra heat for the reactor. This was reflected through the temperature rise while the burner was kept at the same setting.

The data obtained from using the commercial injector showed an even higher H₂/CO ratio (between 1.0 to 1.5 as seen in Figure C.3) but no temperature dependence could be stated due to the limited temperature range. It is clear that the commercial injector drew more recycle gas than the other injector. The higher H₂/CO ratio shift was expected due to this increased recycle rates. However, no significant difference can be found between run 268 and runs 265-267 in either gas quality or gasifier performance. When recycle flow was directed to both injectors, the total flow was not significantly different from the flow using the commercial injector. Since the difference in the H₂/CO ratio is primarily related to the amount of recycle gas, the results between runs using the commercial injector and using both injectors are essentially the same suggesting a recycle flow limit.

The problem of measuring the recycle stream remained; the final decision was to simply condense the steam from the mixture stream of steam and product gas. A part of the mixed stream was carefully and slowly drawn through a condensor collecting the condensible phase while permitting the non-condensable gas (recycled saturated product gas) to flow to a wet test meter for volume measurement. Knowing the molecular weight, the volume of the gas could be converted to mass. From Table C.I, the corresponding steam in the recycled product gas stream could be determined and the ratio of gas/steam found. The condensed steam from the product gas was subtracted from the total amount of condensed water collected in the scrubber. The remaining steam was the fluidization steam. The combined mass of the gas and steam from the recycle stream and the mass of the condensed fluidization steam were used to calculate the mass ratio of the recycle gas stream to the fluidization steam. Since the fluidization steam flow was measured, the recycle gas mass flow rate was obtained using the measured flow rate. This mass flow rate was converted to a volume flow rate with the temperature and pressure measurement at the christmas tree and the molecular weight of the product gas.

During run 267, the procedure discussed above was carried out and 137 ml of condensed water corresponding with 1.5 liters saturated recycle gas were collected. The mass ratio was calculated as follows:

water: 137 ml x 1 g/ml = 137 g

$$\text{gas: } 1.5 \text{ l} \times \frac{\text{g-mole}}{22.4 \text{ l}} \times \frac{19.76 \text{ g}}{\text{g-mole}} = 1.32 \text{ g}$$

steam condensed from recycle stream (from Table C.I):

$$1.32 \text{ g} \times \frac{183.4 \text{ lb/hr steam}}{190.4 \text{ lb/hr gas}} = 1.27 \text{ g}$$

steam condensed from fluidization steam:

$$137 \text{ g} - 1.27 \text{ g} = 135.73 \text{ g}$$

mass ratio of recycle stream to fluidization steam:

$$(1.32 \text{ g} + 1.27 \text{ g}) / 135.73 \text{ g} = 0.0188$$

The fluidization steam rate was 201 lb/hr, thus the mass rate of recycle stream was 3.78 lb/hr with 1.93 lb/hr of the gas and 1.85 lb/hr of steam. The temperature and the pressure near the inlet of the recycle gas were 770°F and 4" W.C. gauge, respectively. Therefore, the volume flow rate of the recycle stream is calculated as follows:

gas:

$$1.93 \frac{\text{lb}}{\text{hr}} \times \frac{\text{lb-mole}}{19.76 \text{ lb}} \times \frac{359 \text{ scf}}{\text{lb-mole}} = 35.1 \text{ scf/hr}$$

$$35.1 \frac{\text{scf}}{\text{hr}} \times \frac{(770 + 460)}{(32 + 460)} \times \frac{406.9}{(406.9 + 4)} = 86.8 \text{ cf/hr}$$

steam (specific volume at this condition is 50 cf/lb):

$$1.85 \text{ lb/hr} \times 50 \text{ cf/lb} = 92.5 \text{ cf/hr}$$

total volume flow rate of the recycle stream:

$$86.8 \text{ cf/hr} + 92.5 \text{ cf/hr} = 179.3 \text{ cf/hr}$$

Alternate Feed Material

An additional run was made to attempt to evaluate the reactor performance on an alternate feed material. The feed material (whole rice hulls) were packaged in 50-lb bags and were placed on a piece of plywood braced on top of the feed bin. The rice hulls were slowly dumped into the live bottom feed bin. The live bottom carried the feed to the injection screw of the reactor. The material was fed smoothly for the first 10 to 15 minutes after switching from sawdust to rice hulls. Smoke began to rise from the feed bin due to the larger void space existed in the loosely packed rice hulls. It was also apparent that the feed rate for the rice hulls was too low, thus the feed rate was increased. The smoke stopped and the consumption of the rice hulls increased. The temperature dropped more than 100°F which confirmed the feed material was pyrolyzed inside the reactor.

The feeder worked fine for another 30 minutes but again smoke began to come out of the feed bin. Temperatures at this time had risen about 100°F. The rate of consumption of the third bag of rice hulls was found to be reduced. At the same time the propane supply developed problems and the pressure at the main control valve could not be maintained at a steady level. The temperature at the flue gas inlet fluctuated more than 400°F.

After two hours it was decided to shut down the system for examination. Three gas samples were collected and

analyzed during this period. The gas compositions are listed in Table C.I with runs no. 269-271. One of the most interesting points is the H_2/CO ratio ranged between 3.1 and 6.1 with hydrogen concentration over 50% (volume) on each of the samples. Although feeding was almost zero during this period, the rice hulls that had already been fed into the reactor were gasified along with some of the sawdust that had been left over. Comparison of the components, as shown in Fig. C.4, between the gas produced with sawdust only and the gas mainly from rice hulls, the hydrogen concentration increased from 32-40% to 54-62%, CO_2 from 18-20% to 23-24%, while methane decreased from 9-11% to 2-5%, and CO from 25-34% to 10-15%. However, the data collected from the upset condition was incomplete and hard to analyze.

Later examination of the feed bin showed the front of the bin to be packed with rice hulls. This packing occurred at point where the hulls are carried to injection screw. After clearing the packed layer, the live bottom screw flightings were found rolled back at the front explaining the lack of consumption of rice hulls and the unusual flow pattern in the bin. The damage of the front end of the transfer screw was judged to have happened after the speed-up of the feed motor. The live bottom screw overloaded the pipes at the drop-out end with rice hulls. The rice hulls packed tighter and tighter until the transfer screw rolled back and slipped. This resulted in a situation where the rice hulls moved upward and around in the feed bin

Vol %

FIGURE C.4 COMPARISONS OF GAS COMPOSITION BETWEEN WOOD AND RICE HULLS

itself. The unstable low pressure of the propane supply reduced the heat input to a point less than normal. Therefore, the reactor temperature did not raise any further even without feeding.

Summary

In summary, the recycle runs showed good improvement on gas quality when higher H_2/CO ratio was desired. The heat input was also decreased due to the exothermic water gas shift reaction accompanying the recycle stream. Further study with higher recycle flows would help in further understanding the relationships and the results obtained in these experiments. It appears that the total efficiency of the gasifier system can be improved while adjusting the H_2/CO ratio to a desired value simply by controlling the recycle flow.

The problem using alternate feed material for this system is expected to be resolved with simple modifications of the feeder design. The modification will not only consider the problem associated with rice hulls in this study, but will also consider the problems that might occur when other alternate feed materials are used. The gas quality is another interesting item worth for further investigation.

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