

Studies on the Gasification of Biomass Fuels With Oxygen-enriched Air

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ABSTRACT

A mathematical model developed for the calculation of gas composition and other performance parameters in the gasification of biomass fuels with oxygen or oxygen-enriched air presented results which are substantially consistent with experimental results. Good comparison of calculated gas compositions with Orsat analyses of gas samples from experimental runs on the gasification of coke, coconut shell and ipil-ipil wood validate the model.

The experiments provided evidence that biomass gasification at atmospheric pressure with pure oxygen is possible without encountering excessively high temperatures and, consequently, alleviating the problem of ash clinkering, if the gas producer is operating in the downdraft mode. High gas calorific values and high rates of gasification were observed to be marked advantages of oxygen-enrichment of the gasifying medium over ordinary air. A trend for cold gas thermal efficiency of the gas producer to improve with increasing oxygen content of the gasification medium was also observed.

INTRODUCTION

An alternative automotive fuel in the form of methanol can be produced from biomass resources. Biomass gasification is the first step in biomethanol production. Biomass is gasified in an oxygen-blown downdraft gasifier into synthetic gas consisting primarily of hydrogen and carbon oxides, and eventually is catalytically reacted to form methanol. Methanol is a good substitute fuel for gasoline engines, either straight or blended with gasoline. It could also be reacted with coconut oil to produce methyl esters which is prospectively a good fuel for diesel engines.

Design of gas producers and optimization of their operation are among recent studies undertaken in the U.P. College of Engineering (5). These were largely empirical processes based on experience and past experimental data. The present study focuses on theoretical considerations using basic principles of thermodynamics and chemical equilibrium (4). One of the problems in conducting experiments in biomass gasification, for instance, is the accurate measurement of the composition of the end products of gasification (CO , H_2 , CH_4 , CO_2). The common method of doing this is by gas chromatography or by absorption of the component gases by various chemicals, as in the Orsat Analyzer. However, high cost of chemicals (some of which have to be obtained overseas) or the unavailability of test equipment does not always allow complete gas analyses to be made in some laboratories. For example, in the present study, measurements of H_2 and CH_4 components were not possible and CO analysis by the Orsat apparatus was difficult, tending to be inaccurate because of stale chemicals. It is desirable, therefore, to be able to predict a complete gas composition when gas analysis is limited to only one or two gas component measurements (like CO_2 and CO). A mathematical model simulating gas producer performance can be developed to accomplish this objective.

The objectives of this study are as follows:

1. To develop a mathematical model based on the principles of thermodynamics and chemical equilibrium for the calculation of gas composition and other performance parameters in gasification of biomass fuels with oxygen or oxygen-enriched air;

2. To study improvements in the gasification of biomass fuels with oxygen or oxygen-enriched air with the aid of the mathematical model; and
3. To validate results of the model with experimental data.

THEORETICAL

The mathematical model assumed a certain percentage approach to chemical equilibrium of the reactions taking place in the gasifier fuel bed at an equivalent reaction temperature obtained through a heat balance of the system. This reaction temperature would depend upon a number of factors, namely: the heating value of the solid fuel, its ultimate analysis, its moisture content, the mode of gasifier operation (updraft or downdraft), the temperature and composition of the gasification medium (percent O₂, N₂ and steam), the heat losses due to radiation and unburnt carbon in the ash residue.

The degree to which chemical equilibrium was attained by each of the pertinent gasification reactions, for instance, the heterogenous water-gas reaction $C + H_2O = CO + H_2$ is defined by certain parameters like \underline{y} in the relation:

$$\underline{y} + K_{pw} = \frac{[CO] * [H_2]}{[H_2O]}$$

where K_{pw} is the equilibrium constant for the heterogenous water-gas reaction at the calculated reaction temperature and the gas components in brackets are the experimental wet gas compositions in percent. Since actual H₂O in the gas could not be experimentally measured, it was estimated to be the same as the calculated equilibrium H₂O. In the calculation of the equilibrium, gas composition takes into account the degree of approach to equilibrium (in percent) defined as \underline{y} in the above equation. The other reactions ($C + CO_2 = 2 CO$, and $C + 2 H_2 = CH_4$) may be assumed to have reached 100% chemical equilibrium if complete gas analyses of the producer gas cannot be measured experimentally.

If complete gas analyses data are available, then the degree of approach to equilibrium of the other two reactions above can be estimated from the following relations:

$$\underline{z} * K_{pb} = \frac{[\text{CO}]^2}{[\text{CO}_2]}$$

$$\underline{x} * K_{pm} = \frac{[\text{CH}_4]}{[\text{H}_2]^2}$$

where z is the percent approach to equilibrium of the Boudouard reaction ($\text{C} + \text{CO}_2 = 2 \text{CO}$), and x that of the methanation reaction ($\text{C} + 2 \text{H}_2 = \text{CH}_4$).

The mathematical model was validated by comparing its results with past experimental data. For instance, Table 1 shows the results of gasification of coke with oxygen and steam in an updraft reactor at the University of New South Wales (2). Tables 2 and 3 summarize experimental and calculated performance of coconut shell gasification with air in updraft and downdraft modes, respectively, at the University of the Philippines (3). These results show that it is possible to estimate the parameters \underline{x} , \underline{y} and \underline{z} (percent approach to equilibrium of the methanation reaction, the heterogenous water-gas reaction and the Boudouard reaction, respectively) so that the calculated gas composition is practically identical to the experimentally-measured gas composition.

A complete discussion of the mathematical model is found in the Appendix.

EXPERIMENTAL

An experimental gasifier used in previous experiments (Figure 1) was modified to allow precise measurements of oxygen and air flows supplied to the reactors during experimental runs. In this small gas producer, the small air inlets (tuveres) were plugged and replaced with a larger (1.5-inch diameter) single air inlet pipe located at the throat of the reactor. Air and oxygen flows were measured by separate rotameters before mixing and being introduced to the gasifier. During an experimental run, the gasifier was mounted on a platform balance so that continuous weight loss of fuel could be monitored during the gasification process. Various biomass fuels were gasified, such as coconut shell, corn cobs, rice husk, coconut husk and ipil-ipil wood. Orsat

analyses of four to five gas samples were made for every run and the averages of these were reported for a given run. Table 4 presents some typical results in the gasification of ipil wood which was the fuel used more extensively in the experiments. Figures 3 to 5 depict graphically the performance of the gasifier at various oxygen enrichment levels of air and total flow of the gasification medium.

DISCUSSION

Table 1 shows the performance of oxygen-steam, pulsed-blast gasification of coke as compared to that when the blast was steady (unpulsed). In the pulsed-blast mode, steam was fed in a steady stream through the reactor grate, and oxygen was introduced in intermittent blasts or pulses. In the unpulsed mode, both steam and oxygen were introduced as a mixture in a steady stream. In the unpulsed mode, therefore, even if H_2 is formed by the reaction $C + H_2O = H_2 + CO$ in the fuel bed near the grate, the presence of O_2 with the product gases will oxidize H_2 to steam again. Therefore, the degree to which the reaction could approach equilibrium will be less compared to gasification in the pulsed-blast mode when O_2 is not present with the product gases between pulses of oxygen. This is verified experimentally by the results shown in Table 1 where the values for \underline{y} and \underline{z} (calculated by means of the mathematical model) are higher for pulsed gasification (e.g. $\underline{y} = 28\%$ and $\underline{z} = 93\%$) than for the unpulsed mode ($\underline{y} = 2\%$ and $\underline{z} = 5\%$). Since no experimental measurement for CH_4 was made, theoretical CH_4 values were calculated by assuming a value for the parameter x of 24% obtained from literature (1).

In Table 2, complete gas analyses, including that of CH_4 , were made for the updraft gasification of coconut shell with air. Thus, estimates for the parameters \underline{x} , \underline{y} and \underline{z} can be calculated from the mathematical model. Approach to equilibrium for the methanation reaction was estimated at $\underline{x} = 100\%$. This was a rough estimate because in Run 1 LUD (Large Up-Draft), for example, the experimental measurement of 0.9% for CH_4 (which might appear inconsistent with the calculated value of $CH_4 = 0.2\%$) was due to the gas sampling and gas composition

measurement techniques which could detect volumetric percentage figures to only within 0.5 percentage point accuracy.

Table 4 shows the typical results of the performance of downdraft gasification of ipil wood with oxygen-enriched air, the percent carbon in the refuse, %c and the losses due to heat radiation and convection in the gasifier. Percent losses are assumed at varying values until the experimental %CO₂ and ratio F/M (ratio of lb dry fuel gasified per mol of gasification medium) equal the calculated %CO₂ and ratio F/M, respectively. (See the Appendix for an example calculation.) It was noted that when the chemicals used in the Orsat Analyzer were fresh, the experimentally-measured CO values were substantially equal to the calculated values. Since there was no way of experimentally measuring the H₂ and CH₄ content of the gases at the time of the experiments, the calculated values provided good estimates of these measurements.

The performance parameters, namely, Higher Heating Value (HHV) in Btu/cubic foot of producer gas, Gasification Rate in Kg/hour and Cold Gas Thermal Efficiency (%), are correlated as functions of Gasification Medium Flow in cubic feet/hour and percent oxygen in the gasification medium. The results are presented as graphs in Figures 2 to 4.

Figure 2 shows a good correlation of HHV against %O₂ in the medium, indicating a significant increase in gas calorific value as the oxygen enrichment of the air gasifying medium is increased.

Figure 3 shows that gasification rate increased markedly both with an increase in the flow rate of the gasifying medium and increase in the oxygen enrichment of the air. This is consistent with considerations of material balance and stoichiometry of reaction involved in the process of gasification.

Figure 4 shows a trend for the cold gas thermal efficiency to increase with oxygen-enrichment. The experimental points, however, are too scattered to indicate any trend as to the effect of flow rate of gasifying medium. The fact that the gasifier was operated in short batch runs, lasting from 1- 3 hours, resulted in significant ungasified carbon in the ash rejects (%c) at varying values (from 88%-97%), which accounts for the scattering of efficiency values, from a low of 26% to a high of 68%. In other words, the gasifier produced a by-product, particularly in the downdraft operation, and this was charged as a loss in the

computation for efficiency. Gasifiers continuous operation, for say at least 10 hours, would result in less unburnt carbon in the ash residue.

One of the concerns when using oxygen or oxygen-enriched air for gasification is the possibility of ash clinkering because of high temperatures reached in the combustion zone. No such problem was encountered when the gasifier was operated in the downdraft mode even when the gasifying medium was 100% oxygen. However, in a single run in the updraft mode, melting and fusion of ash were observed after 2 hours of operation at 30% oxygen content in the gasifying medium. The clinker did not form maybe because the combined water content of wood was considerable ($V_{H_2O} = 0.64$ or 64% mol/mol gasifying medium) and most of this passed through the combustion zone in downdraft mode, cooling the combustion zone in the process. This was not the case in updraft operation.

CONCLUSIONS

The mathematical model developed for the calculation of gas composition and other performance parameters in the gasification of biomass fuels with oxygen or oxygen-enriched air presented results which are substantially consistent with experimental results. More confidence in the validity of the model could be established, however, if more data from longer experimental runs using other biomass fuels are made. A good comparison of calculated gas composition with the complete analysis of gas samples from these experimental runs could further validate the model. Such additional data are recommended to be gathered in further experiments.

Biomass gasification at atmospheric pressure with pure oxygen is possible without encountering the problem of ash clinkering if the gas producer is operating in the downdraft mode. High gas calorific values and high rates of gasification are marked advantages of oxygen-enrichment of the gasifying medium over ordinary air. There is also a trend for cold gas thermal efficiency to improve with increasing oxygen content of the gasification medium.

Table 1. Comparison Between Pulsed and Unpulsed Blast Updraft Gasification of Coke with Oxygen and Steam

Run	33 Pulsed		27 Unpulsed	
Air Rate, cu ft/h	140		150	
% O ₂ , Air-O ₂ Mix	100		100	
% Steam	75		76	
Reaction Temp., deg. C			748	84.7
% Gas	Expt'l	Calcul'd	Expt'l	Calcul'd
% CO ₂	8.5	8.5	19.0	18.8
% CO	55.2	54.2	42.5	41.9
% H ₂	36.3	36.8	38.5	39.0
% CH ₄	0.0	0.2	0.0	0.1
% N ₂	0.0	0.3	0.0	0.3
Higher Heating Value				
Btu/cu ft	313	313	277	277
Cold Gas				
Efficiency, %	90.8	90.9	81.3	81.3
Approach to Equilibrium				
x, %			24	24
y, %			24	3
z, %			100	4
Gasification Rate				
Kg/h	7.93		6.99	
Run	28 Pulsed		29 Unpulsed	
Air Rate, cu ft/h	150		140	
% O ₂ , Air-O ₂ Mix	100		100	
% Steam	66		80	
Reaction Temp., deg. C			770	755
% Gas	Expt'l	Calcul'd	Expt'l	Calcul'd
% CO ₂	7.7	7.7	32.4	32.5
% CO	60.3	59.4	31.7	31.3
% H ₂	32.0	32.5	35.6	35.8
% CH ₄	0.0	0.1	0.0	0.1
% N ₂	0.0	0.3	0.3	0.3
Higher Heating Value				
Btu/cu ft	315	315	230	231
Cold Gas				
Efficiency, %	87.7	87.7	59.2	59.3
Approach to Equilibrium				
x, %			24	24
y, %			28	2
z, %			93	5
Gasification Rate				
Kg/h	7.42		4.64	

Table 2. Updraft Gasification of Coconut Shell with Air (Gas Producer Grate Area = 0.9 Square Meter)

Run	1 LUD	2 LUD		
Air Rate, cu ft/h	13915	26377		
% O ₂ , Air - O ₂ Mix	21	21		
% Steam	0	0		
Reaction Temp., deg. C		703		703
% Gas	Expt'l	Calcul'd	Expt'l	Calcul'd
% CO ₂	8.9	8.1	7.1	6.9
% CO	25.7	22.8	25.6	24.5
% H ₂	11.6	11.5	11.3	11.4
% CH ₄	0.9	0.2	0.5	0.1
% N ₂	52.9	57.5	55.4	57.1
Higher Heating Value Btu/cu ft	137	119	131	123
Cold Gas Efficiency, %	83.1	72.0	78.3	73.6
Approach to Equilibrium				
x, %		100		100
y, %		62		18
z, %		55		18
Gasification Rate kg/h	175.50	342.00		

Table 3. Downdraft Gasification of Coconut Shell with Air (Gas Producer Grate Area = 0.9 Square Meter)

Run	4 LDD	5 LDD		
Air Rate, cu ft/h	19280	21350		
% O ₂ Air-O ₂ Mix	21	21		
% Steam	0	0		
Reaction Temp., deg. C		645		673
% Gas	Expt'l	Calcul'd	Expt'l	Calcul'd
% CO ₂	12.0	12.7	13.3	14.1
% CO	16.0	16.9	12.7	13.6
% H ₂	15.7	15.5	12.9	12.9
% CH ₄	0.0	0.6	0.0	0.3
% N ₂	56.3	54.4	61.1	59.1
Higher Heating Value Btu/cu ft	108	117	88	94
Cold Gas Efficiency, %	67.7	72.8	58.8	62.8
Approach to Equilibrium				
x, %		100		100
y, %		40		14
z, %		68		20
Gasification Rate kg/h	250.20	233.10		

Table 4. Downdraft Gasification of Ipil Wood with Air, Oxygen-Enriched Air and Pure Oxygen

Run	1	2	3	4	5
Air Rate, cu ft/h	150	250	200	400	400
% O ₂ , Air-O ₂ Mix	40	40	21	21	21
% Steam	0	0	0	0	0
Reaction Temp., deg. C	681	667	595	671	640
% Gas					
% CO ₂	12.6	15.3	14.8	8.4	11.1
% CO	34.3	30.7	13.5	23.7	19.5
% H ₂	23.4	24.6	18.0	13.3	15.1
% CH ₄	0.7	1.0	1.1	0.3	0.5
% N ₂	29.1	28.4	52.5	54.3	53.8
Higher Heating Value					
Btu/cu ft	204	200	120	129	123
Cold Gas Efficiency, %	54	67	26	63	43
Carbon in Refuse (c) %	94.5	90.0	97.0	90.0	95.5
Losses (Radiation, etc.), %	3	5	2.5	5	4
Approach to Equilibrium					
x, %	100	100	100	100	100
y, %	15	35	15	15	15
z, %	100	100	100	100	100
Gasification Rate					
kg/h	5.40	7.80	4.05	6.00	7.26
Run	11	12	13	14	15
Air Rate, cu ft/h	300	200	400	150	300
% O ₂ , Air-O ₂ Mix	40	100	40	100	40
% Steam	0	0	0	0	0
Reaction Temp., deg. C	648	701	672	624	671
% Gas					
% CO ₂	17.1	16.3	12.9	28.8	15.5
% CO	27.4	48.4	33.4	27.4	31.4
% H ₂	26.3	34.2	22.8	40.1	19.7
% CH ₄	1.3	1.2	0.6	3.7	0.6
% N ₂	28.0	0.0	30.3	0.0	32.8
Higher Heating Value					
Btu/cu ft	197	295	199	270	181
Cold Gas Efficiency, %	47	66	41	44	55
Carbon in Refuse (c), %	95.5	92.0	96.0	96.0	91.0
Losses (Radiation, etc.), %	3	3	3	2.5	7
Approach to Equilibrium					
x, %	100	100	100	100	100
y, %	23	23	10	40	40
z, %	100	100	100	100	100
Gasification Rate					
Kg/h	12.00	17.20	16.00	16.40	7.00

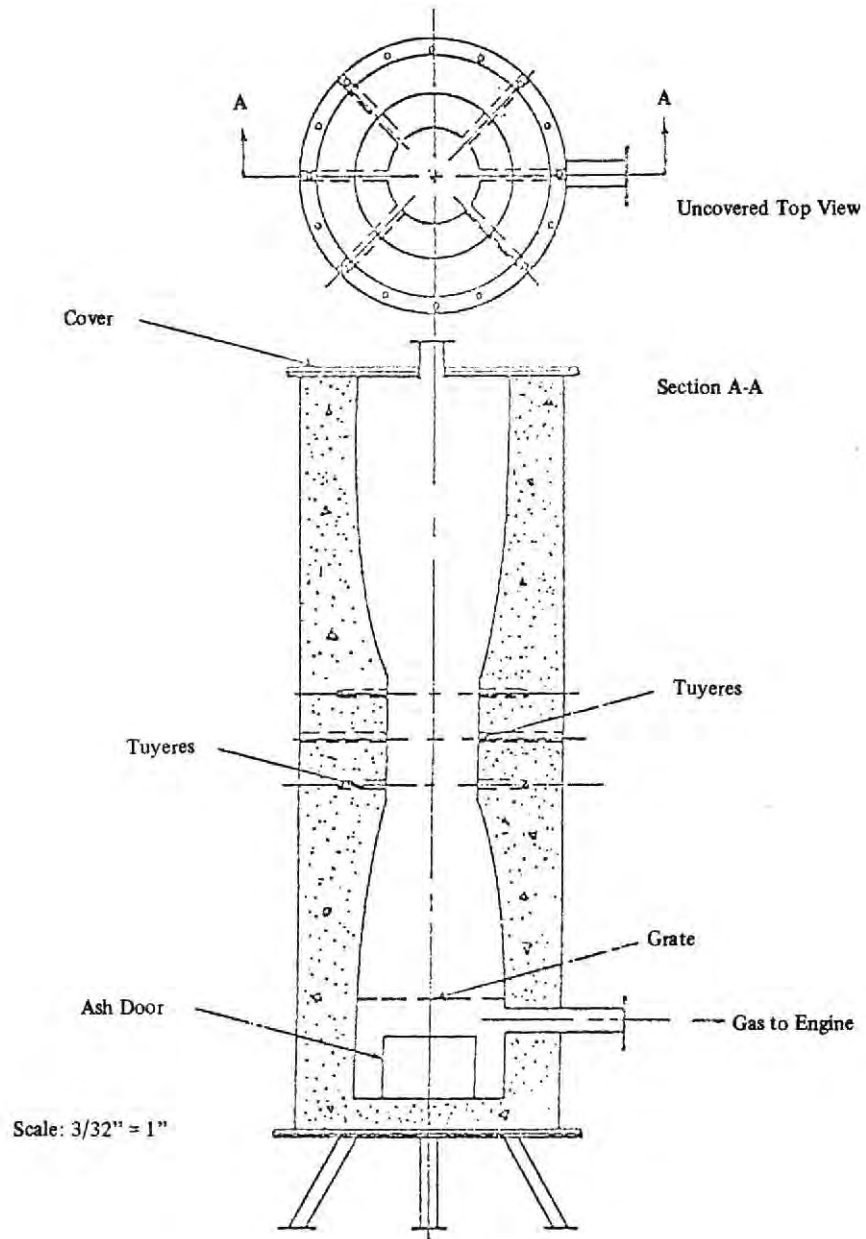


Figure 1. Suction Downdraft Gas Producer

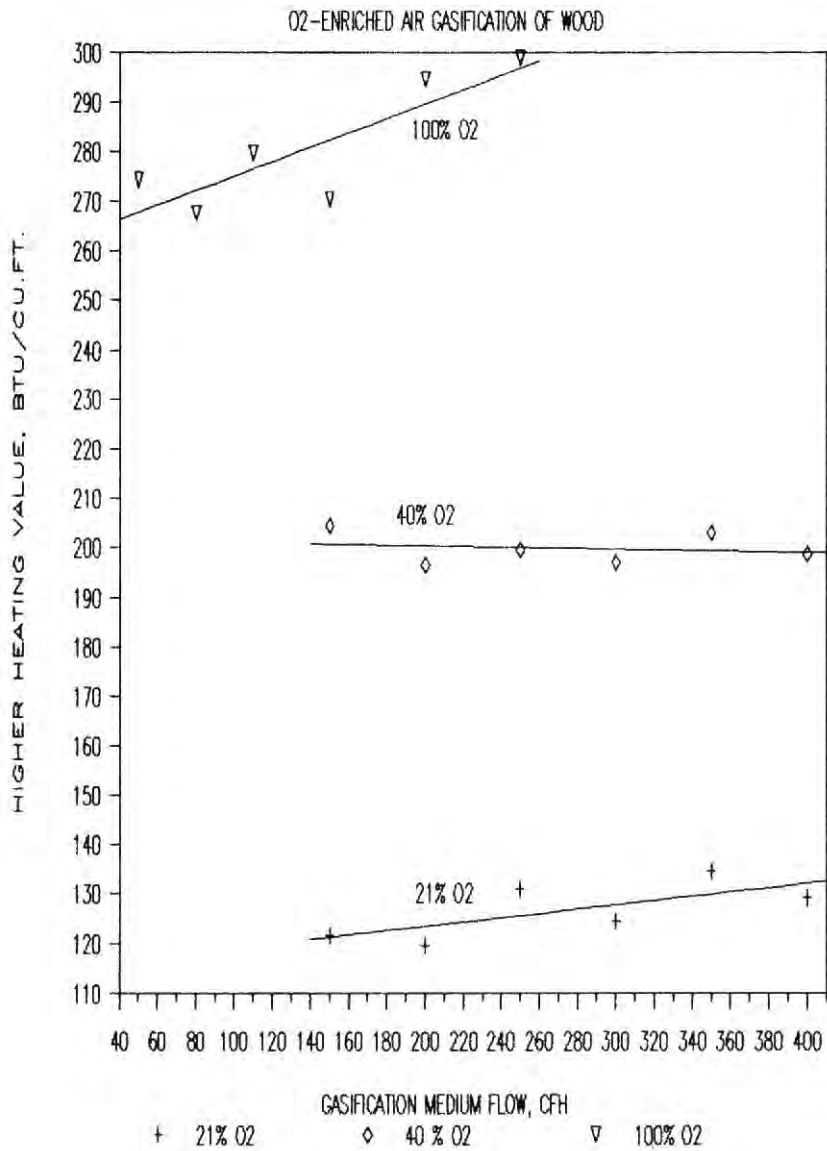


Figure 2. Higher Heating Value

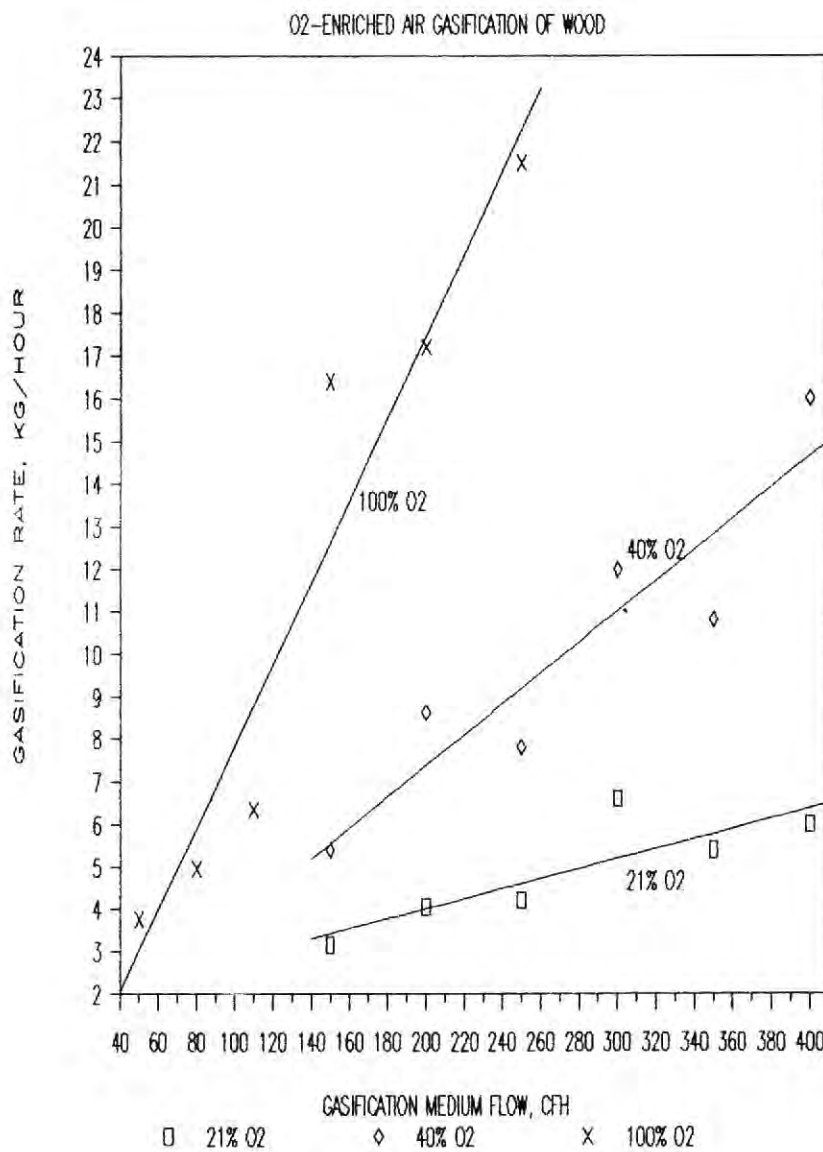


Figure 3. Gasification Rate, kg/hour

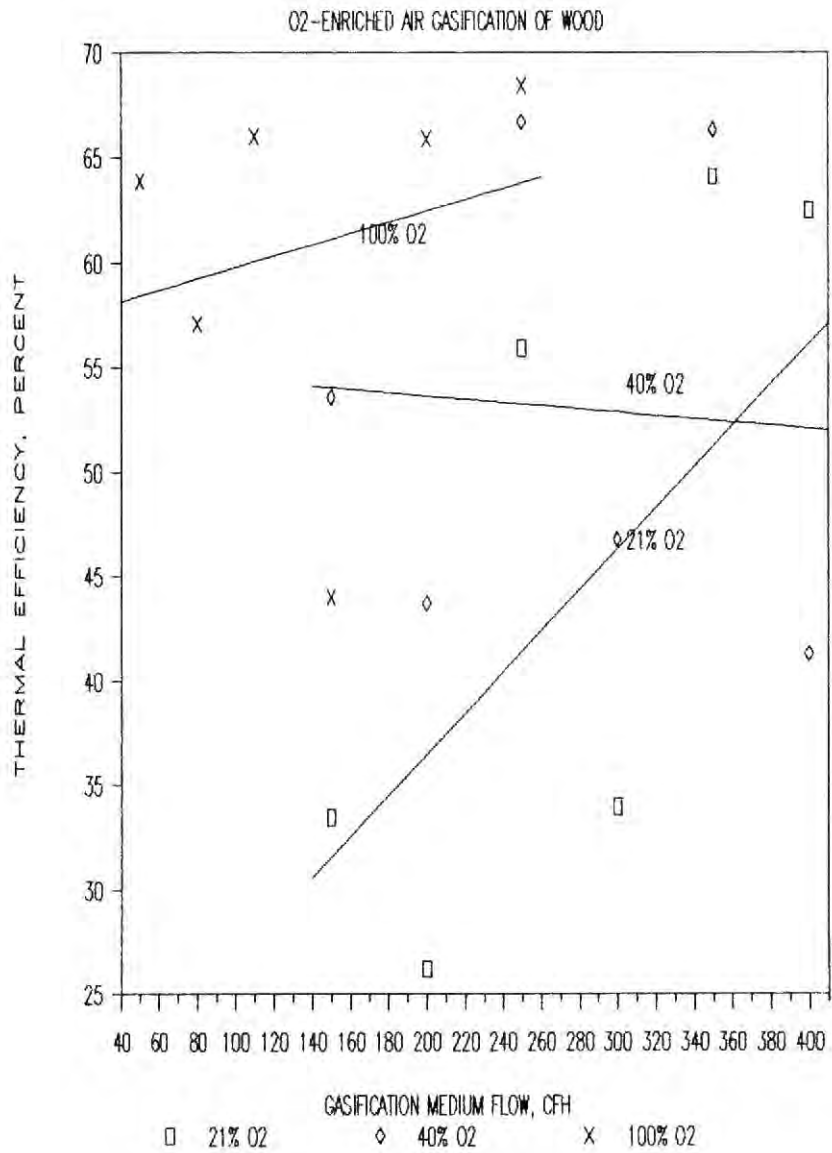


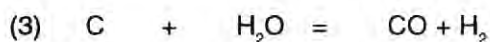
Figure 4. Cold Gas Thermal Efficiency

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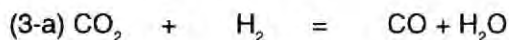
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APPENDIX
DEVELOPMENT OF MATHEMATICAL MODEL FOR THE CALCULATION
OF GAS COMPOSITION AND OTHER PERFORMANCE
PARAMETERS IN THE GASIFICATION OF BIOMASS FUELS WITH
OXYGEN AND STEAM

Consider first the gasification of carbon in a deep fixed bed, where the following reactions are believed to be occurring:



Equation (1) occurs in the so-called oxidation zone while equations (2) to (4) occur in what is termed the reduction zone of the fuel bed. Another reaction, what is known as the homogenous water-gas shift reaction (3-a) can be obtained from (2) and (3):



Let

$V''H_2O, V''O_2, V''N_2$ = Volumetric composition of gasification medium entering the oxidation zone

$V'H_2O, V'O_2, V'N_2$ = Volumetric composition of gases entering the reduction zone

$VH_2O, VO_2, VN_2, VCO, VH_2, VCH_4$
 = Volumetric composition of producer gas

Therefore

$$(5) \quad V'H_2O + V'O_2 + V'N_2 = 1$$

Assuming that in the oxidation zone, the primary reaction is the oxidation of C to CO_2 , then

$$(6) \quad V'H_2O + V'CO_2 + V'N_2 = \text{or} > 1$$

equal to 1 for carbon gasification or greater than 1 in the case of gasification of biomass fuel which contains combined water.

$$(7) \quad VH_2O + VCO_2 + VN_2 + VCO + VH_2 + VCH_4 = 1$$

Consider reaction (2), (also known as the Boudouard reaction) and let

A_b = degree of conversion of CO_2 into CO , with a value varying anywhere between 0 and 1 (negative value indicates reaction in the backward direction).

Therefore

$$(8) \quad C + CO_2 = 2 * A_b CO + (1-A_b) CO_2 + (1-A_b) C$$

If $V'CO_2$ (in mols) is the amount of CO_2 entering the reduction zone, then

$$2 A_b V'CO_2 = \text{Amount of } CO \text{ formed}$$

$$(1-A_b) * V'CO_2 = \text{Amount of } CO_2 \text{ remaining}$$

Consider reaction (3), also known as the heterogenous water-gas reaction) and let

A_w = degree of conversion of steam to CO and H_2
Then,

$$(9) \quad C + H_2O = A_w CO + A_w H_2 + (1-A_w) H_2O + (1-A_w) C$$

If, $V'H_2O$ (in mols) is the amount of steam entering the reduction zone, then

$$A_w V'H_2O = \begin{matrix} \text{amount of } H_2 \text{ formed} \\ \text{amount of } CO \text{ formed} \end{matrix}$$

$$(1-A_w) V'H_2O = \text{amount of steam undecomposed}$$

Consider reaction (4), the methanation reaction, and let:

A_m = degree of conversion of H_2 to CH_4
Then

$$(10) \quad C + 2H_2 = A_m CH_4 + 2 (1-A_m) H_2 + (1-A_m) C$$

The amount of H_2 going into the above reaction comes from reaction (9). This amount is equal to $A_w V H_2O$ mols. Therefore,

$$0.5 A_m V_w V H_2O = \text{amount of } CH_4 \text{ formed}$$

$$(1-A_m) A_w V H_2O = \text{amount of } H_2 \text{ remaining}$$

The amount of gases in the products of gasification are tabulated as follows:

TABLE 1

GAS	AMOUNT, MOLS
CO_2	$(1-A_b) V CO_2$
H_2O	$(1-A_w) V H_2O$
CO	$2A_b V H_2O + A_w V H_2O$
H_2	$(1-A_m) A_w V H_2O$
CH_4	$0.5 A_m A_w V H_2O$
N_2	$V N_2$

Table 1 contains the amounts of gases in terms of the input gas medium ($V CO_2, V H_2O, V N_2$) which are known and the three parameters A_b, A_w and A_m which are unknown. Three equations are needed to solve for these three unknowns. These are provided by the three equations corresponding to the equilibrium constants, respectively, of the homogenous water gas shift reaction (K_w), the methanation reaction (K_m) and the heterogenous water-gas reaction (K_{pw}).

Total mols of products is equal to the sum of the mols of gases in Table 1, or after simplifying:

$$(11) \text{ SUM } V H_2O + V CO_2 + V N_2 + A_b * V CO_2 + (1-0.5 * A_m) A_w * V H_2O \text{ mols/mol blast}$$

An expression for A_b can be obtained from the equilibrium constant equation for the homogenous water-gas reaction:

$$(12) K_w = V CO * V H_2O / V CO_2 / V H_2$$

Substituting the corresponding molal gas concentrations as given in Table 1 and simplifying results in:

$$(13) \quad A_b = \frac{K_w - V'H_2O/V'CO_2 * (1 - A_w) / (1 - A_m)}{K_w + 2/A_w * (1 - A_w)/(1 - A_m)}$$

Calculation is started by assuming values for A_w and A_m and solving equation (13) for A_b . These trial values can be used to solve for gas composition as follows:

$$(14) \quad \begin{aligned} VCO_2 &= (1 - A_b) * V'CO_2 / \text{SUM} \\ VH_2O &= (1 - A_w) * V'H_2O / \text{SUM} \\ VCO &= (2A_b V'CO_2 + A_w V'H_2O) / \text{SUM} \\ VH_2 &= (1 - A_m) * A_w V'H_2O / \text{SUM} \\ VCH_4 &= 0.5 * A_m * A_w * V'H_2O / \text{SUM} \\ VN_2 &= V'N_2 / \text{SUM} \end{aligned}$$

After the volumetric fractions of the gas constituents are computed using equation (14), the equilibrium constants for the heterogenous water-gas reaction are calculated as follows:

$$(15) \quad K_{pw} = \frac{PT * VCO * VH_2}{H_2O}$$

$$(16) \quad K_m = \frac{VCH_4}{VH_2^2 / P_T}$$

where P_T is the total pressure of the product gases, in atmospheres.

If the computed values of K_{pw} and K_m do not equal the correct values of these constants at the given temperature (obtained from data tables or from equations of equilibrium constants expressed in terms of temperature: see Table 2), the whole procedure is repeated with corrected values of A_w and A_m until a solution is obtained.

Gasification with oxygen of biomass fuel is equivalent to gasification of the carbon content of the fuel with accompanying steam equal to the combined water in the biomass. The following sample calculation will illustrate this better.

Sample Calculation

Ipil wood was gasified in a down-draft gasifier with oxygen-enriched air (40% O₂) at a flow rate of 300 cubic feet per hour at 30°C. An initial charge of 27.3 Kg of ipil wood was gasified to heat up and stabilize the gasifier for 125 minutes leaving 4.9 kg of char in the reactor at the end of this time. To start an experimental run, 17.4 Kg of fresh ipil wood was added to the 4.9 Kg of char. The char's ultimate analysis (dry basis) was approximately 96% C and 4% ash; that of ipil wood: 48.6% C, 6.0% H, 44.4% O and 1% ash; the higher heating values (HHV) were 12,971 and 8144 Btu/lb.. The gasifier was mounted on a platform balance and fuel weight loss measurements during gasification were made at 5-minute intervals. The total weight loss during a 150-minute run was 17.5 Kg. Thus, the apparent gasification rate was $17.5/150 \times 60 = 7.0$ Kg/h. The other operating parameters for the 150-minute run were as follows.

1. Equivalent ultimate analysis of mixture of ipil wood and char is weighted average of 17.4 Kg (78%) ipil wood and 4.9 Kg (22%) char, or 59.02% C, 4.68% H, 34.64% O and 1.66% Ash. The weighted average of HHV is 9205 Btu/lb (dry basis). The moisture content was 13.5% for ipil wood and 10.3% for the char.
2. The total weight of ipil wood gasified was $27.3 + 17.4 = 44.7$ Kg. The weight of ash accumulated in the reactor would be $(0.01) \times (44.7) \times (1-0.135) = 0.387$ Kg. The dry ash and char accumulated in the reactor at the end of the run was $3.5 \times (1-0.103) = 3.14$ Kg. Therefore, the percentage carbon (%c) in the dry refuse (ash + char) was $100 \times (3.14-0.387)/3.14 = 88\%$. Carbon in the ash (p), in terms of Kg/Kg (or lb/lb) dry fuel would be: $p = \% \text{ ash} \times \%c / (100 - \%c) / 100 = 0.12173$ or 12.173 lb C/100 lb dry fuel.
3. Since 12.173 lb C/100 lb dry fuel remained with the ash, the net C in the fuel that was gasified was $59.02 - 12.173 = 46.847$ lb/100 lb dry fuel and the amount of refuse or rejects (ash + C) was $1.66 + 12.173 = 13.833$ lb/100 lb fuel. The effective ultimate analysis of the fuel gasified, therefore, was as follows: 46.85% C, 13.83% Refuse (ash + C), 4.68% H and 34.64% O.
4. The effective gasification rate, in pounds dry fuel per hour - (apparent gasification rate) $\times (1 - \%M) / (1 - \% \text{Refuse} / 100)$ or $7.0 \times (1 - 0.135) / (1 - 0.13833) = 7.03$ Kg/h. The gasification medium flow in mols/h was $300 \text{ ft}^3/\text{h} / 359 \times (273/30) = 0.7529$ mol/h. However, since the average Orsat analysis of 8 gas samples showed 0.66% O, (which indicated that not all the oxygen was reacted), a correction factor (less than 1

must be used to reduce the medium flow to its effective rate). If an approximate N_2 content of the producer gas is estimated as 34.3% (see item 8 below), then the correction factor is equal to $(34.4 - 0.66 * 60/40)/34.3 = 0.971$. The effective rate of medium flow would therefore be $0.971 * 0.7529 = 0.7311$ mol/h.

5. The amount of combined water in the dry fuel, in moles/lb was: $0/16/100 = 34/16/100 = 0.02165$. These combined water plus the physical moisture (%M) of the fuel would constitute the effective moisture content in the gasification medium, $V'H_2O$, in mol/mol of gasification medium. Thus,

$$(17) V'H_2O = V'H_2O/100 + \{0/16/100 + \%M/(100 - \%M)/18\} * F/M$$

where F/M is the pounds of fuel gasified per mol of gasification medium: From item 4 above, $F/M = 7.03 * 2.2/0.7311 = 21.15$ lb dry fuel/mol medium. No steam was added to the gasification medium ($V''H_2 = 0$) thus the steam entering the reduction zone, from the above equation would be:

$$\begin{aligned} V'H_2O &= 0 + (0.02165 - 13.5/100 - 13.5)/18 * 21/15 \\ &= 0.641 \text{ mol/mol gasification medium} \end{aligned}$$

In a downdraft reactor, the full amount of combined water and moisture in the fuel is assumed to pass through the reduction zone. In an updraft reactor, only a fraction would pass through the reduction zone because some of the moisture would be distilled off the top of the fuel bed. This fraction is estimated to be a function of the moisture content, %M, of the fuel (equal to $0.5 - 0.006 * \%M$).

6. The gasification medium entering the oxidation zone was 40% O_2 , 60% N_2 and no steam. Thus $V''O_2 = 0.40$, $V''N_2 = 0.60$ and $V''H_2O = 0$ satisfies equation (5):

$$V''H_2O + V''O_2 + V''N_2 = 1$$

7. The gases entering the reduction zone were $V'CO_2 = 0.40$, $V''N_2 = 0.60$ and $V'H_2O = 0.641$. From equation (6):

$$V''H_2O + V'CO_2 + V''N_2 = 0.40 + 0.60 + 0.641 = 1.641$$

8. At an assumed reaction temperature of $667^\circ C$, the producer gas composition was calculated with the following results (see calculations in item 11 below).

	% Dry Basis	% Wet Basis
VH ₂ O	0.0	16.5
VCO ₂	15.9	13.3
VCO	31.1	26.0
VH ₂	18.2	15.2
VCH ₄	0.5	0.4
VN ₂	34.3	28.6
TOTAL	100.0	100.0

9. Orsat analyses of 8 gas samples yielded the following results:

	% Dry Basis	%Wet Basis
VO ₂	0.66	0.0
VCO ₂	15.5	15.9
VCO	26.1	26.7

It is noted that the experimental VCO₂ (air-free basis) is equal to the calculated or theoretical value in Item 8 above, but the experimental VCO value, 26.7%, is somewhat less than the theoretical value shown in Item 8 (31.1%). Measurement of CO by the Orsat Apparatus was, however, tedious and difficult, and it was possible that not all the CO were absorbed by the chemicals particularly when the absorbent became stale after some use.

10. The amount of fuel gasified, F, and the amount of gasification medium used, M, may be computed from the gas analysis and ultimate analysis of the fuel:

$$(18) \quad F = 12 + (VCO + VCO_2 + VCH_4)/(\%C) \text{ lb dry fuel/mol gas}$$

$$(19) \quad M + (VN_2 - N_2 \text{ from fuel})/(V"N_2) \text{ mol/mol gas}$$

Thus,

$F = 12 * (26.0 + 13.3 + 0.4)/46.85 + 10.16$ lb dry fuel/mol wet gas
 $M = (28.6 - 0)/60 = 0.4767$ mol/mol wet gas and $F/M = 10.16/0.477 = 21.29$ lb dry fuel/mol gasification medium. (Compare this with the experimental value of $F/M = 21.15$ lb dry fuel/mol gasification medium obtained in Item 5 above.)

11. Calculations for the gas composition

Composition of gasification medium:

Entering oxidation zone:

$$V''\text{H}_2\text{O} = 0, V''\text{O}_2 = 0.40, V''\text{N}_2 = 0.60$$

Entering reduction zone:

$$V'\text{CO}_2 = 0.40, V'\text{N}_2 = 0.60$$

Substituting $F/M = 21.29$ in equation (17):

$$\begin{aligned} V'\text{H}_2\text{O} &= 0 + \{0.02165 + 13.5/(100 - 13.5)/18\} * 21.29 \\ &= 0.646 \text{ mol/mol gasification medium.} \end{aligned}$$

Use final trial value for reaction temperature, $T_r = 667^\circ\text{C}$ (see calculations in Item 12 below). From appropriate tables or equations for equilibrium constants (see Table 2), obtain the constants for the water-gas reactions (3) and (3-a) and the methanation reaction (4):

$$K_{pw} = 0.9191 \quad K_w = 0.5545 \quad K_m = 0.1948$$

In biomass gasification, after the hydrogen content of the fuel combines with the oxygen to form the combined water, there is usually a net hydrogen left which amounts to:

$$(19) H_{2net} = (H/2 - 0/16)/100 * F/M \text{ mol/mol gasification medium}$$

Therefore, adding H_{net} to the corresponding equations involving hydrogen formation or depletion, the set of equations (14) becomes:

(14-a)

$$\begin{aligned} V\text{CO}_2 &= (1-A_b) * V'\text{CO}_2/\text{SUM} \\ V\text{H}_2\text{O} &= (1-A_w) * V'\text{H}_2\text{O}/\text{SUM} \\ V\text{CO} &= (2 A_b V'\text{CO}_2 + A_w V'\text{H}_2\text{O})/\text{SUM} \\ V\text{H}_2 &= (1-A_m) * (A_w * V'\text{H}_2\text{O} + H_{2net})/\text{SUM} \\ V\text{CH}_4 &= 0.5 * A_m (A_w * V'\text{H}_2\text{O} + H_{2net})/\text{SUM} \\ V\text{N}_2 &= V'\text{N}_2/\text{SUM} \end{aligned}$$

Also equations (11) and (13) become:

$$(11-a) \quad \text{SUM} = V'H_2O + V'CO_2 + V'N_2 + A_b * V'CO_2 + (1 - 0.5 * A_m) * (A_w * V'H_2O + H_{2not})$$

mol/mol gasification medium

$$(13-a) \quad A_b = \frac{K'_w * C_{Hn} - V'H_2O/V'CO_2 * (1 - A_w)/(1 - A_m)}{K'_w * C_{Hn} + 2/A_w * (1 - A_w)/(1 - A_m)}$$

where C_{Hn} is a correction factor which can be derived to give:

$$(20) \quad C_{Hn} = (1 + H_{2not}/A_w/V'H_2O)$$

In biomass gasification, a factor, y , may be used to designate approach to equilibrium of the heterogenous water-gas reaction (3). A value of y less than 1 denotes less than 100% approach to equilibrium. Thus, equation (15) becomes:

$$(15 - a) \quad K'_{pw} = y * K_{pw} = P_1 * VCO * VH_2/VH_2O$$

Consequently, the homogenous water-gas shift reaction equilibrium constant becomes:

$$(12 - a) \quad K'_w = K_w/y = VCO * VH_2O/VCO_2/VH_2$$

A factor x to denote approach to equilibrium of the methanation reaction (4) may also be used. Thus equation (16) becomes:

$$(16 - a) \quad K'_m = x * K_m = P_1 * VCH_4/VH_3^2$$

Assuming a value for $y = 0.26$ (x is usually 1.00 for biomass gasification), the equilibrium constants used were:

$$K'_{pw} = 0.9191 * 0.26 = 0.2391, \quad K'_w = 0.5545/0.26 = 2.1325$$

$$K'_m = 0.1948 * 1.00 = 0.1948$$

Using trial values of $A_w = 0.4640$ and $A_m = 0.0558$ and substituting into equations (19), (20) and (13 - a):

$$\begin{aligned}
 H_{2net} &= (H/2 - 0/16)/100 * F/M \\
 &= (4.68/2 - 34.64/16)/100 * 21.29 \\
 &= 0.03726 \text{ mol/mol gasification medium}
 \end{aligned}$$

$$\begin{aligned}
 C_{Hn} &= (1 + H_{2net}/A_w/V'H_2O) \\
 &- (1 + 0.03726/0.464/0.646) = 1.1244
 \end{aligned}$$

$$\begin{aligned}
 A_b &= \frac{2.1325 * 1.1244 - 0.646/0.40 * (1 - 464)/(1.0558)}{2.1325 * 1.1244 + 2/0.464 * (1 - 0.464)/(1 - 0.0588)} \\
 &= 0.3058
 \end{aligned}$$

From equation (11 - a)

$$\begin{aligned}
 \text{SUM} &= 1.646 + 0.308 * 0.4 + (1 - 0.5 * 0.0558) * (0.464 * 0.646 \\
 &+ 0.03726) = 2.0954 \text{ mols/mol gasification medium}
 \end{aligned}$$

From equation (14 - a)

$$VCO_2 = (1-0.3058)*0.4/2.0954 = 0.1325$$

$$VH_2O = (1-0.464)*0.646/2.0954 = 0.1652$$

$$VCO = (2*0.3058*0.4+0.464*0.646)/2.0954 = 0.2597$$

$$VH_2 = (1-0.0558)*(0.464*0.646+0.03726) = 0.0045$$

$$VCH_4 = 0.5*0.0558*(0.464*0.646+0.03726)/2.0954 = 0.0045$$

$$VN_2 = 0.6/2.0954 = 0.2863$$

To check:

$$(7) VH_2O + VCO_2 + VN_2 + VCO + VH_2 + VCH_4 = 1$$

$$0.1652 + 0.1325 + 0.2863 + 0.2597 + 0.1518 + 0.0045 = 1$$

$$(15 - a) K'_{pw} = P_T * VCO * VH_2/VH_2O = 0.2391$$

$$1 * 0.2597 * 0.1518/0.1652 = 0.2387 \text{ (Check)}$$

$$(12 - a) K'_w = VCO * VH_2O/VCO_2/VH_2 = 2.1325$$

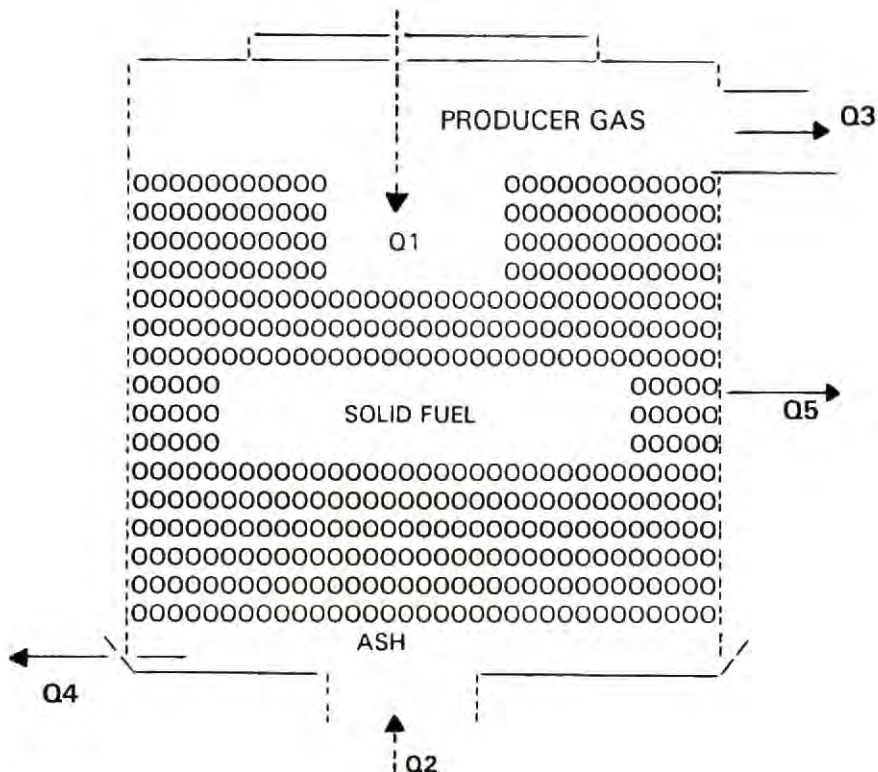
$$0.2597 * 0.1625/0.1325/0.1518 = 0.2387 \text{ (Check)}$$

$$(16 - a) K'_m = P_T * VCH_4/VH_2^2 = 0.1948$$

$$1 * 0.0045/0.158^2 = 0.1945 \text{ (Check)}$$

12. Calculation for the reaction temperature and cold gas efficiency.

- T_r = reaction temperature = 667° C (trial value)
- HHV = higher heating value of dry ipil wood = 9204.6 Btu/lb
- NHV_{dry} = net calorific value of dry ipil wood
 = $HHV - \%M / (100 - \%M) * 1040 - \%H * 9 * 1040 / 100$
 = $9204.6 - 13.5 / (100 - 13.5) * 1040 - 4.68 * 9 * 1040 / 100$
 = 8604 Btu/lb or 8604/1.8 = 4780 Kcal/Kg
- %M = %moisture content of fuel (wet basis) = 13.5%
- NHV^{wet} = net calorific value of wet ipil wood
 = $NHV_{dry} * (1 - \%M / 100) = 4780 * (1 - 13.5) = 4235$ Kcal/Kg
- H_r = enthalpy of fuel preheated to T_r (for updraft operation only) in Kcal/Kg
- H_m = enthalpy of gasifying medium
- $H_{T,w}$ = heating value of product gas
- H_w = enthalpy of product gas
- %L = heat loss in % of NHV_{dry} of gasified fuel = 8.5%
- F = amount of gasified fuel in lb/lb-mol or g/g-mol
- Q_{axd} = external heat losses by radiation and convection
- M = amount of gasifying medium in mols/mol of product gas
- H_{ac} = enthalpy of as including unburnt carbon



The heat balance equation is made with the help of the preceding diagram where:

- Q1 = heat input from the solid fuel
 Q2 = heat input from the gasification medium
 Q3 = heat content of the product gases
 Q4 = heat loss in the ash including unburnt carbon
 Q5 = external heat losses due to radiation and convection

$$Q1 + Q2 = Q3 + Q4 + Q5$$

$$(21) F * (H_f + NHV_{dry}) + M * H_m = H_g + H_{1,g} + F * H_{ac} + Q_{ext}$$

$$(22) H_f = 0.209333 * (T_f - T_o) + 0.2024 * 10^{-9} (T_f^2 - T_o^2) - 50.9333 * 10^{-9} (T_f^3 - T_o^3) \text{ cal/g fuel (assumed the same for all solid fuels),}$$

where T_f and T_o are fuel and ambient temperatures in °C respectively. (Reference 1).

$$(23) H_m = V''H_2O * LH_{H_2O} + V''H_2O * H_{h_2o} + (1 - V''H_2O) * H_{N_2-O_2 \text{ mix}}$$

$$(24) H_{N_2-O_2 \text{ mix}} = H_{N_2} * \{V''N_2 / (V''N_2 + V''O_2)\} + H_{O_2} * \{V''O_2 / (V''N_2 + V''O_2)\}$$

$$(25) H_{1w} = VH_2 * HHV_{H_2} + VCO * HHV_{CO} + VCH_4 * HHV_{CH_4}$$

$$(26) H_g = VH_2 * HH_2 + VN_2 * H_{N_2} + VH_2O * H_{H_2O} + VCO_2 * H_{CO_2} + VCH_4 * H_{CH_4} \text{ cal/g.mol}$$

$$(27) Q_{ext} = F * NHV_{dry} * (\%L/100) / (\text{sqrt}(Pr)) \text{ cal/g.mol}$$

(Reference 2).

where

Pr = total pressure in atmospheres
 H_{H_2}, H_{N_2}, H_{OC} etc. are enthalpy of gases indicated which are obtained from data tables or calculated from appropriate equations (see Table 3). LH_{H_2O} = latent heat of water vapor (from steam tables) $HHV_{CO}, HHV_{H_2}, HHV_{CH_4}$ = higher heating values of indicated gases.

(28) $H_{ax} = 0.209333 * (T_r - T_o) + 0.2024 * 10^{-3} * (T_r^2 - T_o^2) - 50.9333 * 10^{-9} * (T_r^3 - T_o^3) + 14,500 * p/1.8$ cal/g rejects (composed of ash and unburnt carbon), where T_r and T_o are rejects and ambient temperatures in °C, respectively, and p is the gram of carbon/gram of rejects. The value of p is obtained in Item 2: $p = 0.1273$

Calculated before from equations (18) and (19):

$F = 10.16$ lb. dry fuel/lb mol product gas (or g/g mol)

$M = 0.4767$ mol gasification medium/mol product gas

From appropriate data tables or equations (at $T_r = 667^\circ\text{C}$):

LH_{H_2O}	=	11,220.0 cal/g mol steam
NHV_{dry}	=	4,780.2 cal/g dry fuel
HHV_{CO}	=	68,030.5 cal/g mol gas
HHV_{H_2}	=	68,449.3 cal/g mol gas
HHV_{CH_4}	=	212,847.1 cal/g mol gas
H_{CO}	=	4,917.0 cal/g mol gas
H_{2-O_2} mix	=	223.3 cal/g mol gas
H_{H_2O}	=	29,012.1 cal/g mol gas
H_M	=	223.3 cal/g mol gasification medium
$H_{1,g}$	=	29,012.1 cal/g mol gas products
H_{CO_2}	=	7,362.4 cal/g mol gas
H_{N_2}	=	4,873.8 cal/g mol gas
H_{CH_4}	=	6,017.3 cal/g mol gas
H_{H_2}	=	4,688.5 cal/g mol gas
H_G	=	5,356.7 cal/g mol product gas
Q_{ext}	=	4,129.0 cal/g mol product gas
H_f	=	6.5 cal/g dry fuel
H_{ac}	=	1,010.3 cal/g dry fuel

Equation (21) can be written as:

$$(21 - a) H_g = F * (NHV_{dry} + H_f - H_{ac}) - Q_{ext} + M * H_M - H_{1,g}$$

Substituting into equation (21 - a), an identity is obtained:

$$H_g = 5,340.5 \text{ cal/g mol is practically identical to } 5,356.7 \text{ cal/g mol calculated from equation (26).}$$

Therefore, trial value of $T_r = 667^\circ\text{C}$ is the correct value. The choice of approach to equilibrium of the heterogenous water-gas reaction, $y = 0.26$ or 26% and percentage loss to radiation and convection, $\%L = 8.5\%$ resulted in the calculated theoretical $\text{VCO}_2 = 15.9\%$ being identical to the experimentally measured (by Orsat apparatus) $\%\text{CO}_2 = 15.9\%$. Otherwise, other values of y , and $\%L$ (if this is not known) would have to be tried.

The Cold Gas Thermal Efficiency is defined as the ration of the heat content of the cold product gas to the heat inputs from the fuel and the gasification medium. Thus,

$$\begin{aligned}\text{Output} &= H_{1,g} = 29,012.1 \text{ cal/g mol product gas} \\ \text{Inputs} &= F * \text{NHV}_{\text{dry}} + M * H_m = 10.16 * 4780.2 \\ &\quad 0.4767 * 223.3 = 48,673.3 \text{ cal/g mol} \\ &\quad \text{product gas}\end{aligned}$$

$$\text{Cold Gas Thermal Efficiency} = 29,012.1/48,673.3 = 0.596 \text{ or } 59.6\%$$

References

1. Technical Data on Fuels (6th Edition). 1961. Spiers, H.M. (Ed.). The British National Committee World Power Conference.
2. Gumz, W. Gas Producers and Blast Furnaces John Wiley & Sons.
3. V.M. Faires. 1957. Thermodynamics (4th Edition)

Table 2

EQUATIONS FOR EQUILIBRIUM CONSTANTS AT TEMPERATURE T $\text{LOG K} = A_0 + A_1/T + A_2 * T + A_3 * T^2 + A_4 * \text{LOG T}$ (T IN $^\circ\text{K}$)

Constants

K	A_0	A_1	A_2	A_3	A_4
K_{FS}	3.26730	-8820.690	-1.208714 E-3	1.53734 E-7	2.295483
K_{PM}	-33.45778	-4825.986	-5.671122 E-3	8.255484 E-7	14.515760
K_w	36.72508	-3994.704	4.462408 E-3	-6.71814 E-7	-12.220277
K_{pm}	-13.06361	-4682.80	-2.09594 E-3	3.8620 E-7	3.034338

