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OPTIMIZATION STUDIES OF VARIOUS COAL-CONVERSION SYSTEMS. QUARTERLY REPORT, JANUARY--MARCH 1978

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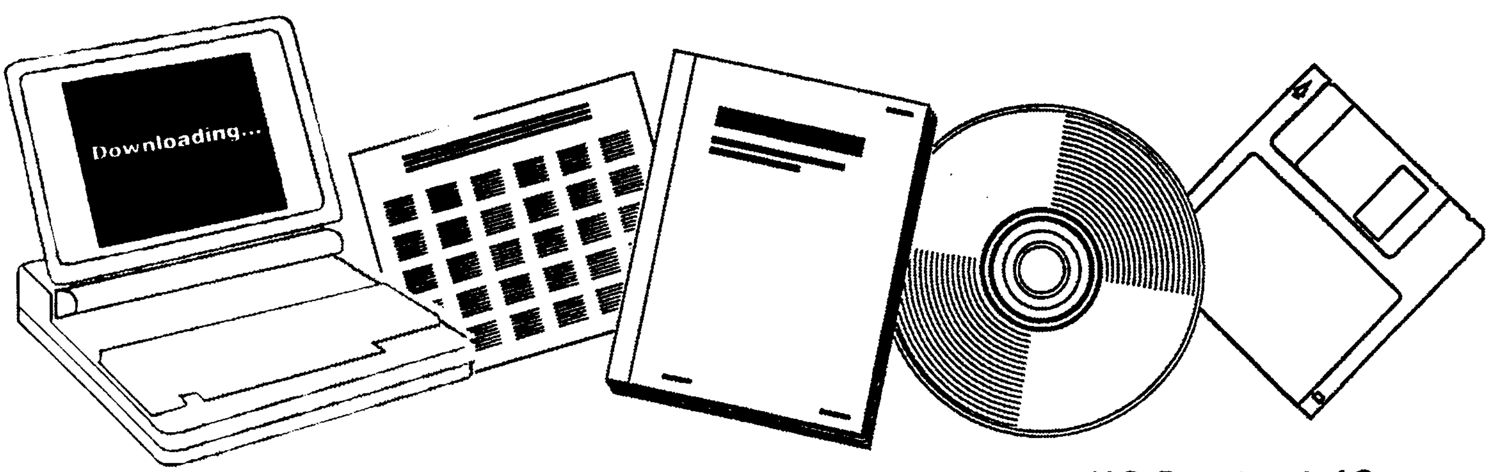

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OPTIMIZATION STUDIES OF VARIOUS COAL-CONVERSION SYSTEMS

Quarterly Report for the
Period January-March 1978

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ABSTRACT

A mathematical model has been developed to simulate the Texaco Downflow Entrained-Bed Pilot Plant Gasifier using coal liquefaction residues as feedstocks. The entrained-bed gasifier was conceptually divided into three zones, i.e., the Pyrolysis and Volatile Combustion Zone, the Gasification and Combustion Zone, and the Gasification Zone. The gas phase was assumed to be completely mixed at the entrance region followed by a region approximating plug flow. The solid phase was assumed to be a plug flow throughout the reactor. Temperature and concentration profiles along the reactor were obtained by solving the material and energy blances and taking into consideration the gasification kinetics, the transport rates and the hydrodynamics of the gasifier. The results of computation from the proposed model were compared with the experimental data.

OBJECTIVE AND SCOPE OF WORK

Optimization Studies of Various Coal-Conversion Systems

The objective of this project is to establish a generalized standard method utilizing mathematical optimization techniques for comparing and evaluating new and existing coal conversion processes and establish the process or processes presenting the most commercial attractiveness. These studies concentrate on the production of high-Btu gas, low-Btu gas, electricity, and coal oil by conversion of caking and noncaking coals containing sulfur.

Major effects will be centered on simulating and optimizing coal conversion processes by making mathematical models of unit operations found in various coal conversion processes. System models produced by this program may accomplish the following goals:

1. Identify costly and thermally inefficient steps in conversion processes illustrating areas where technological development is necessary.

2. Predict the effects of varying operating parameters on the performance of conversion plants.

3. Provide a quick assessment of new conversion processes and predict those processes having the greatest chance of economic success.

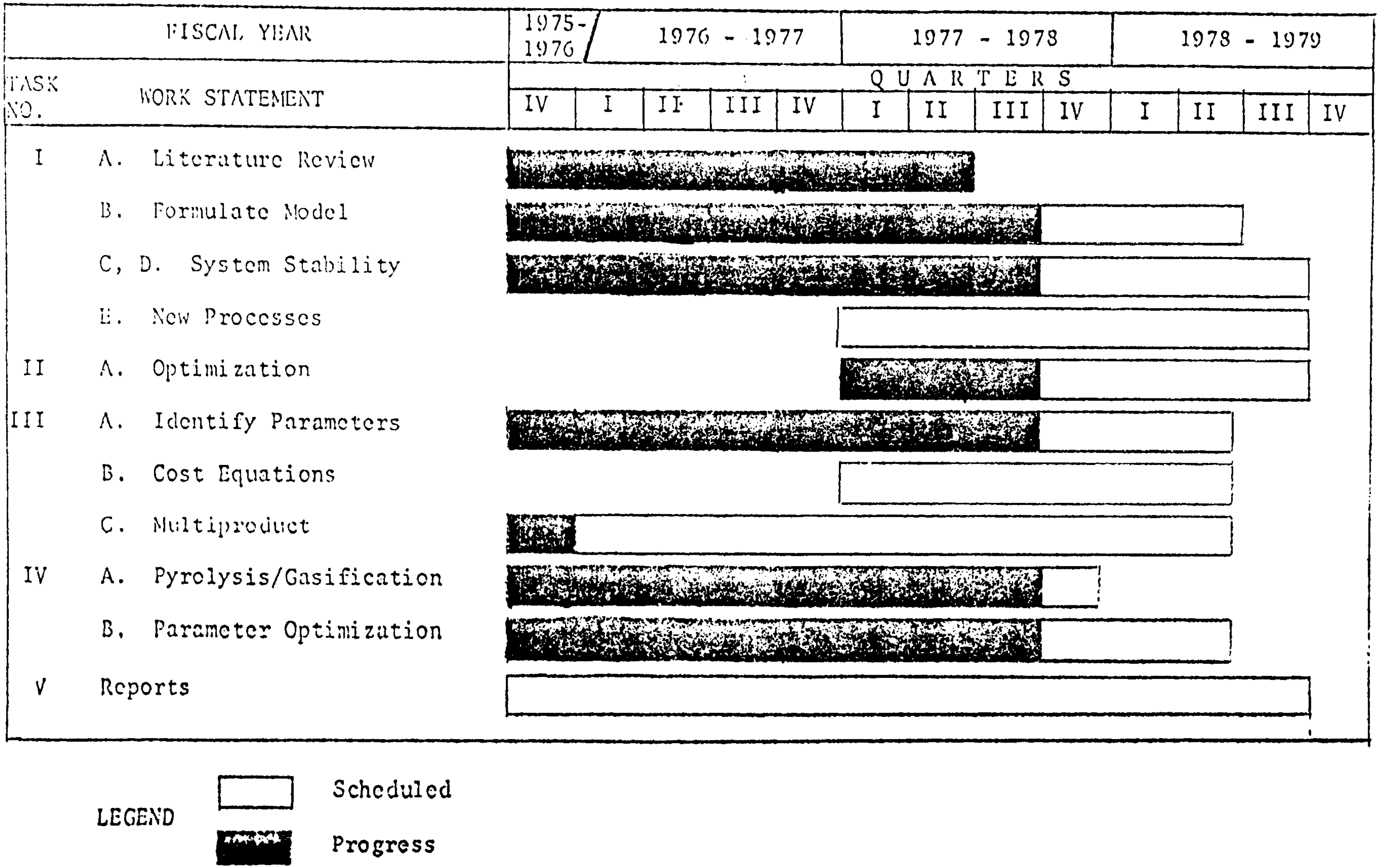
4. Indicate the qualitative and quantitative data to be obtained from experimental process development units.

5. Provide a means for ranking alternative conversion technologies in order of economic promise.

Specifically, these models would provide the basis for examining new or existing coal conversion processes, identify areas of variability, and evaluate the economic feasibility of the processes prior to the costly and time-consuming production of a pilot plant. In addition, optimization and economic studies will be used to identify the most attractive alternatives for producing several energy products such as electricity, synthetic gas, crude oil and methanol in a single coal conversion complex.

Project Plan and Progress Report

◇ Project Start As of Date ____________



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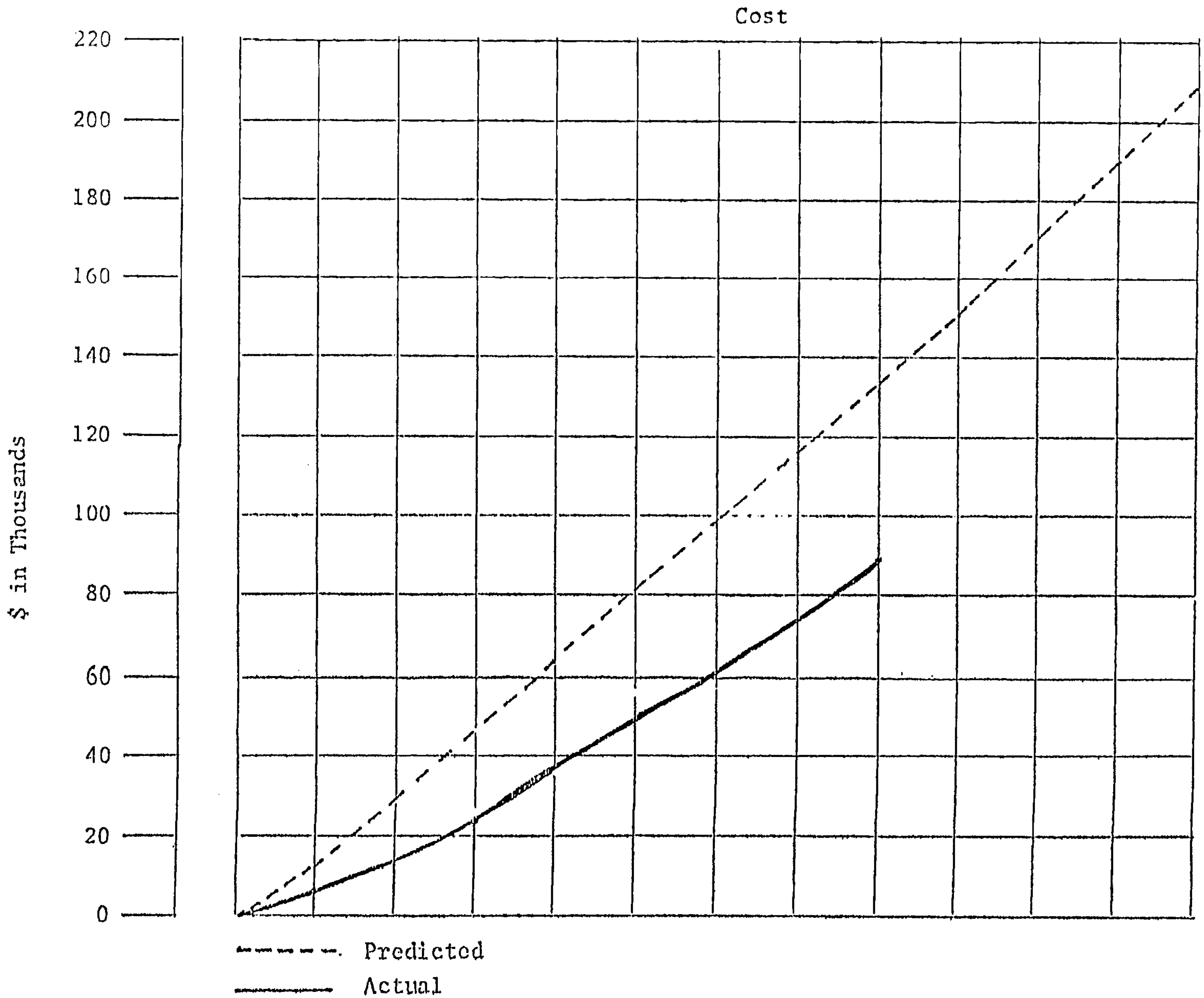

TECHNICAL PROGRESS

Introduction

The primary purpose of this work is to develop a mathematical model for a non-recycling type entrained bed gasification reactor. This model provides the temperature and the concentration profiles for the solids and gas along the bed. Sensitivity tests have been conducted to assess the importance of model parameters on the performance of the gasifier and to gain insight into the scale-up variables.

The current work presents a model to simulate the Texaco Downflow Entrained Bed Pilot Plant Gasifier which uses coal liquefaction residues as feedstock. Detail description of this pilot plant and experimental results has been reported by Texaco's Montebello Research Laboratory [1,2]. Texaco has demonstrated the ability of their downflow entrained bed reactor to gasify light oils, asphalts and coal-water slurries into synthesis gas (H_2+CO) in addition to their interest in testing coal liquefaction residues as the feedstock. The high carbon conversion (about 91-99%) and the fact that a wide range of fossil fuels can be used as feedstocks suggest that the entrained bed gasification is a potentially viable technology for production of high or low BTU gas. In addition, most of the coal liquefaction processes now being developed require hydrogen or synthesis (a mixture of hydrogen and carbon monoxide) for coal liquefaction. There is a need for the development of a mathematical model of a downflow entrained bed gasifier for simulation and scale-up purpose.

Entrained Bed Modeling

The entrained bed gasifier considered is shown schematically in Fig. 1. The reactor is conceptually divided into three zones, i.e., Pyrolysis and Volatile Combustion Zone, Gasification and Combustion Zone, and Gasification Zone. Extremely high temperature is expected in the first two zones because of gas phase combustion. Oxygen is consumed in the first two zones to produce enough heat for the endothermic reactions in the gasification zone. The degree of mixing is high near the inlet nozzle but quickly reaches the state of plug flow. However, no experimental data are available for estimation of the degree of mixing along the reactor. Therefore, it is assumed that gas phase is completely mixed in the first zone while plug-flow in the remaining zones. The solid phase is assumed to be plug-flow throughout the reactor. The solid velocity is calculated based on Stoke's law because of the small particle size and the low slip Renold's number employed. Heat and material balance equations are formulated based on the above assumptions and shown below:

(1) Solid Phase

$$\frac{dT_s}{dt} = \frac{6}{\rho_s C_{p_s} d_p} \left[eF\sigma(T_g^4 - T_s^4) + h_c(T_g - T_s) + \sum_{j(\text{solid})} (-\Delta H_j) r_j \right] \quad (1)$$

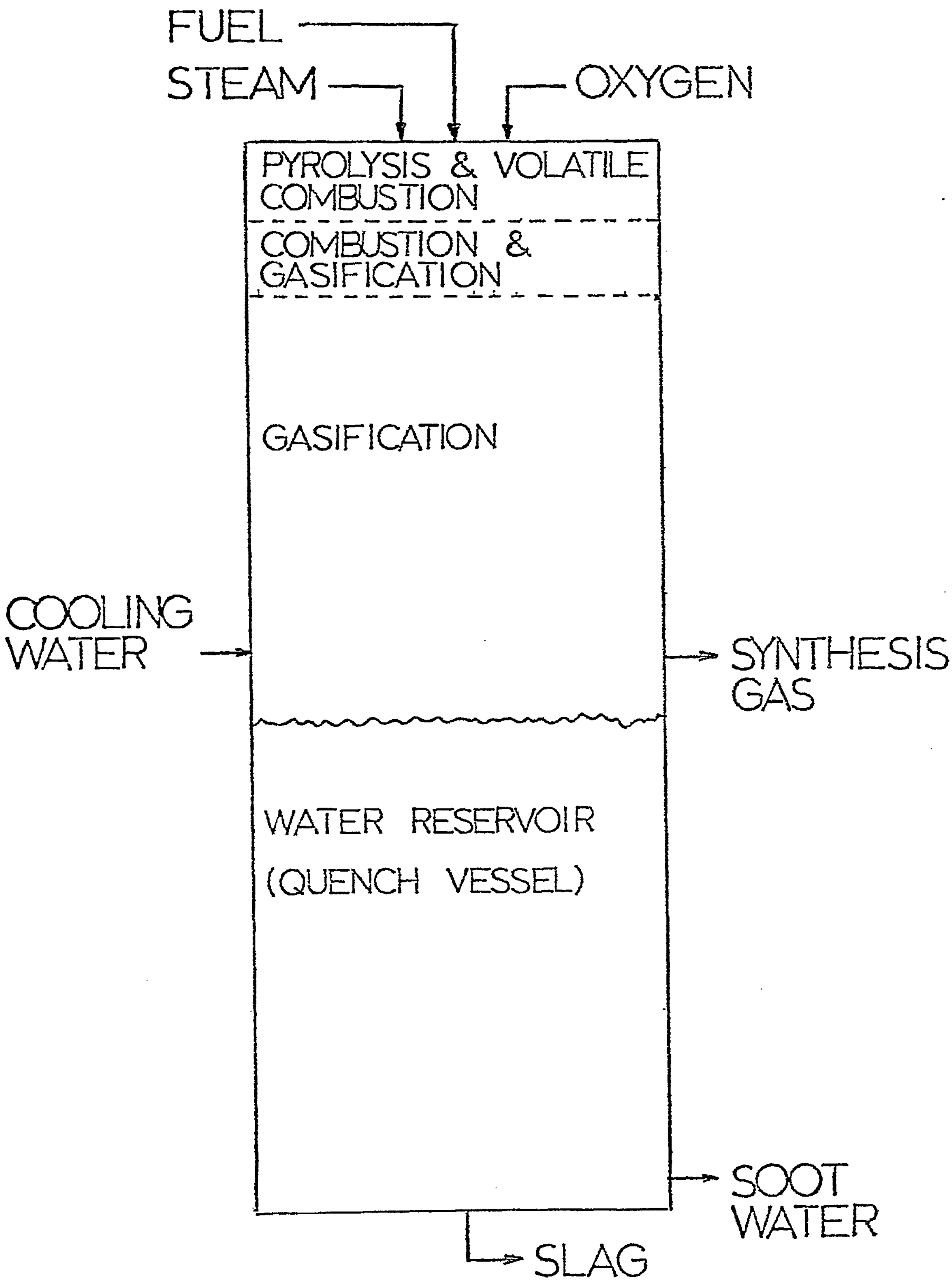

FIGURE 1. TEXACO DOWNFLOW ENTRAINED BED PILOT PLANT GASIFIER

$$\frac{dW_s}{dz} = a\ A_t \sum_{j(\text{solid})} r_j \tag{2}$$

(2) Gas Phase

a) Completely-mixed zone:

$$\Sigma(W_{g_i}\ C_{p_{gi}}\ T_g)_{Z_1} - \Sigma(W_{g_i}\ C_{p_{gi}}\ T_g)_o = (A_t\ a\ Z_1)[eF\sigma(T_g^{\ 4} - T_s^{\ 4}) + h_c\ (T_g - T_s)] + \sum_{k,gas} (-\Delta H_k) r_k\ A_t\ Z_1 - H_{loss,g\text{-}w}\ Z_1 \tag{3}$$

$$W_{g_{i,Z_1}} - W_{g_{i,o}} = A_t\ Z_1 \sum_{k,gas} \nu_{ik}\ r_k \tag{4}$$

b) Plug-flow zone:

$$\frac{d(\sum_i W_{g_i}\ C_{p_{gi}}\ T_g)}{dz} = -\ a\ A_t\ [eF\sigma(T_g^{\ 4} - T_s^{\ 4}) + h_c\ (T_g - T_s) + \sum_{k,gas} (-\Delta H_k)\ r_k\ A_t - H_{loss,g\text{-}w} \tag{5}$$

$$\frac{dW_{g_i}}{dz} = A_t \sum_{k,gas} \nu_{ik}\ r_k \tag{6}$$

where the reaction species a, j and k in each zone are defined in **Appendix 1.**

The solid velocity is calculated by the following formula:

a) Down-flow:

$$v_s = v_{si}\ e^{-b\Delta t} + (v_g + v_t)(1 - e^{-b\Delta t})$$

b) Up-flow

$$v_s = v_{si}\ e^{-b\Delta t} + (v_g - v_t)(1 - e^{-b\Delta t})$$

where

$$b = \frac{18\ \mu}{\rho_s\ d_p^2}$$

$$v_t = \frac{(\rho_s - \rho_g) d_p^2}{18 \mu}$$

v_{si} = initial solid velocity, (cm/sec)

Δt = residence time, (sec)

The rates of reaction of combustion and gasification are shown in Appendix 2. In an entrained bed reactor, the ash layer formed is assumed to remain on the fuel particle during reactions. Since the particle loading in an entrained gasifier is small (less than 1%), particle collisions are likely to be infrequent. The solid-gas reactions are surface reactions and the rates are mostly controlled by diffusion resistances because the temperature in the reactor is high (> 1300°K). It is thus reasonable to assume that the reaction rates may be estimated by the Unreacted-Core Shrinking Model [3] for an entrained-bed gasifier.

Temperature and concentration profiles are obtained by solving the above heat and mass balance equations. A typical result is shown in Appendix 5. Comparison of results between those calculated from proposed model and those obtained from Texaco Pilot Plant Gasifier is shown in Appendix 6. A typical ultimate analysis of the feedstock used and the operating conditions are shown respectively in Appendix 3 and 4.

Discussion

The proposed model is designed to simulate non-recycling type entrained-bed gasifiers. Testing of this model was first evaluated by comparing simulating the experimental results from Texaco Downflow Entrained-Bed Pilot Plant Gasifier using coal liquefaction residues as feedstocks. Good agreement has been obtained between the computation results from the proposed model and the pilot plant results for 26 runs. This model also simulates the reactor temperature and concentration profiles for both the solid and the gas phase. Improvement of this model will be made next quarter by examining other experimental results from entrained-bed pilot plant gasifiers with different coal feed. A comprehensive report on coal dissolution is in progress.

SUMMARY OF PROGRESS

Development of the entrained bed gasification modeling was first focused on the characteristics of Texaco Downflow Entrained Bed Pilot Plant Gasifier using coal liquefaction residues as feedstocks. The mathematical model was set up basically by assuming the gas phase to be completely mixed at the entrance region followed by a region approximating plug flow and the solid phase to be plug flow throughout the reactor. Temperature and concentration profiles for both the solid and the gas phase along the reactor were obtained by solving the material and heat balances taking into consideration the reaction kinetics, the transport rates and the hydrodynamics of the gasifier. Testing of the proposed model was first evaluated by comparing the computation results with the experimental results and good agreement was achieved

Conclusion

Texaco has demonstrated the ability of gasifying coal liquefaction residues, light oils and coal-water slurries into synthesis gas in a downflow entrained-bed gasifier. This process will provide the required hydrogen or synthesis gas for coal liquefaction processes or for fuels and chemical feedstock. The proposed model can provide an insight into the importance of the operating parameters on the reactor performance and a procedure for gasifier scale-up. The computational results from the proposed model were compared with the experimental results and good agreement was achieved.

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Appendix 1

Reaction species in each zone:

The definition of reaction number for j and k used in Eq. (1)-Eq. (6) is given below:

(I) Reactions in the solid phase:

j	name of reaction
1	pyrolysis
2	char-oxygen reaction
3	char-steam reaction
4	char-carbon dioxide
5	char-hydrogen reaction
6	water-gas-shift reaction (catalyzed by the mineral materials in char)
7	methane-steam reforming reaction (catalyzed by the mineral materials in char)

(II) Reactions in the gas phase

k	name of reaction
1	$H_2 + 1/2 \cdot O_2 \rightarrow H_2O$
2	$CO + 1/2 \cdot O_2 \rightarrow CO_2$
3	$CH_4 + 2\ O_2 \rightarrow CO_2 + 2\ H_2O$
4	$C_6H_6 + 15/2 \cdot O_2 \rightarrow 6\ CO_2 + 3\ H_2O$
5	$CO + H_2O \rightleftarrows CO_2 + H_2$
6	$CH_4 + H_2O \rightleftarrows CO + 3\ H_2$

* The first four k's reactions are assumed to be simultaneous and, therefore, the reaction rates are controlled by the formation of H_2, CO, CH_4 and C_6H_6 in solid-involved reactions.

Specific reactions in the three conceptual zones:

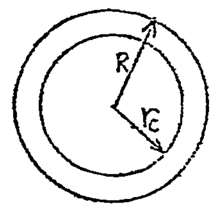
Name of the reaction zone	j	k
Pyrolysis and Volatile Combustion Zone	1	1,2,3,4
Combustion and Gasification Zone	2,3,4	1,2,3
Gasification Zone	3,4,5,6,7	5,6

Appendix 2

Rate Expressions

(I) Surface Reaction Type: Unreacted-Core Shrinking Model [3]

$$\text{Rate} = \frac{1}{\dfrac{1}{k_{diff}} + \dfrac{1}{k_S Y^2} + \dfrac{1}{k_{dash}}\left(\dfrac{1}{Y} - 1\right)} (P_i - P_i^*) \qquad g/cm^2 atm\ sec$$



where

$$Y = \frac{r_c}{R} = \left(\frac{1 - x}{1 - f}\right)^{1/3}$$

f = conversion when pyrolysis is finished, based on original d.m.m.f coal

x = conversion at any time after pyrolysis is completed, based on original d.m.m.f coal

k_{diff} = gas film diffusion constant, $g/cm^2 atm\ sec$

k_S = ash film diffusion constant, $g/cm^2 atm\ sec$

k_{dash} = ash film diffusion constant, $g/cm^2 atm\ sec$

$\doteq k_{diff}\ (\varepsilon^{2.5})$

ε = voidage in the ash layer

P_i = partial pressure of i-component gas

$P_i - P_i^*$ = effective partial pressure of i-component taking account of the reverse reaction effect

(i) Char-O_2 Reaction, $[C + \frac{1}{\phi} O_2 \rightarrow 2\ (1 - \frac{1}{\phi}) CO + (\frac{2}{\phi} - 1) CO_2]$ [4]

$k_S = 8710 \cdot \exp(-17967/T_S)$, T_S in °K

$$k_{diff} = 0.292\ \phi\ \left(\frac{4.26}{T_g}\right)\left(\frac{T_g}{1800}\right)^{1.75} / (P_t\ d_p)$$

ϕ = the mechanism factor based on the stoichiometric relation of CO and CO_2, ϕ can be roughly estimated by the following equations:

$\phi = (2Z + 2)/(Z + 2)$ for $d_p \leq 0.005$ cm

$$\phi = [(2Z + 2) - Z(d_p - 0.005)/0.095]/(Z + 2)$$

$$\text{for } 0.005 \text{ cm} \leq d_p \leq 0.1 \text{ cm}$$

and

$$\phi = 1.0 \text{ for } d_p > 0.1 \text{ cm}$$

where

$$Z = [CO]/[CO_2] = 2500 \exp(-6249/T)$$

d_p in cm and $T = (T_S + T_g)/2$ in °K

$$P_i - P_i^* = P_{O_2}$$

(ii) Char-Steam Reaction, (for temperature greater than 1100°C) [5]

$$k_S = 247 \exp(-21060/T_S)$$

$$k_{diff} = 10 \times 10^{-4} \left(\frac{T}{2000}\right)^{0.75} / (P_t\ d_p)$$

$$k_{eq} = \exp[17.644 - 30260/(1.8 \cdot T_S)]$$

$$P_i - P_i^* = P_{H_2O} - \frac{P_{H_2} P_{CO}}{k_{eq}}$$

(iii) Char-CO_2 Reaction, (for temperature greater than 1100°C) [5]

$$k_S = 247 \exp(-21060/T_S)$$

$$k_{diff} = 7.45 \times 10^{-4} \left(\frac{T}{2000}\right)^{0.75} / (P_t\ d_p)$$

$$P_i - P_i^* = P_{CO_2}$$

(iv) Char-Hydrogen Reaction

This reaction is still in chemical reaction regime even in high temperature as 1600°K, because it has low intrinsic reaction rate but high diffusion characteristics. For the simplicity of calculation, the same expression as that of the Unreacted-Core Shrinking Model is used by changing K_v into K_s according to experimental conditions as follows:

$$k_S = k_v \left(\frac{1-x}{1-x_p}\right)^{1/3} \left(\frac{\rho_s d_p}{6}\right)$$

$$k_S = 0.12 \exp(-17921/T_S) \quad g/cm^2 atm\ sec \qquad [6]$$

$$k_{diff} = 1.33 \times 10^{-3} \left(\frac{T}{2000}\right)^{0.75} / (d_p\ P_t)$$

$$k_{eq} = \frac{0.175}{34713} \exp\ [18400/(1.8\ T_S)]$$

$$P_i - P_i^* = P_{H_2} - \sqrt{P_{CH_4}/K_{eq}}$$

(II) Catalytic Reactions

(i) Water-Gas-Shift Reaction] 7]

$$\text{Rate} = F_W (2.77 \times 10^5)(x_{CO} - x_{CO}^*) \exp\left(-\frac{27760}{1.987\ T}\right) P_t^{(0.5-P_t/250)}$$

$$\exp\left(-8.91 + \frac{5553}{T}\right) \qquad g\ mole/[sec(g\ ash)]$$

The adjustable parameter, F_W, which represents the relative catalytic reactivity of ash to that of iron-base catalyst, is selected to be 0.2 in the model develop

$$x_{CO} = P_{CO}/P_t$$

$$x_{CO}^* = \frac{1}{P_t} \left[\frac{P_{CO_2}\ P_{H_2}}{k_{eq}\ P_{H_2O}}\right]$$

$$k_{eq} = \exp\ (-3.6893 + 7234/(1.8\ T)]$$

P_t is the total pressure

(ii) Methane-Steam Reforming Reaction [8]

$$\text{Rate} = 312 \exp\ [-30{,}000/(1.987\ T)] \qquad 1/sec$$

Appendix 3: Typical ultimate analysis for the feedstock used in Texaco's pilot plant tests. [1,2]

Feedstock Source and Run Number	Dry Fuel Analysis (wt. %)						
	C	H	N	S	O	Ash	Cl
H-Coal residue from Illinois No. 6 coal for Run I-1 [1]	74.05	6.25	0.71	1.77	1.32	15.53	0.37
H-coal residue from Illinois No. 6 coal for Run I-2 [1]	73.04	5.82	0.73	1.37	1.70	16.83	0.48
H-coal residue from Wyodak coal for Run W-1 [1]	78.37	5.79	0.92	0.07	3.70	11.05	0.08
SRC II Vacuum Flash Drum Bottoms [2]	64.90	3.65	1.25	2.96	1.70	25.54	-
Exxon DSP Vacuum Tower Bottoms [2]	70.74	4.67	1.18	2.74	3.95	16.72	-

Appendix 4: Operating conditions for five typical runs of Texaco's pilot plant tests.

Feedstock Source (Run Number)	Feed Rate			Feed Temperature (°K)		
	Fuel Rate (g/sec)	$\frac{H_2O}{Fuel}$	$\frac{O_2}{Fuel}$	Fuel	Steam	Oxygen
H-coal residue from Illinois No. 6 coal (Run I-1) [1]	76.66	0.241	0.866	505.22	696.67	298
H-coal residue from Illinois No. 6 coal (Run I-2) [1]	81.18	0.314	0.7682	496.33	676.33	298
H-coal residue from Illinois No. 6 coal (Run W-1) [1]	86.0	0.318	0.90	513.55	692.44	298
SRC II Vacuum Flash Drum Bottoms [2]	126.11	0.30	0.77	505.22	696.67	298
Exxon DSP Vacuum Tower Bottoms [2]	126.11	0.50	0.79	505.22	696.67	298

Appendix 5: Typical Temperature and Concentration Profiles in the Texaco Downflow Entrained-Bed Gasifier

Comparison of the computational results from the proposed model and the experimental results from the Texaco Pilot Plant Entrained-Bed Gasifier has been carried on for twenty-six runs as shown in Appendix 6. Temperature and concentration profiles calculated from the model for five typical runs are shown in the following figures. The fuel analysis and operating conditions for these typical runs are shown respectively in Appendix 3 and 4.

Appendix 6 Comparison of computational results from the model with experimental results from Texaco Entrained Bed Pilot Plant Gasifier.

(A) Using H-Coal residues from Illinois No. 6 coal as feedstock: [1]

Run No.	Input Condition: Fuel Rate (g/sec)	Input Condition: O_2/Fuel	Input Condition: Steam/Fuel	Dry Product gas / Source	CO Flow Rate g/sec (Vol. %)	H_2 Flow Rate g/sec (Vol. %)	CO_2 Flow Rate g/sec (Vol. %)	CH_4 Flow Rate g/sec (Vol. %)	H_2S Flow Rate g/sec (Vol. %)	N_2 Flow Rate g/sec (Vol. %)	Carbon Conversion %
I-1	76.66	0.866	0.241	Exp.	123.77 (57.57)	6.01 (39.13)	9.985 (2.95)	0.15 (0.12)	0.133 (0.06)	0.53 (0.12)	98.64
				Model	123.94 (56.60)	6.23 (39.84)	10.04 (2.92)	0.20 (0.16)	0.726 (0.27)	0.454 (0.208)	98.88
I-2	81.18	0.768	0.314	Exp.	112.52 (53.06)	6.211 (41.00)	17.2 (5.15)	0.56 (0.46)	0.59 (0.21)	0.1513 (0.07)	90.66
				Model	119.78 (54.11)	6.54 (41.39)	13.54 (3.89)	0.242 (0.193)	0.57 (0.212)	0.451 (0.204)	93.29
I-3	82.202	0.813	0.309	Exp.	121.5 (54.66)	6.24 (39.31)	19.26 (5.51)	0.114 (0.08)	0.67 (0.26)	0.133 (0.11)	98.28
				Model	126.4 (55.89)	6.36 (39.39)	14.56 (4.10)	0.17 (0.131)	0.77 (0.28)	0.496 (0.22)	99.75
I-4A	79.456	0.807	0.323	Exp.	116.0 (55.70)	5.78 (38.90)	16.74 (5.11)	0.15 (0.12)	0.32 (0.11)	0.0124 (0.00)	97.24
				Model	119.96 (55.18)	6.16 (39.69)	15.73 (4.60)	0.148 (0.12)	0.526 (0.20)	0.437 (0.201)	99.76
I-4B	81.846	0.797	0.310	Exp.	119.54 (54.90)	6.107 (39.26)	18.14 (5.30)	0.185 (0.14)	0.70 (0.23)	0.24 (0.11)	97.34
				Model	125.09 (55.71)	6.36 (39.68)	14.37 (4.07)	0.170 (0.132)	0.538 (0.197)	0.449 (0.20)	99.75
I-5A	71.64	0.8263	0.352	Exp.	103.32 (54.02)	5.30 (38.78)	19.972 (6.64)	0.0852 (0.07)	0.57 (0.21)	0.428 (0.22)	98.875
				Model	106.37 (54.89)	5.39 (38.98)	16.824 (5.525)	0.112 (0.101)	0.68 (0.29)	0.421 (0.217)	99.77
I-5B	65.0	0.817	0.392	Exp.	92.3 (52.48)	5.007 (39.85)	19.987 (7.22)	0.086 (0.08)	0.73 (0.31)	0.0034 (0.00)	98.868
				Model	95.45 (53.56)	5.08 (39.92)	16.58 (5.92)	0.100 (0.098)	0.62 (0.29)	0.386 (0.217)	99.75
I-5C	56.264	0.832	0.429	Exp.	81.853 (51.39)	4.53 (39.83)	20.61 (8.23)	0.0434 (0.04)	0.8035 (0.41)	0.0734 (0.04)	98.885
				Model	81.186 (52.75)	4.38 (39.84)	16.461 (6.81)	0.074 (0.084)	0.546 (0.29)	0.335 (0.218)	99.76

Run No.	Input Condition: Fuel Rate (g/sec)	O_2/Fuel	Steam/Fuel	Dry Product gas Source	CO Flow Rate g/sec (Vol. %)	H_2 Flow Rate g/sec (Vol. %)	CO_2 Flow Rate g/sec (Vol. %)	CH_4 Flow Rate g/sec (Vol. %)	H_2S Flow Rate g/sec (Vol. %)	N_2 Flow Rate g/sec (Vol. %)	Carbon Conversion %
I-6	87.73	0.774	0.291	Exp.	125.2 (55.03)	6.41 (39.43)	17.93 (5.01)	0.27 (0.20)	0.3 (0.10)	0.391 (0.17)	97.122
				Model	129.34 (55.85)	6.54 (39.52)	14.35 (3.94)	0.187 (0.142)	0.89 (0.32)	0.53 (0.23)	97.898
I-7A	90.974	0.7757	0.282	Exp.	130.3 (55.33)	6.67 (39.62)	17.1 (4.62)	0.27 (0.20)	0.292 (0.08)	0.213 (0.09)	96.826
				Model	133.58 (55.63)	6.85 (39.96)	14.20 (3.764)	0.202 (0.147)	0.798 (0.274)	0.555 (0.231)	97.735
I-7B	95.392	0.782	0.267	Exp.	139.75 (55.87)	6.975 (39.03)	17.3 (4.40)	0.197 (0.13)	0.866 (0.26)	0.640 (0.25)	96.992
				Model	141.11 (56.223)	7.08 (39.50)	14.29 (3.622)	0.22 (0.153)	0.829 (0.272)	0.583 (0.232)	96.735
I-8A	92.13	0.797	0.247	Exp.	140.7 (57.38)	6.732 (38.43)	14.27 (3.70)	0.13 (0.09)	0.91 (0.30)	0.102 (0.04)	98.641
				Model	140.09 (57.39)	6.735 (38.63)	12.77 (3.54)	0.213 (0.15)	0.797 (0.27)	0.562 (0.230)	97.744
I-8B	95.07	0.8016	0.239	Exp.	145.8 (57.67)	6.913 (38.28)	14.4 (3.62)	0.058 (0.03)	0.80 (0.25)	0.204 (0.08)	98.663
				Model	143.8 (57.56)	7.853 (38.37)	13.43 (3.42)	0.22 (0.152)	0.82 (0.27)	0.576 (0.23)	97.233
I-8C	92.86	0.800	0.246	Exp.	141.073 (57.02)	6.785 (38.39)	15.28 (3.93)	0.071 (0.04)	1.18 (0.37)	0.46 (0.18)	98.605
				Model	141.709 (57.46)	6.43 (38.58)	12.81 (3.31)	0.213 (0.15)	0.81 (0.27)	0.57 (0.23)	98.229
I-9	87.79	0.787	0.268	Exp.	125.4 (57.49)	5.966 (38.29)	12.89 (3.76)	0.096 (0.07)	0.80 (0.26)	0.148 (0.06)	97.45
				Model	131.20 (56.70)	6.43 (38.885)	13.88 (3.82)	0.189 (0.143)	0.65 (0.23)	0.529 (0.23)	97.60
I-10	129.77	0.8346	0.276	Exp.	201.16 (55.18)	10.22 (39.24)	26.92 (4.69)	0.15 (0.26)	0.58 (0.10)	1.73 (0.47)	99.158
				Model	195.918 (55.42)	9.99 (39.56)	24.76 (4.46)	0.284 (0.140)	0.94 (0.22)	0.672 (0.190)	96.01
I-11	132.79	0.8484	0.279	Exp.	204.16 (54.79)	10.57 (39.72)	29.96 (5.11)	0.193 (0.09)	0.68 (0.12)	0.42 (0.11)	99.187
				Model	199.72 (54.95)	10.30 (39.69)	27.50 (4.82)	0.278 (0.134)	0.977 (0.22)	0.697 (0.19)	96.57

(B) Using H-coal residues from Wyodak coal as feedstock: [1]

Run No.	Input Condition: Fuel Rate (g/sec)	Input Condition: O_2/Fuel	Input Condition: Steam/Fuel	Dry Product gas / Source	CO Flow Rate g/sec (Vol. %)	H_2 Flow Rate g/sec (Vol. %)	CO_2 Flow Rate g/sec (Vol. %)	CH_4 Flow Rate g/sec (Vol. %)	H_2S Flow Rate g/sec (Vol. %)	N_2 Flow Rate g/sec (Vol. %)	Carbon Conversion %
W-1	86.0	0.90	0.318	Exp.	143.2 (56.96)	6.8325 (38.04)	19.15 (4.84)	0.0517 (0.03)	0.00 (0.00)	0.19 (0.07)	98.98
				Model	144.1 (56.85)	6.88 (38.00)	17.98 (4.51)	0.58 (0.401)	0.032 (0.010)	0.57 (0.224)	99.75
W-2	87.231	0.859	0.286	Exp.	146.25 (57.67)	6.865 (37.90)	17.1 (4.29)	0.0898 (0.06)	0.0325 (0.00)	0.078 (0.03)	99.17
				Model	148.44 (58.65)	6.741 (37.30)	13.40 (3.37)	0.656 (0.453)	0.028 (0.009)	0.547 (0.216)	99.65
W-3A	128.49	0.844	0.253	Exp.	212.71 (57.80)	9.85 37.47)	25.66 (4.43)	0.22 (0.10)	0.00 (0.00)	0.553 (0.15)	99.393
				Model	212.77 (59.20)	9.45 (36.81)	19.51 (3.454)	0.558 (0.272)	0.088 (0.020)	0.91 (0.251)	97.90
W-3B	128.67	0.86	0.264	Exp.	208.2 (56.02)	10.05 (37.02)	34.016 (5.82)	0.129 (0.06)	0.103 (0.01)	0.68 (0.18)	99.395
				Model	214.8 (58.97)	10.85 (36.86)	21.38 (3.75)	0.303 (0.146)	0.089 (0.020)	0.96 (0.262)	98.85
W-4	125.66	0.857	0.273	Exp.	211.44 (57.61)	10.06 (37.77)	23.77 (4.12)	0.28 (0.13)	0.00 (0.00)	1.20 (0.32)	98.765
				Model	210.58 (58.46)	9.56 (37.16)	20.22 (3.57)	4.215 (0.59)	0.0066 (0.002)	0.769 (0.213	98.215
W-5	129.826	0.854	0.263	Exp.	215.34 (57.00)	10.343 (38.32)	25.45 (4.28)	0.34 (0.15)	0.032 (0.00)	0.734 (0.19)	98.211
				Model	214.78 (57.95)	10.08 (37.56)	21.09 (3.62)	1.32 (0.625)	0.0615 (0.014)	0.837 (0.227)	97.66
W-6	132.82	0.855	0.318	Exp.	215.07 (54.91)	10.97 (39.20)	33.84 (5.49)	0.41 (0.18)	0.00 (0.00)	0.706 (0.18)	97.900
				Model	217.46 (56.17)	10.60 (38.34)	27.65 (4.55)	1.60 (0.762)	0.042 (0.009)	0.837 (0.236)	98.283
W-7	135.64	0.94	0.310	Exp.	220.02 (54.24)	11.00 (37.98)	48.576 (7.62)	0.00 (0.00)	0.00 (0.00)	0.477 (0.11)	99.676
				Model	228.74 (57.23)	10.56 (37.01)	34.12 (5.43)	0.234 (0.105)	0.0072 (0.001)	0.917 (0.229)	99.628

(C) Using SRC II Vaccum Flash Drum Bottoms as feedstock: [2]

Input Condition: Fuel Rate (g/sec)	$\frac{O_2}{Fuel}$	$\frac{Steam}{Fuel}$	Wet Product Gas / Source	CO Flow Rate g/sec (Vol. %)	H_2 Flow Rate g/sec (Vol. %)	CO_2 Flow Rate g/sec (Vol. %)	H_2O Flow Rate g/sec (Vol. %)	CH_4 Flow Rate g/sec (Vol. %)	H_2S Flow Rate g/sec (Vol. %)	N_2 Flow Rate g/sec (Vol. %)	Carbon Conversion %
126.11	0.77	0.30	Exp.	167.584 (53.5)	6.936 (31.0)	32.98 (6.7)	14.50 (7.2)	0.00 (0.00)	4.1 (1.04)	1.566 (0.50)	99
			Model	171.48 (54.01)	7.178 (31.65)	29.54 (5.9)	15.19 (7.44)	0.151 (0.083)	2.016 (0.52)	1.146 (0.361)	99.77

(D) Using Exxon DSP Vaccum Tower Bottoms as feedstock: [2]

126.11	0.79	0.50	Exp.	176.98 (45.6)	9.37 (33.8)	45.87 (7.52)	29.69 (11.9)	0.00 (0.00)	3.81 (0.78)	1.475 (0.38)	99
			Model	175.36 (44.87)	9.57 (34.99)	47.82 (7.74)	29.14 (11.6)	0.173 (0.077)	1.857 (0.39)	1.089 (0.279)	99.00

(E) Coal-Water Slurry Runs:

Coal Type	Input Conditions: Coal Rate (g/sec)	$\frac{O_2}{Coal}$	$\frac{Water}{Coal}$	Dry Product Gas / Source	CO Vol. %	H_2 Vol. %	CO_2 Vol. %	CH_4 Vol. %	H_2S Vol. %	N_2 Vol. %	Carbon Conversion %
Western	186.78	0.91	0.51	Exp.	50.71	35.79	13.14	0.09	0.03	0.24	92.7
				Model	47.86	37.30	14.45	0.055	0.074	0.26	94.87
Eastern	133.5	0.87	0.79	Exp.	41.55	36.15	20.64	0.40	0.85	0.38	85.8
				Model	43.27	36.91	18.91	0.036	0.50	0.373	84.66

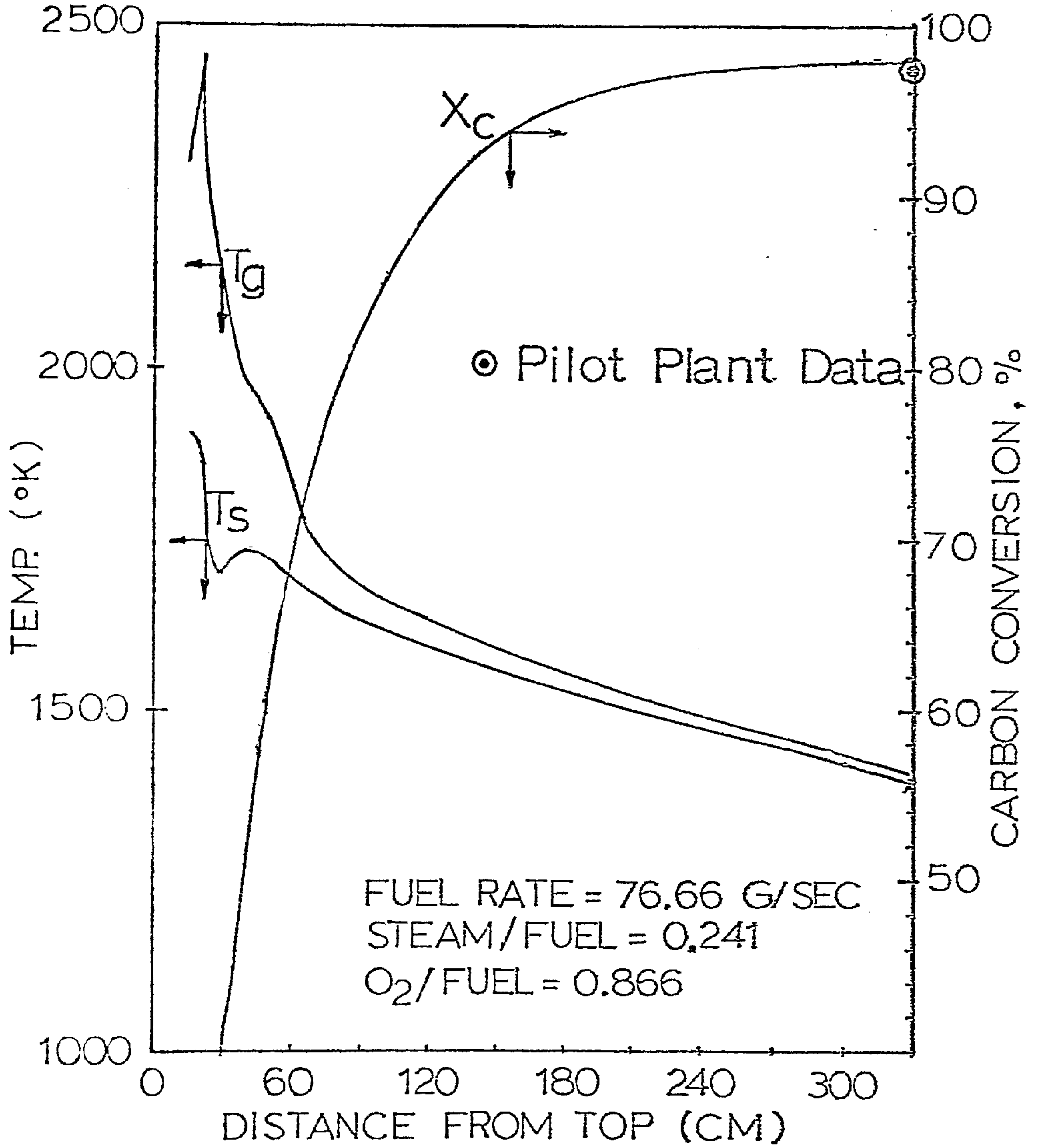


FIGURE 2(A) CALCULATED TEMPERATURE PROFILES AND CARBON CONVERSION PROFILE IN TEXACO PILOT PLANT GASIFIER FOR I-1 RUN.

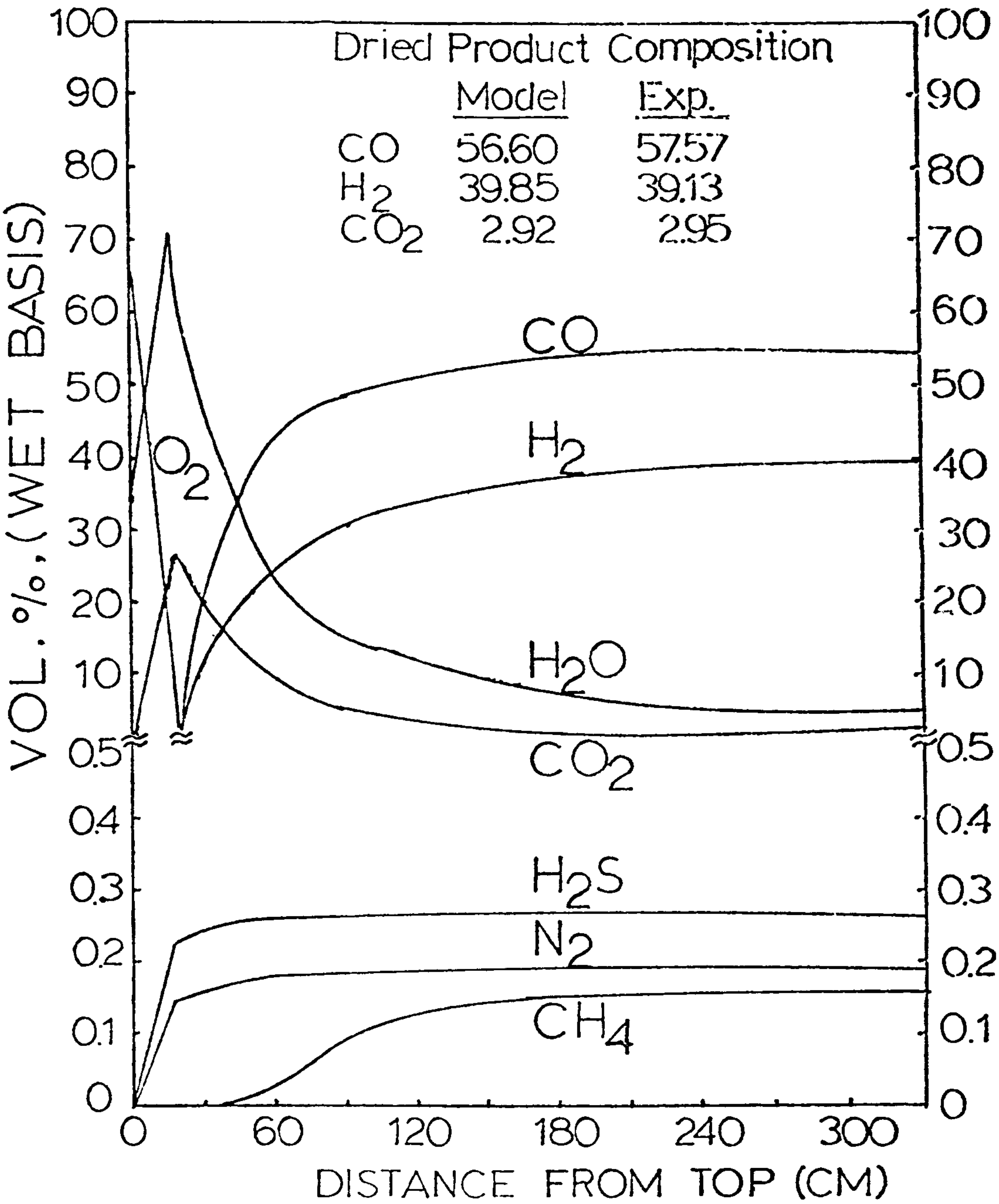


FIGURE 2(B) CALCULATED PRODUCT GAS COMPOSITION PROFILES (WET BASIS) IN TEXACO PILOT PLANT GASIFIER FOR I-1 RUN

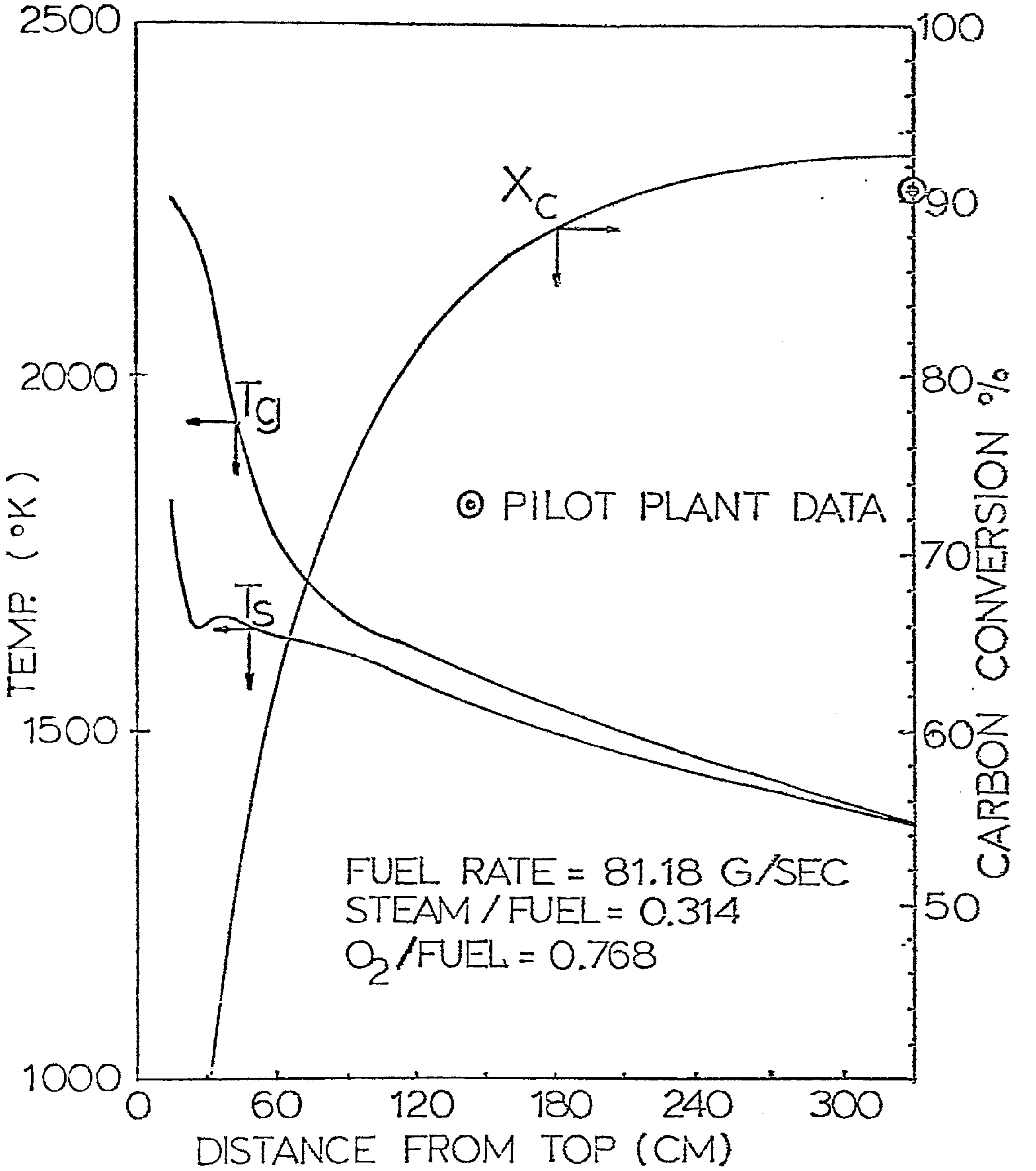


FIGURE 3(A) CALCULATED TEMPERATURE PROFILES AND CARBON CONVERSION PROFILE IN TEXACO PILOT PLANT GASIFIER FOR I-2 RUN

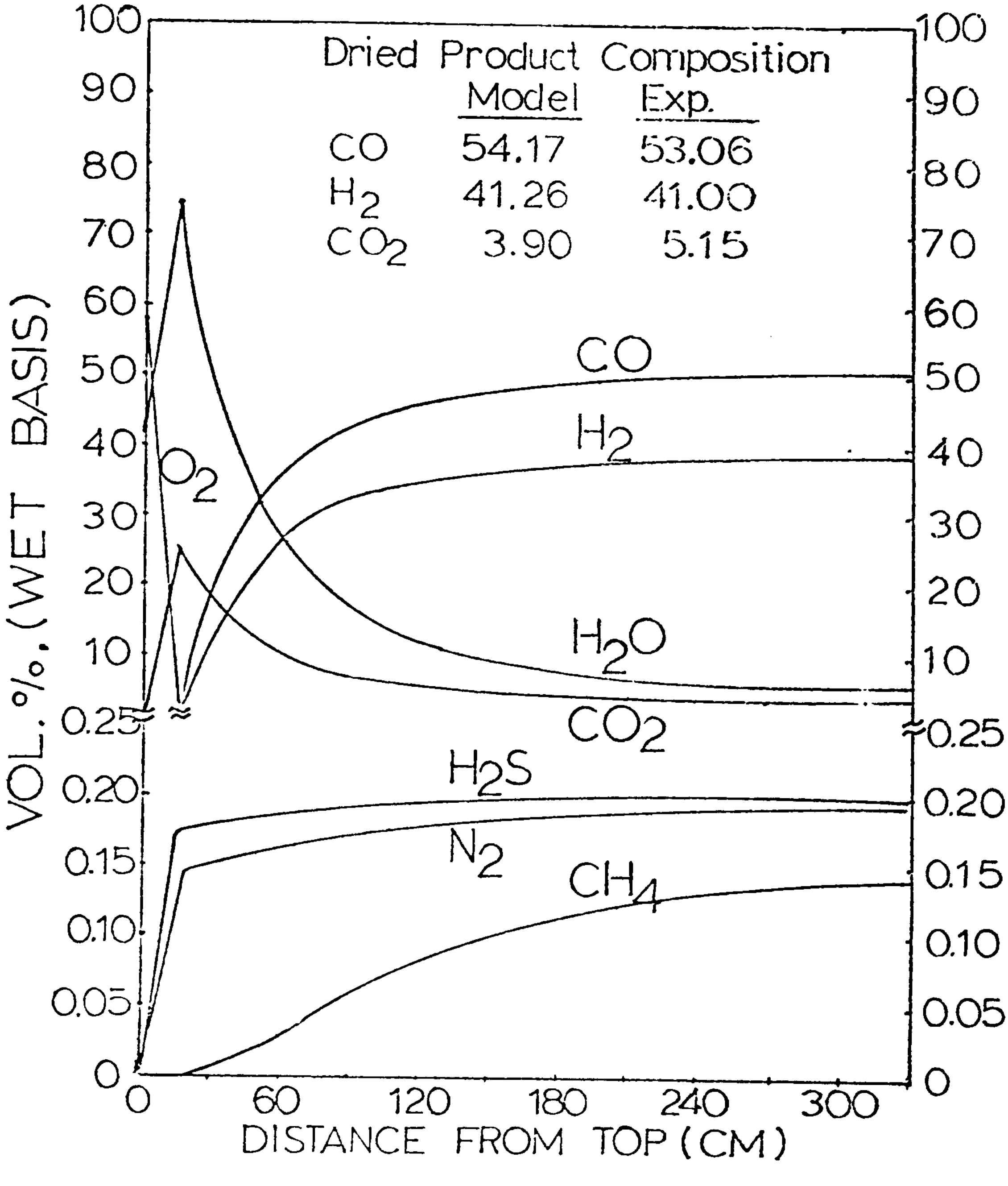


FIGURE 3(B) CALCULATED PRODUCT GAS COMPOSITION PROFILES (WET BASIS) IN TEXACO PILOT PLANT GASIFIER FOR I-2 RUN

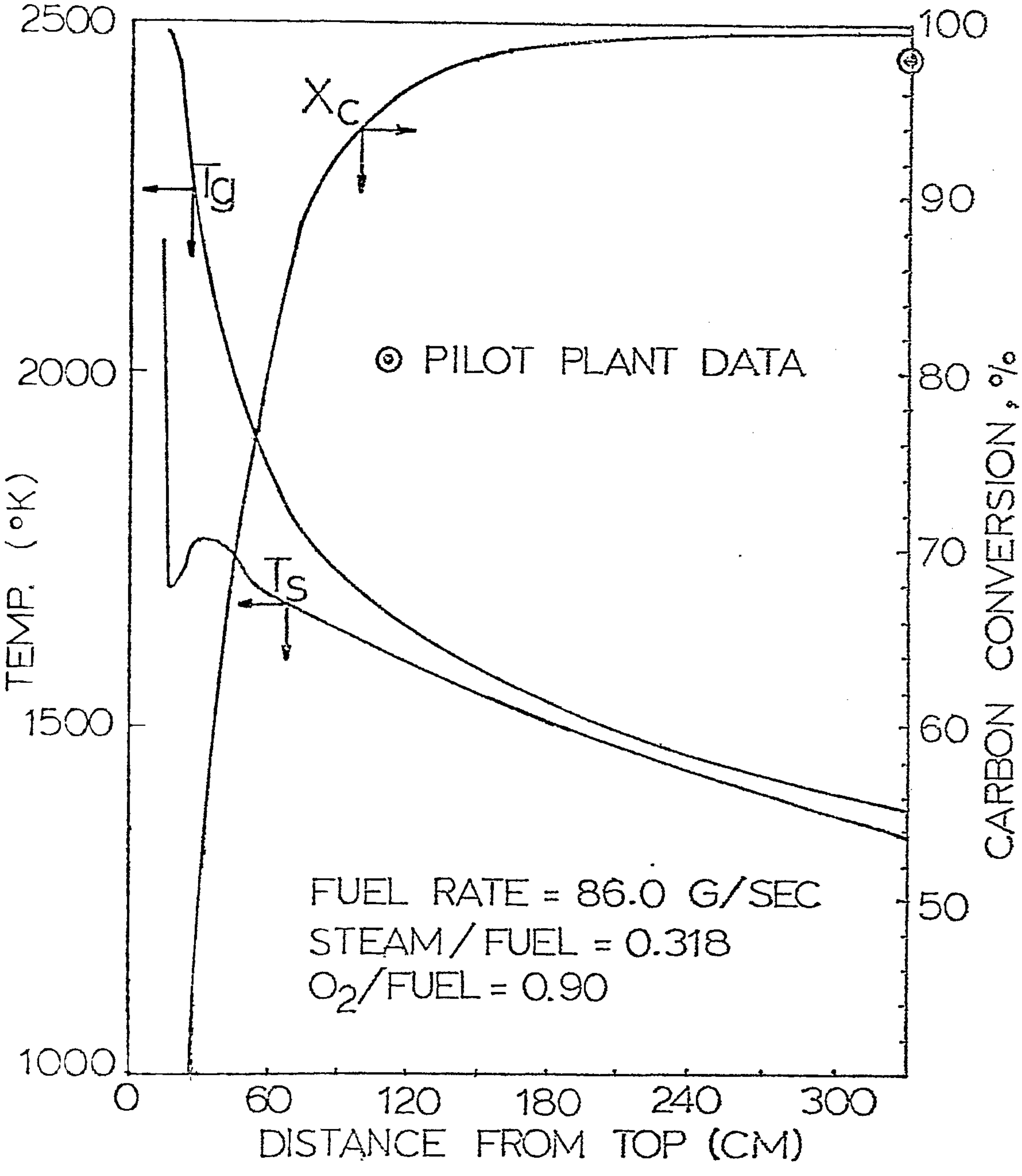


FIGURE 4(A) CALCULATED TEMPERATURE PROFILES AND CARBON CONVERSION PROFILE IN TEXACO PILOT PLANT GASIFIER FOR W-1 RUN

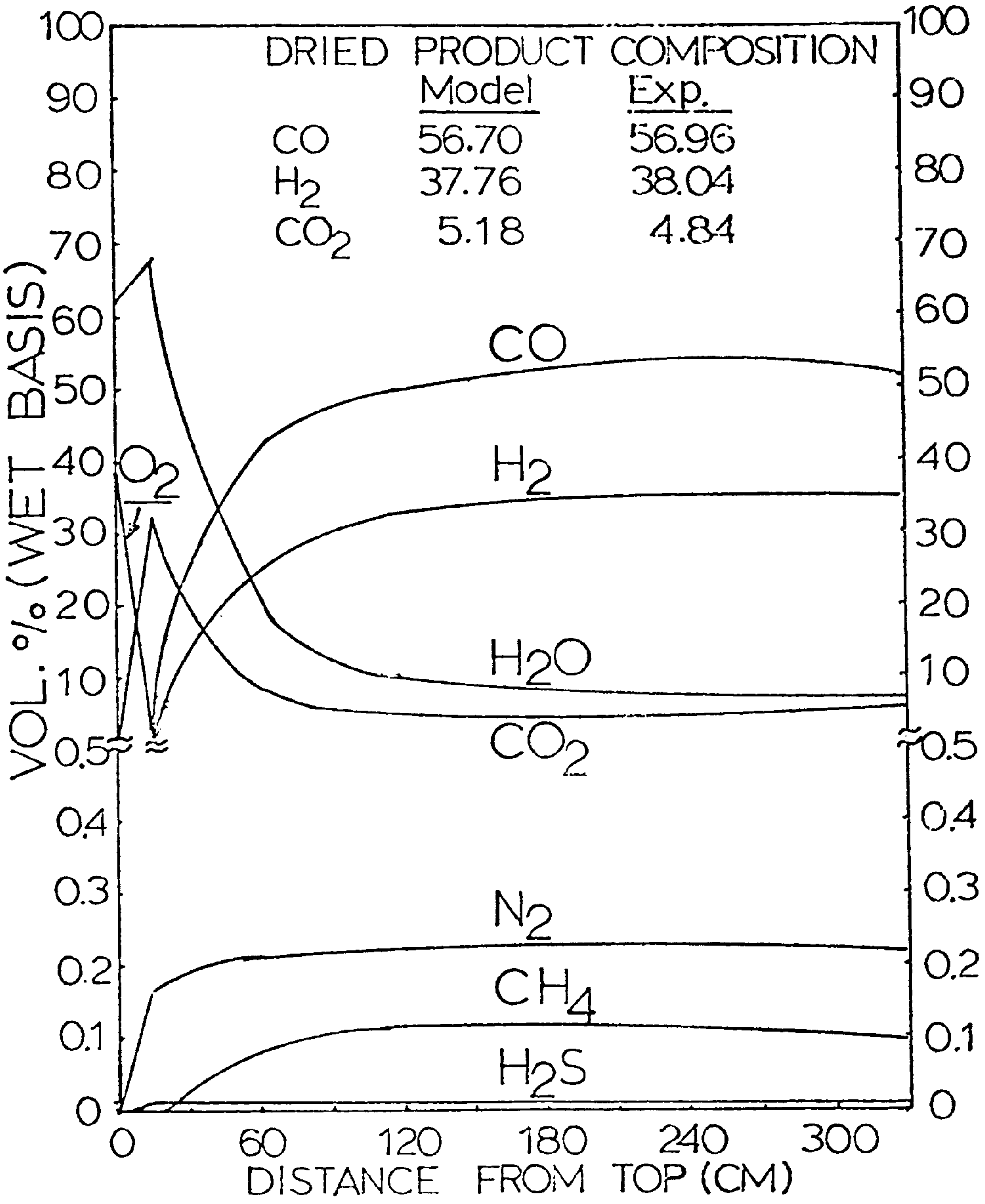


FIGURE 4(B) CALCULATED PRODUCT GAS COMPOSITION PROFILES (WET BASIS) IN TEXACO PILOT PLANT GASIFIER FOR W-1 RUN

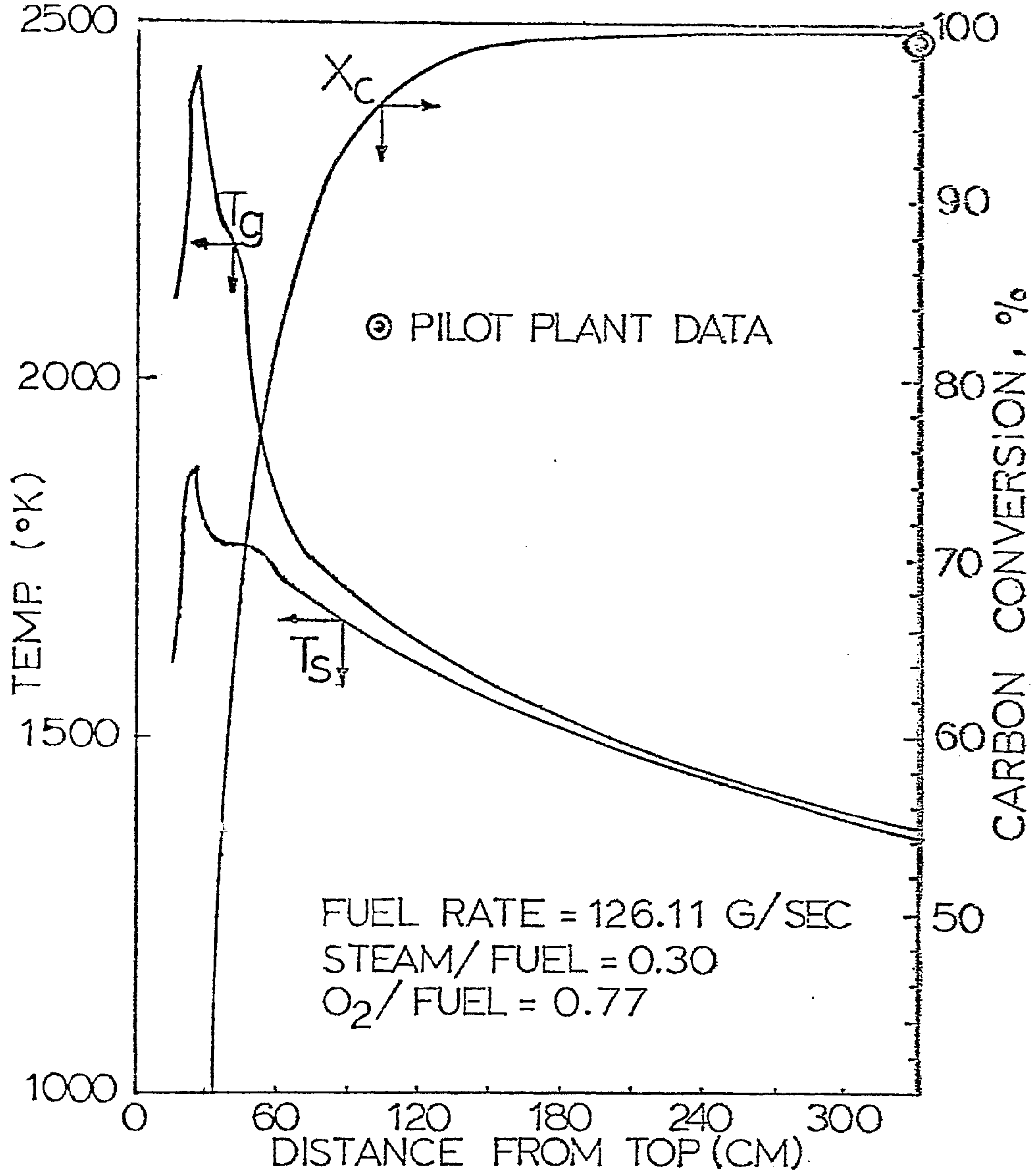


FIGURE 5(A) CALCULATED TEMPERATURE PROFILES AND CARBON CONVERSION PROFILE IN TEXACO PILOT PLANT GASIFIER USING SRC II VACUUM FLASH DRUM BOTTOMS AS FEEDSTOCK

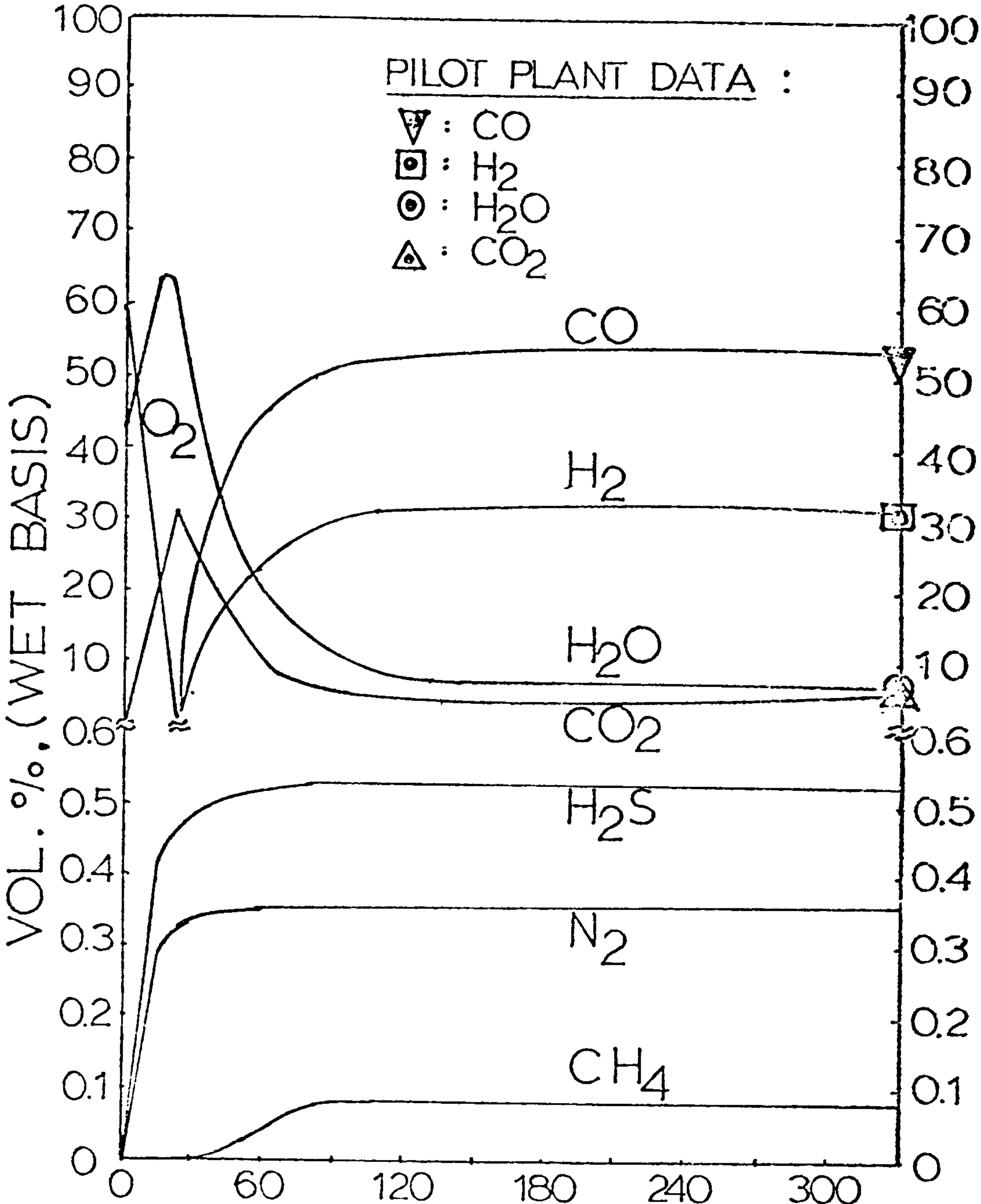


FIGURE 5(B) CALCULATED PRODUCT GAS COMPOSITION PROFILES (WET BASIS) IN TEXACO PILOT PLANT GASIFIER USING SRC II VACUUM FLASH DRUM BOTTOMS AS FEEDSTOCK

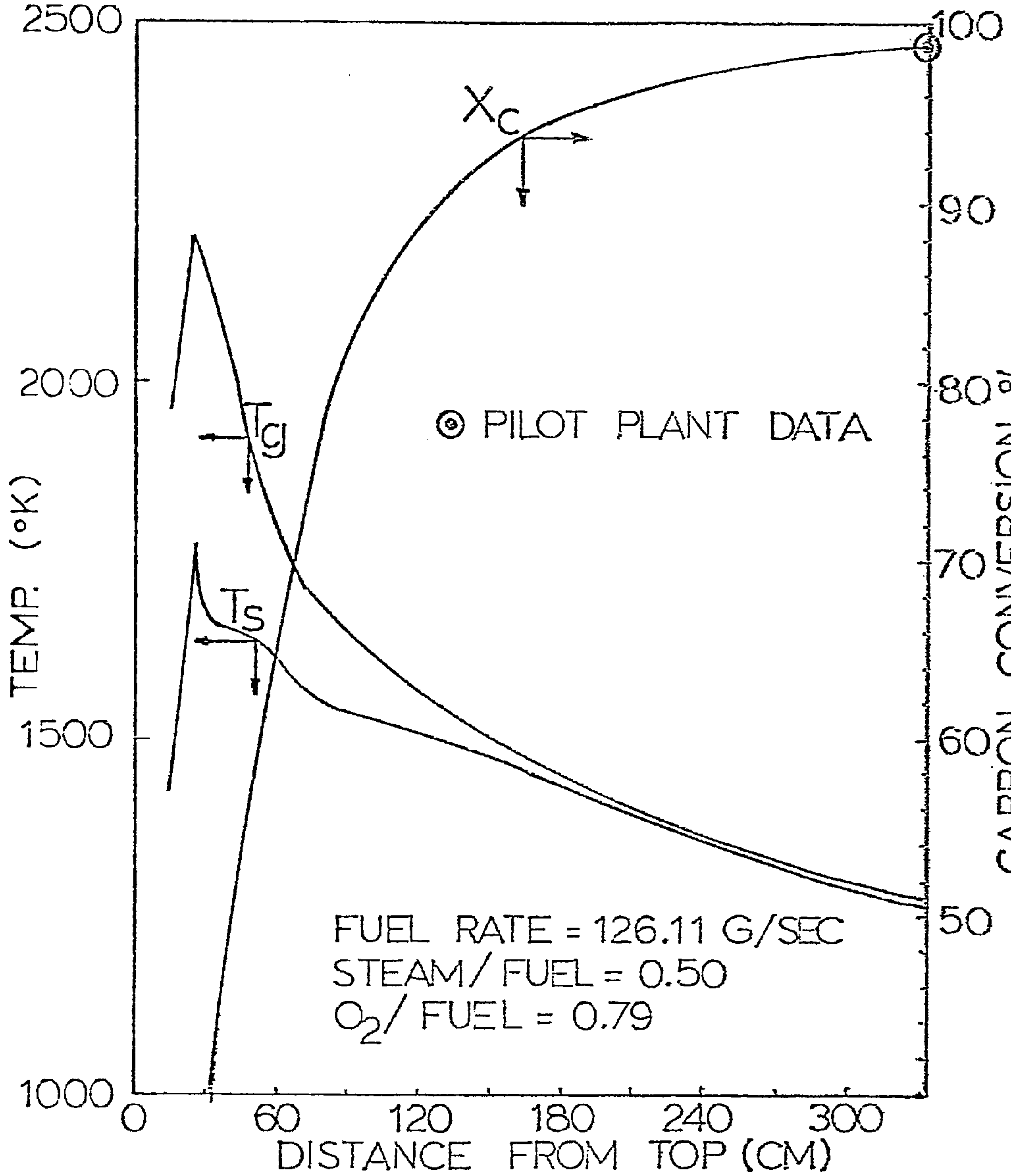


FIGURE 6(A) CALCULATED TEMPERATURE PROFILES AND CARBON CONVERSION PROFILE IN TEXACO PILOT PLANT GASIFIER USING EXXON DSP VACUUM TOWER BOTTOMS AS FEEDSTOCK.

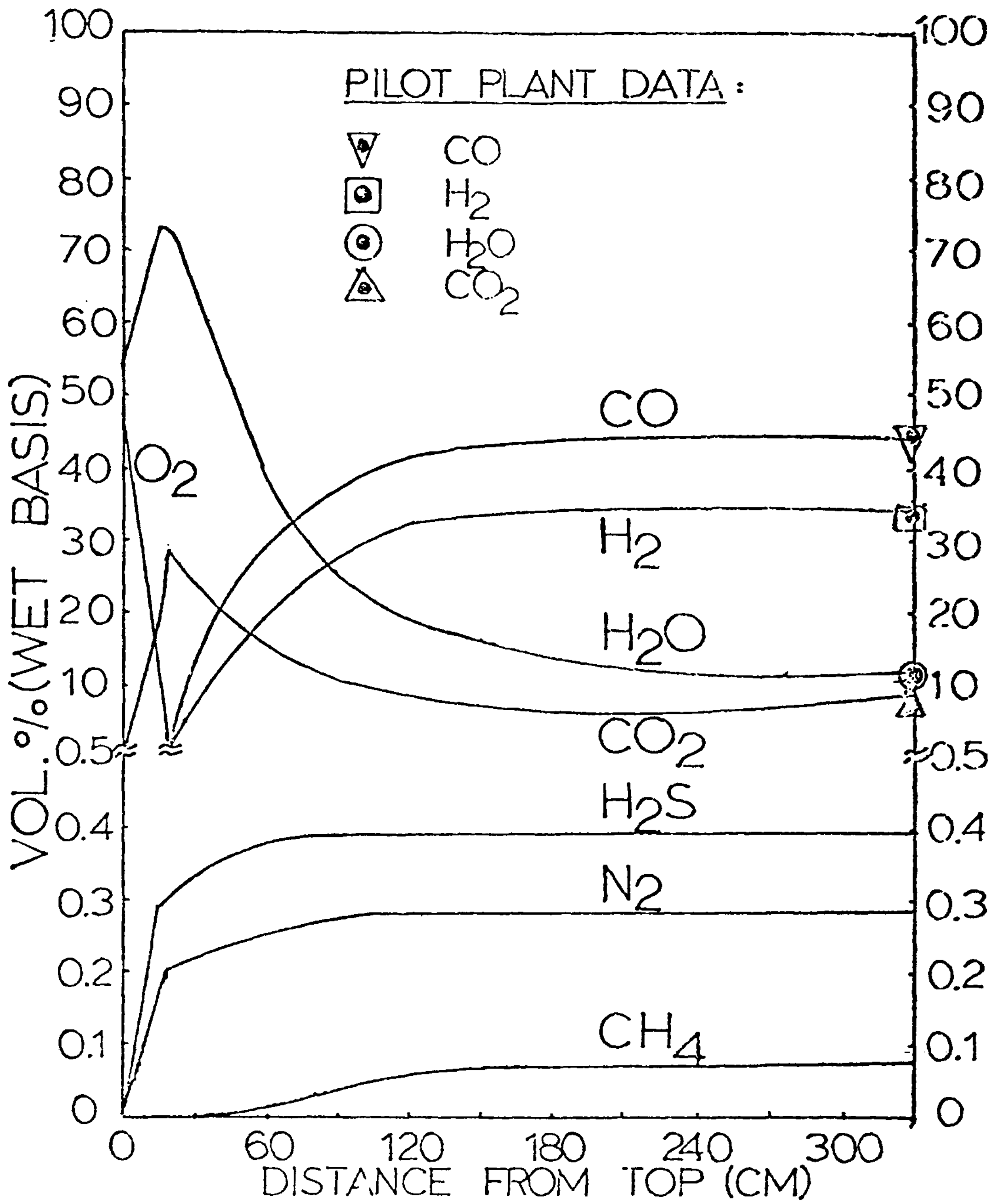


FIGURE 6(B) CALCULATED PRODUCT GAS COMPOSITION PROFILES (WET BASIS) IN TEXACO PILOT PLANT GASIFIER USING EXXON DSP VACUUM TOWER BOTTOMS AS FEEDSTOCK

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