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Analysis of Test Data From IGT High Pressure Fluidized-Bed Gasifier PDU
Volume II

Topical Report

A. Avidan
R. Shinnar

December 1988

Work Performed Under Contract No.: DE-AC21-86MC23077

For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia

By
Foster Wheeler USA Corporation
Clinton, New Jersey

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ANALYSIS OF TEST DATA FROM IGT
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VOLUME - II

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1.0 SUMMARY

Experimental data from IGT's high pressure fluidized-bed PDU for both Phase-1 and Phase-2 were checked for consistency using the invariant technique developed by CUNY. In addition, the usefulness of the IGT data relative to the design and performance projections for a commercial gasifier was assessed.

Of the data from eight Phase-1 test runs and fifteen Phase-2 test runs, only two data points, one from each phase, were found to be reasonably consistent. This is determined by comparing the carbon conversion calculated using the two conventional methods as well as the invariant method. For these two consistent runs carbon conversion calculated by all three methods are approximately equal. Data from the other twenty-one test runs are in varying degrees of inconsistency. CUNY's kinetic model for fluidized-bed gasification was used in an attempt to reconcile the inconsistent data and to determine what could be the reason for their inconsistency. The results indicated that the coal feed rate and/or the product gas methane content was likely in error.

The data also indicated that the PDU had been operated with high oxygen to carbon feed ratios, high steam to oxygen feed ratios and very low solids holdup in the gasifier. All of these conditions contributed towards low carbon conversion at the operating temperatures. In general, the PDU operating conditions were not representative of those expected for practical operation of a commercial gasifier. As a result, the PDU data were judged unsuitable for use in predicting commercial gasifier performance or design of commercial gasifier.

Based on the kinetic model determination, the IGT high pressure gasifier has only a small window for viable commercial operation on Montana Rosebud coal, at reasonable conversions and temperatures. It is recommended that commercialization of the IGT high pressure fluidized bed gasifier be considered only if an additive, such as calcium oxide is included. It is also recommended that the bottom of the IGT PDU gasifier be redesigned to provide more positive control of bed height and solids holdup in the gasifier.

2.0 INTRODUCTION

Under a subcontract with Foster Wheeler USA Corporation, the City University of New York (CUNY), conducted an evaluation of the experimental data obtained by the Institute of Gas Technology (IGT) in a high pressure fluidized-bed gasifier PDU. The specific objectives of CUNY's work were to:

- Check the steady state data, reported by IGT, for consistency using the invariant technique developed by CUNY.
- Compare the experimental results with predictions based on CUNY's kinetic model for fluidized-bed gasification.
- Assess the usefulness of the PDU data for projecting the performance of a commercial scale gasifier.

The scope of this evaluation effort included the experimental data obtained by IGT in both their Phase-1 (1985) and Phase-2 (1986) test programs, which were sponsored by the Gas Research Institute under GRI Contract No. 5084-221-1040. These gasification tests were conducted at pressure levels of 100-450 psig on three coal feedstocks; eastern bituminous coal, Montana subbituminous coal, and North Dakota lignite.

Essentially all of the test data evaluated were contained in IGT's final report on their high pressure PDU test program (1). Additional data delineating the purge nitrogen flow to the fluidized-bed was obtained via correspondence with IGT. Foster Wheeler's report (2) on analysis of IGT's Phase 1 test data provided additional background material for CUNY.

CUNY's evaluation was performed in conjunction with Foster Wheeler's analysis of the Phase-2 test data reported by IGT. The results of Foster Wheeler's analysis are summarized in Volume - I of this report.

3.0 DATA CONSISTENCY

A review of both Foster Wheeler's report on analysis of IGT's Phase-1 data (2) and IGT's final report (1), as well as discussions with IGT during the preparation of this analysis, revealed some serious problems with collection of PDU data. The following problems were noted:

1) Oxygen flow measurement - errors were discovered at the end of Phase-1 operation in the measured oxygen flow. It was necessary to correct the measured Phase-1 oxygen flows by a factor of about 0.65 to obtain the actual flow. IGT was able to identify and correct this problem before the Phase-2 runs.

2) Product gas flow measurement - erroneous product gas flow measurements persisted throughout the Phase-1 and Phase-2 test periods. The problem was attributed to corrosion/erosion of the product gas flow orifice. Product gas flows were therefore indirectly determined based on nitrogen feed rates and product gas compositions and by forcing a nitrogen balance. Foster Wheeler reported that the product gas on a quench free basis calculated by the nitrogen balance method are consistently higher than the measured flows (2).

3) Product gas analyses - the on-line gas chromatograph analyzer was often not in service or unable to produce accurate analysis. As a result, only the gas bomb sample analyses were available for data use. This is a serious problem because bomb data are less reliable and samples were not frequent enough to check out possible fluctuations.

Summaries of the averaged PDU operating conditions for Phase-1 and Phase-2 are given in Tables 1, 2 and 3. In these tables, the reported ratios of nitrogen and oxygen feed were based on purge nitrogen to the fluidized bed and not the total purge nitrogen flow. Similarly, the gasifier superficial velocities were recalculated to include the purge nitrogen flow. The recalculated superficial velocities are substantially higher than that reported by IGT reflecting the significant contribution of purge nitrogen in IGT's PDU operation.

The balances prepared by IGT and used for this analysis are based on forced closure of the nitrogen, hydrogen and sulfur. The nitrogen balance was used to calculate the net dry product gas; the hydrogen balance was used to determine the water content of the product gas; and the sulfur balance was used to determine the hydrogen sulfide content of the product gas.

TABLE 1

SUMMARY OF IGT PHASE-1 DATA

	Montana Rosebud				N.D. Lignite			PGH #8
Run Point	2-6.1	2-6.2	3-1.1	3-2.1	4-1.1	4.2-3	4-2.1	1-2
Pressure, psig	96	195	283	198	292	95	193	97
Temp (freeboard), ^{o}F	1600	1580	1509	1575	1511	1370	1340	1791
Oxygen/Carbon, lbmol/lbmol	0.261	0.299	0.302	0.285	0.277	0.272	0.234	0.369
Steam/Oxygen, lbmol/lbmol	4.37	5.27	6.94	4.96	6.14	8.58	10.31	4.55
Nitrogen/Oxygen, lbmol/lbmol	1.32	1.65	1.86	1.43	1.62	2.51	2.84	1.87
Coal Throughput, lb/hr (dry)	181.7	224.7	273.5	301.5	398.6	110.7	214.4	115.4
Bed Height, ft	2.7	N/A	N/A	2.8	3.0	2.3	2.8	3.1
Superficial Velocity, ft/sec	2.60	2.29	2.39	2.69	2.44	2.58	2.56	3.20
Coal Moisture, lb/lb coal dry	0.15	0.15	0.20	0.18	0.10	0.11	0.10	0.01

TABLE 2

SUMMARY OF IGT PHASE-2 DATA ON MONTANA ROSEBUD COAL

Run Point	5-1.1	5-1.2A	5-1.2B	5-2.1	5-2.2	5-2.3	5-3.1	5-3.2	5-3.3
Pressure, psig	201	300	302	450	450	449	449	448	448
Temp (freeboard), °F	1580	1533	1461	1548	1633	1610	1590	1706	1491
Oxygen/Carbon, lbmol/lbmol	0.284	0.248	0.245	0.332	0.276	0.294	0.352	0.341	0.372
Steam/Oxygen, lbmol/lbmol	6.13	6.14	7.38	7.25	4.75	5.80	6.69	4.37	8.79
Nitrogen/Oxygen lbmol/lbmol	1.19	1.32	1.66	1.82	1.10	1.33	1.32	1.07	2.10
Coal Throughput, lb/hr (dry)	333.8	497.6	426.4	454.3	785.2	601.8	475.2	730.1	334.2
Bed Height, ft	2.2	3.3	2.5	2.7	3.8	3.2	3.1	2.7	4.4
Superficial Velocity, ft/sec	3.36	3.04	3.01	2.88	2.83	2.82	2.86	3.23	2.78
Coal Moisture, lb/lb coal dry	0.06	0.08	0.12	0.10	0.10	0.10	0.12	0.14	0.14

TABLE 3

SUMMARY OF IGT PHASE-2 DATA ON NORTH DAKOTA LIGNITE

Run Point	6-1.1	6-2.1	6-2.2	6-2.3	6-2.4
Pressure, psig	200	300	448	447	447
Temp (freeboard), °F	1431	1419	1413	1426	1531
Oxygen/Carbon, lbmol/lbmol	0.321	0.331	0.313	0.296	0.290
Steam/Oxygen, lbmol/lbmol	6.88	7.87	7.56	6.19	4.47
Nitrogen/Oxygen, lbmol/lbmol	2.05	1.95	2.21	1.98	1.40
Coal Throughput, lb/hr (dry)	230.3	298.2	403.2	478.8	698.7
Bed Height, ft	4.2	3.7	4.2	5.5	2.6
Superfial Velocity, ft/sec	2.62	2.62	2.31	2.19	2.39
Coal Moisture, lb/lb coal dry	0.13	0.19	0.21	0.23	0.24

Carbon, oxygen, ash and energy balance closures are tabulated in Tables 4 and 5. It can be seen that there are significant imbalances in both Phase-1 and Phase-2 data. For example, the carbon closure vary from 74% to 107%.

As shown in Figure 1, the energy balance closure is, in most cases, 2 to 10% higher than the carbon balance closure. This could indicate certain discrepancy in the PDU data due to poor carbon balances, inaccurate product gas composition and flow determinations, or inaccurate determination of the heating value of the coal.

There are two common methods for computing fraction carbon conversion, X. One is based on product gas where the total carbon atoms in the product gas is divided by the carbon atoms in the coal feed. Equation 1 is the mathmatical representation of this method.

$$X(\text{Gas}) = \frac{\text{Carbon in Product Gas}}{\text{Carbon in Coal Feed}} \qquad \text{Equation 1}$$

The other method is based on solid effluent streams where the total carbon atoms in the solid effluent streams are subtracted from the carbon atoms in the coal feed and the result is divided by the carbon atom in the coal feed, i.e.:

$$X(\text{Solid}) = 1- \frac{\text{Carbon in Effluent Solids}}{\text{Carbon in Coal Feed}} \qquad \text{Equation 2}$$

Equation 2 will generally overestimate the carbon conversion level because some of the solids entrained in the product gas are not accounted for.

CUNY has developed another method of calculating carbon conversion which is derived from element balances based on stoichiometric invariants (3, 4 & 5). The invariant is defined, using only the dry product gas composition, as follows:

$$I_s = \frac{Y_{CO} + Y_{H2} + 4\ Y_{CH4}}{Y_{CO} + Y_{CO2} + Y_{CH4}} \qquad \text{Equation 3}$$

Where Y is the mole fraction of components in the dry product gas.

TABLE 4

PHASE-1 DATA CONSISTENCY AND BALANCES CHECK

	PGH #8	Montana Rosebud				N.D. Lignite		
RUN POINT	1-2	2-6.1	2-6.2	3-1.1	3-2.1	4-1.1	4-2.3	4-2.1
CARBON CLOSURE *	90.8	97.4	100.9	106.3	90.9	101.0	98.4	94.9
OXYGEN CLOSURE *	96.7	99.0	98.9	101.5	96.1	99.7	98.7	101.2
ASH CLOSURE *	106.3	98.0	98.9	101.5	96.1	99.7	98.7	101.2
ENERGY BALANCE CLOSURE *	97.0	103.5	105.1	105.5	97.2	104.1	101.8	97.4
CARBON CONVERSION X (GAS), %	79.5	87.8	89.7	85.4	73.4	89.1	72.5	67.4
X (SOLID), %	88.3	90.1	88.9	82.6	82.8	88.1	71.6	70.5
X (INVARIANT), %	94.8	91.5	93.9	74.6	92.8	88.9	76.5	58.6
I_S REPORTED	1.58	1.66	1.59	1.47	1.60	1.51	1.45	1.38

* OUT/IN

TABLE 5

PHASE-2 DATA CONSISTENCY AND BALANCES CHECK

	Montana Rosebud								
Run Point	5-1.1	5-1.2A	5-1.2B	5-2.1	5-2.2	5-2.3	5-3.1	5-3.2	5-3.3
Carbon Closure*	87.7	74.2	94.0	95.3	107.4	86.7	101.8	106.2	98.3
Oxygen Closure*	100.1	88.8	98.8	98.2	101.4	98.5	101.5	97.0	97.3
Ash Closure*	78.2	86.1	90.7	95.1	92.9	97.2	90.3	99.8	91.5
Energy Balance Closure	89.0	88.2	97.4	98.1	106.2	91.1	98.5	106.9	102.1
Carbon Conversion									
X(gas),%	69.5	49.8	61.4	75.1	88.7	71.9	85.3	97.0	72.5
X(solids),%	82.1	76.4	66.3	79.4	81.8	85.3	83.1	90.8	74.8
X(invariant),%	69.2	195.1	68.9	83.9	80.6	78.7	77.3	109.3	89.1
I_s Reported	1.42	1.87	1.54	1.41	1.55	1.47	1.32	1.54	1.34

	N.D. Lignite					Ill. #6
Run Point	6-1.1	6-2.1	6-2.2	6-2.3	6-2.4	7-2.7
Carbon Closure*	95.3	103.8	100.5	105.4	107.7	99.1
Oxygen Closure*	95.7	97.2	99.4	97.7	99.4	99.3
Ash Closure*	98.1	104.7	104.5	108.6	93.0	90.6
Energy Balance Closure	100.9	106.5	100.9	106.4	107.4	98.8
Carbon Conversion						
X(gas),%	73.6	80.7	76.2	79.9	87.6	84.1
X(solids),%	88.4	76.2	75.8	74.7	79.9	84.5
X(invariant),%	91.6	97.7	78.2	90.6	88.5	83.8
I_s Reported	1.505	1.45	1.35	1.48	1.48	1.47

$$I_s = \frac{YCO + YH2 + 4YCH4}{YCO + YCO2 + YCH4}$$

*Out/In

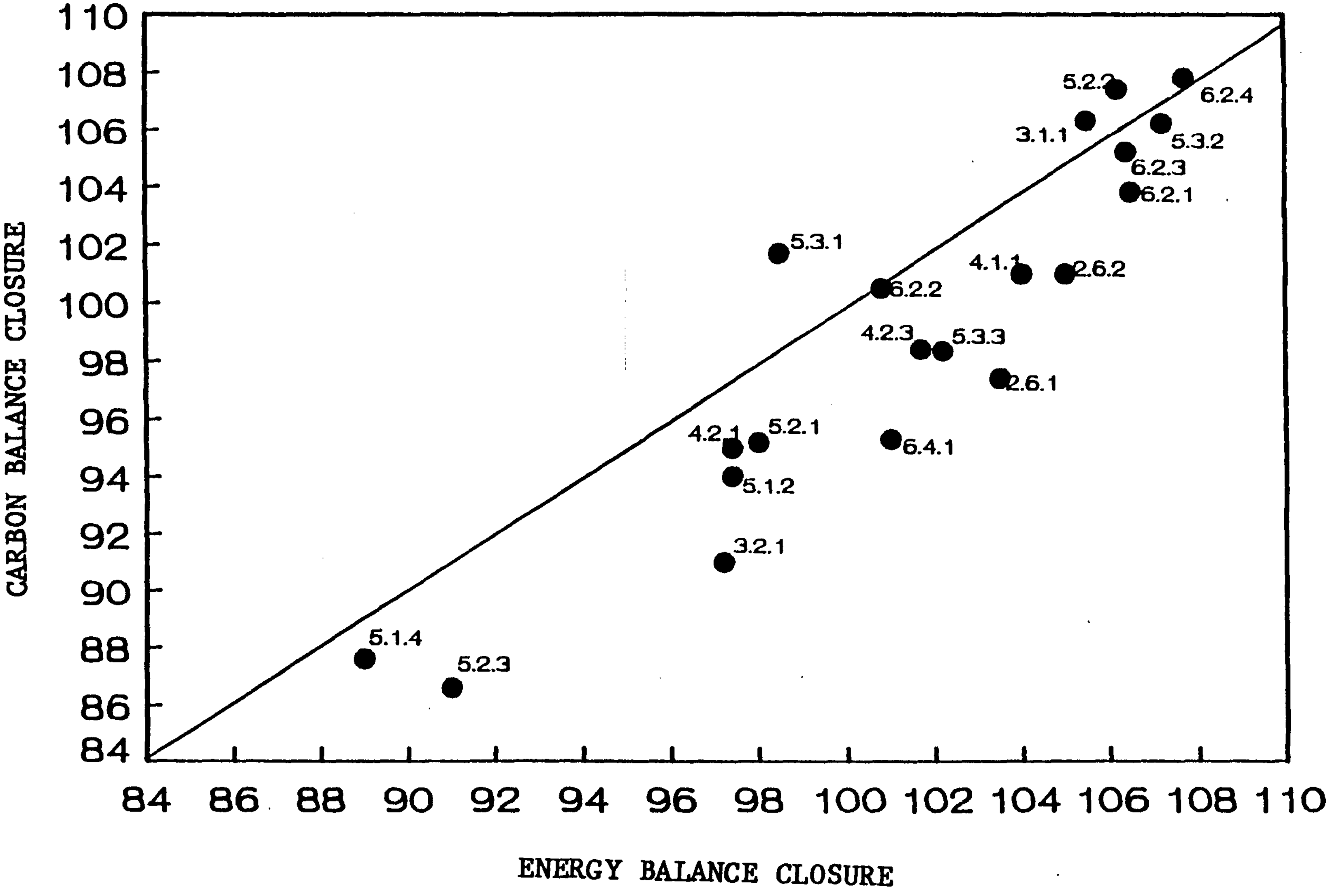


FIGURE 1

ENERGY BALANCE CLOSURE VERSUS CARBON BALANCE CLOSURE

From an element balance, the right hand side of equation 3 can be shown to equal:

$$\frac{Y_{CO} + Y_{H2} + 4Y_{CH4}}{Y_{CO} + Y_{CO2} + Y_{CH4}} = 2 + \frac{a}{2X} - \frac{b}{X} - \frac{2(O_2)}{CX} \qquad \text{Equation 4}$$

Where a and b are the stoichiometric coefficients in the coal composition CH_aO_b and $(O_2)/C$ is the molar ratio of oxygen feed to coal feed.

Equation 4 assumes that the chart does not contain hydrogen or oxygen. If it does, there is an added correction factor (3), but this correction factor is normally very small and can be neglected.

Equation 4 can be used to calculate the carbon conversion, X (invariant), by using the invariant, I_s. When X(Solid) is significantly lower than X(invariant), the data is not consistent. The calculated carbon conversion based on all three methods are given in Tables 4 and 5. The two runs in which the carbon conversions determined by all three methods are reasonably close, hence are considered internally consistent, are runs 4-1.1 and 6-2.2.

It is very difficult to get a set of data that is in complete elemental balance. What is acceptable is a question of technical opinion. However, the inconsistent data must never be used for evaluation and design without rationalization. It is necessary to reconcile the data so that element balances as well as energy balances close exactly.

The kinetic model for the fluidized-bed gasifier developed by CUNY (3) was used in an attempt to reconcile the PDU data. Predicted gasifier performances based on the reported coal properties and feed conditions (rate, pressure and temperature), were developed using the model. Reaction rates derived from the literature (6, 7) for similar coals were used for these calculations. The only assumption needed to use the model was the methane yield. The methane yield was therefore chosen to match IGT's measured results.

Using the kinetic model, the predicted gasifier temperature, product gas composition and carbon conversion were determined as reported on Tables 6, 7 and 8. When carbon conversions determined by the invariant method, equation 4, are reasonably close to those determined using equations 1 and 2 the model predicted values also fit the observed data well. These include the two internally consistent runs and several others such as runs 3-1.1, 4-2.3, 5-2.1 and 5-2.3, where I_s predicted matched very well with that reported. For these runs, the predicted values which are in complete balance, can be used for further evaluations.

TABLE 6

MODELING RESULTS ON PHASE-1 TEST DATA

RUN POINT	2–6.1	2–6.2	3–1.1	3–2.1	4–1.1	4–23	4–2.1
PRESSURE (PSIG)	96	195	283	198	292	95	193
TEMP. REPORTED (F)	1600	1580	1509	1575	1511	1370	1340
TEMP. MODEL (F)	1564	1575	1492	1561	1478	1327	1392
C CONVERSION (GAS)	87.8	89.7	85.4	73.4	89.1	67.4	72.5
C CONVERSION (ASH)	90.1	88.9	82.6	82.8	88.1	70.5	71.6
C CONV. (INVARIANT)	91.5	93.9	74.6	92.8	88.9	58.6	76.5
C CONV. (MODEL)	78.6	79.5	74.5	80.8	84.2	58.2	74.6
R	.332	.376	.405	.353	.329	.402	.369
Rc–R	.047	.010	–.031	.029	.013	–.123	–.051
EL	2.08	2.77	3.80	2.48	2.72	5.19	3.97
CGEF	70.8	67.8	59.2	69.7	72.5	46.1	62.1
Is, Reported	1.66	1.59	1.47	1.60	1.51	1.38	1.45
Is, Model	1.60	1.53	1.47	1.54	1.49	1.38	1.42

TABLE 7

MODELING RESULTS ON PHASE-2 TEST DATA ON MONTANA ROSEBUD COAL

RUN POINT	5-1.1	5-1.2B	5-2.1	5-2.2	5-2.3	5-3.1	5-3.2	5-3.2* COAL+42*	5-3.3
PRESSURE (PSIG)	201	302	450	302	449	449	450	450	448
TEMP. REPORTED (F)	1580	1461	1548	1633	1610	1590	1706	1706	1491
TEMP. MODEL (F)	1585	1464	1552	1635	1588	1582	1805	1685	1485
C CONVERSION (GAS)	69.5	61.4	75.1	88.7	71.9	85.3	97.0	93.5	72.5
C CONVERSION (ASH)	82.1	66.3	79.4	81.8	85.3	83.1	90.8	68.3	74.8
C CONV. (INVARIANT)	69.2	68.9	83.9	80.6	78.7	77.3	109.3	70.3	89.1
C CONV. (MODEL)	77.1	59.7	82.4	80.6	77.0	83.0	93.1	70.3	84.6
R	.368	.410	.403	.342	.382	.424	.365	.341	.440
Rc-R	-.004	-.076	-.029	.029	-.017	-.056	.016	.002	-.069
EL	3.00	3.93	3.97	2.37	3.13	4.08	2.56	2.31	5.21
CGEF	65.2	48.7	64.3	68.7	63.1	63.7	74.4	58.5	63.6
Is, Reported	1.42	1.54	1.41	1.55	1.47	1.32	1.54	1.54	1.34
Is, Model	1.49	1.49	1.41	1.55	1.47	1.37	1.45	1.54	1.34

TABLE 8

MODELING RESULTS ON PHASE-2 TEST DATA
ON NORTH DAKOTA LIGNITE

RUN POINT	6-1.1	6-2.1	6-2.2	6-2.3	6-2.4
PRESSURE (PSIG)	200	300	448	447	447
TEMP. REPORTED (F)	1431	1419	1413	1426	1531
TEMP. MODEL (F)	1430	1423	1415	1432	1536
C CONVERSION, GAS	73.6	80.7	76.2	79.9	87.6
C CONVERSION, ASH	77.4	76.2	75.8	74.7	79.9
C CONV., INVARIANT	91.6	97.7	78.2	90.6	88.5
C CONV., MODEL	85.1	83.6	78.4	83.7	83.7
R	.377	.396	.399	.354	.346
Rc–R	–.039	–.034	–.045	.009	.017
EL	3.62	4.33	4.29	3.14	2.51
CGEF	68.1	64.6	60.1	68.1	69.1
Is, Reported	1.505	1.45	1.35	1.48	1.48
Is, Model	1.39	1.35	1.35	1.44	1.45

However, carbon conversion for all these runs are relatively low. Regrettably, runs with high carbon conversions, in excess of 85%, are all highly inconsistent.

Also given on Tables 6, 7 & 8 are the calculated values of R, Rc-R, EL and CGEF using the model. Definition of these terms were previously given (3,5). They are used to measure the commercial viability of the gasifier. A commercially viable gasifier should have positive Rc-R, low EL and high CGEF. As can be seen from these tables, the IGT test run data are not in the commercially viable range.

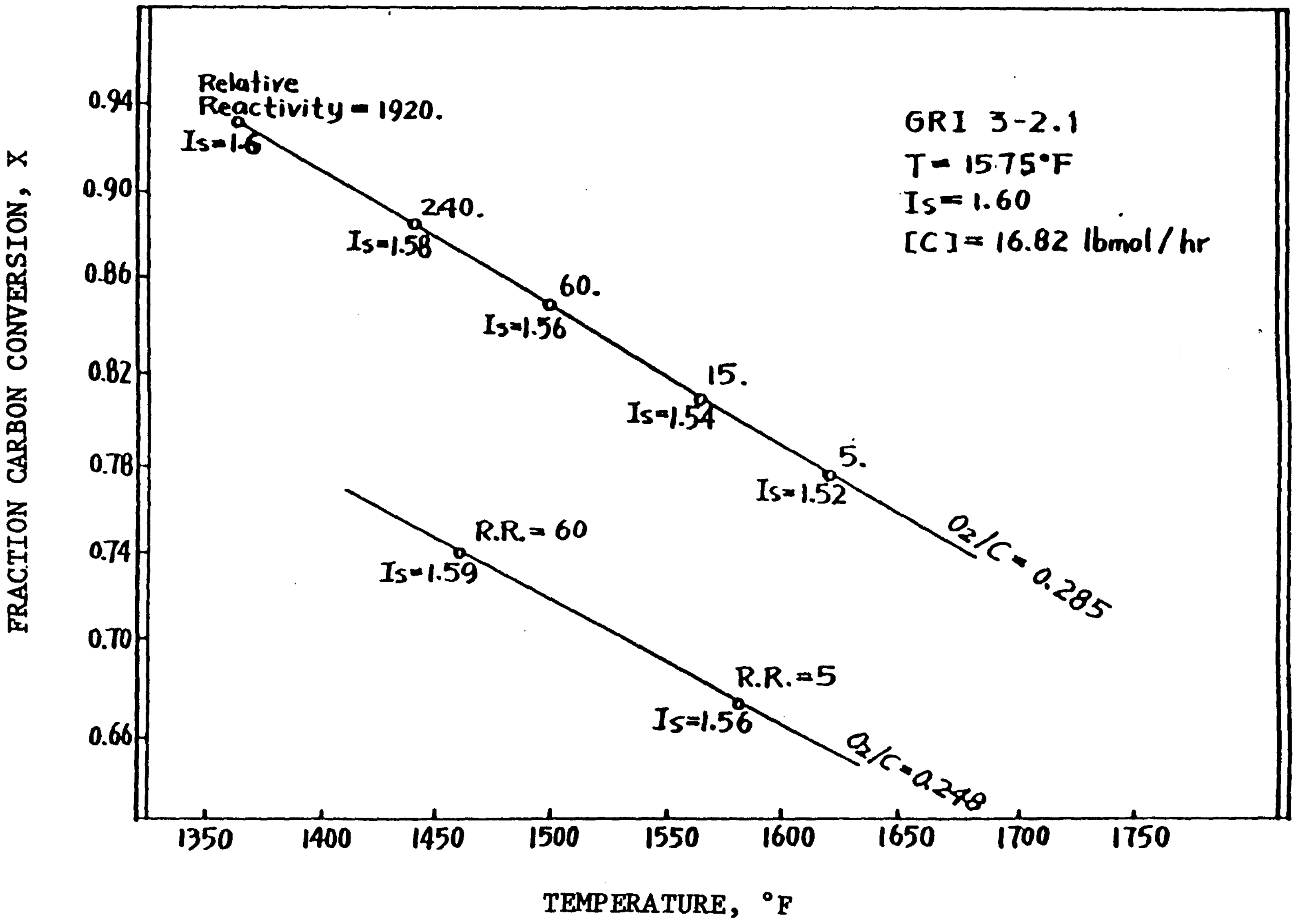
Attempts were made to determine why the predicted values deviated significantly from the reported values. For example, in run 3-2.1, the predicted bed temperature agreed reasonably well with the reported bed temperature. The predicted carbon conversion also matched that calculated using equations 1 and 2, but deviated from that calculated from equations 4. A possible reason is that the coal reactivity obtained from the literature was too low. It is possible to vary the coal reactivity in the model to raise the carbon conversion and match that calculated by equation 4. However, the increased conversion will require an increase in endothermic reaction heat and therefore the bed temperature will drop. This is shown on Figure 2 where carbon conversion is plotted against bed temperature. For the oxygen to carbon molar ratio of 0.285 used for this run, results of varying relative coal reactivity are shown. The relative coal reactivity obtained from the literature was 15 and the corresponding value for I_s was 1.54, X was 80.8% and bed temperature was 1561°F. In order to achieve a carbon conversion of 92.6% it is necessary to raise the relative reactivity to 1920. While I_s or gas composition, thus calculated, matched the reported I_s, the bed temperature fell to 1360°F which is about 200°F lower than the reported temperature.

Another possible reason is that the reported coal feed rate was inaccurate because dry coal feed is a difficult measurement. The effect of changing coal feed rate was examined for run 3-2.1. The coal feed rate for this run was allowed to increase so that the oxygen to carbon molar ratio is decreased to 0.248.

Carbon conversion, bed temperature and I_s were calculated using the model for this O_2/C ratio at different relative reactivities. The results of this determination are also shown in Figure 2. It is evident from this figure that both the reported bed temperature and I_s can be matched by reducing the feed O_2/C ratio drastically. But, it is also evident that the resultant carbon conversion is greatly reduced.

FIGURE 2

EFFECT OF COAL REACTIVITY



One other alternative is to assume that the product gas composition was in error. This could happen when the coal feed rate is not constant. A momentary high coal feed could result in high methane content in the product gas and, if the bomb sample was taken at that moment, the resultant gas analysis will not be consistent with the averaged steady state coal feed. This alternative was examined for run 3-2.1 and the results tabulated in Tables 9 and 10. On Table 9, the impact of methane yield on carbon conversion and bed temperature were tabulated from the PDU data, from the model calculation by assuming the same methane content in the product as that reported, and from the model calculation by assuming methane yield is in equilibrium with the other gas compounds in the product gas stream. On Table 10, the reported and calculated product gas composition were tabulated. When methane yield is assumed to be at equilibrium, the amount of methane produced per mole of carbon in the coal feed is reduced drastically. For this run, the calculated equilibrium methane yield produced is 0.0071 versus 0.092 reported. It can be seen that by reducing methane yield it is possible to bring the reported carbon conversion and bed temperature in line with the reported values. This points to a strong possibility that the source of inconsistency is due to error in the product gas methane content.

Another such case, run 5-3.2, was examined in more detail. For this case, an attempt was made to match both the bed temperature and I_s by varying the feed O_2/carbon ratio by increasing coal feed. The results are shown in Table 7 under run point 5-3.2*. In order to achieve a good match in carbon conversion with that calculated using equations 1, 2 and 4, coal feed must be increased by 42%. This indicated, at least for this run, that an error in coal feed could be the cause of the inconsistent results.

The above demonstrated that the data inconsistency could be caused by error in coal feed measurement and in product gas analysis. By trial and error, the inconsistent runs can all be reconciled. However, there is no way to determine which of the two possible causes, either individually or in combination, contributed to the inconsistency. Therefore, it was deemed futile to make such correction.

TABLE 9

IMPACT OF METHANE YIELD ON CARBON CONVERSION

	PDU data	Model1	Model2
Oxygen to carbon	0.285	0.285	0.285
Steam to oxygen	4.96	4.96	4.96
Methane yield (mole/mole C)	0.092	0.092	0.0071
Temperature (F)	1575	1516	1507
Carbon conversion (%)	73.5	80.8	70.3
Rc − R	0.08	0.03	−0.04
Is	1.60	1.54	1.47

PDU data: For Run 3.2-1

Model 1: Methane yield is the same as PDU data.

Model 2: Methane yield is the equilibrium data at PDU conditions.

TABLE 10

IMPACT OF METHANE YIELD ON GAS COMPOSITION

	PDU data	Model1	Model2
H_2	0.222	0.239	0.281
CO	0.098	0.097	0.095
CO_2	0.156	0.177	0.166
CH_4	0.036	0.036	0.003
H_2O	0.486	0.452	0.456
(Nitrogen free)			
Is	1.60	1.54	1.47

PDU data: For Run 3.2-1

Model 1: Methane yield is the same as PDU data.

Model 2: Methane yield is the equilibrium data at PDU conditions.

4.0 GASIFIER OPERATING REGIMES

It has been shown that the net efficiency of a gasifier as well as its economic attractiveness is strongly dependent on its oxygen and steam requirements (3). Low oxygen and steam requirements result in high efficiency and economic attractiveness. For a commercially attractive fluidized-bed gasifier the maximum steam to oxygen ratio would vary between 2 to 3. The fluidized-bed gasifier is expected to have an advantage over the entrained-flow gasifier because it could operate with considerably lower oxygen to coal ratio while maintaining the steam to oxygen ratio at around 2. The IGT PDU data, however, exhibited an oxygen to coal ratio close to that typically used for an entrained-flow gasifier. Hence, any steam added puts it in an economical disadvantage. The IGT PDU data also indicated that the steam to oxygen ratio used ranged between 4.5 to 8. In addition, large amounts of nitrogen was used for purging. Nitrogen acts as a coolant. Its effect is similar to steam. If the nitrogen added is included as equivalent steam to the gasifier, the effective ratio of steam to oxygen increases to 5.5 to 9.

The effect of varying the oxygen to carbon and steam to oxygen ratios can be calculated by using the CUNY fluidized bed gasifier kinetic model. One such calculation based on Montana Rosebud subbituminous coal at 30 atmospheres gasifier pressure, and 1% heat loss was performed. The result is shown on Figure 4 which is an operating map of the gasifier. The operating map is constructed by assuming a constant product gas velocity in the gasifier. With reasonable steam to oxygen ratios the difference between superficial gasifier velocity based on steam and oxygen feed and that based on the product gas is reasonably small. The operating map based on constant product gas velocity is indistinguishable from that based on constant superficial velocity.

The operating map is also constructed by assuming constant solid holdup in the gasifier. The gasifier solids holdup is equal to the product of bed density and bed height, expressed in pounds of bed material per square foot of bed cross-sectional area. Solids holdup has a pronounced effect on carbon conversion and gasifier temperature. This is shown on Figure 3 for two steam to oxygen feed ratios for a given oxygen to carbon ratio. Figure 3 indicates that, for a given steam/oxygen ratio, when holdup is decreased from about 300 lbs/ft^2 to about 60 lbs/ft^2, the conversion is drastically decreased and the gasifier temperature is increased by nearly 100°F. Figure 3 also shows that when bed holdup is decreased from 300 to 60 lbs/ft^2, the steam to oxygen ratio must be doubled in order to maintain the same bed temperature. Solids holdup could also affect methane content in the product gas, since methane, which is formed by coal devolatilization, is decomposed by reaction with steam at high temperatures. This reaction is catalized by coal ash. As holdup is increased the methane content could reduce.

Figure 4 is constructed with a linear velocity of 3 fps and a solids holdup of 50 lbs/ft^2. The linear velocity and solids holdup used to construct Figure 4 are similar to that used in IGT's PDU operation.

FIGURE 3

EFFECT OF BED HOLDUP

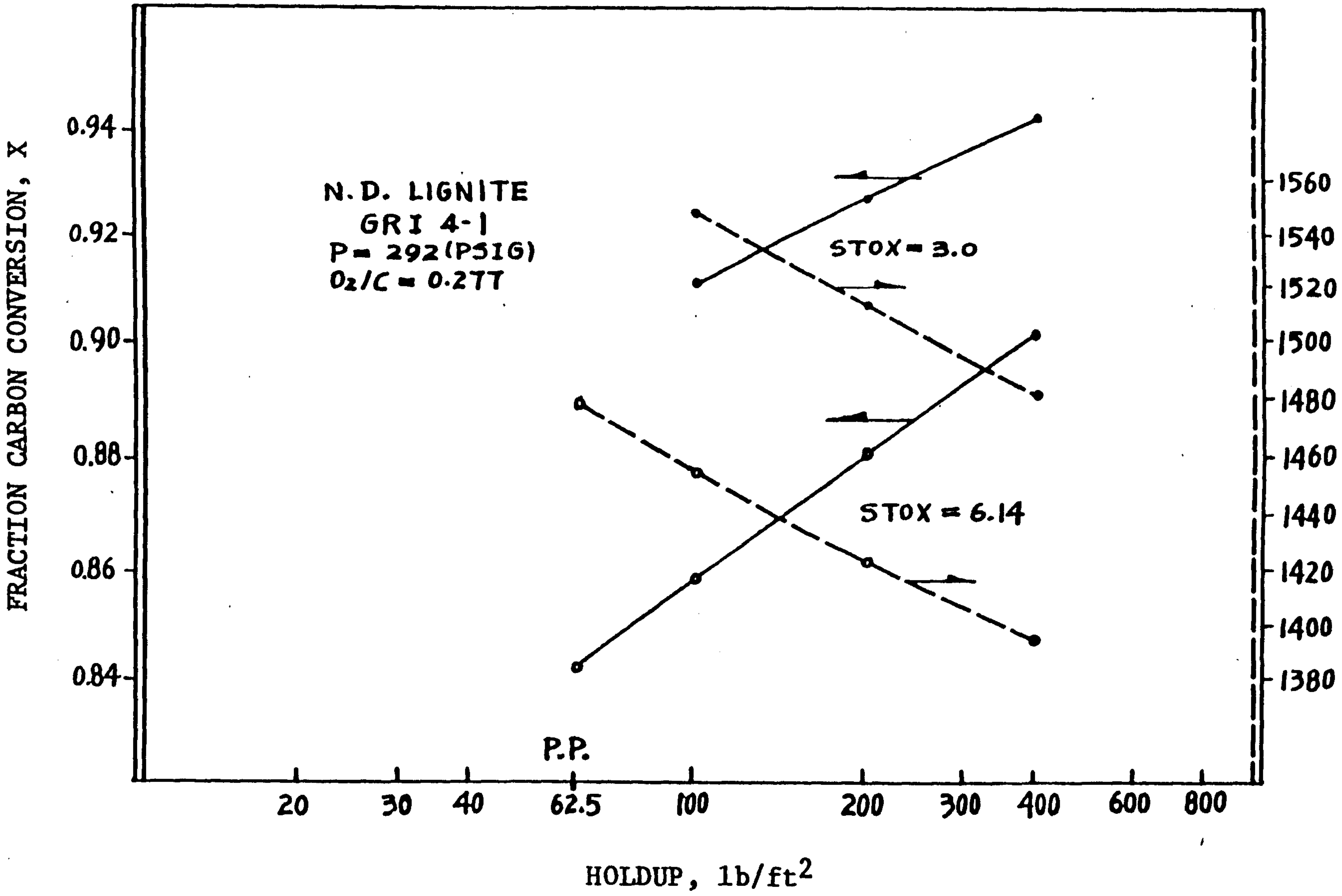
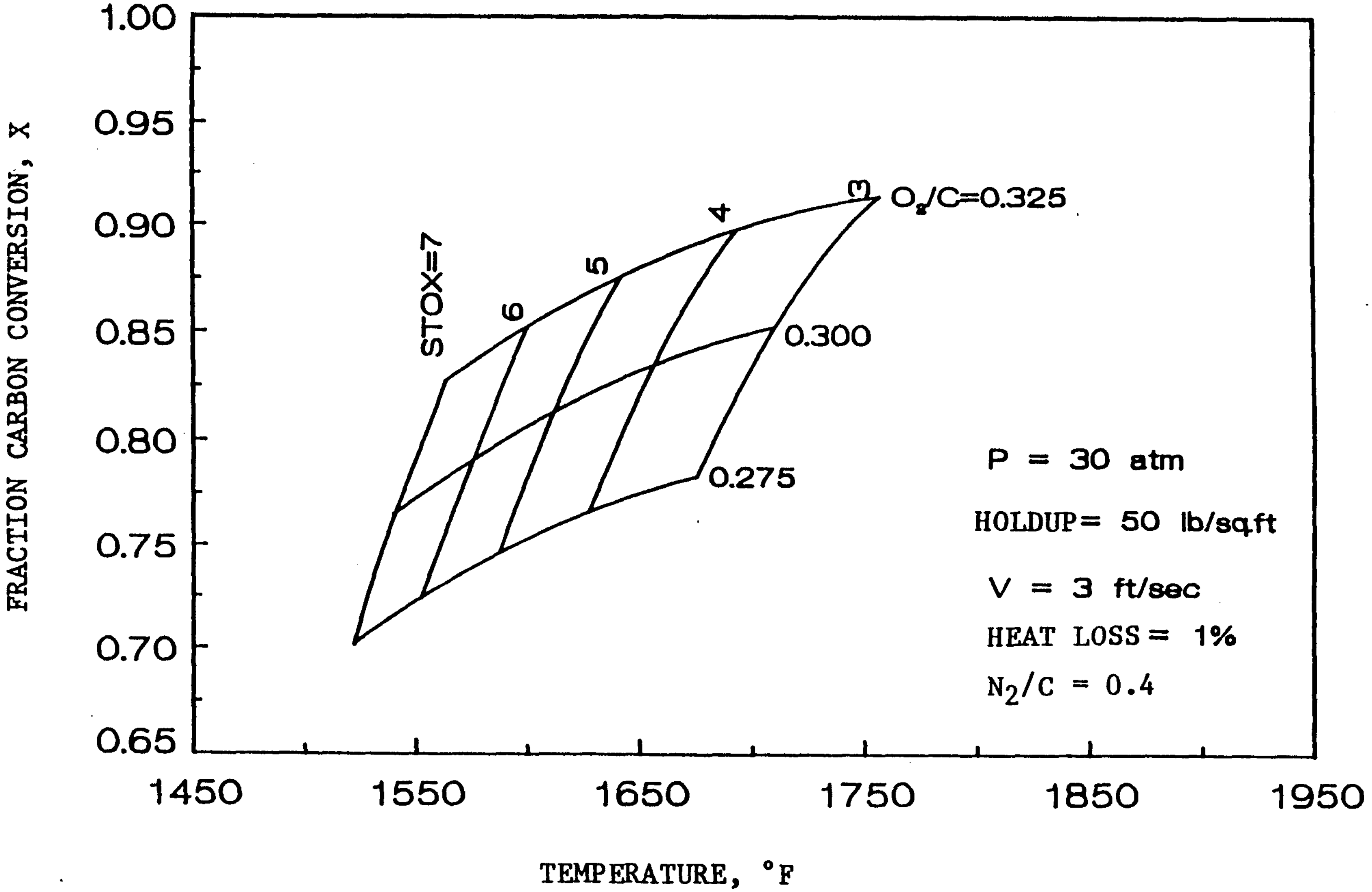


FIGURE 4

OPERATING MAP FOR PDU GASIFIER ON MONTANA ROSEBUD COAL



Consequently Figure 4 represents the operating map of the IGT PDU gasifier when operating on Montana Rosebud Subbituminous coal. To further improve on the simulation of PDU operation, nitrogen was added at a rate equal to 0.4 times the carbon feed in the construction of Figure 4. Figure 4 shows that even with a steam to oxygen ratio as low as 4.0, and an oxygen to carbon ratio as high as 0.325, it is not possible for the PDU gasifier to achieve a carbon conversion in excess of 90%. However, if the solids holdup is increased to 300 lbs/ft^2 as would be expected in a commercial IGT fluidized-bed gasifier, carbon conversion in excess of 95% can be achieved at an oxygen to carbon ratio of 0.3 and a steam to oxygen ratio of less than 3.0. This is shown on Figure 5 which represents the operating map for a commercial pressurized IGT gasifier on Montana Rosebud coal.

Figure 5 was constructed without the addition of nitrogen. Figure 5 shows that, if a steam to oxygen ratio of 2.0 is desired at an O_2/C ratio of 0.3 the bed temperature will be about 1775°F which is considerably higher than that reported for PDU operation. It is doubtful that operating at such a high temperature can be tolerated on account of excessive ash agglomeration and clinkering. The PDU data log reported operating difficulties when temperatures were raised close to the temperature required for operating the gasifier with low steam to oxygen ratio. For a Western coal the temperature for viable operation is lower, but is still about 1700 to 1750°F. This means that the pilot plant has to be able to operate at 1800° to 1850°F to provide a safety margin over clinkering temperature. The actual upper limit seems to be between 1650 to 1750°F. Clinkering is not just a function of gasifier operating temperature; it also depends on the mixing and the local mixing temperature in the fluidized bed. It is therefore a function of the nozzle design, nozzle size and steam to oxygen ratio entering the jet nozzle. The dependence on nozzle design creates difficult problems for the designer and for gasifier scaleup.

The same holds for North Dakota lignite for which a similar operating map has been constructed as shown in Figure 6. The operating map for North Dakota lignite clearly indicates that it has a larger economic window for carbon conversions over 95% at temperatures below 1700°F than Montana Rosebud.

FIGURE 5

OPERATING MAP FOR COMMERCIAL PRESSURIZED IGT GASIFIER ON MONTANA ROSEBUD COAL

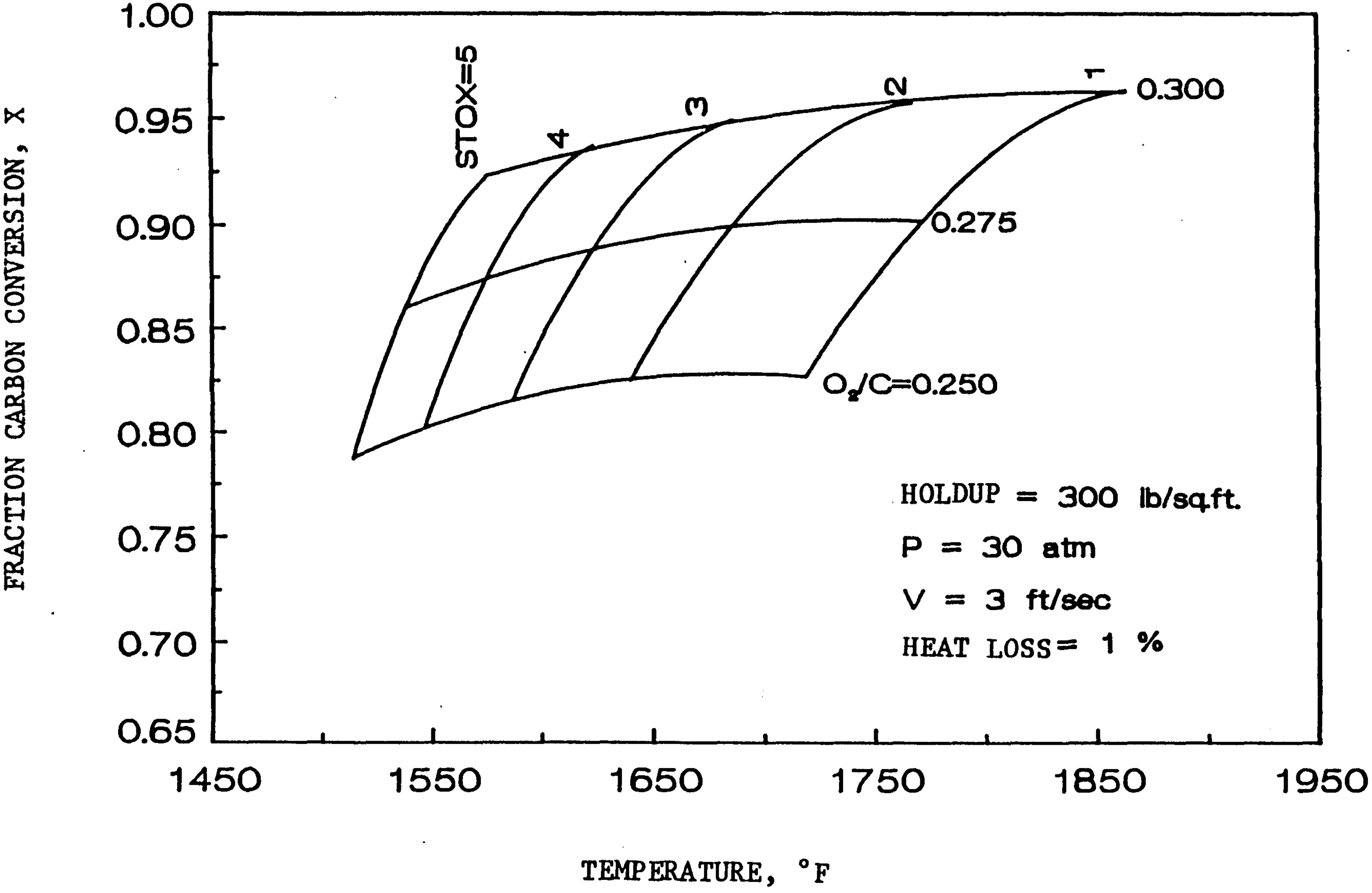
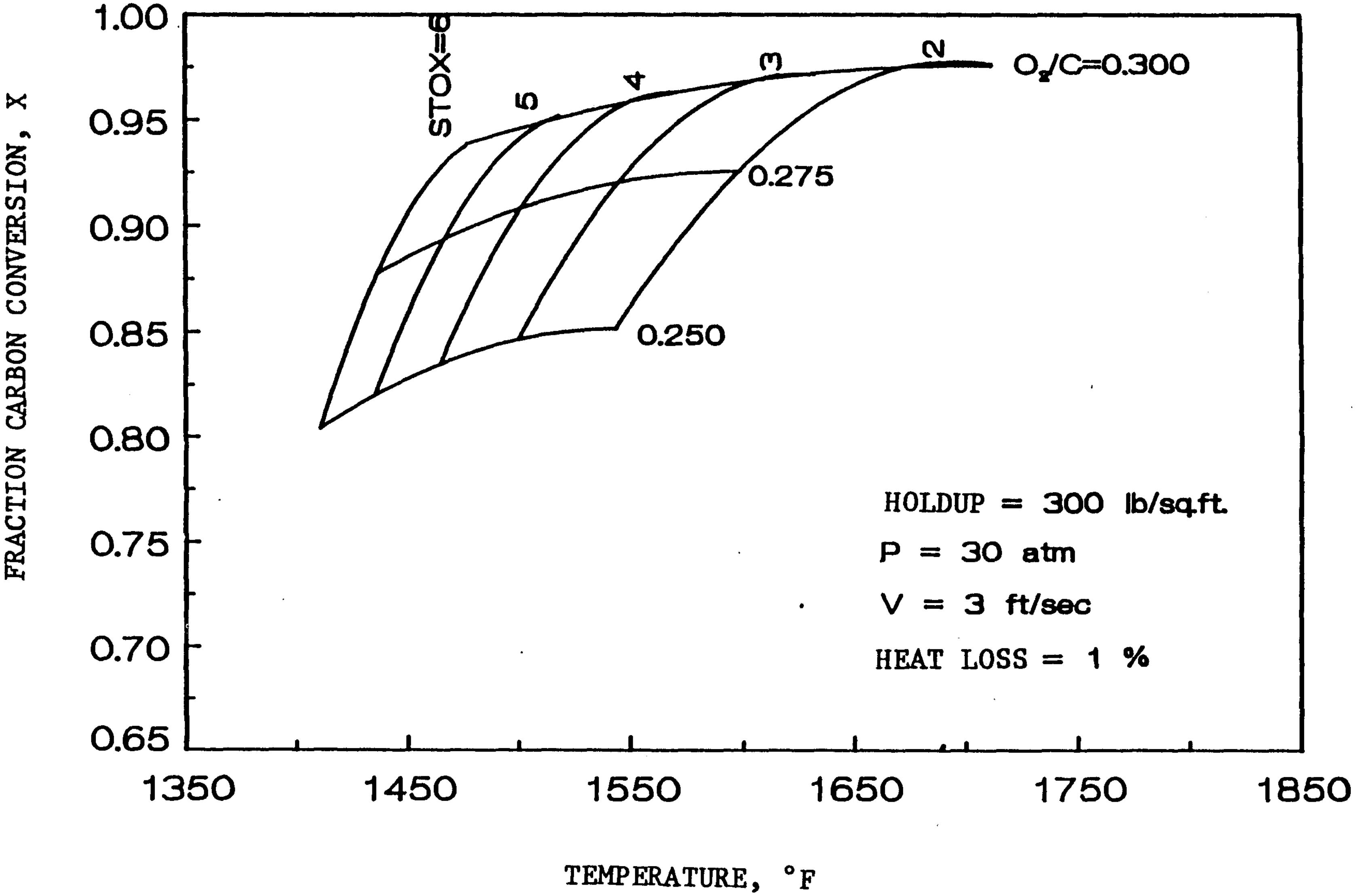


FIGURE 6

OPERATING MAP FOR COMMERCIAL PRESSURIZED IGT GASIFIER ON NORTH DAKOTA LIGNITE



5.0 APPLICABILITY OF PILOT PLANT DATA

The IGT PDU was not able to operate at high solids holdup because, at increased bed height, the gasifier slugged. Slugging is a diameter related phenomenon resulting from high L/D ratios. However, holdup can be significantly increased by reducing the linear velocity. This allows operation at higher bed heights without slugging. IGT claimed that the linear velocity could not be decreased because this caused problems in the bottom of the PDU gasifier. Holdup can also be increased by changing particle size distribution and the addition of fines will reduce slugging.

Even with lower holdup, the PDU could in theory be operated at significantly lower steam to oxygen ratios by increasing the coal feed rate. IGT claimed that the coal feeder was too small and it could not sustain the required coal feed rate in order to operate with a lower steam to oxygen ratio while keeping the linear velocity constant. The PDU data however do not support this argument. There are many runs where coal feed rates were kept very low and steam to oxygen ratio kept very high. The fact that the steam to oxygen ratio was kept high even though there was excess capacity in the coal feed rate for this run, confirmed the impression that IGT's main concern in PDU operation was to prevent clinkering by limiting the operating temperature.

Operation at conditions that will prevent clinkering is a very important requirement which cannot be determined using the CUNY kinetic model. Therefore, it is important to obtain data that will bracket the desirable and operable conditions from the PDU. Furthermore, the steam to oxygen ratios used for the PDU operation were so high, that the PDU was actually operated as a cooled combustor instead of a gasifier. Consequently, all the PDU operating results are in a region that is totally useless for estimating the performance of a commercial gasifier.

At high steam to oxygen ratios, carbon conversion is also suppressed. The reason is that the gasification reaction requires a minimum temperature in order to attain a reasonable rate. High steam to oxygen ratio tends to depress the temperature. Under this condition, heat evolved from oxidation to form CO is used to heat up the excess steam instead of driving the endothermic gasification reaction and therefore result in reduced conversion.

The kinetic relationships of a gasifier are really refined forms of statistical correlations. They allow prediction of results by interpolation, but cannot predict results reliably by extrapolation. It is impossible to predict kinetic performances for a high conversion case with a low oxygen to carbon ratio from data obtained at low conversions and with high oxygen to carbon ratio. For data to be useful for designing a gasifier, they have to cover a wide range of experimental conditions which bracket the design conditions.

It is sometimes claimed that if the pilot plant data can confirm the validity of the model, then the model can be used for extrapolative prediction. Such a claim is totally unrealistic. The CUNY kinetic model is no exception. It is only a learning tool which can be used to estimate the impact of various parameters on reactor performances, and allow a better understanding on how to design meaningful experiments. It cannot substitute for reliable data that are required for design.

There is more to the design of a gasifier than kinetics. The operating constraints due to clinkering and excessive agglomeration also play an important part. This means the designer must have more data on the feed nozzles, the internal mixing pattern, and the different fluidization behavior in order to properly address the gasifier design problem.

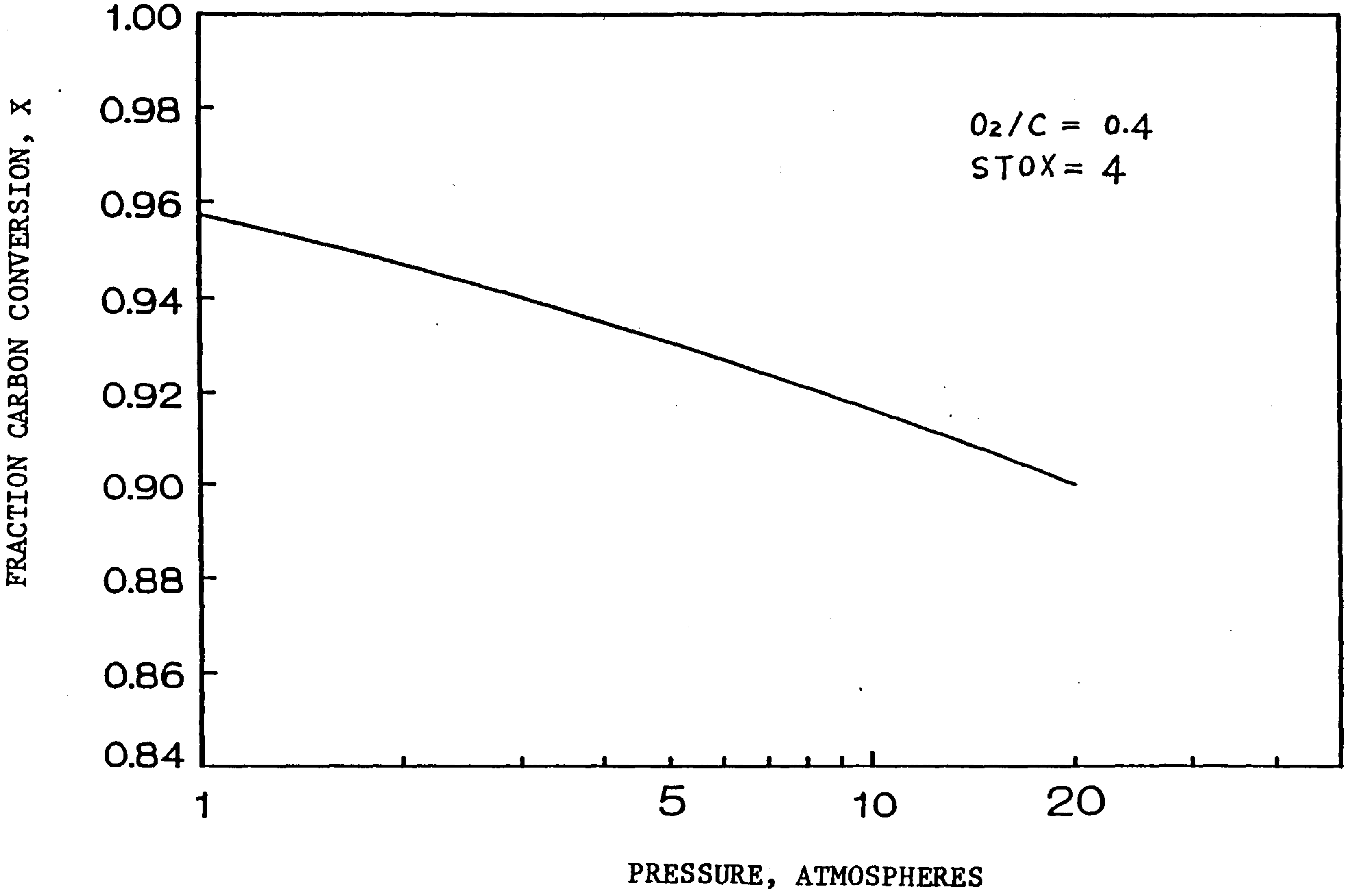
Winkler recognized the problem of requiring a high operation temperature and yet preventing clinkering. That is why the Winkler gasifier employs a hybrid design. The fluidized bed in the Winkler gasifier operates at low temperature to gasify only the active part of the coal, accomplishing only partial gasification. Char gasification in the Winkler gasifier occurs in the entrained flow section in a dilute phase. Part of the oxygen is added to the entrained flow section to raise the reaction temperature substantially.

IGT claimed to have data from the large U-gas pilot plant at acceptable steam to oxygen ratios and at much lower operating pressures. The conversion was however quite low, only about 85%. Furthermore, it is much easier to gasify coal at lower pressure. This is demonstrated on Figure 7 taken from reference (1). In Figure 7 carbon conversion is plotted as a function of gasification pressure for constant solids holdup and linear gas velocity and at fixed ratios of oxygen-to-carbon and steam-to-oxygen. As gasification pressure increases, the conversion decreases. The reason for this is that gas throughput, and therefore coal throughput, increases linearly with pressure, whereas reaction rate increases much slower. Also, the amount of heat generated per unit volume in the jet region increases linearly with pressure. However, the mixing intensity in the jet zone does not increase in the same way thereby making the rate of mixing in the jet even more critical.

The low holdup of the PDU gasifier and its operation at conditions outside the desirable steam to oxygen and oxygen to carbon ratios make it difficult to apply the results to draw any conclusions about the performance of a commercial gasifier, even if the data were consistent. The only conclusion that can be reached is that the IGT pressurized fluidized-bed gasifier has difficulty in operating at temperatures that will allow high carbon conversion.

FIGURE 7

EFFECT OF OPERATING PRESSURE ON CARBON CONVERSION



For the PDU data to be applicable for commercial design, the PDU operation must be at the same temperature, pressure, steam to oxygen ratio, holdup and gas velocity as the anticipated commercial plant. For scoping purposes, but not for design, it is permissible to reduce the linear velocity, as long as the coal feed rate per unit of solids holdup is kept constant. This latter criteria is critical since space velocity cannot be scaled.

An option which may render the IGT gasifier viable is to inject an additive or catalyst into the gasifier. The additive or catalyst must allow operation at higher temperatures without causing clinkering and excessive agglomeration. Calcium oxide is a likely candidate to fulfill this role. The addition of calcium oxide does have its disadvantages. It forms calcium sulfide which is rejected with ash. But calcium sulfide cannot be disposed of directly and requires oxidation to sulfate which is difficult. But even more importantly, it generates a large amount of solid waste which is costly and difficult to deal with. The ash becomes leachable and cannot be handled in the same way as fly ash.

It is more preferable to reject the sulfur as elemental sulfur. This means that either a more efficient additive or catalyst that is only required in smaller concentrations should be tried. In this respect, cellite or cheap natural zeolite are suggested additive for possible testing.

6.0 RECOMMENDATIONS FOR PDU MODIFICATIONS

It is recommended that the PDU design and its operating approach be modified if future data generation is contemplated.

- Future operation should be focused on finding suitable additives which will allow the gasifier to operate at either higher temperatures or higher gasification rates at lower temperature.
- The amount of nitrogen used for instrument tap purging must be reduced by a factor of at least four.
- Product gas flow, preferably after quench, must be accurately measured. This includes better measurement of the solid carbon content in the product gas. Continuous gas chromatographic analysis of the cold product gas is also required.
- Longer sustained runs at constant conditions are required not only for meaningful data, but also to prove the operability of the gasifier.
- Attempts should be made to quantitatively determine the safety factors which need to be applied to operating conditions in order to avoid agglomeration and clinkering.
- The gasifier should be modified to allow operation at much higher solids holdups. Alternatively, but less desirable, the gasifier should be operated with lower gas velocities. A small reduction of gas velocity, about 20 to 30% appears acceptable.
- The gasifier control and design should be changed to allow independent adjustment of bed height. The existing bottom design, featuring an air annules with classifying removal of large ash particles, makes bed height a function of ash agglomeration and does not permit independent control of bed height. A classifying removal design could be incorporated in a commercial plant along with a second removal device to control bed height.
- A larger pilot plant is needed, so that the effect of nozzle size and nozzle configuration in the bottom of the gasifier can be studied, if scale-up to a commercial design is to be seriously considered.

7.0 LITERATURE REFERENCES

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