PRESSURIZED HYDROGASIFICATION OF RAW COAL IN A DILUTE-PHASE REACTOR

Herman F. Feldmann, Joseph A. Mima and Paul M. Yavorsky

Pittsburgh Energy Research Center, U.S. Department of the Interior, Bureau of Mines, 4800 Forbes Avenue, Pittsburgh, Pa. 15213

INTRODUCTION

The direct conversion of raw coal to methane can be achieved by reacting the coal with hydrogen. This approach to converting coal to methane is the basis of the U.S. Bureau of Mines Hydrane Process[1] and the high thermal efficiency resulting from this direct process approach offers substantial potential economic advantages over other routes to pipeline gas.[2] Converting coal directly to methane by reaction with hydrogen was first reported by Dent.[3] The thermodynamic advantages of producing methane directly, rather than by first converting the coal to synthesis gas which is then converted to methane after water-gas shift and methanation, were quickly recognized when investigators in the United States started exploratory work on converting coal to pipeline gas. For example, Channabasappa and Linden[4] concluded that hydrogenating coal to methane with hydrogen which is produced by steam-oxygen gasification of carbon is more thermally efficient than steam-oxygen gasification followed by methanation. However, at first the experimental difficulties in directly hydrogenating raw coal to methane proved to be extreme because of the severe agglomerating properties of most American coals in high-temperature, high-pressure hydrogen atmospheres. This agglomeration problem caused a shift away from attempting to directly hydrogasify raw coal. Instead, the coal was "pretreated" with air or oxygen to destroy its agglomerating properties. While the mild oxidation with air or oxygen was successful in preventing the coal from agglomerating, it adversely reduced its reactivity for methane formation. In fact, the reactivity of the pretreated coal is so reduced that it is impossible to produce a gas by direct hydrogenation of pretreated coal that has a sufficiently high concentration of methane to allow its introduction into a pipeline without costly physical separation of the hydrogen-methane mixture. Thus, while the thermodynamic and chemical advantages of direct hydrogasification of raw coal were clear, the practical difficulties encountered in developing reactor systems to utilize raw coal required that the coal be pretreated before being contacted with hydrogen, but this reduced the process efficiency. Results of directly hydrogasifying pretreated coals in continuous reactors were reported by Institute of gas Technology investigators.[5,6,7]

The problem of processing agglomerating raw coal was solved at the Bureau of Mines when a technique was developed for directly hydrogenating raw coal in a free-fall dilute-phase (FDP) reactor described by Hiteshue.[8] Some results of FDP reactor experiments using raw coal[9,10] have already been presented. However, the results in references 9 and 10 were obtained for rather high pressures of 1,500 and 3,000 psig. Even though operation with raw bituminous coal at 3,000 psig does allow the direct production of raw product gases containing over 80 percent methane and having carbon monoxide contents down to 0.1 vol pct, design considerations indicate pressures around 1,000 psig are more economically attractive. This paper therefore summarizes our FDP reactor data at pressures in the neighborhood of 1,000 psig. These data are useful for the design of the FDP section of the Hydrane process or other processes using similar conditions.

EXPERIMENTAL

Equipment and Procedure

Details of the experimental reactor system and method of operation are given in reference 10. Briefly, the FDP reactor is simply a 3-inch id heated tube contained in a 10-inch pressure vessel. The coal is injected into the top of the 3-inch reactor through a water-cooled nozzle. The coal falls freely through the reactor tube concurrent with the reacting gas which is also injected in at the top of the reactor. Because of rapid heating and a dilute solids phase, agglomeration was avoided; particles were plastic and sticky for only a short time during which particle-particle collisions are few. The heated length of the reactor for all except two of the experiments presented here is 5 ft. The residence time of the coal in the reacting zone is simply the reactor length divided by the average terminal velocity of the coal particles. The char produced in the FDP reactor is collected in a cooled hopper and analyzed after a run. Gas flow rates and compositions are measured over the steady state portion of the run. Ordinarily the capacity of the pressurized char collector allowed a run duration of approximately one hour with approximately fifty minutes of steady state operations during which data could be collected.

Tabular Results

Results of our most recent FDP reactor operations are summarized in table 1 and the analyses of the feed coals used are listed in table 2. The main objectives of these experiments were to: (1) establish the feasibility of directly producing a high-Btu gas by hydrogasifying raw coal in a continuous reactor at economically attractive pressures, (2) measure the yields and distribution of coal hydrogasification reaction products, and (3) provide data for scaling-up the FDP reactor.

Production of High-Btu Gas

The feasibility of producing a gas having a heating value of 900 or more Btu/scf (after cleanup methanation) was established by several experiments. These experiments were designed to simulate the operation of an integrated Hydrane reactor which consists of two stages.[1] In such an integrated reactor, the hydrogen is first fed to a fluid bed where it reacts with char produced by the FDP reactor. The product gas from this fluid bed is the feed gas to the top of the FDP reactor and it consists of about 50 vol pct methane with the remainder hydrogen plus a small amount of carbon monoxide. Thus, the composition of the feed gas to the experimental isolated FDP reactor was adjusted to simulate the fluid-bed product gas from an integrated operation.

Table 3 compares carbon conversion, gas composition, and gas yields for specific experiments with the results used to guide an economic evaluation of the Hydrane process.[2] These results show that the goal of producing a high-Btu gas can be achieved at pressures of 1,000 psig and higher. For all three coals evaluated, Pittsburgh Seam hvab coal, Illinois No. 6 hvcb coal, and lignite, it was shown that the gas produced after methanation to reduce CO to an acceptable level was substitutable for natural gas. Of course, with lignite, the higher oxygen content results in higher yields of CO and this gas will therefore require more methanation than the product gas from the Illinois or Pittsburgh Seam bituminous coals. However, even with lignite the fraction of the total methane that is produced directly rather than by methanation is greater than can be achieved by other process routes using bituminous coal. In an actual plant where hydrogen is produced from the residual char, the catalytic water-gas shift reaction ($CO + H_2O \rightarrow H_2 + CO$) would not be carried to completion. Instead, as the base case analysis of the feed gas to the dilute phase indicates, some CO would be

left in the hydrogen resulting in a somewhat higher CO concentration in the raw product gas from the FDP reactor. This additional CO would result in increased hydrogen consumption during methanation and thereby lower the hydrogen content and increase the heating value of the final product gas. The heating value of the final product gas was therefore calculated based on a constant 4 vol pct CO in the raw dry product gas from the FDP reactor.

Examination of these FDP results indicate the controllable parameters that determine whether the raw product gas will, after methanation of the 4 vol pct CO, have a heating value of at least 900 Btu/scf are the following:

a. Percent methane in the feed gas to the FDP reactor
b. Gas-to-coal feed ratio
c. Reactor pressure
d. Coal residence time (reactor length)

Fortunately, the combination of above variables required to produce 900+ Btu gas are easily achievable in practical reactor systems. For example, the reactor can operate at gas transmission line pressures, both the gas/coal feed ratio and the methane concentration in the feed gas allow operation at carbon conversion levels resulting in balanced plant operation,* and a sufficient coal residence time was achieved in an FDP reactor only 5 ft long. Increases in the percent methane in the feed gas to the FDP reactor, at constant CO concentration, will further increase the heating value of the final product gas.

In our experiments, the reactor pressure shell was pressurized with the feed gas. Because this gas was in direct contact with the reactor electrical heating elements, carbon deposition from methane cracking and subsequent shorting of the electrical elements became a problem when the methane concentration in the feed gas was higher than 50 vol pct. This artificial limitation would not exist in a commercial reactor system where no electric heating elements would be used. Methane concentrations above 50 vol pct would be generated by the fluidized-bed stage of the Hydrane reactor system and therefore allow higher methane concentrations into the FDP reactor. As examples, Pyrcioch and Linden[5] in studying the fluid-bed hydrogasification of a char produced by low temperature pretreatment achieved methane concentration over 50 vol pct, and Lewis and co-workers[9] reported methane concentrations over 60 vol pct from direct moving-bed hydrogasification of chars produced by hydrogasification in an FDP reactor. Thus, the results presented here must be regarded as conservative and higher methane concentration product gases could be produced in commercial FDP reactors where the methane concentration in the intermediate feed gas to the FDP reactor is not limited by artificial constraints.

Product Yields and Distribution

The major products from the dilute-phase hydrogasification of raw coal are gas and char plus smaller amounts of organic liquid products and water. The liquid yield measurement is rather inaccurate because of the relatively small amount of liquids formed and the difficulty of their quantitative recovery. For all the experiments reported in table 1, measured yields of organic liquids varied from less than .01 to .06 lb/lb coal. Attempts to correlate the organic liquid yields with reactor parameters thought to have the greatest effect on these yields, such as reactor wall temperature, hydrogen partial pressure, and gas phase residence time, have been unsucessful. Since, in a base-load pipeline gas plant the organic liquid production will be substantial even at the lowest yields measured, additional work is now going on to characterize these organic liquids.

*Balanced operation means the overall plant produces no surplus char. To achieve balanced operation, the fraction of carbon in the coal converted to methane is regulated so that the remaining carbon is just sufficient to produce process hydrogen and plant power.

Water is produced by both the simple vaporization of moisture in the coal and by the reaction of hydrogen with oxygen in the coal. Recoveries of water both from condensers and as moisture on the char ranged from .01 to .08 lb/lb of coal fed. However, as table 1 indicates, the water recoveries did not exhibit the same degree of fluctuation as did the organic liquid recoveries and most of the measured yields were between .05 and .06 lb/lb coal fed. The water recoveries measured for the Illinois #6 hvcb coal and the single run made with lignite were higher because of the higher oxygen contents of these feeds. For the Illinois coal, water yields varied from about .05 to .09 lb/lb coal. These water yield data indicate that much of the oxygen is either present as bound water or combines with hydrogen to form water during hydrogasification.

Char Yields and Desulfurization

In the overall Hydrane process, the char from the FDP reactor will be further converted in a fluid-bed reactor which is in series with the FDP reactor. The yield of char from the FDP reactor depends on the carbon conversion level as shown in figure 1.

Sulfur is eliminated from the char during hydrogasification as H_2S and the degree of elimination is related to the carbon conversion as shown in figure 2. The scatter may be due to either the influence of parameters other than carbon conversion and/or to inaccuracies in the sulfur determinations. The important point to be demonstrated is that the coal sulfur is extremely reactive under hydrogasification conditions as evidenced by the fact that the coal residence time in the FDP reactor is on the order of one to two seconds. In fact, figure 2 indicates the sulfur in the coal to be more reactive than the carbon in the coal. The Pittsburgh Seam hvab coal contains approximately equal amounts of pyritic and organic sulfur. However, as yet no tests have been made on the char to indicate whether either type is selectively removed during free-fall hydrogasification. Since, in the integrated Hydrane process the char spends additional residence time in a fluid bed at higher hydrogen partial pressures than exist in the FDP reactor, additional char desulfurization will occur in the fluid bed. In preliminary experiments with an integrated FDP fluid-bed reactor system, the sulfur removal from the Pittsburgh Seam coal has been on the order of 85 percent. These results are encouraging because they indicate that char from the Hydrane reactor may be an acceptable fuel to provide the plant's energy and steam requirements without complicated sulfur removal systems and without exceeding air quality restrictions on atmospheric release of sulfur compounds.

Scale-Up of the FDP Reactor

The FDP reactor has two important functions. It must convert the coal to a nonagglomerating char for the subsequent fluid bed and it must convert enough carbon to methane so that the FDP product gas is, after acid gas removal and light methanation, an acceptable pipeline gas.

In the 3-inch id FDP reactor used in our experiments, the coal particles are heated to reaction temperature in the reactor by mixing with the preheated feed gas and by heat transfer from the hot walls of the FDP reactor. However, heat transfer analysis of larger reactors[11] indicates that as one increases the reactor diameter, the amount of heat transfer from hot reactor walls to the particles inside becomes negligible. Therefore, in larger diameter reactors, the coal particles can be raised to reaction temperature only by mixing with the hot methane-hydrogen mixture shunted from the fluid bed. Calculations indicate that the mixing temperature of the hot gas and coal at the top of a large FDP reactor will be on the order of 480° to 540° C. It is therefore important to evaluate the FDP hydrogasification behavior at these

relatively low temperatures. This is difficult to do in the present 3-inch id FDP reactor because the coal quickly heats to the reactor wall temperature and, if the wall temperature is below 725° to 800° C, the coal adheres to the reactor walls and eventually plugs the reactor. In spite of this drawback, experiments were conducted at reduced reactor wall temperatures of 485°, 530°, and 580° C. In these experiments, coal hitting the reactor walls stuck to them and eventually plugged the reactor. However, the coal that did not contact the walls, passed through the reactor and was collected and its conversion and caking properties determined. Results of these lower reactor wall temperature experiments are shown in figures 3 and 4 where the effect of temperature on both the volatile matter and the carbon conversion of the FDP reactor char are shown at the reduced wall temperatures. Figure 3 shows an interior thermocouple temperature at the reactor bottom and figure 4 shows the average reactor wall temperature because the actual particle temperature is not known. The true average particle temperature is probably between the interior thermocouple temperature and the reactor wall temperature. Also shown in figures 3 and 4 is the temperature boundary above which the char is not agglomerating when tested in a fluid bed with hydrogen at 1,000 psig and 900° C. Chars produced at reactor wall temperatures below the boundary temperatures agglomerated when tested at the above conditions. Thus, if one conservatively assumes that the particle temperature is close to the wall temperature, it appears that simply mixing the hot methane-hydrogen mixture produced in the fluid-bed reactor with the coal at the top of the FDP reactor will produce an acceptable non-agglomerating char if the char temperature reaches 580° C even if for residence times of only a second or two.

Previous reaction rate analyses of FDP reactor data[10] at higher hydrogen partial pressures (1,500 to 3,000 psig) and at reactor wall temperatures of 725° and 900° C indicated that the conversion of carbon in raw coal occurs in three stages with each stage having greatly different reactivities toward hydrogen. At the short (one to two seconds) particle residence times in the FDP reactor at 3,000 psig and with wall temperatures of 900° C, all of the first stage (the most reactive) carbon behaved as if it were converted "instantaneously". However, the conversion of the second stage carbon varied with reactor conditions and this variation of rate with reactor conditions was correlated by this simple rate equation:

$$(1) \qquad U_T \frac{dx}{dL} = kp_{H_2}(1-x)$$

where U_T is the average terminal velocity of the particles, k is the rate constant, p_{H_2} the partial pressure of hydrogen; L the location in the reactor (the distance from the coal inlet), and x the carbon conversion level. The fraction of carbon that behaved as though it were instantaneously converted was denoted by E and was determined by finding the value that allowed the best fit of experimental data with the integrated form of equation (1):

$$(2) \qquad \int_E^x \frac{dx}{p_{H_2}(1-x)} = \frac{k}{U_T} L$$

At conditions where the conversion of second stage carbon, x-E, was small, errors or variations in either x or E caused large fluctuations in the value of k, making a kinetic analysis of the data difficult. This occurs where the total carbon conversion is mostly due to the "instantaneous" carbon reaction because conditions are not severe enough to react a substantial fraction of the second stage carbon. As an example of

"severe conditions", operation at 3,000 psig, 900° C wall temperatures and with pure hydrogen feed gas allowed total carbon conversions, x, ranging from 0.40 to 0.50 in the FDP reactor. At these conditions the value of E is approximately 0.14 and the spread between x and E is sufficiently large to allow a reasonable determination of k. For the experiments reported in reference 10, E varied from about 0.15 to 0.20.

At the conditions reported in this paper where the total pressure is closer to 1,000 psig, and the feed gas to the FDP reactor is an approximately equimolar mixture of hydrogen and methane, the total carbon conversions are closer to the fraction of carbon that instantaneously reacts and kinetic interpretation is even more difficult. Therefore, the kinetic analysis is not yet complete. However, for the purposes of FDP reactor simulation, a mathematical model was used that assumed all the carbon reacts at a rate dictated by the equation,

$$(3) \quad \frac{dx}{dL} = \frac{k}{U_T} p_{H_2} (1-x)$$

This equation is felt to be conservative because it does not allow for the fraction of carbon that may react at a considerably faster rate than the final amount of carbon conversion which was used to evaluate the rate constant k. The temperature dependency of k used for our initial reactor simulation studies[11] has been reported.[1] While the more detailed kinetic analysis may result in a modified rate equation, the results of our simulation study[11] indicate that radiant heat transfer plays a dominant role in small FDP reactors such as the one used in this study. Due to the diminishing effect of radiant heat transfer from the reactor walls as the diameter of the reactor increases, temperature profiles in commercial reactors will be considerably different than those existing in our present 3-inch id FDP reactor thus indicating the necessity of using larger diameter pilot plants to obtain reliable scale-up data.

CONCLUSIONS

The feasibility of producing on the order of 95 percent of the total methane production by the direct reaction of raw caking coals with hydrogen in an FDP plus fluid-bed reactor system has been established. The experimental results show that the product gas is of pipeline quality after minor methanation of the small amount of carbon monoxide (only 4 percent) without any need for hydrogen separation.

Further, the experiments indicate the above desirable results can be achieved without encountering agglomeration problems even when unpretreated highly caking coals are used.

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TABLE 1.- Operating data for FDP hydrogasification of raw coal
Feed coal is 50 x 100 mesh except where noted.

Test No., IHR-	146	147	149	151	153	154
Temperature, ° C	900	900	900	900	900	900
Pressure, psig	1000	1000	1200	1100	1100	2000
Coal	hvab	hvab	hvab	hvab	hvab	hvab
Coal rate, lb/hr	12.17	12.44	12.38	11.88	10.92	12.51
Feed gas rate, scfh	153.5	155.2	161.0	158.6	150.6	164.5
Vol %, Hydrogen	50.5	56.0	53.0	48.0	99.2	52.5
do. Methane	41.9	42.3	44.5	49.2	0.2	46.4
do. Nitrogen	6.3	1.8	2.5	2.6	0.6	1.0
Total scf/lb	12.61	11.99	13.00	13.35	13.82	13.15
Hydrogen, scfh	77.5	86.9	85.3	76.1	149.4	86.3
Hydrogen, scf/lb	6.37	6.72	6.89	6.41	13.71	6.90
Product gas, scfh	169.6	171.8	175.5	167.1	143.8	175.0
Vol %, Hydrogen	22.7	25.5	23.6	22.1	49.0	19.8
do. Methane	66.4	67.8	69.7	71.7	46.5	75.9
do. Ethane	0.3	0.4	0.3	0.1	0.3	0.1
do. Carbon monoxide	2.7	3.2	3.0	2.5	3.4	2.2
do. Carbon dioxide	1.0	1.1	0.8	0.5	0.2	0.4
do. Nitrogen	6.4	1.7	2.3	2.7	0.5	1.5
Product yield, Methane	3.97	3.93	4.09	3.52	6.10	4.52
Scf/lb, Ethane	0.04	0.05	0.04	0.01	0.04	0.01
do. CO	0.38	0.43	0.43	0.35	0.45	0.31
do. CO_2	0.00	0.15	0.11	0.04	0.03	0.06
Feed H_2 reacted, scf/lb	3.21	3.33	3.55	2.46	7.23	4.13
Char residue, lb/lb	0.697	0.702	0.698	0.698	0.663	0.694
Condensed liquid, lb/lb						
Water	0.051	0.037	0.033	0.029	0.032	0.038
Oil	0.013	0.009	0.014	0.008	0.005	0.008
Residue moisture, lb/lb	0.009	0.012	0.011	0.008	0.003	0.018
Conversion, maf coal	32.5	32.5	33.1	32.8	36.5	33.0
(Wt pct), Carbon	25.6	25.0	25.5	25.3	28.5	25.1
do. Hydrogen	64.2	66.6	66.4	65.2	70.0	66.6
do. Sulfur	48.8	44.3	43.9	46.5	55.9	46.8
do. Nitrogen	25.1	24.6	27.2	30.6	38.0	26.4
Recovery, overall	96.3	96.0	96.3	93.2	95.7	96.2
(Wt pct) Carbon	94.6	96.3	96.1	94.4	99.0	97.2
do. Hydrogen	98.9	95.2	97.1	92.4	94.5	97.3
do. Ash	100.2	100.2	104.0	99.4	103.4	92.4

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TABLE 1.- (Continued)

Test No., IHR-	156	157	158	160	165	176*
Temperature, ° C	850	850	900	900	850	850
Pressure, psig	1000	2000	2000	1500	1500	1000
Coal	hvab					hvab
Coal rate, lb/hr	12.84	13.00	12.61	12.29	12.94	12.47
Feed gas rate, scfh	157.8	161.7	160.9	161.3	158.8	156.2
Vol %, Hydrogen	49.0	49.9	51.8	53.8	51.3	48.0
do. Methane	49.4	48.4	46.6	43.4	47.0	49.4
do. Nitrogen	1.6	1.7	1.6	2.6	1.7	2.2
Total scf/lb	12.29	12.44	12.76	13.12	12.27	12.53
Hydrogen, scfh	77.3	80.7	82.9	86.8	81.5	75.0
Hydrogen, scf/lb	6.02	6.21	6.61	7.06	6.30	6.24
Product gas, scfh	171.8	180.8	177.0	166.4	173.8	173.1
Vol %, Hydrogen	22.4	18.1	18.0	19.7	21.7	22.8
do. Methane	71.4	79.0	78.7	75.2	73.4	71.6
do. Ethane	0.5	0.1	0.1	0.3	0.1	0.2
do. Carbon monoxide	3.2	0.5	1.1	1.4	2.1	2.5
do. Carbon dioxide	0.7	0.4	0.4	0.8	0.6	0.7
do. Nitrogen	1.4	1.7	1.6	2.4	2.0	1.9
Product yield, Methane	3.48	4.97	5.10	4.49	4.13	3.70
Scf/lb, Ethane	0.06	0.01	0.01	0.04	0.01	0.03
do. CO	0.43	0.07	0.15	0.19	0.28	0.35
do. CO_2	0.09	0.06	0.06	0.08	0.08	0.10
Feed H_2 reacted, scf/lb	3.02	3.69	4.08	4.39	3.38	2.85
Char residue, lb/lb	0.700	0.658	0.703	0.697	0.696	0.706
Condensed liquid, lb/lb						
Water	0.036	0.042	0.050	0.049	0.058	0.048
Oil	0.018	0.015	0.004	0.005	0.012	0.022
Residue moisture, lb/lb	0.015	0.023	0.014	0.019	0.017	0.008
Conversion, maf coal	32.8	37.3	32.2	33.7	33.1	32.2
(Wt pct), Carbon	25.0	30.0	25.0	24.2	23.3	24.0
do. Hydrogen	62.4	65.9	68.0	66.0	62.4	63.5
do. Sulfur	30.0	40.2	45.4	49.2	46.4	45.9
do. Nitrogen	21.3	34.2	29.7	30.6	25.8	24.7
Recovery, overall	95.1	94.1	98.1	95.8	96.8	96.9
(Wt pct), Carbon	95.1	94.1	97.9	96.7	97.6	98.5
do. Hydrogen	95.4	102.0	100.6	96.9	100.9	98.5
do. Ash	106.8	102.2	104.3	101.8	96.9	98.0

*Feed coal particle size range is 100 x 200 mesh.

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TABLE 1.- (Continued)

Test No., IHR-	166	167	172	173	174	177
Temperature, ° C	850	800	850	900*	850*	850
Pressure, psig	1200	1000	2000	1000	1000	1000
Coal	hvab					hvab
Coal rate, lb/hr	12.68	13.21	12.40	12.61	12.86	11.70
Feed gas rate, scfh	156.3	153.1	163.4	149.1	156.8	155.3
Vol %, Hydrogen	49.2	48.4	50.6	52.7	49.6	99.3
do. Methane	48.7	48.2	46.4	45.2	47.0	0.4
do. Nitrogen	2.1	3.3	3.0	2.1	3.3	0.3
Total scf/lb	12.33	11.60	13.18	11.82	12.19	13.18
Hydrogen, scfh	76.9	74.1	82.7	78.6	77.8	154.2
Hydrogen, scf/lb	6.07	5.62	6.11	6.23	6.05	13.11
Product gas, scfh	168.7	170.3	171.9	163.0	174.1	150.9
Vol %, Hydrogen	22.7	27.3	20.4	25.0	27.9	52.9
do. Methane	72.2	67.0	75.8	70.4	66.4	43.6
do. Ethane	0.1	1.0	0.1	0.2	0.8	0.2
do. Carbon monoxide	2.3	1.2	0.7	1.6	1.4	2.1
do. Carbon dioxide	0.5	0.2	0.2	0.4	0.6	0.3
do. Nitrogen	2.3	3.1	2.6	2.1	2.6	0.5
Product yield, Methane	3.60	3.05	4.39	3.76	3.26	5.57
Scf/lb, Ethane	0.01	0.13	0.01	0.03	0.11	0.03
do. CO	0.31	0.15	0.10	0.21	0.19	0.27
do. CO_2	0.07	0.03	0.03	0.05	0.08	0.04
Feed H_2 reacted, scf/lb	3.04	2.10	3.84	3.00	2.27	7.21
Char residue, lb/lb	0.691	0.709	0.674	0.721	0.692	0.646
Condensed liquid, lb/lb						
Water	0.049	0.041	0.053	0.037	0.041	0.042
Oil	0.018	0.029	0.012	0.029	0.030	0.018
Residue moisture, lb/lb	0.019	0.014	0.015	0.018	0.019	0.014
Conversion, maf coal	34.8	30.5	35.0	31.4	33.7	38.2
(Wt pct), Carbon	25.6	25.0	28.0	21.4	23.4	30.8
do. Hydrogen	63.5	59.1	64.3	62.4	59.1	67.1
do. Sulfur	42.8	52.7	44.9	61.8	42.1	45.9
do. Nitrogen	21.7	20.3	32.6	23.7	21.5	27.8
Recovery, overall	94.5	94.1	93.6	96.0	95.0	93.8
(Wt pct), Carbon	92.9	94.0	93.4	99.1	96.9	95.5
do. Hydrogen	97.3	98.9	99.3	98.5	100.8	97.2
do. Ash	110.1	100.7	104.1	99.4	98.0	99.2

* Reactor length - 3 ft.

(Continued on next page)

TABLE 1. (Continued)

Test No., IHR-	178	180	181*	182	183*	186
Temperature, °C	800	900	900	900	850	900
Pressure, psig	1000	1000	1000	1000	1000	500
Coal	hvab					hvab
Coal rate, lb/hr	11.70	12.53	12.72	24.10	3.94	6.73
Feed gas rate, scfh	161.1	441.2	168.0	323.1	151.4	70.0
Vol %, Hydrogen	98.9	99.3	99.2	52.6	51.6	99.0
do. Methane	0.4	0.5	0.5	45.7	46.5	0.4
do. Nitrogen	0.4	0.2	0.3	1.7	1.9	0.6
Total scf/lb	13.77	35.21	13.20	13.41	38.43	10.40
Hydrogen, scfh	159.0	438.1	166.7	170.0	78.1	69.3
Hydrogen, scf/lb	13.59	34.96	13.11	7.05	19.83	10.30
Product gas, scfh	147.5	412.6	155.1	335.0	152.1	76.8
Vol %, Hydrogen	73.0	76.3	44.2	28.5	39.8	50.9
do. Methane	23.9	21.9	50.2	66.9	57.3	44.1
do. Ethane	0.7	trace	0.0	0.2	trace	0.0
do. Carbon monoxide	1.5	1.4	4.7	2.1	1.0	4.2
do. Carbon dioxide	0.2	0.0	0.2	0.3	0.1	0.0
do. Nitrogen	0.4	0.2	0.4	1.7	1.7	0.5
Product yield, Methane	2.96	7.04	6.06	3.17	4.25	5.12
Scf/lb, Ethane	0.09	trace	0.00	0.03	0.0	0.0
do. CO	0.19	0.46	0.57	0.29	0.38	0.47
do. CO_2	0.00	0.00	0.04	0.04	0.04	0.0
Feed H_2 reacted, scf/lb	4.38	9.84	7.71	3.09	4.46	4.59
Char residue, lb/lb	0.705	0.648	0.630	0.708	0.602	0.618
Condensed liquid, lb/lb						
Water	0.039	0.028	0.042	0.038	0.004	0.036
Oil	0.029	0.010	0.007	0.019	0.054	0.017
Residue moisture, lb/lb	0.011	0.013	0.015	0.014	0.010	0.006
Conversion, maf coal	34.7	37.7	40.5	32.8	43.0	40.5
(Wt pct), Carbon	28.1	31.6	33.2	26.0	36.2	33.4
do. Hydrogen	61.0	71.3	71.8	61.5	71.2	74.4
do. Sulfur	48.4	52.5	56.7	49.1	61.6	64.0
do. Nitrogen	21.9	38.1	41.6	25.3	43.7	46.0
Recovery, overall	81.8	96.6	95.9	92.6	91.9	90.6
(Wt pct), Carbon	88.8	98.6	95.2	92.5	90.2	91.0
do. Hydrogen	86.3	99.6	92.9	94.3	97.2	95.5
do. Ash	112.6	97.7	99.4	107.7	101.8	99.6

*Feed coal particle size range is 100 x 200 mesh.

(Continued on next page)

TABLE 1.- (Continued)

Test No., IHR-	189	190	184
Temperature, ° C	850	850	850
Pressure, psig	1000	1000	1000
Coal	hvab	hvab	lignite
Coal rate, lb/hr	12.94	13.01	13.23
Feed gas rate, scfh	166.9	165.1	162.6
Vol %, Hydrogen	52.4	49.4	50.1
do. Methane	44.0	46.0	47.8
do. Nitrogen	3.4	4.3	1.8
Total scf/lb	12.90	12.69	12.29
Hydrogen, scfh	87.5	81.6	81.5
Hydrogen, scf/lb	6.76	6.27	6.16
Product gas, scfh	191.4	184.2	204.1
Vol %, Hydrogen	25.1	23.3	27.9
do. Methane	68.0	69.2	57.5
do. Ethane	0.2	0.2	0.1
do. Carbon monoxide	2.7	2.6	6.3
do. Carbon dioxide	0.8	0.6	5.9
do. Nitrogen	3.0	3.9	2.1
Product yield, Methane	4.38	3.98	3.25
Scf/lb, Ethane	0.03	0.03	0.02
do. CO	0.37	0.34	1.03
do. CO_2	0.12	0.07	0.99
Feed H_2 reacted, scf/lb	3.05	2.99	2.01
Char residue, lb/lb	0.714	0.714	0.494
Condensed liquid, lb/lb			
Water	0.008	0.033	0.102
Oil	0.020	0.017	0.018
Residue moisture, lb/lb	0.009	0.018	0.020
Conversion, maf coal	33.1	30.6	50.7
(Wt pct), Carbon	23.9	22.0	32.1
do. Hydrogen	66.4	60.9	77.4
do. Sulfur	47.5	38.6	51.0
do. Nitrogen	30.1	19.0	51.0
Recovery, overall	95.7	96.7	93.5
(Wt pct), Carbon	98.8	96.9	97.6
do. Hydrogen	98.7	106.9	97.0
do. Ash	109.1	89.6	101.8

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TABLE 1.- (Continued)

Test No., IHR-	161	162	163	164	191	192
Temperature, ° C	900	900	900	900	725	650
Pressure, psig	1000	1500	2000	1200	1000	1000
Coal	Ill.#6	Ill.#6	Ill.#6	Ill.#6	Ill.#6	Ill.#6
Coal rate, lb/hr	10.53	12.31	12.77	11.78	12.19	12.04
Feed gas rate, scfh	157.2	156.2	158.8	165.1	169.8	188.9
Vol %, Hydrogen	54.5	48.8	52.0	50.1	56.5	60.1
do. Methane	44.5	49.3	46.1	47.7	42.8	39.4
do. Nitrogen	1.0	1.8	1.9	2.2	0.7	0.5
Total scf/lb	14.93	12.69	12.94	14.02	13.93	15.69
Hydrogen, scfh	85.6	76.2	82.6	82.7	96.0	113.5
Hydrogen, scf/lb	8.14	6.19	6.73	7.02	7.87	9.43
Product gas, scfh	182.3	185.3	190.1	199.4	190.4	202.0
Vol %, Hydrogen	27.9	20.4	20.0	21.9	43.0	52.2
do. Methane	68.6	75.0	73.8	72.8	52.0	45.3
do. Ethane	trace	trace	0.2	trace	2.3	0.9
do. Carbon monoxide.	1.4	1.9	2.2	2.4	1.6	0.9
do. Carbon dioxide.	0.6	0.8	1.3	0.7	0.6	0.2
do. Nitrogen	1.4	1.7	2.3	1.9	0.4	0.4
Product yield, Methane	5.28	5.03	5.47	5.64	2.16	1.42
Scf/lb, Ethane	trace	trace	0.03	trace	0.36	0.15
do. CO	0.24	0.29	0.34	0.41	0.25	0.15
do. CO_2	0.10	0.11	0.20	0.12	0.09	0.03
Feed H_2 reacted, scf/lb	3.31	3.12	3.63	3.31	1.15	0.67
Char residue, lb/lb	0.622	0.658	0.663	0.653	0.702	0.782
Condensed liquid, lb/lb						
Water	0.079	0.068	0.048	0.063	0.043	0.038
Oil	0.032	0.011	0.006	0.010	0.062	0.035
Residue moisture, lb/lb	0.012	0.013	0.013	0.014	0.010	0.009
Conversion, maf coal	38.0	37.6	36.6	35.5	31.4	24.3
(Wt pct), Carbon	29.8	27.8	26.3	27.8	25.1	19.1
do. Hydrogen	70.1	70.3	72.0	70.6	55.5	47.0
do. Sulfur	43.8	50.9	63.8	51.2	51.8	42.1
do. Nitrogen	26.2	30.5	37.9	34.1	17.4	13.7
Recovery, overall	94.6	97.2	99.0	98.7	95.8	94.5
(Wt pct), Carbon	96.5	96.1	100.0	99.5	97.1	95.1
do. Hydrogen	104.7	106.0	105.4	106.7	100.6	97.3
do. Ash	102.4	100.0	100.6	102.1	101.6	105.7

TABLE 2.- Typical analyses of coals used in this study

	Pittsburgh Seam hvab coal	Illinois #6 hvcb coal	N. Dakota lignite
Proximate Analysis			
Moisture	1.2	1.4	7.8
Volatile matter	36.4	36.8	39.7
Fixed carbon	56.7	55.9	46.9
Ash	5.7	5.9	5.6
Ultimate Analysis (Dry Basis)			
C	79.09	75.45	64.64
H	5.22	5.12	4.48
N	1.60	1.72	0.76
S	1.10	1.32	0.76
O (by difference) ...	7.22	10.41	23.29
Ash	5.77	5.98	6.07
	100.00	100.00	100.00

Hvab coal from U.S. BuMines experimental mine, Bruceton, Pa.
Hvcb coal from Orient #3 mine, Freeman Coal Co., Waltonville, Ill.
Lignite from Baukol-Noonan mine, Burke Co., N. Dakota.

TABLE 3.- Production of pipeline quality gas in FDP reactor

Test No., IHR-	Base	156	176	151	166
Pressure, psig	1000	1000	1000	1100	1200
Coal	hvab	hvab	hvab	hvab	hvab
Feed gas/coal ratio, scf/lb	11.1	12.3	12.5	11.9	12.3
Feed gas composition, vol pct					
Hydrogen	46.1	49.0	48.0	48.0	49.2
Methane	47.5	49.4	49.4	49.2	48.7
Nitrogen	0.0	1.6	2.2	2.6	2.1
Carbon monoxide	4.0	0.0	0.0	0.0	0.0
Carbon dioxide	1.7	0.0	0.0	0.2	0.0
Carbon conversion, wt pct	20.0	25.0	24.0	25.3	--
Product gas (water-free)/coal ratio, scf/lb	14.8	13.4	13.9	14.1	13.3
Product gas composition, water-free					
Hydrogen	21.4	22.4	22.8	22.1	22.7
Methane	68.8	71.4	71.6	71.7	72.2
Ethane	0.0	0.5	0.2	0.1	0.1
Carbon monoxide	4.2	3.2	2.5	2.5	2.3
Carbon dioxide	1.3	0.7	0.7	0.5	0.5
Nitrogen	1.0	1.4	1.9	2.7	2.3
Hydrogen sulfide		0.4	0.2	0.4	0.2
Methane/hydrogen in product	3.21	3.19	3.14	3.24	3.18
Heating value, as-received, Btu/scf	779	817	812	802	815
Heating value w/4% CO Methanation, Btu/scf	927	918	908	903	914
Pct methane equivalent (CH_4+C_2H_6) made directly	94.2	94.7	94.7	94.7	94.8

(Continued on next page)

(TABLE 3.- Continued)

Test No., IHR-	160	165	154	157	158	172
Pressure, psig	1500	1500	2000	2000	2000	2000
Coal	hvab	hvab	hvab	hvab	hvab	hvab
Feed gas/coal ratio, scf/lb	13.1	12.3	12.5	12.4	12.8	13.2
Feed gas composition						
Hydrogen	53.8	51.3	52.5	49.9	51.8	50.6
Methane	43.4	47.0	46.4	48.4	46.6	46.4
Nitrogen	2.6	1.7	1.0	1.7	1.6	3.0
Carbon monoxide	0.0	0.0	0.0	0.0	0.0	0.0
Carbon dioxide	0.2	0.0	0.0	0.0	0.0	0.0
Carbon conversion, wt pct ..	24.2	23.3	25.1	30.0	25.0	28.0
Product gas (water-free)/ coal ratio, scf/lb	13.5	13.4	14.0	13.9	14.0	13.9
Product gas comp., water-free						
Hydrogen	19.7	21.7	19.8	18.1	18.0	20.4
Methane	75.2	73.4	75.9	79.0	78.7	75.8
Ethane	0.3	0.1	0.1	0.1	0.1	0.1
Carbon monoxide	1.4	2.1	2.2	0.5	1.1	0.7
Carbon dioxide	0.8	0.6	0.4	0.4	0.4	0.2
Nitrogen	2.4	2.0	1.5	1.7	1.6	2.6
Hydrogen sulfide	0.1	0.2	0.1	0.2	0.2	0.2
Methane/hydrogen in product.	3.82	3.38	3.83	4.36	4.37	3.72
Heating value, as-received, Btu/scf	835	823	842	863	862	839
Heating value with 4% CO methanation, Btu/scf	928	916	936	948	949	920
Pct methane equivalent (CH_4+C_2H_6) made directly .	95.0	94.8	95.0	95.2	95.2	95.0

(Continued on next page)

(TABLE 3.- Continued)

Test No., IHR-	164	162	163	184
Pressure, psig	1200	1500	2000	1000
Coal	hvcb	hvcb	hvcb	lignite
Feed gas/coal ratio, scf/lb	14.0	12.7	12.9	12.3
Feed gas composition				
Hydrogen	50.1	48.8	52.0	50.1
Methane	47.7	49.3	46.1	47.8
Nitrogen	2.2	1.8	1.9	1.8
Carbon monoxide	0.0	0.0	0.0	0.2
[illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Carbon conversion, wt pct	27.8	27.8	26.3	32.1
Product gas (water-free)/coal ratio, scf/lb	16.9	15.0	15.5	10.6
Product gas composition, water-free				
Hydrogen	21.9	20.4	20.0	27.9
Methane	72.8	75.0	73.8	57.5
Ethane	trace	trace	0.2	0.1
Carbon monoxide	2.4	1.9	2.2	6.3
Carbon dioxide	0.7	0.8	1.3	5.9
Nitrogen	1.9	1.7	2.3	2.1
Hydrogen sulfide	0.3	0.2	0.2	0.1
Methane/hydrogen in product	3.32	3.68	3.69	2.06
Heating value, as-received, Btu/scf	818	833	824	695
Heating value with 4% CO methanation, Btu/scf	914	928	925	902
Pct methane equivalent (CH_4+C_2H_6) made directly	94.8	94.9	94.9	90.1

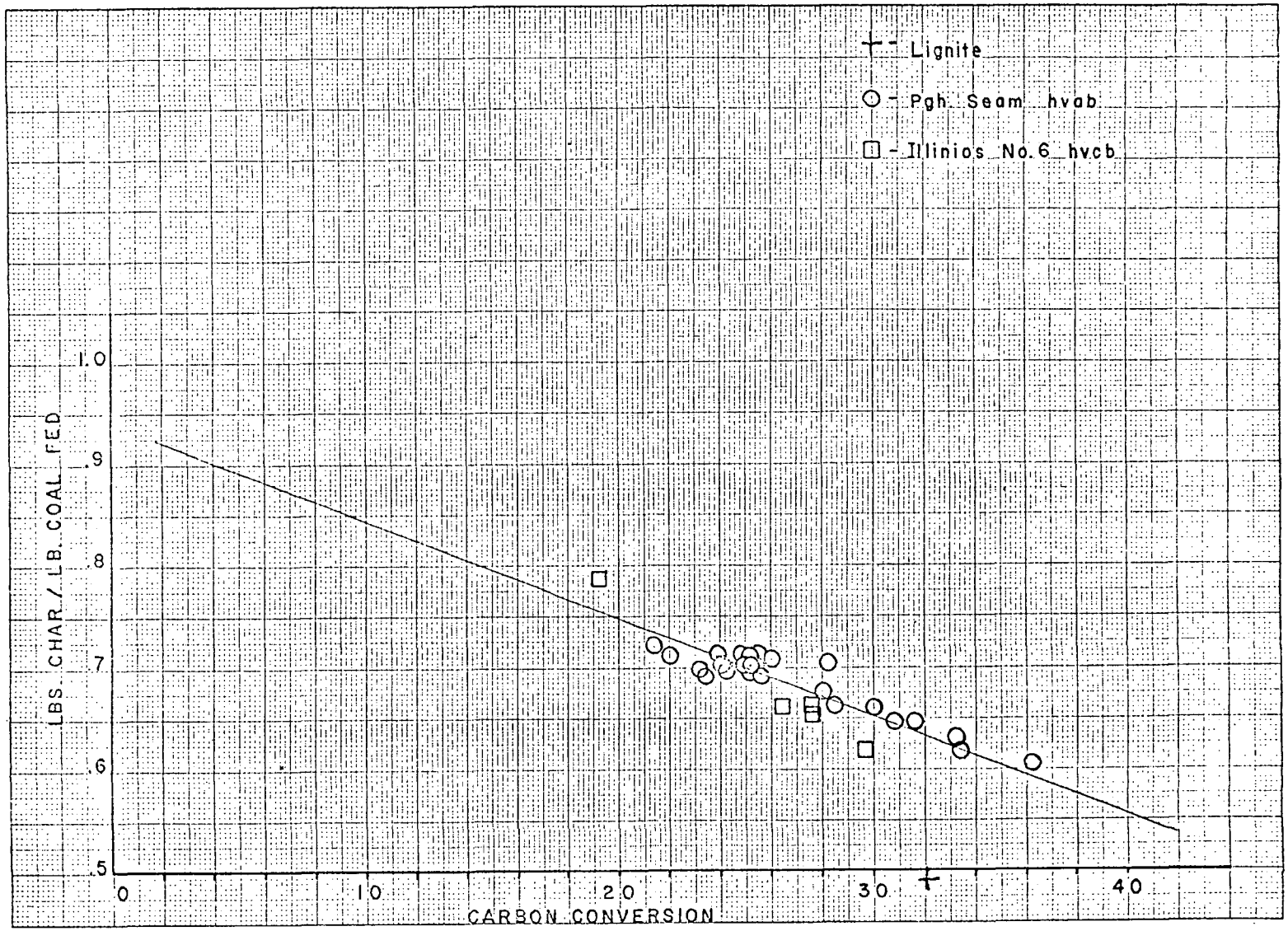

Fig. 1 DEPENDENCE OF FDP REACTOR CHAR YIELD ON CARBON CONVERSION

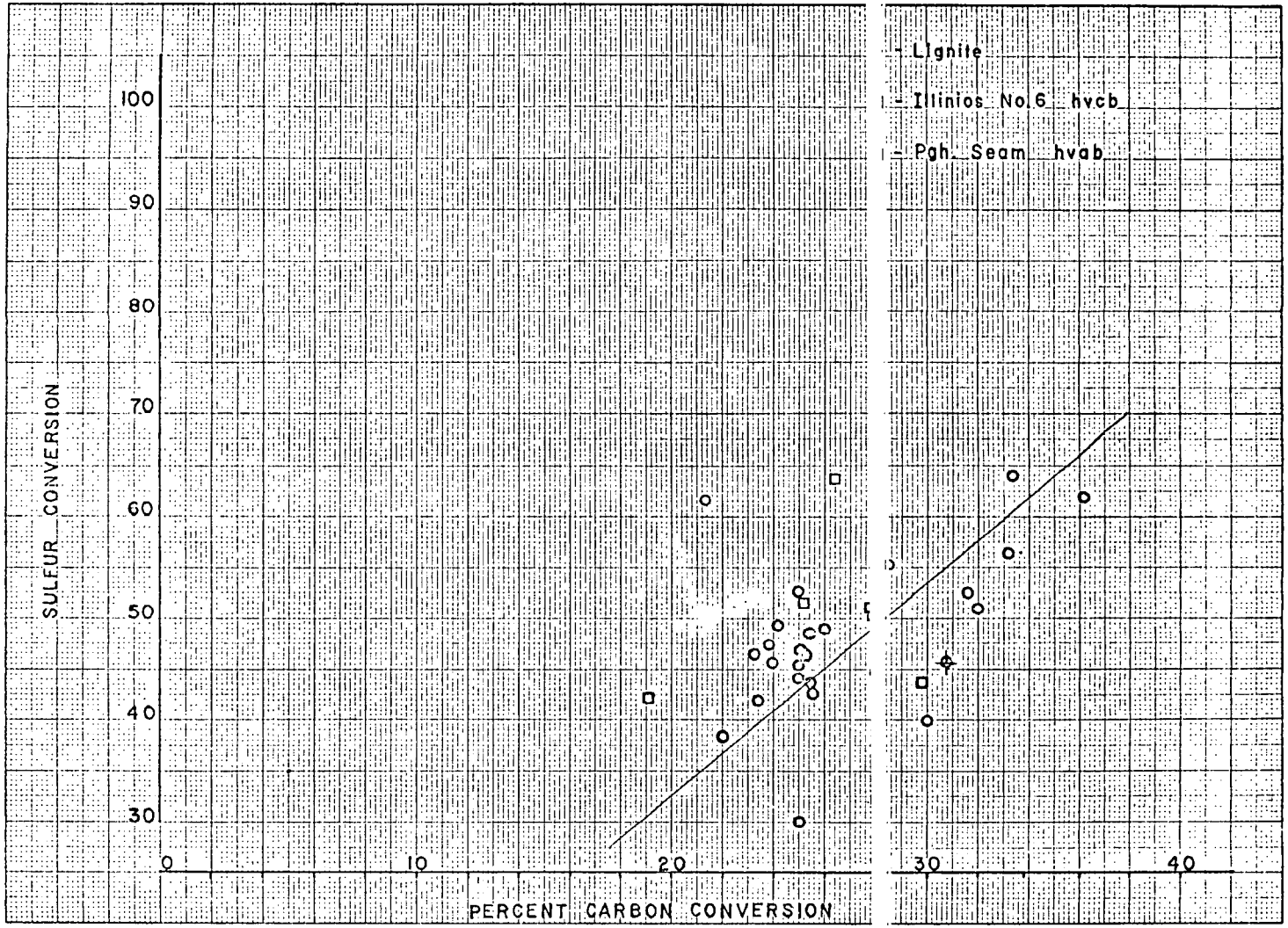

Fig. 2 SULFUR ELIMINATION IN FDP REACTOR

K+E 10 X 10 TO THE CENTIMETER 46 1510
10 X 25 CM MADE IN U.S.A.
KEUFFEL & ESSER CO.

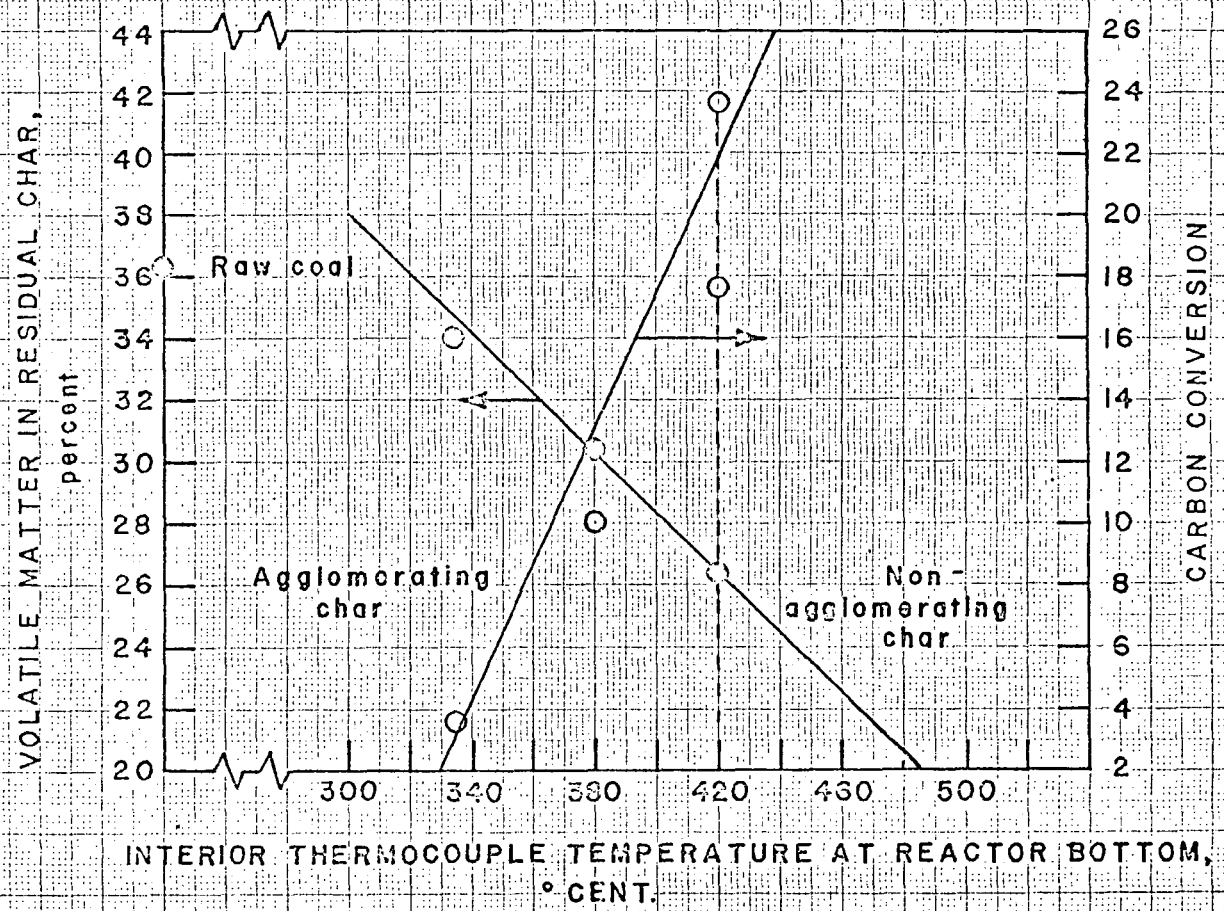

Fig. 3 EFFECT OF ESTIMATED PARTICLE TEMPERATURE ON CARBON CONVERSION AND VOLATILE MATTER

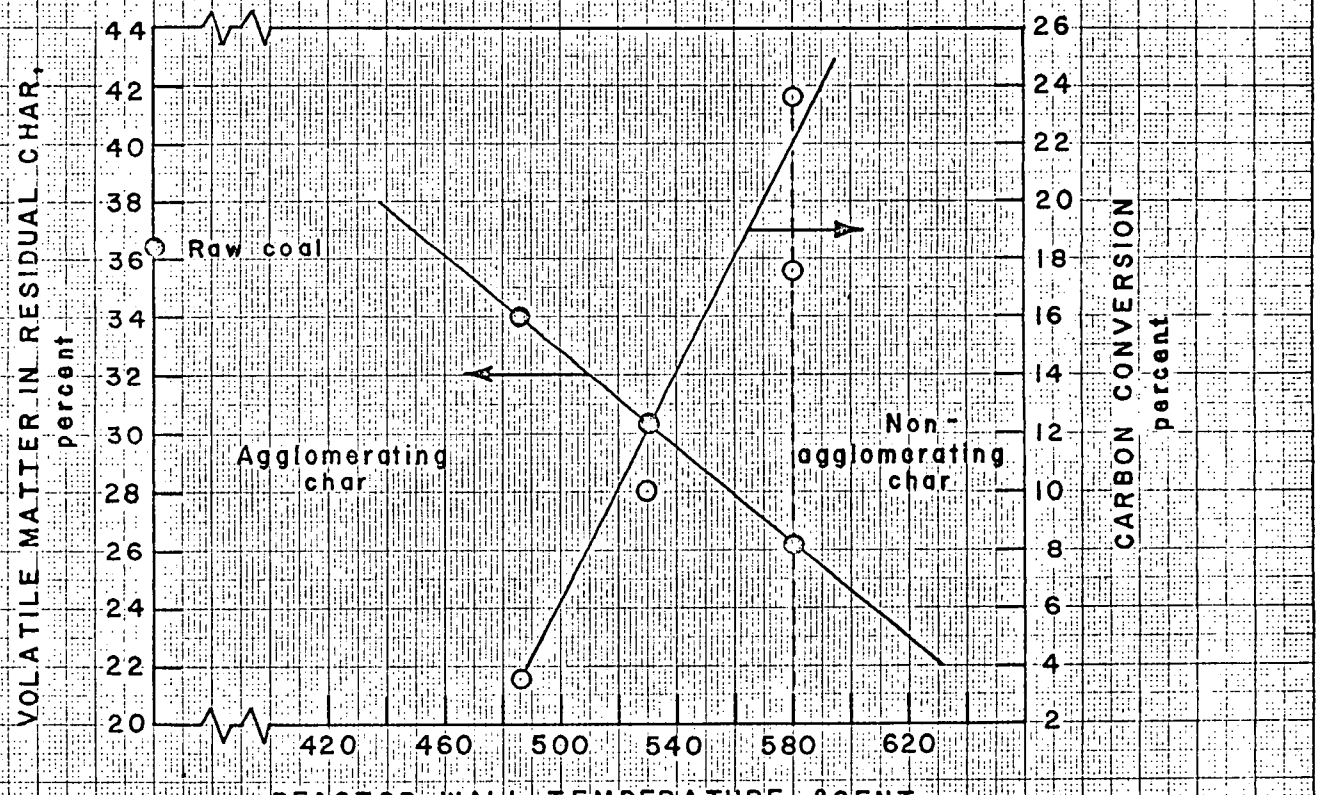

Fig. 4 EFFECT OF ESTIMATED PARTICLE TEMPERATURE ON CARBON CONVERSION AND VOLATILE MATTER

K+E 10 X 10 TO THE CENTIMETER 46 1510 10 X 25 CM MADE IN U.S.A. KEUFFEL & ESSER CO.