A Dispersion Model for the Solvent Refined Coal Process

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In the SRC process, pulverized coal is mixed with an internally generated coal-derived solvent. The resulting slurry and hydrogen gas are pumped and preheated to 780-850°F through a gas-fired preheater connected to a dissolver section for further reaction. The coal is depolymerized by thermal cracking and reacts with hydrogen to produce lower molecular weight materials. At the same time some of the sulfur contained in the coal reacts with hydrogen to produce gaseous H_2S which is then removed by standard technology.

In this paper an axial dispersion model for two-phase flow is applied to study the kinetics of hydrogenation, desulfurization, and coal conversion in the SRC process. In applying the model, the Peclet number for the preheater section is assumed to be infinite, while the Peclet number for the dissolver will usually be around unity. Physically this means that the dissolver will experience a significant amount of axial dispersion. In applying the dispersion model to the SRC process, one needs to know reaction kinetics rate expressions for the pertinent reactions. Considerable research has been conducted concerning the kinetics of coal solvation and sulfur removal. Wen and Han (1975) have determined rate constants using coal liquefaction and desulfurization data gathered primarily from studies by Pittsburgh and Midway Coal Mining Co., the University of Utah, and the Colorado School of Mines. These researchers were able to fit data from the above sources with an empirical expression for the rate of coal dissolution; however, no kinetic expression was obtained for desulfurization and the effect of hydrogen partial pressure on coal dissolution and sulfur removal was not firmly established. It is likely that the different types of reactors and experimental procedures employed in the three laboratories made the data correlation more difficult. Very recently rate expressions describing the hydrogen transfer during the coal dissolution and hydrodesulfurization were determined by Pitts (1976). By performing batch autoclave experiments with Kentucky No. 9/14 coal, hydrogen gas, and creosote oil as solvent, he was able to develop kinetic models for rates of hydrogen consumption, coal dissolution and hydrodesulfurization. His results can be summarized as follows:

A. For hydrogenation: (Guin, et al., 1976)

$$r_{H_2} = -k_{H_2L} e^{-\Delta H/RT} c_{H_2L} \qquad (1)$$

where $k_{H_2L} = 1.06 \times 10^5$/minute and $\Delta H = 21$ Kcal/mole

B. For hydrodesulfurization: (Guin, et al., 1977)

Two parallel independent first order reactions indicative of two different classes of more specific compounds were proposed to represent hydrodesulfurization kinetics, they are:

$$r_{s_1} = -k_{s_1L}\, c_{s_1L} \tag{2}$$

$$r_{s_2} = -k_{s_2L}\, c_{s_2L} \tag{3}$$

where $k_{s_1L} = 5.939 \times 10^{11}\, e^{-40.78\ \mathrm{Kcal/RT}}$

$k_{s_2L} = 1.664 \times 10^{3}\, e^{-18.99\ \mathrm{Kcal/RT}}$

and the initial concentration of two sulfur species are

$c_{s_1}{}^{o} = 1.53 \times 10^{-3}$ grams organic sulfur/cm^3

$c_{s_2}{}^{o} = 6.51 \times 10^{-3}$ grams organic sulfur/cm^3

C. For coal dissolution:

Coal liquefaction was described by a model incorporating dissolved molecular hydrogen. The model assumed there are two differently reactive fractions of coal present in the reaction mixture. The rate equations were formulated to give two independent reactions, they are:

$$r_{c_1} = -k_{c_1L} c_{c_1L} c_{H_2L}^{2} \tag{4}$$

$$r_{c_2} = -k_{c_2L} c_{c_2L} c_{H_2L}^{2} \tag{5}$$

with initial ration of the reactive coal species given by $\frac{c_1{}^{o}}{c_2{}^{o}} = 0.771$ and

$$k_{c_1L} = 3.269 \times 10^{6}\, e^{-4.01\ \mathrm{Kcal/RT}} \tag{6}$$

$$k_{c_2L} = 6.988 \times 10^{10}\, e^{-21.4\ \mathrm{Kcal/RT}} \tag{7}$$

with these available rate expressions, one can now formulate the design equations for the SRC reactor. These equations will be presented below.

For the preheater section:

$$\frac{dC_{H_2g}}{dX} + \frac{k_c^I aL}{\hat{v}_g^I H}\,(C_{H_2g} - C_{H_2L}) = 0 \tag{8}$$

$$\frac{dC_{H_2L}}{dX} - \frac{k_c^I aL}{\hat{v}_L{}^I}\,(C_{H_2g} - C_{H_2L}) + \frac{k_{H_2L}^I L(1-f_g^I)}{\hat{v}_L^I}\, C_{H_2L} = 0 \tag{9}$$

$$\frac{dC_{c_1L}}{dX} + \frac{k_{c_1L}^I\, L(1-f_g^I)}{\hat{v}_L{}^I}\,\left(\frac{c_r}{H}\right)^2 C_{c_1L}\, C_{H_2L}{}^2 = 0 \tag{10}$$

$$\frac{dC_{c_2L}}{dX} + \frac{k_{c_2L}^I\, L(1-f_g^I)}{\hat{v}_L{}^I}\,\left(\frac{c_r}{H}\right)^2 C_{c_2L} C_{H_2L}{}^2 = 0 \tag{11}$$

$$\frac{dC_{s_1L}}{dX} + \frac{L\,(1-f_g^I)}{\hat{v}_L^I}\, k_{s_1L}^I\, C_{s_1L} = 0 \tag{12}$$

$$\frac{dC_{s_2L}}{dX} + \frac{L(1-f_g^{I})}{\hat{v}_L^{I}} k_{s_2L}^{I} C_{s_2L} = 0 \tag{13}$$

with the conditions at X = 0:

$$C_{H_2g} = C_{H_2g}^{o} \qquad C_{c_1L} = C_{c_1L}^{o} \qquad C_{s_1L} = C_{s_1L}^{o}$$
$$C_{H_2L} = C_{H_2L}^{o} \qquad C_{c_2L} = C_{c_2L}^{o} \qquad C_{s_2L} = C_{s_2L}^{o} \tag{14 a-f}$$

For the Dissolver section:

$$\frac{1-X_1}{Pe,g} \frac{d^2C_{H_2g}}{dX^2} = \frac{d\,C_{H_2g}}{dX} + \frac{k_c^{II}aL}{\hat{v}_g^{II}H} (C_{H_2g} - C_{H_2L}) \tag{15}$$

$$\frac{1-X_1}{Pe,L} \frac{d^2C_{H_2L}}{dX^2} = \frac{dC_{H_2L}}{dX} - \frac{k_c^{II}aL}{\hat{v}_L^{II}} (C_{H_2g} - C_{H_2L}) + \frac{k_{H_2L}L(1-f_g^{II})}{\hat{v}_L^{II}} C_{H_2L} \tag{16}$$

$$\frac{1-X_1}{Pe,L} \frac{d^2C_{c_1L}}{dX^2} = \frac{dC_{c_1L}}{dX} + \frac{k_{c_1L}^{II}\,L(1-f_g^{II})}{\hat{v}_L^{II}} \left(\frac{c_r}{H}\right)^2 x\, C_{c_1L}\, C_{H_2L}^2 \tag{17}$$

$$\frac{1-X_1}{Pe,L} \frac{d^2C_{c_2L}}{dX^2} = \frac{dC_{c_2L}}{dX} + \frac{k_{c_2L}^{II}L(1-f_g^{II})}{\hat{v}_L^{II}} \left(\frac{c_r}{H}\right)^2 x\, C_{c_2L}\, C_{H_2L}^2 \tag{18}$$

$$\frac{1-X_1}{Pe,L} \frac{d^2C_{s_1L}}{dX^2} = \frac{dC_{s_1L}}{dX} + \frac{L(1-f_g^{II})}{\hat{v}_L^{II}} k_{s_1L}^{II} C_{s_1L} \tag{19}$$

$$\frac{1-X_1}{Pe,L} \frac{d^2C_{s_2L}}{dX^2} = \frac{dC_{s_2L}}{dX} + \frac{L(1-f_g^{II})}{\hat{v}_L^{II}} k_{s_2L}^{II} C_{s_2L} \tag{20}$$

with the conditions at $X = X_1$:

$$C_{H_2g}(X_1^-) = C_{H_2g}(X_1^+) - \frac{1-X_1}{Pe,g} \frac{dC_{H_2g}(X_1^+)}{dX} \tag{21}$$

$$C_{H_2L}(X_1^-) = C_{H_2L}(X_1^+) - \frac{1-X_1}{Pe,L} \frac{dC_{H_2L}(X_1^+)}{dX} \tag{22}$$

$$C_{c_1L}(X_1^-) = C_{c_1L}(X_1^+) - \frac{1-X_1}{Pe,L} \frac{dC_{c_1L}(X_1^+)}{dX} \tag{23}$$

$$C_{c_2L}(X_1^-) = C_{c_2L}(X_1^+) - \frac{1-X_1}{Pe,L} \frac{dC_{c_2L}(X_1^+)}{dX} \tag{24}$$

$$C_{s_1L}(X_1^-) = C_{s_1L}(X_1^+) - \frac{1-X_1}{Pe,L} \frac{dC_{s_1L}(X_1^+)}{dX} \tag{25}$$

$$C_{s_2L}(X_1^-) = C_{s_2L}(X_1^+) - \frac{1-X_1}{Pe,L} \frac{dC_{s_2L}(X_1^+)}{dX} \tag{26}$$

and the conditions at $X = 1$,

$$\frac{dC_{H_2g}}{dX} = 0 \qquad \frac{dC_{H_2L}}{dX} = 0 \tag{27 a,b}$$

$$\frac{dC_{c_1L}}{dX} = 0 \qquad \frac{dC_{c_2L}}{dX} = 0 \tag{28 a,b}$$

$$\frac{dC_{s_1L}}{dX} = 0 \qquad \frac{dC_{s_2L}}{dX} = 0 \tag{29 a,b}$$

where X_1^- designates the position just inside the exit of the preheater and X_1^+ is the position just inside the inlet of dissolver.

Numerical Solution Technique

The set of first order ordinary differential design equations (8)-(13) for the preheater and transfer line section of SRC reactor can be integrated numerically from $X = 0$ to $X = X_1$ using a Runge-Kutta 4th order method. Once these equations are solved, we then proceed to solve the set of second order differential design equations (15)-(20) for the dissolver section of the SRC reactor. Equations (15)-(20) can also be integrated from $X = X_1$ to $X = 1$ using a 4th order Runge-Kutta procedure. In performing the integration, these second order differential equations have to be transformed into a system of first order differential equations. In order to integrate the system of first order differential equations from $X = X_1$ to $X = 1$, it is necessary to specify as many additional boundary conditions at $X = X_1$ as there are conditions to be satisfied at $X = 1$, we need six additional conditions at $X = X_1$, namely $C_{H_2g}(X_1^+)$, $C_{H_2L}(X_1^+)$, $C_{c_1L}(X_1^+)$, $C_{c_2L}(X_1^+)$, $C_{s_1L}(X_1^+)$, and $C_{s_2L}(X_1^+)$.

These values have to be guessed initially and successively adjusted to satisfy boundary conditions at X = 1. The Newton-Raphson method is used to make this adjustment.

In some cases when some of the parameters appearing in the governing differential equations (8-13) and (15-20) become very large, these differential equations may suffer from a difficulty known as stiffness when numerical solution of the equations is attempted. In such cases, the classical fourth order Runge-Kutta integration technique may fail to integrate the system of stiff equations due to numerical instability. However, it has been found that Gear's method (1971 a,b) is very powerful for solving stiff differential equations, and therefore, its use is recommended when the fourth order Runge-Kutta integration technique fails.

Results and Conclusions

In this section, data from the Wilsonville SRC pilot plant will be compared with results predicted using the model proposed in this paper. The operating conditions and reactor configuration at Wilsonville will be employed in the calculation of the extent of coal dissolution, hydrodesulfurization and hydrogen consumption based on the present proposed model.

The following operating conditions and reactor configuration were used in the simulation:

Coal Type	Kentucky 9/14 mixture
Hydrogen Pressure	1700 psig
Dissolver Temperature	820° F
Slurry Concentration	25%
Coal Feed Rate	500lb./hr.
Gas Feed Rate	11.7 scf/lb. of coal
Hydrogen Concentration	85%
Dissolver Diameter	12 in.
Preheater Diameter	1½ in.
Transfer line Diameter	1½ in.
Preheater Volume*	1.6 $ft.^3$
Transfer line Volume	1.5 $ft.^3$
Dissolver Volume	18.0 $ft.^3$
Mean Reaction Temperature of Preheater and Transfer line	700° F

* The actual coal conversion reaction is initiated in the preheater coil. The temperature in this coil increases from 100° F to 800° F. Since the rate of conversion below 600° F is expected to be low, the reaction volume for the preheater was arbitrarily defined as the volume of that portion of the coil in which the temperature exceeds 600° F.

Based on the above conditions the following data can be obtained from literature correlating equations, which were needed in order to solve the design equations:

	Preheater/Transfer Line		Dissolver
$\hat{v}_L$	21.3 cm/sec	$\hat{v}_L$	0.335 cm/sec
$\hat{v}_g$	102.7 cm/sec	$\hat{v}_g$	1.80 cm/sec
Z_1	253 ft.	Z_2	23 ft.
$N_{Pe,L}$	1.18×10^3	$N_{Pe,L}$	0.97
$N_{Pe,g}$	4.83×10^2	$N_{Pe,g}$	4.66
f_g	0.44	f_g	0.06
$k_c a$	3.57×10^{-1}/sec	$k_c a$	5.53×10^{-2}/sec

Preheater/Transfer Line (cont.)		Dissolver (cont.)	
Henry's Constant at 700°F	$2.8 \frac{g/cc \text{ of } H_2 \text{ in vapor}}{g/cc \text{ of } H_2 \text{ in liquid}}$	Henry's Constant at 820°F	$2.5 \frac{g/c.c \text{ of } H_2 \text{ in vapor}}{g/c.c \text{ of } H_2 \text{ in liquid}}$

Several cases were studied in this paper, and the parameters chosen for these are listed in Table 1.

Case A: simulates the Wilsonville SRC pilot plant. The yield of coal dissolution predicted by the present proposed model is 92% which is in good agreement with that measured under similar conditions in the Wilsonville pilot plant. The actual yield in the pilot plant was also around 92%.

Cases B & C: investigate the effect of mass transfer on the SRC process. In these two cases the mass transfer coefficients were chosen as 1/10 and 1/50 of that of Case A. The coal dissolution is reduced from 92% for Case A to 90.7% for Case B and to 85.5% for Case C. The reason for this is obvious because for the lower mass transfer coefficient, $k_c a$, the dissolved hydrogen concentration in the solvent will also be lowered. Since the coal dissolution rate depends on the second power of the dissolved hydrogen concentration as can be seen from eqs. (4) and (5), the reduction in coal dissolution is expected. However, the reduction of the hydrogen mass transfer coefficient does not effect the extent of hydrodesulfurization, because the desulfurization rate does not depend on the dissolved hydrogen concentration (eqs. 2 and 3).

Cases D & E: investigate the effect of Peclet number in the dissolver section on the yield of coal dissolution, hydrodesulfurization and hydrogen consumption. For a given reactor and given operating conditions, there will be a corresponding Peclet number. The higher the value of the Peclet number, the closer a plug flow condition will be approached; the lower the value of the Peclet number, the closer the well-mixed condition will be approached. If the Peclet number is in the intermediate range, say, $0.1 < N_{Pe} < 20$, then neither of the above mentioned idealized situations obtains. In this intermediate region, the axial dispersion model may be used. In general, the yield for a reactor with higher Peclet number is higher than that with lower Peclet number. This can be seen from Cases D & E where both the gas phase and liquid phase Peclet number in the dissolver section were chosen as 0.1, and the mass transfer coefficients were chosen respectively as equal to Cases A and C. The coal liquefaction yield for Case D is 89.1%, which is about 2.9% lower than that for Case A; for Case E the coal dissolution is 82%, which shows a 3.5% decrease compared to Case C. The Peclet numbers for both Cases A & C are $N_{Pe,L} = 0.97$ and $N_{Pe,g} = 4.66$. From the above comparisons, we see that the effect of Peclet number on coal dissolution becomes more significant when the mass transfer coefficient is low.

The detailed concentration profiles are given in Figs. 1-4. Figs. 1 and 2 show the hydrogen concentration profile in the vapor phase and liquid phase, respectively, along the reactor. The coal concentration, $C_c = C_{c_1L} + C_{c_2L}$, profiles are shown in Fig. 3, and also shown in Fig. 4 is the organic sulfur concentration, $C_s = C_{s_1L} + C_{s_2L}$, profiles. The yields of coal liquefaction, hydrodesulfurization and hydrogen consumption rate for Cases A to E are also tabulated in Table 2.

In addition to the above five cases, cases for which the mass transfer coefficient was twice and ten times that of Case A were also simulated. The results showed that the hydrogen consumption and coal dissolution were not significantly different from Case A. This implies that the Wilsonville SRC pilot plant is already operating in a kinetically limited regime and that mass

transfer is not the rate controlling step. These two cases were not plotted on Figs. 1-4 because they were essentially indistinguishable from Case A. This result is reasonable, since no external catalyst, other than the naturally occuring coal mineral matter, is present in the SRC reactor. It is known, however, that coal minerals can and do catalyze both hydrogenation and desulfurization reactions in the SRC process (Henley, et al, 1976; Prather, et al, 1976; Guin, et al, 1977). Under operating conditions near those of Case A studied here, the organic sulfur removal and the hydrogen consumption at the Wilsonville SRC pilot plant averaged around 50% and 2% based on weight of maf coal feed, respectively. As observed from Table 2, the corresponding values from the model are about 50% lower than those determined experimentally in the Wilsonville SRC pilot plant under apparently similar operating conditions. There are two reasonable explanations for this deviation. First, it should be noted that in the present model the hydrogen consumption only includes the actual amount of hydrogen chemically reacting in the preheater and dissolver, i.e., no other losses such as from flash systems, waste vents, imcomplete separations, leaks, etc., are included. Secondly and perhaps more importantly, it has been observed that because of the very low upward liquid velocity, an accumulation of coal mineral matter occurs in the Wilsonville SRC reactor. This would cause increased desulfurization and hydrogenation because of the catalytic nature of these minerals. Recent lower hydrogen consumption with an initially clean reactor at Wilsonville tends to verify this latter result. Also it has recently been shown (Guin, et al, 1977), that the addition of iron compounds, predominant in coal minerals, can increase the extent of organic sulfur removal by around 100%. Finally, it should be noted that the kinetics rate expressions used herein are strictly valid only for the Kentucky 9/14 coal with a creosote oil solvent. For other coals and solvents, different rate expressions would be required for the most accurate results.

CONCLUSIONS

1. The mixing and flow pattern of the multi-phase reactor is taken into account through the axial dispersion term, which has not been previously utilized for the SRC reactor modeling.
2. The mass transfer of hydrogen from the gas phase into the liquid phase is also accounted for in the model, the effect of hydrogen mass transfer rate on the hydrogen consumption and coal dissolution rate can be easily simulated; and the influence and extent of mass transfer control can be assessed. In particular, the Wilsonville SRC reactor is found to operate in a kinetically, rather than a mass transfer limited regime.
3. All of the parameters arising in the model are obtainable independently from literature correlation; these parameters have definite physical meanings based upon the flow conditions, therefore, by changing the relative magnitude of each parameter, the general trend of the effect of each operating variable on the SRC process can be assessed, thus providing a better engineering design basis for simulation and optimization of the SRC process.
4. The hydrogen consumption and organic sulfur removal at the Wilsonville SRC pilot plant are higher than predicted by the present model, which uses kinetics rate expressions from laboratory experiments. This is probably due to the accumulation of catalytic coal minerals in the Wilsonville reactor, giving higher rate constants.

Nomenclature:

D_g	gas phase axial dispersion coefficient
D_L	liquid phase axial dispersion coefficient
v_{slip}	two phase slip velocity
v_g	bubble phase velocity, m/sec
v_L	liquid phase velocity, m/sec
$\bar{v}_g$	bubble phase superficial velocity, m/sec
$\bar{v}_L$	liquid phase superficial velocity
$\bar{\epsilon}_g$	gas holdup
a	specific gas-liquid interfacial area
Pe, g	$= \frac{v_g (L-z_1)}{D_g}$, gas phase Peclet number in the dissolver section
Pe,L	$= \frac{v_L(L-z_1)}{D_L}$, liquid phase Peclet number in the dissolver section
z_1	length of preheater plus transfer line section
z_2	length of dissolver
X_1	dimensionless axial length of preheater plus transfer line z_1/L
X	dimensionless axial coordinate, z/L
K_c	liquid phase mass transfer coefficient
$K_c a$	volumetric liquid phase mass transfer coefficient
z	coordinate in the axial direction
H_o	Henry's law constant of hydrogen in creosote oil
c_{ij}	feed concentration of component i in phase j
L	total length of reactor, $L = Z_1 + Z_2$
C_{ij}	$= c_{ij}/(H_j^k c_r)$ dimensionless concentration for ith component in the j phase.
c_{ij}	concentration of component i in the j phase
c_r	reference concentration

Superscript:

o	denotes initial condition
I	in the preheater section
II	in the dissolver section
k	= 0 for gas phase; = 1 for liquid phase

Subscript:

g	gas phase
L	liquid phase

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TABLE 1 Values of Parameters Appearing in Design Equations for Cases A, B, C, D and E

Parameters	$K_c aL/\hat{v}_g H$		$K_c aL/\hat{v}_L$		$k_{H_2L}(1-f_g)/\hat{v}_L$		$[k_{c_1L}(1-f_g)/\hat{v}_L](\frac{c_r}{H})^2$	
Cases	Preheater	Dissolver	Preheater	Dissolver	Preheater	Dissolver	Preheater	Dissolver
A	10.33	104	139.2	137.4	2.878×10^2	14.4	1.343×10^3	2.369×10^5
B	1.033	10.4	13.92	13.74	2.878×10^2	14.4	1.343×10^3	2.369×10^5
C	0.207	2.08	2.78	27.48	2.878×10^2	14.4	1.343×10^3	2.369×10^5
D	10.33	104	139.2	137.4	2.878×10^2	14.4	1.343×10^3	2.369×10^5
E	0.207	2.08	2.78	27.48	2.878×10^2	14.4	1.343×10^3	2.369×10^5

Parameters	$[k_{c_2L}L(1-f_g)/\hat{v}_L](\frac{c_r}{H})^2$		$k_{S_1L}L(1-f_g)/\hat{v}_L$		$k_{S_2L}L(1-f_g)/\hat{v}_L$		$N_{Pe,g}$	$N_{Pe,L}$	c_r
Cases	Preheater	Dissolver	Preheater	Dissolver	Preheater	Dissolver			
A	35.2	2.283×10^4	1.8×10^{-2}	67.35	1.68×10^{-3}	0.941	4.66	0.97	0.141
B	35.2	2.283×10^4	1.8×10^{-2}	67.35	1.68×10^{-3}	0.941	4.66	0.97	0.141
C	35.2	2.283×10^4	1.8×10^{-2}	67.35	1.68×10^{-3}	0.941	4.66	0.97	0.141
D	35.2	2.283×10^4	1.8×10^{-2}	67.35	1.68×10^{-3}	0.941	0.1	0.1	0.141
E	35.2	2.283×10^4	1.8×10^{-2}	67.35	1.68×10^{-3}	0.941	0.1	0.1	0.141

TABLE 2. Calculated Yields for the SRC Process

	Case A	Case B	Case C	Case D	Case E
Coal liquefaction yield	92%	90.7%	85.5%	89.1%	82%
Organic sulfur removal	25%	25%	25%	23.9%	23.9%
Hydrogen consumption (g H_2/g maf coal)	0.87%	0.82%	0.61%	0.66%	0.50 %

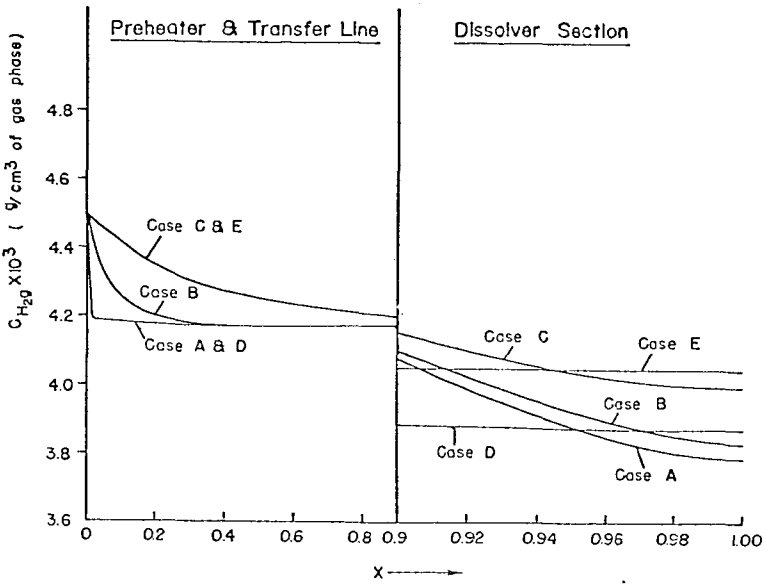

Figure 1. Vapor Phase Hydrogen Concentration vs. X

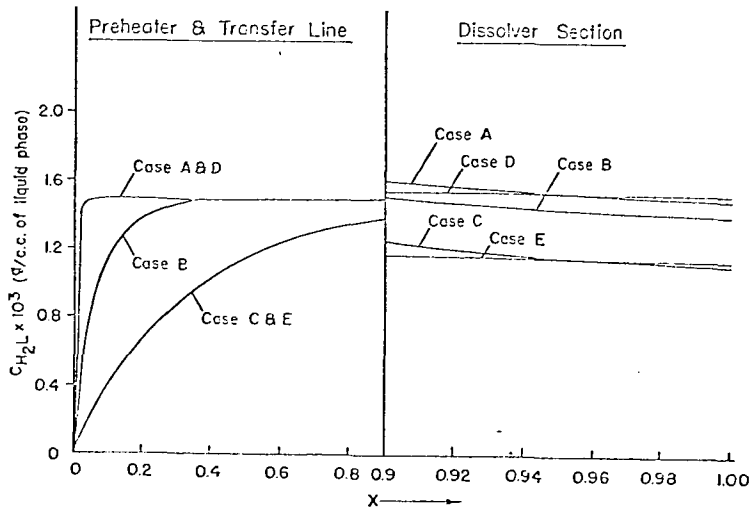

Figure 2. Liquid Phase Hydrogen Concentration vs. X

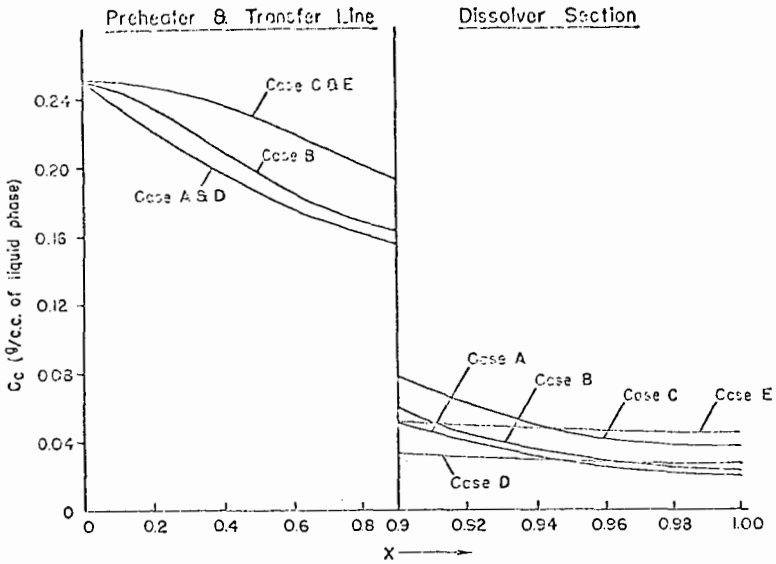

Figure 3. Coal Concentration vs. X

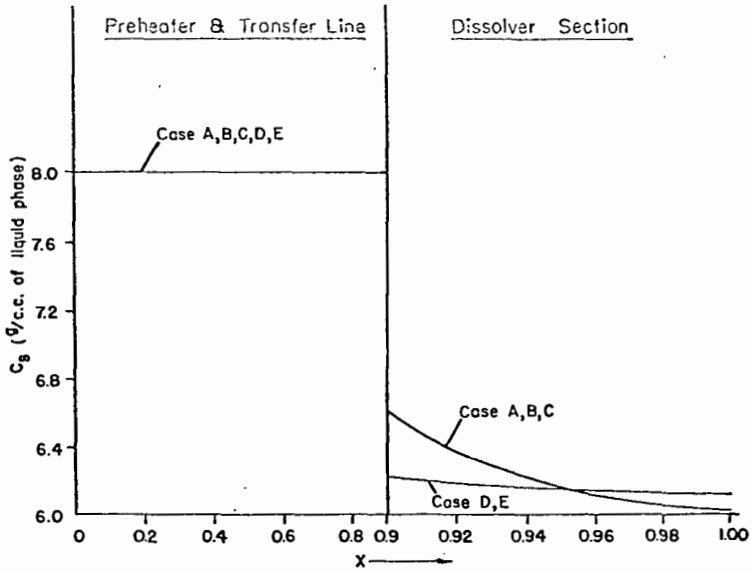

Figure 4. Sulfur Concentration vs. X

CONTINUOUS LIQUEFACTION OF LIGNITE IN A PROCESS DEVELOPMENT UNIT

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INTRODUCTION

The decline in domestic production and reserves of petroleum and natural gas has highlighted the necessity of supplemental sources. Presently nearly 50 percent of our petroleum requirements are imported, a situation which is economically and politically dangerous. A major source of replacement energy is our abundant coal reserves. Direct utilization of most coals is environmentally unacceptable, and in many applications the solid form is not suitable for use and liquid or gaseous phases are required.

Conversion of coal to liquid or gaseous fuels can be accomplished by hydrogenation to increase the H to C ratio. Direct liquefaction from coal without gasification and synthesis steps allows reduced hydrogen requirements, fewer process steps, and consequently, lower costs.[1] Ash and sulfur contents can be reduced and heating value per unit weight increased. A low polluting solvent refined coal or synthetic fuel oil could be used in stationary power generating facilities. Liquid fuels suitable for use in mobile transportation can be produced by additional hydrogenation.

Many hydrogenation-liquefaction processes have been proposed for coal. Most supported research on liquefaction has been directed toward high rank coal and has been summarized by Friedman et al[1]. A modification appropriate for hydrogenation of low rank coals is the use of carbon monoxide in the presence of a hydrogen donor solvent.[2] Natural moisture content in the low rank coal is more effective than added moisture. Synthesis gas (H_2 and CO) also gives high conversion and is less expensive than either pure carbon monoxide or hydrogen.[3] Although naturally occurring ash constituents in the lignite are known water gas shift reaction catalysts, the contribution of carbon monoxide seems to be more complex than this and is ash catalyzed.[4] Carbon monoxide appears to be effective because of removal of a source of cross linkage rather than by bond cleavage.

Initial work on liquefaction of lignite at the University of North Dakota was batch autoclave tests under sponsorship of the Burlington Northern Railroad. A major expansion occurred in 1972 when a five year, research contract was signed with the U.S. Office of Coal Research (now the Energy Research and Development Administration). Extension of the contract is being negotiated.