IX) MODEL APPLICATION

Use of the Bhatia Epstein model for prediction of volume fractions from operating conditions was discussed previously (1). The three parameters (U_{tb}, K_o, and X_k) are provided to the model as input, along with gas and liquid velocities. Other necessary data, such as the Richardson Zaki parameters (U_t and n) may be predicted using the technique in Section VIII.

The application of this model to an operating liquefaction reactor is complicated by the variation of slurry density and viscosity; over the 46 operating periods of PDU-10, significant changes in these two slurry properties were observed. Inducing these changes and observing their effect was an objective of the PDU-10 run. In a full scale commercial operation, the ranges of fluid properties would not be as broad after the initial startup.

For modelling purposes, Amoco Test 6, coming at the end of the PDU-10 normal run may be taken as most representative. The values of bubble rise velocity, wake volume ratio, and wake concentration ratio from this test are recommended for use. Implicit in these values, however, are the effects of the Amocat I-A catalyst properties, coal and ash concentration, and the reactivity and yield structure of the Wyodak coal tested in PDU-10. Application of this model to other coals or catalysts should proceed only after suitable tests, such as physical inspection and micro-autoclave experiments provide the basis for estimation of slurry product properties.

X) CONCLUSIONS

The overall goal of our research efforts is to develop a fundamental understanding of the fluid dynamics occurring in the three-phase, ebullated-bed H-Coal system, as well as obtain data useful in the design, operation, and control of the system. Several important areas have been investigated under this DOE contract, and the results are summarized here.

A) The fluid dynamics occurring in HRI's H-Coal process development unit (PDU) during Run PDU-10 were measured and compared with Amoco Oil cold-flow fluidization results. It was found that catalyst bed expansions and gas holdups are higher in the PDU than those observed in the cold-flow tests for slurries having the same nominal viscosity.

B) Analysis of the PDU results shows that the differences in A) can be explained by assuming that the viscosity in the reactor is effectively four times greater than the viscosity of the slurry sample obtained from the pressure letdown vessel. It may be that viscosity gradients rather than a uniform viscosity exist in the reactor. It is known from results published in the literature that a small amount of asphaltenes added to an oil has a dramatic effect on the viscosity of the oil.

C) Comparison of PDU results with cold-flow results shows that the bulk of the operating reactor gas flow lies in the ideal bubbly regime. It also appears that the gas bubbles in these PDU tests are rising quite slowly. Only two of the operating points in our test program on the PDU were found to lie in the churn turbulent regime.

D) Two- and three-phase fluidization experiments were carried out in Amoco's cold-flow fluid dynamics unit. The data base now includes coal char/kerosene slurry concentrations of 4.0, 9.8, and 20.7 vol% in addition to the 15.5 and 17.8 vol% data from our earlier work. Both HDS-2A and Amocat-1A catalysts were used in the tests. Bed expansion is primarily a function of slurry velocity, with gas velocity having only a weak effect. Bed contractions have been observed in some cases at sufficiently high gas velocity.

E) Gas and liquid holdups were found to be uniform across the cross-section of the Amoco cold-flow fluid dynamics pilot plant.

F) A significant degree of backmixing was found to occur in the H-Coal system as measured in the Amoco cold-flow unit. Dispersion coefficients and Peclet numbers were found to lie in the ranges 70 to 130 cm^2/sec and 3.6 to 9.9, respectively.

G) Three techniques were developed for the study of the bubble dynamics occurring in three-phase fluidized beds: 1) A laser light beam probe for measuring the behavior of bubbles greater than 120 microns; 2) A laser holographic technique for determining the size, shape, and position of bubbles in the bed; and 3) A resistivity probe for determining bubble frequencies and a qualitative measure of the flow regime.

H) The bubble size studies in G) showed that the design of the inlet distributor played an important role in the breakup of bubbles entering the fluidized bed. It was shown that smaller bubbles can be generated by this particular inlet distributor than could be produced by the breakup of bubbles by solid particles within the bed.

I) An equilibrium bubble size model was developed based on data obtained using the light beam probe.

J) A viscometer was adapted for measurement of the viscosity of coal slurries at high temperature and pressure.

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NOMENCLATURE

A	Cross-sectional area of the column
BE	Catalyst bed expansion, %
C	Concentration, ppm
d	Particle diameter, cm
d_m	Sauter mean bubble diameter
d_3	Particle diameter
d_t	Column diameter
C(s)	Laplace transform of C
C_D	Drag coefficient
D	Reactor diameter, feet
D_l	Dispersion coefficient, cm^2/sec
F(s)	Laplace transfer function
g	Gravitational acceleration constant, 980.6 cm/sec^2
Ga	Galileo number
H	Catalyst bed height
I_o	Gamma-ray intensity through empty reactor, counts/sec
I_l	Gamma-ray intensity through kerosene, counts/sec
I_m	Gamma-ray intensity through test conditions, counts/sec
K_o	Wake volume ratio, volume of wake/volume of bubble
L	Distance between sampling trays, 157.5 cm or material thickness
M	Mass of dry catalyst
n	Richardson-Zaki index
Pe	Peclet number
Q	Amount of tracer injected into the system
R	Radial position index, R/R_o
R_i	Radial sampling position
R_o	Radius of fluidization column
Re_t	Particle Reynolds number
S	Standard deviation or square root of the variance in log scale
ΔT	PDU reactor temperature minus ambient temperature, ^oF
T	Sampling time corresponding to C, sec
T'	Sampling time corresponding to log-normal mean
T''	Sampling time corresponding to one log-normal standard deviation
T_m	Mean residence time, sec
Ug	Superficial gas velocity, cm/sec
U_l	Superficial liquid velocity, cm/sec

U_t	Particle terminal velocity, cm/sec
U_{tb}	Bubble terminal velocity, cm/sec
U_s	Gas/liquid slip velocity, cm/sec
U_z	Average linear liquid velocity, cm/sec
$\dot{Y}$	Volumetric liquid feed rate, ft^3/sec
V_{CD}	Drift flux, cm/sec
X_k	Solid concentration ratio in wake
z	Axial position, cm
Z	Axial position, dimensionless (z/L)

Greek

α	Thermal expansion coefficient, in/in ^oF
θ	Sampling time, dimensionless
σ	Second moment or square root of the variance, or surface tension
σ_θ	σ/T_m
μ	First moment or mean of the RTD curve, or viscosity in centipoise
ϵ_c	Volume fraction catalyst, dimensionless
ϵ_{co}	Settled catalyst volume fraction, dimensionless
ϵ_f	Volume fraction fines, dimensionless
ϵ_g	Volume fraction gas, dimensionless
ϵ_l	Volume fraction liquid, dimensionless
ϵ_w	Volume fraction wake, dimensionless
ϵ_{sl}	Volume fraction slurry = $\epsilon_l + \epsilon_f$
ρ_p	Catalyst dry particle density, g/cm^3
ρ_c	Catalyst soaked particle density, g/cm^3
ρ_f	Fines density, g/cm^3
ρ_l	Liquid density, g/cm^3
ρ_{sl}	Slurry density, g/cm^3
$\Delta\rho$	Solid/liquid density difference
ρ_B	Catalyst bulk density, g/cm^3
μ_c	Catalyst mass absorption coefficient
μ_f	Fines mass absorption coefficient
μ_l	Liquid mass absorption coefficient
ω_f	Weight percent fines
τ	Shear stress, gm/cm-sec
τ_o	Yield stress, gm/cm-sec
η_{pl}	Platic viscosity, cp
$\dot{\gamma}$	Shear rate, sec^{-1}
Δ	Difference (in predicted and observed volume fractions)