Greek Letters

ϵ = bed porosity
θ = dimensionless time ($= t/\bar{t}$)
μ = fluid viscosity
ρ = fluid or bed density
Λ = distribution parameter ($= \rho_b q_0/c_0$)

Subscripts

b = bed, breakthrough
0 = normalizing or input value

Literature Cited

Bieber, H., F. E. Steidler, and W. A. Selke, "Ion exchange rate mechanism," *Chem. Eng. Prog. Symp. Ser.*, **50**(14), 17–21 (1954).

Boyd, G. E., A. W. Adamson, and L. S. Myers, "The exchange adsorption of ions from aqueous solutions by organic zeolites—II Kinetics," *J. Am. Chem. Soc.*, **69**, 2836–2848 (1947).

Brauch, V. and E. V. Schlunder, "The scale up of activated carbon columns for water purification based on results from batch tests—II," *Chem. Eng. Sci.*, **30**, 539–548 (1975).

Carberry, J. J. "A boundary layer model of fluid particle mass transfer in fixed beds," *AIChE J.*, **6**, 460–463 (1960).

Cooney, D. O., W. Infantolino, and R. Kane, "Comparative studies of hemoperfusion devices. I. Invitro clearance characteristics," *Biomat. Med. Dev. Art. Org.*, **6**, 199–213 (1978).

Dunlop, E. H., S. G. Gazzard, P. G. Langley, M. J. Weston, L. R. Cox, and R. Williams, "Design features of hemoperfusion columns containing activated carbon," *Med. Biol. Eng.*, **14**, 220–226 (1976).

Kataoka, T., H. Yoshida, and K. Ueyama, "Mass transfer in laminar region between liquid and packing material surface in packed bed," *J. Chem. Eng. Japan*, **5**, 132–136 (1972).

Miyauchi, T. and T. Kiruchi, "Axial dispersion in packed beds," *Chem. Eng. Sci.*, **30**, 343–348 (1975).

Moison, R. L. and H. A. O'Hern, "Ion exchange kinetics," *Chem. Eng. Prog. Symp. Ser.*, **55** (24), 71–85 (1959).

Radcliffe, D. F., "Mass transfer in hemoperfusion columns and other sorbent based devices for blood detoxification," Ph.D. Thesis Uni. of Strathclyde (1978).

Selke, W. A., Y. Bard, A. D. Pasternak, and S. K. Aditya, "Mass transfer rates in ion exchange," *AIChE J.*, **2**, 468–470 (1956).

Updhyay, S. N. and G. Tripathi, "Liquid phase mass transfer in fixed and fluidized beds of large particles," *J. Chem. Eng. Data*, **20**, 20–26 (1975).

Manuscript received June 6, 1980; revision received December 2, and accepted December 10, 1980

Holdup Correlations in Slurry-Solid Fluidized Beds

I. A. VASALOS
and D. N. RUNDELL
Amoco Research Center
Naperville, IL
K. E. MEGIRIS
and G. J. TJATJOPOULOS
Department of Chemical Engineering
University of Thessaloniki, Greece

The use of slurry-solids fluid beds plays an important role in the development of synthetic fuels and the hydrogen processing of petroleum resids. In the H-Coal process in particular, coal oil slurries are processed over extrudates of hydrodesulfurization catalyst. The objective of this note is to present a model which describes the volume fraction occupied by the slurry phase. The accurate prediction of the liquid holdup is important not only for the calculation of the bed height in liquid solid fluidized beds, but also is required for the development of a model predicting the holdup in three-phase fluidized beds.

Although several correlations have been considered in the literature (Richardson and Zaki, 1954; Ramamurthy and Sabbaraju, 1973) for pure liquids, very little has been published for slurry systems. The objective of this paper is to extend the work of previous investigators for systems of particular interest to the H-Coal process. A summary of correlations describing liquid-solid fluidized beds has been presented elsewhere (Vasalos et al., 1979, 1980).

In the present publication, the application of two correlations is considered: 1) the Richardson and Zaki; and 2) Ramamurthy and Sabbaraju. The first relates the liquid volume fraction (ϵ_1) due to bed expansion to the ratio of the superficial velocity (U_1) to the terminal velocity of a single particle (U_t):

$$\epsilon_1 \eta = U_1/U_t \tag{1}$$

The second is based on an empirical approach where a function of Reynolds and Galileo numbers is related to the function $[1 - 1.21(1 - \epsilon_1)^{2/3}]^{-1}$. This function relates the velocity of a fluid in minimum cross-section of the bed to the superficial velocity. It is derived by assuming that the particles are arranged in an imaginary node of a simple cubical lattice.

EXPERIMENTAL

Data used in this work were obtained in a cold flow unit, details of which have been reported elsewhere (Vasalos et al., 1979). The unit consists of a glass reactor 15 cm in diameter and 5 m in height. Extrudates of hydrodesulfurization catalyst (1.8-mm diameter, 5.1-mm length) were fluidized with coal fines-kerosene slurries and mineral oil. The density and viscosity of these fluids are reported in Table 1. It has been found that this slurry has about the same viscosity as the H-Coal liquids at actual operating conditions.

For each set of conditions, the bed height was measured by scanning the entire reactor length with an elevator on which was mounted a 10 mc gamma-ray source and a detector. From the change in the gamma-ray absorption, the bed height can easily be inferred. The liquid or slurry holdup is then determined from the equation:

$$1 - \epsilon_1 = M/\rho_c AH \tag{2}$$

I. A. Vasalos is currently with Department of Chemical Engineering, University of Thessaloniki, Greece.

ISSN0001-1541-82-5046-0346-$2.00 © The American Institute of Chemical Engineers, 1982.

TABLE 1. SUMMARY OF EXPERIMENTAL RUNS

Run No.	Liquid	Fines, vol %	Temp, °K	Density, 10^{-3} kg/m³	Particle Density: Catalyst*	Particle Density: Coal Fines**	Fluid Viscosity (10^{-3} Ns/m²)
204	Kerosene	5.1	294	0.835	1.63	1.7	3.5
211	Kerosene	17.8	299	0.952	1.70		8.2
214	Kerosene	15.5	338	0.927	1.69		6.7
422	Mineral Oil	0	338	0.820	1.61		6.1
428	Mineral Oil	0	326	0.830	1.61		9.1

* Soaked with liquid.
** Density distribution; 90% (0.8–2.2) × 10^3 kg/m³, even distribution; particle size, 70%, 325 mesh.

RESULTS AND DISCUSSION

The application of the Richardson and Zaki correlation to coal char-kerosene slurries and mineral oil is shown in Figures 1 and 2, respectively. From plots of this type the particle terminal velocity, U_t, and the exponent, n, are derived with the application of standard techniques. The extrapolated particle terminal velocities and the calculated particle Reynolds and Galileo numbers are listed in Table 2. For the calculation of the dimensionless groups (Re_t, Ga), the equivalent particle diameter (defined as the diameter of a sphere having the same volume) is used. The terminal velocity Reynolds number was found to depend on Galileo number with the following correlation, which is also illustrated in Figure 3:

$$Re_t = 0.132\ Ga^{0.714} \tag{3}$$

where:

$$Re_t = d_s U_t \rho / \mu \tag{4}$$

$$Ga = d_s{}^3 \rho\ (\rho_s - \rho) g / \mu^2 \tag{5}$$

Figure 3 indicates that both coal slurry and mineral oil data fall on the same line, supporting an earlier conclusion (Vasalos et al., 1979) that coal char-kerosene slurries behave as a simple liquid of similar viscosity.

From Eq. 3, the particle terminal velocity is easily derived by the following equation:

$$U_t = 0.132\ g^{0.714} d_s{}^{1.142} (\rho_s - \rho)^{0.714} / \rho^{0.286} \mu^{0.428} \tag{6}$$

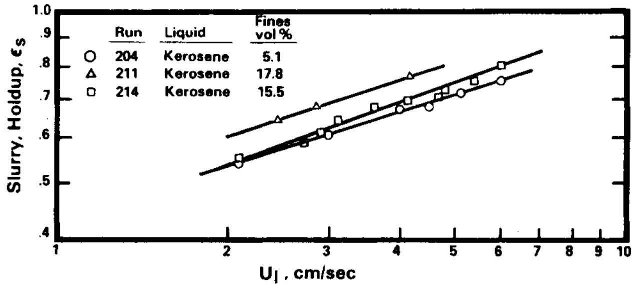

Figure 1. Application of Richardson-Zaki correlation to coal-char kerosene slurry/catalyst.

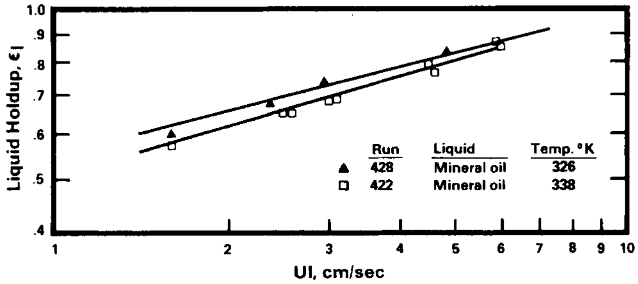

Figure 2. Application of Richardson-Zaki correlation to mineral oil/catalyst.

The range of Re_t and Ga covered is:

$$25 < Re_t < 100$$

$$10^3 < Ga < 10^5$$

Equation 6 is similar to that reported by Lewis and Bowerman (1978) for spheres; however, it is necessary to include the effect of particle sphericity. Sphericity is defined as:

$$\Phi_s = (\text{surface area of sphere/surface area of particle) both of the same volume} \tag{7}$$

In this study the sphericity factor was 0.67.

Following the determination of the particle terminal velocity, the exponent n was related to Re_t by the following correlation:

$$n = 4.64\ (Re_t)^{-0.055} \tag{8}$$

The small change of n with Re_t as previously reported by Richardson and Zaki is expected for a range of Reynolds numbers of 1 to 200.

Another approach was also followed in correlating the data. It is based on the use of the Galileo and Reynolds numbers as previously discussed by Ramamurthy. In relating a function of Re and Ga with $[1 - 1.21(1 - \epsilon_l)^{2/3}]^{-1}$, it is expected that as $\epsilon_l \to 1$ the correlation should approach Equation 3. For this reason, all data were correlated using the function Re/(0.132 × $Ga^{0.714}$). The results are shown in Figure 4. It is observed that the following correlation exists:

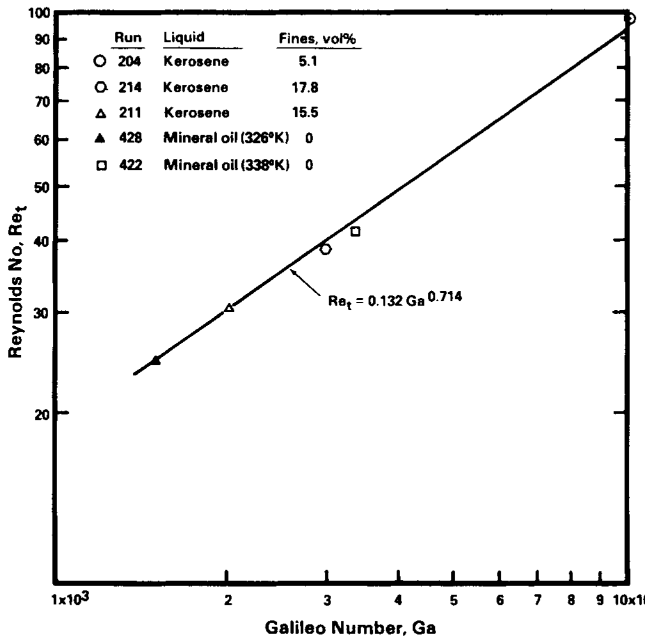

Figure 3. Terminal velocity Reynolds number.

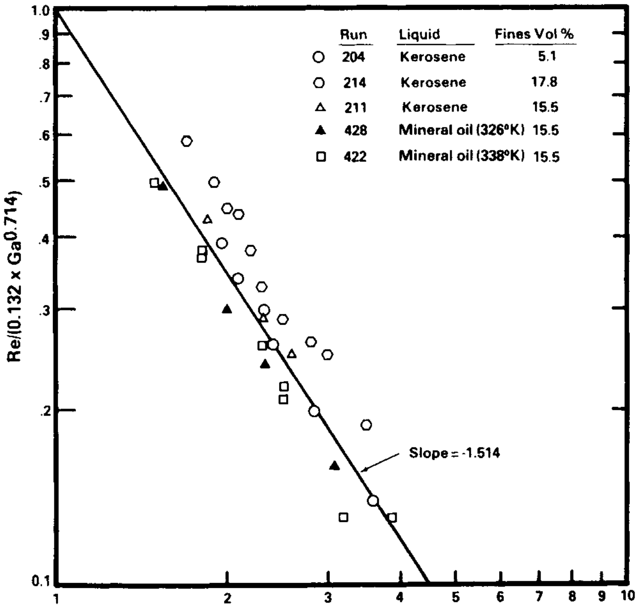

Figure 4. Correlation of liquid holdup data.

$$Re/0.132\ Ga^{0.714} = \Phi_\epsilon{}^{-1.514} \quad (9)$$

where:

$$\Phi = [1 - 1.21\,(1 - \epsilon_l)^{2/3}]^{-1} \quad (10)$$

This relationship is slightly different from that reported by Ramamurthy and Sabbaraju (1973) for the intermediate flow region.

A comparison of the Richardson and Zaki approach with Eq. 9 was made for a bed of catalyst particles (1.8 mm × 5.1 mm) fluidized with mineral oil at 353°K. At these conditions, the density and viscosity of mineral oil were found equal to 810 kg/m^3 and 0.0042 Ns/m^2, respectively. Using Equations 6 and 8, the particle terminal velocity and index n were found equal to 14.2 cm/s and 3.65, respectively. A comparison of experimental values of ϵ_l with the correlations presented in this study is shown in Table 3. Although both correlations predict the liquid holdup fairly accurately, Eq. 9 offers a more general approach for those cases where n changes rapidly with Reynolds number and other systems parameters. Because these correlations are based on only one particle size, future work will investigate the effect of sphericity factor Φ_s on the structure of the equations.

ACKNOWLEDGMENT

This work was conducted under a contract from the Department of Energy (Contract DE-AC05-77ET-10149). Permission to publish this work is gratefully acknowledged.

NOTATION

A = cross-sectional area of reactor, m^2
d = particle diameter, cm
d_s = equivalent diameter of a sphere, cm
g = acceleration of gravity, cm/s^2
Ga = Galileo number
H = catalyst bed height, m
M = mass of dry catalyst to reactor, kg
n = Richardson-Zaki index n or exponent,
Re_t = particle Reynolds number based on U_t
U_l = superficial liquid velocity, cm/s
U_t = terminal velocity of an isolated particle, cm/s

Greek Letters

ϵ_l = volume fraction of liquid
ρ = density of liquid or slurry, kg/m^3
ρ_c = density of a dry catalyst particle, kg/m^3
ρ_s = density of liquid-soaked catalyst particle, g/cm^3
μ = liquid or slurry viscosity, Ns/m^2
Φ = correlation parameter (Eq. 10)
Φ_s = sphericity of catalyst particle (Eq. 8)

TABLE 2. CATALYST PARTICLE TERMINAL VELOCITIES

Run No.	U_t, cm/s	Re_t	Ga
204	15.37	99	10,405
211	9.78	31	2,033
214	10.63	39	2,985
422	11.52	42	3,342
428	10.14	25	1,501

TABLE 3. COMPARISON OF RICHARDSON-ZAKI WITH EQ. 1: LIQUID-MINERAL OIL

U_l, cm/s	ϵ_l, Experimental	ϵ_l Predicted R-Z	ϵ_l Predicted Eq. 8
2.5	0.60	0.62	0.58
3.0	0.65	0.65	0.61
4.5	0.73	0.73	0.70
6.0	0.82	0.79	0.78
1.6	0.51	0.55	0.49

LITERATURE CITED

Lewis, E. W., E. W. Bowerman, "Fluidization of Solid Particles in Liquids," *Chem. Eng. Prog.*, **48**, 603 (1952).

Richardson, J. F., W. N. Zaki, "Sedimentation and Fluidization: Part I," *Trans. Inst. Chem. Eng.*, **32**, 35 (1954).

Ramamurthy, K., K. Sabbaraju, "Bed Expansion Characteristics of Annular Liquid Fluidized Beds," *Ind. Eng. Chem. Proc. Des. Devel.*, **12**, 184 (1973).

Vasalos, I. A., E. M. Bild, D. F. Tatterson, "Modeling the Fluid Dynamics of the H-Coal Reactor," presented at the 87th Annual AIChE Meeting, Boston (August 19–22, 1979).

Vasalos, I. A., E. M. Bild, D. N. Rundell, and D. F. Tatterson, "Study of Ebullated Bed Fluid Dynamics for H-Coal," Final Technical Progress Report, Contract DE-AC05-77ET-10149 (April 16, 1980).

Manuscript received October 20, 1980; revision received January 21, and accepted February 10, 1981