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THE EFFECT OF ELEVATED PRESSURE ON
FLUIDIZATION PHENOMENA

Final Report

By
Wen-ching Yang
Dale L. Keairns

Work Performed Under Contract No. AC22-82PC53036

Westinghouse Electric Corporation
Pittsburgh, Pennsylvania

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Final Report for Contract DE-AC22-82PC53036

to

Pittsburgh Energy Technology Center
Department of Energy
Pittsburgh, Pennsylvania

by

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1. INTRODUCTION AND SUMMARY

Under contract to the Pittsburgh Energy Technology Center (PETC), Department of Energy, the Westinghouse Electric Corporation, at its R&D Center in Pittsburgh, has studied the effect of elevated pressure on fluidization phenomena. The program consists of five tasks:

- Task 1 - Review of Literature on Fluidization Velocities
- Task 2 - Extension of Existing Fluidization Correlations
- Task 3 - State-of-the-Art Review Concerning Flow Regime Transitions
- Task 4 - Flow Regime Data Analysis
- Task 5 - Identification of Data Gaps

This report summarizes the results obtained during the contract period.

A critical review of the literature concerning the effect of elevated pressure and particle characteristics on minimum fluidization velocity, beginning fluidization velocity, minimum bubbling velocity, total fluidization velocity, and complete fluidization velocity was performed. Fluidization data from various sources were compiled and compared. A fundamental correlation with improved methodology was developed for estimating the minimum fluidization velocity at elevated pressure with good agreement. The developed correlation was found to be applicable at elevated temperature as well.

The state-of-the-art of the effect of pressure and particle characteristics on flow regime transitions from bubbling to turbulent fluidization and from turbulent to fast fluidization were reviewed. Available data on the effect of elevated pressure and particle characteristics on flow regime transitions were analyzed, however, the information available in the literature was meager at best.

The critical fluidization development areas were identified as the instrumentation development and the experiments carried out with reactive systems, under elevated temperature and pressure, and in large scale fluidization units. Potential research areas which are within the capability of the existing PETC high pressure test facility were also summarized. This was compiled on the basis of a literature survey on currently available literature information in those areas.

2. LITERATURE REVIEW

The available computerized literature search was obtained from the Engineering Index from 1970 to 1982 and from Chemical Abstracts from 1972 to 1982. For publications earlier than 1970, the author's personal file was used. The results are summarized here.

2.1 Effect of Pressure on Minimum Fluidization and Minimum Bubbling Velocities

2.1.1 Review of Experimental Studies

The experimental evidence of the effect of pressure on minimum fluidization velocity and minimum bubbling velocity reported in the literature was reviewed, and the results are summarized in Table 1. There are some conflicting observations, as is usual, involving a complex fluidizing system with different particle characteristics under different pressures. The following observations can be reported, however.

- o For fine particles similar to Geldart's Group A powders (Geldart, 1973), the U_{mf} is independent of pressure. The U_{mf} decreases with increasing pressure for particles well within Group B. For particles in the transition region, the pressure has slight effect. Most conflicting observations were reported for those particles at the transition region and were due primarily to the uncertainty of the effect of pressure on the boundary between Group A and Group B powders and to the possible effect of particle size distribution.
- o The effect of pressure on the U_{mf} can be estimated using Ergun's equation (1952) or its variations, such as that suggested by Wen and Yu (1966_b). If the voidage at minimum fluidization and the particle shape are known, the original Ergun equation gives a better correlation. Otherwise, the Wen and Yu type of correlations with correlation constants suggested by different authors (Wen and Yu, 1966_b; Babu et al., 1978; and Chitester et al., 1982) can be used.

Table 1

SUMMARY OF LITERATURE INFORMATION ON THE EFFECT OF PRESSURE ON THE
MINIMUM FLUIDIZATION VELOCITY, U_{mf} , AND THE MINIMUM BUBBLING VELOCITY, U_{mb}

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPa	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Chitester et al. (1982)	10.2 cm dia. 3-D bed	(1) Coal (1.25 g/cm^3) $(d_p) = 88 \mu\text{m}$ $d_{pH} = 140 \mu\text{m}$ $181 \mu\text{m}$ $182 \mu\text{m}$ $358 \mu\text{m}$ $361 \mu\text{m}$	101, 2170, 4240, 6300	$\neq f(P)$ $\neq f(P)$ $\neq f(P)$ Decreases slightly with P Decreases monotonically with P Decreases nonmonotonically with P	N.A.
		(2) Char (1.12 g/cm^3) $157 \mu\text{m}$ $156 \mu\text{m}$ $374 \mu\text{m}$		Decreases slightly with P Decreases slightly with P Decreases monotonically with P	
		(3) Ballotini (2.47 g/cm^3) $56 \mu\text{m}$ $102 \mu\text{m}$		$\neq f(P)$ Decreases slightly with P	
Rowe et al. (1982)	7.6 cm dia. 3-D bed	(1) FCC catalyst as received $(d_p) = 70 \mu\text{m}$ $d_{pH} = (0.819 \text{ g/cm}^3)$	up to 2000	$\neq f(P)$	$(U_{mb} - U_{mf})$ Increases with P
		(2) FCC catalyst, fines free $(d_p) = 82 \mu\text{m}$ $d_{pH} = (0.819 \text{ g/cm}^3)$		$\neq f(P)$	$(U_{mb} - U_{mf})$ Increases with P

Table 1 (continued)

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPa	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Sobreiro & Monteiro (1982)	4.5 cm dia. column	(1) Ballotini (2.3 g/cm ³) (d_p) = 125 μ m 177 μ m 250 μ m	100 to 3500	$\neq f(P)$ Decreases slightly with P Decreases with P	$\neq f(P)$ Decreases slightly with P Decreases with P
		(2) Alumina (2.8 g/cm ³) (d_p) = 88 μ m 125 μ m 177 μ m		$\neq f(P)$ $\neq f(P)$ Decreases with P	Increases with P $\neq f(P)$ $\neq f(P)$
		(3) Pyrrhotite (4.0 g/cm ³) (d_p) = 125 μ m		Decreases with P	Decreases with P
Banczyk & Wozniak (1981)	2.4 cm dia. column	Glassbeads (2.467 g/cm ³) d_p = 1200 μ m	up to 606	Decreases with P	N.A.
King (1979)	10 cm dia.	(1) Ballotini (2.82-2.98 g/cm ³) d_p = 61 μ m, N ₂ 64 μ m, CO ₂ 101 μ m, N ₂ 475 μ m, N ₂	up to 2500	$\neq f(P)$ Decreases slightly with P Decreases slightly with P Decreases with P	N.A. Increases with P N.A. N.A.
		(2) Sand (2.66 g/cm ³) d_p = 81 μ m, N ₂ 88 μ m, N ₂		N.A. Decreases slightly with P	Increases with P $\neq f(P)$
		(3) Polymer (0.55 g/cm ³) d_p = 688 μ m, N ₂		Decreases with P	N.A.

Table 1 (continued)

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPa	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Varadi & Grace(1978)	31 cm x 1.6 cm 2-D bed	Sand(2.65 g/cm ³) $d_p = 250-295 \mu m$	up to 2200	Decreases slightly with P	Increases with P; ($U_{mb} - U_{mf}$) Increases linearly with P
Guedes de Carvalho et al. (1978)	10 cm dia. column	(1)Ballotini(2.9 g/cm ³) (d_p) = 64 μm 211 μm	up to 2800	N.A.	$\neq f(P)$ Decreases slightly with P
		(2)Sand(2.69 g/cm ³) (d_p) = 74 μm		N.A.	$\neq f(P)$
2-4 Crowther and Whitehead (1978)	2.7 cm dia. column	(1)Synclust (0.9 g/cm ³) (d_p) = 63 μm	up to 6900	N.A.	Increases slightly with P; particulate fluidization at 6900 kPa
		(2)Coal (1.3 g/cm ³) (d_p) = 63 μm		N.A.	Increases slightly with P; particulate at 5600 kPa
		57 μm			Increases slightly with P; particulate at 5600 kPa
		27 μm			Particulate fluidization
		19 μm			Particulate fluidization

Table 1 (cont'd)

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPa	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Saxena & Vogel (1977)	15.2 cm dia. column	dolomites(1.90-3.19g/cm ³) (d_{pH}) = 717 μ m	179-834	Decreases with P	N.A.
Knowlton (1974)	29.2 cm dia. column	(1)anthracite(1.47 g/cm ³) (d_{pH}) = 780 μ m	up to 6900	Decreases with P	N.A.
		(2) Illinois No. 6 coal (1.29 g/cm ³) (d_{pH}) = 290 μ m		Decreases with P	N.A.
		(3)Lignite(1.17 g/cm ³) (d_{pH}) = 270 μ m		Decreases with P	N.A.
		(4)Hydrogasified Lignite (1.12 g/cm ³) (d_{pH}) = 230 μ m		Decreases with P	N.A.
		(5)FMC Char(1.18 g/cm ³) (d_{pH}) = 240 μ m		Decreases with P	N.A.
		(6)Coke Breeze (1.57 g/cm ³) (d_{pH}) = 320 μ m		Decreases with P	N.A.
		(7)Siderite (3.91 g/cm ³) (d_{pH}) = 290 μ m		Decreases with P	N.A.

Table 1 (continued)

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPa	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Godard and Richardson (1968)	10 cm dia. x 90 cm high	(1) Diakon (1.184 g/cm ³) (d_p) _H = 125 μ m (d_p) _H = 186 μ m (2) Phenolic resin (0.24 g/cm ³)	up to 1414	$\neq f(P)$ $\neq f(P)$	Increase markedly between 101 to 810 kPa; less rapidly at higher pressures ($U_{mb} - U_{mf}$) increases with P
Chekhov et al., (1964)	2.5 cm dia column	(1) Coke (N.A.) $d_p = 80 \mu$ m $d_p = 240 \mu$ m $d_p = 410 \mu$ m $d_p = 830 \mu$ m (2) Bronze (N.A.) $d_p = 119 \mu$ m	up to 30300	$\neq f(P)$ Decreases slightly with P Decreases markedly with P between 101 to 8000 kPa Decreases markedly with P between 101 to 3000 kPa $\neq f(P)$	N.A. N.A.

Table 1 (continued)

SOURCE	BED DIMENSIONS	PARTICLE CHARACTERISTICS	PRESSURE RANGE, kPA	EXPERIMENTAL OBSERVATION	
				U_{mf}	U_{mb}
Sechenov and Altshuler (1959)	4 cm dia. x 50 cm high	Aluminum Silicate Catalyst	up to 1515		
		$d_p = 200 \mu m$ (1.72 g/cm ³)		Decreases with P	N.A.
		$d_p = 670 \mu m$ (1.66 g/cm ³)		Decreases with P	N.A.
		$d_p = 1750 \mu m$ (1.34 g/cm ³)		Decreases with P	N.A.

Key for Table 1:

N.A. = Not Available

* f(P) = independent of pressure

P = pressure

- o The data scatter on U_{mf} reported in the literature is probably due to differences in particle shape, size distribution, and their resulting differences in voidage at minimum fluidization. Several researchers have reported difficulty in reproducing the ϵ_{mf} experimentally.
- o There is more disagreement among different authors about the minimum bubbling velocity, U_{mb} , than there is about U_{mf} , presumably because subjective observation of initial bubbling in the bed is involved. The U_{mb} of particles of similar characteristics has been reported both to increase and to decrease with increasing pressure. It is generally agreed, however, that the excess gas required for bubbling ($U_{mb} - U_{mf}$) increases with increasing pressure.

2.1.2 Review of Existing Correlations

The effect of pressure on the U_{mf} was usually estimated using Ergun's equation or its variations, such as that suggested by Wen and Yu (1966b). The Ergun equation can be written as

$$\frac{\Delta P}{L} = \frac{150(1-\epsilon)^2 \mu U}{\phi^2 \epsilon^3 d_p^2} + \frac{1.75(1-\epsilon)U^2 \rho_f}{\epsilon^3 \phi d_p} \quad (1)$$

At the minimum fluidization condition, the pressure drop across the bed can be expressed as

$$\frac{\Delta P}{L} = (\rho_s - \rho_f) (1 - \epsilon_{mf}) g \quad (2)$$

Equating Equations (1) and (2), a quadratic equation in $(Re)_{mf}$ can be obtained

$$1.75 A (Re)_{mf}^2 + 150 B (Re)_{mf} = Ar \quad (3)$$

(kinetic term) (viscous term)

where

$$A = \frac{1}{\phi \epsilon_{mf}^3} ; \quad B = \frac{1 - \epsilon_{mf}}{\phi^2 \epsilon_{mf}^3} \quad (4)$$

The shape-voidage functions A and B have been approximated by Wen and Yu (1966b) to be

$$A \approx 14 ; \quad B \approx 11 \quad (5)$$

Equation (3) can be rearranged into

$$(Re)_{mf} = \sqrt{(C_1)^2 + C_2 Ar} - C_1 \quad (6)$$

where

$$(Re)_{mf} = \frac{d_p U_{mf} \rho_f}{\mu}, \quad \text{the Reynolds number} \quad (7)$$

$$Ar = \frac{d_p^3 \rho_f (\rho_s - \rho_f) g}{\mu^2}, \quad \text{the Archimedes number} \quad (8)$$

The two constants, C_1 and C_2 , in the original Wen and Yu equations are 33.7 and 0.0408 respectively. Several other sets of constants are also proposed. They are summarized below.

<u>Reference</u>	<u>C_1</u>	<u>C_2</u>
Wen and Yu (1966b)	33.7	0.0408
Richardson (1971)	25.7	0.0365
Babu et al., (1978)	25.25	0.0651
Grace (1982)	27.2	0.0408
Chitester et al., (1982)	28.7	0.0494

For low Reynolds number where $(Re)_{mf} < 20$, U_{mf} can be calculated from Equation (3) by neglecting the kinetic term.

$$U_{mf} = \frac{d_p^2 (\rho_s - \rho_f) g}{1650 \mu} \quad (Re)_{mf} < 20 \quad (9)$$

Since the gas viscosity is usually independent of the pressure and the particle density is much larger than the fluid density, $\rho_s \gg \rho_f$, in most cases, the U_{mf} of small particles do not usually change with the pressure as verified from the experimental evidence cited in Table 1.

For large Reynolds number, the viscous term is negligible and

$$U_{mf}^2 = \frac{d_p (\rho_s - \rho_f) g}{24.5 \rho_f} \quad (Re)_{mf} > 1000 \quad (10)$$

For large or dense particles, increasing the pressure will reduce the minimum fluidization velocity of the particles.

If the voidage at minimum fluidization and the particle shape are known, the original Ergun equation (1952) gives a better prediction. Unfortunately, in most instances, neither information is available. The accuracy of U_{mf} predictions based on the Ergun type of equations depends on the selection of the constants C_1 and C_2 in Equation (6). Several sets of constants C_1 and C_2 have been suggested by different authors based on different data bases as summarized earlier. The most recent set of constants was suggested by Chitester et al. (1982), based on the data obtained for pressures up to 6300 kPa for bed materials of coal, char, and ballotini. In view of so many sets of constants already suggested so far, it is safe to expect that more sets of constants will be proposed in the future for different data bases. For fundamental predictions of the fundamental fluidization property such as the minimum fluidization velocity, a more fundamental approach is necessary. Other correlations available in the literature for prediction of the minimum fluidization velocity have been surveyed by Babu et al. (1978) and Grewal and Saxena (1980).

Geldart and Abrahamsen (1978) proposed to calculate the minimum bubbling velocity as follows

$$\frac{U_{mb}}{U_{mf}} = \frac{4.125 \times 10^4 \mu^{0.9} \rho_f^{0.1}}{(\rho_s - \rho_f) g d_p} \quad (11)$$

Equation (11) predicts that the ratio of U_{mb}/U_{mf} is proportional to pressure to the 0.1 power. This prediction agreed well with the data by Richardson (1971). It also indicates a much more dependence of U_{mb}/U_{mf} on the viscosity which is consistent with the data of de Vries et al., (1972). Geldart and Abrahamsen (1980) modified Equation (11) later to become

$$\frac{U_{mb}}{U_{mf}} = \frac{2300 \rho_f^{0.126} \mu^{0.523} \exp(0.716F)}{\bar{d}_p^{0.8} g^{0.934} (\rho_s - \rho_f)^{0.934}} \quad (12)$$

The numerical constant in Equation (12) has units of $\text{kg}^{0.285} \text{m}^{-0.167} \text{s}^{-1.345}$. They also proposed an equation for calculating the minimum bubbling velocity as shown below

$$U_{mb} = 2.07 \bar{d}_p \rho_f^{0.06} \exp(0.716F)/\mu^{0.347} \quad (13)$$

For small alumina powders, Sobreiro and Monteiro (1982) found that the minimum bubbling velocity increased with increasing pressure and could be correlated by the following equation.

$$U_{mb} = 43 (\rho_f/\mu)^{0.1} d_p \quad (14)$$

Equation (14) has the units of m for d_p , kg/m^3 for ρ_f , and Ns/m^2 for μ . This is very much comparable to that suggested by Geldart and Abrahamsen (1978) as shown below

$$U_{mb} = 33 (\rho_f/\mu)^{0.1} d_p \quad (15)$$

2.2 Beginning Fluidization and Total Fluidization Velocities

A typical pressure drop-gas velocity profile for a mixture with a size distribution or a density difference will look like that shown in Figure 1 with a characteristic hysteresis between the branch with increasing gas velocity and the branch with decreasing gas velocity

because of particle segregation. Generally, three fluidization velocities can be defined as shown in Figure 1, where U_{bf} is the beginning fluidization velocity and U_{tf} , the total fluidization velocity. The U_{bf} and the U_{tf} were originally proposed by Kondukov and Sosna (1965) and Gelperin et al. (1967; 1971). They defined the velocity at which fine or light particles started to fluidize as the velocity of beginning fluidization, U_{bf} , and the velocity at which all the large or heavy particles began to fluidize as the velocity for total fluidization, U_{tf} . Segregation of particles occurs at a velocity between these two values. Figure 1 shows the determination of these two velocities from the fluidization curve of a mixture.

Recent publications by Chiba et al. (1979) and Yang and Kearns (1982) indicated that the pressure profiles shown in Figure 1 are not unique and are actually dependent on the segregating tendency of the mixture. Idealized pressure profiles for a binary mixture are shown in Figure 2. For an ideal system, where the particles are of small size difference and of equal density, both the ascending and descending portions of the fluidization curve will coincide, shown as curve (a) in Figure 2. For a highly segregating mixture where the particle segregation rate is fast, the fluidization curve will follow curve (a) for the ascending portion but will descend along curve (b), because the mixture will already have separated into two distinct layers of flotsam and jetsam when the gas velocity is higher than the minimum fluidization velocity of the jetsam. The descending portion of the fluidization curve of any real systems lying between these two extreme cases will follow any curves between (a) and (b), such as curve (c), depending on the rate of separation and the time spent on obtaining the fluidization curve. A unique U_{bf} , thus, may not be generally obtainable from the fluidization curve as shown in Figure 1. A bed operated at U_{mf} as defined in Figure 1 is clearly not completely fluidized, i.e., a portion of the bottom is still stagnant. In this case, the minimum fluidization velocity, U_{mf} , loses its physical significance because it no longer

Curve 651878-A

Mixtures are initially well mixed as a fixed bed. These mixtures are transformed to a well mixed fluidized bed when the gas velocity is increased. Curve OBC2 describes the changes of ΔP with the gas velocity. When the gas velocity is gradually reduced, the ΔP follows the lower curve and the mixture is transformed back to a fixed bed

Figure 1 - Pressure Drop versus Gas Velocity for a Bed of Two Particle Sizes

Figure 2 - Effect of Mixing/Segregation State on Descending Portion of the Fluidization Curve

- (a) — · — Perfectly Mixed
- (b) - - - Completely Segregated
- (c) — — Partially Mixed/Segretated

represents the hydrodynamic demarcation between a packed bed and a fluidized bed.

Usually, if the particles in the bed have a size ratio of more than 1.3, there is a tendency for the particles to segregate during fluidization, especially at close to the minimum fluidization velocity (Wen and Yu, 1966a). The particle segregation can generally be detected by visual observation if the vessel wall is transparent or by comparison of the maximum pressure drop obtained with W/A where W is the weight of the bed and A is the cross-sectional area of the bed. Theoretically, when all the particles in the bed are suspended by the gas stream, the measured pressure drop should be equal to W/A . Because of imperfection in the distributor plate design and channelling, the measured pressure drop may be slightly less. If the measured pressure drop is substantially less, particle segregation must be suspected.

Vaid and Sen Gupta (1978) suggested two equations for estimating both U_{bf} and U_{tf} for a mixture of up to five components in liquid fluidized beds as shown in Equations (16) and (17).

$$(Re)_{bf} = [(18.1)^2 + 0.0192 Ar]^{0.5} - 18.1 \quad (16)$$

$$(Re)_{tf} = [(24)^2 + 0.0546 Ar]^{0.5} - 24 \quad (17)$$

where

$$Re = \bar{d}_p U_p \rho_f / \mu \quad (18)$$

$$Ar = \bar{d}_p^3 (\bar{\rho}_s - \rho_f) \rho_f g / \mu^2 \quad (19)$$

$$\bar{d}_p = 1 / \sum (x_i / d_{pi}) \quad (20)$$

$$\bar{\rho}_s = \sum (x_i \rho_{si}) \quad (21)$$

No study was found in the literature on the effect of pressure on U_{bf} and U_{tf} . Equations (16) and (17) were derived primarily on the

basis of data from liquid fluidized beds. Their application to gas fluidized beds needs to be verified. A recent study by Yang and Keairns (1982) indicated that U_{bf} calculated from Equation (16) followed the trend of but was not equal to the true minimum fluidization velocity of the mixture. The U_{tf} calculated from Equation (17), however, did not have any physical significance in the acrylic/dolomite system examined in that the bed was not fluidized at U_{tf} because of the segregation of the heavier component. U_{tf} is better defined for the highly segregating systems, such as the acrylic/dolomite system, by setting the U_{tf} equal to the U_{mf} of the heavier or larger component. When a wide size distribution of bed material is involved, the definition of U_{bf} and U_{tf} is uncertain.

2.3 Effect of Pressure on Complete Fluidization Velocity

In studying the fluidization characteristics of particles of wide size distribution, Knowlton (1977) noted that the minimum fluidization velocity determined according to Figure 1 is a misnomer, because the bed was not completely fluidized at U_{mf} . He then visually determined the so-called complete fluidization velocity (U_{cf}) to be the velocity at which the bed became completely fluidized (i.e., all particles are mobile). The determined U_{cf} is usually considerably larger than the U_{mf} determined through Figure 1. He suggested that the complete fluidization velocity of a mixture of wide size distribution be calculated using the following equation.

$$U_{cf} = \sum X_i U_{mfi} \quad (22)$$

The minimum fluidization velocity of the individual particle size fraction, $(U_{mf})_i$, in Equation (22) was calculated using the Wen and Yu equation (1966b).

For estimating the complete fluidization velocity, U_{cf} , the method shown in Equation (22) works reasonably well for pressure up to ~ 7 MPa (Knowlton, 1977). Yang (1981) suggested an alternative method to calculate U_{cf} using the Ergun equation expressed in the following way.

$$Ar = 150 \frac{1 - \epsilon_{cf}}{\epsilon_{cf}^3} (Re)_{cf} + 1.75 \frac{1}{\epsilon_{cf}^3} (Re)_{cf}^2 \quad (23)$$

where

$$Ar = \frac{\rho_f(\rho_s - \rho_f)g\bar{d}_p^3}{\mu^2} ; \quad (Re)_{cf} = \frac{\bar{d}_p U_{cf} \rho_f}{\mu} \quad (24)$$

Equation (24) was first fitted to the experimental data at the lowest pressure to calculate the voidage at complete fluidization, ϵ_{cf} . The voidage was then assumed to be independent of pressure and Equation (24) was used to calculate the U_{cf} at any pressure. The U_{cf} calculated using this method agreed with the data obtained by Knowlton (1977) very well.

The complete fluidization velocity, U_{cf} , first defined by Knowlton (1977) was based totally on visual observation and thus tended to be subjected to subjective error. It may be an important design limitation but hardly a well-defined characteristic of a fluidized bed. Except for the references cited here, his definition has not received a wide acceptance in fluidization research. A more quantitative and well-defined complete fluidization velocity can be derived by setting it equal to the U_{mf} of the heavier or larger component in the bed material, as in the case of U_{tf} discussed earlier.

3. DEVELOPMENT OF A NEW CORRELATION FOR ESTIMATING THE MINIMUM FLUIDIZATION VELOCITY AT ELEVATED PRESSURE AND TEMPERATURE

It was pointed out in the last section that use of the Wen and Yu equation (1966b) for calculating the minimum fluidization velocity at elevated pressure is not satisfactory in that more than 5 sets of empirical constants have been proposed for use in the equation. The accuracy is uncertain for particles with characteristics outside the range of the data correlated by different authors. An approach with more consistent results is required.

3.1 Development of a Correlation for Estimating the Minimum Fluidization Velocity at Elevated Pressure

Although the Wen and Yu equation has been widely used in the fluidization literature, it is hardly recognized that the equation is simply an expression of Reynolds number — drag coefficient ($Re - C_D$) relationship, first pointed out by Zenz and Othmer (1960). By equating the drag force and the gravitational force for a single spherical particle, we have

$$\frac{\pi}{8} C_D \rho_f U^2 d_p^2 = \frac{\pi d_p^3}{6} (\rho_s - \rho_f) g \quad (25)$$

or

$$C_D = \frac{4}{3} \frac{d_p (\rho_s - \rho_f) g}{\rho_f U^2} \quad (26)$$

It can be readily found that

$$Ar = \frac{4}{3} Re^2 C_D \quad \text{for spherical particles} \quad (27)$$

The Wen and Yu type of equation shown in Equation (6) can be rearranged to give

$$C_D = \frac{(Re + C_1)^2 - C_1^2}{\frac{3}{4} C_2 Re^2} \quad (28)$$

Equation (28) is simply another expression for relationship between Re and C_D .

In 1973, Barnea and Mizrahi proposed a general correlation for pressure drop through fixed beds of spherical particles based on a discrete particle model (rather than pipe flow analogy) corrected for particle interaction. They extended the standard C_D vs. Re curve for single sphere to multi-particle systems by incorporating proper functions of the volumetric particle concentration. The modified Reynolds number and drag coefficient they suggested are

$$(Re)_\epsilon = Re \left[\frac{1}{\epsilon \exp \{5(1-\epsilon)/3\epsilon\}} \right] \quad (29)$$

$$(C_D)_\epsilon = C_D \frac{\epsilon^3}{1 + (1-\epsilon)^{1/3}} \quad (30)$$

By substituting the standard Re - C_D curve for single sphere with the modified $(Re)_\epsilon$ and $(C_D)_\epsilon$ expressed in Equations (29) and (30), the standard Re - C_D can be applied for multi-spherical particle systems. The suggested correlation was extended by Barnea and Mednick (1975) for calculation of the minimum fluidization velocity. For easy graphical presentation, they suggested to plot the dimensionless diameter $(Re^2 \cdot C_D)_{mf}^{1/3}$ vs. the dimensionless velocity $(Re/C_D)_{mf}^{1/3}$. At minimum fluidization condition, these two dimensionless groups can be expressed as

$$\left(\frac{(Re)_\epsilon}{(C_D)_\epsilon} \right)_{mf}^{1/3} = \left(\frac{Re}{C_D} \right)_{mf}^{1/3} \left\{ \frac{[1 + (1 - \epsilon_{mf}^{1/3})]^{1/3}}{\epsilon_{mf}^{4/3} \exp \left(\frac{5}{9} \frac{(1 - \epsilon_{mf})}{\epsilon_{mf}} \right)} \right\} \quad (31)$$

$$\left((Re)_\epsilon^2 (C_D)_\epsilon \right)_{mf}^{1/3} = (Re^2 C_D)_{mf}^{1/3} \left\{ \frac{\epsilon_{mf}^{1/3}}{\left[1 + (1 - \epsilon_{mf})^{1/3} \right]^{1/3} \exp \left(\frac{10(1 - \epsilon_{mf})}{9 \epsilon_{mf}} \right)} \right\} \quad (32)$$

Equations (31) and (32) were combined with the standard drag correlations recommended by Clift et al. (1978) and summarized in Table 2 for multi-particle systems at different voidage fractions at minimum fluidization shown in Figures 3 and 4. The Wen and Yu correlation (1966b) and the correlation by Chitester, et al. (1982) are also included for comparison.

The experimental minimum fluidization velocity data obtained at the PETC using different sizes of coal, char and ballotini as bed material and operated at pressures of 101, 2169, 4238 and 6306 kPa are also presented in Figures 3 and 4. The coal sizes of harmonic mean diameters of 88 μm and 140 μm can best be correlated with the curve for $\epsilon_{mf} = 0.48$ (see Figure 3). The experimentally observed voidages at minimum fluidization for these particles range from 0.45 to 0.51 as shown in Table 3. The ballotini particles can be correlated fairly well with the curve for $\epsilon_{mf} = 0.44$ (Figure 4). This compares with the experimental voidages of 0.40 to 0.45. The rest of the data can be correlated with the curve for $\epsilon_{mf} = 0.46$ as compared to experimental values of 0.44 to 0.47 (except that of coal particles of 361 μm and all char particles). All char particles have consistently higher experimental voidage at minimum fluidization. It is not known whether the difficulty in determining the particle density of the porous char particles has contributed to this consistent discrepancy.

The fact that the minimum fluidization velocities at different pressures for the same particles can be correlated with a single curve

Table 2

RECOMMENDED DRAG CORRELATIONS: STANDARD DRAG CURVE, $w = \log_{10} Re$

Range	Correlation
(A) $Re < 0.01$	$C_D = 3/16 + 24/Re$
(B) $0.01 < Re \leq 20$	$\log_{10} \left[\frac{C_D Re}{24} - 1 \right] = -0.881 + 0.82w - 0.05w^2$ $\text{i.e., } C_D = \frac{24}{Re} [1 + 0.1315 Re^{0.82 - 0.05w}]$
(C) $20 \leq Re \leq 260$	$\log_{10} \left[\frac{C_D Re}{24} - 1 \right] = -0.7133 + 0.6305w$ $\text{i.e., } C_D = \frac{24}{Re} [1 + 0.1935 Re^{0.6305}]$
(D) $260 \leq Re \leq 1500$	$\log_{10} C_D = 1.6435 - 1.1242w + 0.1558w^2$
(E) $1.5 \times 10^3 \leq Re \leq 1.2 \times 10^4$	$\log_{10} C_D = -2.4571 + 2.5558w - 0.9295w^2 + 0.1049w^3$
(F) $1.2 \times 10^4 < Re < 4.4 \times 10^4$	$\log_{10} C_D = -1.9131 + 0.6370w - 0.0636w^2$

From Bubbles, Drops, and Particles by Clift, R., Grace, J. R., & Weber, M. E., Academic Press, New York (1978)

Table 3

PRESSURE EFFECT ON BED VOIDAGE AT
MINIMUM FLUIDIZATION FOR VARIOUS PARTICLES

ϵ_{mf}		P, (kPa)			
		101	2169	4238	6306
Harmonic Mean Diameter for Coal, (μm)	88	0.49	0.50	0.51	0.50
	140	0.45	0.45	0.46	0.46
	181	0.46	0.46	0.47	*
	182	*	0.45	0.45	*
	358	0.44	0.46	0.46	0.47
	361	0.44	0.53	0.53	0.53
Harmonic Mean Diameter for Char, (μm)	157	0.51	0.53	0.53	0.53
	196	0.50	*	0.53	0.53
	374	0.55	0.57	0.58	0.58
Harmonic Mean Diameter for Ballotini, (μm)	96	0.40	0.41	0.42	0.42
	102	0.45	0.45	0.44	0.45

* Data are not available.

at a constant ϵ_{mf} provides a simple methodology for estimating the minimum fluidization velocity at elevated pressure accurately. The following procedure is suggested. The material of interest is first subjected to experimentation at atmospheric pressure to determine its minimum fluidization velocity. This data point is then located on the proper curve of constant ϵ_{mf} in a plot similar to that shown in Figures 3 and 4. The minimum fluidization velocities at any pressure for the same material can be calculated readily from the same curve.

The coal, char and ballotini particles employed by the PETC for their minimum fluidization velocity studies are by no means of narrow size distribution. The harmonic mean particle sizes are used for the present analysis. There is legitimate justification for using the harmonic mean particle size in the discrete particle model employed here because the drag on the particles is proportional to the specific particle surface area per unit bed volume. The harmonic mean particle size can be theoretically derived to be the reciprocal of the specific particle surface area per unit bed volume. The successful correlation obtained with materials of reasonably wide size distribution here implies that the proposed methodology can also be applied for bed materials of relatively wide size distribution as long as the harmonic mean particle size is used in the correlation.

The curves shown in Figures 3 and 4 are obtained from $Re-C_D$ relationship for spherical particles. For particles of irregular shape, the correction for $Re-C_D$ is available. The determination of shape factor for different materials is difficult. This is especially true when particles of wide size distribution are involved where the shape of the particles may be different for different sizes of particles. Judging from the fact that the experimental voidages at minimum fluidization conditions seem to agree fairly well with the voidages of the particular curves fitting the particular sets of data as shown in Figures 3 and 4, the effect of the particle shape may very well be of second order when the bed is at its loosest-packed condition at minimum fluidization.

3.2 Application of the Correlation to Estimating the Minimum Fluidization Velocity at Elevated Temperature

The fundamental correlation developed for estimating the U_{mf} at elevated pressure was also found to be applicable at elevated temperature. Although this is not specifically in the contract, testing of the developed correlation at elevated temperature is important in identification of future potential areas of fluidization research.

The correlation was applied to four sets of minimum fluidization velocity data obtained at elevated temperature, those by Singh et al. (1973), Mii et al. (1973), Yoshida et al. (1976), and Pattipati and Wen (1981). They are presented in Figures 5 through 7 with encouraging results. The experimental conditions and the particle characteristics employed by these authors are summarized in Table 4. It can now be concluded that the correlation and the methodology proposed to estimate the minimum fluidization velocity at elevated pressure is a fundamental approach and can be equally applicable for estimating the minimum fluidization velocity at elevated temperature for spherical particles and relatively roundish particles such as catalyst and sand. For particles substantially deviate from the spherical particles such as the graphite particles with a sphericity of 0.65 used by Mii et al. (1973), the correlation does not apply as well.

TABLE 4 - SUMMARY OF MINIMUM FLUIDIZATION VELOCITY DATA AT ELEVATED TEMPERATURE
EMPLOYED FOR TESTING THE CORRELATION

<u>References</u>	<u>Bed Material</u>			<u>Fluidizing Medium</u>	<u>Temperature Range (°K)</u>
	<u>Material</u>	<u>Size (μm)</u>	<u>Density (kg/m³)</u>		
Mii et al. (1973)	Graphite	300	1970	air and nitrogen	293 - 1073
Singh et al. (1973)	Sand	-250+150	2650	air, argon, steam acetylene, hydrogen, natural gas	288 - 973
Yoshida et al. (1976)	Catalyst	151, 214	1540	air	293 - 673
	Sand	151	2580	air	293 - 673
Pattipati and Wen (1981)	Sand	240	2630	air	293 - 1073
		462	2630	air	293 - 1023
		1310	2630	air	373 - 1123
		3376	2630	air	318 - 773

4. REVIEW OF EXPERIMENTAL AND THEORETICAL STUDIES ON FLOW REGIME TRANSITIONS AND DATA ANALYSIS

4.1 Different Regimes of Fluidization

There are at least five different regimes of fluidization observed experimentally. Within each particular regime, there are also fine distinctions easily recognized as distinct subregimes. If fast fluidization, which resembles very much to that of the dense phase pneumatic transport requiring continuous feed of solids into the bed, can be recognized as a separate fluidization regime, the dilute phase pneumatic transport (entrained bed) can also be classified as a distinct fluidization regime. The different regimes of fluidization are summarized in Table 5. The particulate fluidization which usually occurs only in liquid fluidized beds is not included in this discussion. From the experimental evidence available in the literature so far, the regimes of fluidization for coarse particles and fine particles appear to be substantially different, and they will be discussed separately.

4.1.1 Fluidization Regimes of Coarse Particles

For gas fluidization of large particles, Catipovic et al (1978), divided the regimes into slow bubbles, fast bubbles, and rapidly growing bubbles regimes. The qualitative flow regime diagram is shown in Figure 8.

Demarcation between fast and slow bubbles - the transition between the fast and the slow bubbles regimes is marked by the following equation.

$$\frac{U_{mf}}{\epsilon_{mf}} = U_B = \left(\frac{g D_B}{2} \right)^{1/2} \quad (33)$$

Figure 8 - Regimes of Fluidization for Large Particles

Table 5 REGIMES OF FLUIDIZATION

<u>Regime</u>	<u>Velocity Range</u>	<u>Principal Features</u>
Fixed Bed	$0 \leq U < U_{mf}$	Particle stationary; flow through interstices
Particulate Fluidization	$U_{mf} \leq U < U_{mb}$	Bed expands homogeneously; no apparent bubbles; bed surface well defined with little pressure fluctuation
Bubbling Bed	$U_{mb} \leq U < U_{ms}$	Visible bubble voids; bubble growth and coalescence apparent; bed surface and pressure fluctuate with bubbles
Slugging Bed	$U_{ms} \leq U < U_k$	Bubble voids occupy most of bed cross-section; bed surface and pressure fluctuate with regularity of slugs
Turbulent Fluidization	$U_k \leq U < U_{tr}$	No well-defined discrete bubble voids; solids move in clusters or streamers; small pressure fluctuation
Fast Fluidization	$U_{tr} \leq U$	Solids are transported and have to be recycled; no apparent bubbles; solids move in clusters or strands
Dilute Phase Pneumatic Transport	$U_{tr} < U$	Solids are transported in dilute phase with voidage usually > 0.94 for large particles and $> 0.75 - 0.80$ for fine particles

Regime of rapidly growing bubbles - In this regime, the bubble growth rate is of the same magnitude as the bubble rising velocity

$$\frac{d(D_B)}{dt} = \frac{d(D_B)}{dh} \cdot \frac{dh}{dt} = U_B \quad (34)$$

for shallow beds

$$\frac{d(D_B)}{dh} \approx 1 \quad (35)$$

If the bubble diameter vs the bed height relationship is available, the transition to this fluidization regime can be obtained. For example, if the bubble diameter correlation by Cranfield and Geldart (1974) shown in Equation (36) is used, the boundary of the rapidly growing bubble regime can be expressed as in Equation (37)

$$D_B = 0.0326 (U - U_{mf})^{1.11} h^{0.81} \quad (36)$$

$$U - U_{mf} = 21.85 h^{0.17} \quad (37)$$

where the units are D_B in cm, h in cm, and U in cm/sec. By this definition, the fluidized beds for certain bed materials may be in different fluidization regimes at different bed heights. The turbulent regime was not discussed by Catipovic et al. (1978), however.

Canada et al. (1978), and Canada and McLaughlin (1978), studied the fluidization modes of two different sizes of glass beads (650 μm and 2600 μm) in different sizes of bed at pressures up to 10 atmospheres and found that there were four definable flow regimes, the bubbly flow, the plug flow, the apparent slug flow, and the turbulent flow regimes. The turbulent flow regime was defined based on both visual observation of the bed and the bed pressure fluctuation measurement. Visually, no identifiable gas bubbles of familiar shape, but irregular gas voids were observed in the turbulent regime. Transition to turbulent regime also corresponded to decreases in bed pressure fluctuation from the maximum

which usually occurred during the apparent slug flow regimes. The transition to turbulent flow was observed to occur at

$$U = 0.55 \text{ to } 0.65 U_t \quad \text{for } 650 \text{ } \mu\text{m particles} \quad (38)$$

and

$$U = 0.30 \text{ to } 0.35 U_t \quad \text{for } 2600 \text{ } \mu\text{m particles} \quad (39)$$

These transition velocities were independent of the operating pressure and the bed height. Since the velocity ratio of U_{mf}/U_t for 650 μm particles is about 0.07 and for 2600 μm particles is about 0.1, transition to turbulent fluidization can occur at about 8 to 9 times the U_{mf} for 650 μm particles and at 3 to 4 times the U_{mf} for 2600 μm particles. This is qualitatively consistent with the photographic determination of flow regime transitions conducted by PETC as shown in Figure 9 for pressures up to about 1500 kPa. At pressures higher than 3000 kPa, the PETC data show that the transition velocity is relatively constant and equal to about 2 to 3 times the U_{mf} of the particles. The PETC data also indicate that transition to the turbulent regime does not have to pass the slug flow regime at pressures higher than about 2500 kPa. The transition from the bubbly flow to the turbulent flow regimes directly is possible. Of course, this transition point may depend on the bed height and bed size which should be studied further.

Another interesting observation by Canada et al. (1978) was that the average bed void fraction at the transition to turbulent flow was always constant at about 75% with both the small and large particles employed in their experiment. There is underlying importance for this observation which deserves further experimental and theoretical investigation.

4.1.2 Fluidization Regimes of Fine Particles

A qualitative fluidization map for fine particles was initially proposed by Yerushalmi et al. (1978) and refined later in Yerushalmi and Cankurt (1979). A more recent and more general fluidization regime

Figure 9 - Flow Regime Transitions Observed at PETC at Elevated Pressure for Char Particles

diagram was presented by Li and Kwauk (1980) with actual experimental data using fine alumina particles as bed material. A schematic presentation of the flow regime diagram following that of Li and Kwauk (1980) is shown in Figure 10. The log-log coordinates are used here, instead of the original semilogarithmic presentation, to show the approximately straight-line relation in different regimes.

Use of the voidage as the ordinate and the gas velocity as the abscissa allows a convenient presentation for fluidization regimes from incipient fluidization all the way to dilute-phase pneumatic transport, as shown in Figure 10. Different fluidization regimes simply exhibit different voidage-gas velocity relations.

The Transition between Bubbling and Turbulent Fluidization - In a recent analysis Yang (1982) incorporated the theory of particle agglomeration proposed by Capes (1974) and the continuity wave theory to arrive at a transition criterion between the bubbling and turbulent fluidization regimes. Capes (1974) modified the Richardson and Zaki equation (1954) to include an "effective" porosity to give

$$\frac{U}{U_i} = \epsilon_e^m = (1 - KC)^m \quad (40)$$

The "effective" porosity, ϵ_e , includes the void space between aggregates (or clusters) only while ϵ , the apparent voidage, includes also the void space between solid particles inside the clusters. K is the volume of aggregates (clusters) per unit volume of solid and C is the volumetric fraction of solids. The continuity wave velocity in a dispersion of particles was derived by Wallis (1969) as

$$v_w = j + \frac{\partial}{\partial \epsilon_e} (j_{fs}) \quad (41)$$

where

$$j = j_f + j_s \quad (42)$$

Figure 10 - Schematic Fluidization Regime Diagram (following Li and Kwauk, 1980)

Combining Equations (40) to (42), we have

$$V_w = U_{im} (1 - \epsilon_e) \epsilon_e^{m-1} . \quad (43)$$

For detail derivation, refer to the paper by Yang (1982) which is included here as an appendix.

Yerushalmi and Cankurt (1979) observed experimentally that the transition from bubbling fluidization to turbulent fluidization is reflected in the fluctuations of both the local dynamic pressure and the total pressure drop across the bed. The transition can be characterized by two velocities, U_c and U_k . U_c is the velocity at which the pressure fluctuation is the maximum, and U_k marks the onset of turbulent fluidization, where the pressure fluctuations begin to level off. The pressure fluctuations during this transition observed by Yerushalmi and Cankurt (1979) correspond to the turbulent fluidization regime recorded by Li and Kwauk (1980).

In this analysis the bubbling-to-turbulent fluidization transition was assumed to occur at the maximum continuity wave velocity. Differentiating equation (43) and setting it to zero, we have:

$$\frac{dV_w}{d\epsilon_e} = U_{im} [(m-1) \epsilon_e^{m-2} (1-\epsilon_e) - \epsilon_e^{m-1}] = 0 , \quad (44)$$

or

$$\epsilon_e = \frac{m-1}{m} \text{ at the bubbling-to-turbulent-fluidization transition.} \quad (45)$$

At an "effective" porosity, ϵ_e , less than $(m-1)/m$, the continuity wave velocity, V_w , increases with increases in ϵ_e . When ϵ_e is larger than $(m-1)/m$, V_w decreases with increases in ϵ_e . The bubbling-to-turbulent fluidization transition thus occurs at an "effective" porosity, shown in equation (45), or at an apparent porosity, shown in equation (46):

$$\epsilon = 1 - \frac{1-\epsilon_e}{K} \text{ at bubbling-to-turbulent-fluidization transition.} \quad (46)$$

The calculated voidages from Equation (46) at transition between the bubbling and turbulent fluidization regimes were compared with that experimentally observed with reasonable agreement (Yang, 1982).

Empirical Correlations for Transition between Bubbling and Turbulent

Fluidization - The mechanistic model proposed for transition between the bubbling and turbulent fluidization outlined in equation (46) cannot be applied a priori because the dependence of the cluster size on operating conditions and particle characteristics is unknown. We found, however, that for the three finer particles used by Yerushalmi and Cankurt (Solids A, B and C), the transition points between bubbling and turbulent fluidization can be correlated with high accuracy in the following equations:

$$K = 0.883 (Re)_t^{-0.485} \quad \begin{array}{l} \text{correlation coefficient} \\ = -0.9994 \end{array} \quad (47)$$

$$m = 2.31 (Re)_t^{-0.0547} \quad \begin{array}{l} \text{correlation coefficient} \\ = -0.9979 \end{array} \quad (48)$$

$$\frac{U_i}{U_t} = 25.49 (Re)_t^{-0.582} \quad \begin{array}{l} \text{correlation coefficient} \\ = -0.9880 \end{array} \quad (49)$$

In the absence of other correlations, equations (47) to (49) can be used as quantitative criteria to predict the transition between bubbling and turbulent fluidization for fine particles.

The Transport Velocity - When the operating gas velocity is higher than a so-called transport velocity, a sharp increase in carry-over occurs and the bed becomes essentially an entrained bed. New feed or recycling of solids is required to maintain operation of the bed. Depending on the operating gas velocity and the solid loading, the transport reactor can be either in dilute-phase pneumatic transport or in dense-phase pneumatic transport, also called fast fluidized bed, as shown in Figure 10. In this analysis we have assumed the transport velocity of each system to be equal to the terminal velocity of the cluster, U_i . The results are in fair agreement with those observed experimentally (Yang, 1982).

Demarcation between Dilute-Phase Pneumatic Transport and Fast Fluidization - From all the experimental evidence reported in the literature, Yang (1982) concluded that the transition between the dilute-phase pneumatic transport and the fast fluidization regime could be approximated by the choking correlations developed for pneumatic transport.

Yang (1983) recently proposed an empirical correlation for choking using data from Yousfi and Gau (1974), who experimentally determined the gas velocity and the voidage at choking. The correlation is shown below

$$\frac{U}{U_t} = \epsilon_c^q \quad (50)$$

where

$$q = -33.2 + 1.18 (\text{Re})_t \text{ for } (\text{Re})_t < 23 \quad (51)$$

and

$$(\text{Re})_t = \frac{d_p U_t \rho_f}{\mu} . \quad (52)$$

The results from Equations (50) - (52) compared favorably with the experimental data obtained by Li and Kwauk (1980) on the transition between dilute-phase pneumatic transport and fast fluidization. A very recent advance on hydrodynamic correlations for fast fluidization was reported by Li, et al. (1982).

5. HIGH PRIORITY EXPERIMENTAL PROGRAMS AND DATA GAPS

In the authors' opinion, the following areas of fluidized bed technology are in need of priority studies.

A. Instrumentation Development

Rowe and Masson (1981) compared the conclusions on the bubble properties drawn based on results obtained by X-ray and by the intrusive probe under similar operating conditions and found that there were substantial differences. The discrepancy might be due to improper design of the probe and/or the wrong conclusions inferred from the probe measurement. Thus the development direction should be toward the miniaturization of the probe to lessen the disturbance the probe might have on the fluidization behavior and to improve the interpretation of the measurement results. To attend these objectives the probe development should be in conjunction with a non-intrusive technique such as X-ray or movies such that the probe results can be confirmed independently. Instruments are also required for measurement in hostile environment such as in elevated temperature and pressure, and in systems with chemical reactions.

B. Elevated Temperature and Pressure Environment

Most of the fluidization phenomena were studied at ambient conditions. Substantial effort should be directed toward studies under elevated temperature and pressure, especially under the conditions where both high temperature and high pressure are present at the same time.

C. Studies on Reactive Systems

Studies on hydrodynamics of the fluidized beds were usually carried out in non-reactive systems. The local generation of heat and exchange of mass are usually taken care of mathematically without experimental verification. If proper instrumentation and technique could be developed, direct studies of reactive systems might be possible.

D. Large Scale Experimental Units

Before scaleup criteria can be thoroughly understood, experiments in large scale units are absolutely essential. Studies using both large and small units will help to identify the similarity criteria.

Other critical areas of investigation were also reviewed by Wen and Dutta (1977) recently.

A literature search was conducted during this contract to identify the potential areas of research which are within the capability of the current PETC high-pressure test facility. The results of the search is summarized in Table 6. The lack of information at elevated pressure is apparent in all areas of fluidization. Transitions between different fluidization regimes at elevated pressure are completely absent. This obviously is a high priority research area which can be studied effectively using the PETC facility. Other high priority experimental programs which can be studied effectively at PETC are summarized in Table 7.

TABLE 6

Fluidization Phenomena and Transition Data Base

Area	Literature Base	Literature (Elevated Pressure)	Representative Elevated Pressure Data	
			Pressures Up To	Solids
Fluidization Velocities	241*	20	53 atm	char, coal, catalyst, siderite, sand
Spouting Phenomena	184	19	20 atm	catalyst, sand, ballotini
Bubble Phenomena	368	12	28 atm	catalyst, ballotini, sand
Jet Phenomena	77	4	53 atm	char, siderite, sand
Slugging Transition	2	0	atmospheric	glass, plastics, sand
Bubbling-Turbulent Transition	5	0	atmospheric	catalyst, dicalite, sand, glass
Particle Separation	61	1	5 atm	char/dolomite

- NOTES: 1. The literature search was conducted for the Chemical Abstracts from 1967-1982 and the Engineering Index from 1970-1981.
2. Since the search was conducted based on keywords, a single paper can be cited several times. Also discussion for a particular phenomenon is superficial for some references. Only about half of the cited references can be expected to deal with the particular phenomenon in depth.
3. The literature at elevated pressure has been considered by Westinghouse. The pressure data on phenomena related to coal gasification has been incorporated into present correlations and models. This provides a basis for identifying data needs for the PETC program.

* number of references

TABLE 7

PETC Candidate Experimental Programs

Area	Property	Variables	Measurements
● Fluidization Phenomena	● fluidizing velocities (minimum fluidization velocity, minimum bubbling velocity, total fluidization velocity, complete fluidization velocity	● pressure ● particle type ● particle size	● pressure drop profiles and fluctuations ● visual observation
	● Spouting Phenomena (minimum spouting velocity, minimum spouting bed height)	● pressure ● bed height ● particle characteristics	● pressure drop profiles and fluctuations ● movies
	● Bed Expansion/Deaeration Time	● pressure ● bed height ● particle characteristics	● bed height = $f(t)$ ● pressure drop profiles ● visual observations
	● Bubble/Jet Phenomena (bubble size, frequency, and velocity; jet penetration)	● pressure ● bed height ● particle characteristics ● jet velocity ● solid loadings (jets) ● jet nozzle	● movies (including top?) ● capacitance probe ● pressure drop profiles ● optical probe (Note: 2-D and 3-D units)

TABLE 7 (Contd)

Area	Property	Variables	Measurements
● Fluid- Particle System Transitions	● Bubbling-Turbulent	<ul style="list-style-type: none"> ● pressure ● gas velocity ● particle characteristics 	<ul style="list-style-type: none"> ● pressure drop profiles and fluctuations ● bed expansion ● movies <p>(Note: 2-D and 3-D units)</p>
	● Bubbling-Slugging	<ul style="list-style-type: none"> ● pressure ● bed height ● particle characteristics 	<ul style="list-style-type: none"> ● slugging bed height ● slugging frequency ● minimum bed height for non-slugging ● pressure drop profiles and fluctuations ● visual observation/movies
	● Particle Separation	<ul style="list-style-type: none"> ● pressure ● gas velocity ● particle density, size, shape 	<ul style="list-style-type: none"> ● particle concentration profiles (e.g., resistivity probe) ● visual observations

6. NOTATION

A, B	= constants in Equations (3) and (4)
Ar	= Archimedes number, $Ar = d_p^3 \rho_f (\rho_s - \rho_f) g / \mu^2$
C	= volumetric fraction of solids, $C = 1 - \epsilon$
C_1, C_2	= two constants in Wen and Yu type equation shown in Equation (6)
C_D	= drag coefficient
$(C_D)_\epsilon$	= modified drag coefficient defined in Equation (30)
d_p	= particle diameter
$(d_p)_i$	= particle diameter of the fraction i
$\bar{d}_p$	= mean particle size from standard sieve analysis = $1/\sum (\lambda_i / (d_p)_i)$
D_B	= bubble diameter
F	= fines fraction less than 45 μm
g	= gravitational acceleration
h	= height of the bed
j	= volumetric flux
j_f	= volumetric flux of fluid
j_s	= volumetric flux of solids
j_{fs}, j_{sf}	= drift fluxes defined in the Appendix
K	= volume of aggregates (clusters) per unit volume of solid
m	= exponent in the modified Richardson and Zaki equation [equation (40)]
q	= exponent in equation (50) and defined in equation (51)
Re	= Reynolds number, $Re = d_p U_{pf} / \mu$
$(Re)_\epsilon$	= modified Reynolds number defined in Equation (29)
$(Re)_{bf}$	= Reynolds number based on beginning fluidization velocity
$(Re)_{cf}$	= Reynolds number based on complete fluidization velocity
$(Re)_{mf}$	= Reynolds number based on minimum fluidization velocity defined in Equation (7)

$(Re)_t$	= Reynolds number defined in Equation (52)
$(Re)_{tf}$	= Reynolds number based on total fluidization velocity
U	= superficial fluidization velocity
U_B	= bubble velocity
U_{bf}	= superficial beginning fluidization velocity
U_c	= transition velocity (see Appendix for definition)
U_{cf}	= superficial complete fluidization velocity
U_i	= terminal velocity of a cluster
U_k	= transition velocity (see Appendix for definition)
U_{mb}	= superficial minimum bubbling velocity
U_{mf}	= superficial minimum fluidization velocity
U_t	= terminal velocity of a single particle
U_{tf}	= superficial total fluidization velocity
V_w	= continuity wave velocity defined in Equation (43)
X_i	= weight fraction of particle size fraction i
ρ_f	= fluid density
ρ_s	= particle density
μ	= fluid viscosity
ϵ	= voidage
ϵ_c	= voidage at choking
ϵ_{cf}	= voidage at complete fluidization
ϵ_e	= "effective" voidage; $\epsilon_e = 1 - KC$
ϵ_{mf}	= voidage at minimum fluidization

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APPENDIX

MECHANISTIC MODELS FOR TRANSITIONS BETWEEN REGIMES OF FLUIDIZATION

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MECHANISTIC MODELS FOR TRANSITIONS BETWEEN REGIMES OF FLUIDIZATION

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INTRODUCTION

Understanding the flow regime transitions is important because different flow regimes provide vastly different solid circulation patterns, solids and gas mixing rates, and thus the chemical reaction rates. They, in turn, dictate the reactor design considerations. There are very few theoretical discussions and correlations available in the literature on the criteria of flow transitions between these different regimes. This paper attempts to develop mechanistic models to predict these flow transitions for fine particles generally belonging to Geldart's Class A. (Geldart, 1973).

A GENERAL FLUIDIZATION REGIME DIAGRAM

A qualitative fluidization map for fine particles was initially proposed by Yerushalmi et al. (1978) and refined later in Yerushalmi and Cankurt (1979). A more recent and more general fluidization regime diagram was presented by Li and Kwauk (1980) as shown schematically in Figure 1. The log-log coordinates are used here, instead of the original semilogarithmic presentation, to show the approximately straight-line relationship in different regimes, which will become more apparent later. Use of the voidage as the ordinate and the gas velocity as the abscissa allows a convenient presentation for fluidization regimes from incipient fluidization all the way to dilute-phase pneumatic transport. Different fluidization regimes simply exhibit different voidage-gas velocity relations.

THE THEORY OF CLUSTER FORMATION

The application of the concept of cluster formation was first suggested by Yerushalmi et al. (1978). Matsen (1982) also used it in his analysis of choking and entrainment mechanisms. A quantitative theory,

however, was first proposed by Capes (1974) for application to the fluidization of fine particles. He followed the development of the concept of flocs during sedimentation by Scott (1968) and modified the Richardson and Zaki equation (1954) to include an "effective" porosity, as shown in Equation (1).

$$\frac{U}{U_i} = \epsilon_e^m = (1 - KC)^m. \quad (1)$$

The "effective" porosity, ϵ_e , includes the void space between aggregates (or clusters) only while ϵ , the apparent voidage, includes also the void space between solid particles inside the clusters. He attributed the large values of n in the Richardson and Zaki equation often reported in the literature for fine particles (ranging up to a value of 10) to the increased effective particle volume caused by agglomeration or to the irregular particle shape, which immobilized a layer of fluid in the surface irregularities. Using Equation (1) with the "effective" voidage brings the exponent m within the range of the original Richardson and Zaki equation.

The bed density reported by Yerushalmi and Cankurt (1979) and the voidage-gas velocity relation reported by Li and Kwauk (1980) were processed to give the best fit using the original Richardson and Zaki equation and the modified one shown in Equation (1). The results are summarized in Table 1. The U_i can be considered to be the terminal velocity of a cluster with the "effective" particle diameter. In fitting Equation (1) to the experimental data, the procedure Capes (1974) suggested was followed. The calculation was carried out until the value of m and n agreed to within 0.02. The successful correlation of these experimental data indicates that, hydrodynamically, these fluid beds are similar to beds with "clusters" of finite sizes.

THE CONCEPT OF CONTINUITY WAVE AND THE TRANSITION BETWEEN BUBBLING AND TURBULENT FLUIDIZATION

The continuity wave exists as a quasi-steady-state phenomenon whenever there is a relation between the steady equilibrium flow rate of a substance and its concentration. The development of the basic equations was provided by Wallis (1969). The same approach is now employed here to predict the transition between bubbling and turbulent fluidization.

The drift flux between the fluid and the solids is defined as:

$$j_{fs} = -j_{sf} = (1 - \epsilon_e) j_f - \epsilon_e j_s \quad (2)$$

with the positive direction to be upward. The solids are assumed to form clusters with an "effective" porosity ϵ_e . The continuity wave velocity in a dispersion of particles was derived by Wallis (1969) as:

$$v_w = j + \frac{\partial}{\partial \epsilon_e} (j_{fs}); \quad j = j_f + j_s \quad (3)$$

In the bubbling and turbulent fluidization regimes no net transport of solids occurs and thus:

$$j_s = \frac{W_s}{\rho_s A} = 0 \quad \text{and} \quad j_f = \frac{Q_f}{A} = U. \quad (4)$$

The continuity wave velocity [Equation (3)] becomes

$$v_w = U + \frac{\partial}{\partial \epsilon_e} [(1 - \epsilon_e) U]. \quad (5)$$

Combining equations (1) and (5), we have the continuity wave velocity for the bubbling and turbulent fluidization regimes as

$$v_w = U_i^m (1 - \epsilon_e) \epsilon_e^{m-1}. \quad (6)$$

Yerushalmi and Cankurt (1979) observed experimentally that the transition from bubbling fluidization to turbulent fluidization is reflected

in the fluctuations of both the local dynamic pressure and the total pressure drop across the bed. The bubbling-to-turbulent-fluidization transition with maximum pressure fluctuation was assumed to occur at a voidage corresponding to the maximum continuity wave velocity in this analysis. Differentiating Equation (6) and setting it to zero, we have, at the bubbling-to-turbulent-fluidization transition,

$$\epsilon_e = \frac{m-1}{m} \quad \text{or} \quad \epsilon = 1 - \frac{1-\epsilon_e}{K} \quad (7)$$

At an "effective" porosity, ϵ_e , less than $(m-1)/m$, the continuity wave velocity, V_w , increases with increases in ϵ_e . When ϵ_e is larger than $(m-1)/m$, V_w decreases with increases in ϵ_e . Substituting the values of m , K , and U_i in Table 1 into Equations (1) and (7), we can calculate the gas velocity and voidage at transition and compare them with those observed experimentally in Table 2. With our current understanding of this phenomenon, the reasonable agreement shown in Table 2 is encouraging. The average diameter of the clusters at transition, the number of individual particles making up each cluster, and the density of each cluster, are also calculated and reported in Table 2.

THE TRANSPORT VELOCITY

When the operating gas velocity is higher than a so-called transport velocity, a sharp increase in carry-over occurs, and the bed becomes essentially an entrained bed. New feed or recycling of solids is required to maintain operation of the bed. Depending on the operating gas velocity and the solid loading, the transport reactor can be either in dilute-phase or in dense-phase pneumatic transport, also called fast fluidized bed, as shown in Figure 1. In this analysis we have assumed the transport velocity

of each system to be equal to the terminal velocity of the cluster, U_i . The results are compared with those observed experimentally in Table 2.

DEMARCATIION BETWEEN DILUTE-PHASE PNEUMATIC TRANSPORT AND FAST FLUIDIZATION

Yerushalmi and Cankurt (1979) observed pistonlike movement of solids for the larger solids they used in their experiments, a phenomenon similar to the slugging transition between dilute- and dense-phase pneumatic transport and predictable from existing choking correlations (Yang, 1976). For fine particles, Yousfi and Gau (1974) observed experimentally that the transition between dilute- and dense-phase transport did not produce solids slugs extending across the cross section of the pipe, but rather solids "bundles." It is also clear from the fluidization regime map experimentally produced by Li and Kwauk (1980) and shown in Figure 1 that the dilute-phase pneumatic transport and the fast fluidization (dense-phase transport bed) have similar characteristics in the voidage-gas velocity plot. The difference in dependence of the voidage and the gas velocity is due to the difference in the gas-solids slip velocity created by cluster formation. It is logical, then, to expect that the transition between the dilute-phase pneumatic transport and the fast fluidization regime can be approximated by the choking correlations developed for pneumatic transport.

Yang (1982) recently proposed an empirical correlation for choking using data from Yousfi and Gau (1974), who experimentally determined the gas velocity and the voidage at choking. The correlation is presented in Equation (8).

$$\frac{U}{U_t} = \epsilon^q ; \quad q = -33.2 + 1.18 (Re)_t \text{ for } (Re)_t < 23 \quad (8)$$

The results from Equation (8) compared favorably with the experimental data obtained by Li and Kwauk (1980) on the transition between dilute-phase pneumatic transport and fast fluidization, as shown in Table 3.

EMPIRICAL CORRELATIONS FOR TRANSITION BETWEEN BUBBLING AND TURBULENT FLUIDIZATION

The mechanistic model proposed for transition between the bubbling and turbulent fluidization cannot be applied a priori because the dependence of the cluster size on operating conditions and particle characteristics is unknown. We found, however, that for the three finer particles used by Yerushalmi and Cankurt (Solids A, B, and C), the transition points between bubbling and turbulent fluidization can be correlated with high accuracy by the following equations:

$$K = 0.883 (Re)_t^{-0.485} \text{ correlation coefficient} = -0.9994 \quad (9)$$

$$m = 2.31 (re)_t^{-0.0547} \text{ correlation coefficient} = -0.9979 \quad (10)$$

$$\frac{U_i}{U_t} = 25.49 (Re)_t^{-0.582} \text{ correlation coefficient} = -0.9880 \quad (11)$$

In the absence of other correlations, equations (9) to (11) can be used for quantitative criteria to predict the transition between bubbling and turbulent fluidization for fine particles.

NOTATIONS

- A = cross-sectional area of a fluidized bed or a pipe
- C = volumetric fraction of solids, $C = 1 - \epsilon$
- d_p = particle diameter
- j = mixture volumetric flux, $= j_f + j_s$

j_f, j_s = volumetric fluxes of fluid and solids respectively
 j_{fs}, j_{sf} = drift fluxes defined in Equation (2)
 K = volume of aggregates (clusters) per unit volume of solid
 m = exponent in the modified Richardson and Zaki equation [Equation (1)]
 n = exponent in the Richardson and Zaki equation
 q = exponent in Equation (8)
 Q_f = volumetric flow rate of the fluid
 $(Re)_t$ = Reynolds number, $= d_p U_t \rho_f / \mu$
 U = superficial fluid velocity
 U_i = terminal velocity of a cluster
 U_t = terminal velocity of a single particle
 v_w = continuity wave velocity defined in Equation (3)
 W_s = solid flow rate
 ϵ = voidage
 ϵ_e = "effective" voidage defined in Equation (1), $\epsilon_e = 1 - KC$
 μ = fluid viscosity
 ρ_f = density of the fluid
 ρ_s = density of the solids

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Fig. 1 -- Schematic fluidization regime diagram (following Li and Kwauk, 1980)

TABLE 1

RESULTS OF FITTING EXPERIMENTAL DATA WITH ORIGINAL AND MODIFIED RICHARDSON AND ZAKI
EQUATIONS (EQUATIONS 1 AND 2)

Data Source	Bed Material	n	Correlation Coefficient	m	K	U_i , cm/s	Correlation Coefficient
Yerushalmi & Cankurt (1979)	Dicalite 4200	16.34	0.9618	2.72	3.76	336	-0.9420
	FCC Catalyst	6.32	0.9891	2.50	1.75	398	-0.9876
	HFZ-20 Catalyst	5.15	0.9854	2.44	1.45	545	-0.9906
	No. 85 Sand	4.07	0.9584	2.39	1.39	814	-0.9703
Li & Kwauk (1980)	Fine Alumina	8.31	0.9061	3.03	1.90	232	-0.8797

Table 2

COMPARISON OF CALCULATED AND EXPERIMENTAL PARAMETERS AT BUBBLING-TURBULENT
FLUIDIZATION TRANSITION

Data Source	Bed Material	Transition Velocity, cm/s		Transition Voidage		Cluster Size, μm	Number of Particles in a Cluster	Density of the Cluster, g/cm ³	Transport Velocity, cm/s	
		Exp.	Calc.	Exp.	Calc.				Exp.	Calc.
Yerushalmi & Cankurt (1979)	Dicalite 4200	53-107	97	0.877-0.910	0.902	595	1.6×10^3	0.44	-	336
	FCC Catalyst	61	108	0.692	0.771	1461	1.5×10^4	0.61	120-150	398
	HFZ-20 Catalyst	91-137	151	0.640-0.716	0.718	1138	8.6×10^3	1.00	~210	545
	No. 85 Sand	274-550	223	0.750-0.860	0.699	1223	68	1.91	-	814
Li & Kwauk (1980)	Fine Alumina	~100	69	~0.85	0.826	272	67	1.66	~230	232

Table 3

COMPARISON OF EXPERIMENTAL AND CALCULATED TRANSITION POINT BETWEEN DILUTE-PHASE
PNEUMATIC TRANSPORT AND FAST FLUIDIZATION (DENSE-PHASE TRANSPORT)

Material - Fine Alumina (Li and Kwauk, 1980)

Superficial Particle Velocity, cm/s	Gas Velocity at Transition, cm/s	Voidage at Transition	
		Experimental	Calculated
0.076	250	0.985	0.949
0.491	310	0.975	0.942
1.81	415	0.945 - 0.970	0.933
2.57	450	0.935 - 0.950	0.931
4.05	475	0.930	0.929