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Gas Fluidization of Solids in a Stationary Liquid\*

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Abstract

Critical gas flow rates were measured for fluidizing initially static beds of particles and for settling of fluidized beds in a stationary pool of liquid. Experiments were conducted with beds of glass, nickel and  $UO_2$  particles ranging in size from 11 to 548  $\mu m$  in pools of water, ethanol, Freon-113 and water-glycerine solutions. Beds of particles smaller than 328  $\mu m$  were fluidized by a mechanism of individual particles being carried off of walls of channels in the bed, and beds of particles larger than 328  $\mu m$  slugged before breaking up to a fluidized state. The data were empirically correlated.

The suspension or fluidization of solid particles in a two-phase gas-liquid fluid has been of considerable interest in chemical processes and recently in nuclear safety analysis. Dhir and Catton [1] and Cho, et al. [2] analyzed heat removal from nuclear core debris assuming fluidization of the particulate debris in boiling coolant. Extensive work has been conducted on three phase fluidization with both the liquid and gas phases flowing [3]. However, in batch processes and certain special situations, such as boiling in a bed of nuclear core debris, gas bubbling through a bed of particles in a non-flowing or stationary pool of liquid must be considered.

The critical gas velocity required to completely suspend or fluidize all of the particles in a stationary pool has had some limited investigation. Kato [4] measured the air velocity to completely fluidize glass spheres (74 to 295  $\mu\text{m}$ ) and magnetite particles (74 to 175  $\mu\text{m}$ ) in 0.2 N sodium sulfite solution. Kato correlated the data with the following functional relationship:

$$\frac{U_{mf}}{U_m} = f [H U_m (\rho_s - \rho_l) / \mu_l] \quad (1)$$

Roy, et al. [5] correlated their data in terms of the maximum quantity of solids suspended as a function of the bubble velocity. Their empirically developed correlation depended on the superficial gas Reynolds number,  $Re_T$ , based on the tube diameter.

$$H_{max} = 6.84 \times 10^{-4} [Re_T N_B^{-0.23} \left(\frac{U_t}{U_B}\right)^{-0.18} (\gamma')^{-3} C_\mu] \quad Re_T < 500 \quad (2)$$

$$H_{max} = 1.072 \times 10^{-1} [Re_T^{0.2} N_B^{-0.23} \left(\frac{U_t}{U_B}\right)^{-0.18} (\gamma')^{-3} C_\mu] \quad Re_T < 600 \quad (3)$$

where  $C_{\mu} = 1 - 0.5892 \log \mu_{\ell} + 0.1026 \log^2 \mu_{\ell}$  ( $\mu_{\ell}$  in centipoise).

The data were obtained for various solid materials ranging in size from 127 to 675  $\mu\text{m}$  and water, alcohol, and various oils.

Narayanan et al. [6] considered the minimum gas fluidization velocity as a pickup velocity in which energy is transmitted from the gas phase to the liquid. The velocity to initiate particle suspension was theoretically determined to be:

$$U_{mf} + \frac{1}{3} \left[ 2g H_g H_{sl} \left( \frac{\rho_{\ell} - \rho_g}{\rho_{\ell}} \right) \right]^{1/2} = \left\{ 2g(\rho_s - \rho_{\ell}) \left[ \frac{2 D_p}{3 \rho_{\ell}} + \frac{H_s H_{sl}}{\rho_s + H_s \rho_{\ell}} \right] \right\}^{1/2} \quad (4)$$

A correction factor to adjust the theoretical value of  $U_{mf}$  to that experimentally observed was empirically correlated as a function of solids concentration.

The gas holdup,  $H_g$ , was empirically determined to be:

$$H_g = 0.062 U \quad U < 6.7 \text{ cm/s} \quad (5)$$

$$H_g = 0.133 U^{0.38} \quad 6.7 < U < 21.34 \text{ cm/s} \quad (6)$$

Their data was based on 84 to 508  $\mu\text{m}$  quartz in water.

Imafuku et al. [7] determined the critical gas velocity to completely suspend the particles in liquid by observing the pressure drop. As for the case of single phase fluidization the pressure drop across the bed increases with gas velocity until the particles are fluidized. After incipient fluidization the pressure drop is independent of gas velocity. They worked with

various solid particles ranging from 64 to 180  $\mu\text{m}$  in size and water and glycerine solutions.

While the above cited efforts make a significant contributions in analysis and data, correlations of incipient fluidization in the three-phase system are basically empirical. The mechanisms of transfer of momentum from the gas phase to the liquid and particulate phases are not as simple as for a single-fluid phase flowing through a particle bed. Scheidegger [8] has summarized a considerable amount of work on two-phase flow through packed beds. Tutu, et al. [9] recently experimentally and theoretically analyzed the forces involved in momentum transfer from a flowing gas to beds of 3.18, 6.35, and 12.7 mm stainless steel spheres in a pool of water. These forces include the gas-solid drag, liquid-solid drag, gas-liquid drag, and the surface binding force. This work on two-phase flow through packed beds is not completely applicable to conditions of incipient fluidization. The gas flow through unconfined beds of smaller and less dense particles will force open channels in the bed (this phenomena is described below). This changes the original packing structure and alters the flow distribution through the bed. This structural change as well as the various interacting forces involved severely complicates modelling concepts. However, if the packing remains fixed, the modeling can be simplified. Cho, et al. [2] observed flooding to occur in their bed of rather large dense particles (387 to 650  $\mu\text{m}$  copper) before fluidization began. Therefore, a correlation developed by Wen and Yu [10] for single-phase fluidization could be used since the flooded bed contained essentially no liquid phase. The flooding velocities were calculated using the correlation of Sherwood and Lobo as reformulated by Wallis [11] and were consistently below the observed and calculated incipient fluidization velocities for their copper particle beds.

Where structural changes occur in the bed, i.e., channel formation, correlations of incipient three-phase fluidization are primarily empirical. Implicit in an empirical correlation is its limitation to the range of variables in the data base. This work contributes additional data to expand the data base and observations which should aid in developing models of incipient three-phase fluidization.

### Experimental

Experiments were conducted with Lucite columns 49.3, 50 and 100 mm in diameter (see Fig. 1). Beds of 127 to 548- $\mu\text{m}$  glass, 311- $\mu\text{m}$  nickel, and 11- $\mu\text{m}$   $\text{UO}_2$  particles were fluidized. Water, Freon-113, glycerine-water solutions, and ethanol were used for the liquid phase. The liquid level was at least 50 mm above the bed under static conditions (no air flow). In these tests, the incipient fluidization velocities remained the same with this liquid level or higher levels. Air was dispersed at the base of the beds by a 0.312 mm thick sintered stainless steel plate with an average pore opening of 10  $\mu\text{m}$ . Pressure taps were in the plenum beneath the gas distributor plate and at the base of the bed just above the distributor plate. The pressure above the plate was measured with a Statham Laboratories PM6TCb pressure transducer, and the pressure below the plate with a Data Sensors, Inc. PB415 D-10 pressure transducer.

The bed pressure drop was determined with the pressure tap just above the distributor plate. The line to this tap had to be continuously flushed with liquid to keep air bubbles from backing into it. The pressure drop across the distributor plate proved inconsistent from experiment to experiment apparently because of partial capillary clogging of the pores in the distributor plate. The air flow was gradually increased until the bed was fluidized. Fluidization was indicated by observation as well as by the pressure drop leveling off without further increase as the air flow was increased. This behavior, which

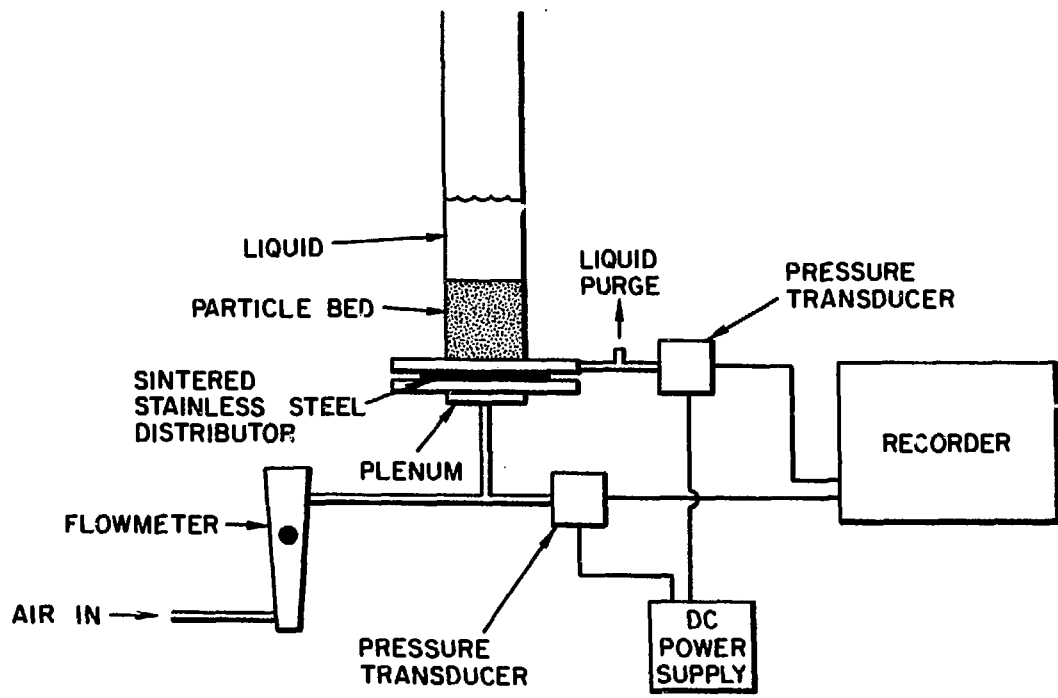


Figure 1. Apparatus for Determining Minimum Fluidization Velocities.

was also observed by Imafuku et al. [7], is identical to fluidization by a single phase fluid. Assuming negligible resistance to air flow by the liquid phase the pressure drop for the fluidized bed is equivalent to the weight of the particle bed. During fluidization the effective density of the liquid phase is increased by the weight of the suspended particles and this is reflected in the pressure at the base.

### Results

The minimum fluidization velocity is plotted as a function of static bed depth in Figures 2 to 4. Fig. 2 describes glass-water systems, Fig. 3 glass-Freon-113 systems, and Fig. 4 11- $\mu\text{m}$   $\text{UO}_2$  in water. A complete data tabulation is given in the Appendix. There is a significant difference in behavior between glass particles 327.5  $\mu\text{m}$  in diameter and larger in water and for particles 310  $\mu\text{m}$  and smaller. The incipient fluidization velocity for particles 310  $\mu\text{m}$  in diameter and smaller is independent of bed depth, whereas it is dependent on bed depth for particles 327  $\mu\text{m}$  in diameter and larger.

The air flow forced open channels (~ 5 mm wide) in the particle beds of both large and small particles. For beds of the smaller particles, the particles were stripped off the channel walls until all the particles were suspended. Since the particles were individually removed from the channel walls and carried up into suspension in the liquid pool above the bed, the velocity required for incipient fluidization was independent of the bed depth. However, with larger particles the entire bed was observed to slug up and down off the base before complete breakup to minimum fluidization. Since the entire bed was elevated before large particle fluidization, the incipient fluidization velocity was dependent on bed depth (or bed weight).



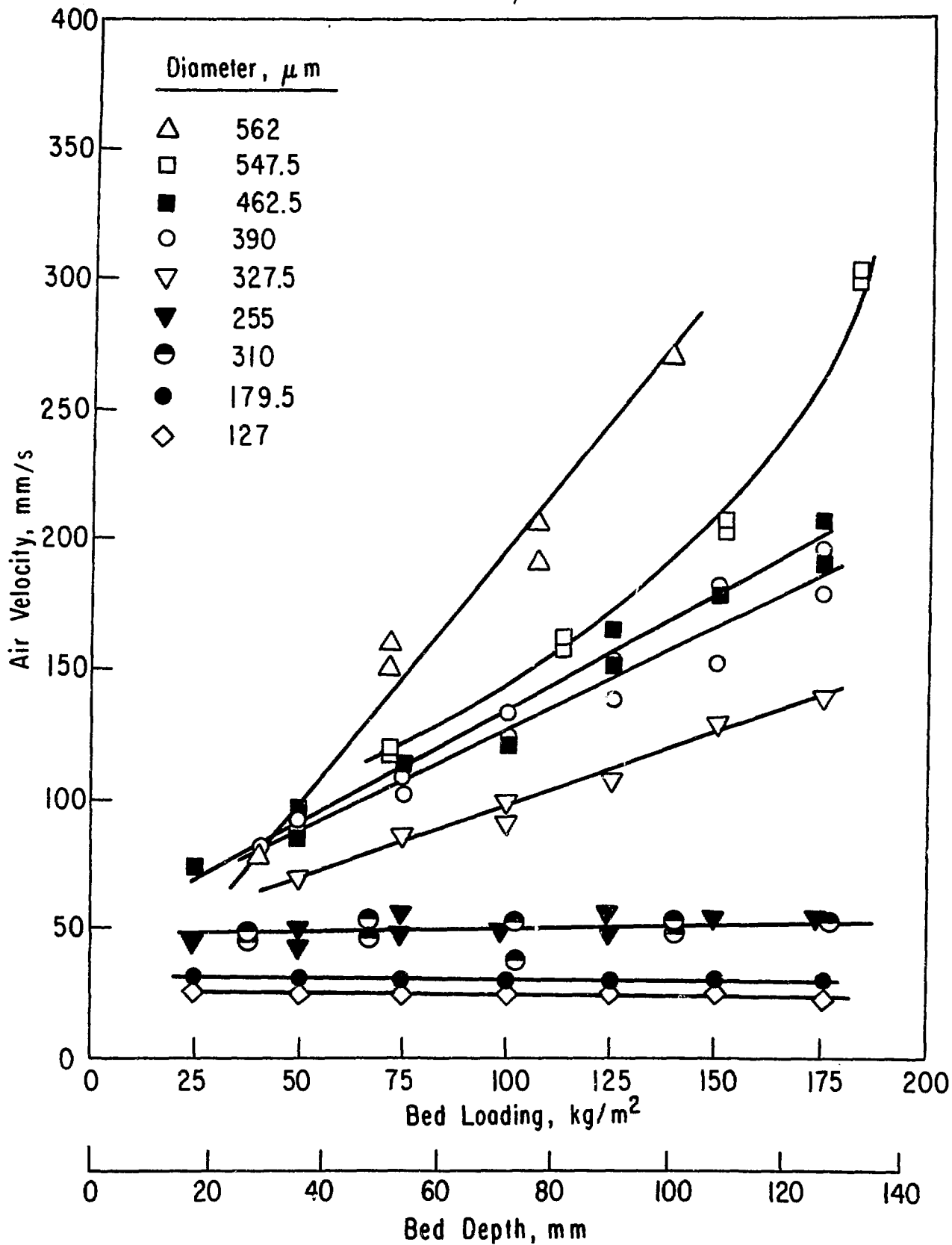


Figure 2. Air Velocities Required to Fluidize Water-Glass Beds.

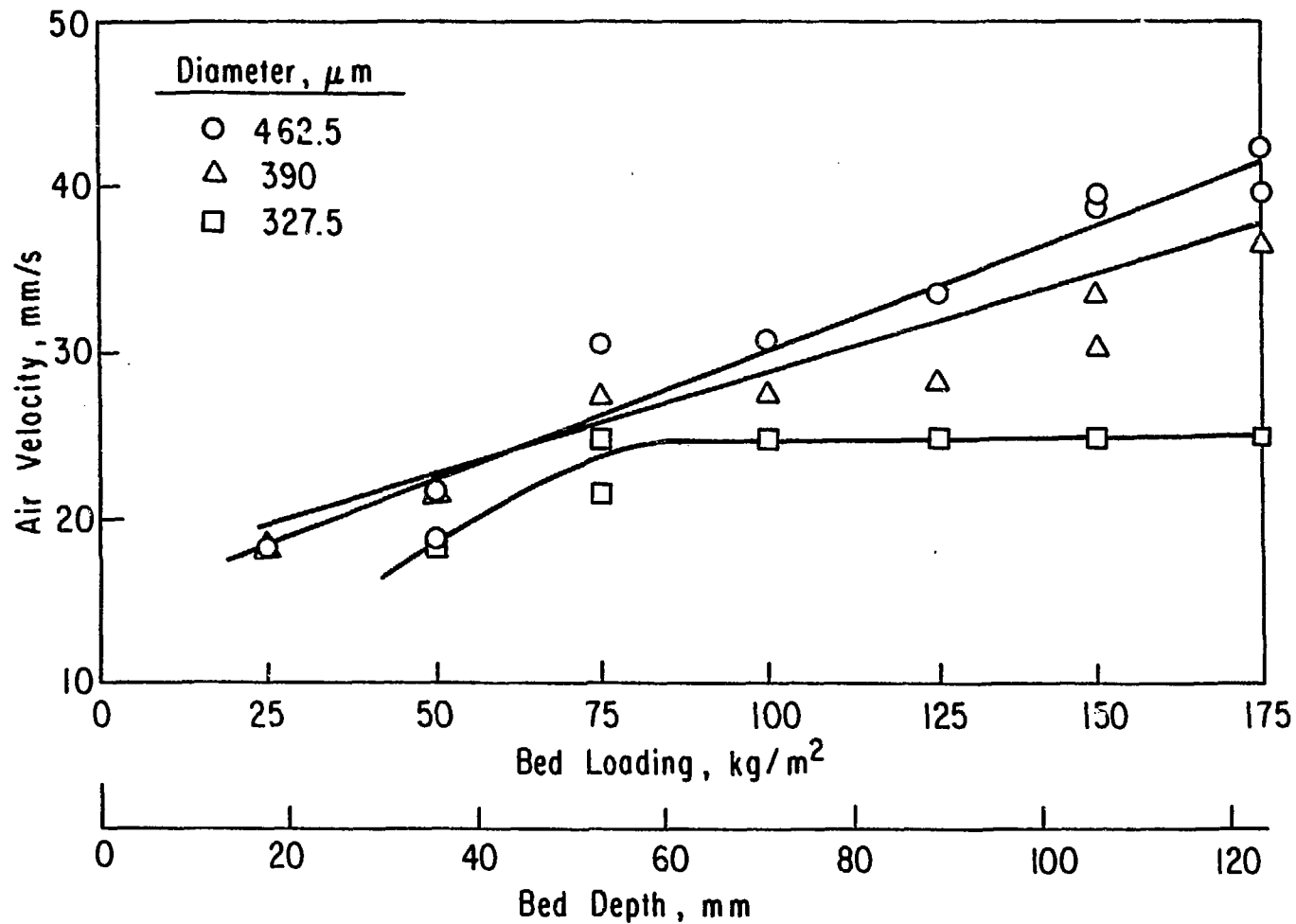


Figure 3. Air Velocities Required to Fluidize Freon-113-Glass Beds.

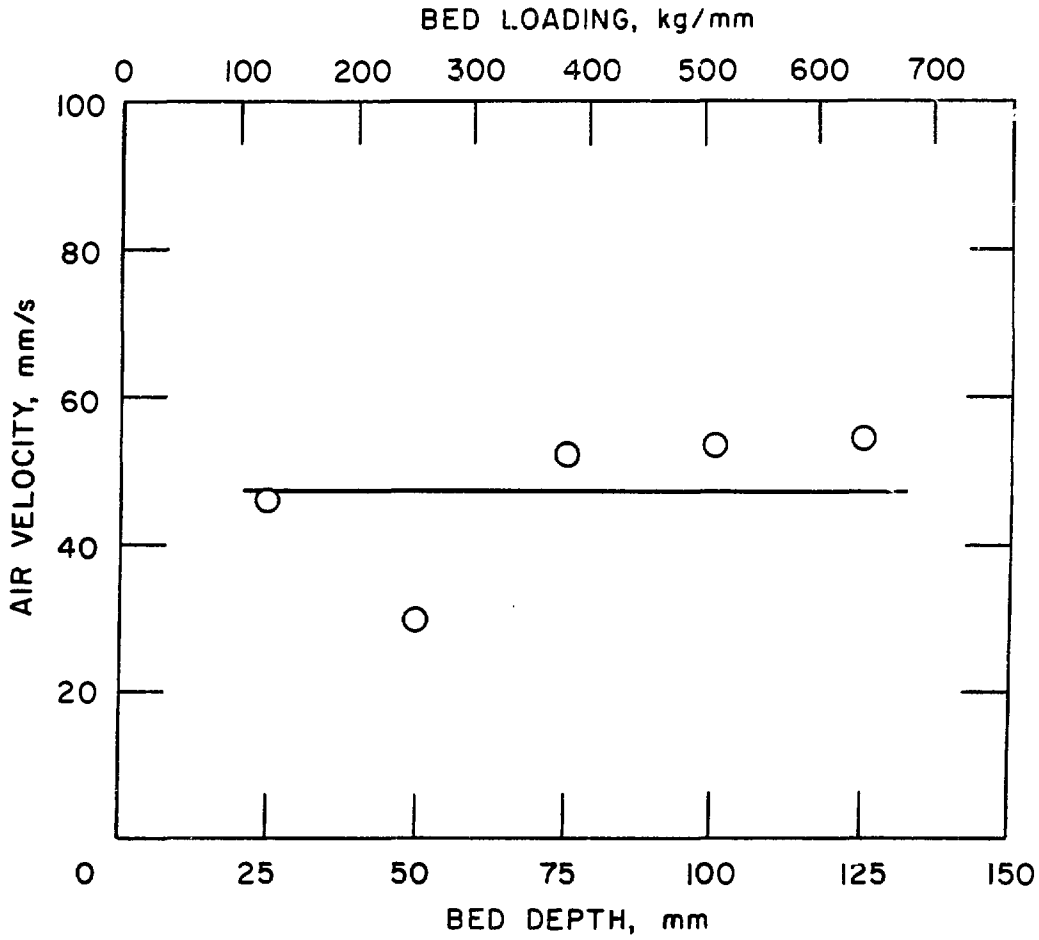


Figure 4. Air Velocities Required to Fluidize Beds of 11  $\mu\text{m}$   $\text{UO}_2$  in Water.

The same behavior was observed with the glass-Freon-113 beds. However, the critical particle diameter of 390  $\mu\text{m}$  was somewhat larger than the 327.5  $\mu\text{m}$  value for the water system. Buoyancy would be greater in the Freon-113, since Freon-113 is denser (1635  $\text{Kg}/\text{m}^3$ ) than water (998  $\text{Kg}/\text{m}^3$ ).

The settling velocity was determined by gradually decreasing the air flow after the particulate phase was completely suspended in the liquid pool. There was a hysteresis effect in that the critical velocity for the transition from a suspended bed to a settled bed was lower than that required to initially fluidize a settled bed.

The fluidization mechanism is too complex for theoretical analysis of incipient fluidization at present. The data were analyzed empirically by determining the effect of each parameter on the minimum fluidization velocity. These parameters were combined in the form of dimensionless groups. For the smaller particles in which the minimum fluidization velocity is independent of bed height the data were correlated by (see Fig. 5):

$$\left(\frac{U_t}{U_{mf}}\right) \left(\frac{U_{mf} D_p \rho_l}{\mu_l}\right)^2 = 8.2 \times 10^{-4} \left[\frac{D_p^3 \rho_l g(\rho_s - \rho_l)}{\mu_l^2}\right]^{0.75} \left[\frac{\sigma}{g D_p^2 (\rho_s - \rho_l)}\right]^{0.375} \quad D_p < 328 \mu\text{m}$$

or in terms of dimensionless numbers:

$$\left(\frac{U_t}{U_{mf}}\right) \text{Re}_{mf}^2 = 8.2 \times 10^{-4} N_f^{1.5} N_{Eo}^{-0.375} \quad D_p < 328 \mu\text{m} \quad (7)$$

The coefficient  $8.2 \times 10^{-4}$  was determined by regression analysis for a curve with a slope of 1. The best least squares fit resulted in a curve with a slope of 0.97 and a coefficient of  $9.5 \times 10^{-4}$ . However, the data scatter does not justify this precision.

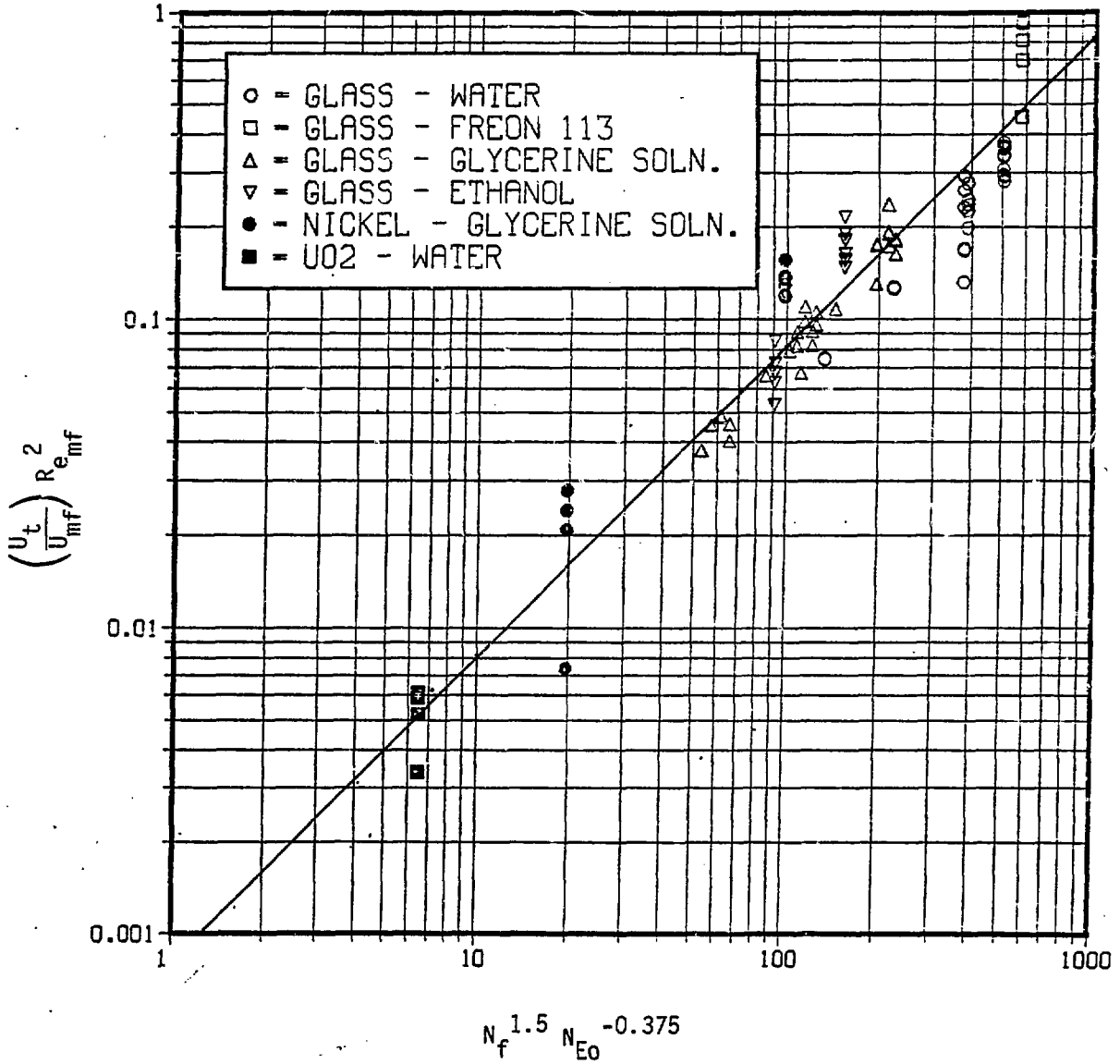


Figure 5. Correlation of Minimum Fluidization of Particles Smaller than 328 μm.

The critical velocity for settling of the particles was correlated by (see Fig. 6):

$$\left(\frac{U_t}{U_{st}}\right) Re_{st}^2 = 8.67 \times 10^{-5} N_f^2 N_{Eo}^{-0.50} \quad D_p < 328 \mu m \quad (8)$$

The critical velocity for settling was not as precisely determined as that for initial fluidization since some judgment was required as to when the fluidized bed began to collapse.

The terminal velocity was calculated from the following:

$$U_t = 0.153 \frac{g^{0.71} D_p^{1.14} (\rho_s - \rho_l)^{0.71}}{\rho_l^{0.29} \mu_l^{0.43}} \quad (9)$$

The correlations (7) and (8) are somewhat similar to

$$C_D Re^2 = \frac{4}{3} N_f^2 \quad (10)$$

used for determining terminal velocities. Rowe [12] used this functional relationship for correlating the minimum fluidization velocity for the single phase flow of either liquid or gas by correcting the coefficient of friction,  $C_D$ , by a factor of 68.5 to account for the presence of the neighboring particles. This is consistent with the observed mechanism of small particle incipient fluidization in which individual particles are stripped from the channel walls until the entire bed is suspended.

The 311  $\mu m$  nickel particles were somewhat smaller than the smallest size copper particles (387  $\mu m$ ) used in the tests by Cho, et al. [12]. Since both copper and nickel were the same density, it was of interest to apply the flooding criteria given in Wallis [10] to the 311  $\mu m$  nickel particle bed. The flooding velocity was calculated to be about 0.10 m/s, which is less than

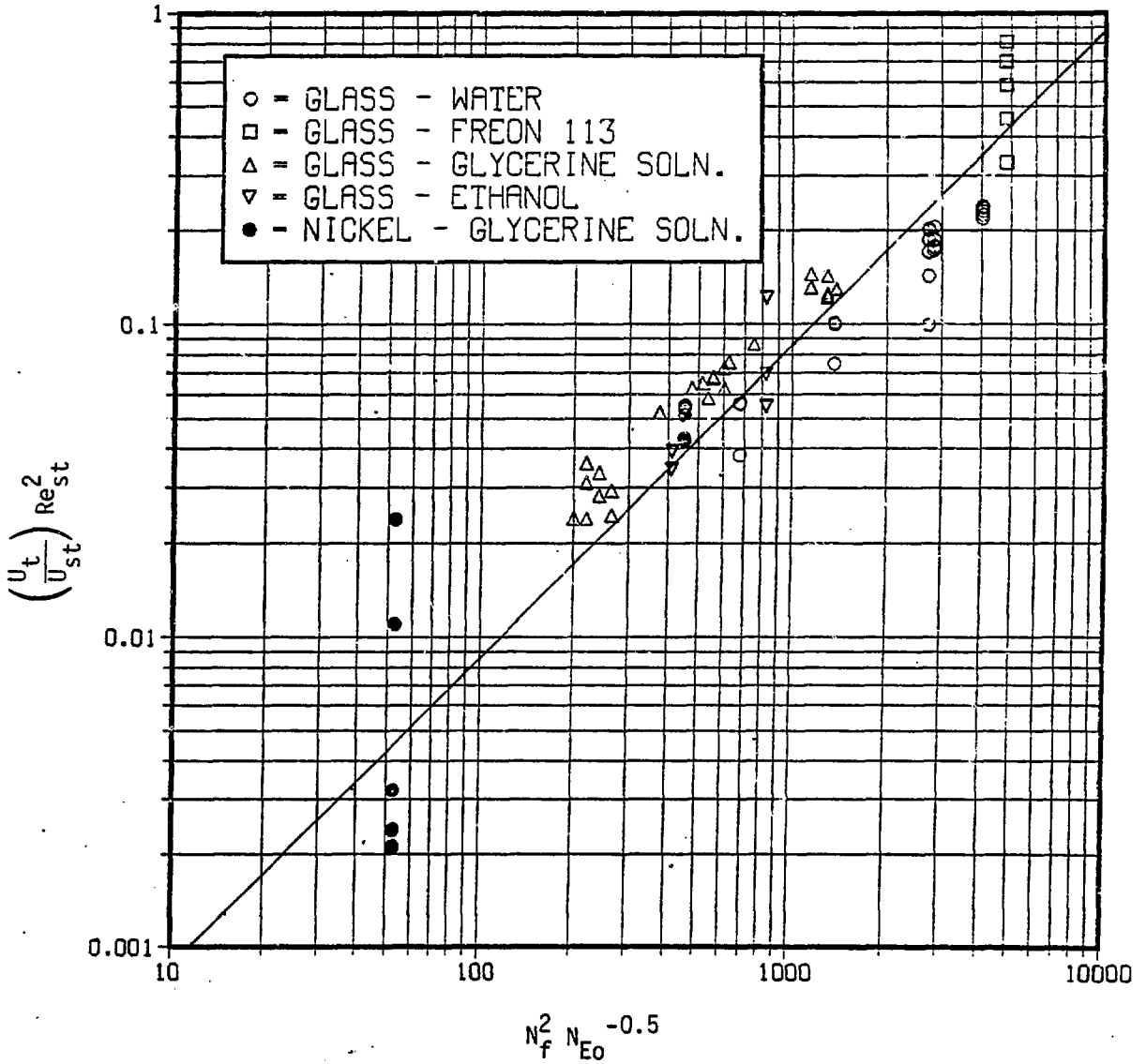


Figure 6. Correlation of Critical Settling Velocities of Particles  
Less Than 328 μm.

observed fluidization velocities. It should be noted that 311  $\mu\text{m}$  is well below the smallest packing size (6 mm) used for the flooding correlation. The minimum fluidization velocity for the 311  $\mu\text{m}$  nickel particles based on a flooded bed with only air flow through the particle bed, using the Wen and Yu [11] correlation is 0.26 m/s. This was below the observed 0.4 to 0.6 m/s range for the three phase system.

For the larger beds in which the minimum fluidization velocity is dependent on bed height the data were correlated by (see Fig. 7):

$$\left[ \frac{\rho_L U_{mf}^2}{D_p g(\rho_s - \rho_L)} \right]^{1.75} \left[ \frac{D_t^2 g(\rho_s - \rho_L)^{0.5}}{U_{mf} \mu} \right] \left[ \frac{g D_p}{U_{mf}^2} \right]^{1.0} = 77.0 H^{0.5} \quad D_p > 328 \mu\text{m}$$

or in terms of dimensionless groups:

$$j_{mf}^{*3.5} U_{mf}^{*-0.5} Fr_{mf}^{-1.0} = 77.0 H^{0.5} \quad D_p > 328 \mu\text{m} \quad (11)$$

The above expression can be simplified to:

$$U_{mf} = 77.0 \frac{H^{0.5} g^{0.25} D_p^{0.75} (\rho_s - \rho_L)^{1.25}}{\rho_L^{1.75} D_t \mu^{0.5}} \quad D_p > 328 \mu\text{m} \quad (12)$$

A least squares analysis indicated that the data would be better correlated with H to the 0.55 power and a constant of 67.9. However, it was felt that the data scatter did not justify this precision. Therefore, a least squares fit was forced to H to the 0.5 power with a resulting constant of 77.0.

The critical velocity for settling of fluidized particles was correlated by (see Fig. 8):



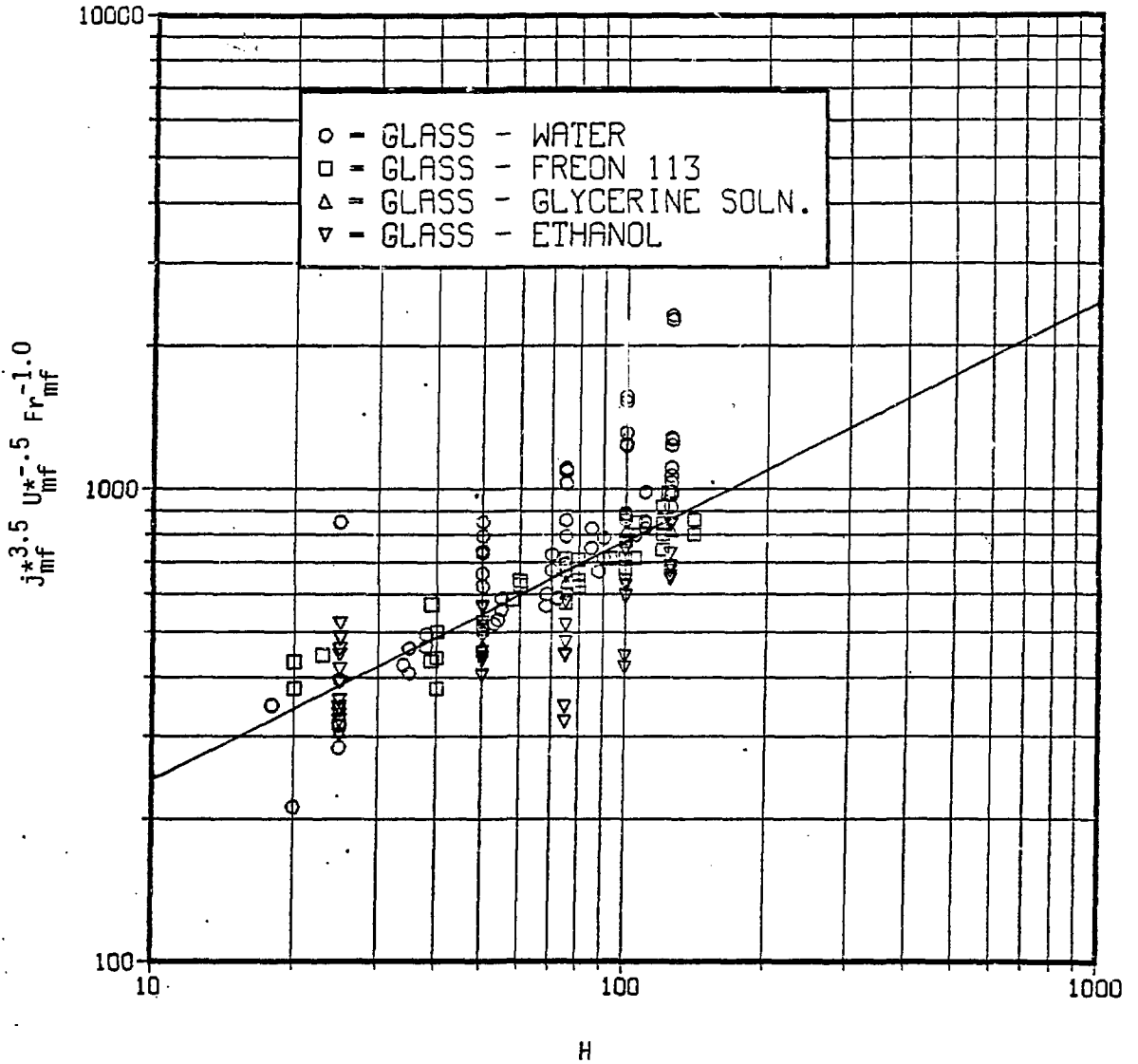


Figure 7. Correlation of Minimum Fluidization of Particles Greater than 328  $\mu\text{m}$ .

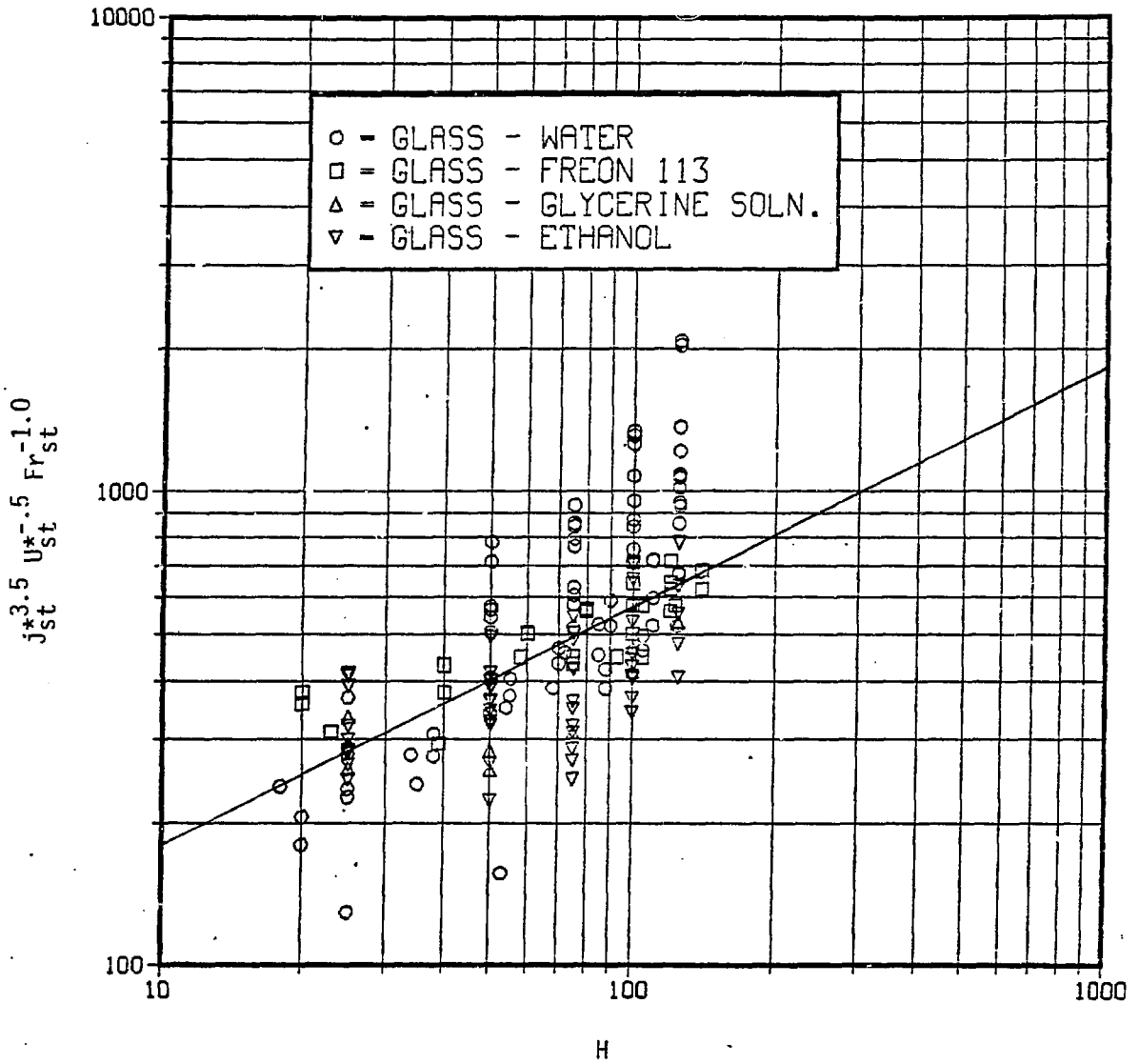


Figure 8. Correlation of Critical Settling Velocities of Particles Greater than 328 μm.

$$j_{st}^{3.5} U_{st}^{-.5} Fr_{st}^{-1.0} = 57.7 H^{.5} \quad D_p > 328 \mu m \quad (13)$$

which can be rearranged to:

$$U_{st} = 57.7 \frac{H^{.5} g^{.25} D_p^{.75} (\rho_s - \rho_l)^{1.25}}{\rho_l^{1.75} D_t^{.5}} \quad D_p > 328 \mu m \quad (14)$$

Without force fitting, the least squares analysis yielded H to the 0.57 power and a constant 43.1.

The data scatter was greater for the critical settling velocity than for the minimum fluidization velocity, and the data scatter was greater for the larger particles than for the smaller particles. The onset of fluidization was more catastrophic with the larger particles. The entire bed slugged before breaking up into the fluidized state. The breakup of the slugging bed was not as sharply defined as the smoother process of fluidizing the smaller particles individually. The stability of the particulate slugs was sensitive to the column diameter. Within the range of data scatter, the minimum fluidization velocity was observed to be inversely proportional to the column diameter (to the first power). Narayanan et al. [6] assumed a power of 0.5 for particles greater than 200  $\mu m$  for their correction factor based on work at slurry transport through pipes. Additional data are needed to confirm the column diameter dependency for three-phase fluidization since the current data are limited to one series of measurements with 548  $\mu m$  glass particles for two different column diameters.

## Conclusions

1. Bed pressure drop increased with increased gas flow and then leveled off after incipient fluidization in a manner similar as for fluidization with a single phase fluid.
2. The critical gas velocity required for fluidization was independent of particle bed depth for particles less than 328  $\mu\text{m}$ . The particles were individually stripped off the channel walls until the entire bed was fluidized.
3. The minimum fluidization gas velocity required for beds of particles larger than 328  $\mu\text{m}$  was dependent on particle bed depth. These beds slugged before breaking up into a fluidized state.
4. Further work is needed on the effect of tube diameter.
5. Further analysis is needed to determine the criterion for beds fluidized by stripping of individual particles off the channel walls and for beds fluidized by breakup of slugging.

NOMENCLATURE

$C_D$	drag coefficient
$C_\mu$	viscosity correlation factor
$D_p$	particle diameter
$D_t$	tube diameter
$Fr$	Froude number, $U^2/g D_p$
$g$	acceleration due to gravity
$H$	static particle bed depth
$H_g$	gas hold up
$H_l$	static liquid depth
$H_{max}$	maximum quantity of solids that can be held in suspension
$H_s$	solids concentration, Kg of solid/Kg of liquid
$H_{sl}$	static slurry height
$j^*$	$U \rho_l^{1/2} [g D_p (\rho_s - \rho_l)]^{-1/2}$
$L$	bed loading, kg/m <sup>2</sup>
$N_B$	modified bubble flow number, $\sigma/U_B \mu_l$
$N_{E\ddot{o}}$	Eötvös number, $g D_p^2 (\rho_s - \rho_l)/\sigma$
$N_f$	inverse viscosity, $[D_p^3 \rho_l g (\rho_s - \rho_l)]^{1/2}/\mu_l$
$Re$	Reynolds number, $U D_p \rho_l/\mu_l$
$Re_T$	$U D_T \rho_l/\mu_l$
$U$	superficial gas velocity
$U^*$	$U \mu/D_t^2 g (\rho_s - \rho_l)$
$U_B$	bubble velocity
$U_m$	terminal velocity of largest particle
$U_t$	particle terminal velocity in liquid phase
$\varepsilon$	particle bed voidage

$\gamma$  wettability factor  
 $\gamma'$   $\gamma$  less wettable solid/ $\gamma$  most wettable solid  
 $\mu_l$  liquid density  
 $\rho_g$  gas density  
 $\rho_l$  liquid density  
 $\rho_s$  particle density  
 $\sigma$  surface tension  
 $\psi$  shape factor

Subscripts

$m_f$  minimum fluidization  
 $st$  settling

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TABLE A1. (cont'd)

Test No.	D <sub>p</sub> ,mm	P <sub>s</sub> , Kg/ms	γ	Dt,mm	H <sub>1</sub> ,mm	L <sub>1</sub> <sup>2</sup> , Kg/m <sup>2</sup>	c	H <sub>1</sub> ,mm	ρ <sub>1</sub> , <sup>3</sup> Kg/m <sup>3</sup>	ν <sub>1</sub> ,Cp	σ <sub>1</sub> , dynes/cm	U <sub>inf</sub> <sup>o</sup> , Cm/s	U <sub>st</sub> <sup>o</sup> , Cm/s
388	547.5	2600	1.0	50.3	25.0	42.80	.4	75.0	785.1	1.100	21.50	22.12	16.50
389	547.5	2600	1.0	50.3	50.0	82.33	.4	100.0	785.1	1.100	21.50	22.25	16.68
390	547.5	2600	1.0	50.3	50.0	82.33	.4	100.0	785.1	1.100	21.50	26.75	16.90
391	547.5	2600	1.0	50.3	75.0	117.41	.4	125.0	785.1	1.100	21.50	27.39	20.48
392	547.5	2600	1.0	50.3	75.0	117.41	.4	125.0	785.1	1.100	21.50	25.23	22.89
393	547.5	2600	1.0	50.3	100.0	152.94	.4	150.0	785.1	1.100	21.50	31.46	27.52
394	547.5	2600	1.0	50.3	100.0	152.94	.4	150.0	785.1	1.100	21.50	31.16	29.59
395	547.5	2600	1.0	50.3	125.0	185.25	.4	175.0	785.1	1.100	21.50	35.35	32.88
396	547.5	2600	1.0	50.3	125.0	185.25	.4	175.0	785.1	1.100	21.50	35.56	32.56