Chapter 5. Fluidization

5.1 Fundamental

* Δp $vs.$ U Figure 5.1

Minimum (incipient) fluidization, U_{mf}

From force balance

Net downward force

$$\Delta p = (1-\varepsilon)(\rho_p - \rho_g)H \qquad (1)$$

Net upward force

$$\frac{\Delta p}{H} = 150\frac{(1-\varepsilon)^2}{\varepsilon^3}\frac{\mu U}{x_{sv}^2} + 1.75\frac{1-\varepsilon}{\varepsilon^3}\frac{\rho_g U^2}{x_{sv}} \qquad (2)$$

Equating (1) and (2) at $U = U_{mf}$

$$Ar = 150\frac{(1-\varepsilon)}{\varepsilon^3}Re_{mf} + 1.75\frac{1}{\varepsilon^3}Re_{mf}^2$$

where $Ar \equiv \frac{\rho_g x_{sv}^3(\rho_p - \rho_f)g}{\mu^2}$, Archimedes number

$$Re_{mf} = \frac{\rho_f U_{mf} x_{sv}}{\mu}$$

$\varepsilon = 0.4$, usually

More practically,

Wen and Yu(1966) for $x_{sv} > 100\mu m$

$$Ar = 1056 Re_{mf} + 159 Re_{mf}^2$$

Baeyens and Geldart(1974) for $x < 100\mu m$

$$U_{mf} = \frac{(\rho_p - \rho_f)^{0.934} g^{0.934} x^{1.8}}{1110\mu^{0.87}\rho_f^{0.066}}$$

5.2 Relevant Powder and Particle Properties

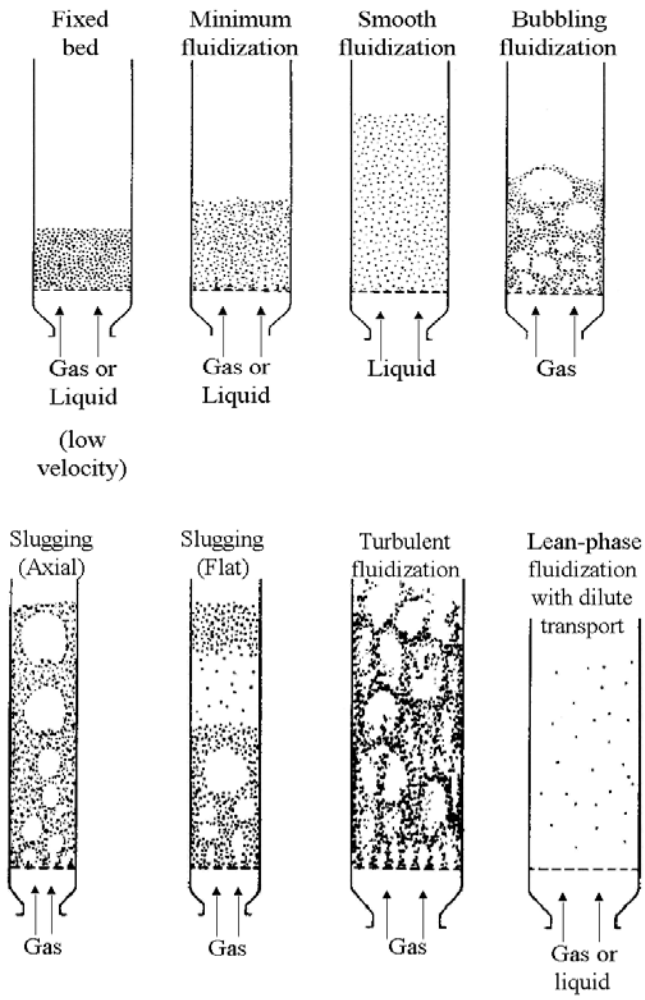
- Absolute density: materials property
- Particle density: Figure 5.2
- Bed density

※ Sieve diameter, x_p , $x_v = 1.13x_p$

$$mean\ x_p = \frac{1}{\sum m_i / x_i}$$

5.3 Bubbling and Non-Bubbling Fluidization

Types of Fluidization



Various types of fluidized beds

- Bubbling fluidized bed : Figure 5.3 for Group B particles

- Liquid fluidization: Figure 5.4

5.4 Classification of Powders

Geldart(1974) Figure 5.6

Table 5.1

Group A : Nonbubbling for $U_{mf} < U < U_{mb}$

where

$$U_{mb} = 2.07\exp(0.716F)\left[\frac{x\rho_g^{0.06}}{\mu^{0.347}}\right]$$

where F : fraction of powder less than 45 μm

Bubbling for $U > U_{mb}$

Maximum bubble size

$$d_{Bv,\max} = \frac{2}{g}(U_{T2.7})^2$$

Group B : Bubbling for $U > U_{mf}$

No maximum in bubble size

Slug**ging** ($d_B \geq \frac{1}{3}D$) - Figure 5.8 (right)

Tagi and Muchi (1952)

In order to avoid slug formation

$$\left(\frac{H_{mf}}{D}\right) \leq \frac{1.9}{(\rho_p x_p)^{0.3}}$$

Slug forms when $\left(\frac{H_{mf}}{D}\right) > \frac{1.9}{(\rho_p x_p)^{0.3}}$ and

$$U > U_{mf} + 0.16(1.34D^{0.175} - H_{mf})^2 + 0.07(gD)^{1/2}$$

Group D : Spoutable

Group C : Subject to *channeling* in large diameter-bed

5.5 Expansion of a Fluidized Bed

(1) Nonbubbling Fluidized Bed

Upward superficial fluid velocity(or volumetric flux)

$$U = U_T \varepsilon^2 f(\varepsilon) = U_T \varepsilon^n$$

For $Re_p \leq 0.3$, $n = 4.65$

For $Re_p \geq 500$, $n = 2.4$

Assuming conservation of bed mass, M_B

$$M_B = (1-\varepsilon)\rho_p AH$$

$$\therefore (1-\varepsilon_1)\rho_p AH_1 = (1-\varepsilon_2)\rho_p AH_2$$

$$\therefore \frac{H_2}{H_1} = \frac{1-\varepsilon_1}{1-\varepsilon_2}$$

WorKed Example 5.1

(2) Bubbling Fluidized Bed

The bed is assumed to consist of two phases:

- *Dense* (particulate, emulsion) phase : a state of minimum fluidization
- *Lean* (bubble) phase : flow of gas in excess of minimum fluidization as bubbles

Gas flow as bubbles $= Q - Q_{mf} = (U - U_{mf})A$

Gas flow in the emulsion phase $= Q_{mf} = U_{mf}A$

$$\varepsilon_B = \frac{H - H_{mf}}{H} = \frac{Q_B}{AU_B} = \frac{U - U_{mf}}{U_B}$$

* Mean bed voidage, $1-\varepsilon = (1-\varepsilon_B)(1-\varepsilon_{mf})$

* Bubble Size and Rise Velocity

$$U_B = \Phi_B (g d_{Bv})^{1/2}$$

where $\Phi_B = function\ of\ D$

For group B

$$d_{Bv} = \frac{0.54}{g^{0.2}}(U - U_{mf})^{0.4}(L + 4N^{-1/2})^{0.8}$$

where N : the number of holes/m^2

L : distance above the distributor

5.6 Entrainment

- Carryover, elutriation

Zone above fluidized bed (Freeboard) - Figure 5.11

- Splash zone
- Transport disengagement zone

  Transport disengagement height,

  TDH from Figure 5.12 or

  $TDH = 4.47 d_{Bvs}^{0.5}$

- Dilute-phase transport zone

Worked Example 5.1

Instantaneous entrainment rate of size x_i

$$R_i = -\frac{d}{dt}(M_B m_{Bi}) = K_{ih}^{*} A m_{Bi}$$

where M_B: total mass of solids in the bed

A: area of bed surface

m_{Bi}: fraction of bed mass with size x_i at time t

K_{ih}^{*}: elutriation constant

Total rate of entrainment

$$R_T = -\frac{d}{dt}[\sum(M_B m_{Bi})] = \sum K_{ih}^{*} A m_{Bi}$$

$K^{*}_{i\infty}$: K_{ih}^{*} above TDH - (5.45) and (5.46)

Worked Example 5.2

5.7 Heat transfer in Fluidized Bed

Heat transfer between particles and gas

$$Nu = 0.03\,Re_p^{1.3} \quad (Re_p < 50)$$

where $Nu = \frac{h_{gp}x}{k_g}$

$L_{0.5}$: L for $\frac{T_g - T_s}{T_{g0} - T_s} = 0.5$

Very short(0.95mm～5mm)

Heat transfer between surface and bed

$$h = h_{pc} + h_{gc} + h_r$$

particle convection, gas convection, radiation

Figure 5.14: h_{pc} is controlled by k_g

Figure 5.15: maximum occurs by blanket of bubbles

When $U > (2 \sim 3) \times U_{mf}$

For group B powders

$$h_{max} = 35.8 k_g^{0.6} \frac{\rho_p^{0.2}}{x^{0.36}}, \ \mathrm{W/m^2K}$$

For group A powders

$$Nu_{max} = 0.157\,Ar^{0.475}$$

5.8 Applications of Fluidized Beds

Advantages

- *Liquid-like behavior*, easy to control and automate
- *Rapid mixing*, uniform temperature and concentration
- *Resists rapid temperature changes*, hence responds slowly to changes in operating conditions and avoids temperature runaway

with exothermic reactions

- *Circulate solids* between fluidized beds for heat exchange
- *Applicable for large or small scale operations*
- *Heat and mass transfer rates are high*, requiring smaller surfaces

Disadvantages

- *Bubbling beds* are *difficult* to predict and are *less efficient*
- Rapid mixing of solids causes *nonuniform residence times* for continuous flow reactors
- *Particle comminution(breakup)* is common
- *Pipe and vessel walls erode* to collisions by particles

(1) Physical Processes

Drying / Mixing / Granulation / Coating / Heat exchanger/ Adsorption(Desorption)

Figure 5.17

(2) Chemical Processes

Table 5.2

Figure 5.18 Fluidized catalytic cracker

5.9 A Simple Model for the Bubbling Fluidized Bed Reactor

Figure 5.19

Orcutt et al(1962)

Form the overall mass balance on the reactant + shell mass balance

$$1-\frac{C_H}{C_0}=(1-\beta e^{-\chi})-\frac{(1-\beta e^{-\chi})^2}{\frac{kH_{mf}(1-\varepsilon_p)}{U}+(1-\beta e^{-\chi})}$$

$$C_{BH} = C_P + (C_0 - C_P)\exp[-\chi]$$

$$C_P = \frac{C_0[U-(U-U_{mf})e^{-\chi}]}{kH_{mf}(1-\varepsilon_P)+[U(U-U_{mf})e^{-\chi}]}$$

$$C_H = \frac{U_{mf}C_P + (U-U_{mf})C_{BH}}{U}$$

where C_0 : concentration of reactant at distributor

C_H : concentration of reactant leaving the reactor(H)

C_B : concentration of the reactant in the bubble phase

C_P : concentration of the reactant in the particulate phase

$\beta = \frac{U-U_{mf}}{U}$ and $\chi = \frac{K_cH}{U_B}$

where K_c: mass transfer coefficient of the reactant from particulate phase

Figure 5.20

Worked Example 5.3