Chapter 5. Fluidization

5.1 Fundamentals

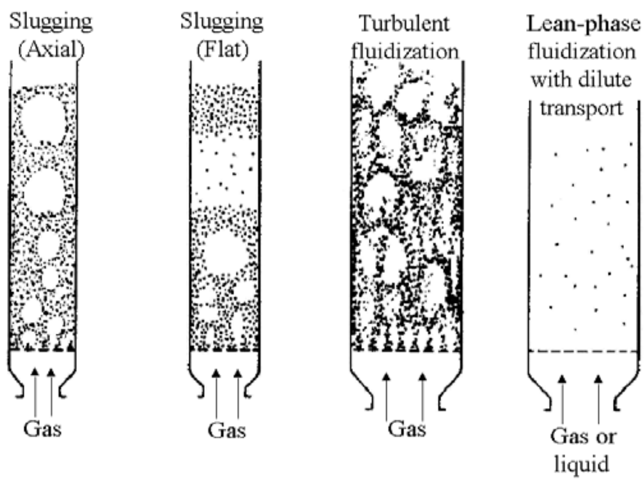
* Δp *vs.* U

유체가 분체층을 통과할 때 유속이 증가할수록 분체나 벽과의 마찰의 증가로 압력강하가 증가한다. 그러나 유속이 증가하여 항력이 중력과 비등해지면 분체층은 들어올려지는 힘을 받아 부풀려진다. (공극률ε의 증가) 이때부터 분체는 층에서 비교적 자유로이 움직이게 되고 유체의 압력강하는 상수값에 머문다. 유속이 더욱 증가하여 분체의 최종침강속도 근처가 되면 분체는 유체를 따라 이송되는 시점에 이른다.

Minimum (incipient) fluidization

U_{mf}: minimum fluidization velocity

Various Types of Fluidization



Various types of fluidized beds

From force balance

Net downward force

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

Net upward force

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

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

$$Re_{mf} = \frac{\rho_f U_{mf} d_{p,sv}}{\mu}$$

$\varepsilon = 0.4$, usually

More practically,

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

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

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

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

5.2 Relevant Powder and Particle Properties

Particle density Figure 5.1

particle mass per hydrodynamic particle volume

Absolute density

particle mass per true particle volume

Bed density or bulk density

bed mass per bed volume including voids

Mean diameter of the powder from sieve analysis

$$mean\ d_p = \frac{1}{\sum \frac{m_i}{d_{pi}}}$$

5.3 Bubbling and Non-Bubbling Fluidization

At superficial velocities above the minimum fluidization velocity, bubbling or nonbubbling fluidization appears depending on properties of gas and particles

In most case of liquid fluidization, only nonubbling fluidization occurs

Figure 5.3 and Figure 5.4

5.4 Classification of Powder Groups

Geldart(1974)   Figure 5.6

Table 5.1

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

where

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

where F : fraction of powder less than 45 μm

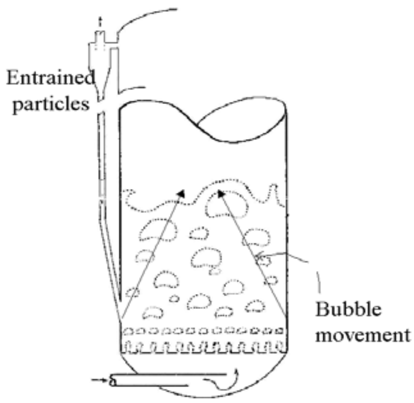
Bubbling for $U > U_{mb}$

Maximum bubble size

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

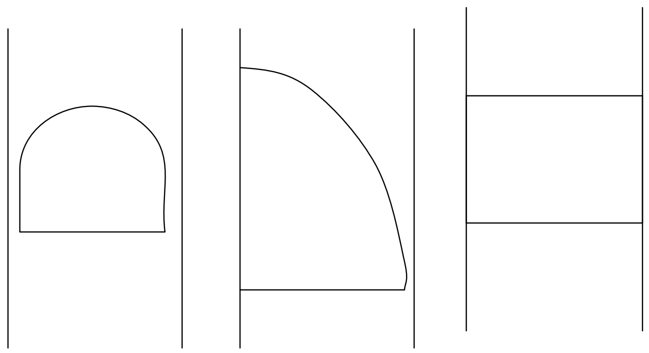
Group B : **Bubbling for $U > U_{mf}$**

No maximum in bubble size



Bubble growth and movement

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



Types of slugs

Tagi and Muchi (1952)

In order to avoid slug formation

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

or

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

Group D : Spoutable Figure 5.5

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

5.5 Expansion of a Fluidized Bed

1) Nonbubbling Fluidized Bed

Upward superficial fluid velocity

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

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

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

Assuming conservation of bed mass, M_B

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

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

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

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

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

5.7 Heat transfer

Heat transfer between particles and gas

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

Heat transfer between surface and bed

$$\underbrace{h}_{} = \underbrace{h_{pc}}_{\text{particle convection}} + \underbrace{h_{gc}}_{\text{gas convection}} + \underbrace{h_r}_{\text{radiation}}$$

Figure 5.14

Figure 5.15

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

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

Applications

- Physical Processes :

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

  Figure 5.17
- Chemical Processes

  Table 5.2

  Figure 5.18 Fluidized catalytic cracker