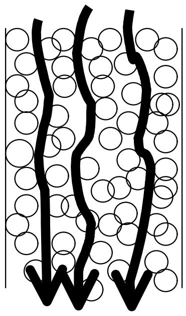
Chapter 4 Movement of Particles by Fluid Flow

4.1 Fluid Flow through Packed Bed of Particles (Chapter 4)

(1) Pressure Drop - Flow Relationship

1) Laminar Flow

Fluid flow through a packed bed: simulated by fluid flow through a hypothetical tubes



$$\therefore \frac{(-\Delta p)}{H} = \frac{32\mu U}{D^2}$$

$$\Rightarrow \frac{(-\Delta p)}{H_e} = \frac{K_1 \mu U_i}{D_e^2}$$

Hagen-Poiseille equation

Substituting suitable relations for H_e *(equivalent height) and* D_e *(equivalent diameter)*

$$\therefore \frac{(-\Delta p)}{H} = 180 \frac{\mu U}{x^2} \frac{(1-\varepsilon)^2}{\varepsilon^3}$$

Carman-Kozeny equation

2) General Equation for Turbulent and Laminar Flow

Ergun equation

$$\frac{(-\Delta p)}{H} = 150 \frac{\mu U}{x^2} \frac{(1-\varepsilon)^2}{\varepsilon^3} + 1.75 \frac{\rho_f U^2}{x} \frac{(1-\varepsilon)}{\varepsilon^3}$$

Laminar *Turbulent*

Laminar flow for $Re* = \frac{xU\rho_f}{\mu(1-\varepsilon)} < 10$

Turbulent flow for $Re* = \frac{xU\rho_f}{\mu(1-\varepsilon)} > 2000$

or

$$f* = \frac{150}{Re*} + 1.75$$

where $f* \equiv \frac{(-\Delta p)}{H} \frac{x}{\rho_f U^2} \frac{\varepsilon^3}{(1-\varepsilon)}$

Friction factor

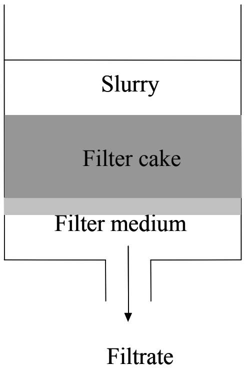
Figure 4.1

3) Nonspherical Particles

x_{sv} *(surface-volume diameter) instead of* x

Worked Example 4.1

(2) (Liquid) Filtration



(1) Introduction

Filter media : Canvas cloth, woolen cloth, metal cloth, glass, cloth, paper, synthetic fabrics

Filter aids : To avoid cake plugging

e.g. Diatomaceous silica, perlite, purified woolen cellulose, other inert porous solids

- By either adding slurry (increasing cake permeability)

or precoating the filter media surface

1) Incompressible Cake

For cake filter

From laminar part of Ergun equation

Δ μ
ε

$$\frac{(-\quad p)}{H} = \frac{150\quad U(1-\quad)^2}{x^2\quad ^3}$$

where L : cake thickness

x : surface-volume diameter of particle

* For compressible filter cake,

$$\frac{dp}{dL} = r_c \mu U$$

where r_c: a function of pressure difference

By defining cake resistance r_c

$$r_c = \frac{150(1-\varepsilon)^2}{x^2 \varepsilon^3},$$

$$\frac{(-\Delta p)}{H} = r_c \mu U$$

$$\text{where} \quad U = \frac{1}{A}\frac{dV}{dt}$$

V: volume of slurry fed to filter

Also defining ϕ (volume formed by passage of unit volume filtrate)

$$\phi = \frac{HA}{V},$$

$$\frac{dV}{dt} = \frac{A^2(-\Delta p)}{r_c \mu \phi V}$$

Including the resistance of filter medium,

since the resistances of the cake and the filter medium are in series,

$$(-\Delta p) = (-\Delta p_m) + (-\Delta p_c)$$

$$\downarrow$$

$$\frac{1}{A}\frac{dV}{dt} r_c \mu H_c$$

By analogy for the filter medium

$$(-\Delta p_m) = \frac{1}{A}\frac{dV}{dt} r_m \mu H_m$$

$$\therefore (-\Delta p) = \frac{1}{A}\frac{dV}{dt}(r_m \mu H_m + r_c \mu H_c)$$

Defining equivalent height of filter cake and volume of filtrate

$$r_m H_m = r_c H_{eq} \text{ and } H_{eq} = \frac{V_{eq}}{A}$$

$$V_{eq} = \frac{A}{\phi}\frac{r_m H_m}{r_c}$$

where V_{eq}*: volume of filtrate passing to create a cake of thickness* H_{eq}

$$\therefore \frac{1}{A}\frac{dV}{dt} = \frac{(-\Delta p)A}{r_c \mu \phi (V + V_{eq})}$$

Constant rate filtration

$$\frac{1}{A}\frac{dV}{dt} = \frac{(-\Delta p)A}{r_c \mu \phi (V + V_{eq})} = constant$$

Constant pressure filtration

Integrating

$$\frac{t}{V} = \frac{r_c \mu \phi}{A^2(-\Delta p)}\left(\frac{V}{2} + V_{eq}\right)$$

Worked Example 4.2

3) Washing the Cake

Figure 4.2

4.2 Fluidization (Chapter 5)

(1) Fundamental

* Δp vs. U *Figure 5.1*

Minimum (incipient) fluidization, U_{mf}

From force balance

Net downward force

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

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 $U = U_{mf}$

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

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

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

$$\varepsilon = 0.4, \text{ usually}$$

More practically,

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

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

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

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

* *Densities of particles*

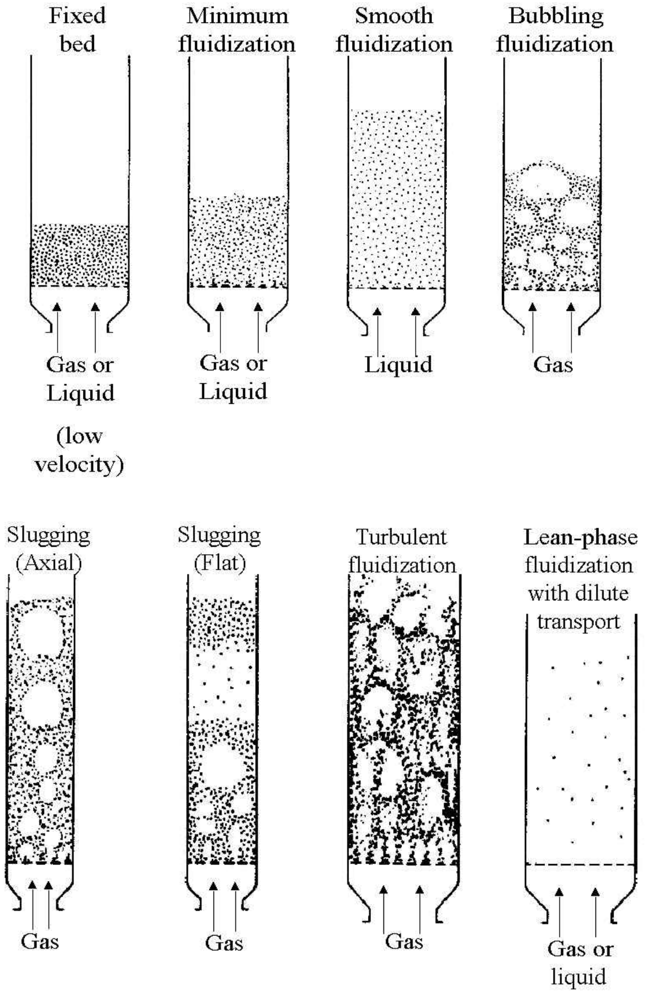
- *Absolute density: materials property*
- *Particle density: Figure 5.2*
- *Bed density*

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

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

(2) Bubbling and Non-Bubbling Fluidization (5.3)

Types of Fluidization



Various types of fluidized beds

- *Bubbling fluidized bed : Figure 5.3 for Group B particles*

- *Liquid fluidization: Figure 5.4*

Worked Example 5.1

(3) Classification of Powders (5.4)

Geldart(1974) Figure 5.6

Table 5.1

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

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

No maximum in bubble size

Group D : Spoutable

Group C : Subject to channeling in large diameter-bed

(4) Applications of Fluidized Beds(5.8)

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

Figure 5.17

2) Chemical Processes

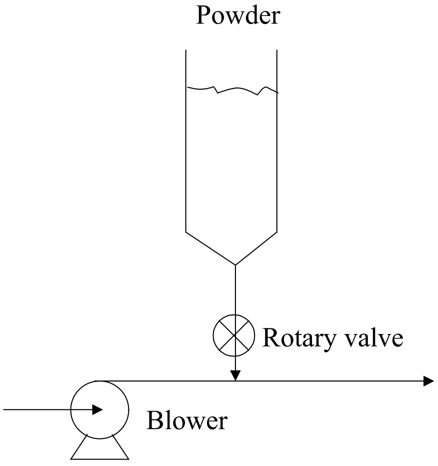
Table 5.2

Figure 5.18 Fluidized catalytic cracker

4.3 Pneumatic Transport (Chapter 6)

(1) Pneumatic Transport

- Use of a gas to transport a particulate solid through pipeline



- Three major variables for pneumatic conveying

  - solid mass flow rate
  - gas mass flow rate
  - pressure gradient(pressure drop per unit length)

1) Dilute-Phase and Dense-Phase Transport

Dilute-Phase	*Dense-Phase*
High gas velocity (> 20 m/s)	*Low-gas velocity (1-5 m/s)*
Low solids concentration (< 1 % by volume)	*High solids concentration (> 30 % by volume)*
Low pressure drop (<5 mbar/m)	*High pressure drop (> 20 mbar/m)*
Short-route, continuous transport(< 10 ton/h)	*Batch or semibatch transport*
Capable under negative pressure	
Particles behave as individuals	
Fully suspended in gas	*Not-fully suspended in gas*
Fluid-particle : dominant	*Much interaction between particles and between particle and wall*

2) The Choking Velocity in Vertical Transport

Figure 6.1 - Δp/ΔL vs. U (gas superficial velocity)

at various solids flow flux G

Static head of solids → friction resistance

Choking velocity, U_{CH}

The lowest velocity at which the dilute-phase transport can operate at G given

Punwani et al (1976)

$$\frac{U_{CH}}{\varepsilon_{CH}} - U_T = \frac{G}{\rho_p(1-\varepsilon_{CH})}$$

$$\rho_f^{0.77} = \frac{2250D(\varepsilon_{CH}^{-4.7}-1)}{\left[\frac{U_{CH}}{\varepsilon_{CH}} - U_T\right]^2}$$

3) Saltation Velocity in Horizontal Transport

Figure 6.2 - Δp/ΔL vs. U(gas superficial velocity)

at various solids flow flux G

Saltation velocity, U_{SALT}

The gas velocity at which the solids to begin to settle out

Boundary between dilute phase flow and dense phase flow

Rizk(1973)

$$\underbrace{\frac{M_p}{\rho_f U_{SALT} A}}_{\text{solid loading}} = \left\{\frac{1}{10^{(1440x+1.96)}}\right\}\underbrace{\left\{\frac{U_{SALT}}{\sqrt{gD}}\right\}}_{\text{Froude number at saltation}}^{(1100x+2.5)} \quad \text{in SI}$$

where M_p *: particle mass flow rate*

D *: pipe diameter*

4) Fundamentals

Gas and particle velocity

Superficial velocity

$$U_{fs}=\frac{Q_f}{A}\quad and\quad U_{fp}=\frac{Q_p}{A}$$

Actual velocity

$$U_f=\frac{Q_f}{A\varepsilon}=\frac{U_{fs}}{\varepsilon}\quad and\quad U_p=\frac{Q_p}{A(1-\varepsilon)}=\frac{U_{ps}}{1-\varepsilon}$$

* *Slip velocity* U_{slip}

$$U_{rel}=U_f-U_p\equiv U_{slip}$$

Continuity

Gas mass flow rate

$$M_f=AU_f\varepsilon\rho_f$$

Particle mass flow rate

$$M_p=AU_p(1-\varepsilon)\rho_p$$

Solid loading

$$\frac{M_p}{M_f}=\frac{U_p(1-\varepsilon)\rho_p}{U_f\varepsilon\rho_f}$$

Pressure drop

From Newton's 2nd law of motion *Figure 6.3*

Rate of momentum for flowing gas-solid mixture

= Net force exerting on the mixture

↓

$$p_1-p_2=\underbrace{\frac{1}{2}\varepsilon\rho_f U_f^2}_{\text{gas acceleration}}+\underbrace{\frac{1}{2}\rho_p(1-\varepsilon)U_p^2}_{\text{solids acceleration}}+\underbrace{F_{fw}L}_{\text{gas-wall friction}}+\underbrace{F_{pw}L}_{\text{solids-wall friction}}$$

$$+\underbrace{\rho_f L\varepsilon g\sin\theta}_{\text{gas gravity}}+\underbrace{\rho_p L(1-\varepsilon)g\sin\theta}_{\text{solids gravity}}$$

5) Design for Dilute Phase Transport

Gas velocity

$U_f \sim 1.5 U_{SALT}$ *since* $U_{SALT} > U_{CH}$

for systems comprising both vertical and horizontal lines

$U_f \sim 1.5 U_{CH}$

for vertical line only

Table. Approximate air velocity for powder transport

Powder	U, m/s
Wheat, rice, plastic pellets	16 - 24
Grains, limestone powder	16 - 23
Soda ash, sugar	15 - 20
PVC powder	20 - 26
Carbon powder	18 - 24
Cement	18 - 28
Alumina powder	24 - 32
Sand	23 - 30

Pipeline pressure drop

$$F_{pw}L = 0.057GL\sqrt{\frac{g}{D}} \qquad \text{for vertical transport}$$

$$F_{pw}L = \frac{2f_p(1-\ )\rho_p U_p^2 L}{D} = \frac{2f_p G U_p L}{D} \qquad \text{for horizontal transport}$$

$$\text{where} \quad U_p = U_f(1-0.0638\,x^{0.3}\rho_p^{0.5}) \quad \text{and}$$

$$f_p = \frac{3}{8}\frac{\rho_f}{\rho_p}C_D\frac{D}{d_p}\left(\frac{U_f-U_p}{U_p}\right)$$

C_D: *drag coefficient (fn of* Re_p*)*

Bend

~ *7.5 m of vertical section pressure drop*

* *Downflow through vertical-to-horizontal bend :*

- *greater tendency for saltation*
- *avoided if possible.*

** Blinded tee bend : Figure 6.4 with respect to radius elbow*

- prolonging service life due to cushioning effect

- with the same pressure drop and solid attrition rate

Worked Example 6.1

Equipment

Figure 6.5 Positive pressure system

Figure 6.6 Negative Pressure system

** Centrifugal blowers(fan)*	*vs.*	*Positive displacement blower*
low pressure		*high pressure*
small amount of dust allowed		*no dust is allowed*

Some problems in pneumatic transport

	Possible	*Avoided by*
Blocking	*at high concentration region(around solid feeder and bend*	*feeding at dispersed state sufficient acceleration length and adequate bend curvature*
Adhesion	*with moisty, low-melting or electrically charged powder*	*adequate range of gas velocity*
Attrition	*at bend*	*- low gas velocity* *- higher solid load* *- changing collision angle and bend material.*

6) Dense Phase Transport

Flow Patterns

- Horizontal - Figure 6.7

Saltating flow - unstable, bad flow pattern

*Discontinuous dense phase flow**

*Dune Flow / Discrete Plug Flow / Plug Flow**

Continuous Dense Phase Flow - requires high pressure
adequate for short-pipe transport

Equipment

Blow tanks : with fluidizing element (Figure 6.13)
without fluidizing element (Figure 6.14)

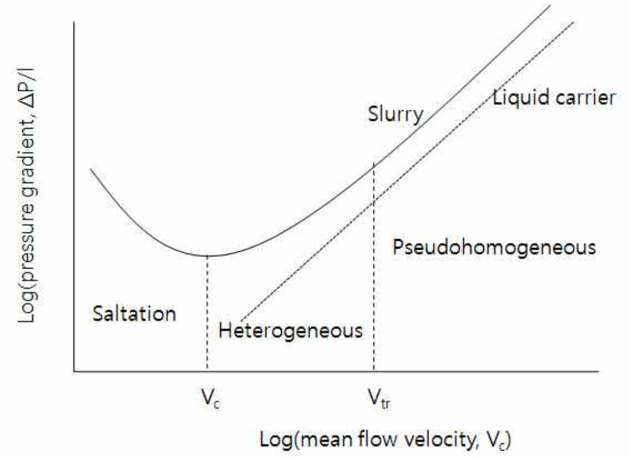
Plug formation : air knife (Figure 6.10)
air valve (Figure 6.11)
diaphragm (Figure 6.12)

Plug break-up : bypass (Figure 6.8)
pressure actuated valves (Figure 6.9)

Design and Operation

- *Use of test facilities + past experience for pipe size, air flow rate and type of dense phase system*
- *Group A, D better than Group B, C for dense phase conveying*
- *Higher permeability: more suitable for plug flow type conveying*
- *Higher air retention: more suitable for dune mode flow*

4.3 Flow of Liquid-Solid Suspension (Slurries)(Supplement)



Characteristics of hydraulic transport

Transition velocity

Durand(1953)

$$U_{tr} = 11.9(U_T D)^{1/2} x^{1/4}$$

where D: pipe diameter

Critical(saltation) velocity

Durand(1953)

ρ $U_c = F_L[2gD({}_p/\ {}_f - 1)]^{1/2}$

ε *where F_L: function of x and*

Hanks (1980)

ρ $U_c = 3.12(1-\ )^{0.186}\left(\frac{x}{D}\right)^{1/6}[2gD({}_p/\ {}_f - 1)]^{1/2}$