UNIVERSITEIT TWENTE.

Multiphase Reactor Technology

Fluidization (summary)

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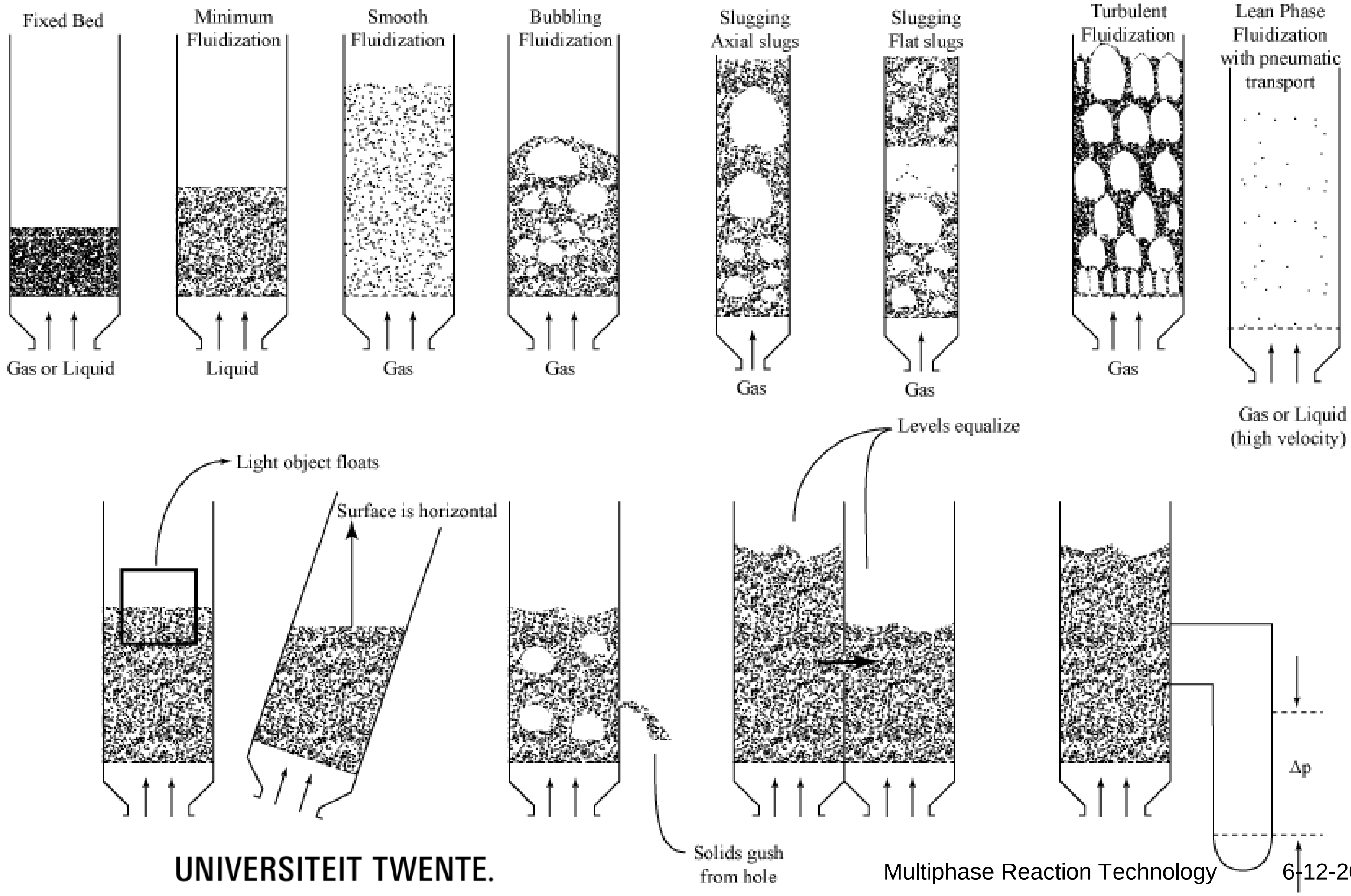
Further reading

- Detailed slide packages (blackboard)
- Froment & Bischoff, chapter 13 (blackboard)
- Van Deemter model
- Kuni & Levenspiel

Introduction
Phenomenon of fluidization



Application

- Types of reactions: catalytic (heterogeneous) reactions, reacting solids
- Physical processes: drying, heat exchange, adsorption (also chemisorption), coating, granulation

✓ Heat exchange and drying

✓ Coating and granulation

✓ Gas purification via adsorption

✓ Chemical synthesis(acrylonitrile, maleic and phtalic anhydride)

✓ Polymerization of lower olefines(propylene)

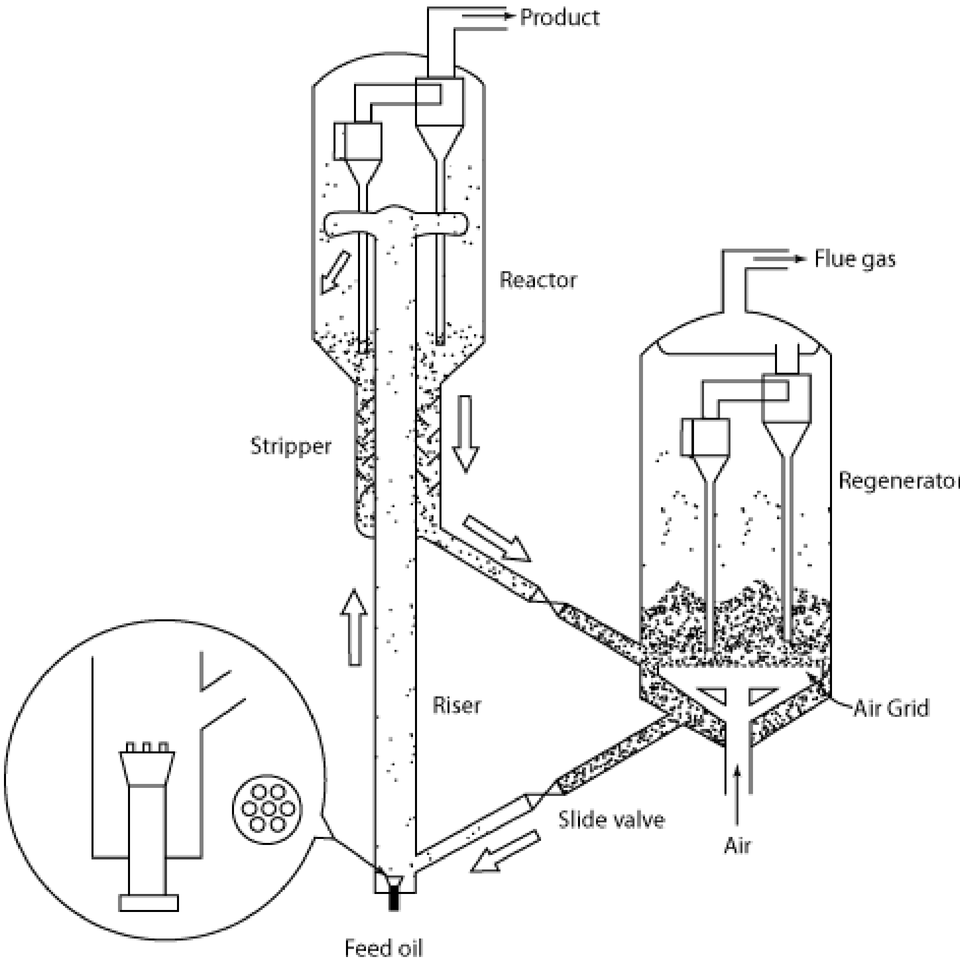
✓ Fischer-Tropsch synthesis

✓ Fluid coking and Flexi-Coking

✓ Combustion and incineration

✓ Fluid Catalytic Cracking (FCC)

Fluid Catalytic Cracking (FCC)



Zeolite catalyst (very active)

Reaction takes place in the riser

High conversion in short contact time

Plug flow ➔ high yield of gasoline
(no overcracking)

Reactor at 470-550°C up to 3.5 bar
Regenerator at 580-700°C up to 4 bar

Riser 1.5 m ID
Regenerator 8 m ID

Catalyst circulation rate 15-30tons/min

Characterisation of particulate solids

- Equivalent spherical diameter d_{sph}

$$\frac{\pi}{6} d_{sph}^3 = V_{particle}$$

- Sphericity ϕ_s (particle and sphere have same volume):

$$\phi_s = \frac{A_{sphere}}{A_{particle}}$$

- Specific surface of particle a'

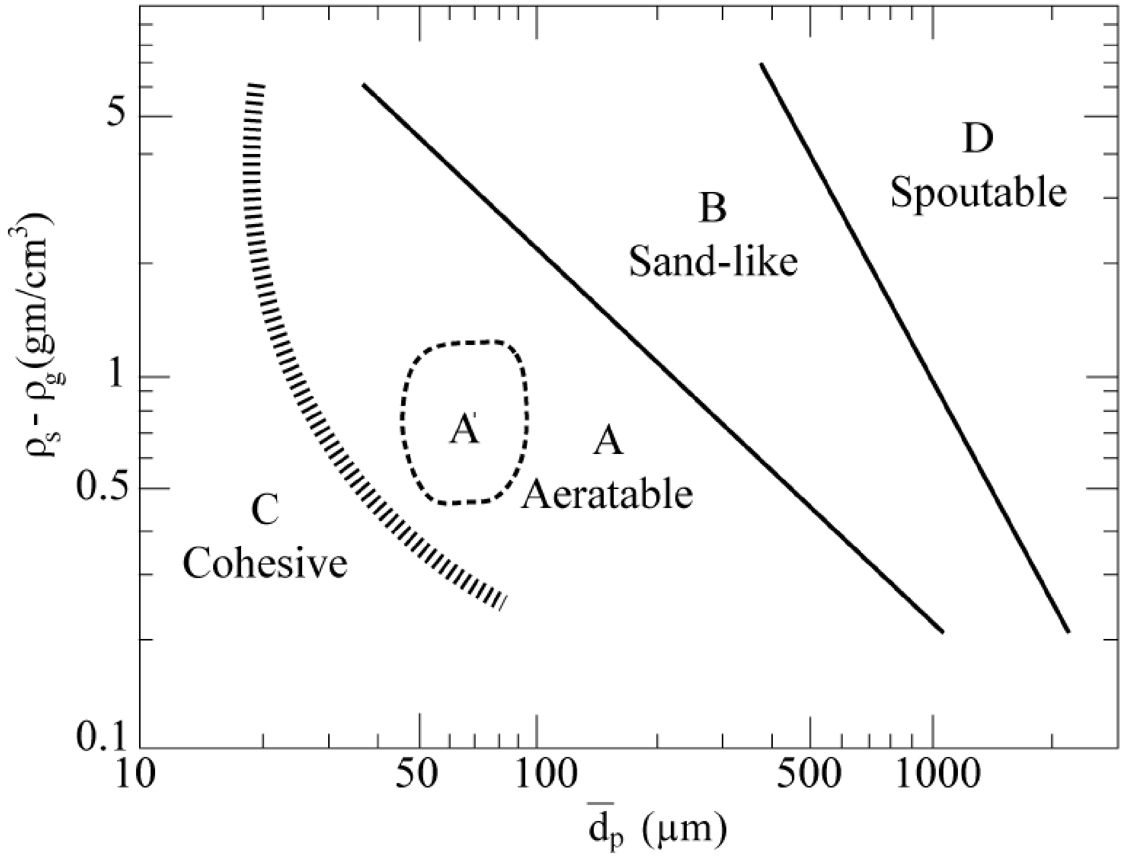
$$a' = \frac{A_{particle}}{V_{particle}} = \frac{6}{\phi_s d_{sph}}$$

- Specific bed surface a:

$$a = \frac{6\,(1-\varepsilon_b)}{\phi_s d_{sph}}$$

Fluidization Regimes
Geldart's classification

- Four types of fluidization behaviour (A, B, C and D)



Fluidization Regimes
Geldart's classification

- Group A: Aeratable, or particles with a small size and/or density (<1.4 g/cm^3). Smooth fluidization at low u and controlled bubbling (small bubbles). FCC is a typical example

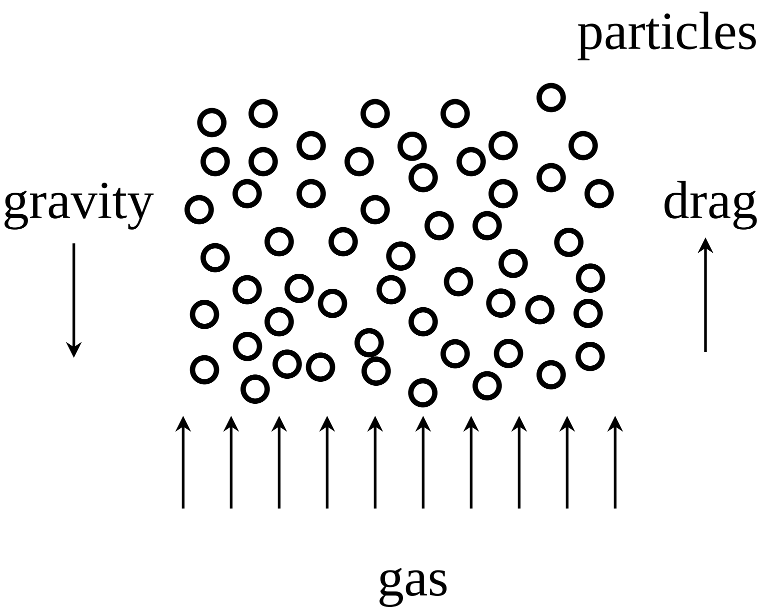
- Group B: Sandlike, most particles of size 40 μm<d_p<500 μm and density 1.4<ρ_s<4 g/cm^3. These particles fluidize well with vigorous bubbling action and bubbles that grow large (coalescence)

- Group C: Cohesive, or very fine powders. Normal fluidization is extremely difficult due to strong interparticle forces which exceed the gas drag. Examples: flour and starch

- Group D: Spoutable, or large and/or dense particles. Deep beds of these particles are difficult to fluidize due to "explosive" bubble growth (coalesence). Examples: peas and coffee beans

note: Geldart's classification effectively includes only ρ_s and d_p !!!

Minimum fluidization velocity u_{mf}



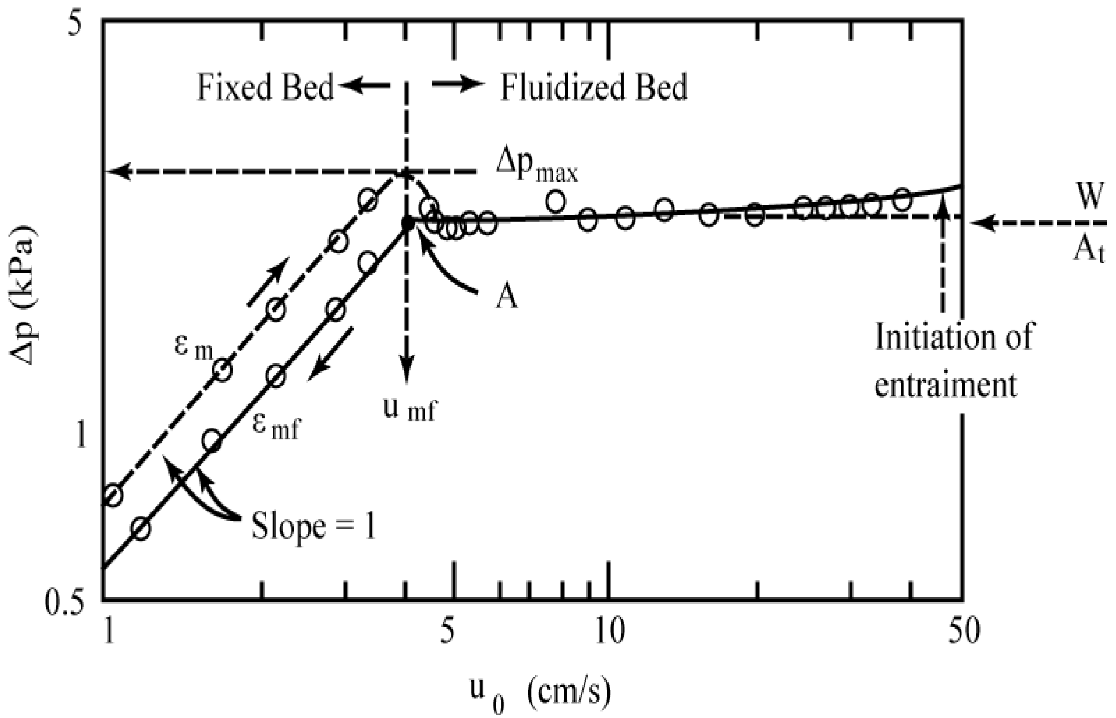
$F_d > F_g$: entrainment

$F_d = F_g$: fluidization

$F_d < F_g$: defluidization

Minimum fluidization velocity u_{mf}

- Pressure drop Δp versus fluidization velocity u_0



- Incipient or minimum fluidization: weight of particles is equal to the drag force

Minimum fluidization velocity (u_{mf})

drag force due to gas flow = weight of particles

Or

(pressure drop) (cross sectional area) = (volume of bed) (fraction of solid) (specific weight of solid)

$$\Delta p_b A_t = A_t L_{mf}\left(1-\varepsilon_{mf}\right)\left[\left(\rho_s-\rho_g\right)g\right]$$

Minimum fluidization velocity u_{mf} and terminal velocity u_t

- Expression for u_{mf}:

$$\frac{1.75}{\varepsilon_{mf}^3 \phi_s}\left(\frac{d_p u_{mf} \rho_g}{\mu}\right)^2 + 150\frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3 \phi_s^2}\left(\frac{d_p u_{mf} \rho_g}{\mu}\right) = \frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}$$

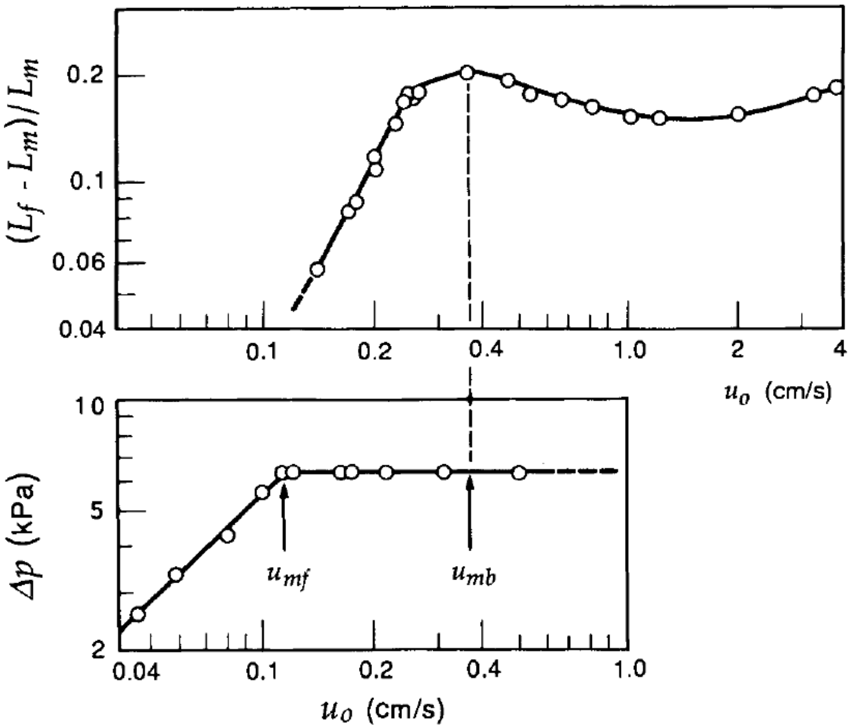
- Dimensionless form:

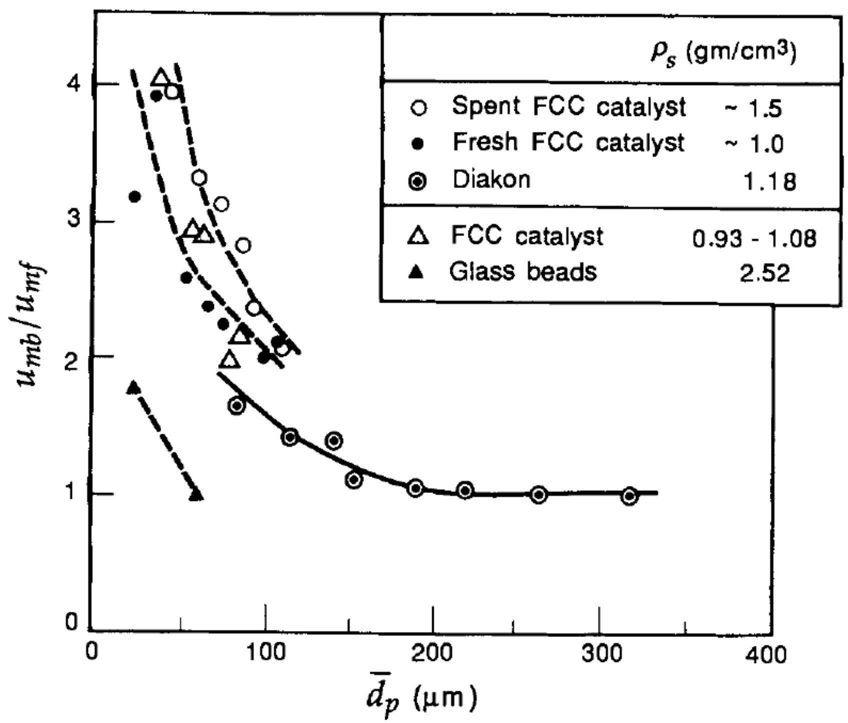
$$\frac{1.75}{\varepsilon_{mf}^3 \phi_s}\left(Re_{p,mf}\right)^2 + 150\frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3 \phi_s^2}\left(Re_{p,mf}\right) = Ar$$

Archimedes number

Particle Reynolds number

Minimum bubbling velocity





Characteristics of selected particles

	Size, d_p (mm)						
Particles	0,02	0,05	0,07	0,1	0,2	0,3	0,4
Sharp sand,Φ_s=0,67	-	0,6	0,59	0,58	0,54	0,5	0,49
Round sand, Φ_s=0,86	-	0,56	0,52	0,48	0,44	0,42	-
Mixed round sand	-	-	0,42	0,42	0,41	-	-
Coal and glass powder	0,72	0,67	0,64	0,62	0,57	0,56	-
Anthracite coal, Φ_s=0,63	-	0,62	0,61	0,6	0,56	0,53	0,51
Absorption carbon	0,74	0,72	0,71	0,69	-	-	-
Fischer-Tropsch catalyst, Φ_s=0,58	-	-	-	0,58	0,56	0,55	-
Carborundum	-	0,61	0,59	0,56	0,48	-	-

Terminal velocity u_t

- Terminal velocity of particle of size dp falling through a fluid :

$$u_t = \left[\frac{4d_p(\rho_s - \rho_g)g}{3\rho_g C_d}\right]^{\frac{1}{2}}$$

- Drag coefficient C_d given by: Powder Technology 58 (1989) 63

$$C_d = \frac{24}{Re_p}\left[1 + \left(8.1716e^{-4.0655\phi_s}\right)Re_p^{0.0964+0.5565\phi_s}\right] + \frac{73.69Re_p e^{-5.0748\phi_s}}{Re_p + 5.378e^{6.2122\phi_s}}$$

- Drag coefficient for spherical particles (ϕ_s=1):

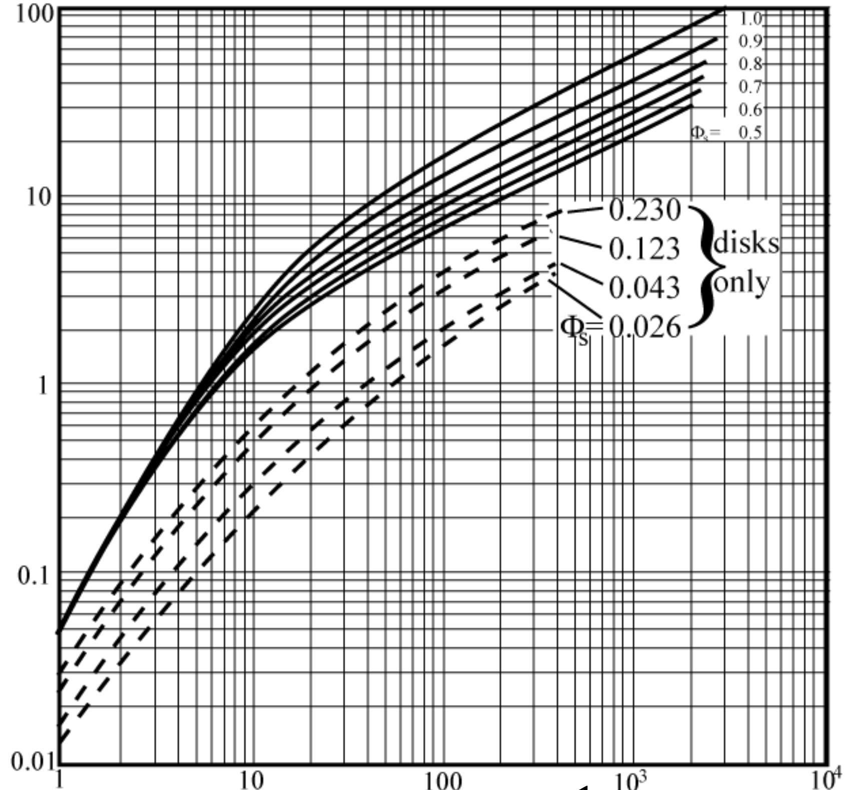
$$C_d = \frac{24}{Re_p} + 3.3643Re_p^{0.3471} + \frac{0.4607Re_p}{Re_p + 2682.5}$$

$$Re_p = \frac{d_{sph}u_t\rho_g}{\mu}$$

Terminal velocity u_t

- Graphical determination of terminal velocity

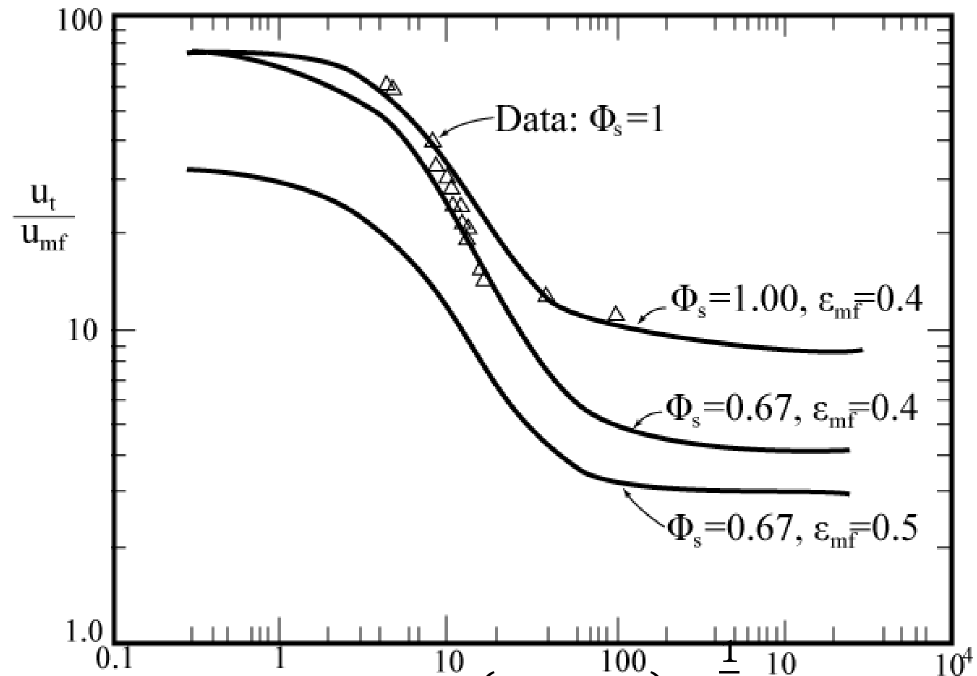
$$u_t^* = u_t\left[\frac{\rho_g^2}{\mu(\rho_s - \rho_g)g}\right]^{\frac{1}{3}}$$



$$d_p^* = d_p\left[\frac{\rho_g(\rho_s - \rho_g)g}{\mu^2}\right]^{\frac{1}{3}}$$

Fluidization Regimes

Ratio of terminal velocity u_t and minimum fluidization velocity u_{mf}



To avoid carryover of particles the velocity should be between u_{mf} and u_t

$$d_{sph}^{*} = d_{sph}\left[\frac{\rho_g(\rho_s - \rho_g)g}{\mu^2}\right]^{\frac{1}{3}}$$

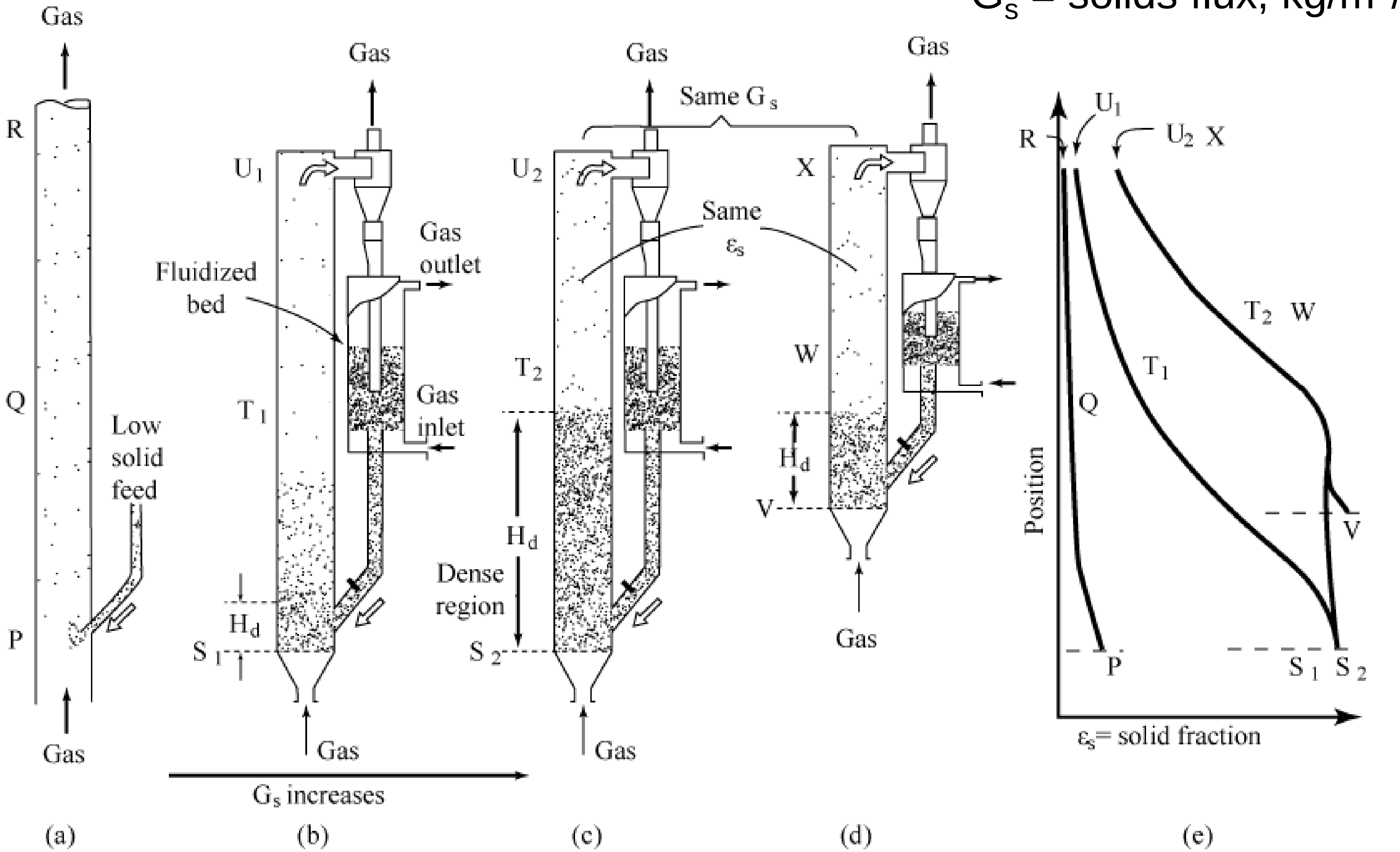
- Useful velocity range for large particles is much smaller than that for small particles (reflected in applications)

High velocity fluidization

- Pneumatic conveying [a] and fast fluidization [b], [c] and [d]

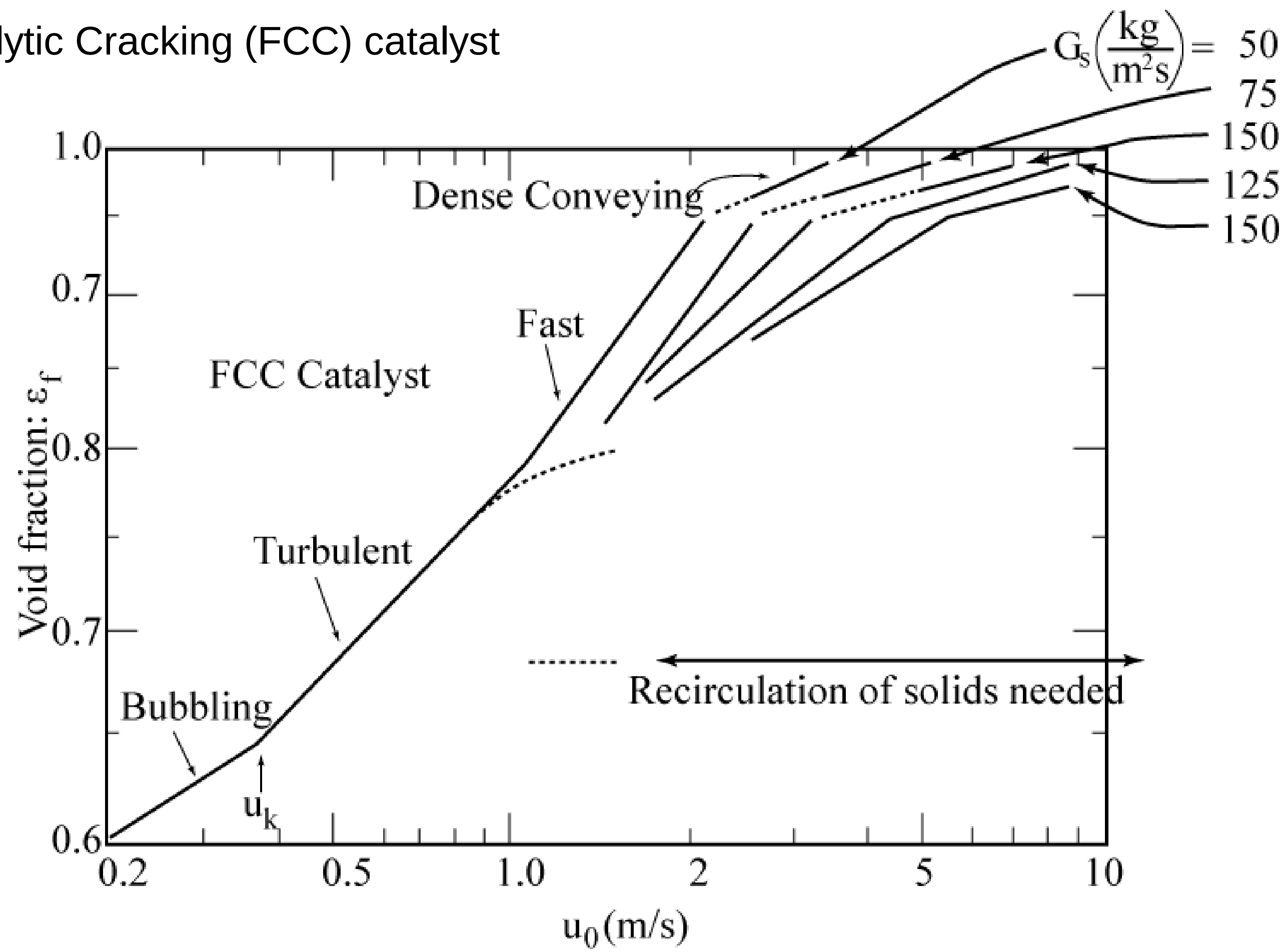
G_s = solids flux, kg/m²/s

u=20u$_t$



Fluidization Regimes

- Void fraction ε_f and fluidization regimes as a function of the superficial gas velocity u_0 for Fluid Catalytic Cracking (FCC) catalyst



Bubbling fluidized beds

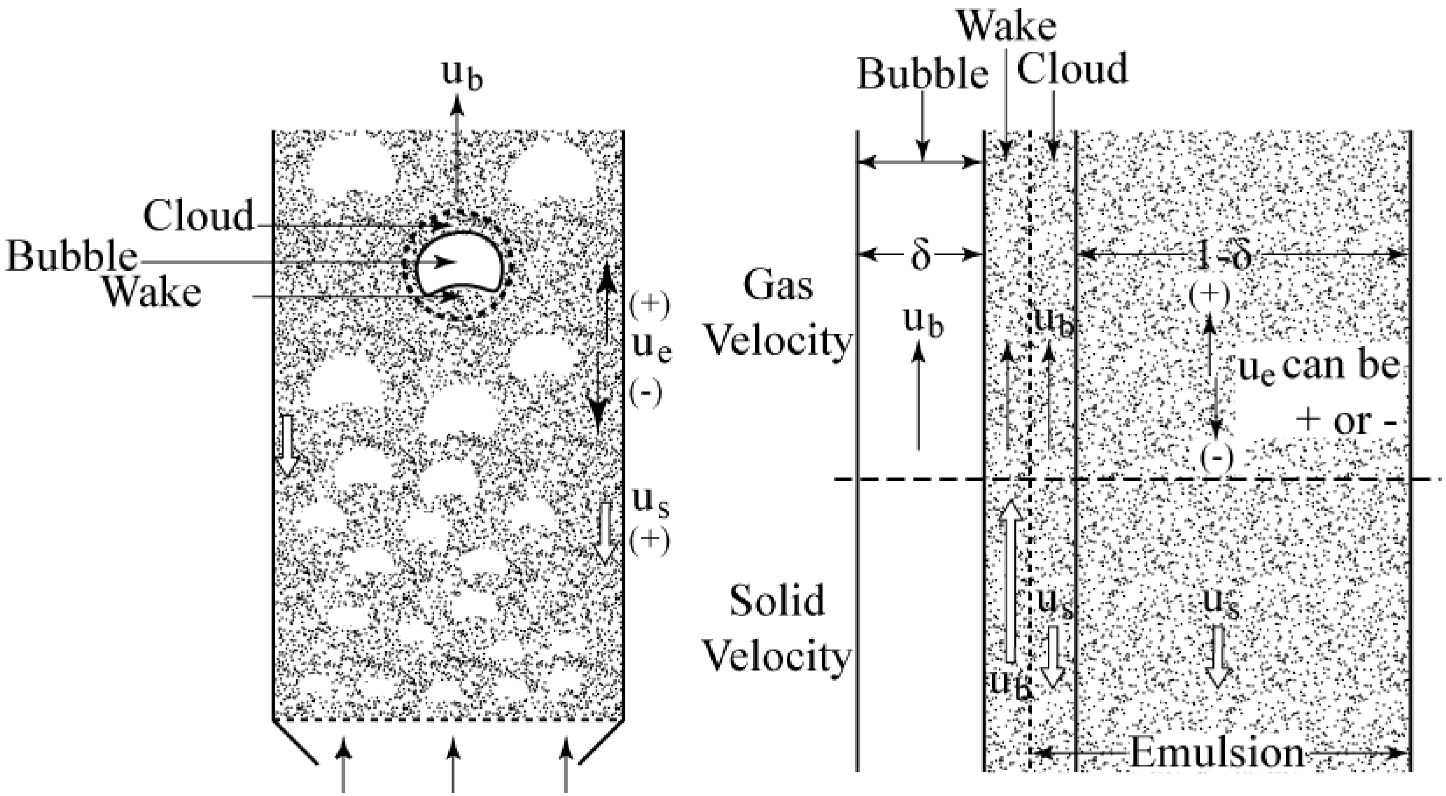
- Are bubbles actually bubbles?
- Bubbles are actually void fractions through which gas flows
    - This flow is characterized by a mass transfer K

$$K = \frac{volumetric\ flow\ through\ bubble\ m^3/s}{m^3\ bubble}$$

$$K = \frac{volumetric\ flow\ through\ bubble\ m^3/s}{m^3\ bed}$$

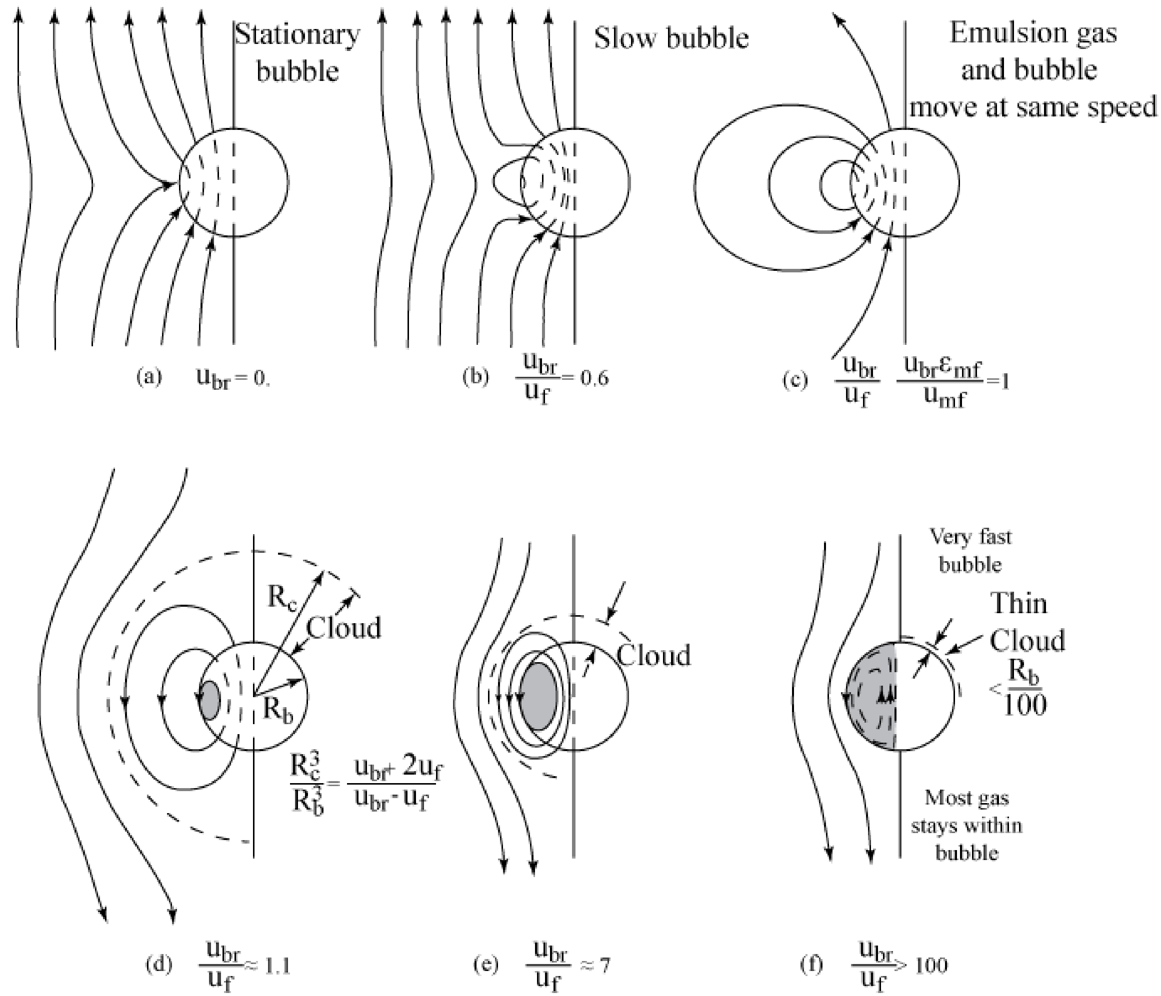
Unit is 1/s

Schematic representation



Gas in bubble = PFR, Gas in emulsion CSTR to PFR
Solids is CSTR – axially dispersed with / without circulation flow

Gas bubble behaviour



Gas dispersion and gas exchange

- Gas exchange between bubble and emulsion phase

U_b is the bubble rise velocity
= linear velocity
V_b = bubble volume[#]

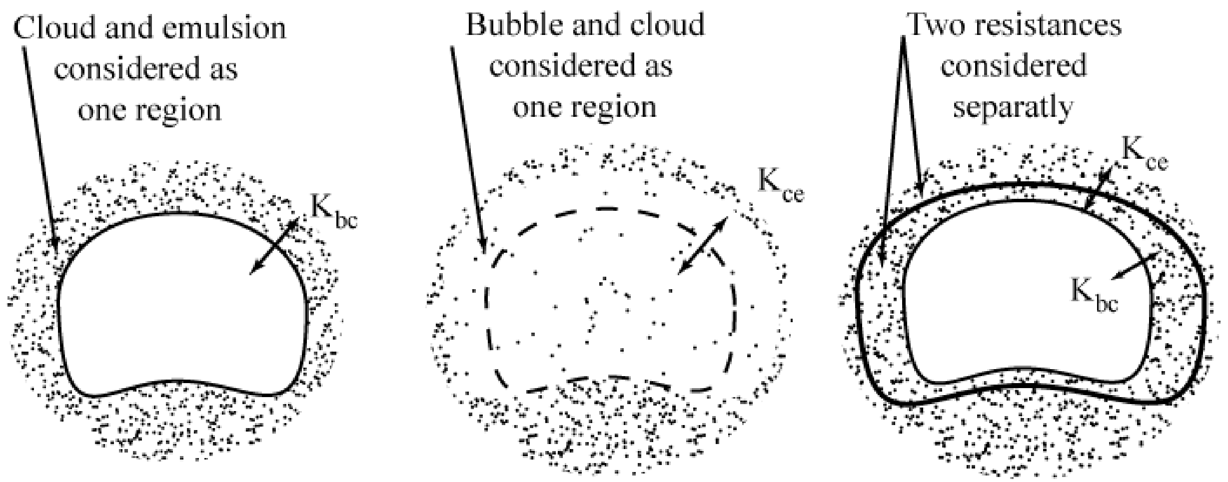
  - Definitions of gas exchange:

Local parameter

$$-\frac{1}{V_b}\frac{dN_{Ab}}{dt} = -u_b\frac{dC_{ab}}{dz} = K_{be}(C_{Ab} - C_{Ae}) = K_{bc}(C_{Ab} - C_{Ac}) = K_{ce}(C_{Ac} - C_{Ae})$$

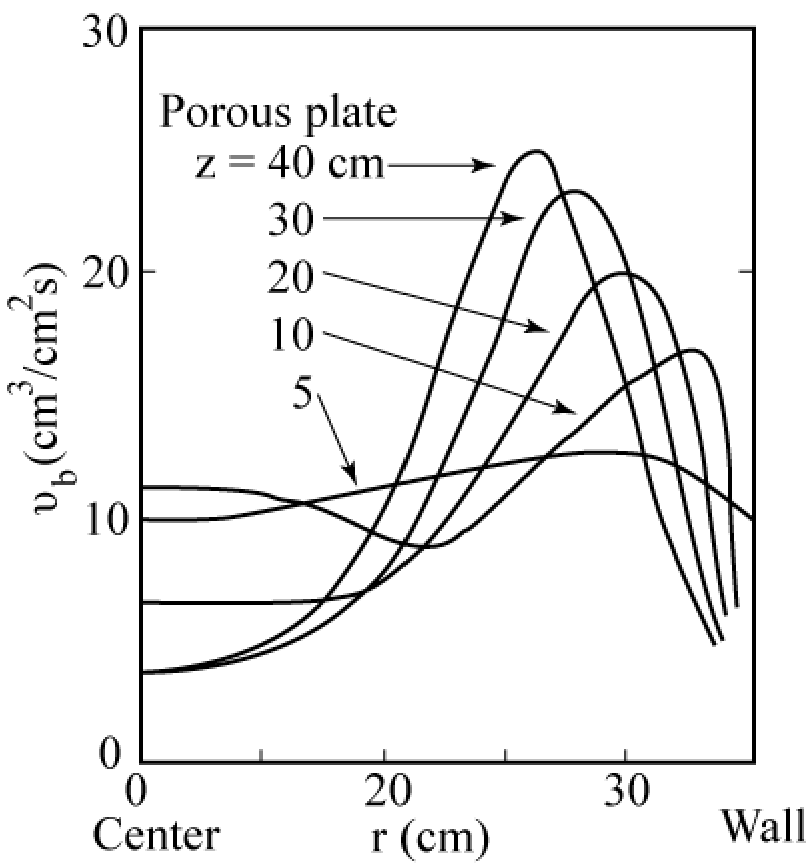
  - Relationship between interchange coefficients:

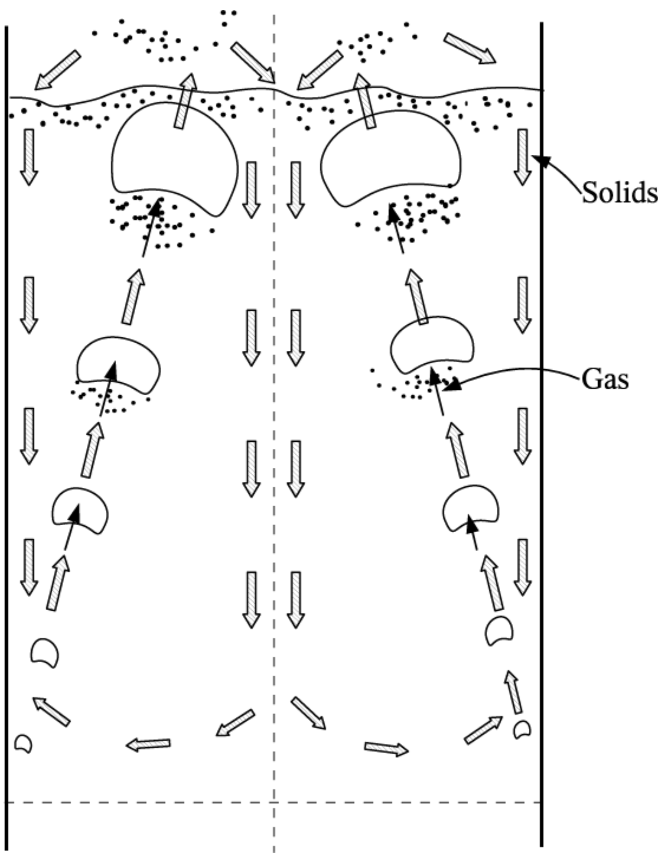
$$\frac{1}{K_{be}} = \frac{1}{K_{bc}} + \frac{1}{K_{ce}}$$



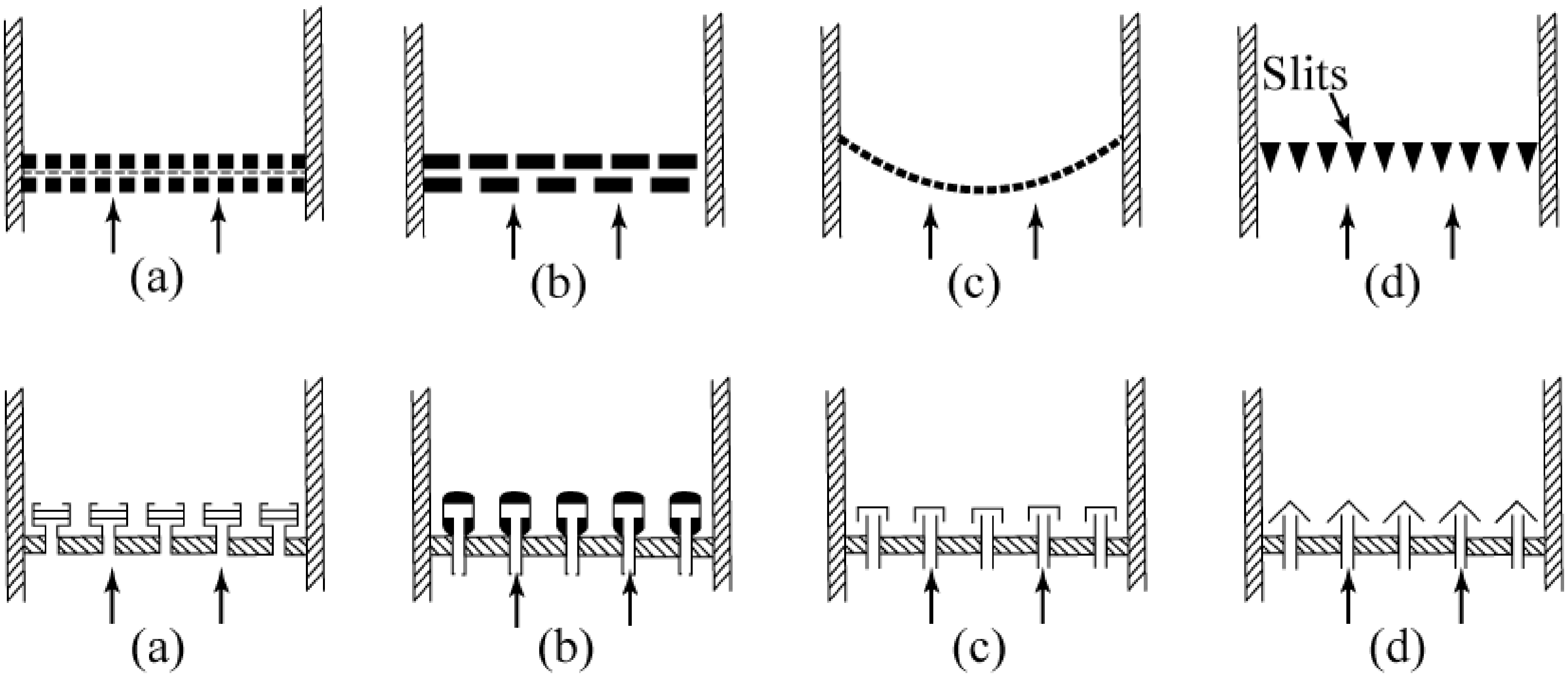
[#]Vb also used for rise velocity

Bubble flow patterns





Gas distributors



Pressure drop over distributor needs to be ~ 50% of pressure drop over bed

Gas distributor design

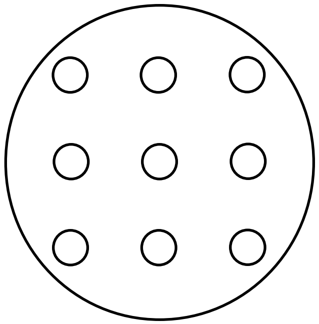
- Initial bubble diameter (cm) for distributor with N_{or} orifices per unit area (cm^{-2}) of plate where distance between orifices equals l_{or}:

$$d_{bo} = \frac{1.30}{g^{0.2}}\left[\frac{u_0 - u_{mf}}{N_{or}}\right]^{0.4}$$

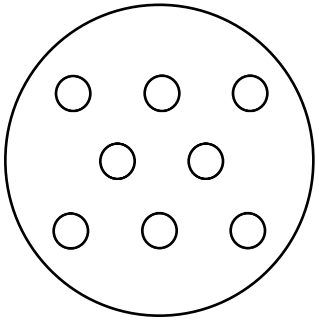
low gas flow rate
non-touching bubbles
$d_{bo} < l_{or}$

$$d_{bo} = \frac{2.78}{g}\left(u_0 - u_{mf}\right)^2$$

high gas flow rate
touching bubbles
$d_{bo} > l_{or}$



$$N_{or} = \frac{1}{l_{or}^2}$$



$$N_{or} = \frac{2}{\sqrt{3}\,l_{or}^2}$$

Gas bubble behaviour

- Emulsion gas flow and voidage

$$\left(\frac{\varepsilon_e}{\varepsilon_{mf}}\right)^3 = \frac{1-\varepsilon_{mf}}{1-\varepsilon_e} = \left(\frac{u_e}{u_{mf}}\right)^{0.7}$$

- Bubble size correlation due to Mori and Wen (size in cm):

$$\frac{d_{bm}-d_b}{d_{bm}-d_{bo}} = e^{-0.3\left(\frac{z}{d_t}\right)}$$ with $$d_{bm} = 0.65\left[\frac{\pi}{4}d_t^2(u_0-u_{mf})\right]^{0.4}$$

  - range of experimental conditions:

d_t<=1.3 m 0.5<=u_{mf}<=20 cm/s

60<=d_p<=450 μm u_o-u_{mf}<=48 cm/s

Gas bubble behaviour

- Bubble size correlation due to Werther (size in cm):

$$d_b = 0.853\left[1 + 0.272\left(u_0 - u_{mf}\right)\right]^{\frac{1}{3}}(1 + 0.0684z)^{1.21}$$

  - range of experimental conditions (valid for porous plate)

$d_t >= 20$ cm $\quad 1 <= u_{mf} <= 8$ cm/s

$100 <= d_p <= 350$ μm $\quad 5 <= u_0 - u_{mf} <= 30$ cm/s

- Werther correlation can be adapted for gas distributor with orifices

(initial bubble size not zero in this case):

Gas bubble behaviour

- Bubble rise velocity (linear velocity) correlations

  - Geldart A solids with d_t<=1 m (velocity in m/s):

$$u_b = 1.55\{(u_0 - u_{mf}) + 14.1(d_b + 0.005)\}d_t^{0.32} + u_{br}$$

  - Geldart B solids with d_t<= 1m (velocity in m/s):

$$u_b = 1.6\{(u_0 - u_{mf}) + 1.13d_b^{0.5}\}d_t^{1.35} + u_{br}$$

  - Note that u_{br} denotes the rise velocity of single bubbles, given by:

$$u_{br} = 0.711\sqrt{gd_b}$$

Gas exchange: models / correlations

- Estimation of gas exchange coefficients (Bubble model + Higbie penetration model)

  - bubble to cloud exchange coefficient K_{bc}:

$$K_{bc} = 4.5\left(\frac{u_{mf}}{d_b}\right) + 5.85\left(\frac{D^{\frac{1}{2}} g^{\frac{1}{4}}}{d_b^{\frac{5}{4}}}\right)$$

  - cloud to emulsion exchange coefficient K_{ce}:

$$K_{ce} = 6.77\left(\frac{D\,\varepsilon_{mf}\,(0.711)(g d_b)^{\frac{1}{2}}}{d_b^3}\right)^{\frac{1}{2}} = 6.77\left(\frac{D\,\varepsilon_{mf}\,u_{br}}{d_b^3}\right)^{\frac{1}{2}}$$

note: K_{bc} contains both convective and diffusive contributions

Gas exchange: RTD measurements

- Mass balances for two-zone model

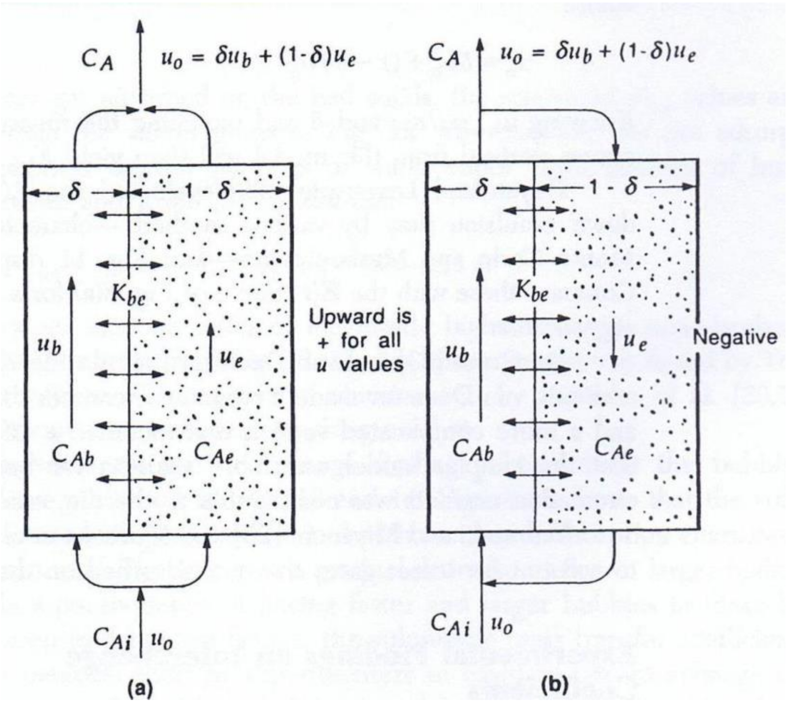
  - bubble phase:

$$\frac{\partial C_{Ab}}{\partial t} + u_b \frac{\partial C_{Ab}}{\partial z} = K_{be}(C_{Ab} - C_{Ae})$$

  - emulsion phase:

$$\frac{\partial C_{Ae}}{\partial t} + \frac{u_e}{\varepsilon_e}\frac{\partial C_{Ae}}{\partial z} = \frac{\delta}{1-\delta} K_{be}(C_{Ab} - C_{Ae})$$

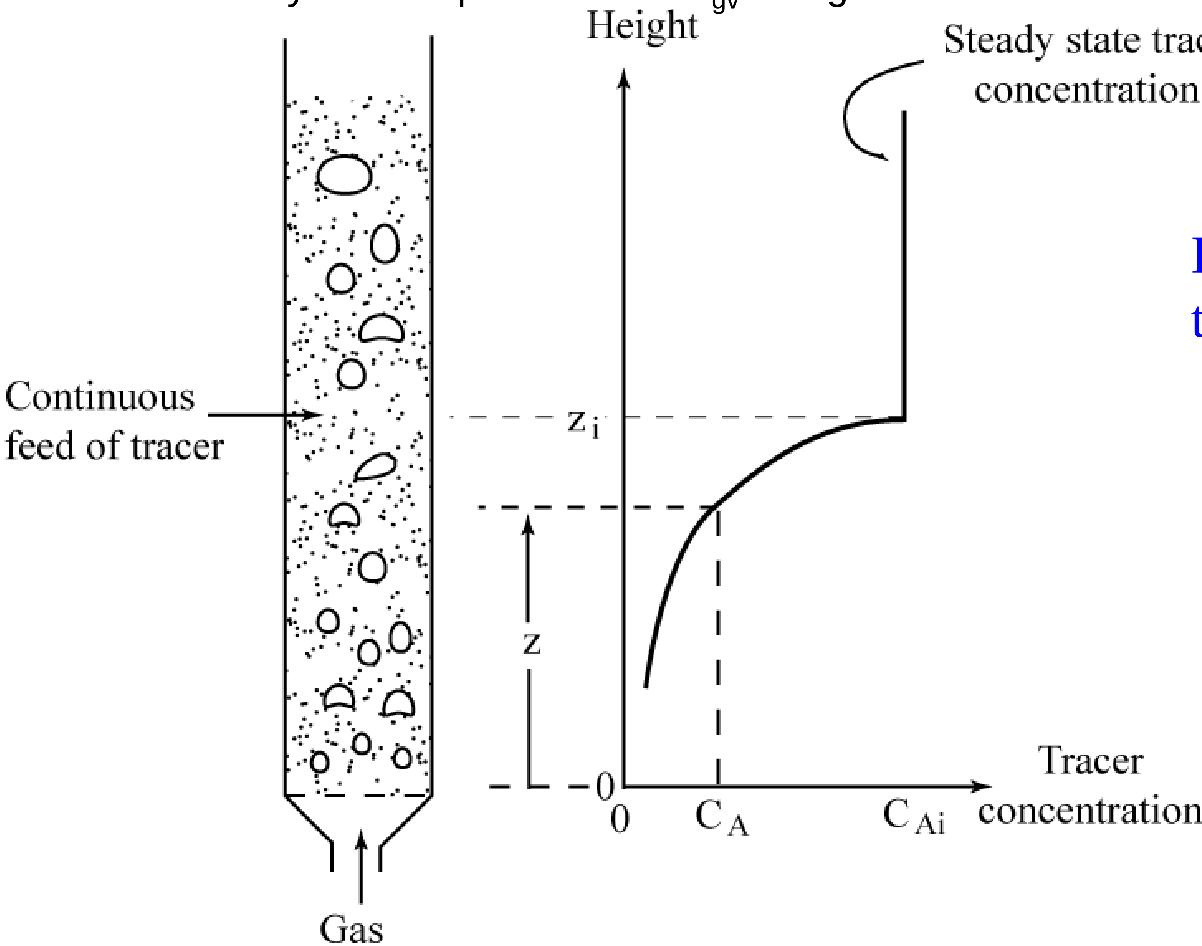
integral balance: $u_0 = \delta u_b + (1-\delta)u_e$



empirical information on bubble velocity u_b, emulsion velocity u_e, emulsion voidage ε_e and bubble holdup δ required (from integral balance one of the quantities can be computed from the others)

Gas dispersion

- Steady state experiment for D_{gv} in a gas-fluidized bed



Differential equation governing the axial dispersion process

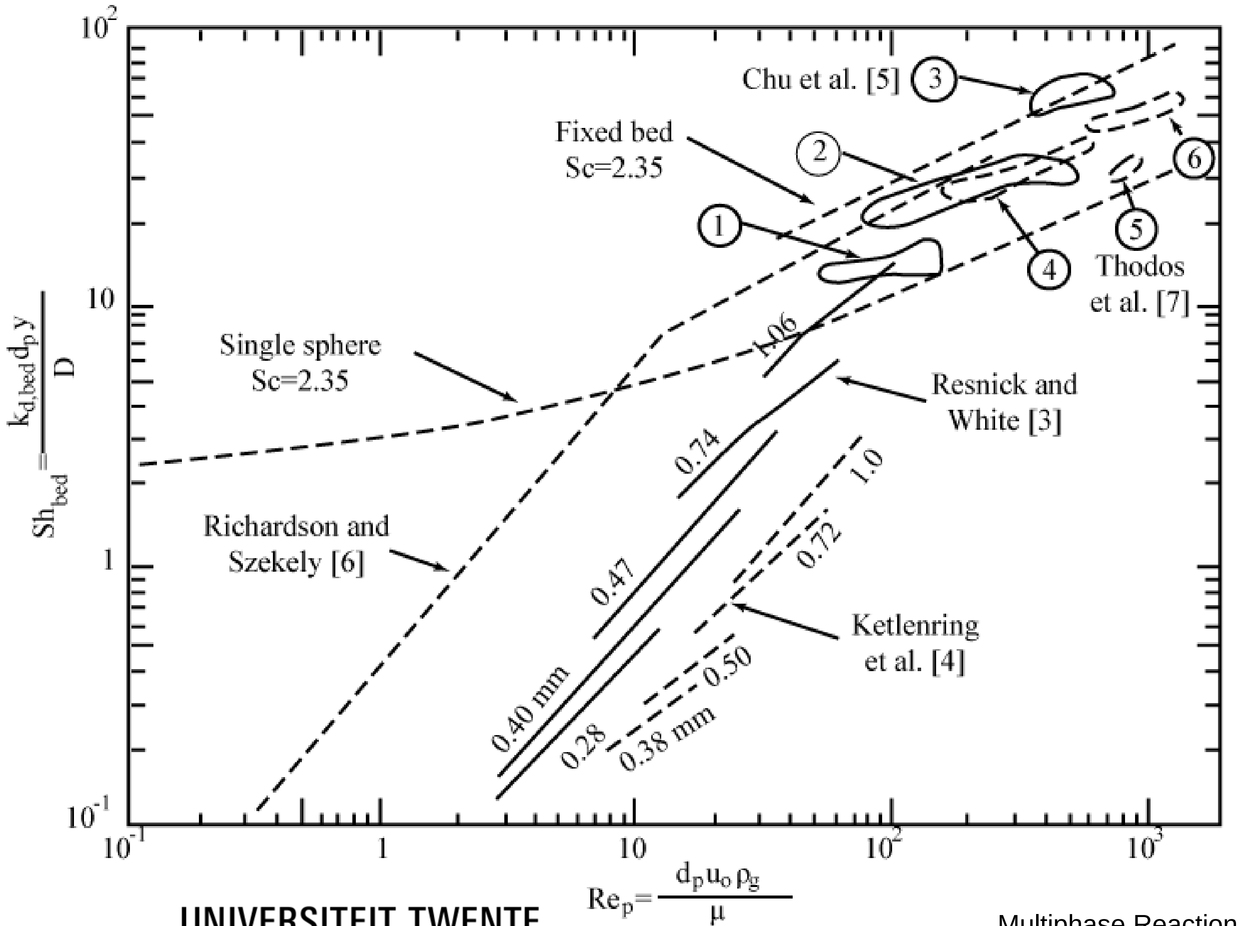
$$D_{gv}\frac{d^2C_A}{dz^2} - u_0\frac{dC_A}{dz} = 0$$

$$\frac{C_A}{C_{Ai}} = \exp\left[\frac{-u_0(z_i - z)}{D_{gv}}\right]$$

At Umf, we measure the dispersion in the emulsion phase

Particle-to-gas mass transfer

- Experimental findings on mass transfer in fluidized beds



Note: Sh_{bed} does not approach 2 at low Re_p

Heat transfer

- Correlations for (see detailed slides)
  - Particle to gas
  - Immersed object to bed
  - Bed to wall

Models for BFB reactors

- Kuni & Levenspiel

- Bubble/clould/emulsion
- Bubble is pfr
- U_b from theory / correlation
- K_{be} from theory / correlation
- U_b & K_{be} are / can be a function of the height of the bed
- Emulsion gas: pfr with zero velocity

- Van Deemter / Van Swaaij

- Bubble/emulsion
- Bubble is pfr
- Gas exchange from RTD / tracer tests
- Disperson from RTD / tracer tests
- Gas exchange and dispersion are constant
- Emulsion gas axial dispersed with/without gas flow

Kuni & Levenspiel fine particle model

$u_o/u_{mf} >> 1$ and $u_b/u_{mf} >> 1$

- ✓ Fresh entering feed gas containing reactant A enters bed and on contact with the fine catalyst powder reacts according to a first order reaction
- ✓ Bed consists of three regions: bubble [b], cloud [c] and emulsion [e] region (bubble wake is part of the cloud)
- ✓ All feed gas passes through the bed as bubbles ($u_o >> u_{mf}$)
- ✓ gas interchange rate between bubble and cloud and between cloud and emulsion are given by K_{bc} and K_{ce} respectively

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- definition of solids distribution in bubble, cloud and emulsion:

γ_b, γ_c and γ_e denote respectively volume of solids dispersed in bubble, cloud and emulsion divided by volume of bubble

- Balance formulation for reactant A:

overall disappearance in bubble	=	reaction in bubble	+	transfer to cloud-wake
transfer to cloud-wake	=	reaction in cloud-wake	+	transfer to emulsion
transfer to emulsion	=	reaction in emulsion		

Dense Fluidized Beds
Conversion of gas due to catalytic reactions (1st order)

- Mass balance for reactant A: $-\frac{dC_{A,b}}{dt} = -u_b \frac{dC_{A,b}}{dz} = \gamma_b K_r C_{A,b} + K_{b,c}(C_{A,b} - C_{A,c})$

- additional equations $K_{b,c}(C_{A,b} - C_{A,c}) = \gamma_c K_r C_{A,c} + K_{c,e}(C_{A,c} - C_{A,e})$

$$K_{c,e}\left(C_{A,c} - C_{A,e}\right) = \gamma_e K_r C_{A,e}$$

- Upon eliminating concentrations of A in emulsion and cloud we get:

$$-u_b \frac{dC_{A,b}}{dz} = K_f C_{A,b}$$

$$K_f = \gamma_b K_r + \cfrac{1}{\cfrac{1}{K_{b,c}} + \cfrac{1}{\gamma_c K_r + \cfrac{1}{\cfrac{1}{K_{c,e}} + \cfrac{1}{\gamma_e K_r}}}}$$

K_f represents an overall rate constant for chemical reaction accounting for all relevant mass transfer resistances in fine particle gas-fluidized beds

Inspection: combination of resistances in series/parallel

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Integration of mass balance between inlet and certain position in bed:

$$\frac{C_{A,b}}{C_{A,inlet}} = \frac{C_{A,b}}{C_{A,i}} = \exp\left[-K_f \frac{z}{u_b}\right]$$

assumption: bubble size remains approximately constant in bed

- Since “all” gas fed to the bottom of the bed passes in the form of bubbles we can write for the reactor as a whole:

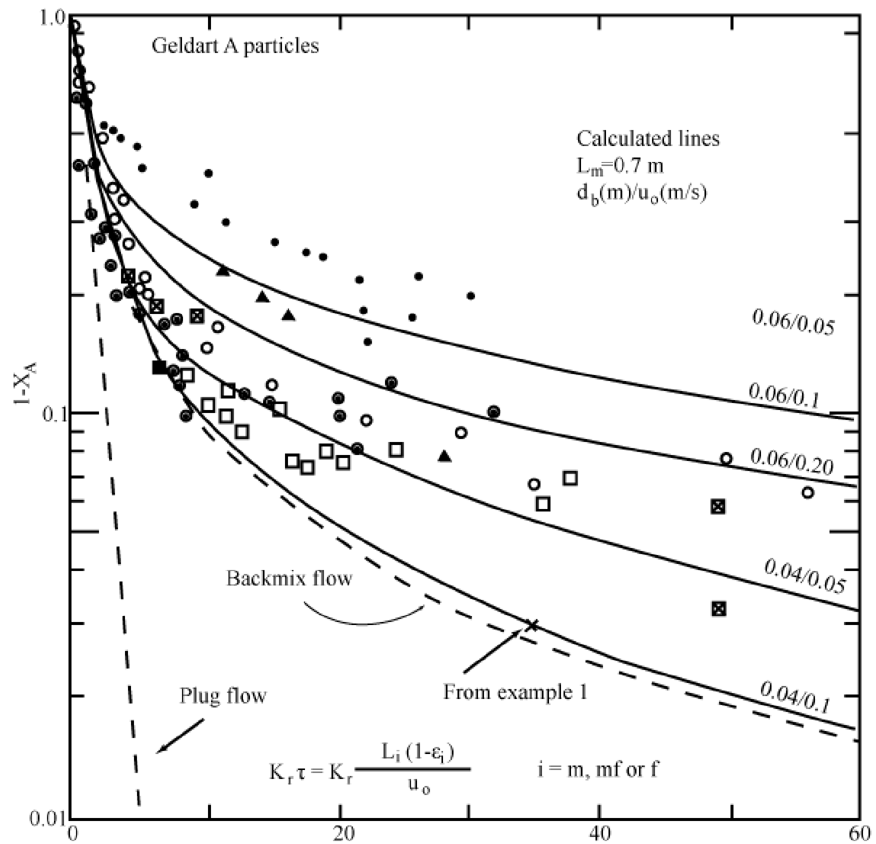
$$1 - X_A = \frac{C_{A,0}}{C_{A,i}} = \exp\left[-K_f \frac{z}{u_b}\right]$$

contribution of emulsion is negligible (not valid for coarse particles)

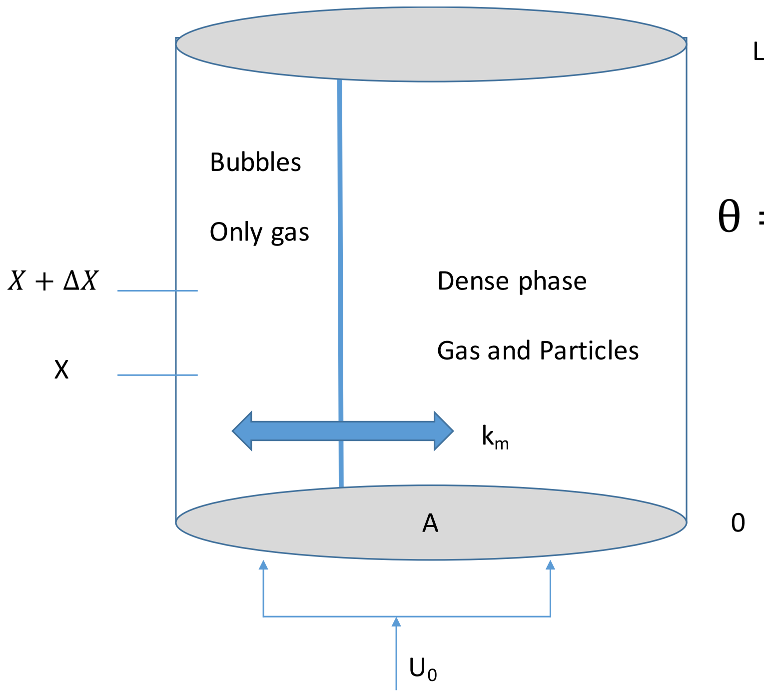
- Note that all parameters (except Kr of course) can be obtained from the Kunii and Levenspiel (K-L) model discussed before in detail !!!

estimates of extent of conversion can be made for fine particle beds

Experimental findings (fine particles)



Van Deemter / van Swaaij model



$$\theta = \frac{X}{L}$$

$$N_T = \frac{k_m L}{U_0} \qquad \mathrm{H} = \frac{L}{N}$$

$$N_E = \frac{U_0 L}{D_e}$$

$$N_R = \frac{K_r \rho_c f_s L}{U_0} \qquad f_s = \frac{m^3\ solids\ in\ bed}{m^3_{bed}}$$

First order reaction, no gas through dense phase

$$\frac{dC^b}{d\theta} = -\frac{k_m L}{U_0}\left(C^b - C^d\right) = -N_T\left(C^b - C^d\right)$$

$$0 = \frac{1}{N_E}\frac{d^2C^d}{d\theta^2} + N_T\left(C^b - C^d\right) - N_R C^d$$

Riser (CFB)

- Simples model:
  - Plug flow for fluid
  - Plug flow for solids
  - Hold-up of solids from correlations
  - Sh & Nu relations with Re_p based on slip velocity (V_f-V_s)

- In reality:
  - Radial velocity profiles, radial solids profiles, uniform solids hold-up → clusters.