UNIVERSITEIT TWENTE.

Multiphase Reactor Technology
HC 4: Fluidization

Fausto Gallucci

Sascha Kersten

Contents

- Dense fluidized beds

  - ✓ Gas distributor design
  - ✓ Gas bubbles behaviour
  - ✓ Flow models
  - ✓ Entrainment and elutriation
  - ✓ Mixing and segregation of solids

Dense Fluidized Beds
Gas distributor design

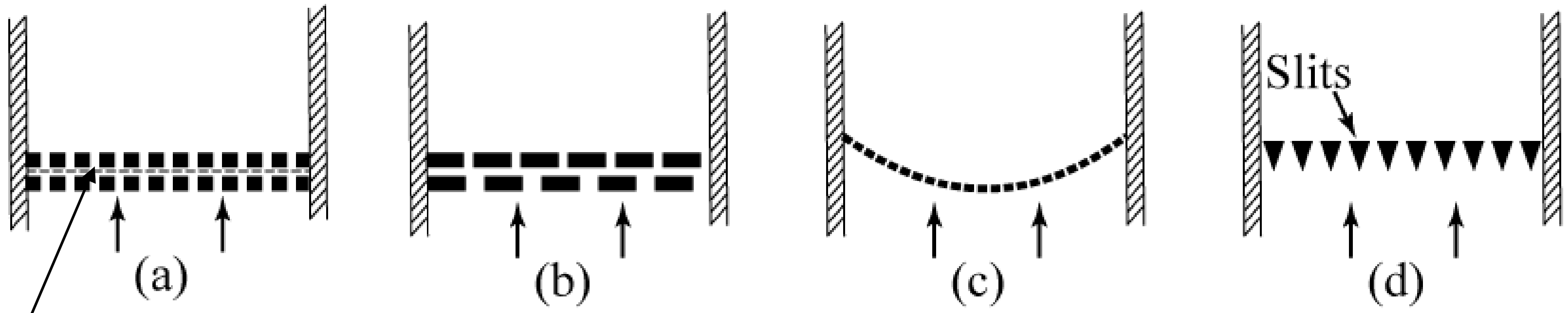
- Gas distributors are required to properly feed the gas through the bed.

Ideally they need to have the following characteristics

- ✓ Low pressure drop
- ✓ High strength
- ✓ High resistance against thermal stresses
- ✓ Resistance to clogging
- ✓ Low cost

Dense Fluidized Beds
Gas distributor design

- Plate and grate distributors



Metal screen
(in some cases not preferred)

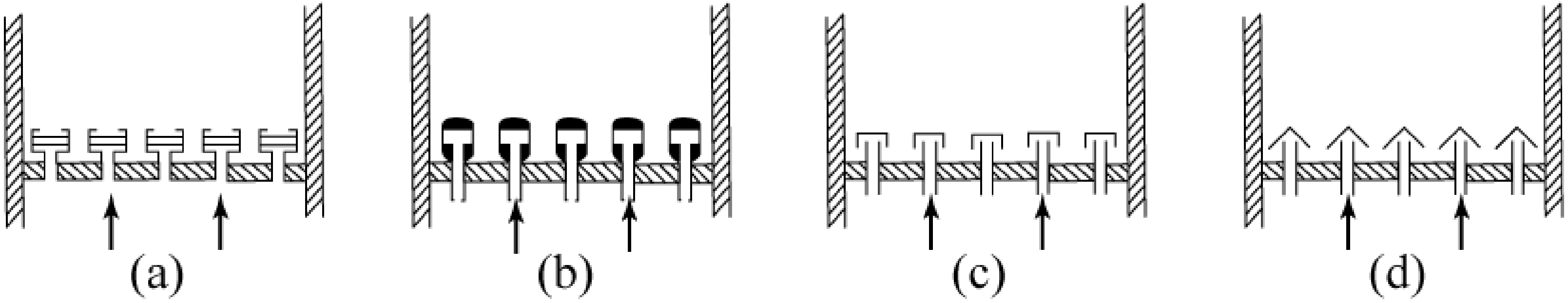
Cheap but low resistance
for heavy loads

Higher resistance
for heavy loads

Orifice diameters:
1-2 mm small scale up to 50 mm large scale

Dense Fluidized Beds
Gas distributor design

- Tuyere distributors (bottom)



Used in high temperature or highly reactive conditions

Much more expensive than perforated plates
Problems with settling and sintering of particles on the distribution plate

Dense Fluidized Beds
Gas distributor design

- Summary of experimental findings (d=distributor; b=bed)

  - Even gas distribution for u close to u_{mf}: $\Delta p_d >= 0.15 \Delta p_b$

  - Ratio of Δp_d and Δp_b decreases with increasing u/u_{mf} ratio

  - $\Delta p_d / \Delta p_b$ roughly independent of bed height

  - For similar conditions: $$\left(\frac{\Delta p_d}{\Delta p_b}\right)_{\substack{porous\\plate}} > \left(\frac{\Delta p_d}{\Delta p_b}\right)_{\substack{orifice\\plate}}$$

    difference decreases to zero for $u >> u_{mf}$

  - Practical “rule of thumb”: $\Delta p_d = (0.2 \text{ to } 0.4) \Delta p_b$

Dense Fluidized Beds
Gas distributor design

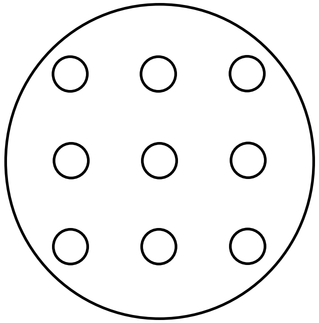
- Initial bubble diameter (cm) for distributor with N_{or} orifices per unit area (cm^{-2}) of plate where distance between orifices equals l_{or}:

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

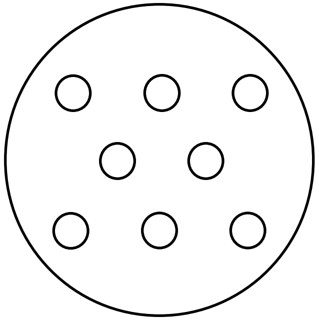
low gas flow rate
non-touching bubbles
$d_{bo}<l_{or}$

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

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



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



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

Dense Fluidized Beds
Gas bubbles behaviour

- Davidson model for isolated bubbles
  - ✓ Postulate 1: A gas bubble is solid-free and circular (2D) or spherical (3D) in shape.
  - ✓ Postulate 2: The emulsion phase flowing in the vicinity of the bubble behaves like an incompressible, inviscid fluid with macroscopic density $\rho_s(1-\varepsilon_{mf})$.
  - ✓ Postulate 3: The relative motion between the gas phase and the emulsion phase is given by Darcy's law:

emulsion phase velocity

$$\bar{u} - \bar{v} = -K\nabla p$$

gas phase velocity

bed permeability

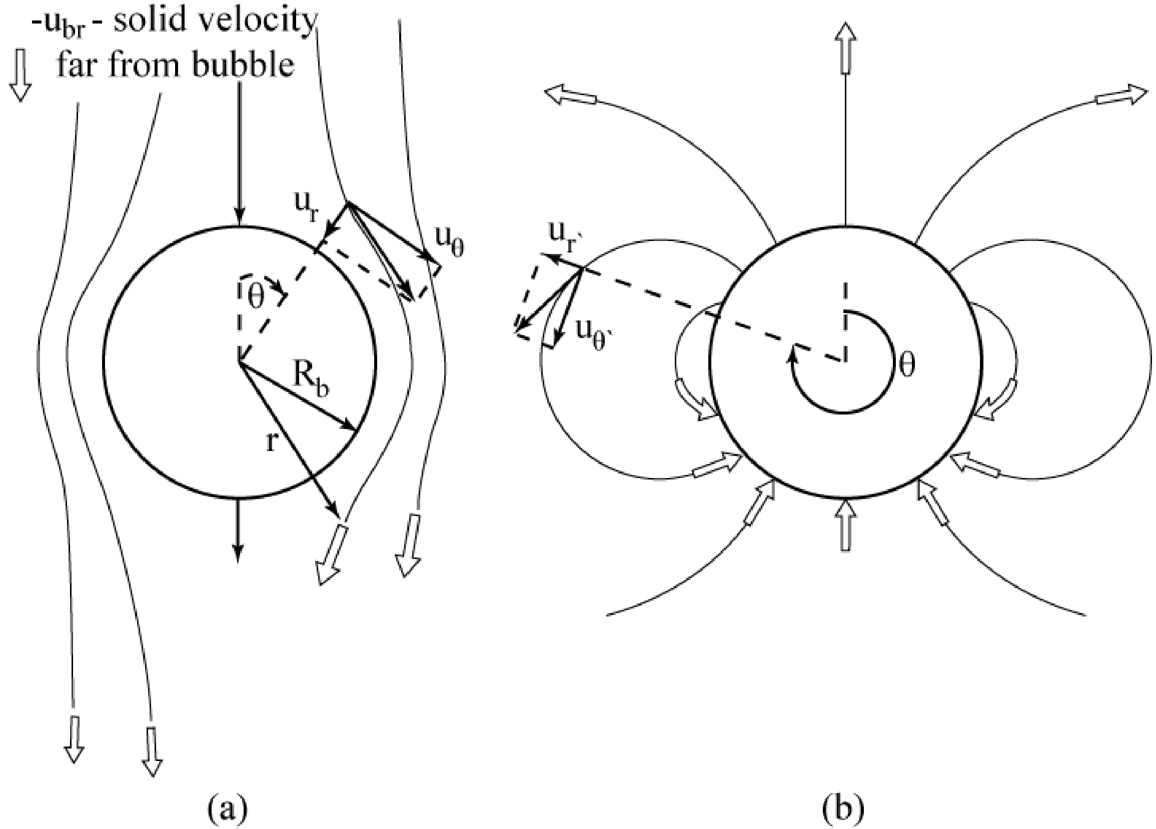
Boundary conditions

- Far from bubbles the pressure gradient in undisturbed
- The pressure in a bubble is constant

Dense Fluidized Beds
Gas bubbles behaviour

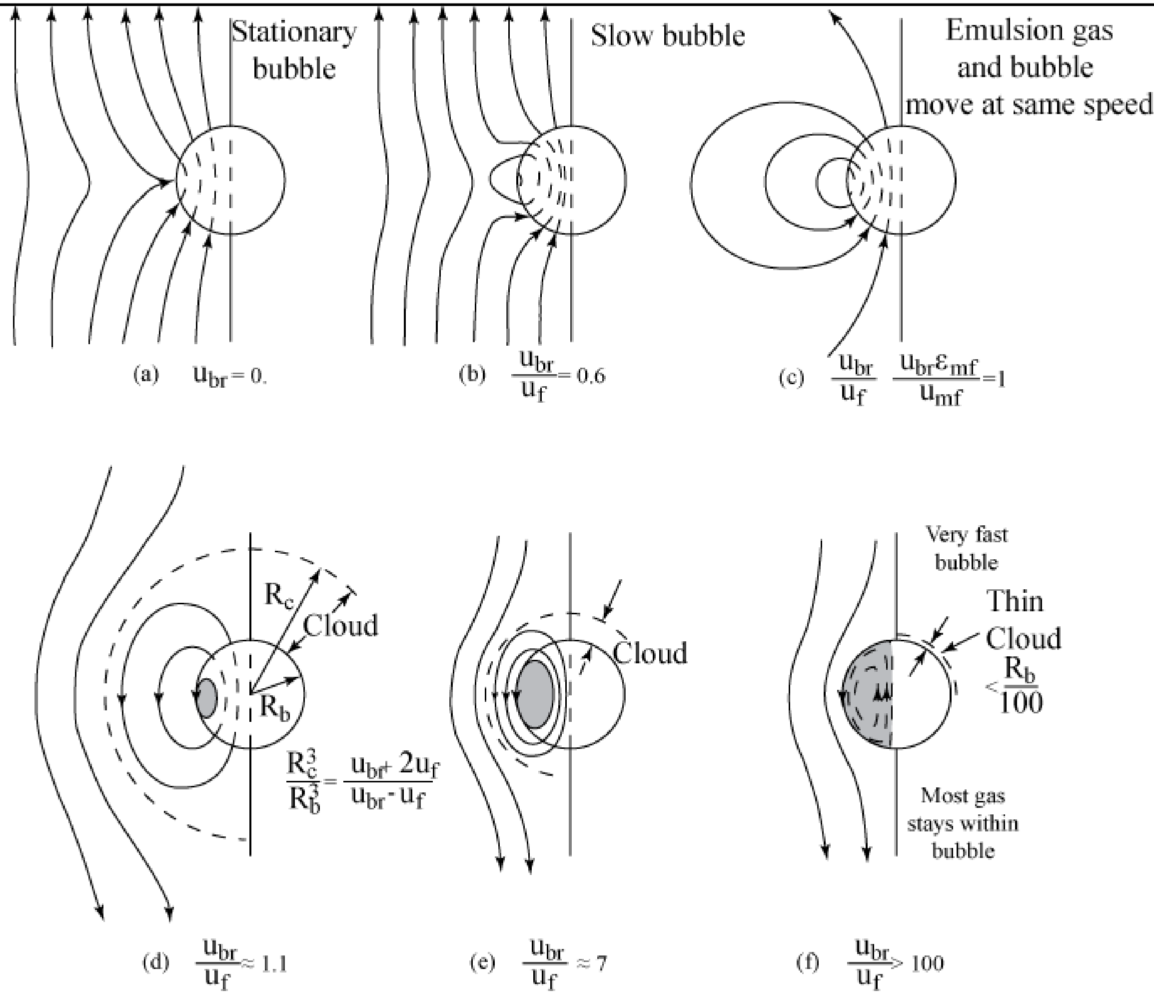
- Motion of solids in the vicinity of a rising bubble viewed by observer moving with the bubble (a) and stationary observer (b)

Velocity profile



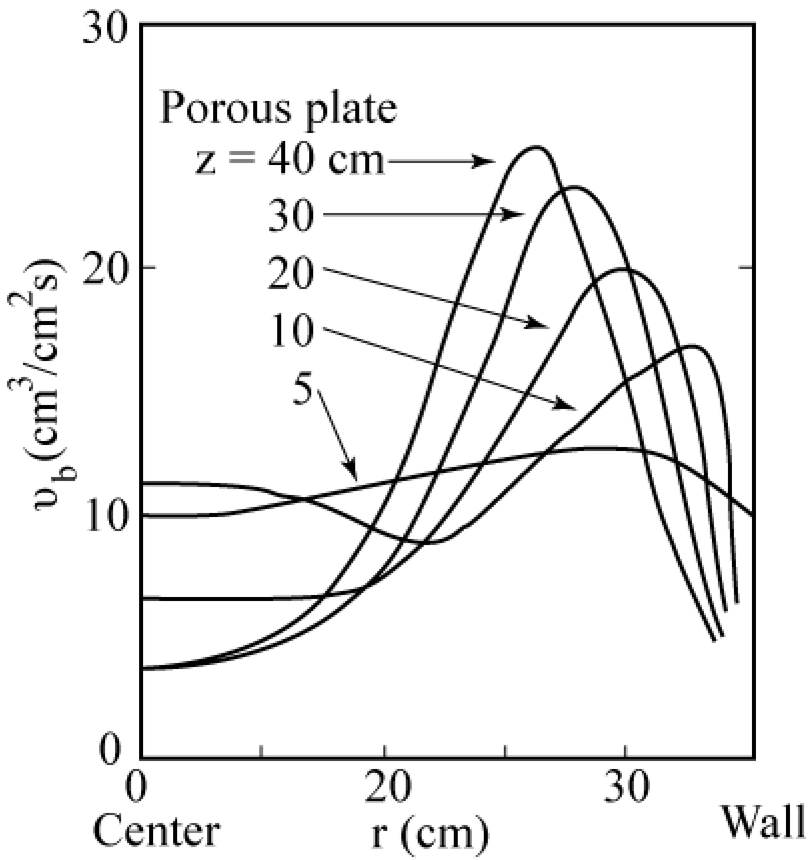
Stream function

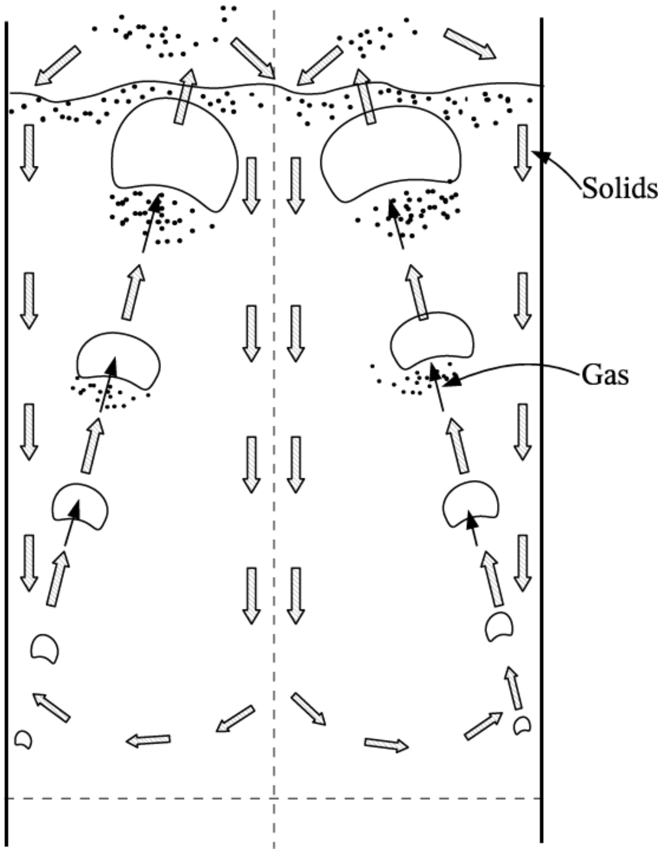
Dense Fluidized Beds
Gas bubbles behaviour



Dense Fluidized Beds
Gas bubbles behaviour

- Visible bubble flow pattern for Geldart B solids (d_t=1m, d_p=103 μm, u_{mf}=1.35 cm/s and u_0=20 cm/s) and general emulsion flow pattern





Dense Fluidized Beds
Gas bubbles 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):

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

  - 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

Dense Fluidized Beds
Gas bubbles 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 $1 <= u_{mf} <= 8$ cm/s

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

- Werther correlation can be adapted for gas distributor with orifices

(initial bubble size not zero in this case):

strategy: fit initial bubble size d_{b0} at initial height of bubble formation z_0

Dense Fluidized Beds
Gas bubbles behaviour

- Bubble rise velocity correlations

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

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

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

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

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

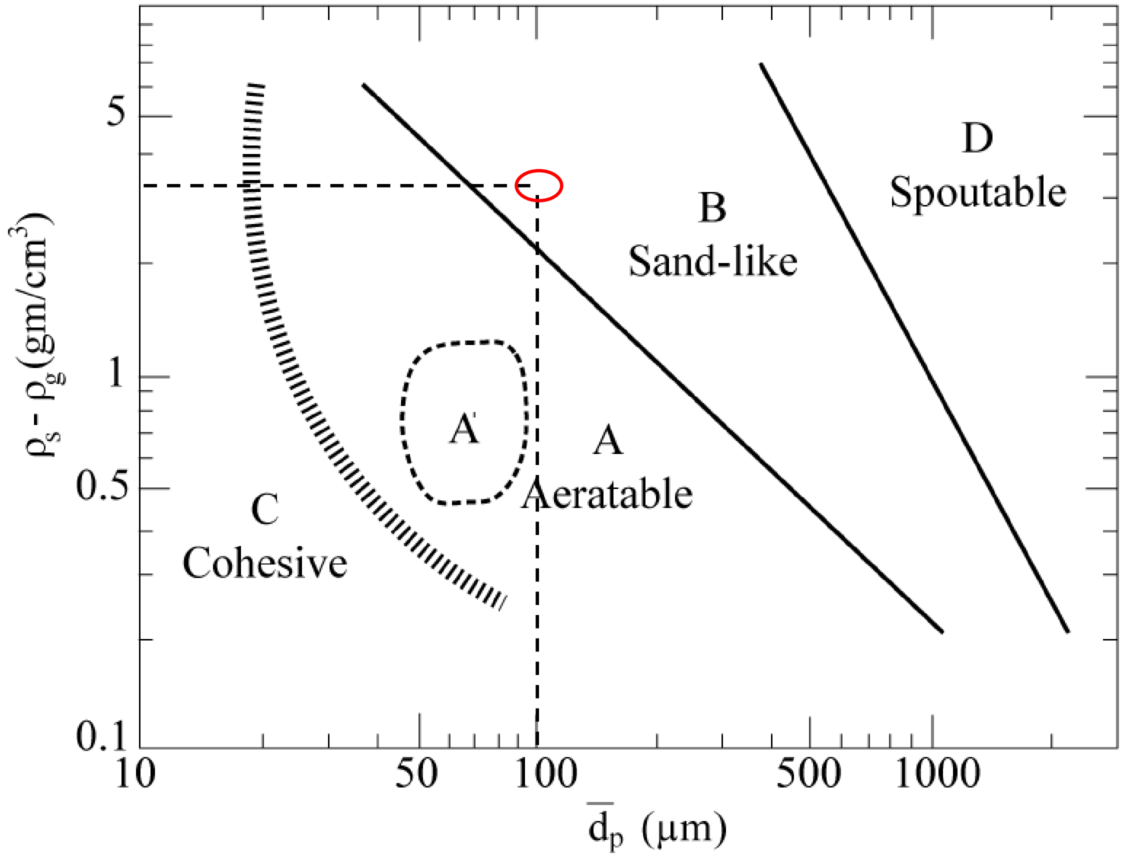
$$u_{br} = 0.711\sqrt{gd_b}$$

Exercise (10 minutes)

- Estimate bubble diameter and bubble velocity at height z = 0.5 m in a bed (d_t = 0.5m) of sand (ρ_s-ρ_g = 3 g/$cm_{3,}$ d_p=100μm, u_{mf} = 1cm/s, u_0 = 0.45 m/s) supported by a perforated plate distibutor (triangular arrangement, d_{or} = 2mm, l_{or} = 30 mm)

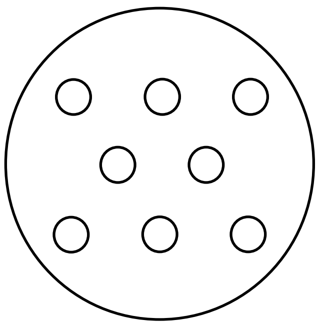
Solution

- Check which kind of particle we have (from the graph they are Geldart B)



Solution

- Calculate the bubble size.



$$N_{or} = \frac{2}{\sqrt{3} l_{or}^2} = 1.3 \cdot 10^3 m^{-2}$$

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

$$d_{bo} = \frac{2.78}{g}\left(u_0 - u_{mf}\right)^2$$
$$= \frac{2.78}{9.81}(0.45 - 0.01)^2 = 0.055m$$
$$= 5.5cm$$

Solution

- Calculate the bubble size.

$$d_{bm} = 0.65\left[\frac{\pi}{4} d_t^2 \left(u_0 - u_{mf}\right)\right]^{0.4} = 0.65\left[\frac{\pi}{4} 50^2 (45 - 1)\right]^{0.4}$$
$$= 61.3cm$$

$$\frac{d_{bm} - d_b}{d_{bm} - d_{b0}} = e^{-0.3\left(\frac{z}{d_t}\right)} = \frac{61.3 - d_b}{61.3 - 5.5}$$
$$= e^{-0.3\left(\frac{50}{50}\right)}$$
$$d_b = 20cm$$

Solution

- Calculate the bubble velocity.

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

$$u_b = 1.6\{(u_0 - u_{mf}) + 1.13 d_b^{0.5}\} d_t^{1.35} + u_{br}$$
$$u_b = 1.6\{(0.45 - 0.01) + 1.13 \cdot 0.2^{0.5}\} 0.5\text{^}1.35 + 0.711(9.81 \cdot 0.2)^{0.5}$$
$$u_b = 1.59 m/s$$

Dense Fluidized Beds
Flow models

- Necessity for flow models:

provide conceptual framework to estimate a.o. volume fractions
and phase velocities insight in gas-solids contacting
predict performance of bubbling beds in physical and chemical applications

- Many types of models have been proposed:
  - Toomey and Johnstone simple two-phase model
  - Kunii and Levenspiel model

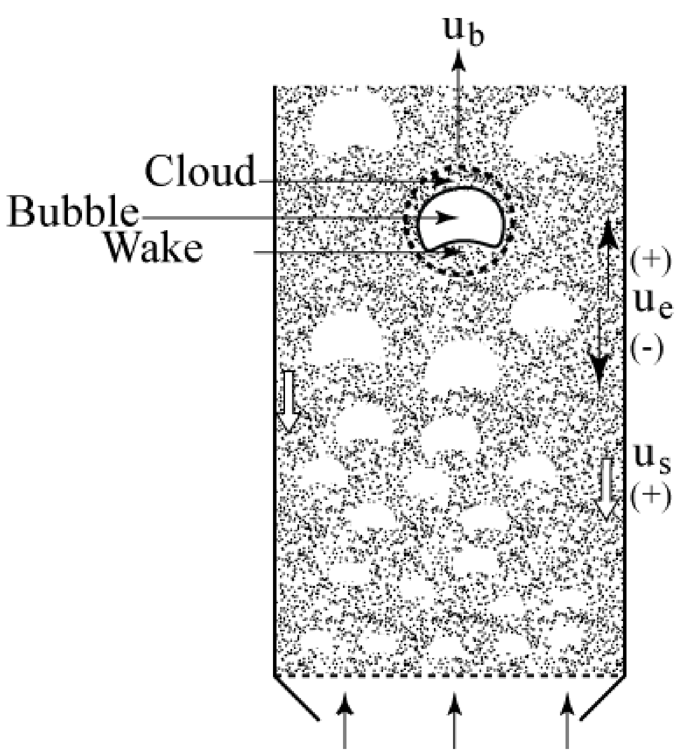
Dense Fluidized Beds
Flow models

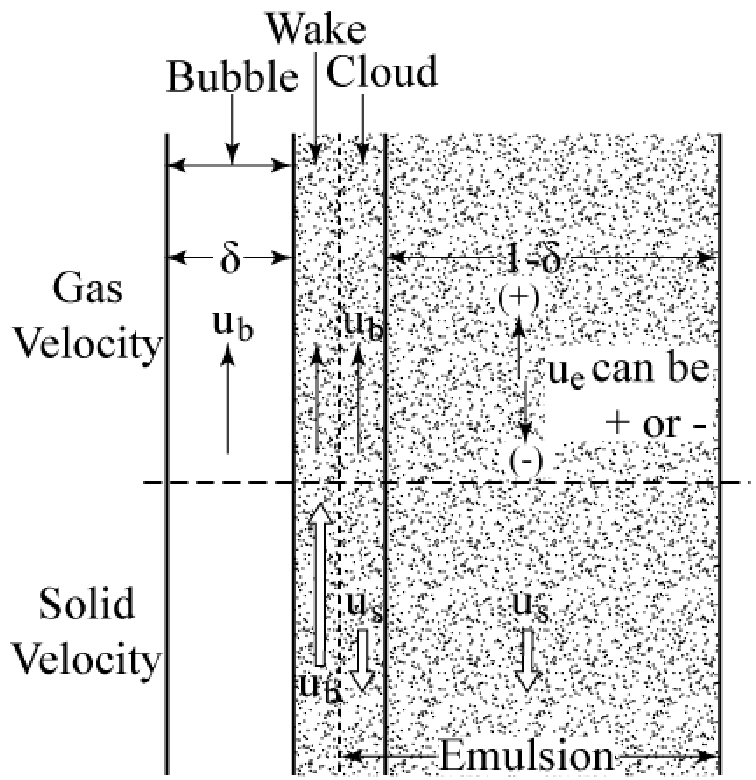
- Simple two-phase model:
  - Rise velocity of bubbles: $u_{br} = 0.711\sqrt{gd_b}$
  - Rise velocity of emulsion gas: $u_e = \dfrac{u_{mf}}{\varepsilon_{mf}}$
  - Superficial rise velocity of emulsion gas: u_{mf}
  - Rise velocity of solids: $u_s = u_{s,up} = u_{s,down} = 0$
  - Fraction of bed in bubbles: $\delta = \dfrac{u_0 - u_{mf}}{u_b - u_{mf}}$
  - Fraction of bed in emulsion: $1 - \delta = \dfrac{u_b - u_0}{u_b - u_{mf}}$

Dense Fluidized Beds
flow models

- Kunii and Levenspiel (K-L) model with Davidson bubbles and wakes

✓ Distinction between slow cloudless bubbles and fast clouded bubbles and accounts for bubble wakes





Dense Fluidized Beds
flow models

- Bubble rise velocities

✓ small laboratory beds of Geldart A and B solids and any size bed of Geldart D solids use:

$$u_b = u_0 - u_{mf} + u_{br} = u_0 - u_{mf} + 0.711\sqrt{g d_b}$$

✓ large diameter beds:

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

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

with u_{br} given by: $$u_{br} = 0.711\sqrt{g d_b}$$

Dense Fluidized Beds
Flow models

- Fraction of bubbles in the bed δ:

  - ✓ Slow bubbles without clouds, or $u_b<u_{mf}/e_{mf}$:

$$\delta = \frac{u_0 - u_{mf}}{u_b + 2u_{mf}}$$

  - ✓ Intermediate bubbles with thick clouds $u_{mf}/e_{mf} <u_b<5u_{mf}/e_{mf}$:

$$\delta = \frac{u_0 - u_{mf}}{u_b + u_{mf}} \quad \text{(lower boundary)} \qquad \delta = \frac{u_0 - u_{mf}}{u_b} \quad \text{(upper boundary)}$$

  - ✓ Fast bubbles with thin clouds, or $u_b>5u_{mf}/e_{mf}$:

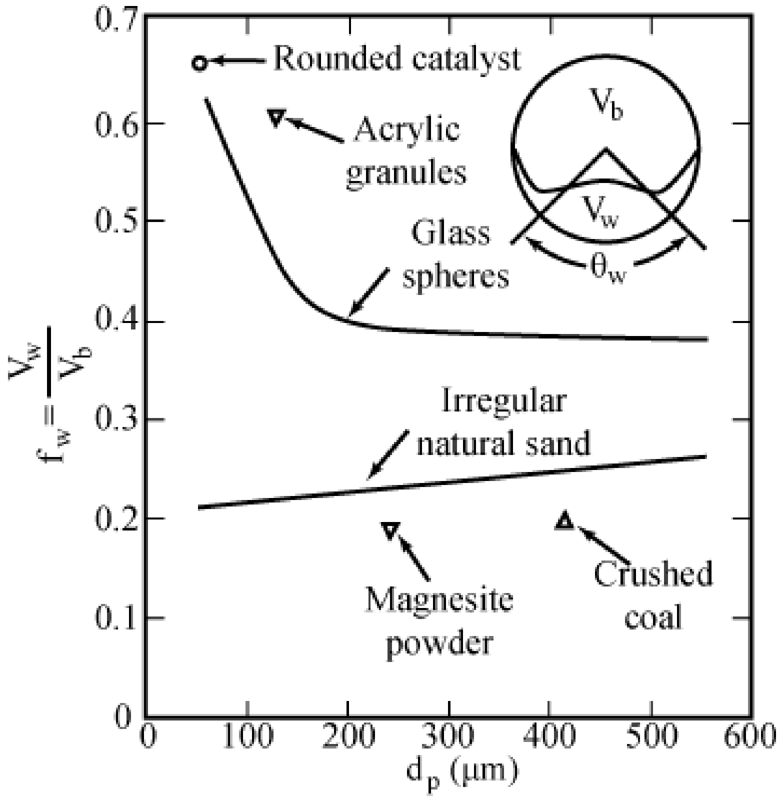
$$\delta = \frac{u_0 - u_{mf}}{u_b - u_{mf}}$$

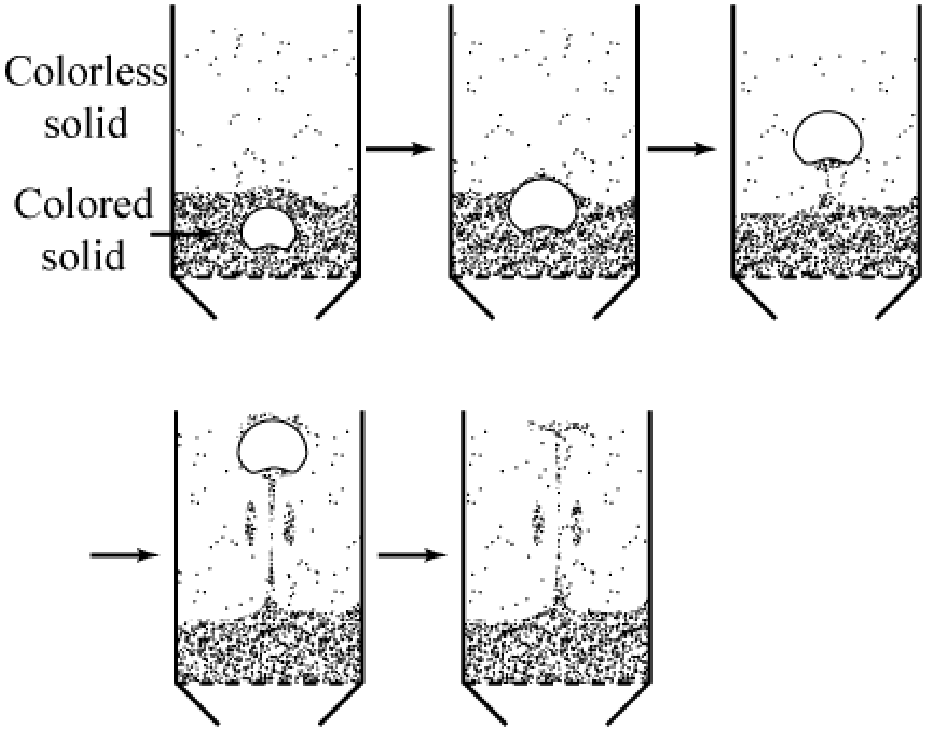
Dense Fluidized Beds
Flow models

- Cloud volume to bubble volume f_c:

$$f_c = \frac{3}{\frac{u_{br}\varepsilon_{mf}}{u_{mf}} - 1}$$

- Wake volume to bubble volume f_w:





Dense Fluidized Beds
Flow models

- Fraction of bed in emulsion:

$$f_e = 1 - \delta - f_w \delta$$

- Definition of solids distribution:

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

$$\gamma_b, \gamma_c \text{ and } \gamma_e = \frac{volume\ of\ solids\ dispersed\ in\ b, c, e}{volume\ of\ bubble}$$

Dense Fluidized Beds
Flow models

Volume of solids in emulsion divided by bubble volume γe:

$$\gamma_e = \frac{\left(1-\varepsilon_{mf}\right)(1-\delta)}{\delta} - \gamma_b - \gamma_c$$

Volume of solids in cloud (+ wake) divided by bubble volume γ_c:

$$\gamma_c = \left(1-\varepsilon_{mf}\right)(f_c + f_w)$$
$$= \left(1-\varepsilon_{mf}\right)\left[\frac{3}{\frac{u_{br}\varepsilon_{mf}}{u_{mf}} - 1} + f_w\right]$$

Dense Fluidized Beds
Flow models

- Volume of solids in bubble divided by bubble volume γ_b:

$$\gamma_b = 0.005$$ from experimental observations

- Rise velocity of wake solids $u_{s,wake}$: $$u_{s,wake} = u_b$$

- Downflow velocity of emulsion solids $u_{s,down}$: $$u_{s,down} = \frac{f_w \delta}{1 - \delta - f_w \delta} u_b$$

- Rise velocity of emulsion gas through bed u_e:

$$u_e = \frac{u_{mf}}{\varepsilon_{mf}} - u_{s,down}$$

criterion for downflow of emulsion gas

$$\frac{u_b}{u_{mf}} > \frac{1 - \delta - f_w \delta}{f_w \varepsilon_{mf} \delta}$$

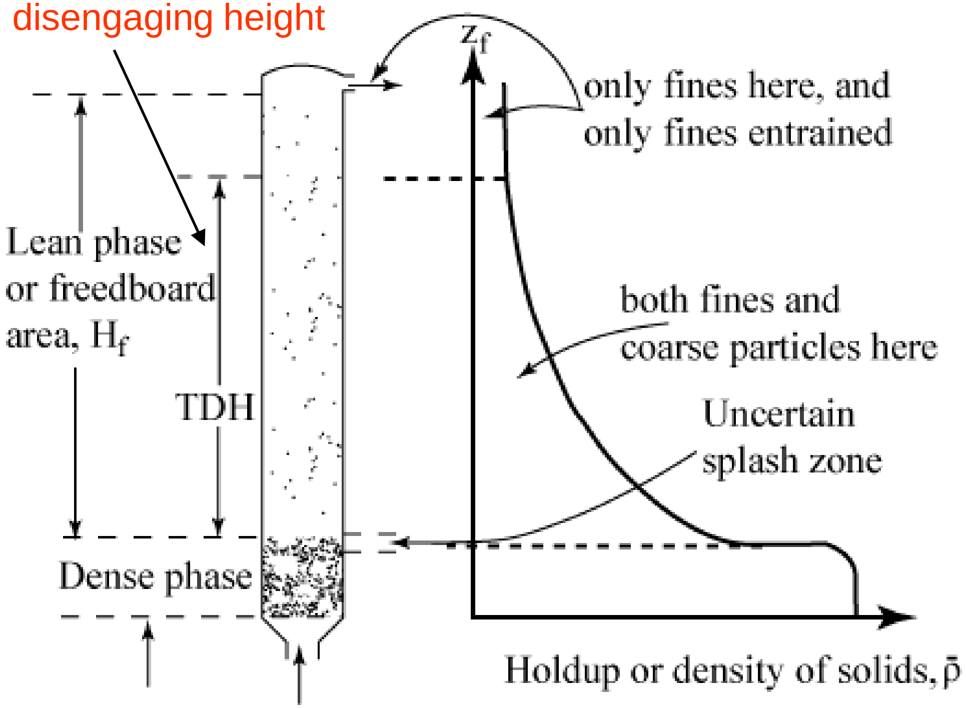
Entrainment and Elutriation

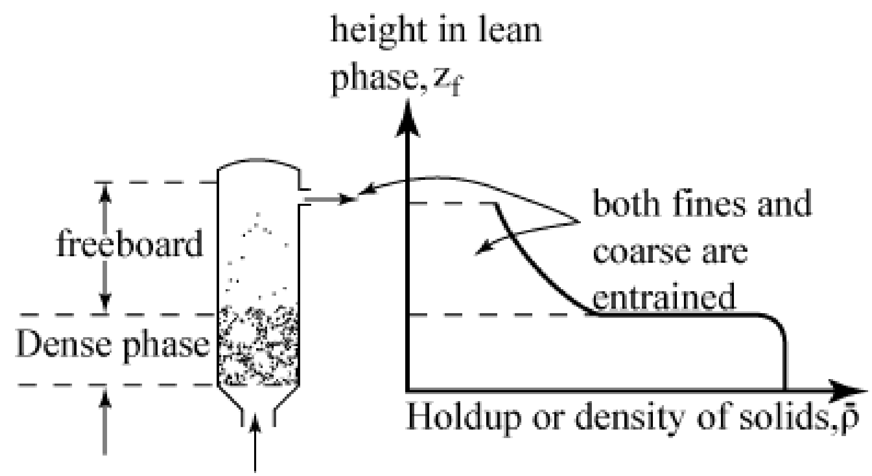
- ✓Entrainment: solids flux Gs (kg/m^2.s) carried by gas at exit of vessel

- ✓Elutriation: separation or removal of fines from mixture

Dense Fluidized Beds
Entrainment and elutriation

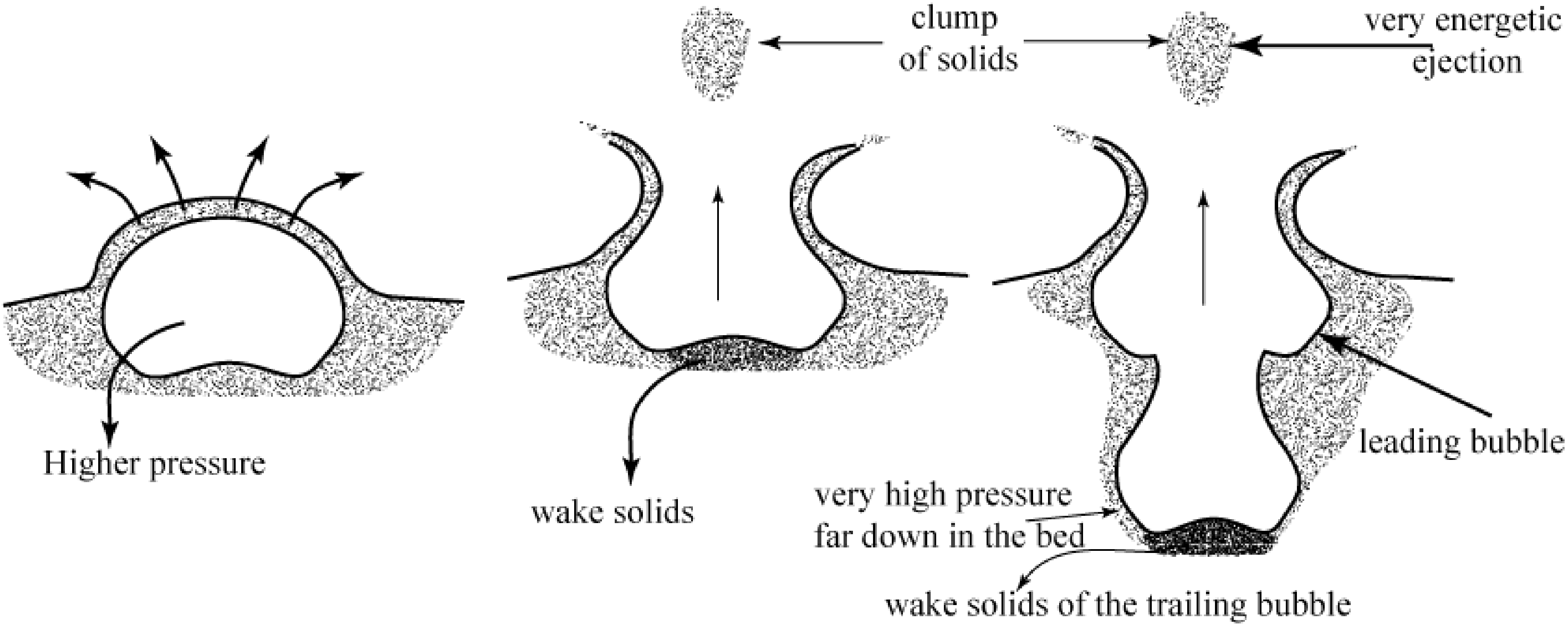
Transport disengaging height





Dense Fluidized Beds
Entrainment and elutriation

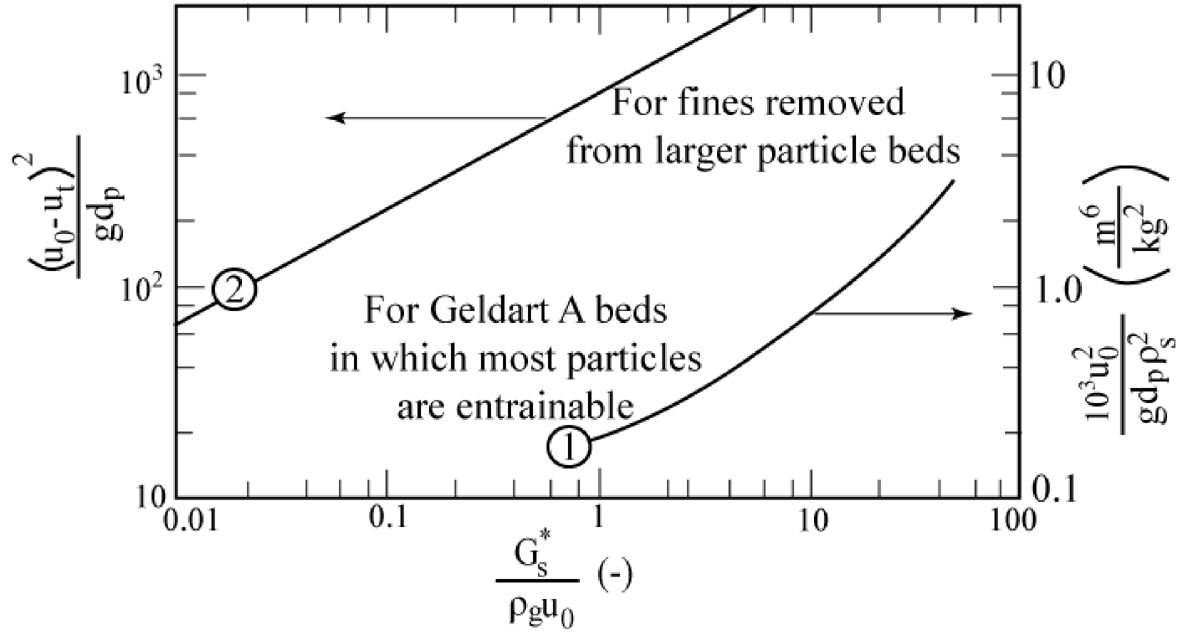
- Mechanism of solids ejection into freeboard



Dense Fluidized Beds
Entrainment and elutriation

- Procedure for calculating the entrainment flux G_s (kg/m^2.s) from tall vessels (H_f>TDH)

  - Divide PSD in narrow size interval and find size intervals with: $u_t<u_o$
  - Find $G^*_{s,i}$ from figure below for relevant system



Dense Fluidized Beds
Entrainment and elutriation

- Total entrainment flux G_s (kg/m^2.s):

$$G_s = \sum_{\substack{elutriable \\ intervals}} x_i G^*_{s,i} \qquad\qquad G_s = \int_{\substack{elutriable \\ intervals}} G^*_s(d_p) p d(d_p)$$

discrete PSD   continuous PSD

x_i: mass fraction for size interval i

- Entrainment flux G_s (kg/m^2.s) from short vessels (H_f< TDH)

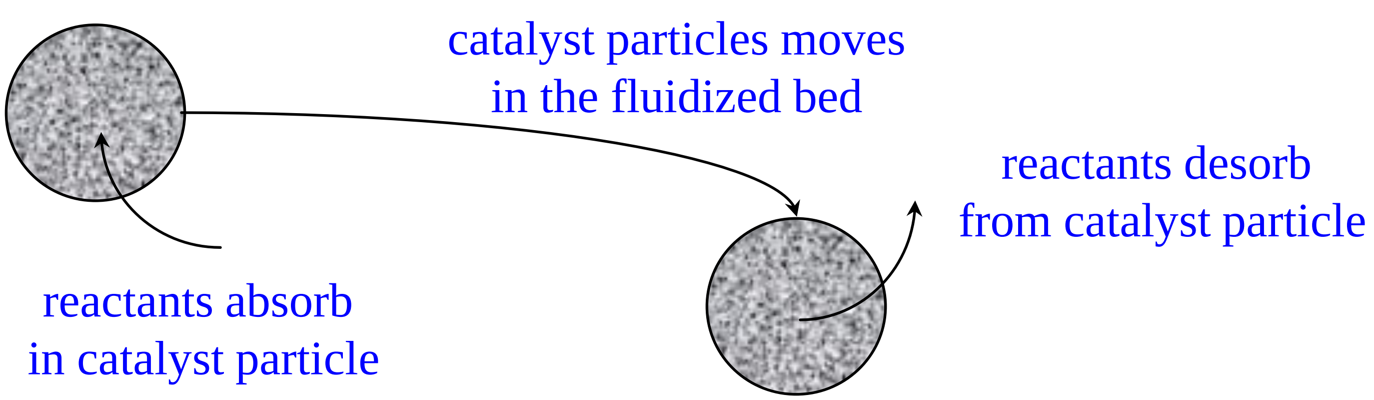
Kunii and Levenspiel freeboard-entrainment model
(domain of Fast Fluidization)

Dense Fluidized Beds
Mixing and segregation of solids

- Topics covered:

  - ✓ Vertical mixing and segregation of solids
  - ✓ Horizontal mixing and dispersion of solids
  - ✓ Mixing-segregation equilibrium

- Relevance of solid movement/mixing for catalytic reactors

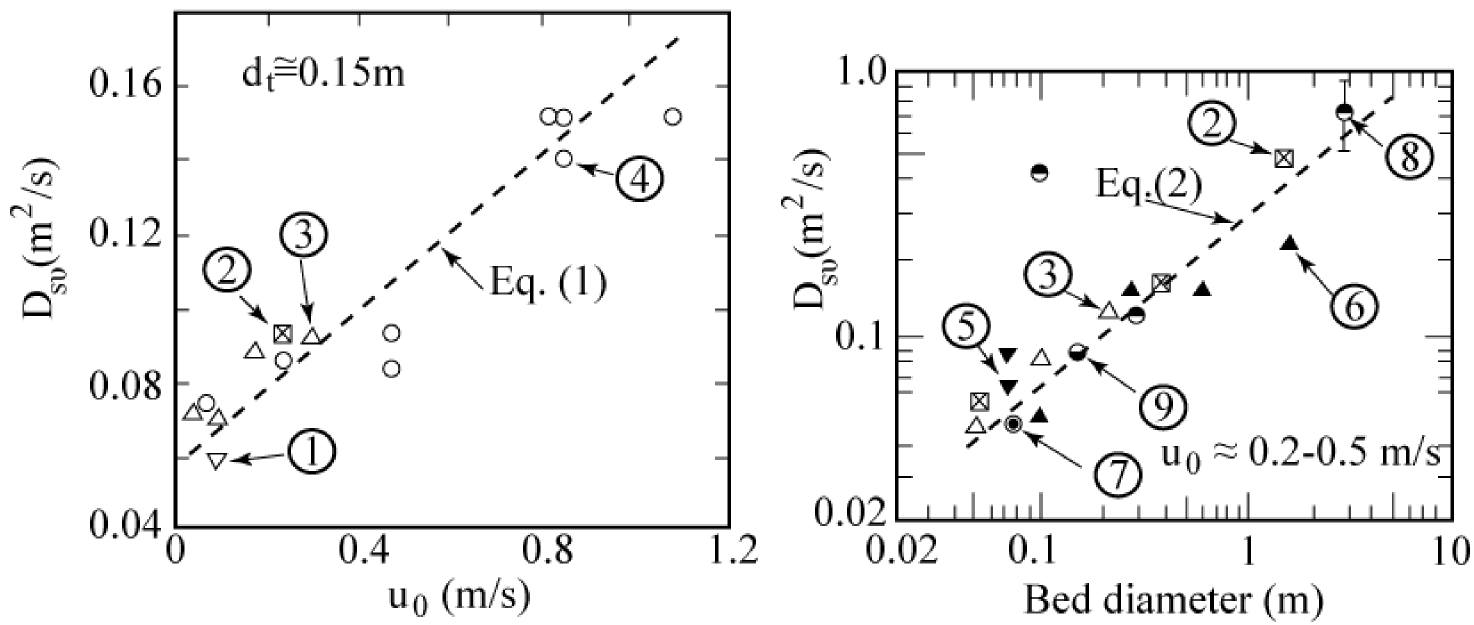


Dense Fluidized Beds
Mixing and segregation of solids

- Vertical solids mixing: often reasonably well described by the (axial) dispersion model:

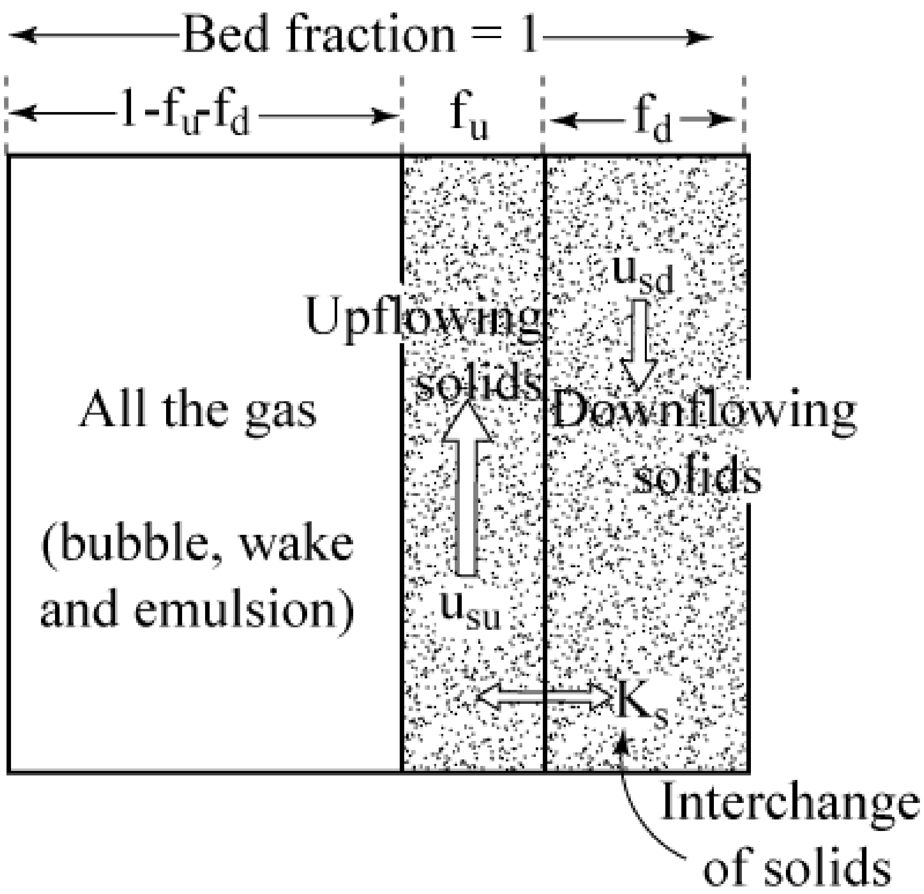
$$\frac{\partial}{\partial t} C_s = D_{sv} \frac{\partial^2}{\partial x^2} C_s$$

- Experimental results for axial dispersion coefficient for solids D_{sv}:



Dense Fluidized Beds
Mixing and segregation of solids

- Counterflow solid circulation model (van Deemter):



Dense Fluidized Beds
Mixing and segregation of solids

- Counterflow solid circulation model (van Deemter):

  - Mass balance downflowing solids (fraction f_d):

$$f_d \frac{\partial}{\partial t} C_{sd} + f_d u_{sd} \frac{\partial}{\partial z} C_{sd} + K_s (C_{sd} - C_{su}) = 0$$

  - Mass balance upflowing solids (fraction f_u):

$$f_u \frac{\partial}{\partial t} C_{su} + f_u u_{su} \frac{\partial}{\partial z} C_{su} + K_s (C_{su} - C_{sd}) = 0$$

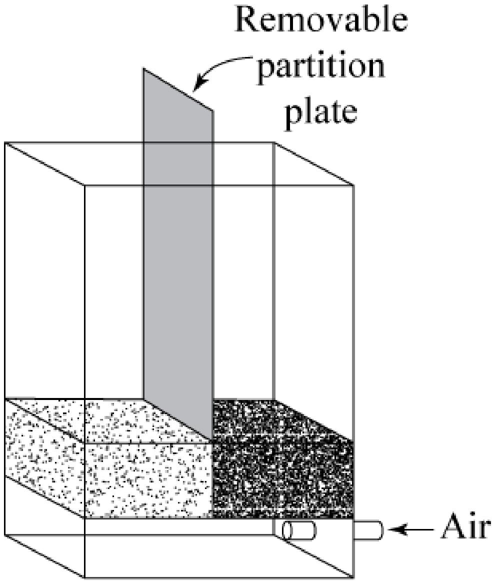
  - Effective dispersion coefficient for tall beds of fine particles:

$$D_{sv} = \frac{f_d^2 u_{sd}^2}{K_s (f_d + f_u)} = \frac{f_d^2 u_{sd}^2}{K_s \left(1 - \delta\right)\left(1 - \varepsilon_f\right)}$$

Dense Fluidized Beds
Mixing and segregation of solids

- Horizontal mixing and dispersion of solids

  ✓ Experimental setup for horizontal solids dispersion coefficient D_{sh}



- General finding: vertical mixing rate considerably faster than horizontal mixing rate

  D_{sv}(0.1-0.4 m^2/s) - D_{sh}(10-30 cm^2/s)

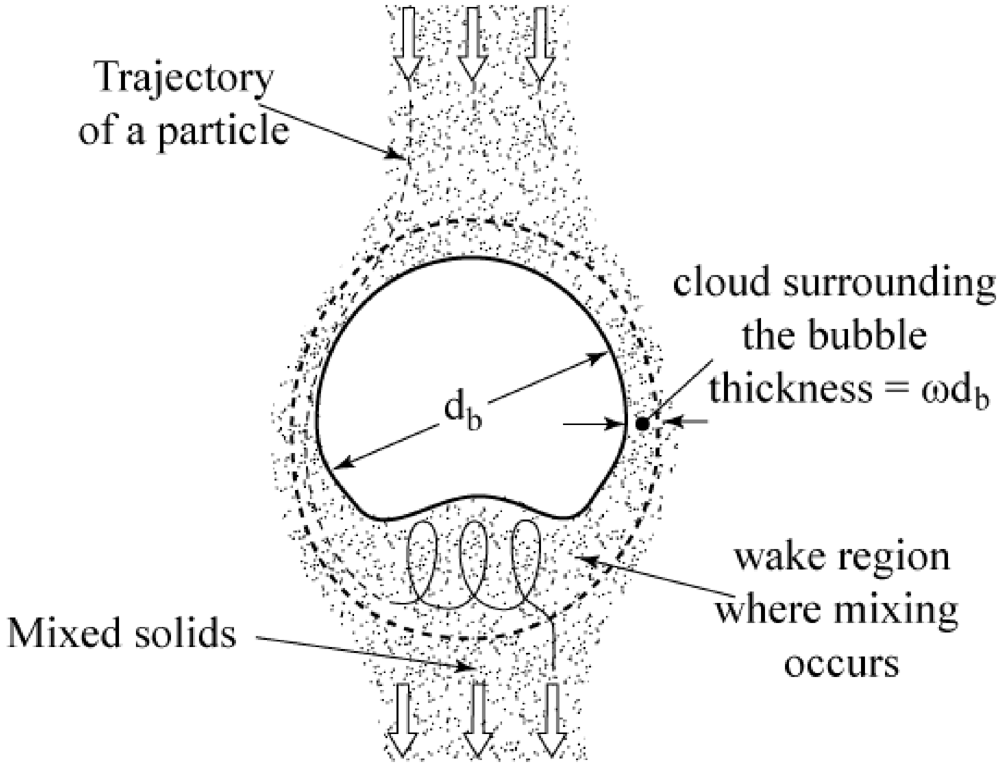
Dense Fluidized Beds
Mixing and segregation of solids

- Mechanistic model for D_{sh} based on the Davidson bubble

  - Particles moving in the cloud region are caught in the bubble wake

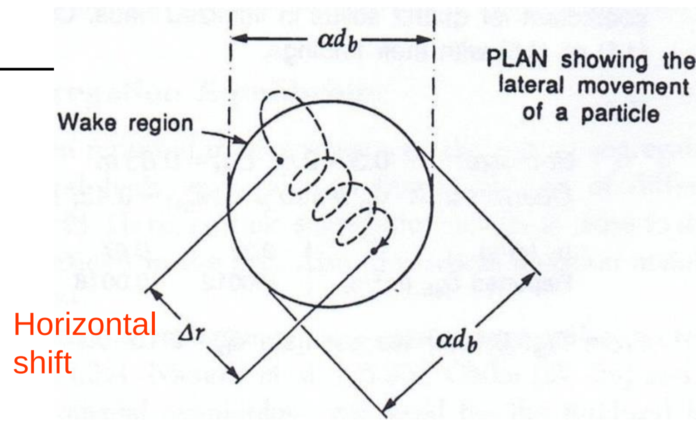
  - Uniform mixing of these particles in the bubble wake

  - Finally particles leave bubble wake at random positions



Dense Fluidized Beds
Mixing and segregation of solids

- Probability theory: $$\overline{\Delta r^2} = \frac{(\alpha d_b)^2}{4}$$



αd_b represents the effective wake diameter

- Final expression for D_{sh} for both fast and intermediate bubbles:

$$D_{sh} = \frac{3}{16}\frac{\delta}{1-\delta}\frac{\alpha^2 u_{mf} d_b}{\varepsilon_{mf}}$$

- Comparison with experimental data:

✓ For Geldart A and AB solids: good fit of experiments with $\alpha=1$

✓ For Geldart BD solids: good fit of experiments with $\alpha=0.77$

Dense Fluidized Beds
Mixing and segregation of solids

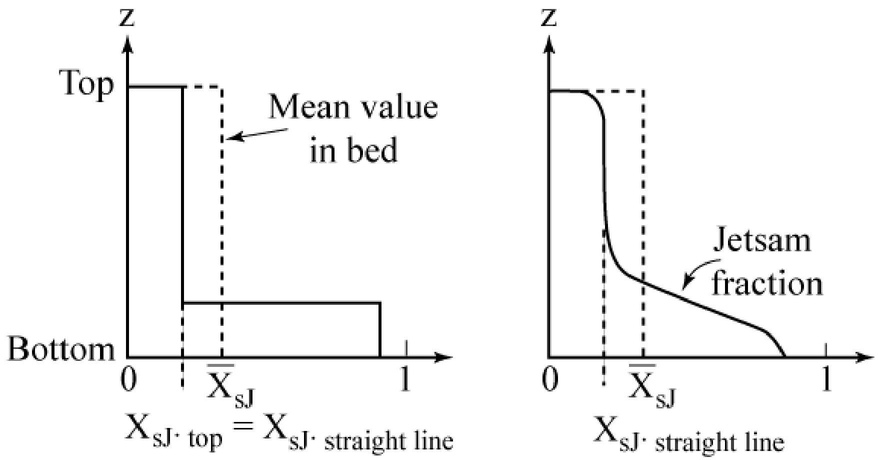
- Mixing-segregation equilibrium
  - Relevant for many industrial applications involving (large) size and/or density differences in the fluidized bed
  - The large and/or more dense particles tend to reside at the bottom of the fluidized bed (especially near umf of these large/more dense particles)
- Terminology:
  - Jetsam: component that ultimately sinks
  - Flotsam: component that floats to the top of the bed

Dense Fluidized Beds
Mixing and segregation of solids

- Definition of solids mixing index M (ratio of jetsam in top portion of the bed and (average) jetsam fraction in well-mixed bed):

$$M = \frac{X_{sJ,top}}{\bar{X}_{sJ}}$$

- Solids distribution in a binary mixture:
  - left: perfect segregation (at low u)
  - right: segregation under vigorous bubbling conditions (at high u)

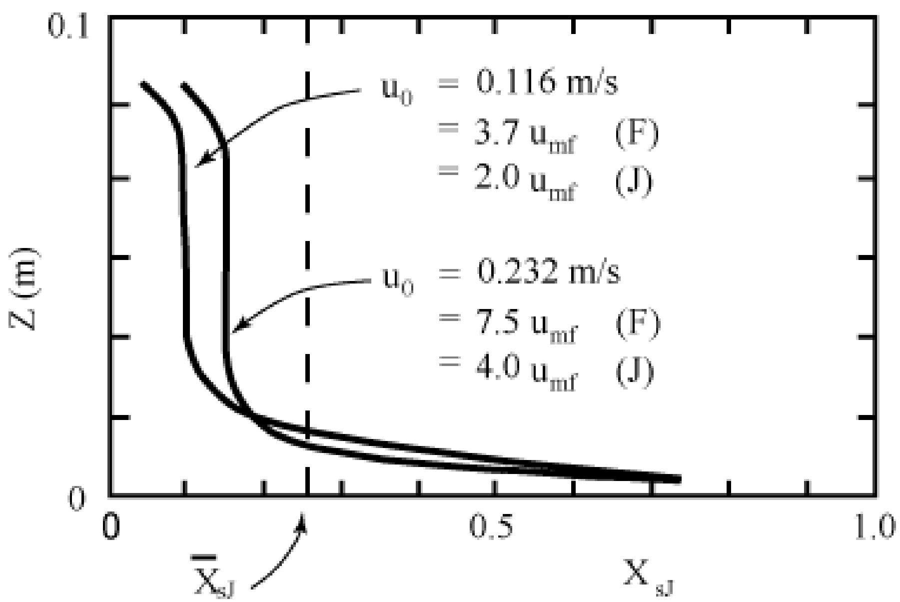


Dense Fluidized Beds
Mixing and segregation of solids

- Vertical segregation of commercial solids (d_t=0.141 m, L_m=0.10-0.15 m). left: density difference; right: size difference

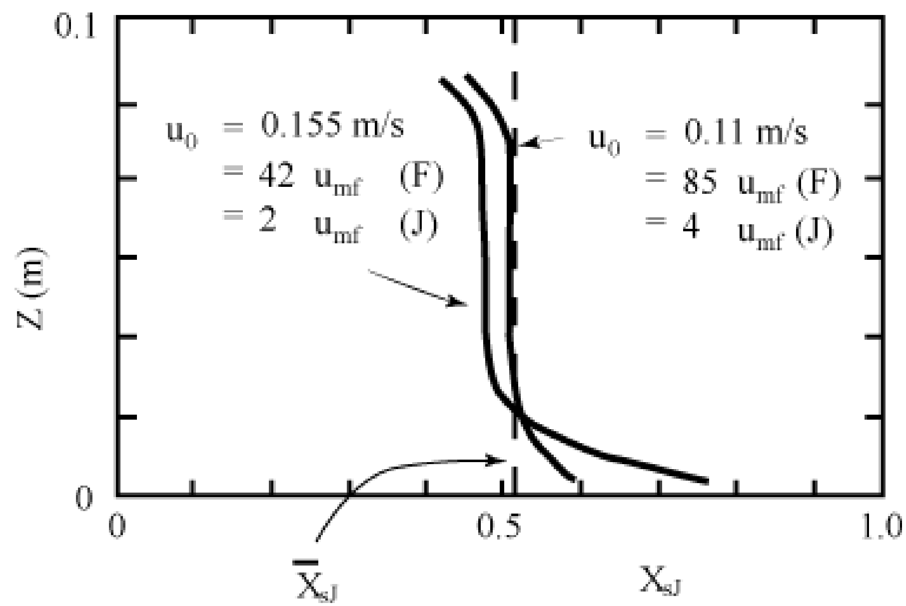
Persi washing powder base 237 μm, 1020 kg/m^3

Sodium perborate 261 μm, 1790 kg/m^3



Cracking catalyst 60.4 μm, 2198 kg/m^3

Alumina catalyst 210 μm, 2100 kg/m3



Resume

- Different types of distributors
- Distributor determines the initial bubble size
- Bubble velocity/dimensions change depending on the position and on the characteristics of solids
- Bubbles have wakes which are responsible of solid movements/mixing
- Entrainment of solids should be taken into account when designing FB
- Mixing and segregation of solids can be important for chemical reactions in FB