JMBC course Particle Technology 2015

Fluidization

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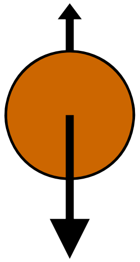
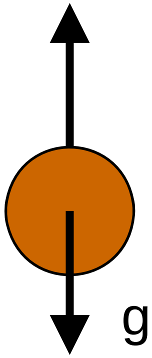
Intro gas-solids fluidized bed

Packed bed:
particles are
stagnant

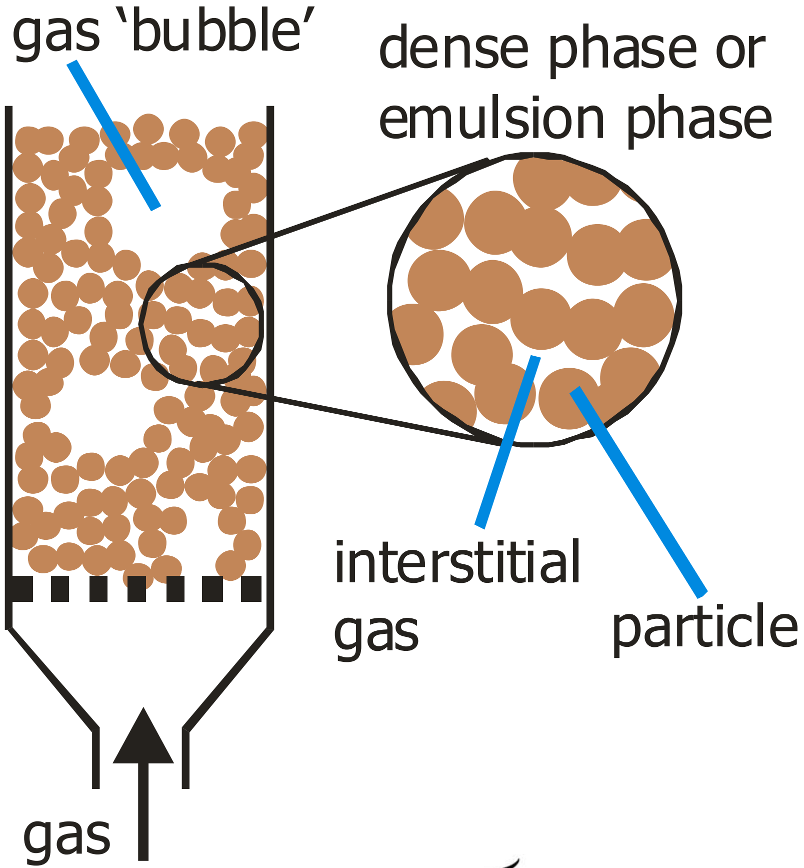


drag force
smaller than
gravitational
force

Fluidized bed:
particles
suspended in
an upward
gas stream;
they move



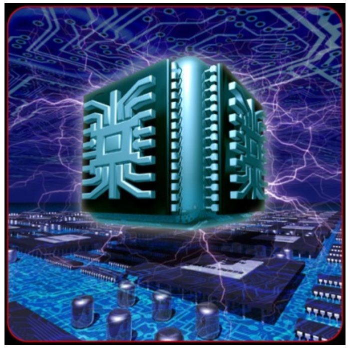
drag force
equals
gravitational
force

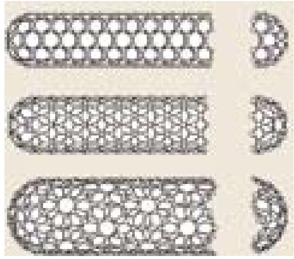


Significance of Fluidized beds

Advanced materials

- Silicon production for semiconductor and solar industry
- Coated nanoparticles
- Nano carbon tubes





Chemical and Petrochemical

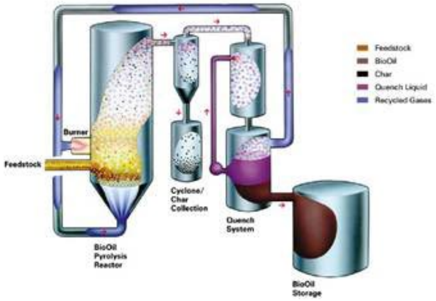
- Cracking of hydrocarbons
- Gas phase polymeric reactions

Physical operations

- Coating of metal and glass objects
- Drying of solids
- Roasting of food
- Classify particles

Combustion / pyrolysis

- Combustion/gasification of coal
- Pyrolysis of wood waste
- Chemical looping comubstion



Pharmaceutical

- Coating of pills
- Granulation
- Production of plant and animal cells

http://www.chemsoc.org/timeline/pages/1961.html

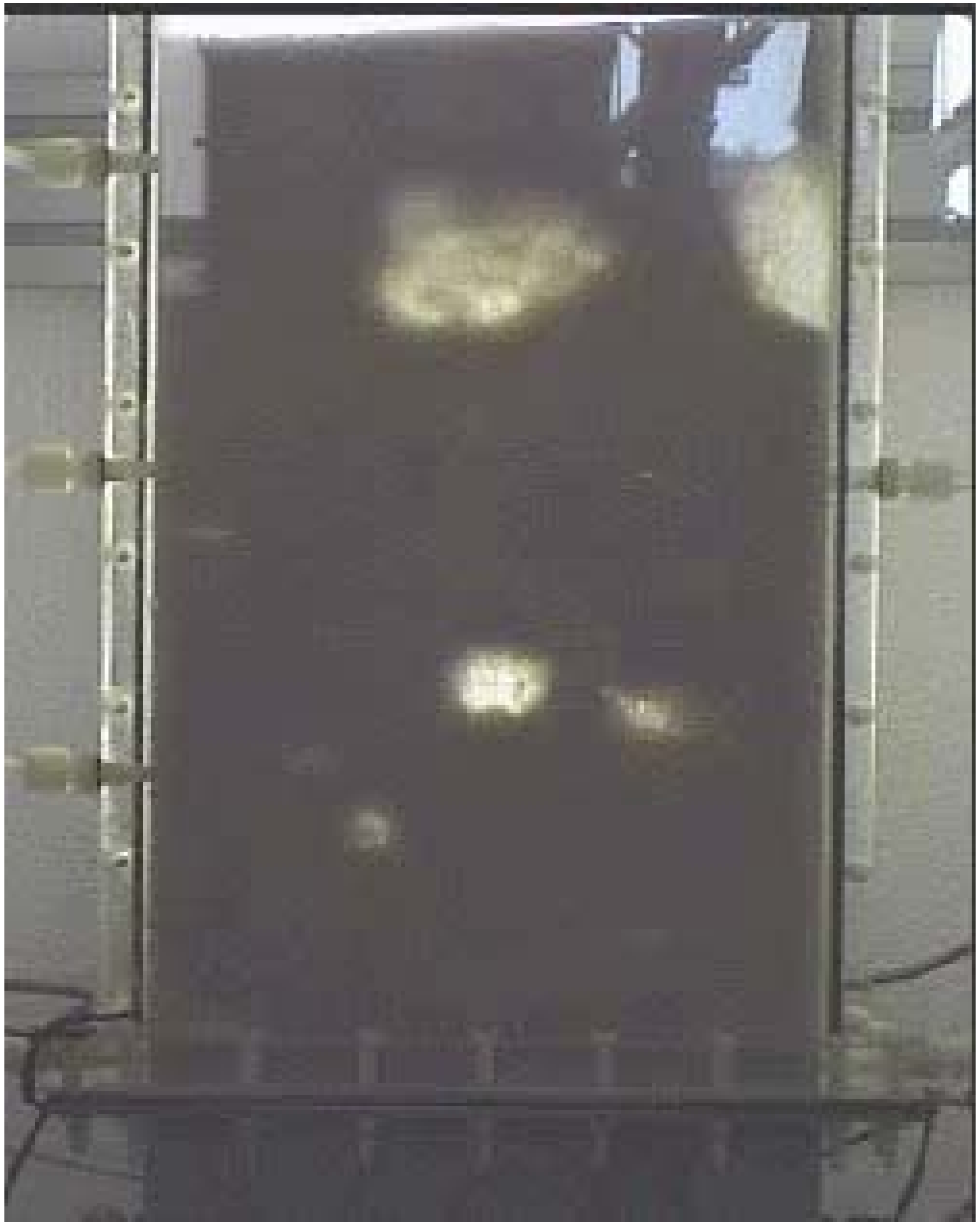
http://physicsweb.org/article/world/11/1/9

www.unb.ca/che/che5134/ fluidization.html

http://www.niroinc.com/html/drying/fdfluidtype.html

http://www.dynamotive.com/biooil/technology.html

Gas-Solid Fluidized Bed



Characteristics of Gas Fluidized Beds

- Primary Characteristics:
  - Bed behaves like liquid of the same bulk density – can add or remove particles.
  - Rapid particle motion – good solids mixing.
  - Very large surface area available

  What is the surface area of 1 m^3 of 100 μm particles?

(Dis)advantages of fluid beds

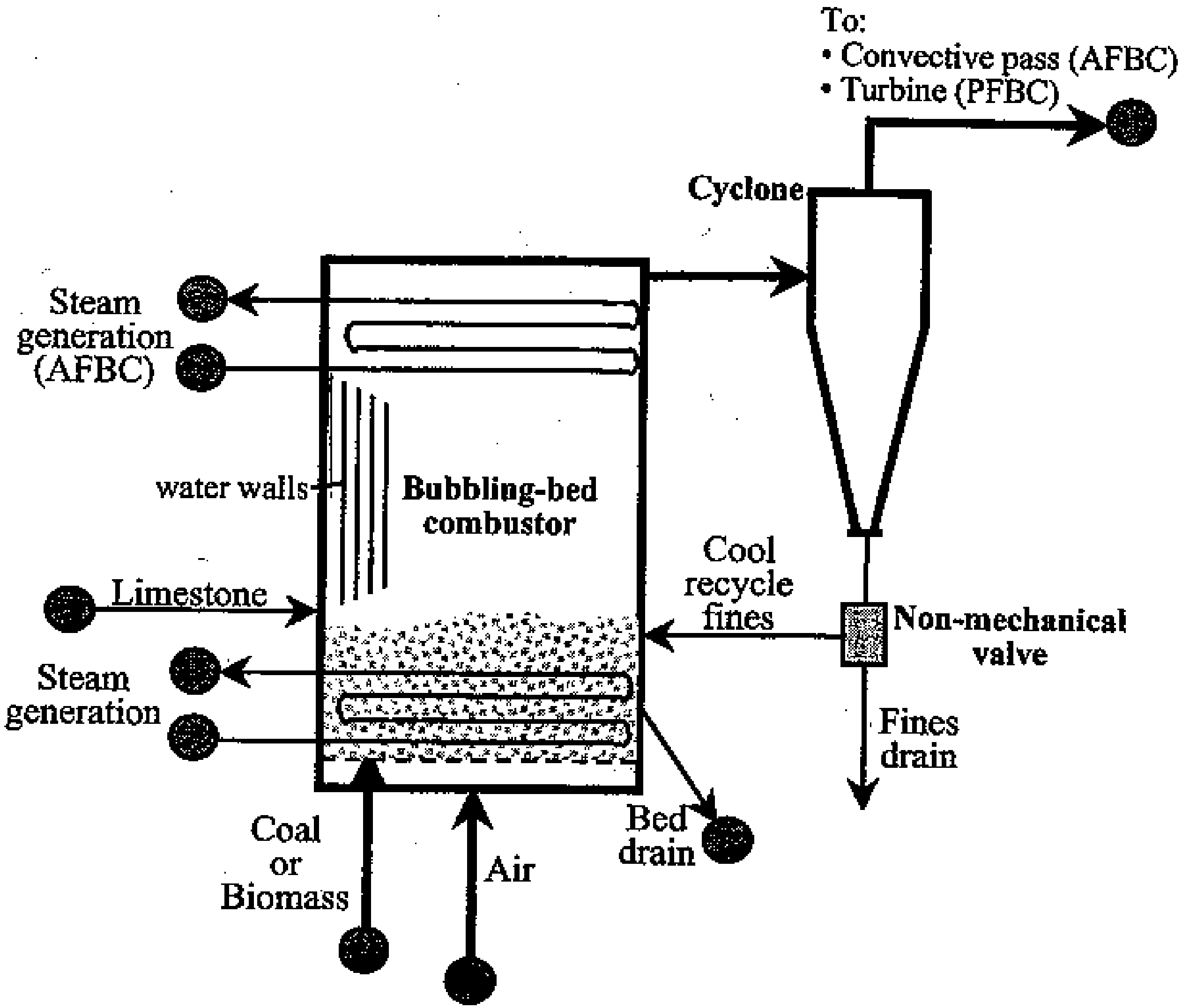
Advantages

- good G-S mass transfer in dense phase
- good heat transfer
- easy solids handling
- low pressure drop

Disadvantages

- bypass of gas in bubbles
- broad RTD gas and solids
- erosion of internals
- attrition of solids
- difficult scale-up

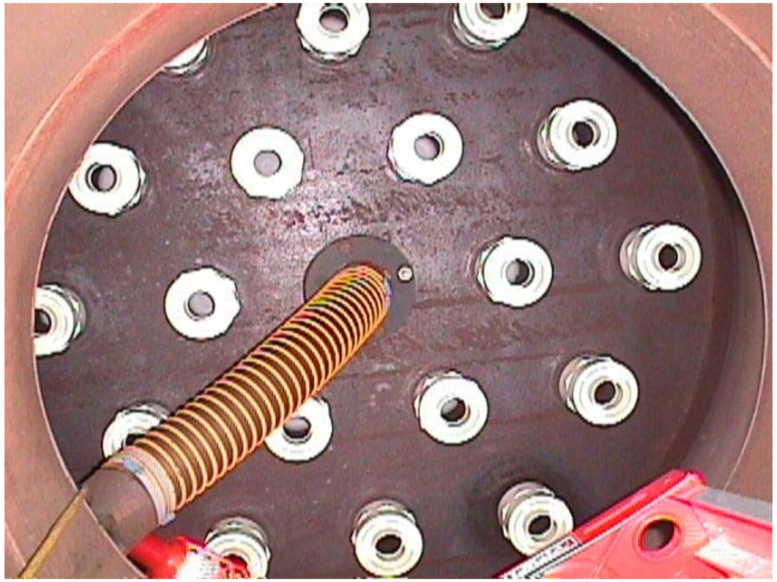
Basic Components



Yang W. Bubbling fluidized beds (Chapter 3). In: *Handbook of Fluidization and Fluid-Particle Systems. Yang W (Ed.). Marcel* Dekker, Inc., NY, NY, USA, 53–113 (2003).

1.56 m Diameter Column

Industrial Scale



Solid offtake

A Fluid Catalytic Cracking Unit. Photo courtesy of Grace Davison.

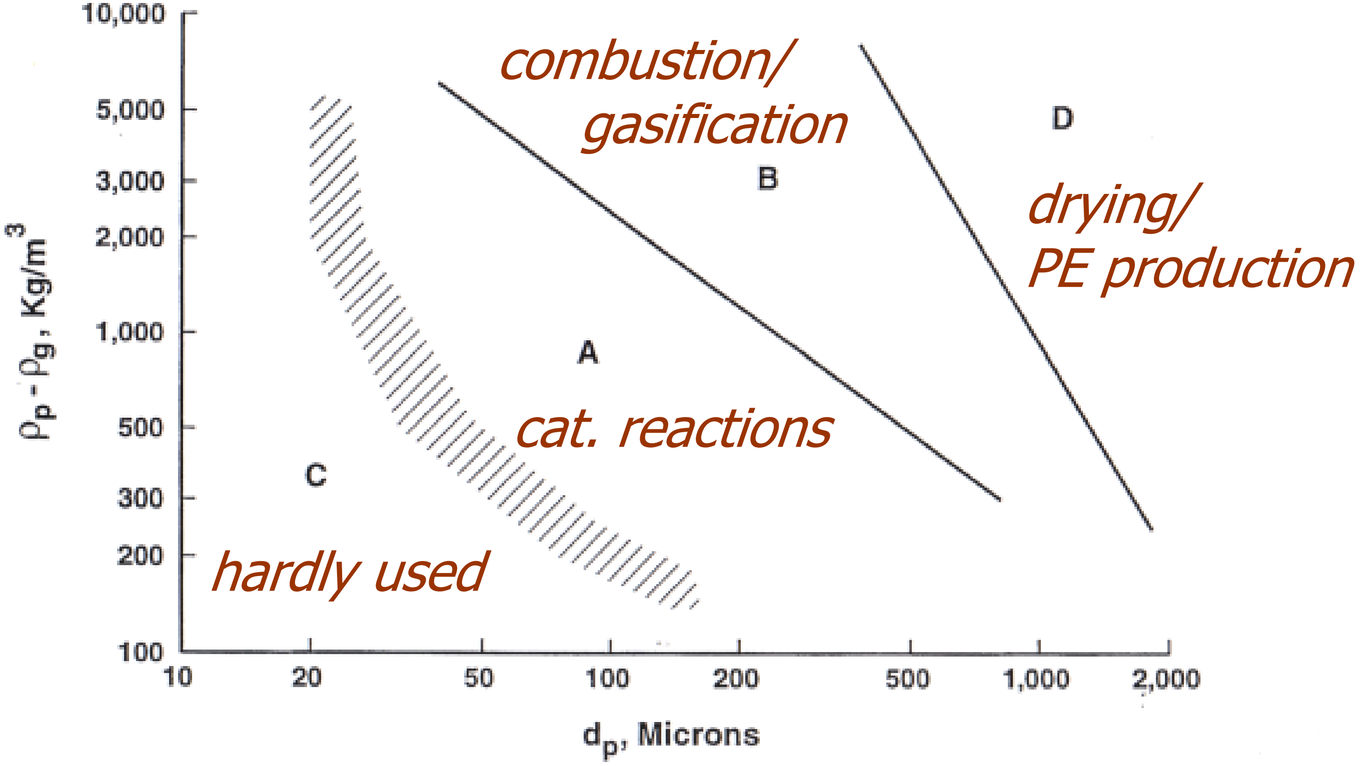
Approaches to the Study of Particulate Systems

- Totally empirical (leading to dimensional correlations)
- Empirical guided by scientific principles (e.g. Buckingham Pi Theorem to obtain dimensionally consistent correlations)
- Semi-empirical, i.e. some mechanistic basis, but with one or more empirical constants
- Mechanistic physical model without any empiricism, numerical solutions of governing equations of motion and transport

Geldart’s powder classification

(Geldart, D. Powder Technology, 1, 285, 1973)

Applies at Ambient Conditions



A: Aeratable ($U_{mb} > U_{mf}$) Material Has a Significant Deaeration Time *(FCC Catalyst)*

B: Bubbles Above U_{mf} ($U_{mb} = U_{mf}$) *(500-micron Sand)*

C: Cohesive *(Flour, Fly Ash)*

D: Spoutable *(Wheat, 2000-micron Polyethylene Pellets)*

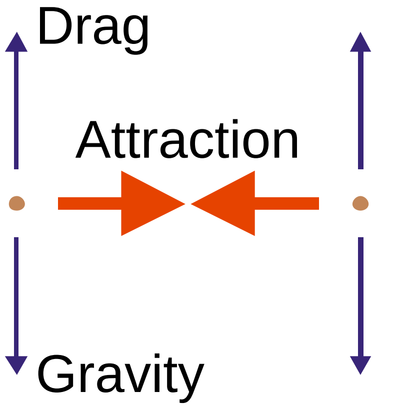
Geldart's powder classification

C

Cohesive

0-30 μm

flour

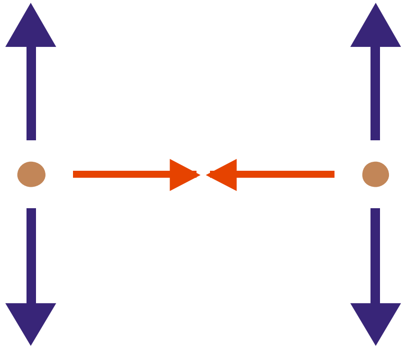


A

Aeratable

30-100 μm

milk powder

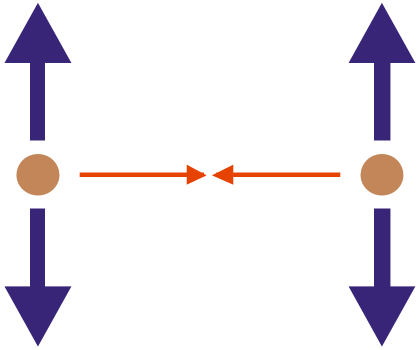


B

Bubbling

100-1000 μm

sand

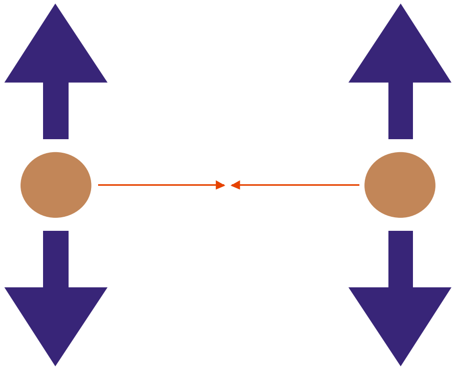


D

Spoutable

>1000 μm

coffee beans



Group C

- Cohesive
- Difficult to fluidized, and channeling occurs
- Interparticle forces greatly affect the fluidization behaviour of these powders
- Mechanical powder compaction, prior to fluidization, greatly affected the fluidization behaviour of the powder, even after the powder had been fully fluidized for a while
- Saturating the fluidization air with humidity reduced the formation of agglomerates and greatly improved the fluidization quality. The water molecules adsorbed on the particle surface presumably reduced the van der Waals forces.
- $d_p \sim 0\text{-}30\ \mu m$
- Example: flour, cement

Group A

- Aeratable
- Characterized by a small d_p and small ρ_p
- U_{mb} is significantly larger than U_{mf}
- Large bed expansion before bubbling starts
- Gross circulation of powder even if only a few bubbles are present
- Large gas backmixing in the emulsion phase
- Rate at which gas is exchanged between the bubbles and the emulsion is high
- Bubble size reduced by either using a wider particle size distribution or reducing the average particle diameter
- There is a maximum bubble size
- $d_p \sim 30\text{-}100\ \mu m$
- Examples: FCC, milk flour

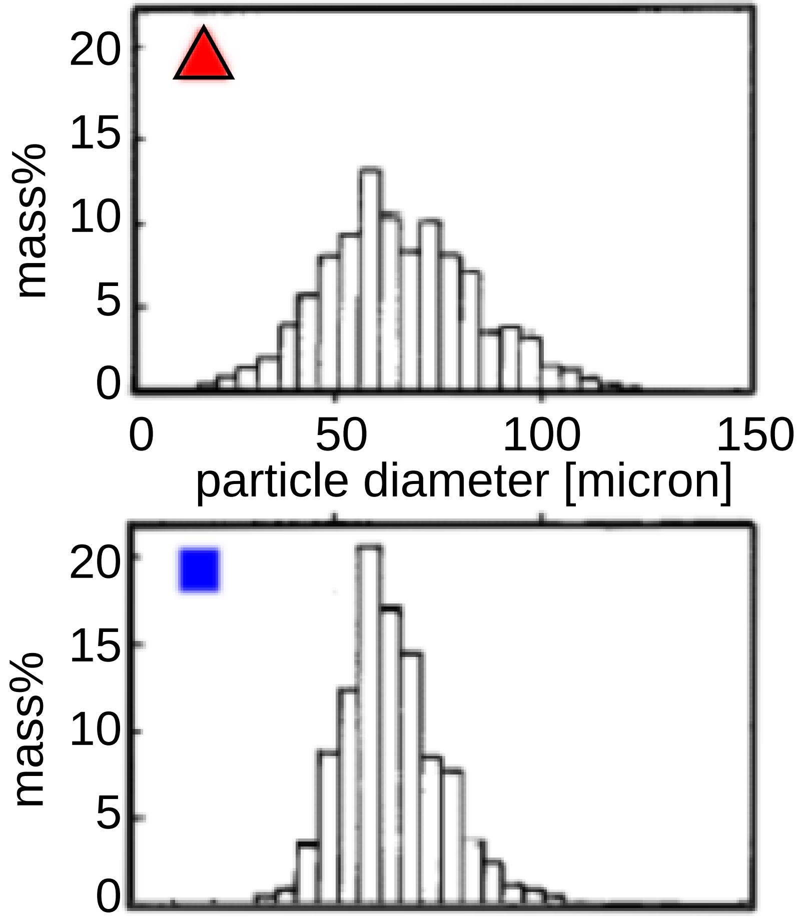
Group B

- Bubbling
- U_{mb} and U_{mf} are almost identical
- Solids recirculation rates are smaller
- Less gas backmixing in the emulsion phase
- Rate at which gas is exchanged between bubbles and emulsion is smaller
- Bubbles size is almost independent of the mean particle diameter and the width of the particle size distribution
- No observable maximum bubble size
- $d_p \sim 100\text{-}1000\ \mu m$
- Example: sand

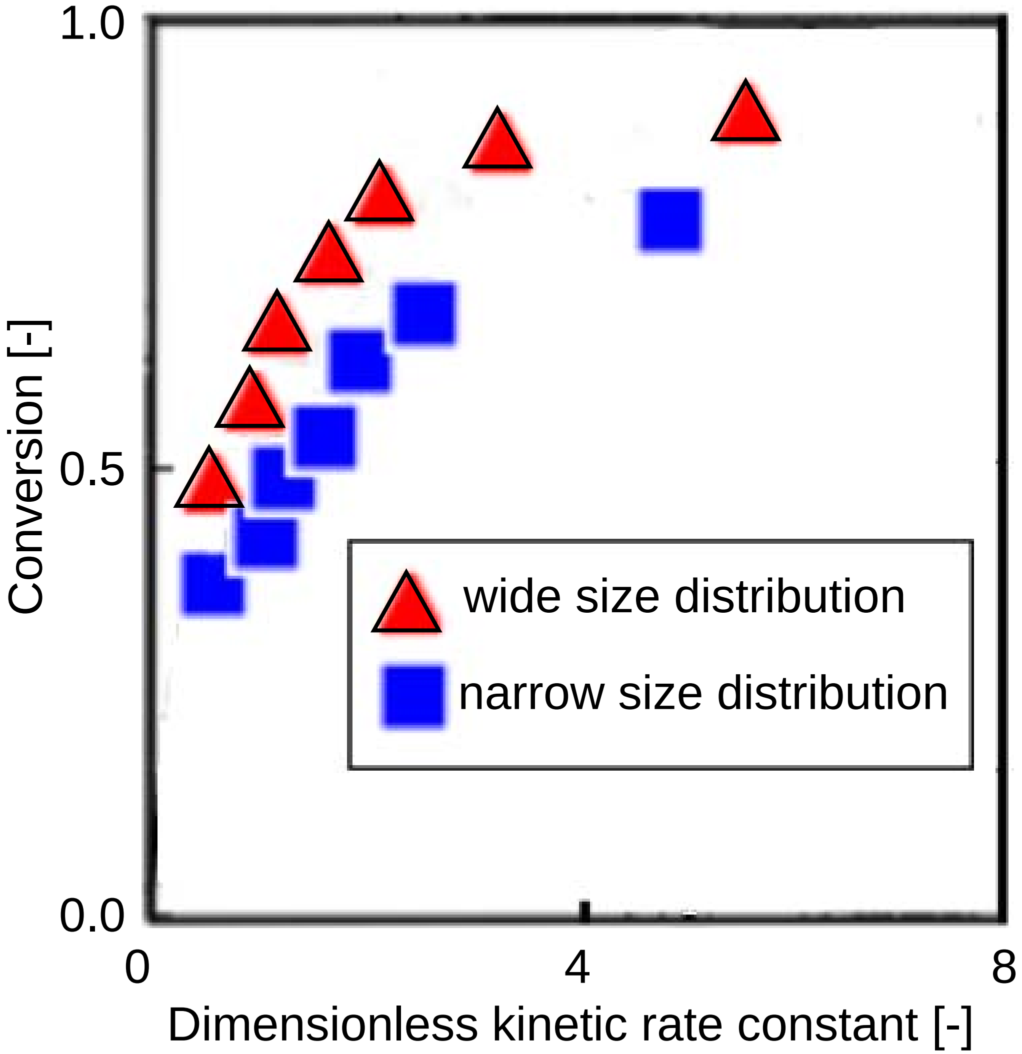
Group D

- Spoutable
- Either very large or very dense particles
- Bubbles coalesce rapidly and flow to large size
- Bubbles rise more slowly than the rest of the gas percolating through the emulsion
- Dense phase has a low voidage
- $d_p \sim >1000$ mm
- Examples: Coffee beans, wheat, lead shot

Influence of particle size distribution

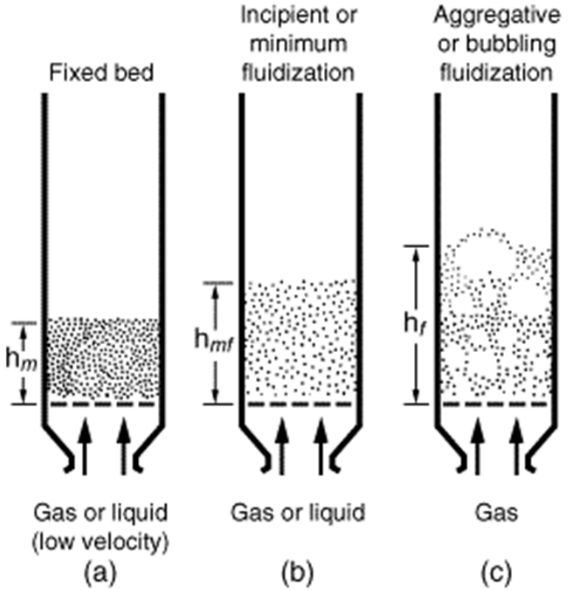


Adapted from Sun & Grace (1990)



Implication of U_{mb}/U_{mf}

- U_{mb}/U_{mf} could be used as an important index for the fluidization performance of fine particle fluidized beds on local hydrodynamics
- Geldart particle classification:
  - Group A powders with $U_{mb}/U_{mf}>1$
  - Group B powders with $U_{mb}/U_{mf}=1$



Demarcation

Demarcation between Group A and B powders

$$\frac{U_{mb}}{U_{mf}} \geq 1 \qquad U_{mb} = K\,d_p$$

For air at room T and P, K = 100 (Yang, W.-C., 2003)

$$\frac{8 \times 10^{-4}\, g\, d_p \left(\rho_p - \rho\right)}{K\,\mu} \leq 1$$

Pressure and temperature effect: (Grace, 1986)

$$\left(d_p^*\right)_{AB} = 101\left(\frac{\rho_p - \rho}{\rho}\right)^{-0.425}$$

If : $\left(d_p^*\right) < \left(d_p^*\right)_{AB}$ → powder belongs to Group A or C

If : $\left(d_p^*\right) > \left(d_p^*\right)_{AB}$ → powder belongs to Group B or D

Demarcation between Group B and D powders

For Group D powders $U_B \leq \frac{U_{mf}}{\varepsilon_{mf}}$

Demarcation

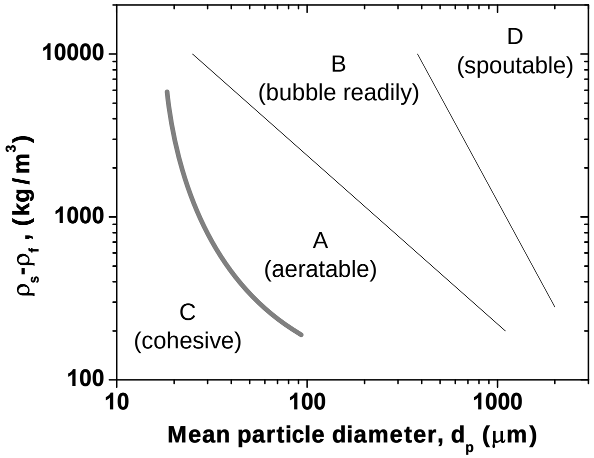
Demarcation between Group C and A powders (Molerus, 1982)

$$10(\rho_p - \rho_f)d_p^3 g / F_H = 0.01$$

F_H is the adhesion force determined experimentally.
(F_H=8.76x10^{-8} N for glass beads and FCC catalysts)

Demarcation between Group A and B powders

$$\frac{U_{mb}}{U_{mf}} \geq 1$$

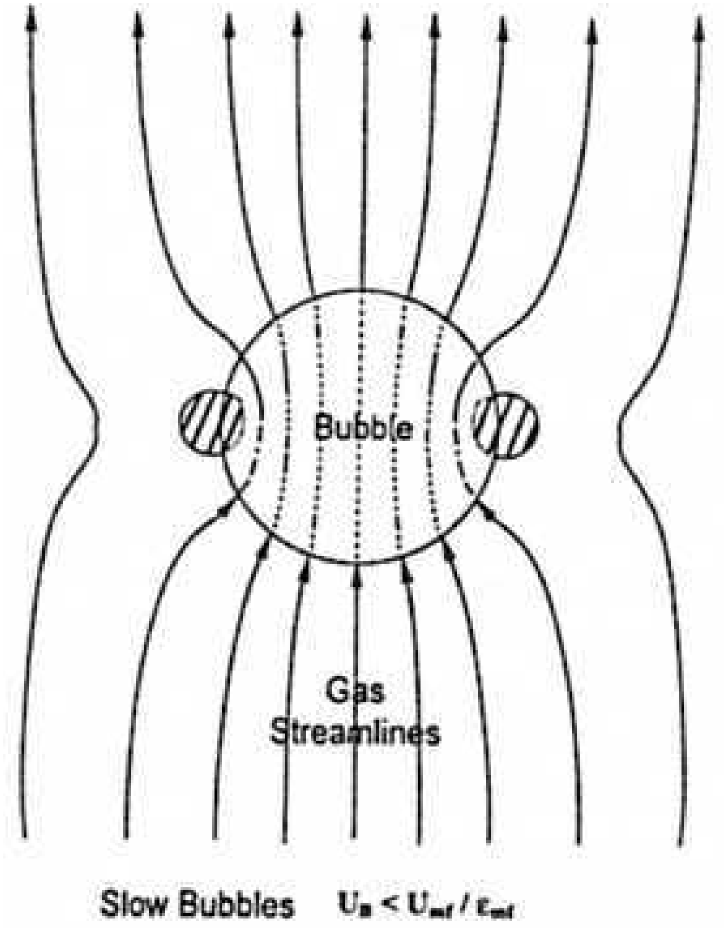


Demarcation between Group B and D powders

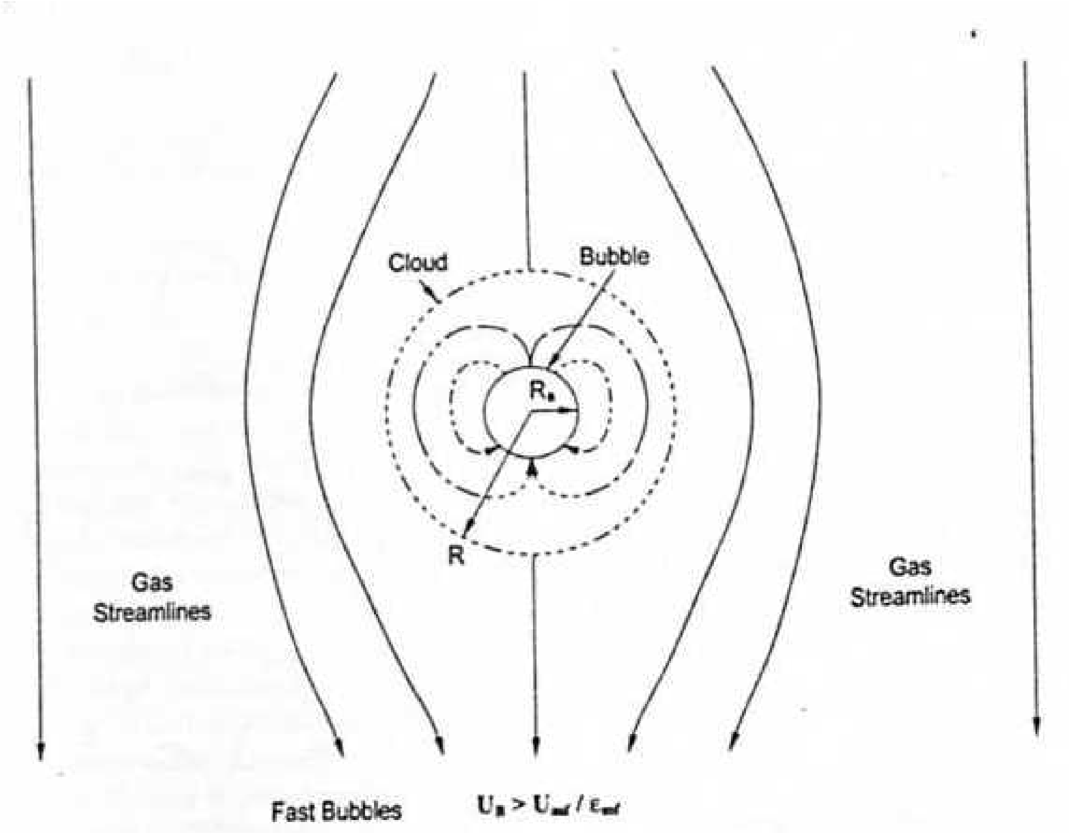
For Group D powders $U_B \leq \dfrac{U_{mf}}{\varepsilon_{mf}}$

Characteristics of single bubble: Slow vs. fast bubbles (Davidson model)

slow bubble



fast bubble



group D

group A & B

Simple demarcation criteria accounting for T, P effects

Goosen's classification:

- C/A boundary: Ar=9.8
- A/B boundary: Ar=88.5
- B/D boundary: Ar=176,900

Grace's classification:

- A/B boundary: Ar=125
- B/D boundary: Ar=145,000

$$Ar = \frac{g\rho_g\left(\rho_p - \rho_g\right)d_p^3}{\mu^2}$$

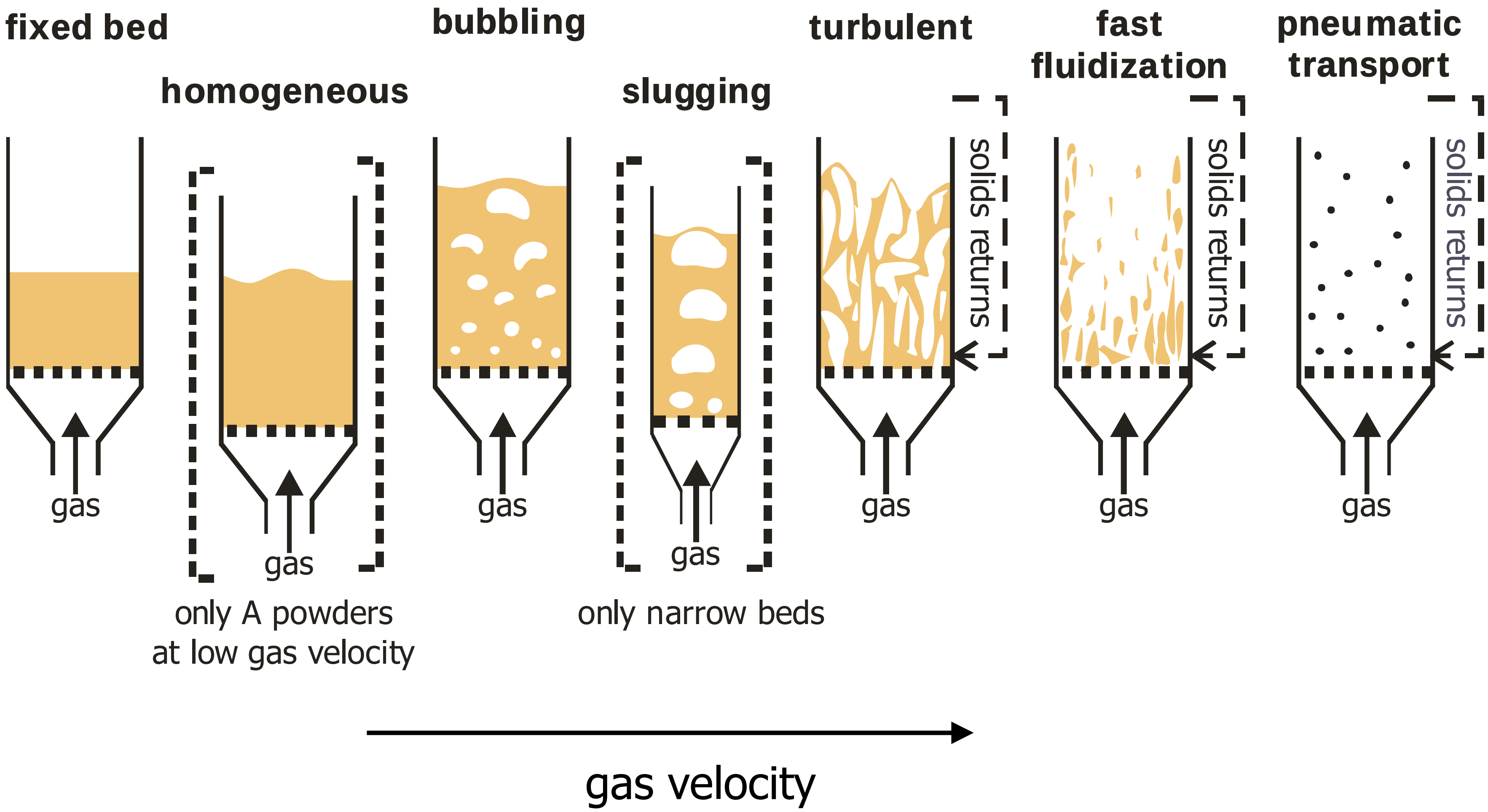
Correlations for U_{mb}

- Abrahamsen and Geldart (1980)

$$\frac{U_{mb}}{U_{mf}} = \frac{2300\rho_g^{0.126}\mu_g^{0.523}\,exp\left(0.716F_{45}\right)}{d_p^{0.8}g^{0.934}\left(\rho_p-\rho_g\right)^{0.934}}$$

Where F_{45} is the fraction of solids which are less than 45μm.

Flow Regimes



fixed bed

homogeneous

bubbling

slugging

turbulent

fast fluidization

pneumatic transport

solids returns

solids returns

solids returns

gas

gas

gas

gas

gas

gas

gas

only A powders
at low gas velocity

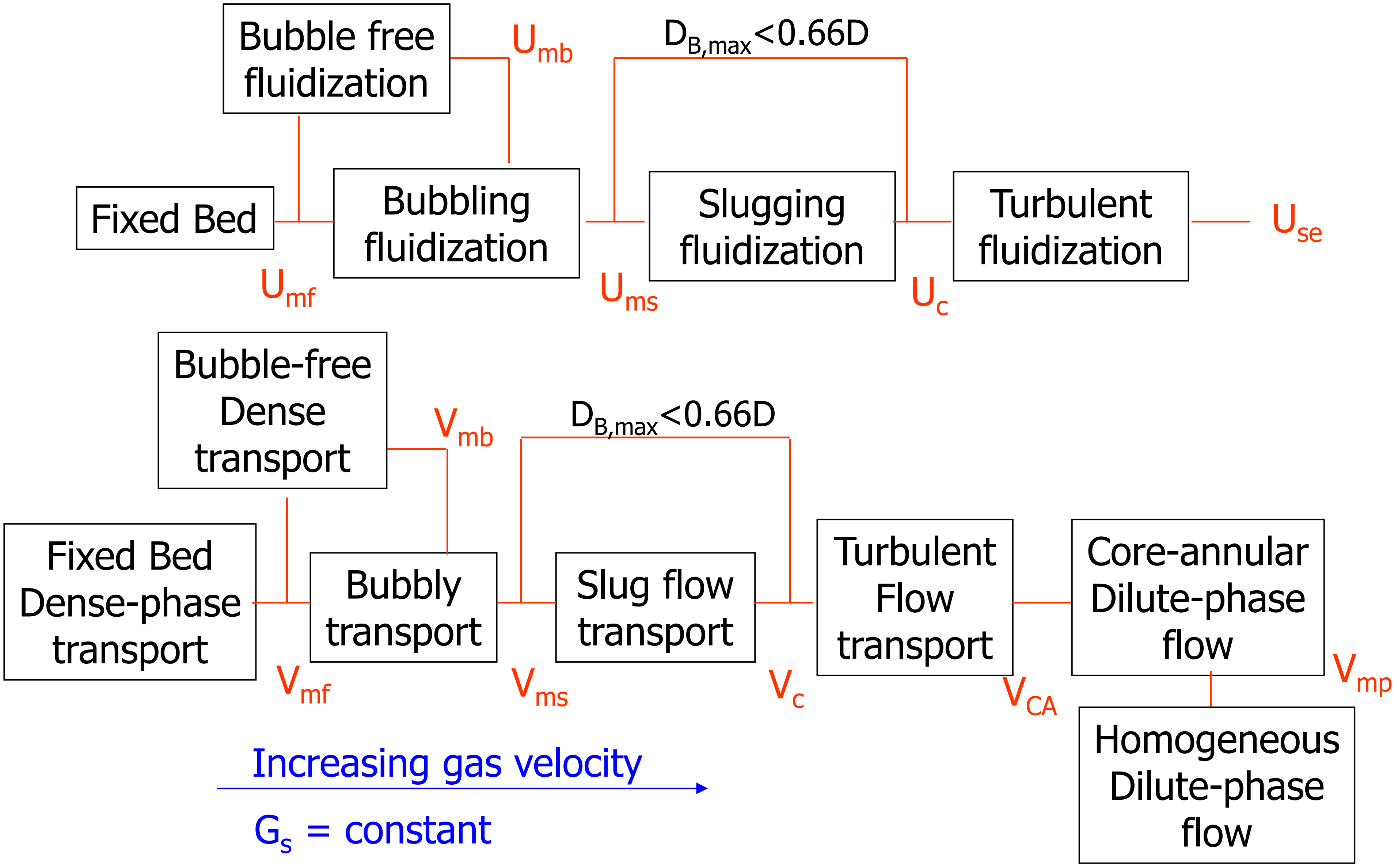
only narrow beds

gas velocity

Flow Regimes

U range	**Regime**	**Appearance and Principal Features**
$0 \le U < U_{mf}$	Fixed Bed	Particles are quiescent; gas flows through interstices
$U_{mf} \le U < U_{mb}$	Particulate regime	Bed expands smoothly in a homogeneous manner; top surface well defined, small-scale particle motion
$U_{mb} \le U < U_{ms}$	Bubbling regime	Gas voids form near distributor, coalesce and grow; rise to surface and break through
$U_{ms} \le U < U_c$	Slug flow regime	Bubble size approaches column cross-section. Top surface rises and collapses with regular frequency.
$U_c \le U < U_k$	Turbulent fluidization flow regime	Pressure fluctuations gradually decrease until turbulent fluidization flow regime is reached.
$U_k \le U < U_{tr}$	(Turbulent Regime)	Small gas voids and particle clusters dart to and fro. Top surface is difficult to distinguish.
$U \ge U_{tr}$	Fast Fluidization	No upper surface to bed, particles transported out the top in clusters and must be replaced.
$U >> U_{tr}$	Pneumatic conveying	No bed. All particles fed in are transported out the top as a lean phase.

Regime Transition Flow Chart



Bubble free fluidization

U_{mb}

$D_{B,max} < 0.66D$

Fixed Bed

Bubbling fluidization

Slugging fluidization

Turbulent fluidization

U_{se}

U_{mf}

U_{ms}

U_c

Bubble-free Dense transport

V_{mb}

$D_{B,max} < 0.66D$

Fixed Bed Dense-phase transport

Bubbly transport

Slug flow transport

Turbulent Flow transport

Core-annular Dilute-phase flow

V_{mp}

V_{mf}

V_{ms}

V_c

V_{CA}

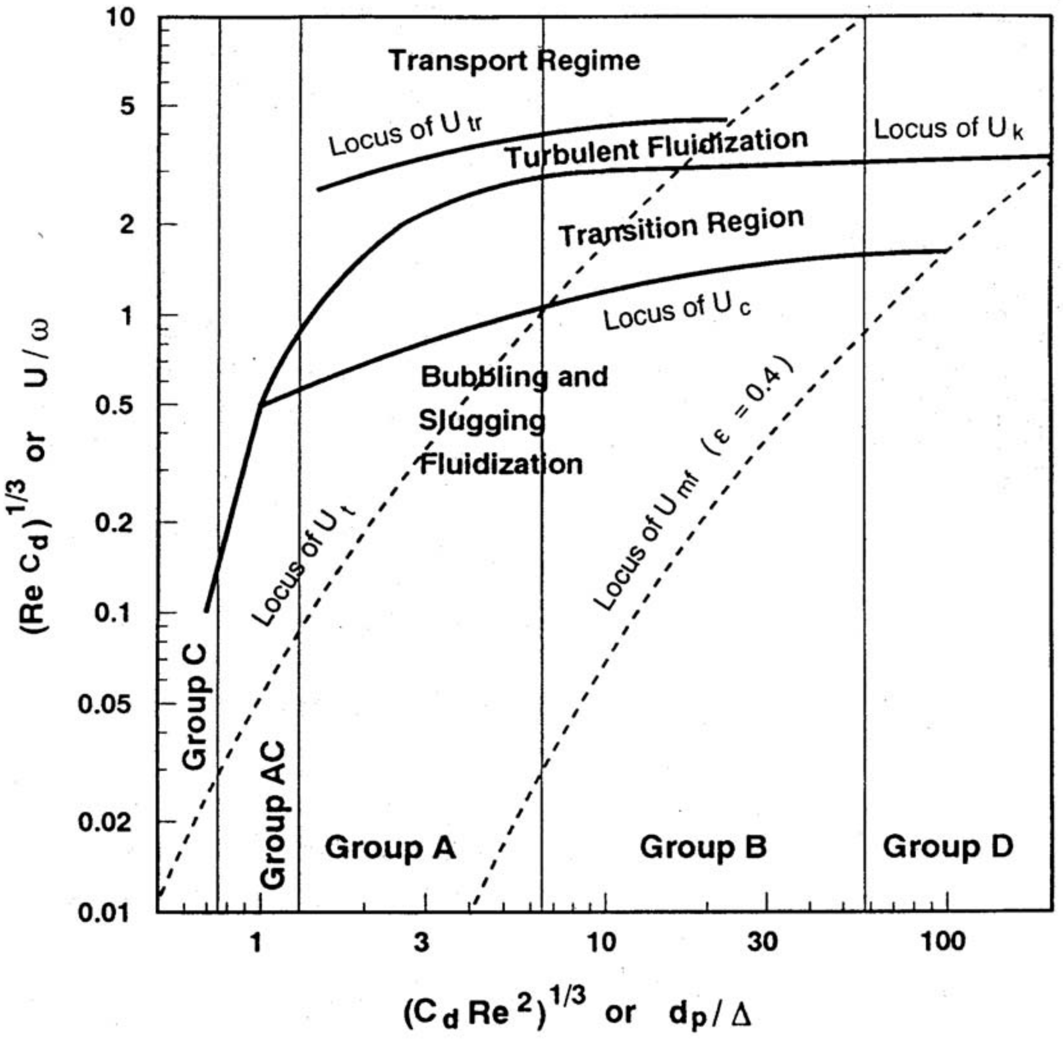
Homogeneous Dilute-phase flow

Increasing gas velocity

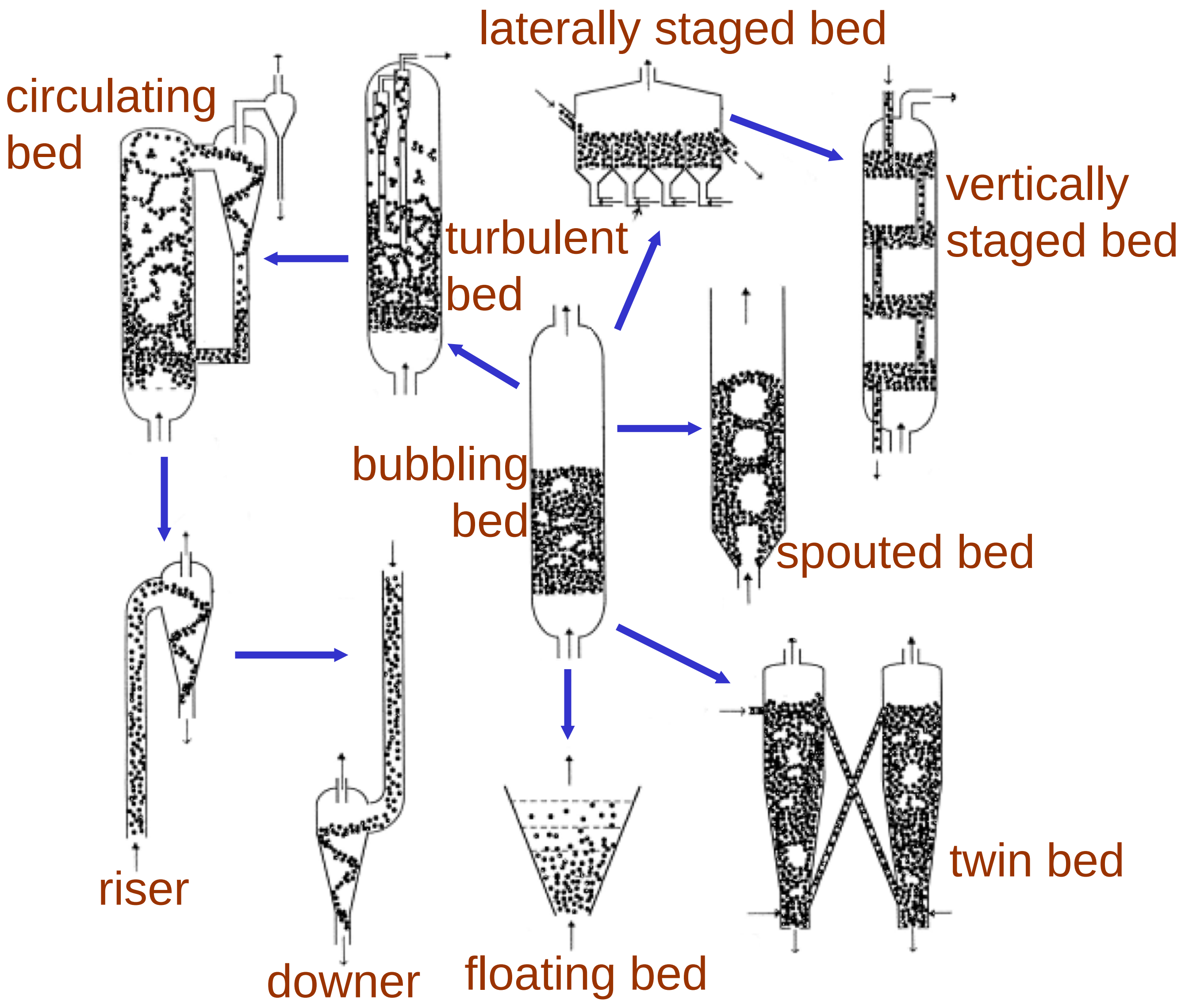
G_s = constant

Bi & Grace, 1995

Unifying Fluidization Regime Diagram



Fluidized bed lay-out



laterally staged bed

circulating bed

turbulent bed

vertically staged bed

bubbling bed

spouted bed

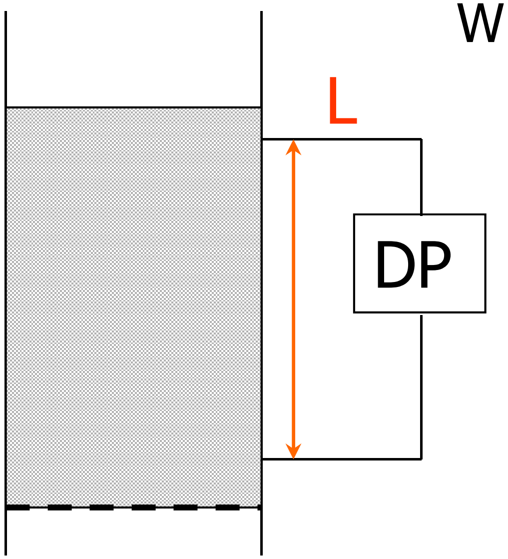
riser

twin bed

downer

floating bed

Static Head of Solids



When acceleration, friction and gas head are negligible

$$\Delta P = \frac{\text{weight of particles} - \text{upthrust on particles}}{\text{bed cross} - \text{sectional area}}$$

$$\frac{\Delta P}{L} = (1 - \varepsilon)(\rho_p - \rho)g$$

Time-averaged pressure measurements

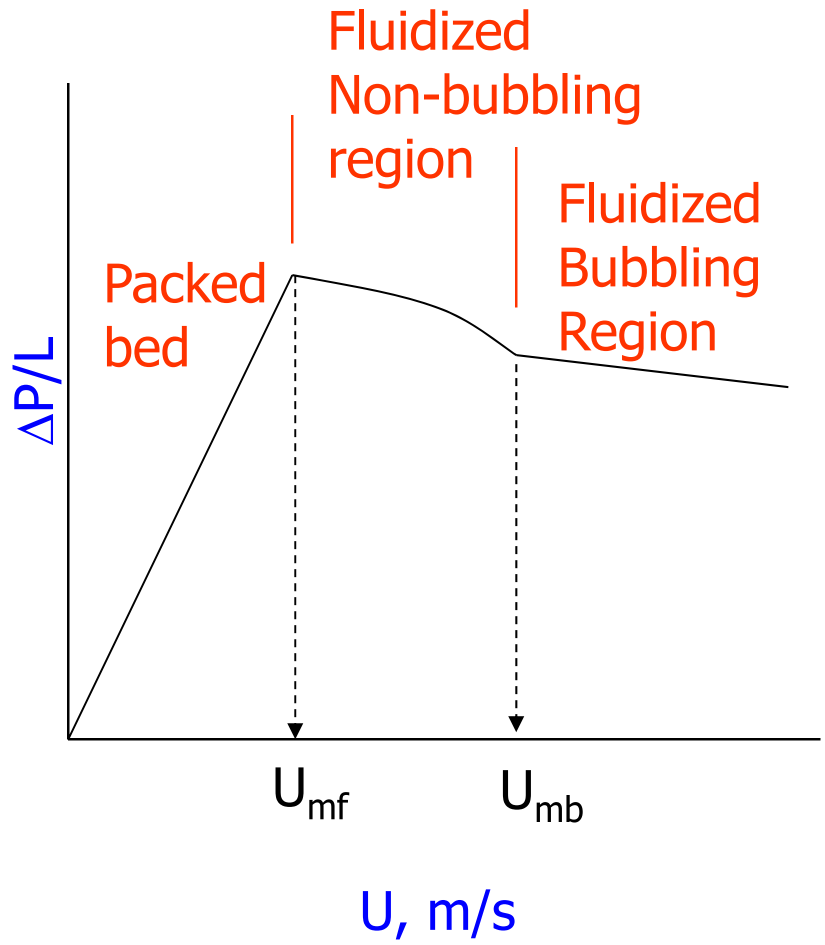
- Most industrial and pilot plant fluidized beds have pressure taps.
- There should be at least 2 or 3 taps within the fluidized bed.
- Pressure measurement from plenum chamber must be from where it will not be affected by either the gas expansion or the contraction
- For hot units, back flushed taps are often used.

(gas flowrate must be regulated)

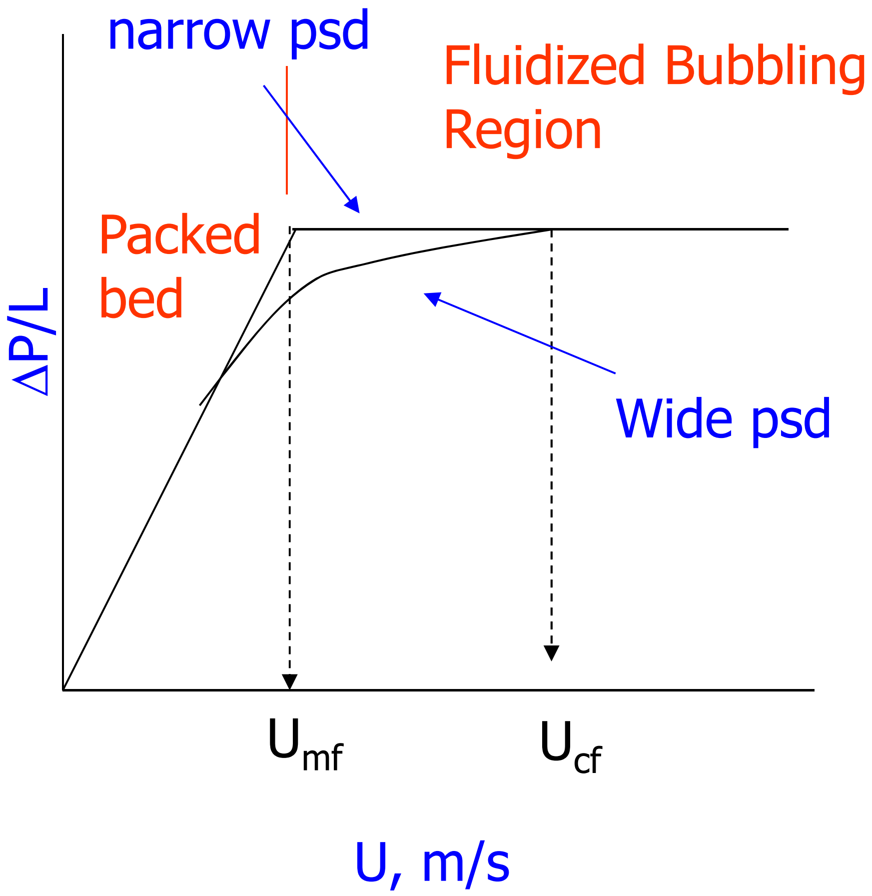
For other measurement techniques in fluidized beds:
van Ommen & Mudde, Int J Chem Reactor Eng 6 (2008) R3

Fluidization Curve

Group A particles



Group B particles



Minimum Fluidization

The frictional pressure drop at the point of minimum fluidization equalizes the bed mass per unit of cross-sectional area.

$$-\Delta P_{friction} = 150\frac{U_{mf}\mu\left(1-\varepsilon_{mf}\right)^2 \Delta x_{mf}}{D_{sv}^2\varepsilon_{mf}^3} + 1.75\frac{U_{mf}^2\rho\left(1-\varepsilon_{mf}\right)\Delta x_{mf}}{D_{sv}\varepsilon_{mf}^3}$$

$$= g\Delta x_{mf}\left(\rho_p-\rho\right)\left(1-\varepsilon_{mf}\right)$$

The frictional pressure drop at the point of minimum fluidization (U_{mf} , ε_{mf} , Δx_{mf}), can be considered equal to the frictional pressure drop in a fixed bed (Ergun)

Minimum Fluidization Velocity (U_{mf})

Dimensionless relationship following from equation on previous slide

$$\mathrm{Re}_{mf} = \sqrt{C_1^2 + C_2 Ar} - C_1$$

$$Re_{mf} = U_{mf} \rho D_{sv} / \mu$$

$$Ar = g\rho\left(\rho_p - \rho\right) D_{sv}^3 / \mu^2$$

$$C_1 = 300\left(1 - \varepsilon_{mf}\right) / 7$$

$$C_2 = \varepsilon_{mf}^3 / 1.75$$

Minimum Fluidization

Estimation of ε_{mf}

$0.4 < \varepsilon_{mf} < 0.55$

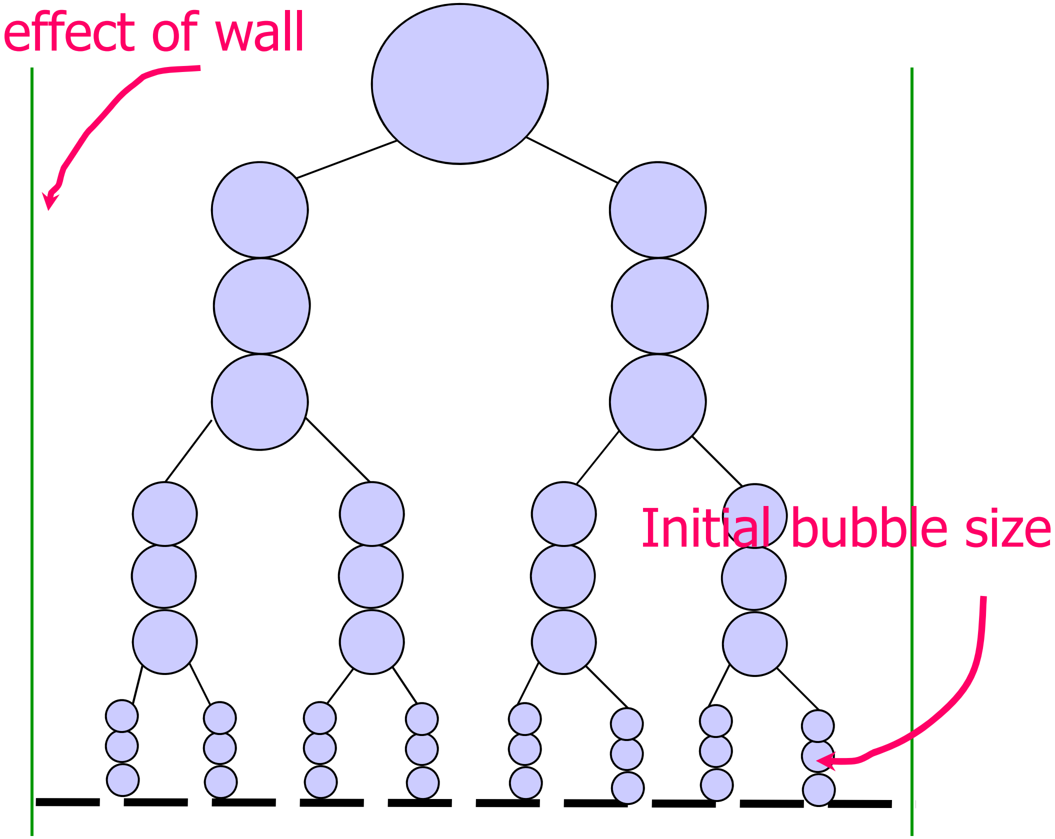
- $\varepsilon_{mf} \approx \varepsilon_{\text{fixed bed}}$
- $\varepsilon_{mf} \approx (14\,\phi)^{-1/3}$ where ϕ is the particle sphericity

Authors	C_1	C_2
Wen and Yu (1966)	33.7	0.0408
Richardson (1971)	25.7	0.0365
Saxena and Vogel (1977)	25.28	0.0571
Babu et al. (1978)	25.25	0.0651
Grace (1982)	27.2	0.0408
Chitester et al. (1984)	28.7	0.0494

Freely Bubbling Beds: Bubble Growth

Why grow?

1) The hydrostatic pressure on the bubbles decreases as they rise up the bed;
2) Bubbles may coalesce by one bubble catching up with another;
3) Bubbles which are side by side may coalesce;
4) Bubbles may grow by depleting the continuous phase locally.



Mean bubble size = f(type of distributor, distance above the distributor plate, excess gas velocity)

Freely Bubbling Beds: Bubble Size

Darton et al. (1977)

$$D_b(z) = 0.54\,g^{-0.2}\left(U-U_{mf}\right)^{0.4}\left(z+4\sqrt{\frac{A}{N_{or}}}\right)^{0.8}$$

where A/N_{or} is the area of distributor plate per orifice

Mori & Wen (1975)

$$\frac{D_{be}-D_b}{D_{be}-D_{b0}} = \exp\left(-0.3\frac{z}{D_T}\right)$$

Ranges of key variables:

$5 \le U_{mf} \le 200\,\mathrm{mm/s}$

$60 \le d_p \le 450\,\mu\mathrm{m}$

$U - U_{mf} \le 0.48\,\mathrm{m/s}$

$D_T \le 1.3\,\mathrm{m}$

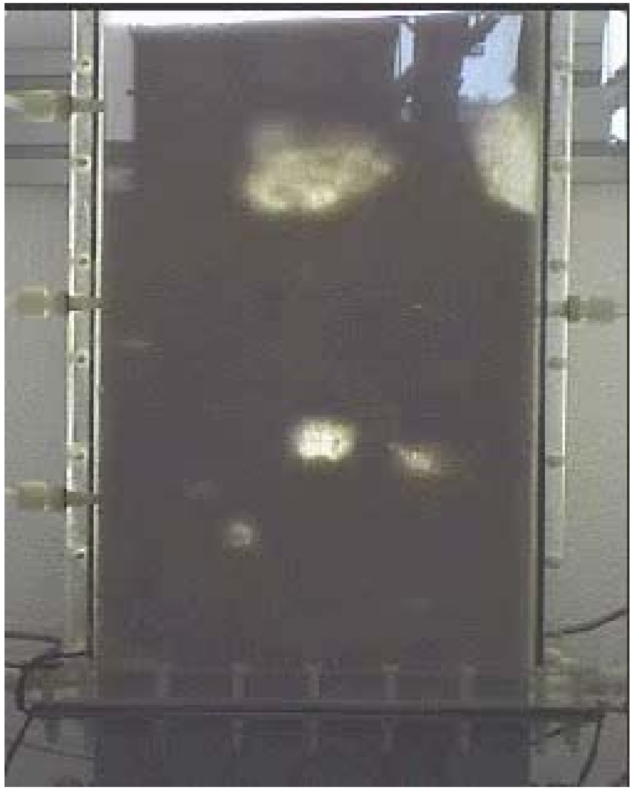
where:

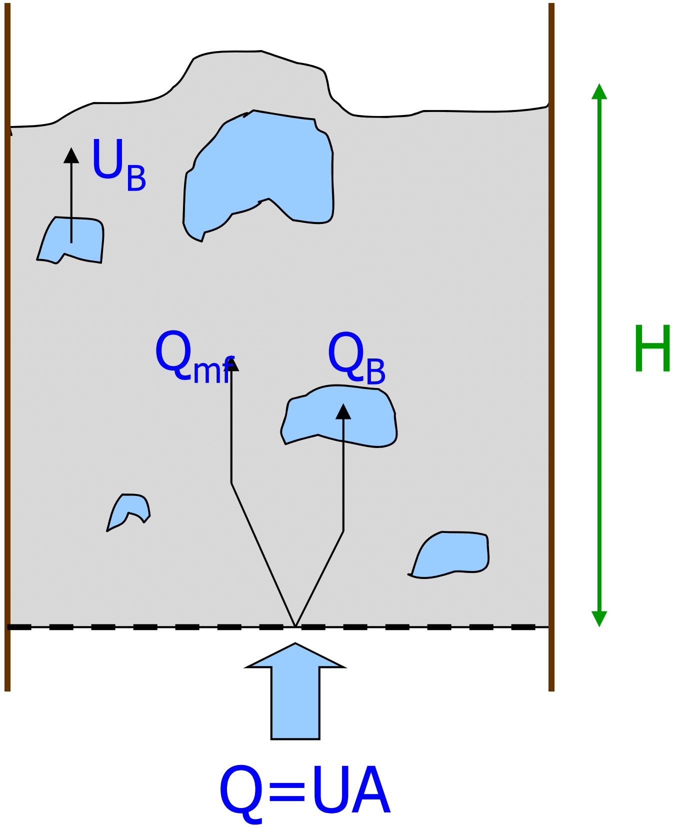
$$D_{b0} = \frac{1.38}{g^{0.2}}\left[\frac{A}{N_{or}}\left(U-U_{mf}\right)\right]^{0.4}$$ for perforated plates

$$D_{b0} = 0.376\left(U-U_{mf}\right)^2$$ for porous plates

$$D_{be} = 1.64\left[A\left(U-U_{mf}\right)\right]^{0.4}$$

Two-phase theory





Two-phase theory

Total gas flow rate: $Q = UA$

Flow rate in dense phase: $Q_{mf} = U_{mf} A$

Gas passing through the bed as bubbles: $Q - Q_{mf} = (U - U_{mf})A$

Fraction of the bed occupied by bubbles:

$$\delta_b = \frac{H - H_{mf}}{H} = \frac{Q - Q_{mf}}{AU_B} = \frac{U - U_{mf}}{U_{\mathbf{B}}}$$

The distribution of the gas between the bubbles and dense phase is of interest because it influences the degree of chemical conversion.

In practice, the two-phase theory overestimates the volume of gas passing through the bed as bubbles

$$\frac{Q_B}{A} = Y(U - U_{mf})$$

Visible bubble flow rate:

where $0.8 < Y < 1.0$ for Group A powders

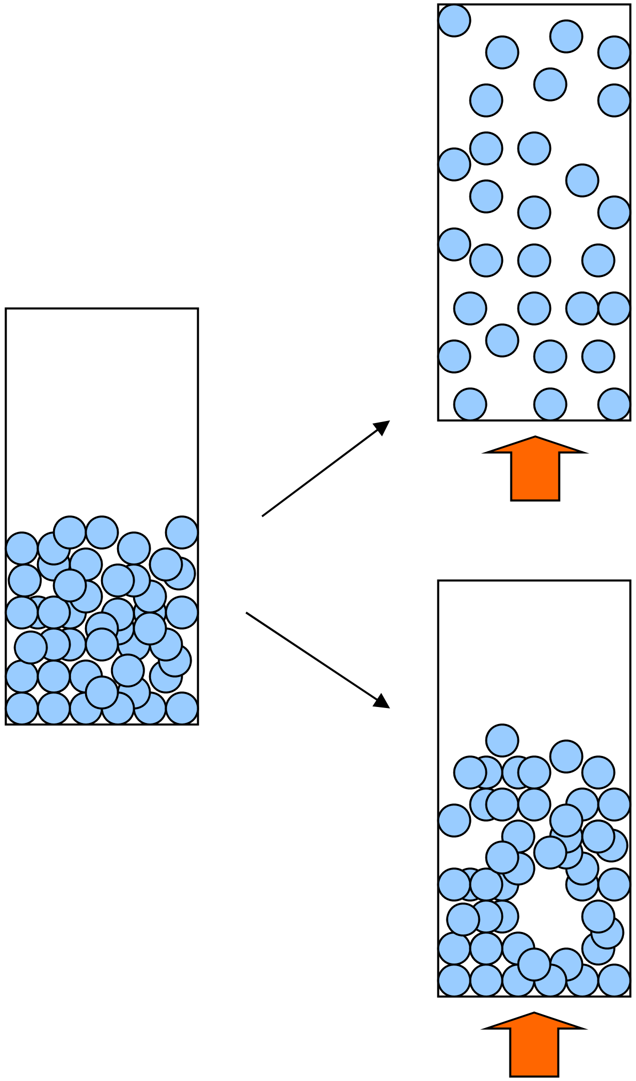
where $0.6 < Y < 0.8$ for Group B powders

where $0.25 < Y < 0.6$ for Group D powders

Baeyens and Geldart (1985)

$$\boxed{Y = 2.27 Ar^{-0.21}}$$

Homogeneous fluidization



Non-bubbling fluidization
Particulate or homogeneous fluidization

Mechanism???
Delay caused in the adjustment of the mean particle velocity to a change in the local concentration resulting from the larger particle to fluid phase inertia (Didwania, 2001)

Bubbling fluidization
Aggregative or heterogeneous fluidization

Steady-state expansion of fluidized beds

$$V_B = L_B (1-\varepsilon)A$$

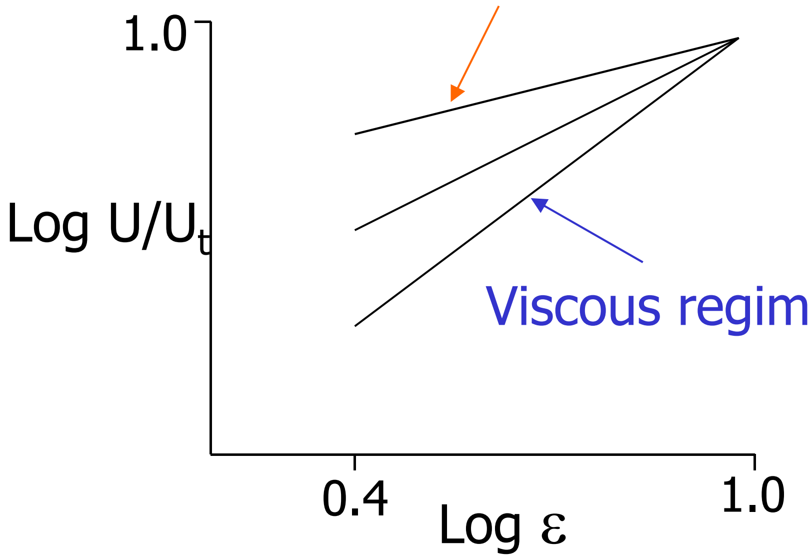
$$\Delta P_B = \left(\rho_p - \rho_g\right) V_B\, g = \text{constant}$$

V_B: total volume of particles

Homogeneous bed expansion: Richardson-Zaki relation

$$\frac{U}{U_t} = \varepsilon^n$$

Inertial regime: n=2.4



Viscous regime: n=4.8

mass of paticles in the bed :

$$M_B = (1-\varepsilon)\,\rho_p\, A\, H$$

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

Heat and mass transfer

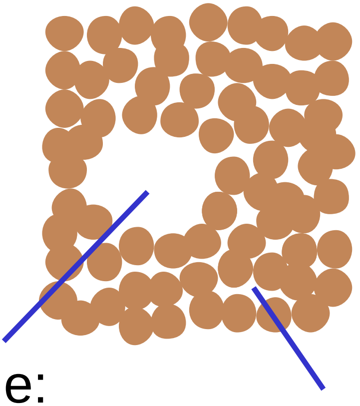
Heat transfer: particle to wall or internal
Mass transfer: gas to particle

Fluidized beds show an excellent heat transfer

Mixing of solids by (large) bubbles →
almost constant temperature throughout the reactor

However, large bubbles
decrease the mass transfer

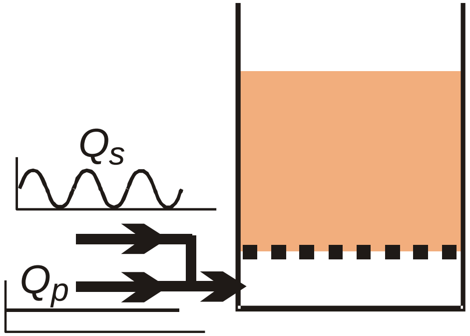
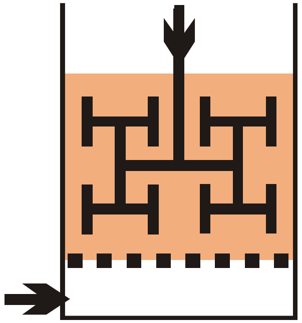
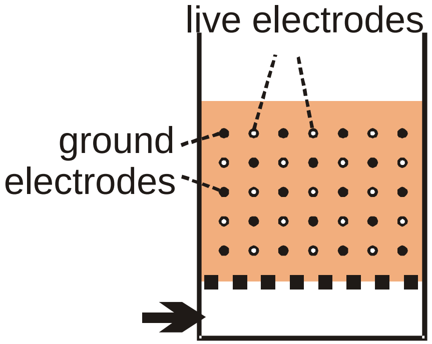
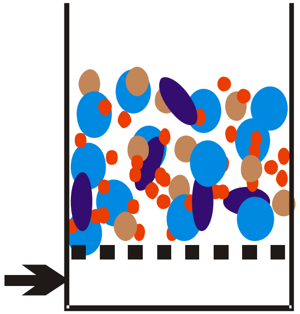
Research → decrease bubble size



Bubble:
shortcut
of gas

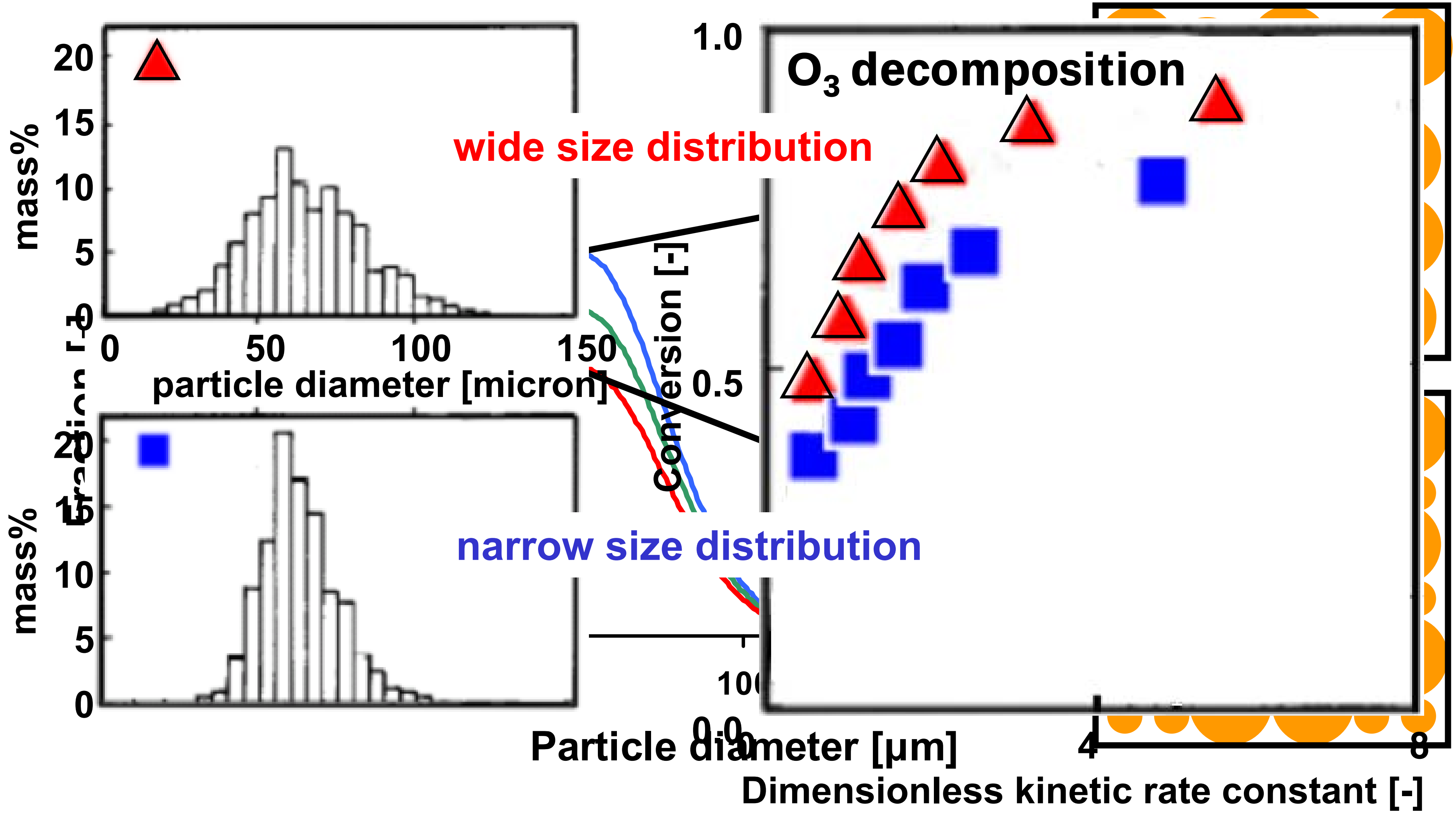
Interstitial gas:
effective

Ways to structure a fluidized bed

	Dynamics	**Geometry**
Gas		
Particles		

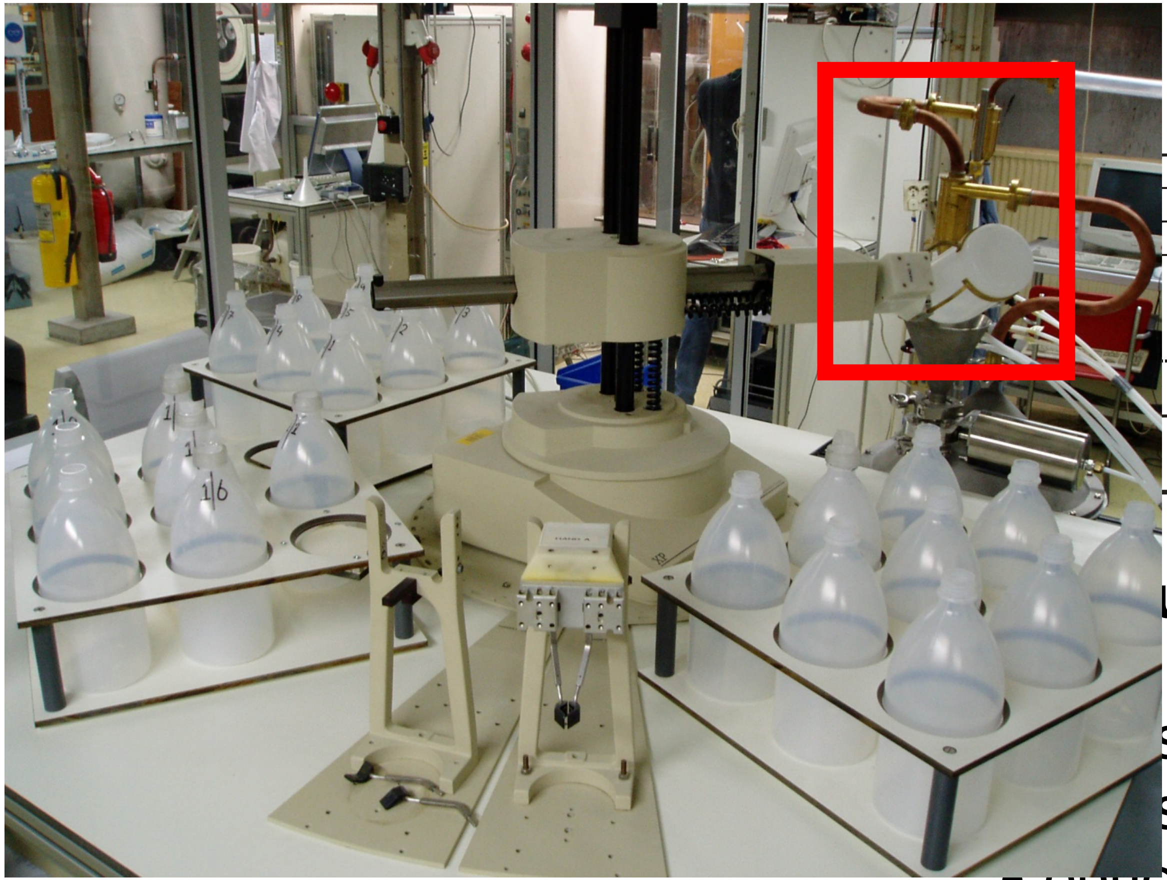
van Ommen *et al.*, *Ind. Eng. Chem. Res.*, 46 (2007) 4236

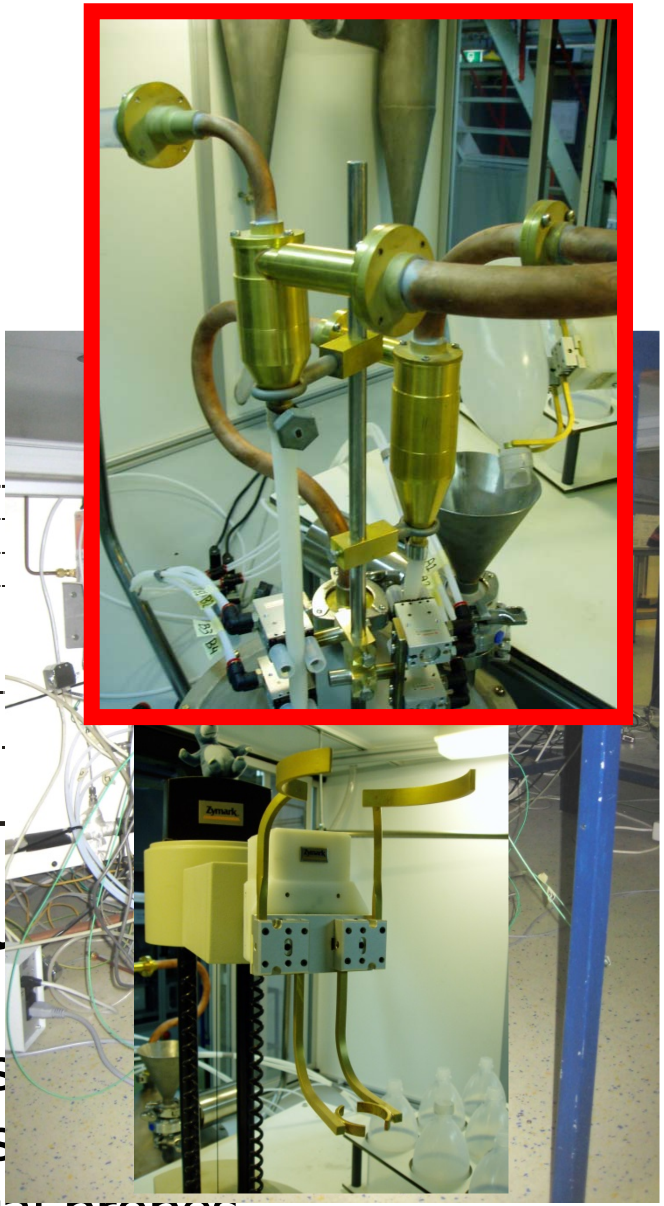
Tailoring the particle size distribution


wide size distribution

narrow size distribution

High Throughput Experimentation



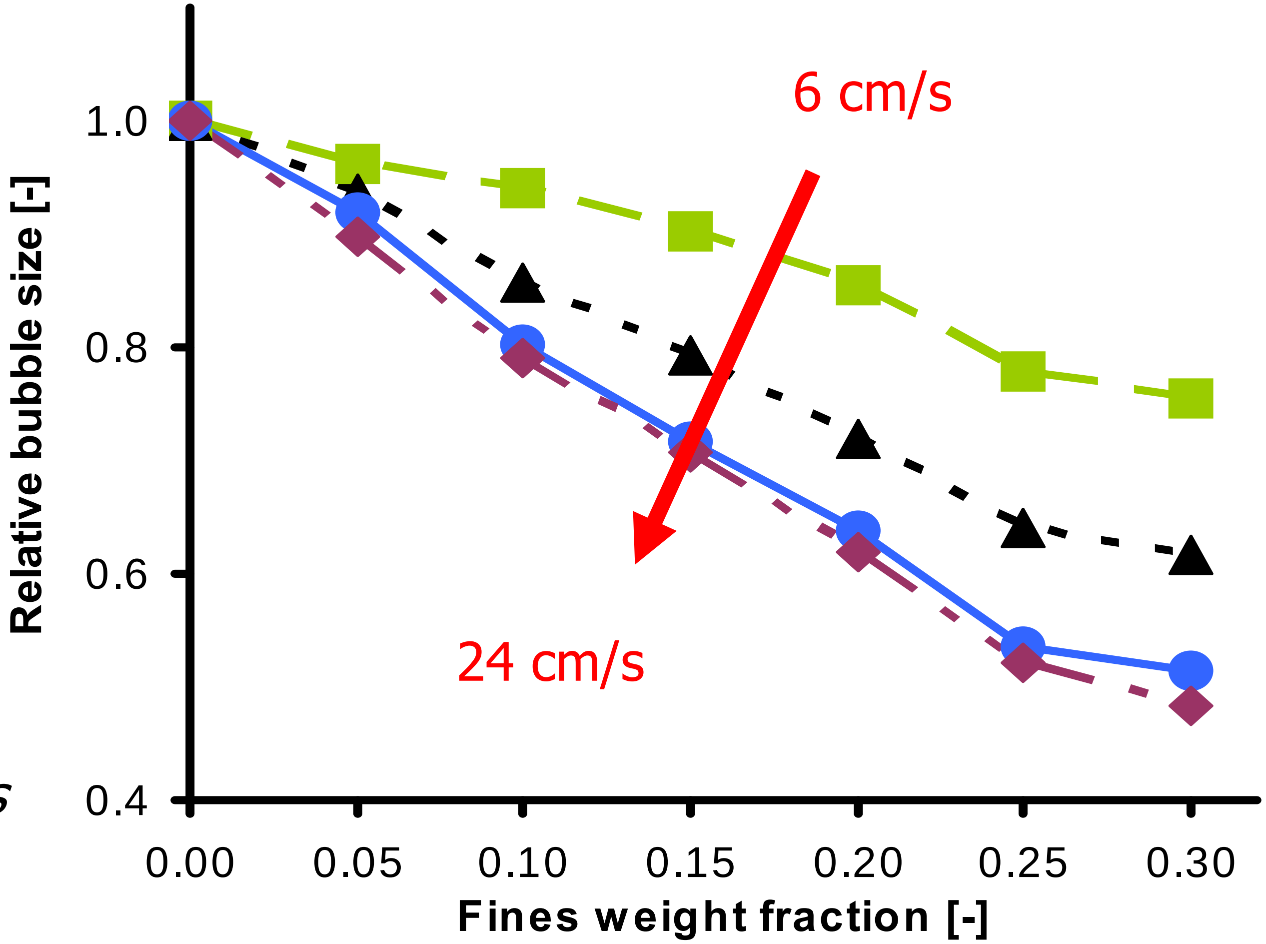


- optical probes

Fines effect on bubble size

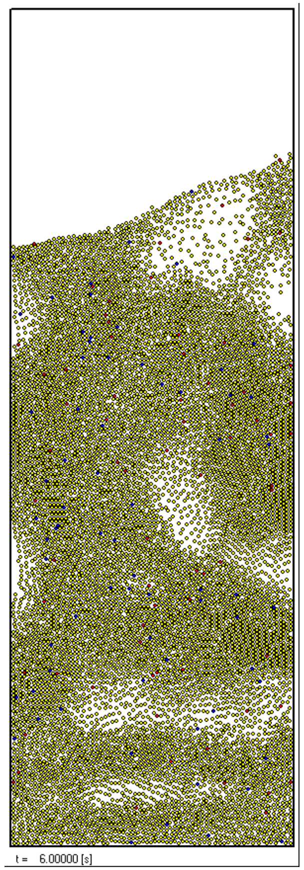
Fines are added to a powder with a D_{50} of 70 µm

Based on pressure fluctuations

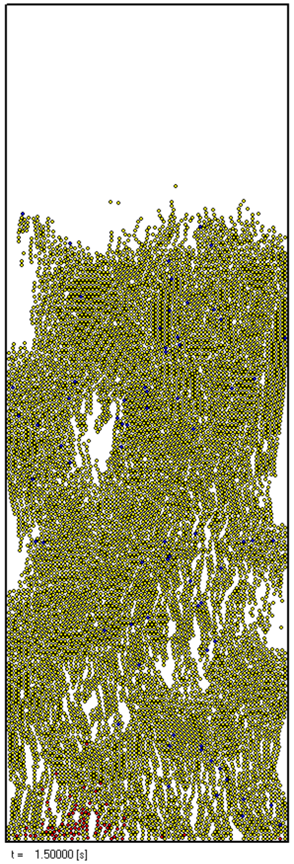


Beetstra *et al., AIChE J.* 55 (2009) 2013

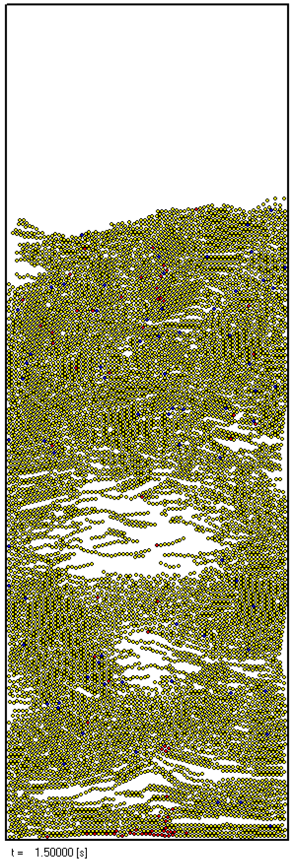
AC electric field: CFD simulations



No Field



Vertical
0.7 kV/mm, 30Hz



Horizontal
0.7 kV/mm, 30Hz

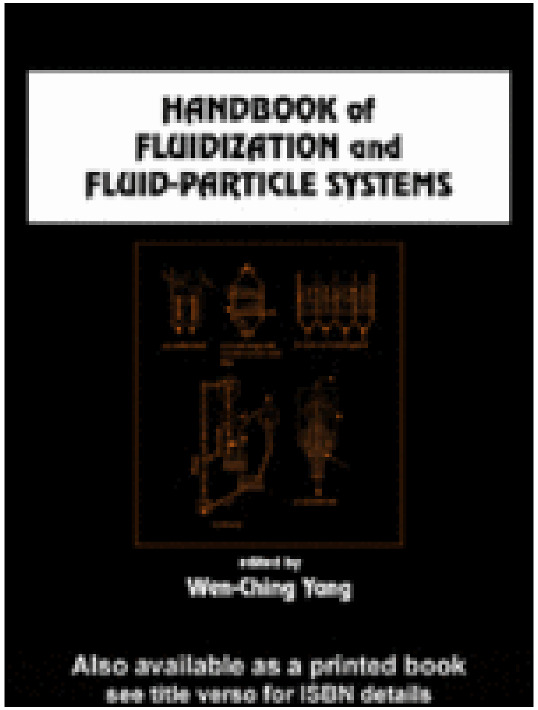
Discrete particle model with 10000 particles.
Uniform AC electric fields.

At experimental field strength: ~50 % reduction of bubble area.

Kleijn van Willigen *et al., Powder Technol.* 183 (2008) 196

Textbooks

Fluidization Engineering
Kunii, D. & Levenspiel, O.
ISBN: 8131200353
Pub. Date: Jan 2005 , 2nd ed.
Publisher: Elsevier



Handbook of Fluidization and Fluid-Particle Systems
Ed. Wen-Ching Yang
ISBN: 978-0-8247-0259-5
Pub. Date: March 2003
Publisher: Routledge, USA

Course Material

- Additional Resource Material
  - http://www.erpt.org/
  - Rhodes, M., Introduction to Particle Technology, Wiley, Sussex England, ISBN: 0471984833, 1999.
  - Yang, W.-C., Marcel Dekker, Handbook of Fluidization and Fluid-particle Systems, New York ISBN: 082470259, 2003.
  - Fan, L.-S., Gas-Liquid-Solid Fluidization Engineering, Butterworth-Heinemann, Boston, 1989.
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  - Grace, J.R., A. Avidan and T.M. Knowlton, "Circulating Fluidized Beds," Blackie Academic press, Boston, 1997.
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