JMBC Particle Technology 2019

Fluidization

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With material from Naoko Ellis**

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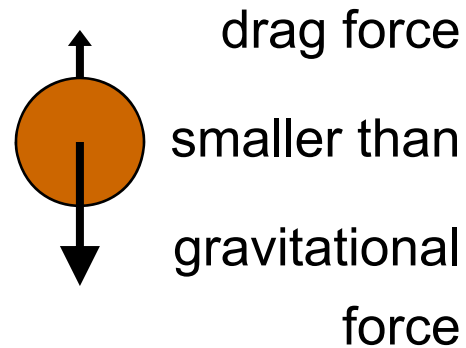# Powder Flow   versus   Gas-Particle Flow

- First particle-particle interaction, then gas-particle interaction

- Research area:
    granular matter

- First gas-particle interaction, then particle-particle interaction
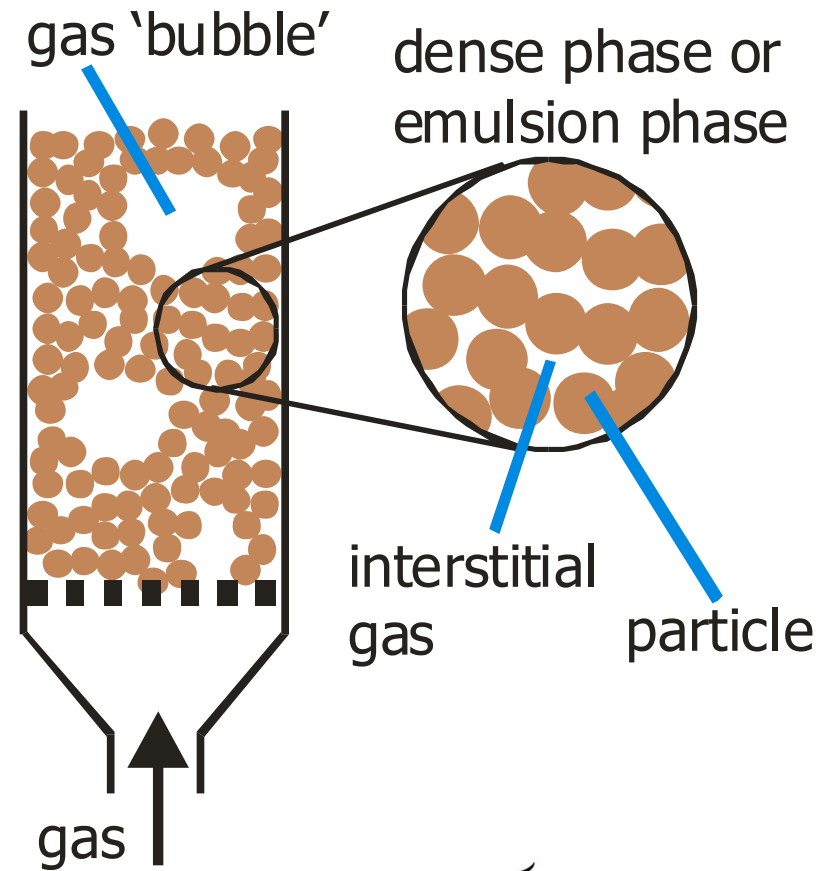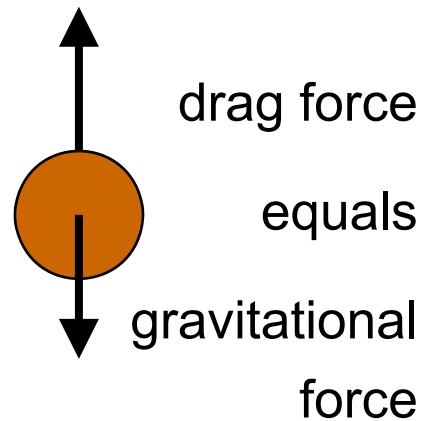
- Research area:
    fluid mechanics
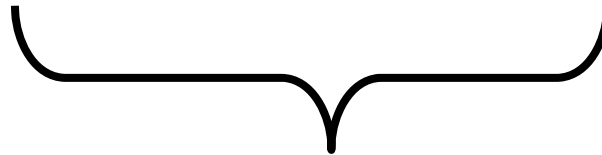
Chemical Engineering / Particle Technology

Intro gas-solids fluidized bed

Significance of Fluidized beds
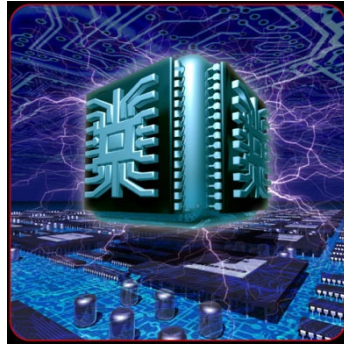
Advanced materials

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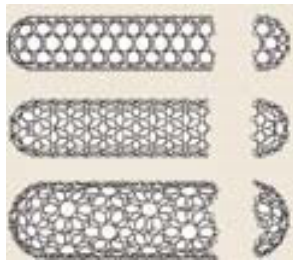
Chemical and Petrochemical

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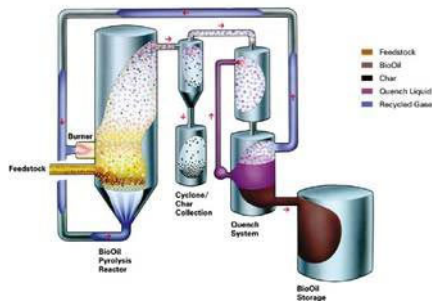
www.unb.ca/che/che5134/ fluidization.html

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

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

Gas-Solid Fluidized Bed
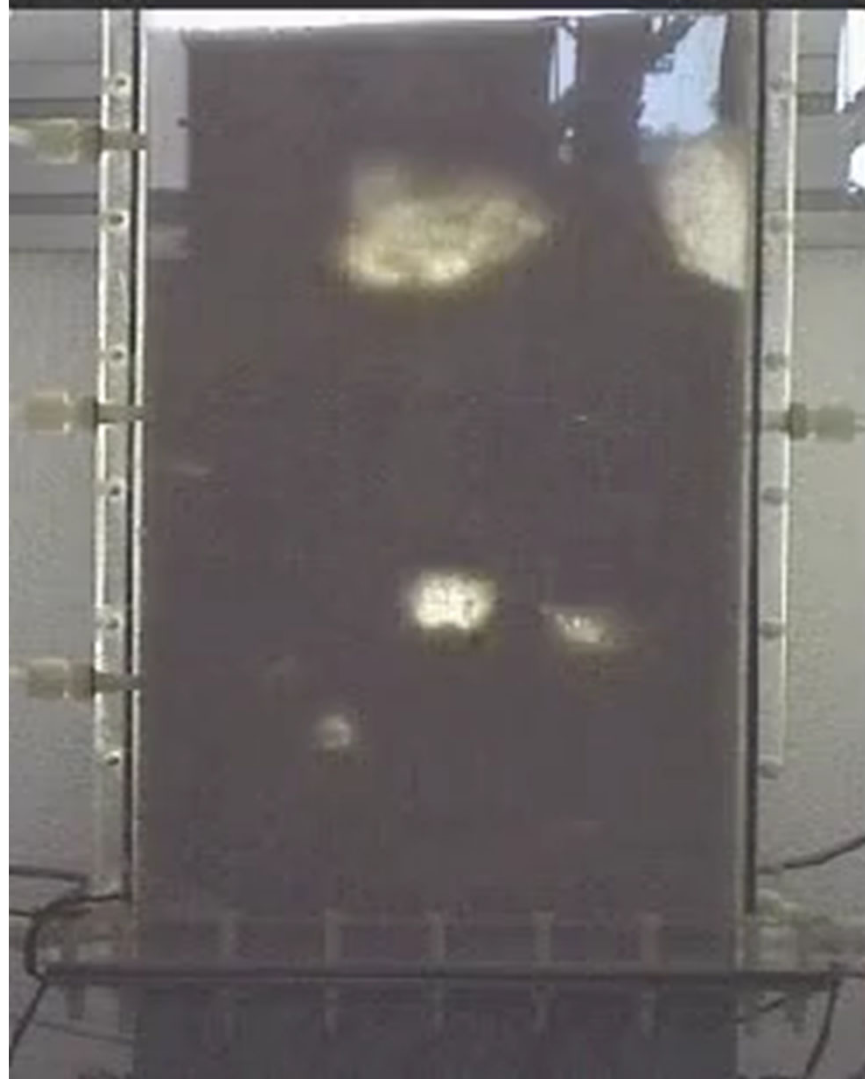
Characteristics of Gas Fluidized Beds

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

TUDelft

Question

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

Characteristics of Gas Fluidized Beds

Secondary Characteristics:

– Good heat transfer from surface to bed, and gas to particles

– Isothermal conditions radially and axially

– Pressure drop depends only on bed depth and particle density –
does not increase with gas velocity (ideal case)

– Particle motion usually streamlines → erosion and attrition

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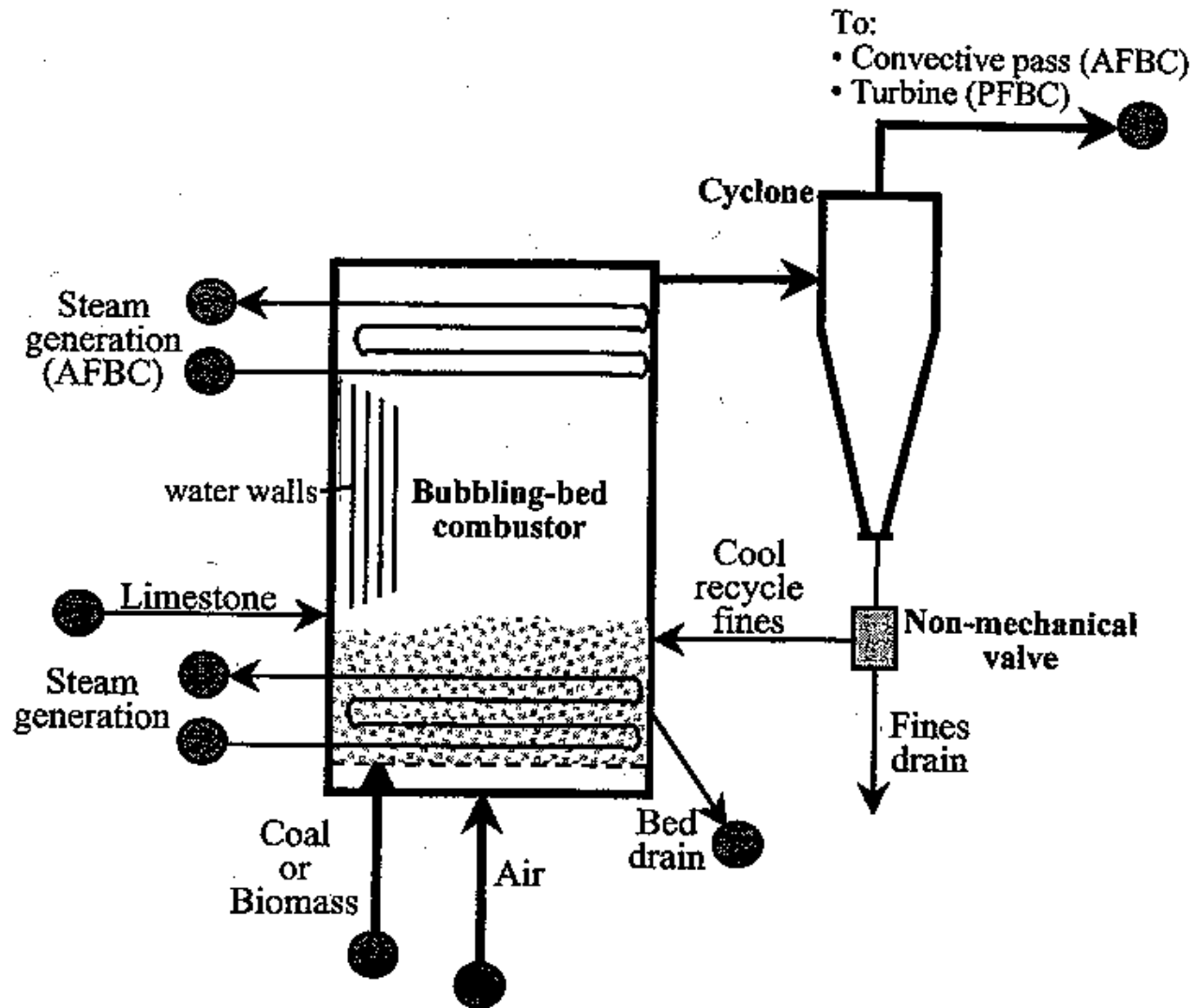
Yang W. Bubbling fluidized beds (Chapter 3). In: *Handbook of Fluidization and Fluid-Particle Systems. Yang W (Ed.). Marcel* Dekker, Inc., NY, NY, USA, 53–113 (2003).

1.56 m Diameter Column

Industrial Scale

Solid offtake

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

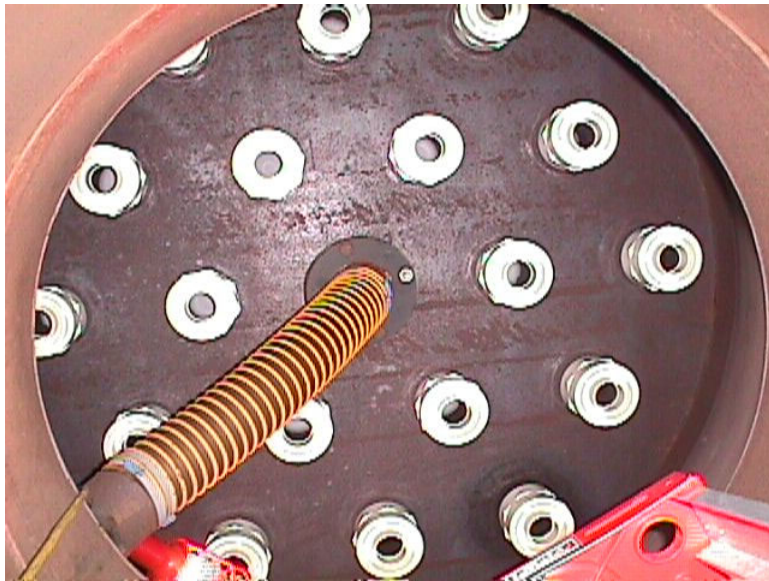# Approaches to the Study of Particulate Systems

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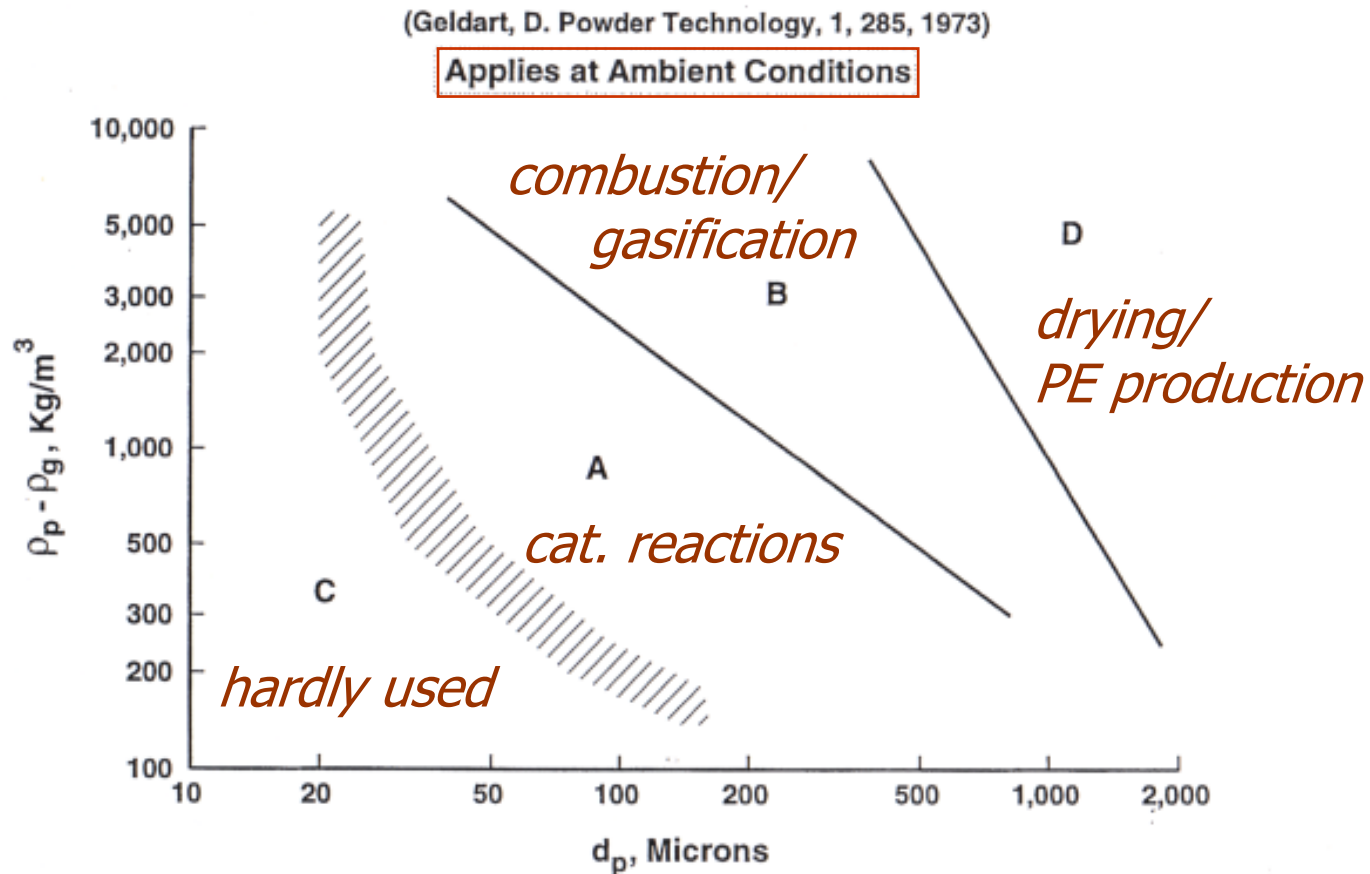
(Geldart, D. Powder Technology, 1, 285, 1973)

Applies at Ambient Conditions

combustion/ gasification

drying/ PE production

cat. reactions

hardly used

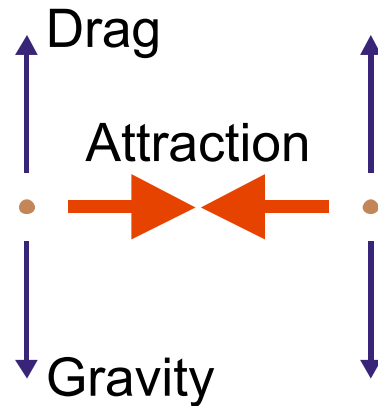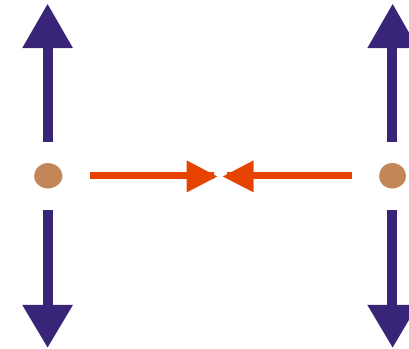A: Aeratable ($U_{mb} > U_{mf}$) Material Has a Significant Deaeration Time *(FCC Catalyst)*

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
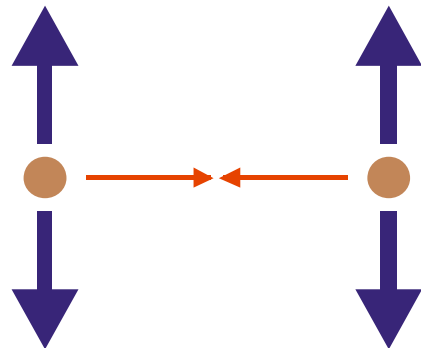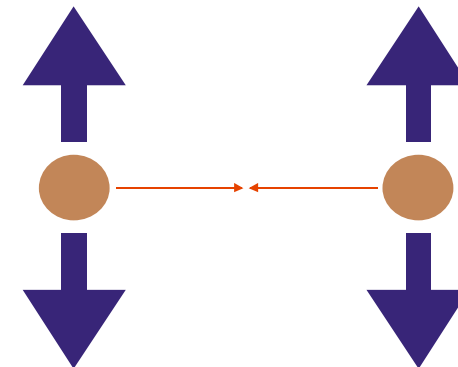
Group C

- 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.
- $d_p \sim$ 0-30 μm
- 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
- $d_p \sim$ 100-1000 μm
- 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
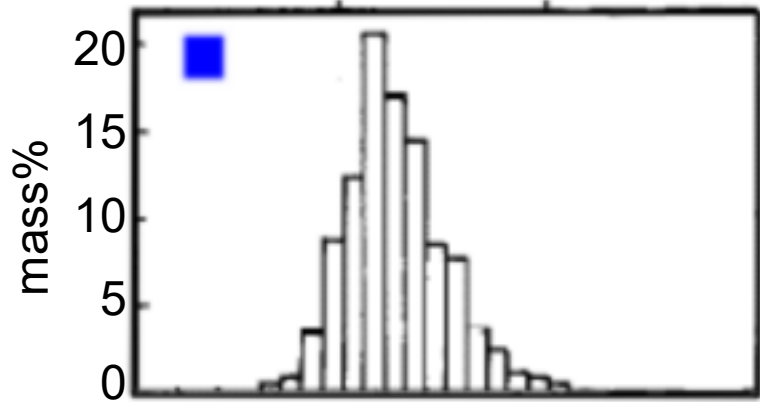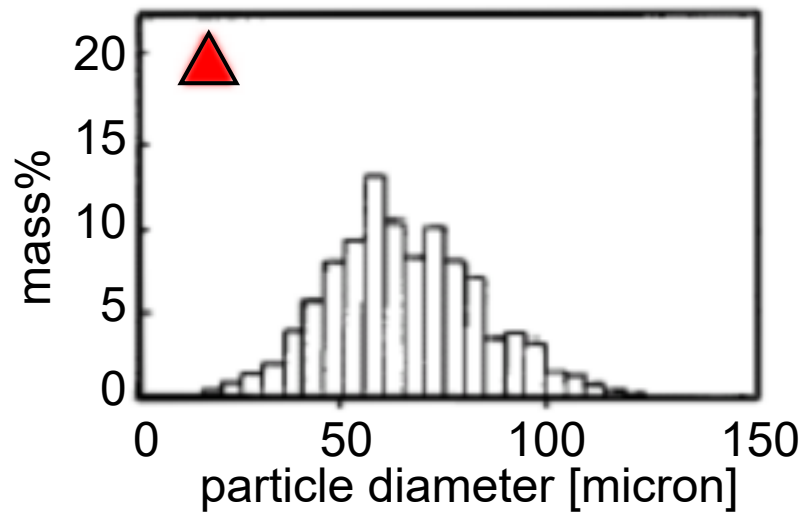
Adapted from Sun & Grace (1990)
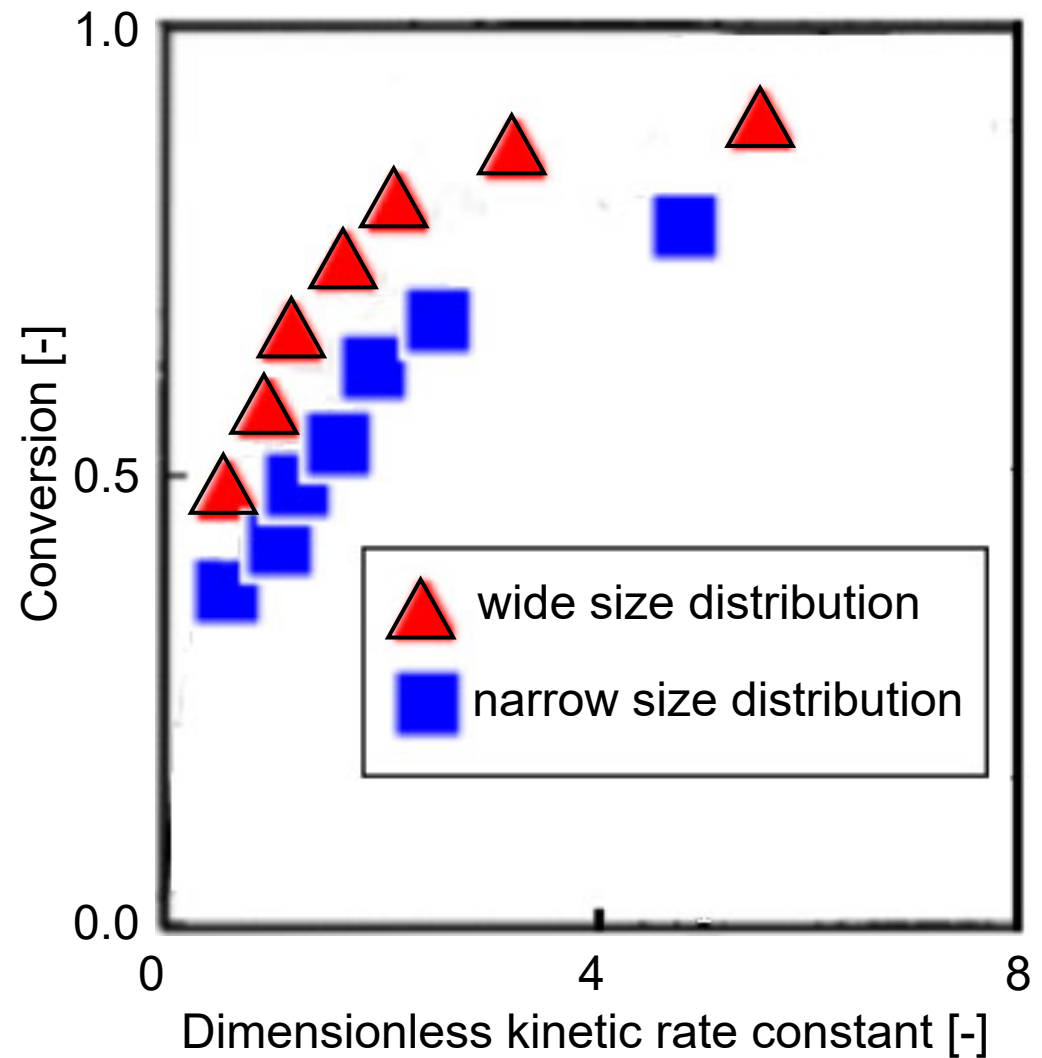
Implication of U_{mb}/U_{mf}

- 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 $U_{mb}/U_{mf}=1$

TU Delft

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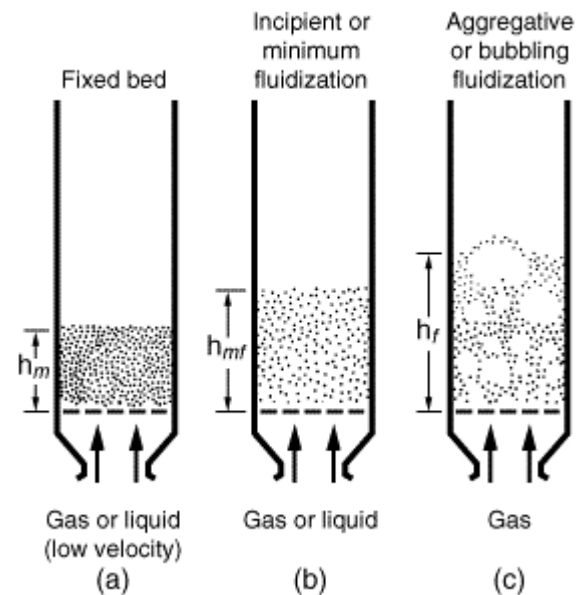
$$\left(d_p^*\right)_{AB} = 101 \left(\frac{\rho_p - \rho}{\rho}\right)^{-0.425}$$

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

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

Demarcation between Group B and D powders

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

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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.
($F_H = 8.76 \times 10^{-8}$ 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 $U_B \leq \dfrac{U_{mf}}{\varepsilon_{mf}}$
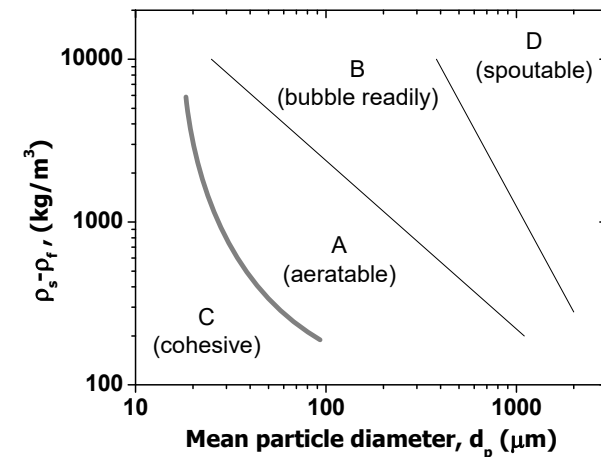
Characteristics of single bubble: Slow vs. fast bubbles (Davidson model)

slow bubble

fast bubble

Bubble

Cloud · Bubble

Gas Streamlines

Gas Streamlines

Gas Streamlines
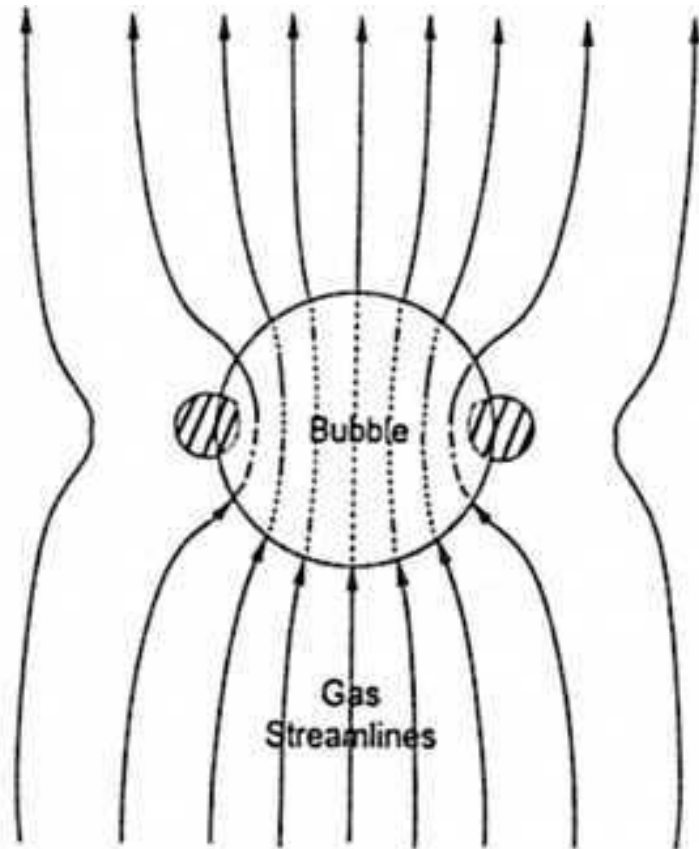
Slow Bubbles $U_B < U_{mf} / \varepsilon_{mf}$
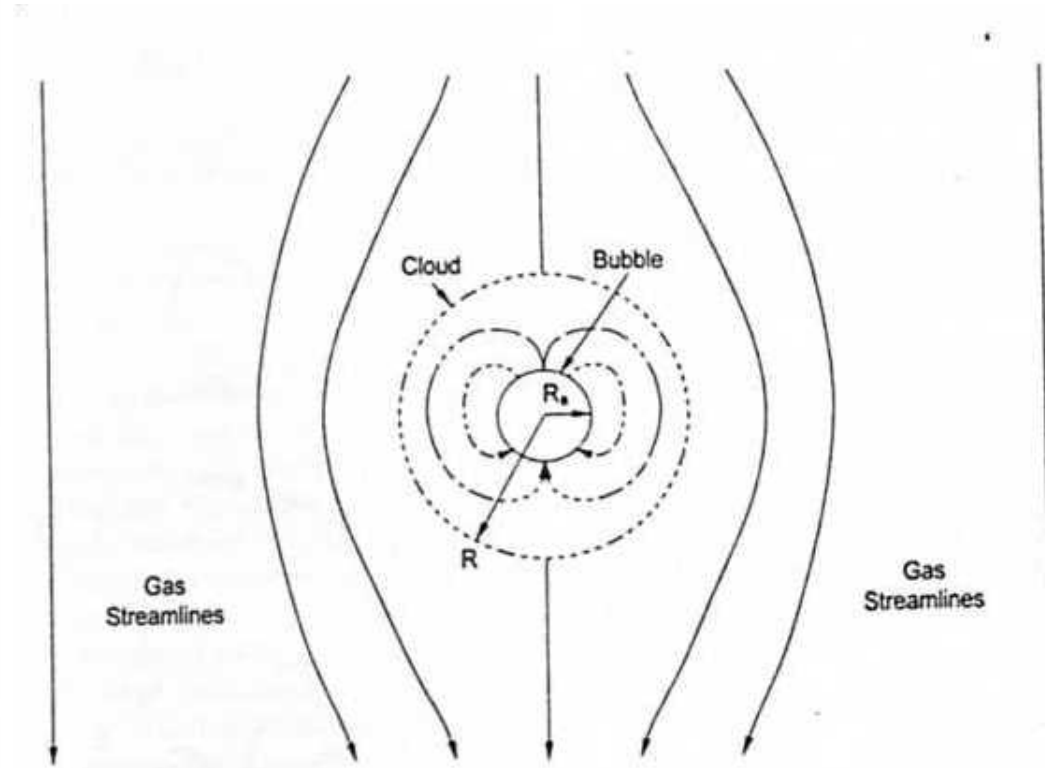
Fast Bubbles $U_B > U_{mf} / \varepsilon_{mf}$

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

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

Correlations for U_mb

Abrahamsen and Geldart (1980)

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

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

gas

gas

gas

gas

gas

gas

gas

solids returns

solids returns

solids returns
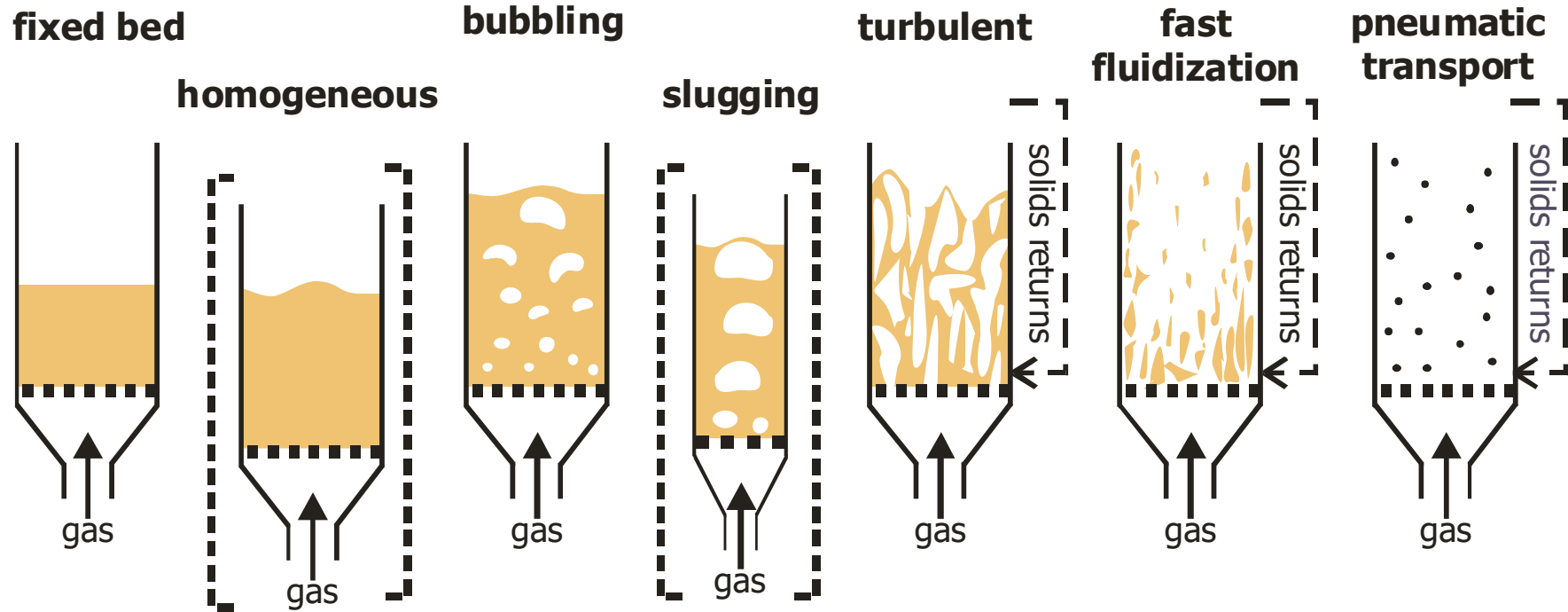
only A powders
at low gas velocity

only narrow beds

gas velocity

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Flow Regimes

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

Regime Transition Flow Chart
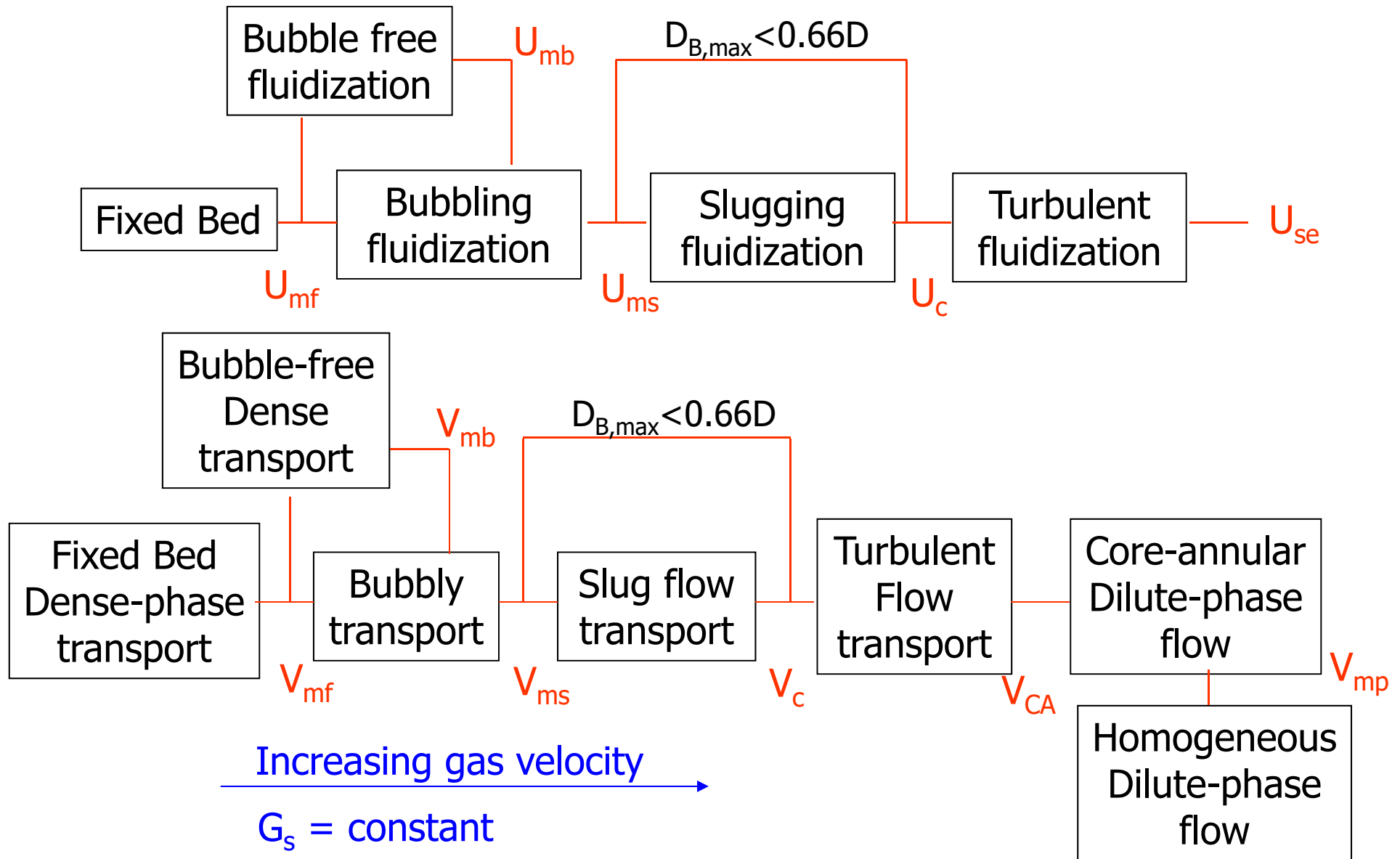
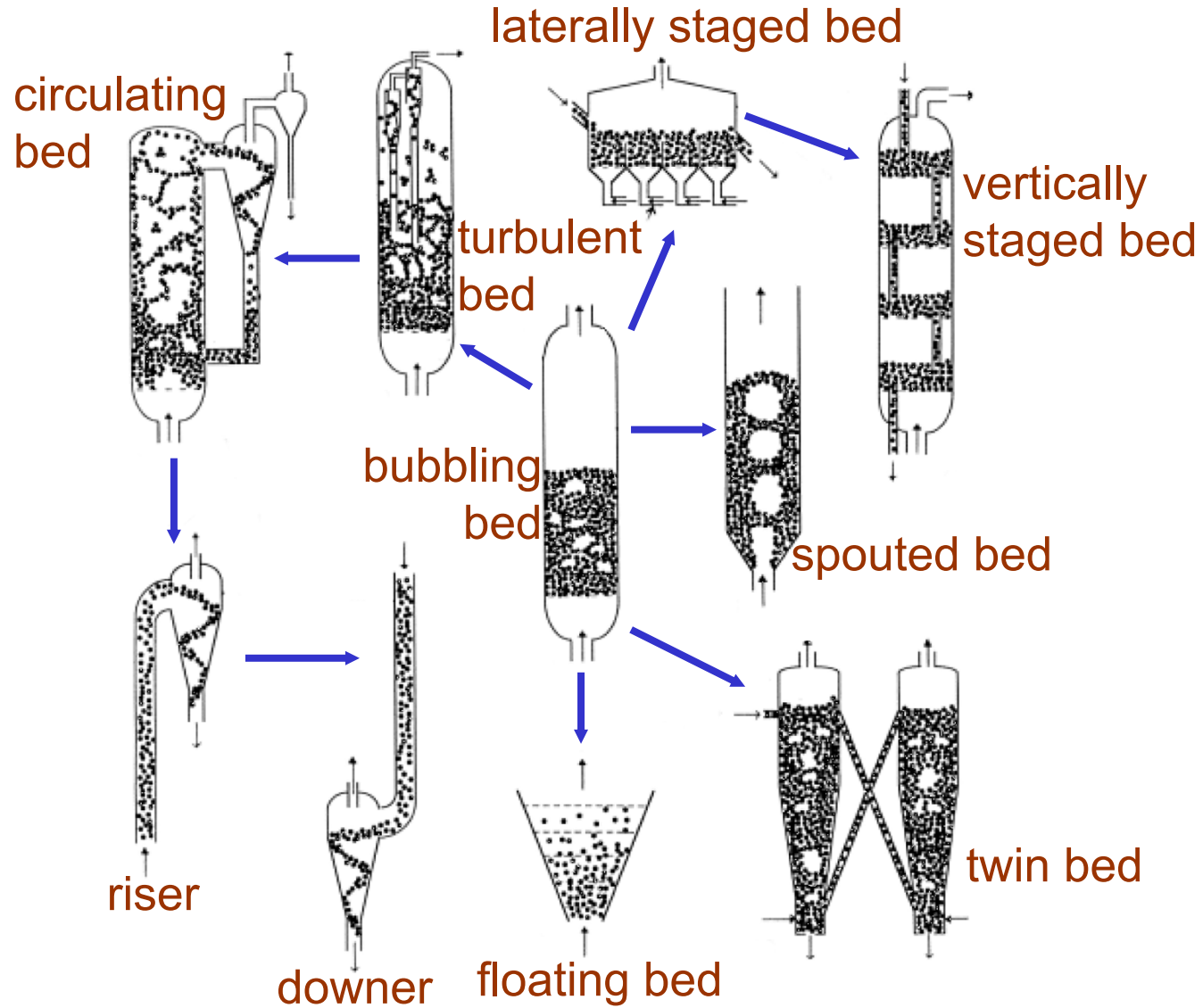

Bi & Grace, 1995
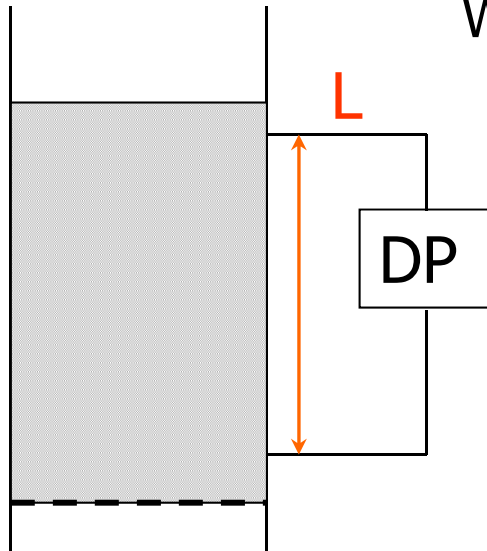
Unifying Fluidization Regime Diagram

Fluidized bed lay-out

circulating bed

laterally staged bed

turbulent bed

vertically staged bed

bubbling bed

spouted bed

riser

downer

floating bed

twin bed
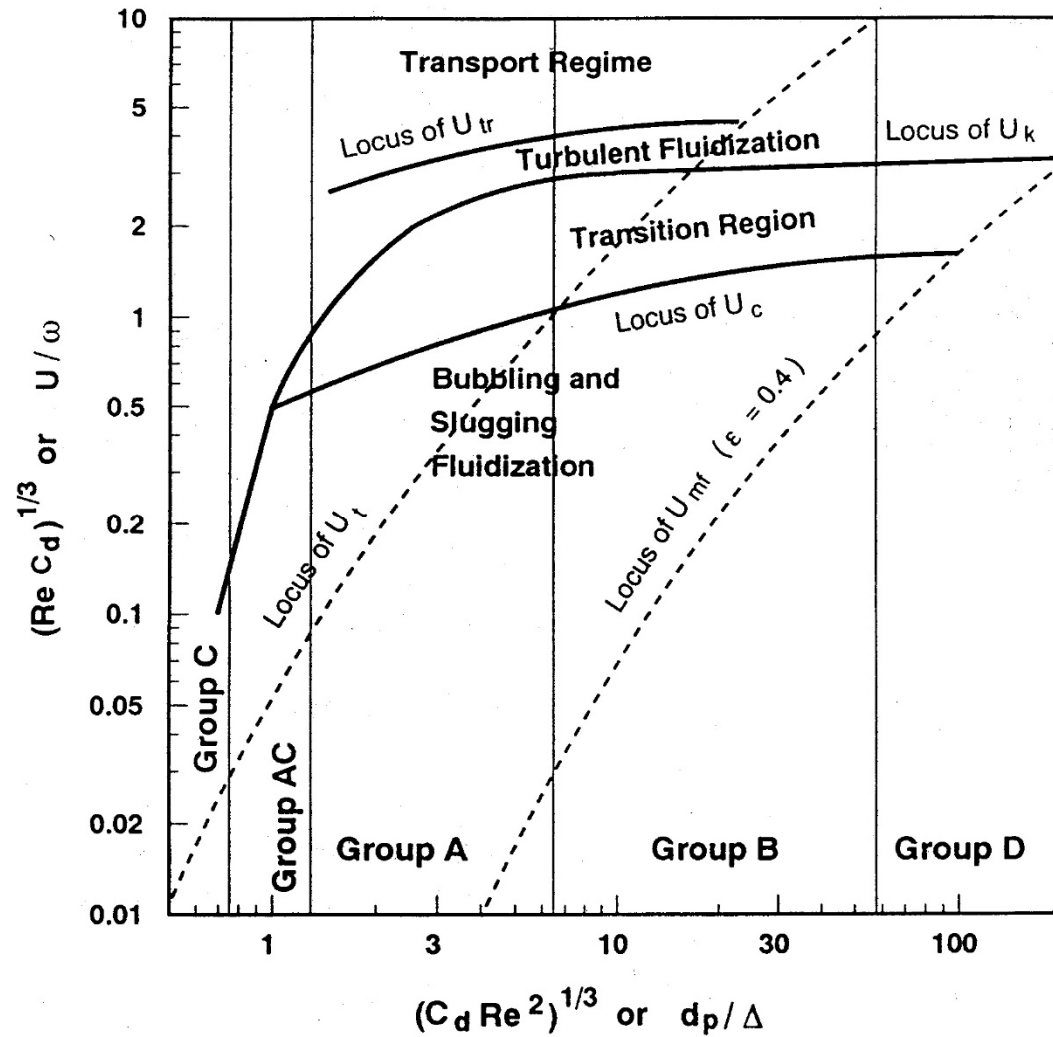
Static Head of Solids

When acceleration, friction and gas head are negligible

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

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

L

DP

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

Fluidized Non-bubbling region

Fluidized Bubbling Region

Packed bed

$\Delta P/L$

U_{mf} U_{mb}

U, m/s

Group B particles

narrow psd

Fluidized Bubbling Region

Packed bed

Wide psd

$\Delta P/L$

U_{mf} U_{cf}
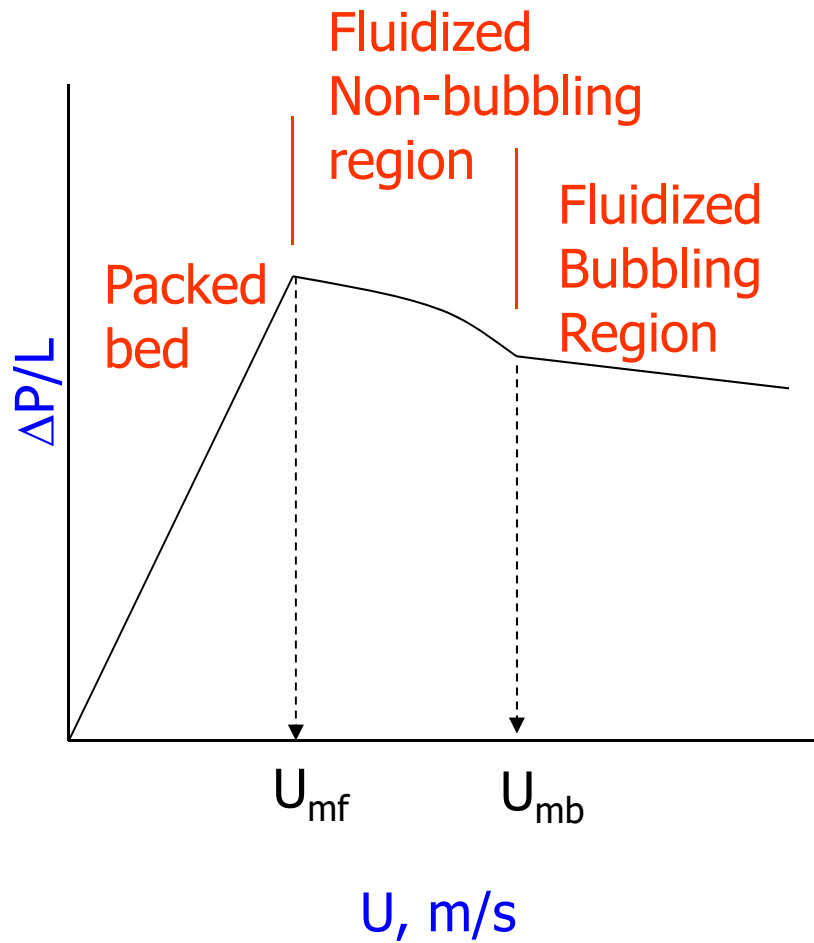
U, m/s
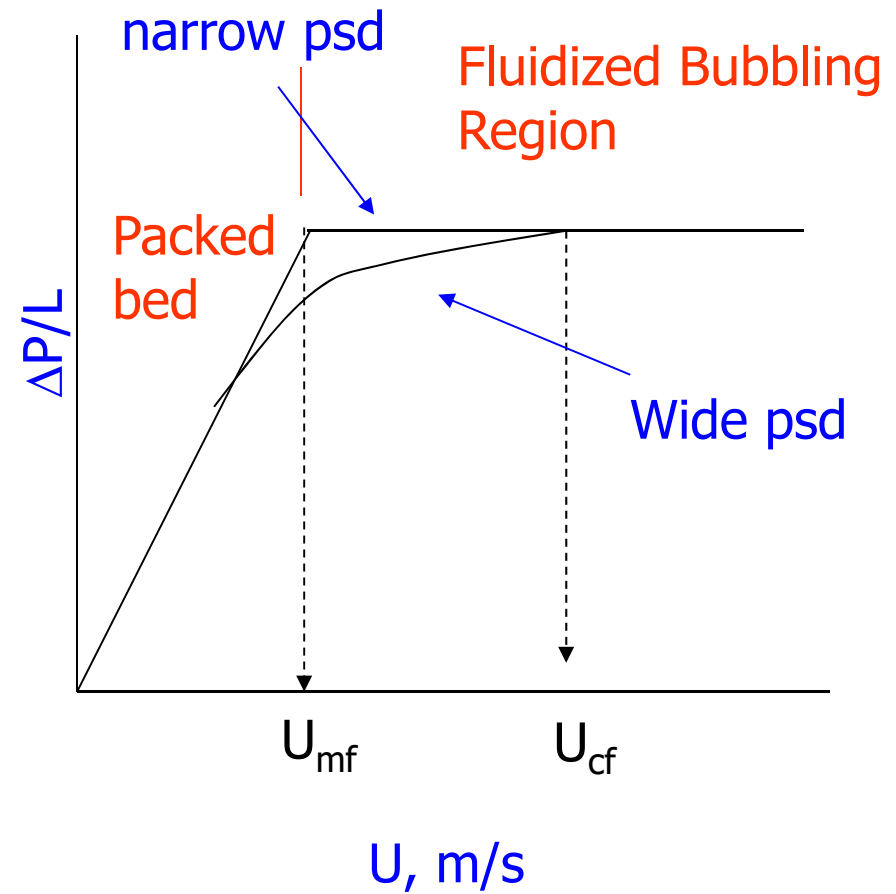
Minimum Fluidization

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

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

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

$$C_1 = 300\left(1 - \varepsilon_{mf}\right)/7 \qquad\qquad C_2 = \varepsilon_{mf}^3/1.75$$

Minimum Fluidization

Estimation of ε_{mf} $0.4 < \varepsilon_{mf} < 0.55$

- $\varepsilon_{mf} \approx \varepsilon_{fixed\ bed}$
- $\varepsilon_{mf} \approx (14\ \phi)^{-1/3}$ 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)

effect of wall

Initial bubble size
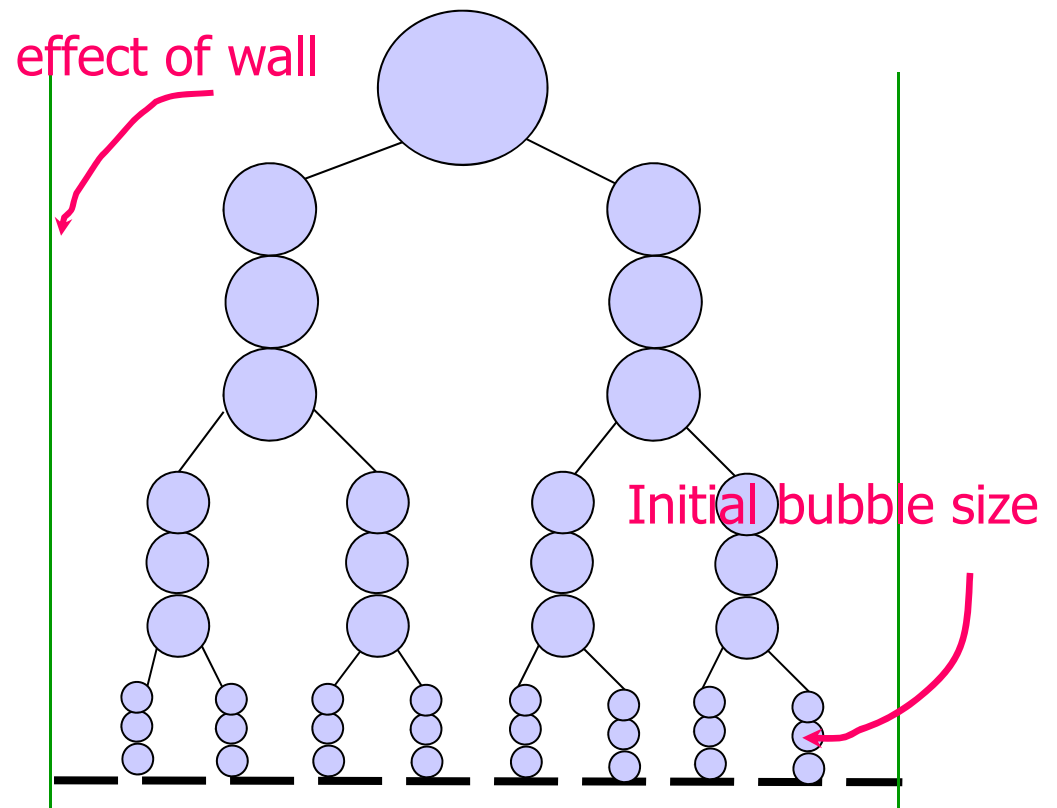
Freely Bubbling Beds: Bubble Size

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:

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

$60 \le d_p \le 450 \, \mu m$

$U - U_{mf} \le 0.48 \, m/s$
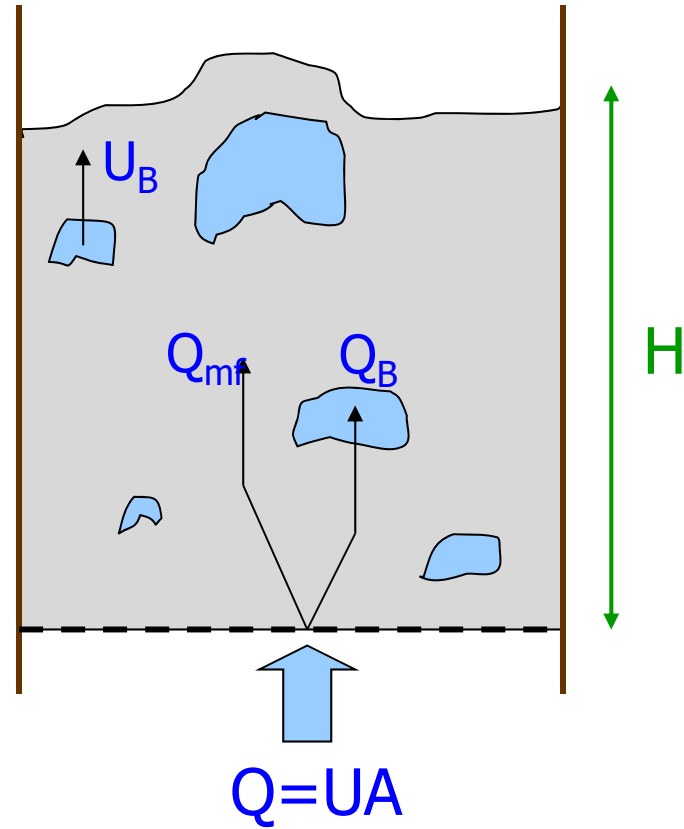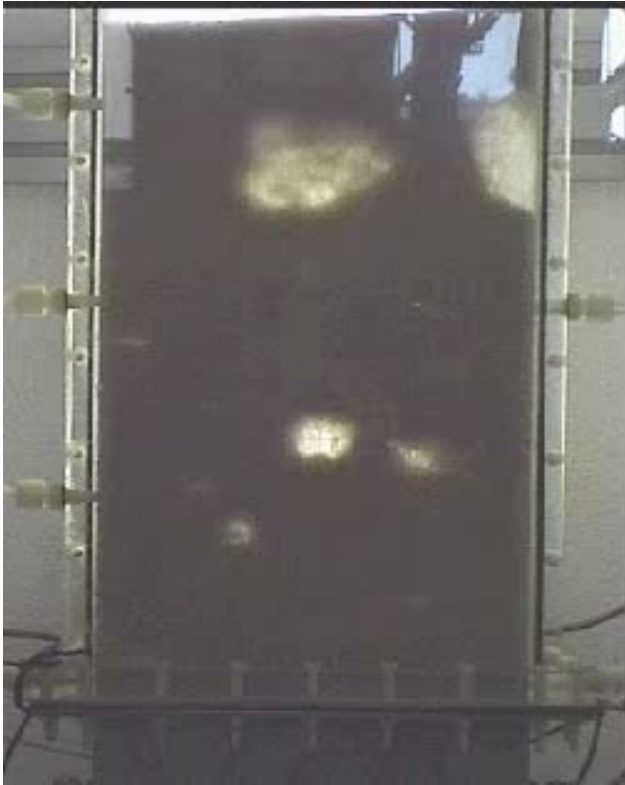
$D_T \le 1.3 \, m$

where:

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

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

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

Two-phase theory

U_B

Q_{mf} Q_B

H

$Q=UA$

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:

$$\delta_b = \frac{H - H_{mf}}{H} = \frac{Q - Q_{mf}}{A U_B} = \frac{U - U_{mf}}{U_B}$$

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

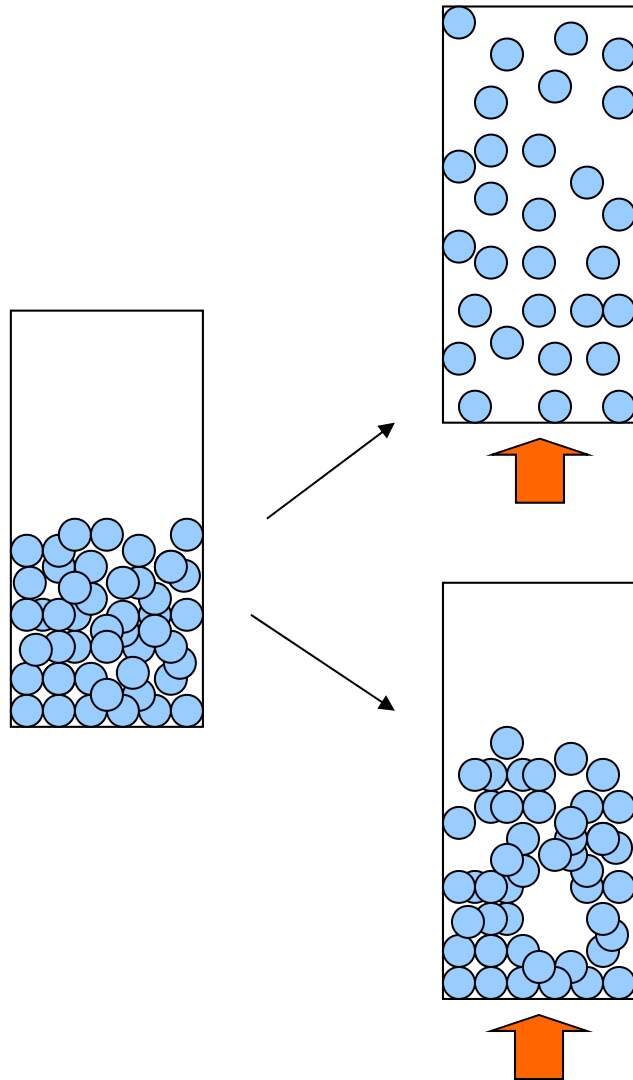where $0.6 < Y < 0.8$ for Group B powders

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

Baeyens and Geldart (1985)

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

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

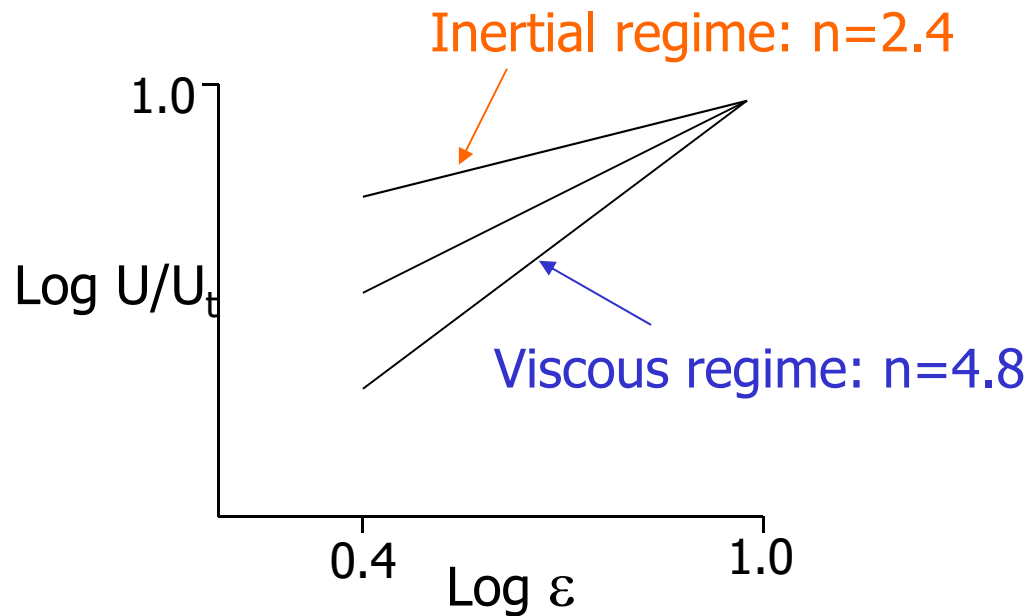# Steady-state expansion of fluidized beds

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

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

V_B: total volume of particles

Homogeneous bed expansion: Richardson-Zaki relation

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

Inertial regime: n=2.4

mass of paticles in the bed :

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

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

Viscous regime: n=4.8

Log U/U$_t$

0.4   Log ε 1.0

1.0

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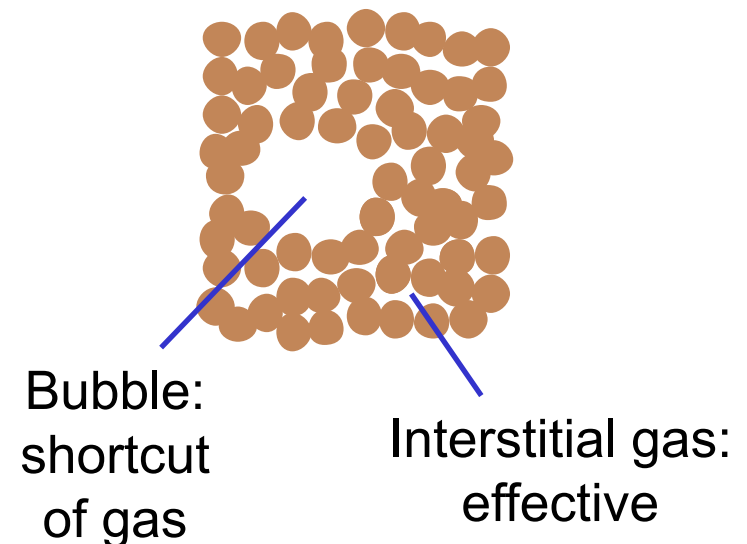
Heat transfer: particle to wall or internal
Mass transfer: gas to particle

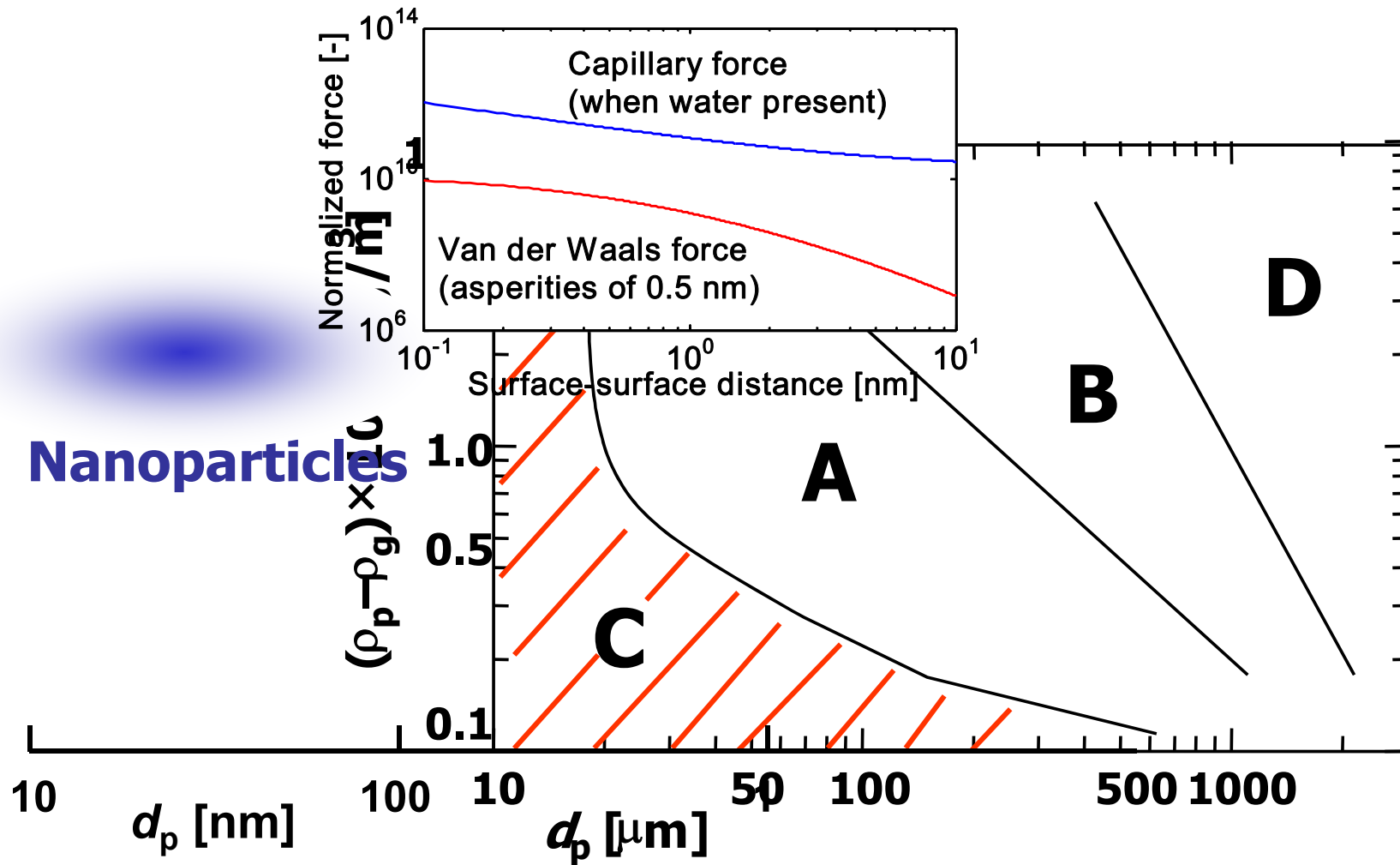Fluidized beds show an excellent heat transfer

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

However, large bubbles
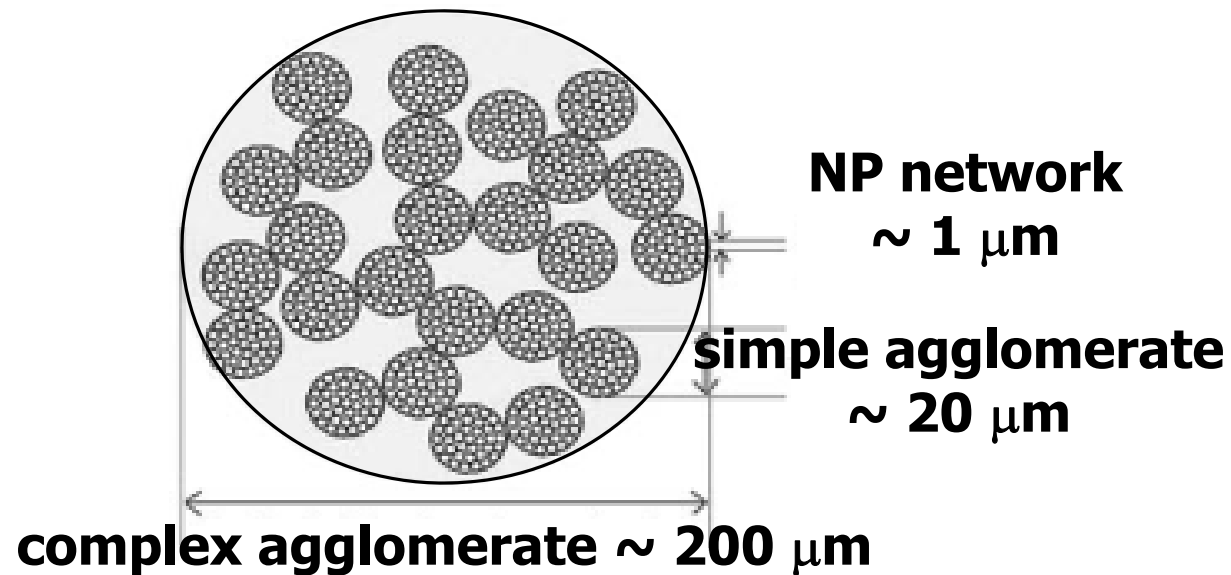decrease the mass transfer
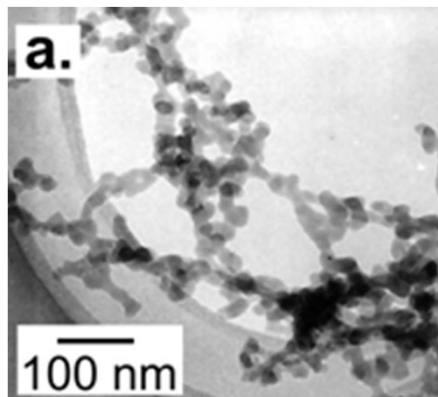
Research → decrease bubble size

Bubble:
shortcut
of gas

Interstitial gas:
effective

Geldart's Fluidization Map still valid?

Nanoparticles

Inset: Normalized force [-] vs Surface-surface distance [nm]

Capillary force (when water present)

Van der Waals force (asperities of 0.5 nm)

Main plot axes: $(\rho_p - \rho_g) \times \times$ [g/m^3] vs d_p [nm] and d_p [μm]

Regions: A, B, C, D

Nanoparticles are fluidized as agglomerates!

NP network ~ 1 μm

simple agglomerate ~ 20 μm

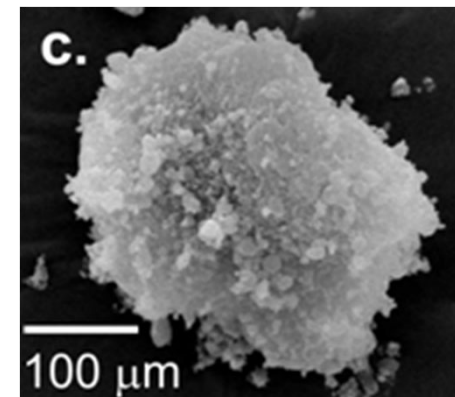**complex agglomerate ~ 200 μm**

Wang et al., Powder Technol. 124 (2002) 152:
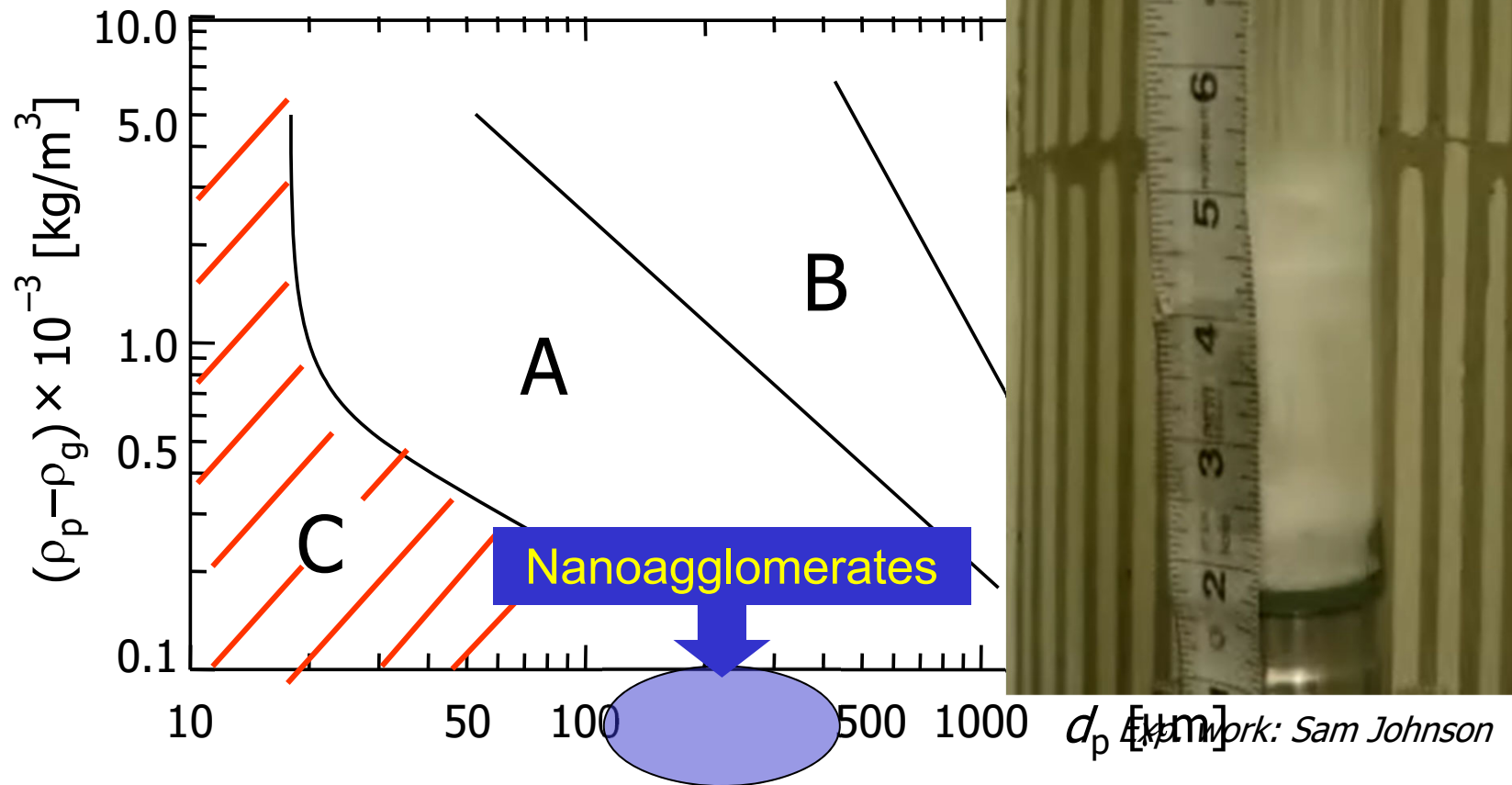

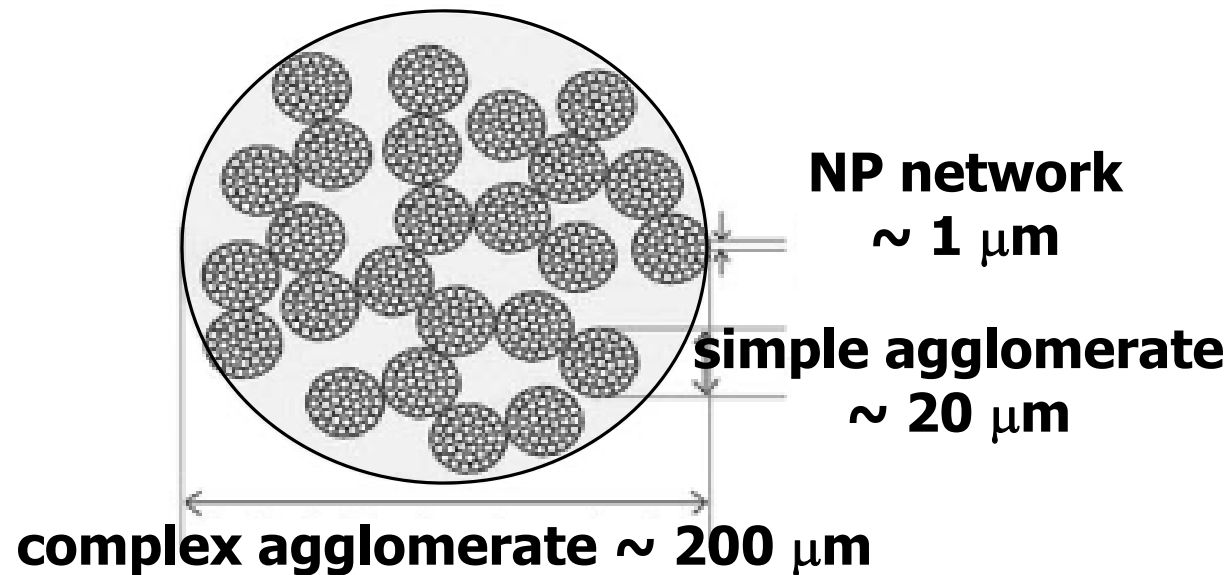TEM NP network

SEM simple agglomerate

SEM complex agglomerate

TUDelft

Geldart's Fluidization Map still valid?

- Primary particle size – 10-100 nm
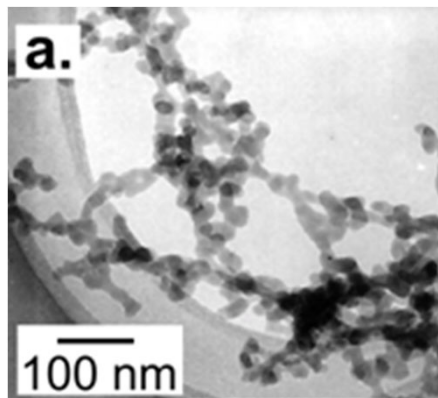- Agglomerate size – 100-400 μm
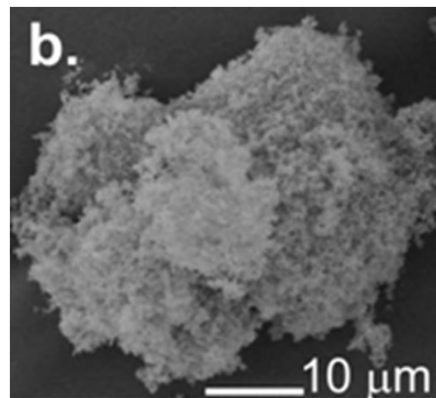- Agglomerate density – 20-120 kg/m^3

25 nm TiO_2 particles

Nanoagglomerates

d_p [μm]

Work: Sam Johnson

Nanoparticles are fluidized as agglomerates

NP network
~ 1 μm

simple agglomerate
~ 20 μm

complex agglomerate ~ 200 μm

Wang et al., Powder Technol. 124 (2002) 152:

100 nm

TEM NP network

10 μm

SEM simple agglomerate

100 μm

SEM complex agglomerate
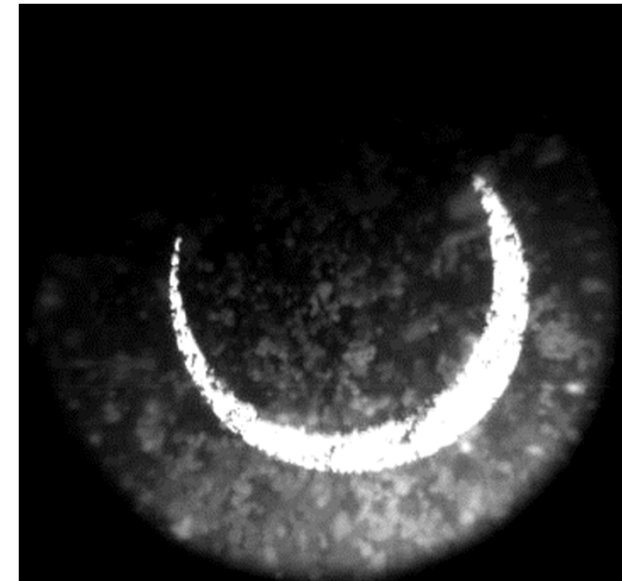
In-situ movies of nanoparticle agglomerates

High-speed camera with boroscope, slowed down 70x
Primary particles:
SiO_2, 10-20 nm diameter

splash zone

bed

U_0=4 cm/s

4 mm

De Martín, Chem Eng Sci 112 (2014) 79
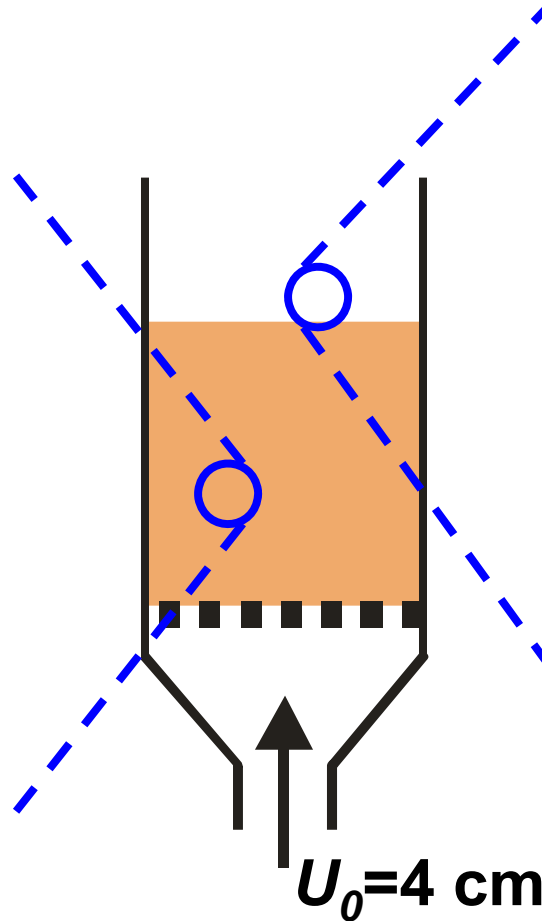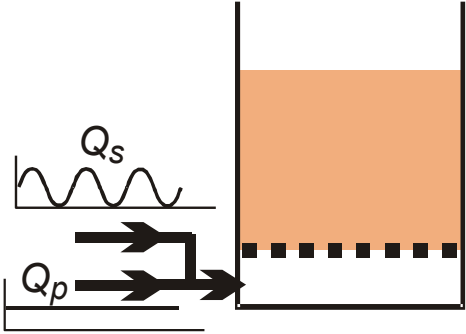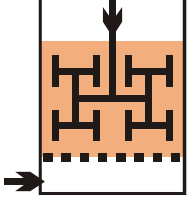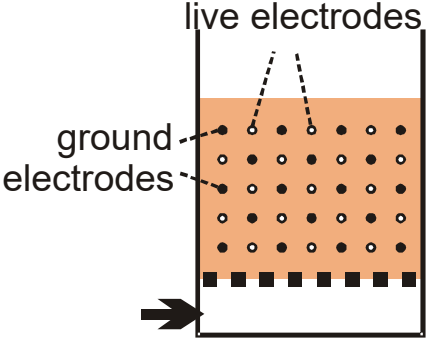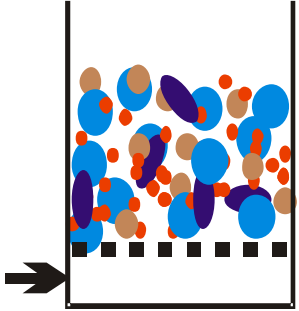
Ways to structure a fluidized bed

	Dynamics	**Geometry**
Gas	Q_s Q_p	
Particles	live electrodes, ground electrodes	

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

Tailoring the particle size distribution

mass%

20
15
10
5
0

0   50   100   150
particle diameter [micron]

wide size distribution

mass%

20
15
10
5
0

narrow size distribution

Conversion [-]

1.0

0.5

0.0

100

Particle diameter [μm]

O₃ decomposition

O_3 decomposition

4   8
Dimensionless kinetic rate constant [-]
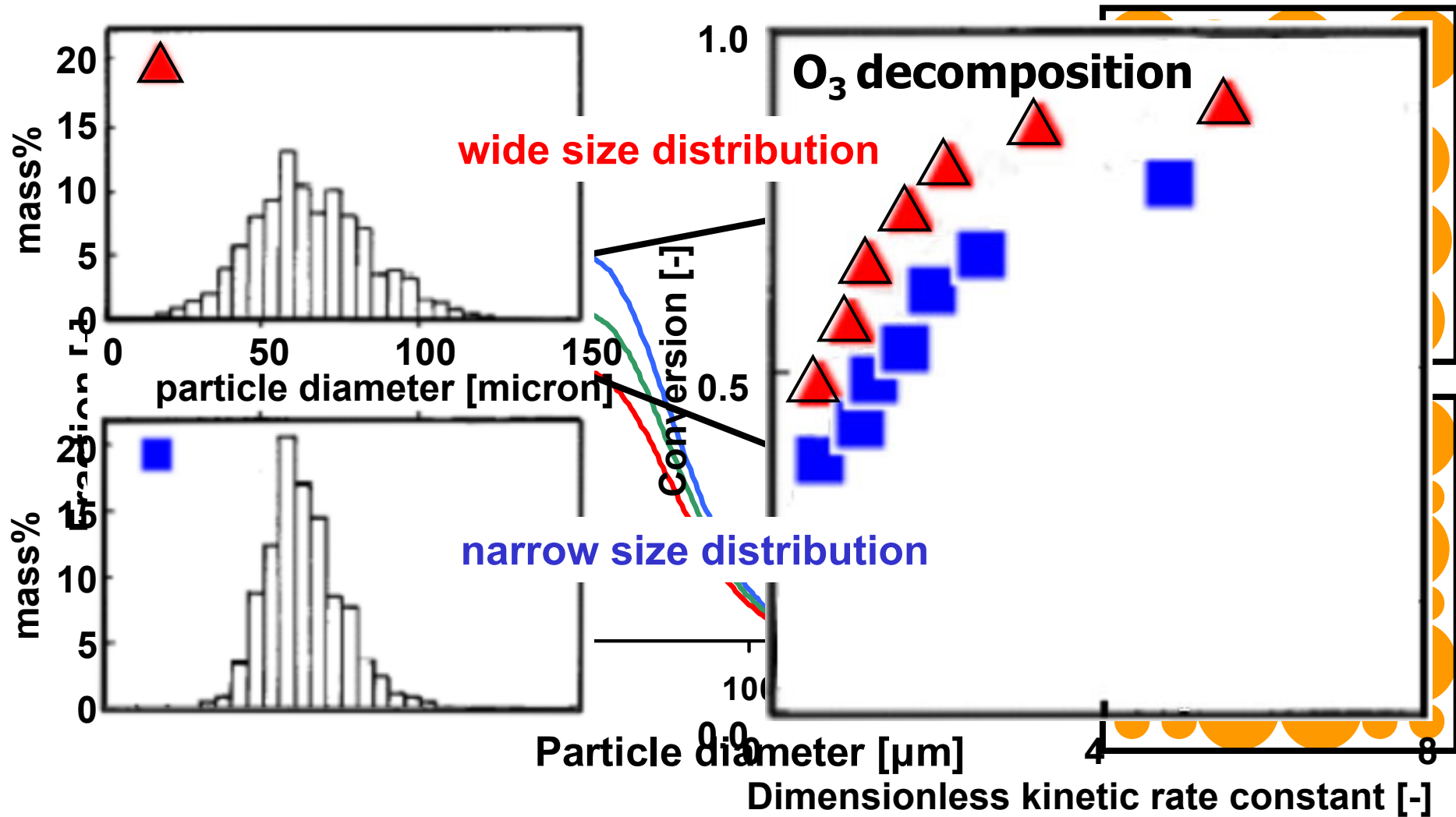
Fines effect on bubble size

Fines are added to a powder with a D_{50} of 70 μm
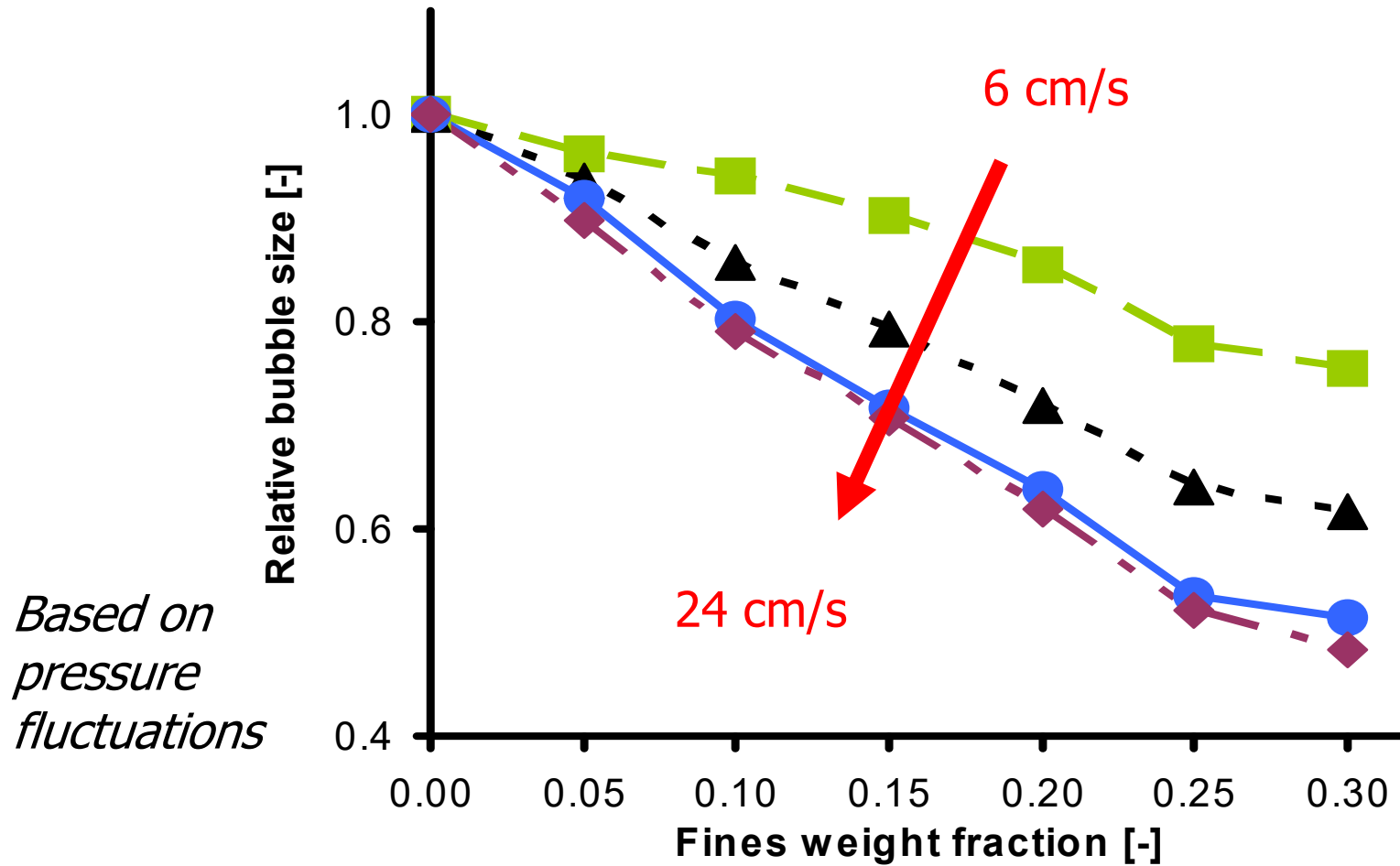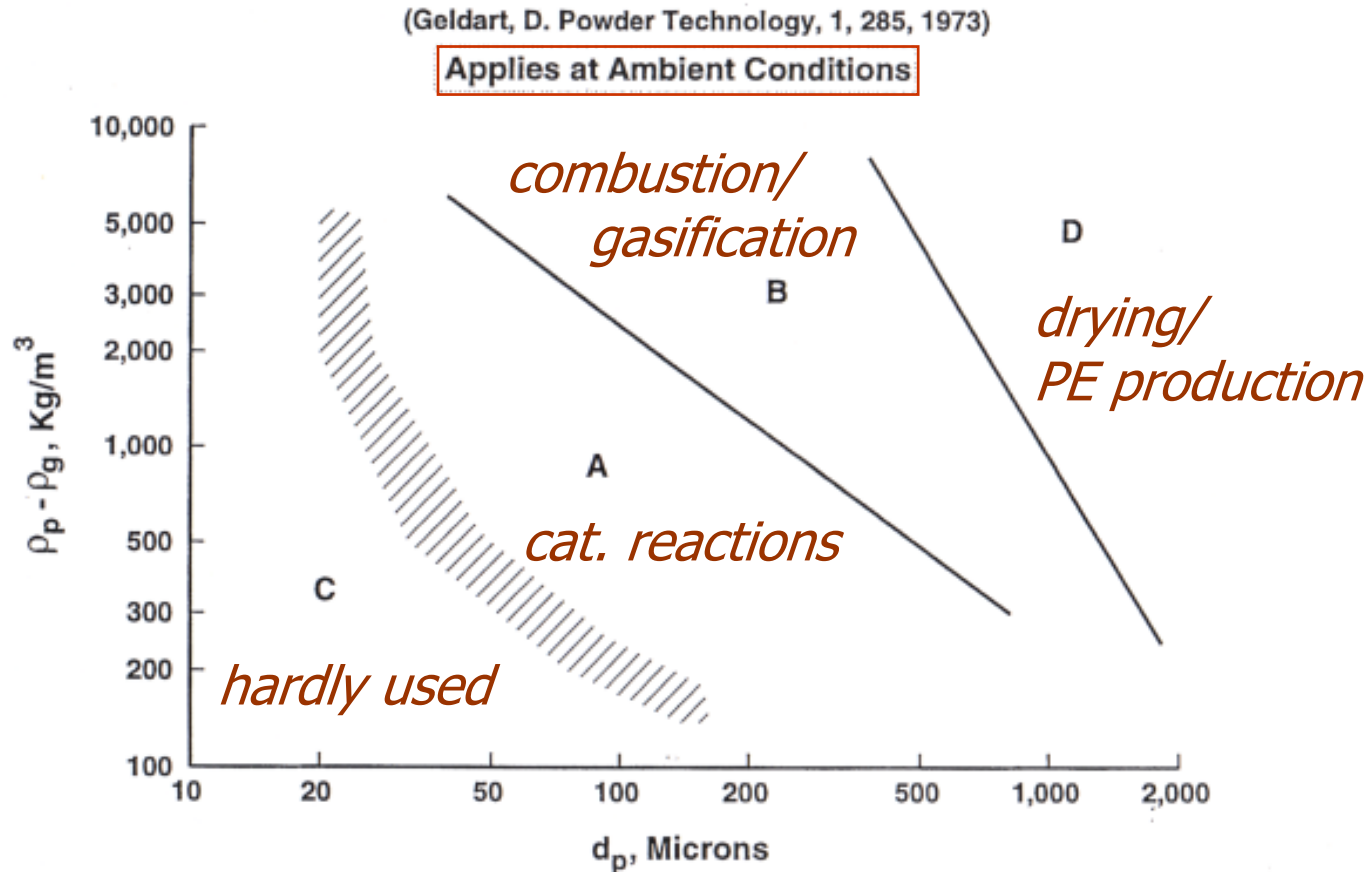
Based on pressure fluctuations

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

Geldart's powder classification

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

Applies at Ambient Conditions

combustion/ gasification

drying/ PE production

cat. reactions

hardly used

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)*

Textbooks
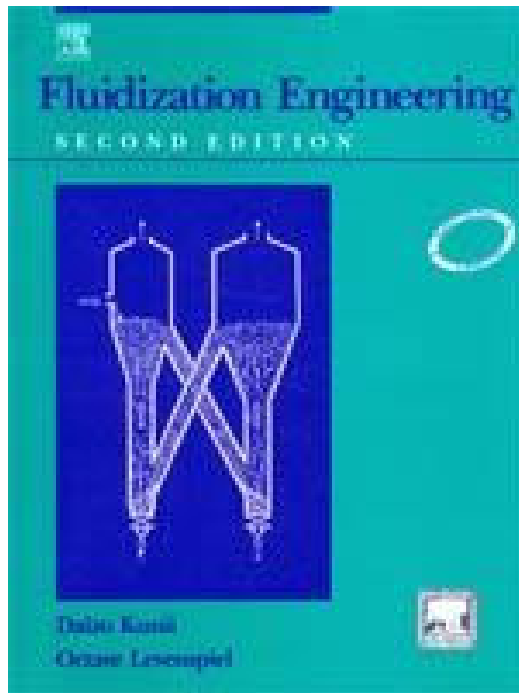
Fluidization Engineering
Kunii, D. & Levenspiel, O.
ISBN: 8131200353
Pub. Date: Jan 2005 , 2nd ed.
Publisher: Elsevier
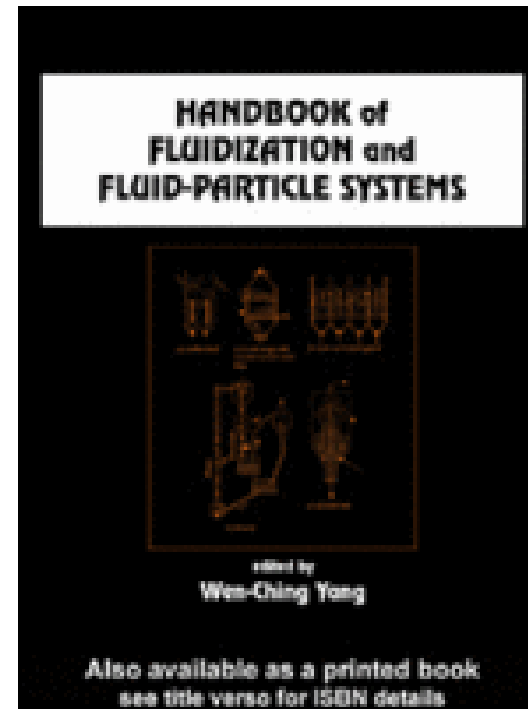
Handbook of Fluidization and Fluid-
Particle Systems
Ed. Wen-Ching Yang
ISBN: 978-0-8247-0259-5
Pub. Date: March 2003
Publisher: Routledge, USA

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.
  - Perry, R.H. and D.W. Green, Perrry's Chemical Engineers' Handbook, 7th Ed., McGraw-Hill, New York, 1997.
  - Fan, L.S and C. Zhu, "Principles of gas-solid flows", Cambridge Press, Cambridge, 1998.
  - Grace, J.R., A. Avidan and T.M. Knowlton, "Circulating Fluidized Beds," Blackie Academic press, Boston, 1997.
  - Kunii, D. and O. Levenspiel, "Fluidization Engineering", 2nd edition, Butterworth-Heinemann, Boston, 1991.
  - Geldart, D., "Gas Fluidization Technology", John Wiley & Sons, New York, 1986.

References

- Rhodes, M., and Zakhari, A., "Laboratory Demonstrations in Particle Technology", CD, Monash University, Melbourne, Australia (1999).
- Geldart, D., "Types of Gas Fluidization", Powder Tech., 7, 285-292 (1973).
- Yang, W.-C., "Bubbling Fluidized Beds", in *Handbook of Fluidization and Fluid-Particle Systems*, Ed. Yang, W.-C., Marcel Dekker, N.Y., pp. 53-111 (2003).
- Ellis, N., "Hydrodynamics of Gas-Solid Turbulent Fluidized Beds", Ph.D. Dissertation, UBC (2003).
- Bi, H.T., and Grace, J.R., "Flow Regime Diagrams for Gas-Solid Fluidization and Upward Transport", Int. J. Multiphase Flow, 21, 1229-1236 (1995).
- Grace, J.R., "Contacting Modes and Behaviour Classification of Gas-Solid and Other Two-Phase Suspensions", Can. J. Chem. Eng., 64, 353-363 (1986).
- Gidaspow, D., "Multiphase Flow and Fluidization: Continuum and Kinetic Theory Descriptions", Academic Press, San Diego (1994).
- Kunii, D., and Levenspiel, O., "Fluidization Engineering", Butterworth-Heinemann, Newton, MA (1991).

References

- Bird, R.B., Stewart, W.E., and Lightfoot, E.N., "Transport Phenomena", John Wiley & Sons, N.Y., Chapter 6: Interphase Transport in Isothermal Systems (2002).
- Wen, C.Y., and Yu, Y.H., "Mechanics of Fluidization", Chem. Eng. Prog. Symp. Series, 62, 100 - 111 (1966).
- Gidapsow, D., and Ettehadieh, B., "Fluidization in Two-Dimensional Beds with a Jet, Part II: Hydrodynamics Modeling", I&EC Fundamentals, 22, 193-201 (1983).
- Didwania, A.K., "A New Instability Mode in Fluidized Beds", Europhysics Letters, 53, 478-482 (2001).
- DallaValle, J.M., "Micromeritics, the technology of fine particles", Pitman Pub., N.Y. (1948).
- McCabe W.L., Smith, J.C., and Harriot P., "Unit operations of Chemical Engineering", McGraw-Hill, New York (1993)
- Rhodes, M., "Introduction to Particle Technology", Wiley, Chichester (1998)
- Grace, J.R., "Fluidized-Bed Hydrodynamics", in *Handbook of Multiphase Systems*, Ed. Hetsroni, G., Hemisphere, Washington, pp., 8:5-64 (1982).