UNIVERSITEIT TWENTE.

Multiphase Reactor Technology
HC 5: Fluidization

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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

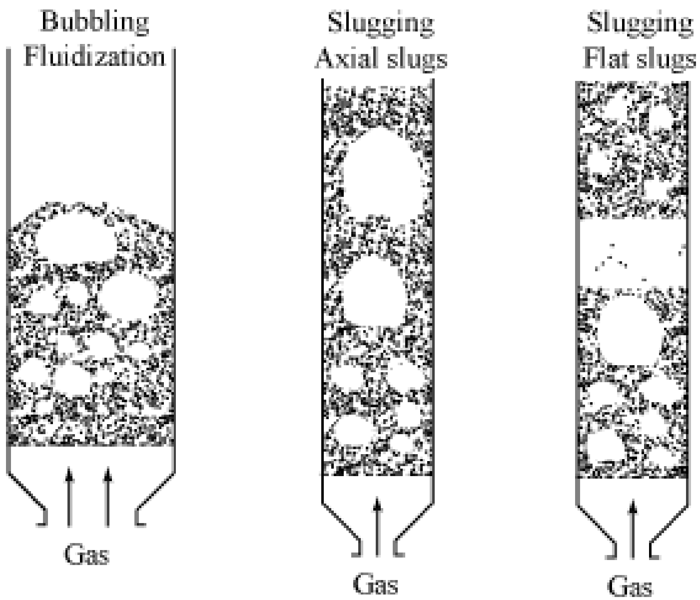
Contents

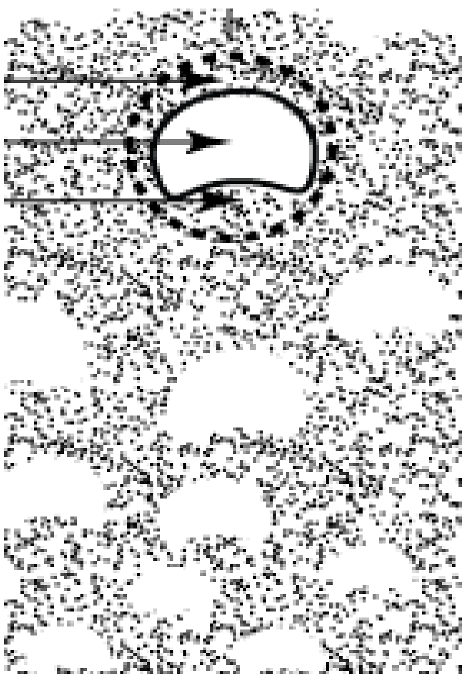
- ✓ Gas dispersion and gas exchange
- ✓ Particle-to-gas mass and heat transfer
- ✓ Heat transfer between fluidized beds and surfaces
- ✓ Conversion of gas due to catalytic reactions

- Recent developments

- ✓ Applications of fluidized bed chemical reactors
- ✓ Computational fluid dynamics (CFD) based modeling

Dense Fluidized Beds
Gas dispersion and gas exchange





Dense Fluidized Beds
Gas dispersion and gas exchange

- Relevance: especially for gas-solid contacting (catalytic reactions) gas dispersion and bubble-emulsion exchange of gas are very important

- Dispersion of gas in fluidized beds: both vertical and horizontal dispersion of gas occurs. General experimental finding: $D_{gv} > D_{gh}$

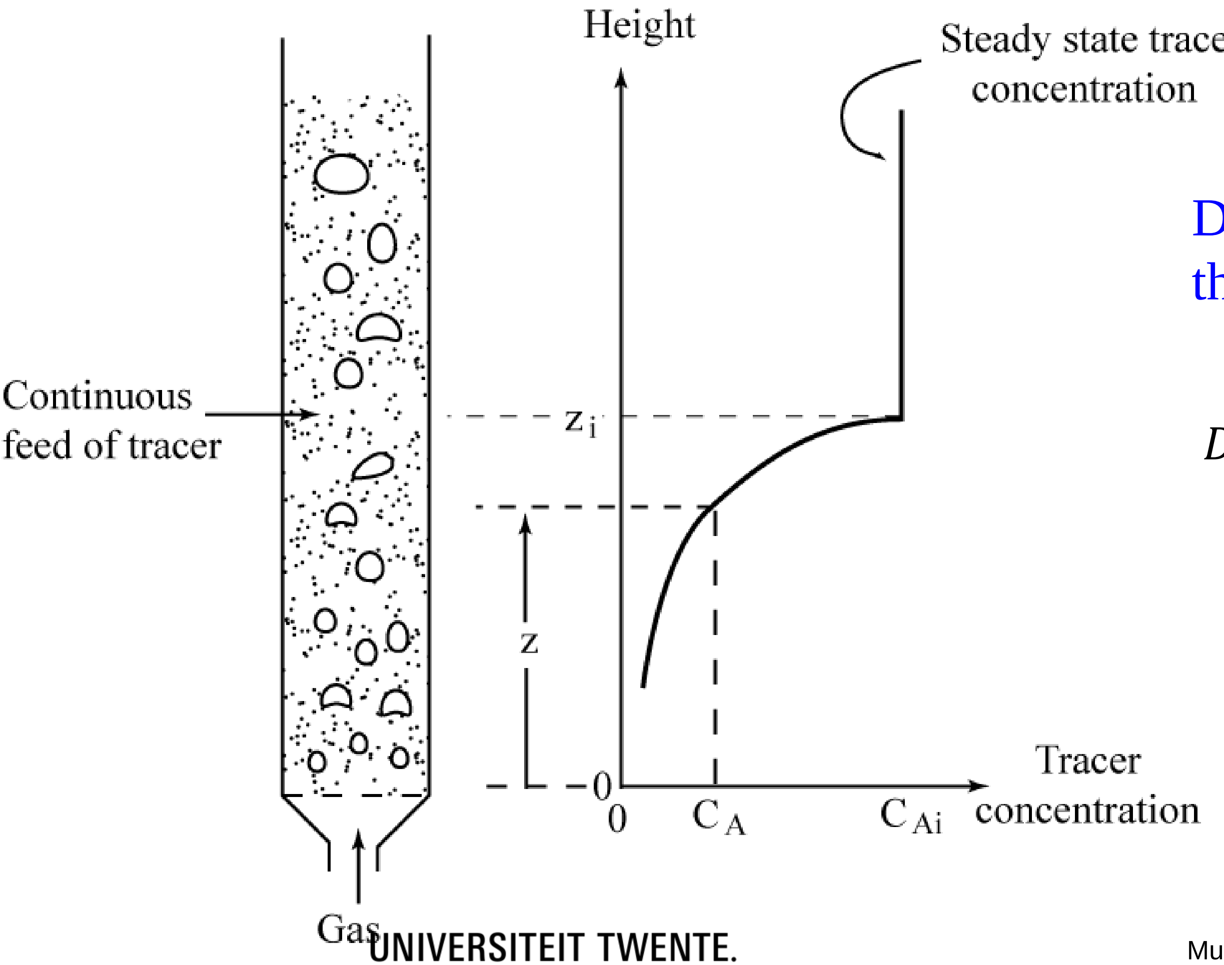
- Experimental techniques for dispersion coefficients:

from steady state gas tracer experiments + fitting experimental results to diffusion equation

from stimulus-reponse experiments + fitting experimental results to diffusion equation

Dense Fluidized Beds
Gas dispersion and gas exchange

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



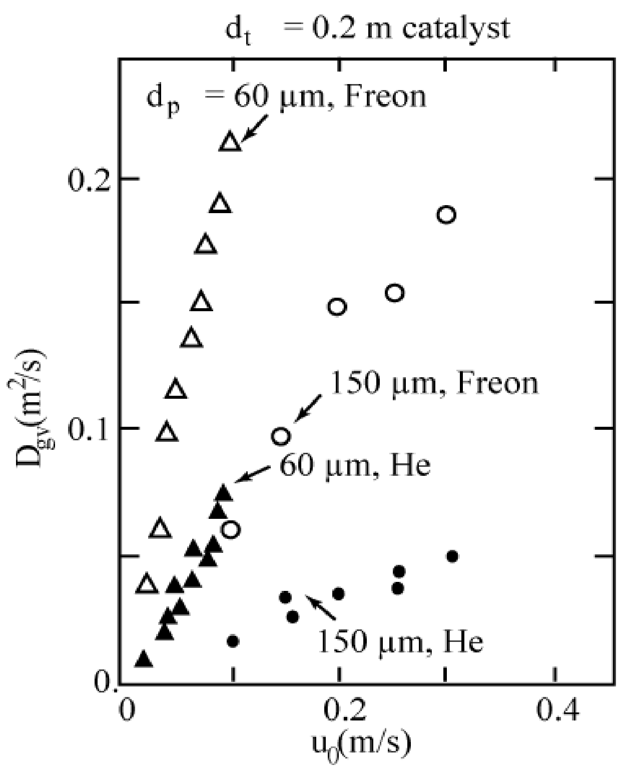
Differential equation governing the axial dispersion process

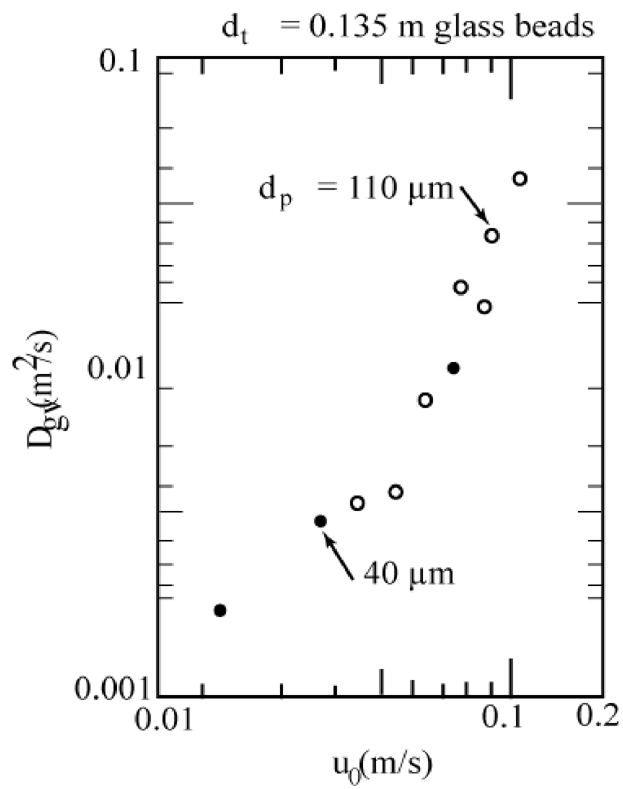
$$D_{gv}\frac{d^2C_A}{dz^2} - u_0\frac{dC_A}{dz} = 0$$

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

Dense Fluidized Beds
Gas dispersion and gas exchange

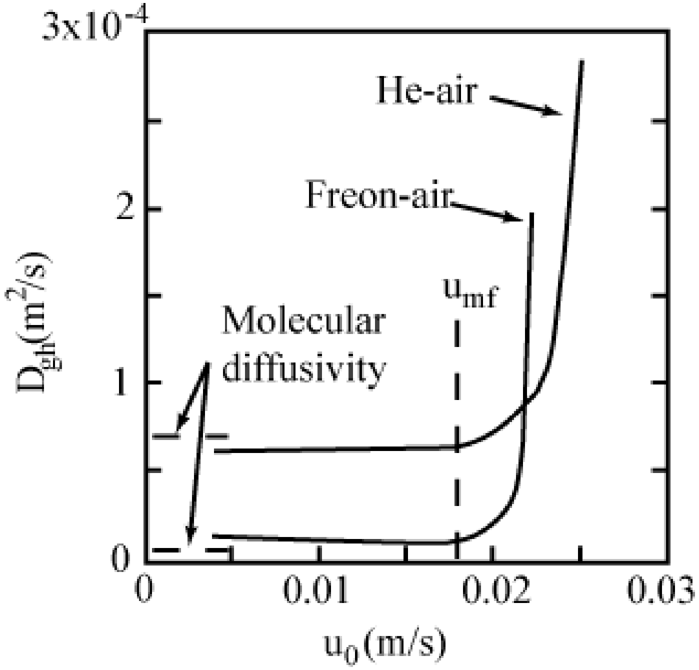
- Experimental results for D_{gv} (steady state backmixing (left) and stimulus response (right) experiments) for microspherical catalyst d_p=150 μm; triangular points for FCC catalyst 60 μm

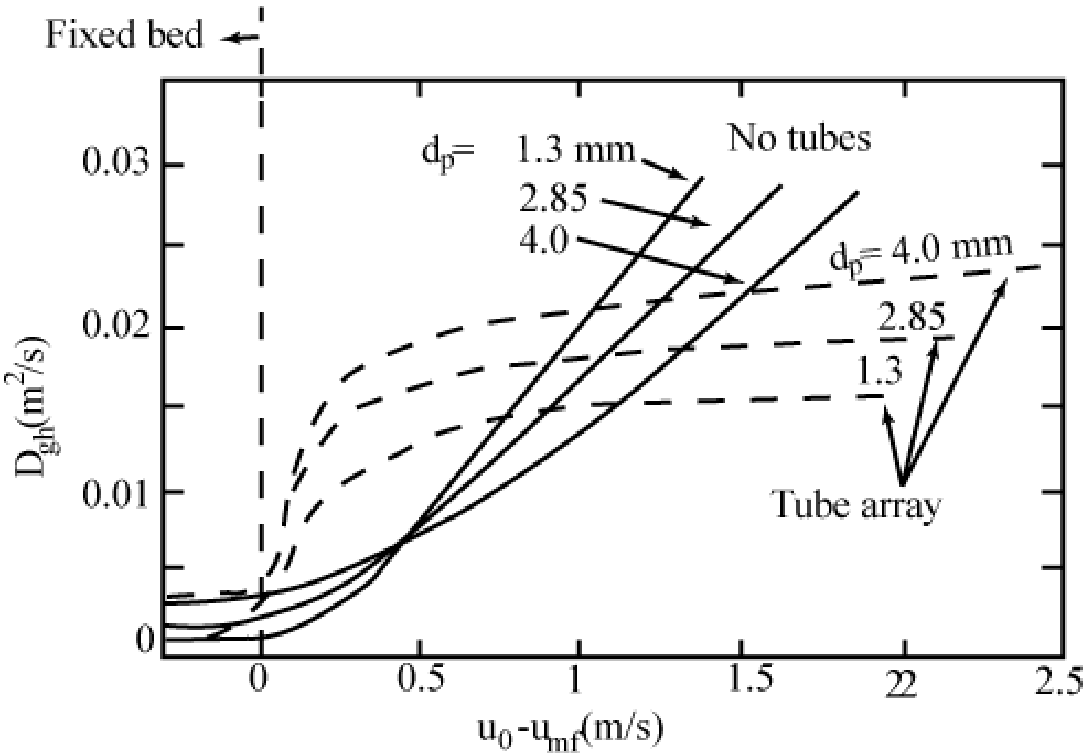




Dense Fluidized Beds
Gas dispersion and gas exchange

- Experimental results for D_{gh}: near u_{mf} in beds of fine solids (d_t=0.2 m, microspherical catalyst, d_p=150 μm), the effect of a tube array in beds of coarse particles, u_{mf}=0.73-1.83 m/s





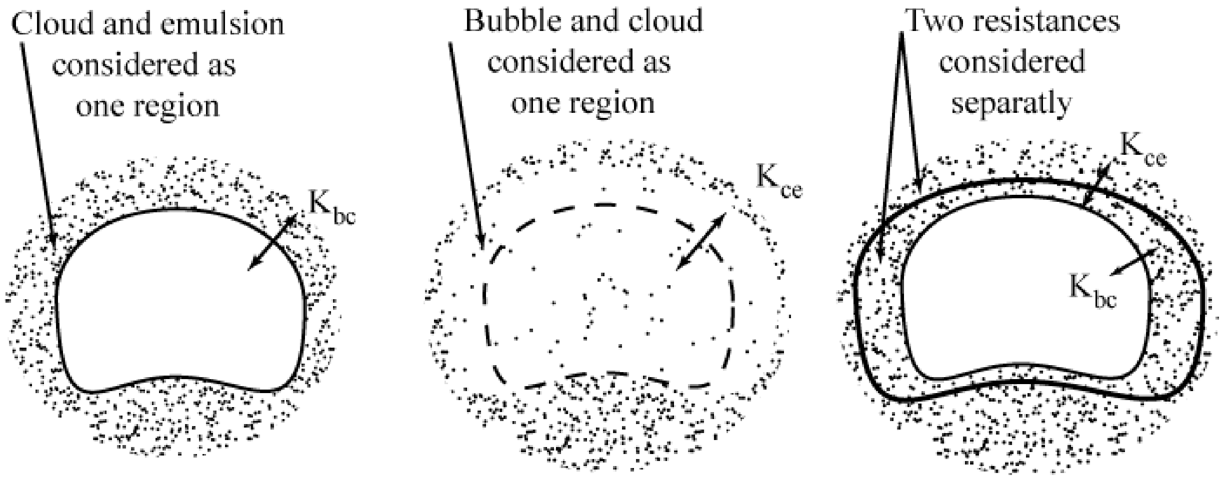
Dense Fluidized Beds
Gas dispersion and gas exchange

- Gas exchange between bubble and emulsion phase

  - Definitions of gas exchange:

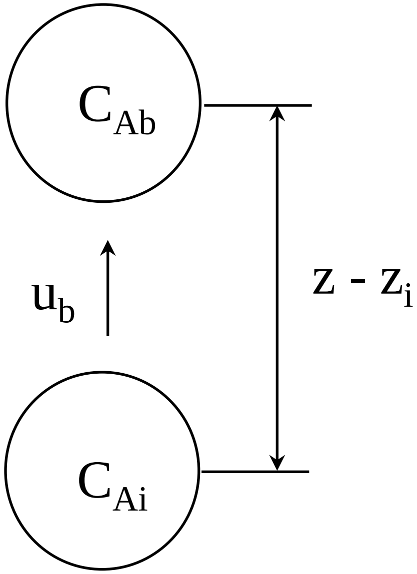
$$-\frac{1}{V_b}\frac{dN_{Ab}}{dt} = -u_b\frac{dC_{ab}}{dz} = K_{be}(C_{Ab}-C_{Ae}) = K_{bc}(C_{Ab}-C_{Ac}) = K_{ce}(C_{Ac}-C_{Ae})$$

  - Relationship between interchange coefficients: $$\frac{1}{K_{be}} = \frac{1}{K_{bc}} + \frac{1}{K_{ce}}$$



Dense Fluidized Beds
Gas dispersion and gas exchange

- Experimental methods for K_{be}: single bubble method (injection of a single bubble with tracer at z=zi and initial tracer concentration C_{Ai})



measurement of concentrations in the bubble at two z-levels yields K_{be} (bubble rise velocity known)

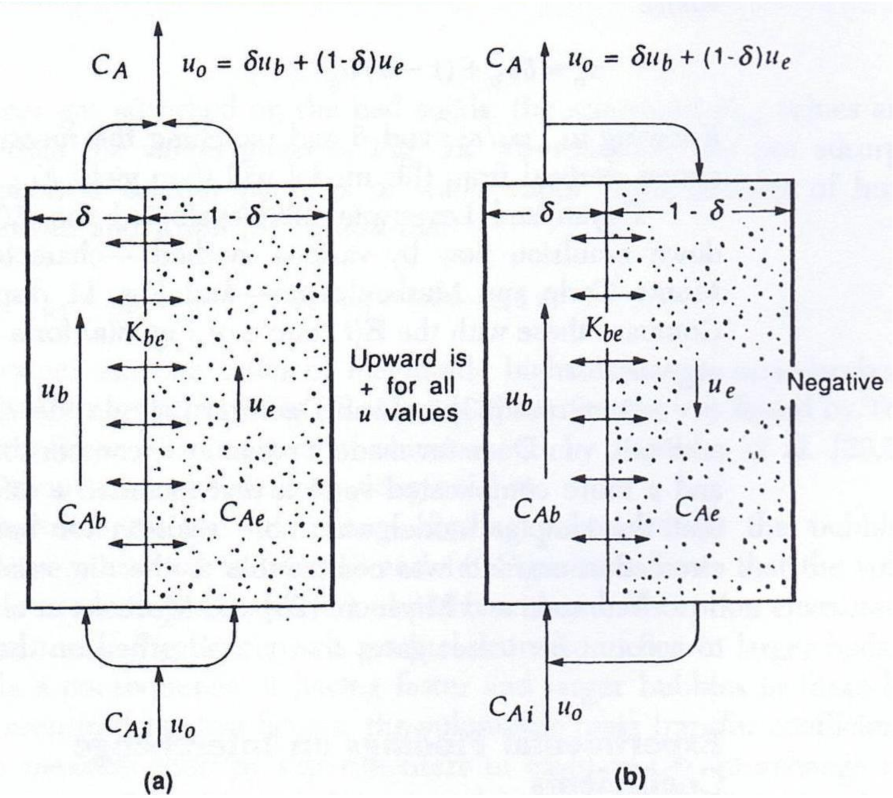
$$-u_b \frac{dC_{Ab}}{dz} = K_{be}(C_{Ab} - C_{Ae})$$

$$\ln(C_{Ab} - C_{Ae})\Big|_{zi}^{z} = -\frac{K_{be}(z - z_i)}{u_b}$$

$$\frac{C_{Ab} - C_{Ae}}{C_{Ai} - C_{Ae}} = exp\left[-\frac{K_{be}(z - z_i)}{u_b}\right]$$

Dense Fluidized Beds
Gas dispersion and gas exchange

- Experimental methods for Kbe
  - Bubbling bed method (pulse- or step-response of tracer + fitting experimental concentration profiles against two-zone model)



Dense Fluidized Beds
Gas dispersion and gas exchange

- 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)

Dense Fluidized Beds
Gas dispersion and gas exchange

- Estimation of gas exchange coefficients (Davidson 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

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Relevance: mass and/or heat transfer between fluidized particles and fluidizing agent occur frequently in a great variety of processes such as adsorption, drying, granulation, gas phase polymerization of C_2H_4

> accurate prediction of transfer rates of mass and/or heat is required for design purposes

- Topics covered:
  - ✓ mass transfer: experimental
  - ✓ interpretation of mass transfer coefficients
  - ✓ heat transfer: experimental
  - ✓ interpretation of heat transfer coefficients

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Mass transfer: experimental results for single spheres (Froessling):

$$Sh^* = \frac{k_d^* d_{sph} y}{D} = 2 + 0.6\left(\frac{\rho u_0 d_{sph}}{\mu}\right)^{\frac{1}{2}}\left(\frac{\mu}{\rho D}\right)^{\frac{1}{3}}$$
$$= 2 + 0.6\left(Re_{sph}\right)^{\frac{1}{2}}(Sc)^{\frac{1}{3}}$$

with y the logarithmic mean fraction of inert non-diffusing component

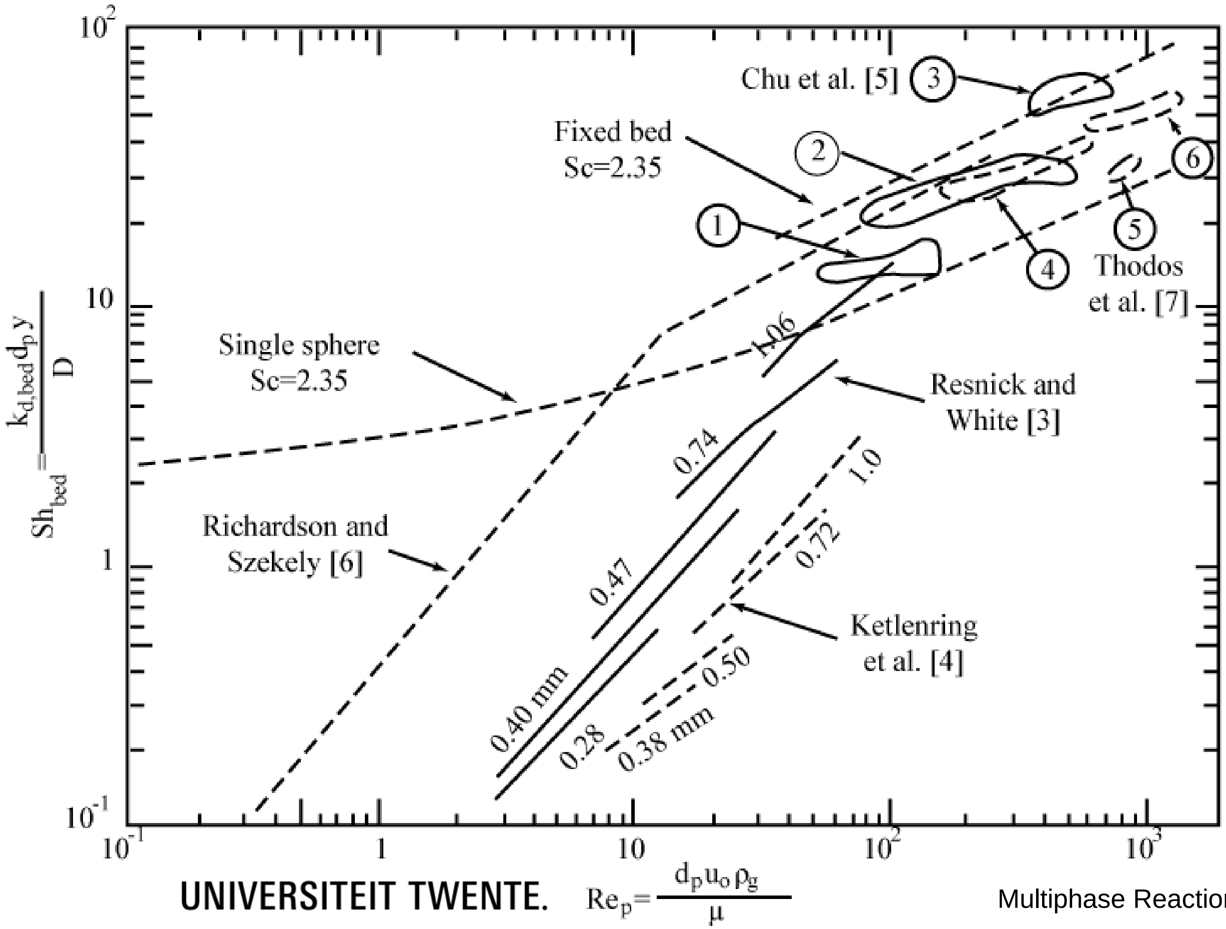
$$Sh^* = 2 + 1.8\left(Re_p\right)^{\frac{1}{2}}(Sc)^{\frac{1}{3}}$$

- Mass transfer: experimental results for fixed beds (Ranz):

- Mass transfer: experimental results for fluidized beds: difficult to measure due to experimental problems (high volumetric mass transfer rate + proper driving force)

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Experimental findings on mass transfer in fluidized beds



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

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Interpretation of mass transfer coefficients: distinction between single particle mass transfer coefficient $k_{d,p}$ and bed mass transfer coefficient $k_{d,bed}$

- Single particle mass transfer coefficient $k_{d,p}$:

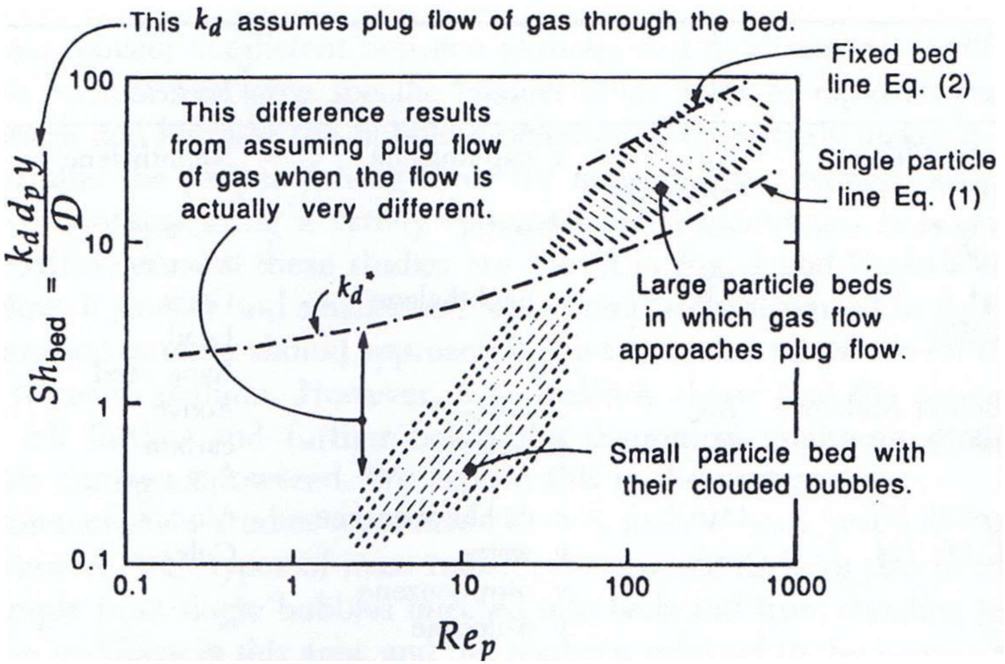
Consider a single particle containing A immersed in a bed of other particles free of component A. In this case we measure the mass coefficient of a single particle

$$-\frac{1}{S_{particle}}\frac{dN_{A,particle}}{dt} = k_{d,p}(C_{A,p} - C_{A,bed})$$

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

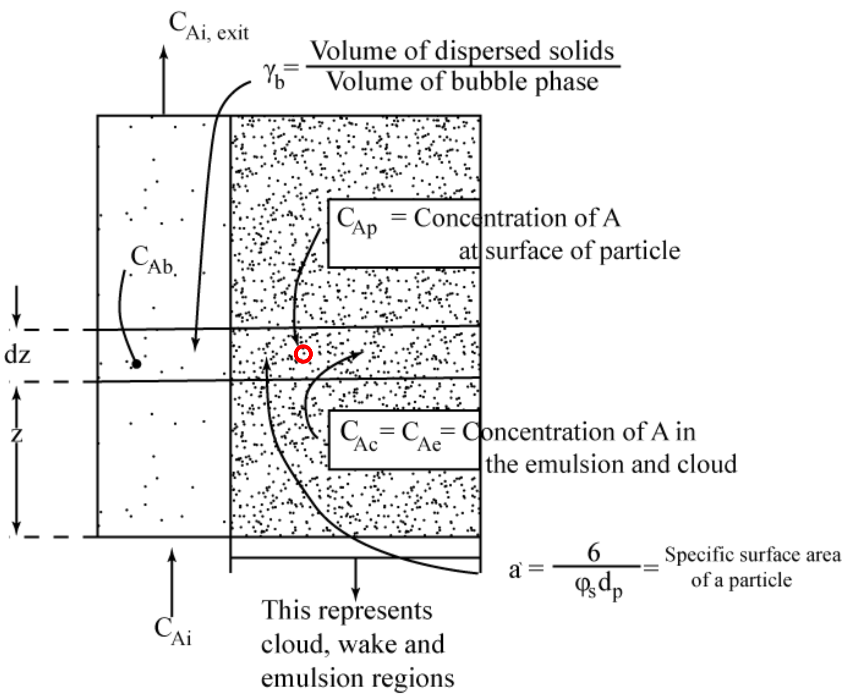
- Bed mass transfer coefficient $k_{d,bed}$: especially for fine particle systems single particle method impractical and therefore whole bed of particles is studied instead

literature results suffer from incorrect interpretation (incorrect flow models)



Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Mass transfer rate from bubbling bed model:



- Differential mass balance for (transferred) component A):

$$\frac{1}{S_{particles}} \frac{dN_A}{dz} = k_{d,bed}(C_{A,p} - C_{A,b})$$

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Differential mass balance for bubble phase:

specific particle surface a'

$$\frac{dC_{A,b}}{dt} = u_b \frac{dC_{A,b}}{dz} = \frac{k_{d,bed}\left(1-\epsilon_f\right)a'}{\delta}\left(C_{A,p}-C_{A,b}\right) \qquad a' = \left(\frac{A_{particle}}{V_{particle}}\right) = \frac{6}{\phi_s d_{sph}}$$

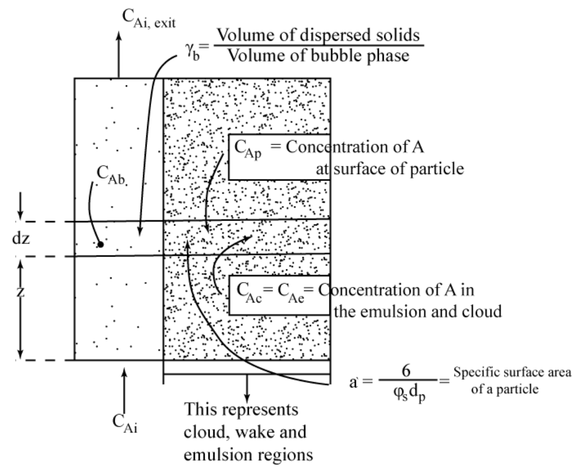
- Mass transfer in terms of interchange coefficient K_d:

$$u_b \frac{dC_{A,b}}{dz} = K_d\left(C_{A,p}-C_{A,b}\right)$$

- Comparison of above expressions leads to:

$$\frac{k_{d,bed}\left(1-\epsilon_f\right)a'}{\delta} \Longrightarrow \qquad Sh_{bed} = \frac{k_{d,bed}d_p y}{D} = \frac{y\phi_s d_p^2\delta}{6D\left(1-\varepsilon_f\right)}K_d$$

Dense Fluidized Beds
Particle-to-gas mass and heat transfer



$$K_d = \begin{pmatrix} \text{added from particles} \\ \text{dispersed in the bubbles} \end{pmatrix} + \begin{pmatrix} \text{transfer across the} \\ \text{bubble - cloud boudary} \end{pmatrix}$$

$$K_d = \gamma a' k_d^* + K_{bc} = \gamma_b \frac{6}{\phi_s d_p} k_d^* + K_{bc}$$

$$= \gamma_b \frac{6\, Sh^* D}{\phi_s d_p^2 y} + K_{bc}$$

K_{bc} given earlier

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Combining previous equations leads to final expression for Sh_{bed}:

$$Sh_{bed} = \frac{\delta}{1-\varepsilon_f}\left[\gamma_b Sh^* + \frac{y\phi_s d_p^2}{6D}\ K_{bc}\right]$$

- From Kunii-Levenspiel model:

$$\frac{\delta}{1-\varepsilon_f} = \frac{u_0 - u_{mf}}{u_{br}\left(1-\varepsilon_{mf}\right)}$$

- For a given bed of solids and constant bubble size above equation can be written as (using equation from Kunii-Levenspiel model):

$$Sh_{bed} = A\ Re_p - B$$

Equation can fit large portion of experimental data for fluidized beds

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Heat transfer: experimental results for single spheres (Ranz):

$$Nu^* = \frac{h^* d_{sph}}{k_g} = 2 + 0.6 \left(\frac{\rho u_0 d_{sph}}{\mu}\right)^{\frac{1}{2}} \left(\frac{C_p \mu}{k_g}\right)^{\frac{1}{3}} = 2 + 0.6 \left(Re_p\right)^{\frac{1}{2}} (Pr)^{\frac{1}{3}}$$

- Heat transfer: experimental results for fixed beds (Ranz):

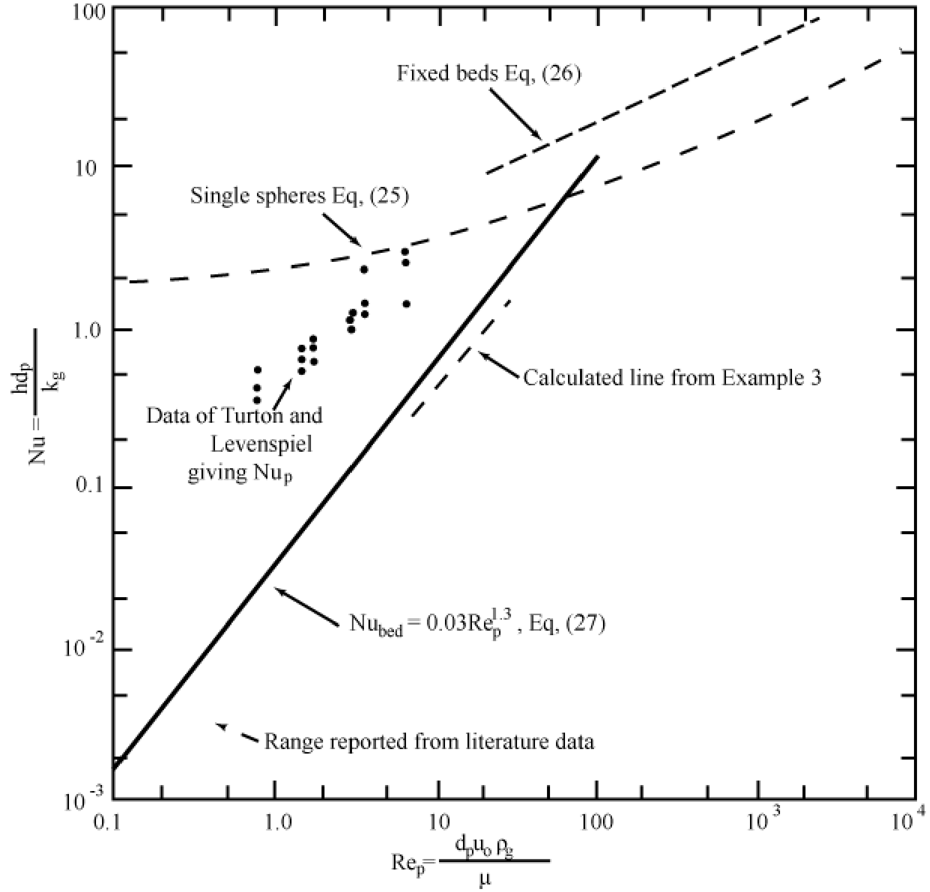
$$Nu^* = 2 + 1.8 \left(Re_p\right)^{\frac{1}{2}} (Pr)^{\frac{1}{3}}$$

- Heat transfer: experimental techniques for gas-fluidized beds: both steady state techniques and unsteady state techniques have been used by many investigators !!!

General problem: flow model used for interpretation of the experimental data

Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Experimental findings on heat transfer in fluidized beds



Dense Fluidized Beds
Particle-to-gas mass and heat transfer

- Nusselt number at low particle Reynolds number (Re_p=0.1 to 100):

$$Nu_{bed} = \frac{hd_p}{k_g} = 0.03\left(\frac{\rho u_0 d_p}{\mu}\right)^{\frac{1}{3}} = 0.03\left(Re_p\right)^{\frac{1}{3}}$$

- Note: data of Turton and Levenspiel correspond to heat transfer coefficients for individual particles and not to bed average quantities !!!

- Expression for bed average Nusselt number from bubbling bed model:

$$Nu_{bed} = \frac{hd_p}{k_g} = \frac{\delta}{1-\varepsilon_f}\left[\gamma_b Nu^* + \frac{\phi_s d_p^2}{6k_g} H_{bc}\right]$$

$$H_{bc} = 4.5\left(\frac{u_{mf}\rho_g C_{p,g}}{d_b}\right) + 5.85\,\frac{\left(k_g \rho_g C_{p,g}\right)^{\frac{1}{2}} g^{\frac{1}{4}}}{d_b^{\frac{5}{4}}}$$

Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- Relevance: temperature control required in many physical and chemical operations ⇒ for design purposes prediction of heat transfer rates between fluidized beds and surfaces is required

- Topics covered:
  - Definition of heat transfer coefficient h
  - Experimental findings vertical tubes
  - Experimental findings horizontal tubes
  - Bed conductivity models
  - General expression for h at a heat exchanger surface

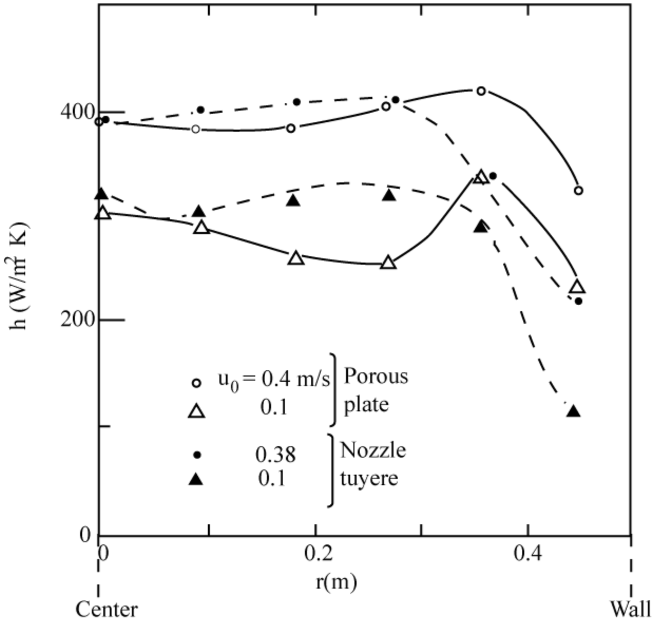
Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- Definition of heat transfer coefficent h:

$$q = A_w h \Delta T$$

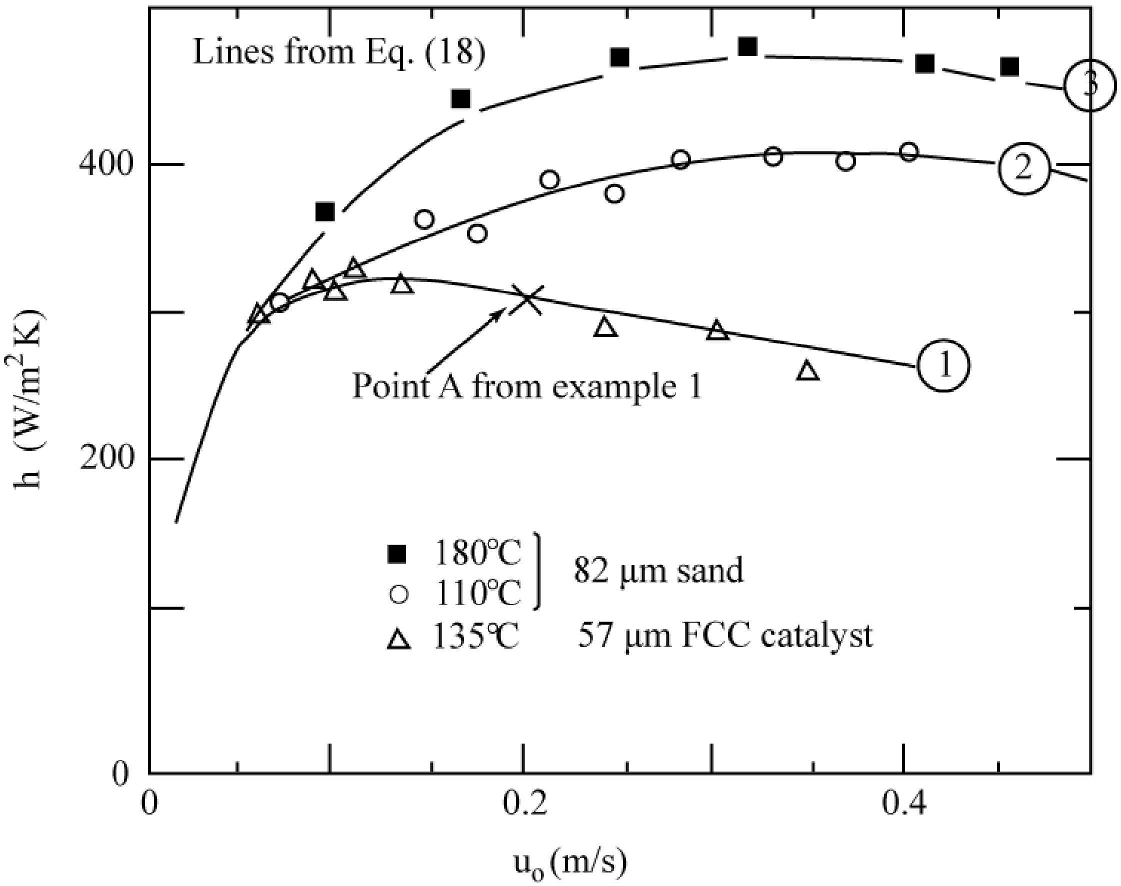
h for fluidized beds is 1-2 orders of magnitude higher than *h* for gas

- Experimental findings vertical tubes (z=0.85 m, d_t=1 m, Lm=1.36 m, bed of quartz sand d_p=96 $_m$m) + effect of gas distributor plate



Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- Experimental findings horizontal tubes in a 0.3 x 0.3 m bed:



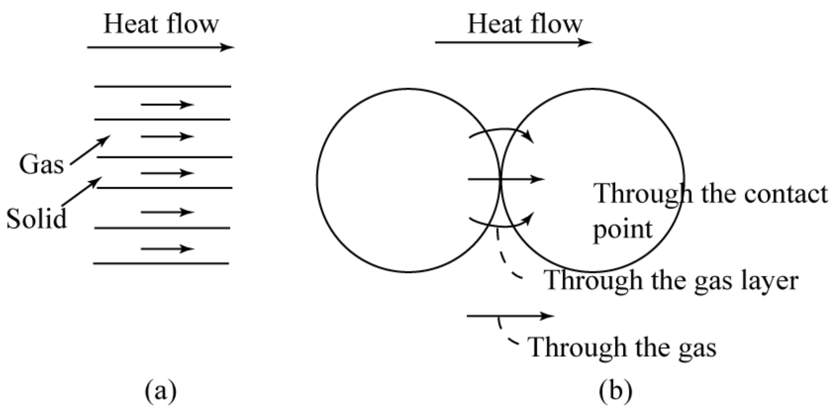
Note: with increasing superficial gas velocity u_o we have two opposing effects:

Decreasing ε_s: lower h

Increasing bubble frequency: higher h

Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- Bed conductivity models



- Case (a): heat flows in parallel paths:

$$k_e^0 = \varepsilon_{mf} k_g + (1 - \varepsilon_{mf}) k_s$$

superscript "o" refers to stagnant gas conditions

- Case (b): modification of parallel path model:

$$k_e^0 = \varepsilon_{mf} k_g + \left(1 - \varepsilon_{mf}\right) k_s \left[\frac{1}{\phi_b \left(\frac{k_s}{k_g}\right) + \frac{2}{3}}\right]$$

$\phi_b = d_{eqv}/d_p$ with d_{eqv} the equivalent thickness of the gas film near the contact points which aids the interparticle heat transport

Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

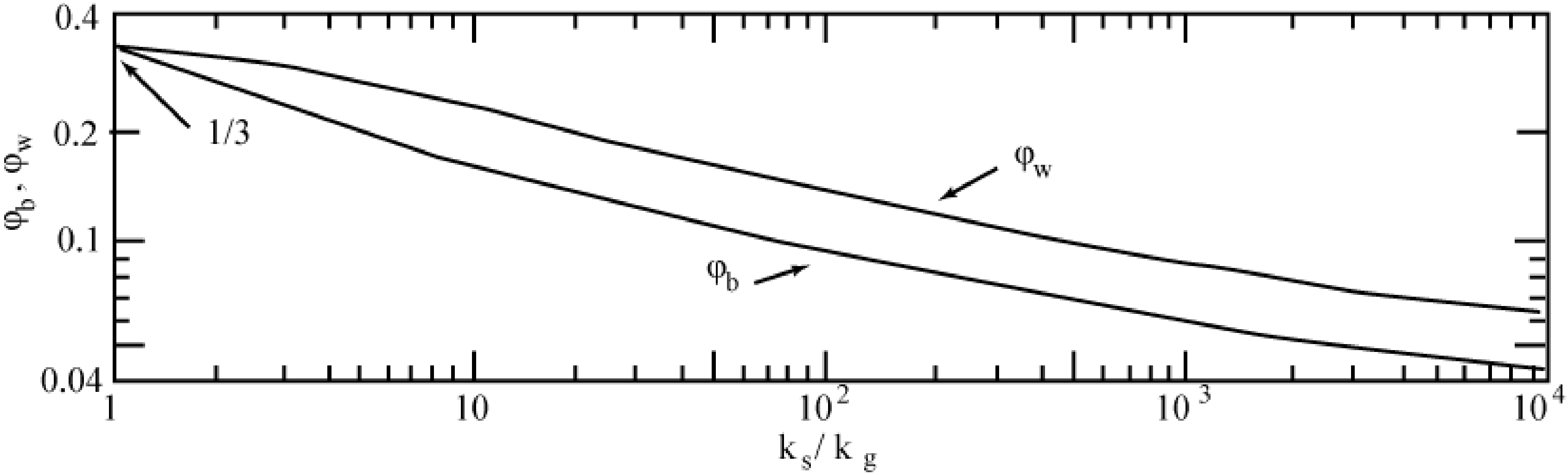
- Modification of parallel path model for wall layer:

$$k_{e,w}^{0} = \varepsilon_w k_g + (1-\varepsilon_w) k_s \left[\frac{1}{\phi_w \left(\frac{k_s}{k_g}\right) + \frac{1}{3}}\right]$$

ε_w represents the mean void fraction for the wall layer

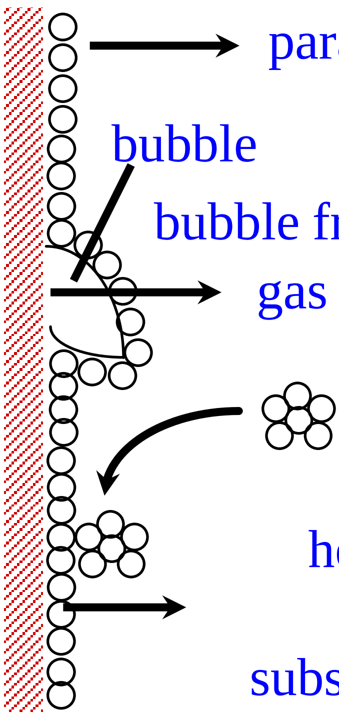
- Figure for determination of ϕ_b and ϕ_w:

$$h_w^0 = \frac{2k_{e,w}^0}{d_p}$$



Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- Conceptual picture of heat transfer between gas-fluidized beds and immersed surfaces (container walls or submerged heat exchange tubes)



Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- General expression for h at a heat exchanger surface:

$$h = \left[\delta_w\left(h_r + h_g\right)\right] + \left[\frac{1-\delta_w}{\dfrac{1}{h_r + h_g} + \dfrac{1}{h_{packet}}}\right]$$

first contribution: bubble at surface; second contribution: emulsion at surface

gas convection contribution in bubble h_g can often be neglected

- Additional equations for general h expression

Contribution due to radiation h_r

$$h_r = \frac{\sigma\left(T_s^4 - T_w^4\right)}{\left(\dfrac{1}{e_s} + \dfrac{1}{e_w} - 1\right)\left(T_s - T_w\right)}$$

Stefan Boltzmann constant
$\sigma = 5.67.10^{-8}$ W/(m^2.K^4)

Dense Fluidized Beds
Heat transfer between fluidized beds and surfaces

- contribution due to gas convection h_w:

$$h_w = h_w^0 + \alpha_w\left(C_{p,g}\rho_g u_0\right) = \frac{2k_{e,w}^0}{d_p} + \alpha_w\left(C_{p,g}\rho_g u_0\right)$$

α_w=0.05 (fitted constant)

- contribution due to packet renewal h_{packet}:

$$h_{packet} = 1.13\left[\frac{k_e^0\rho_s\left(1-\varepsilon_{mf}\right)C_{p,s}n_w}{\left(1-\delta_w\right)}\right]^{\frac{1}{2}}$$

n_w: bubble frequency

- Extreme of fine particles at low temperatures:

$$h = 1.13\left[k_e^0\rho_s\left(1-\varepsilon_{mf}\right)C_{p,s}n_w\left(1-\delta_w\right)\right]^{\frac{1}{2}}$$

surface renewal of
packets emulsion phase

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Relevance: for rational research and the development of new chemical processes tools to predict the performance of fluidized bed reactors are required

- Topics covered:
  - measures of reaction rate and reactor performance
  - experimental findings for fine particle bubbling beds
  - reactor model for fine particle bubbling beds

- Restriction to fine particle bubbling beds due to (relative) significance in large number of industrial processes

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Measures of reaction rate and reactor performance

  ✓ conversion rate for solid-catalyzed reaction:

$$-\frac{1}{V_b}\frac{dN_A}{dt} = K_r C_A$$

- Integration of conversion rate equation

valid for non-porous solid and independent of bed voidage and particle size

  ✓ plug flow:

$$1 - X_A = \frac{C_{A,0}}{C_{A,i}} = \exp(-K_r\tau) \qquad \tau = \frac{V_s}{\phi_g}$$

i=m, f or mf

  ✓ mixed flow:

$$1 - X_A = \frac{C_{A,0}}{C_{A,i}} = \frac{K_r\tau}{1 + K_r\tau} \qquad K_r\tau = K_r\frac{K_i(1-\varepsilon_i)}{u_0}$$

- Equations for plug flow and mixed flow are useful for reference purposes to interprete experimental data and more advanced reactor models

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Experimental findings
  - In general fluidized bed reactors require more catalyst than a fixed bed to achieve a given degree of conversion due to less effective gas-solid contacting

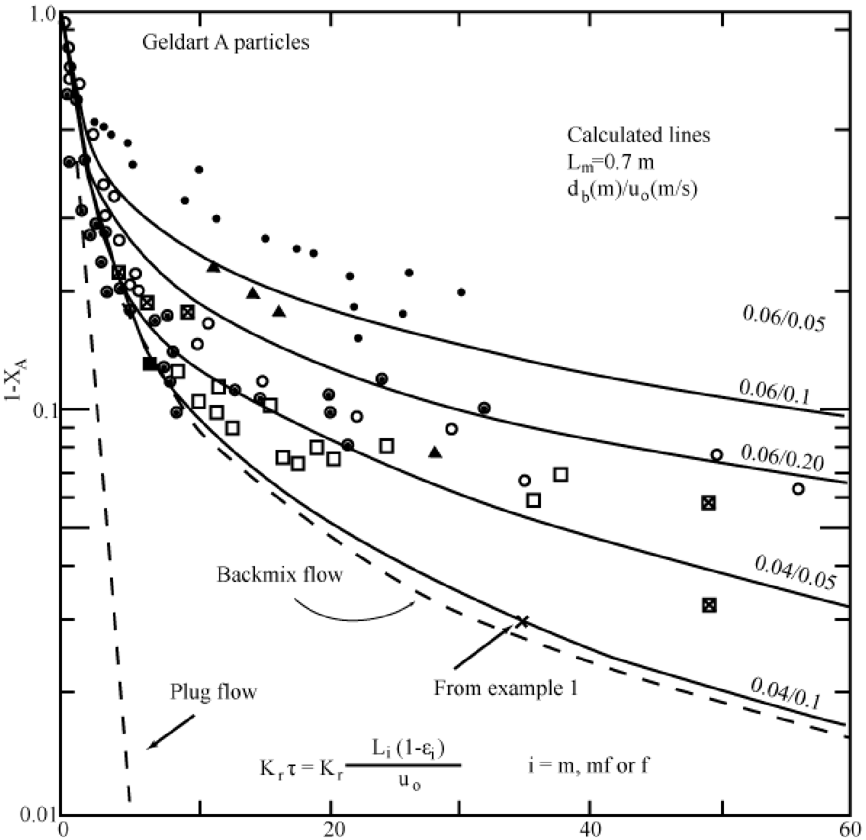
- Behaviour has been studied extensively in literature (often using beds with a too small diameter: hydrodynamics sensitive to bed diameter) using model reactions:

- + Ozone decomposition
- + Oxidation of carbon monoxide

fast irreversible reactions
first order kinetics
simple reaction scheme

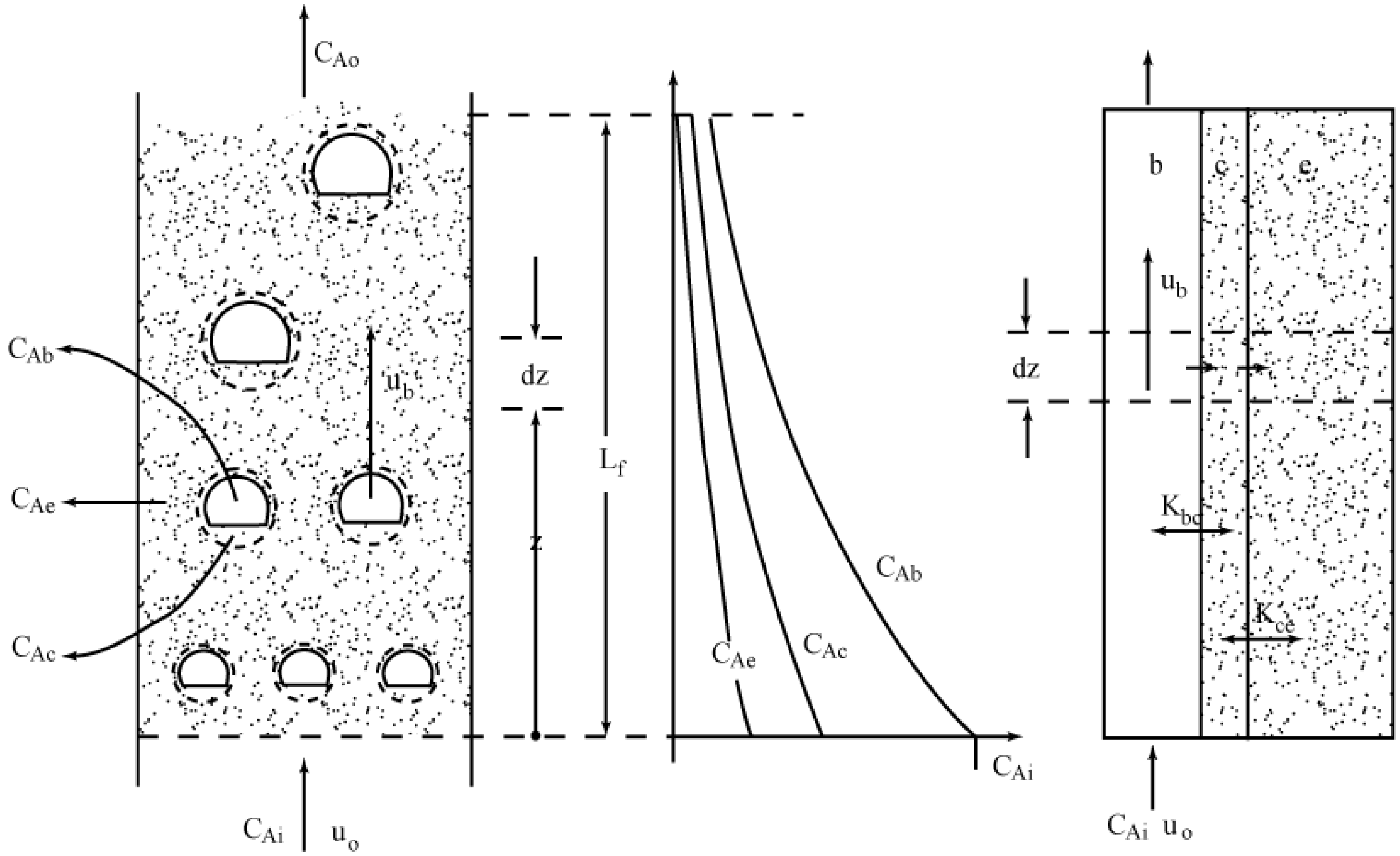
Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Experimental findings for very fine Geldart A catalyst



Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Reactor model for fine particle bubbling beds (vigorously bubbling)



Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Assumptions for “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

- Recall 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

- 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:

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

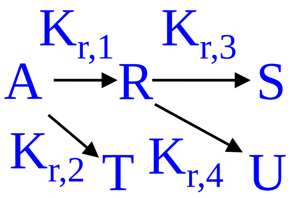
contribution of emulsion is negligible (not valid for coarse particles)

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

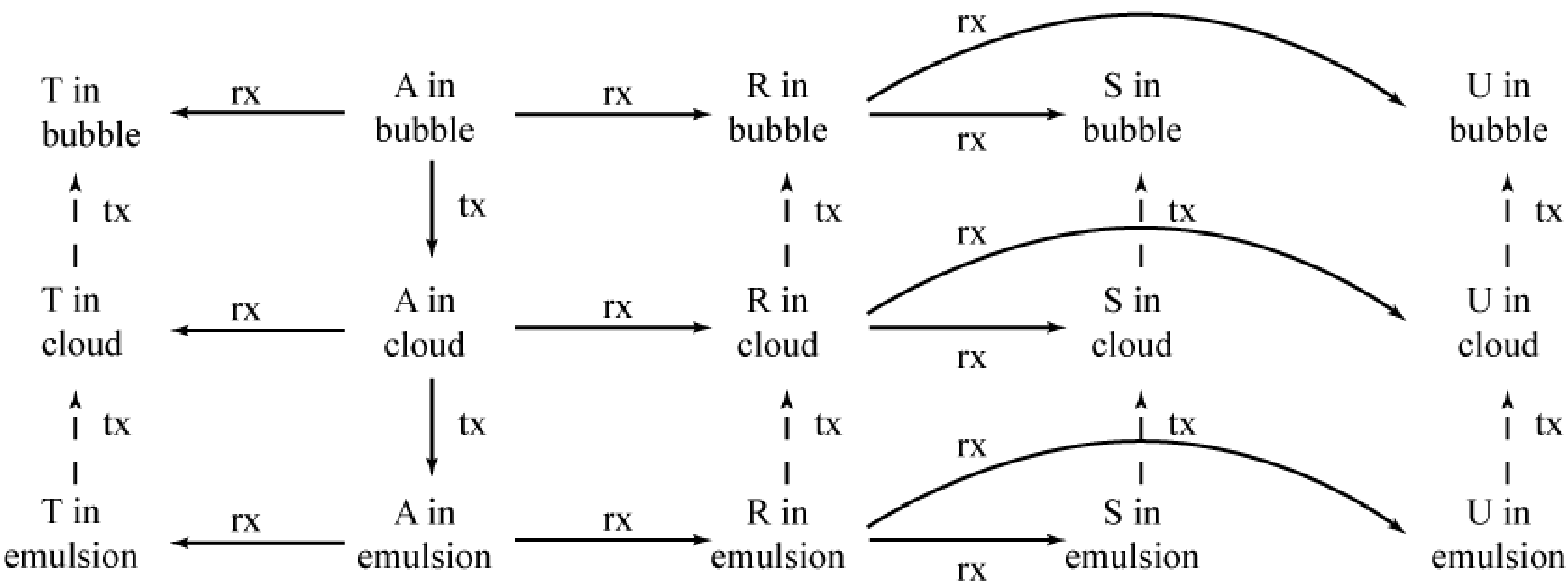
estimates of extent of conversion can be made for fine particle beds

Dense Fluidized Beds
Conversion of gas due to catalytic reactions

- Extension to multiple reactions is straightforward although the algebra becomes (much) more involved !!!



- Schematic representation for a so-called Denbigh reaction



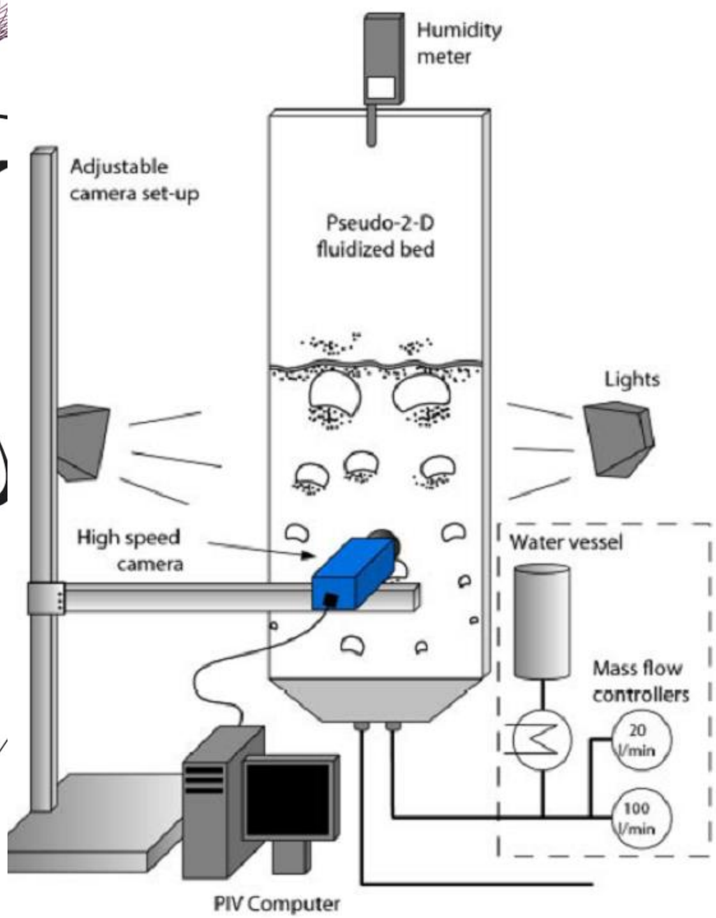
Recent Developments

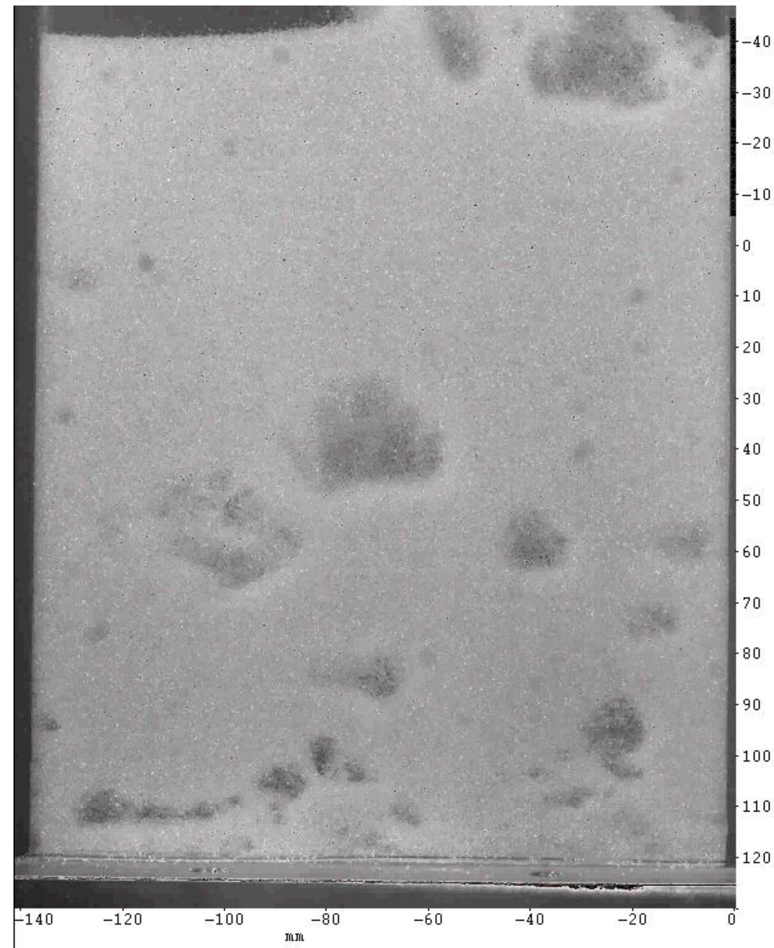
- Applications

  - Fluidization of fine particles (including C-powders)
  - Multi-scale modeling

Recent Developments

- Experimental validation



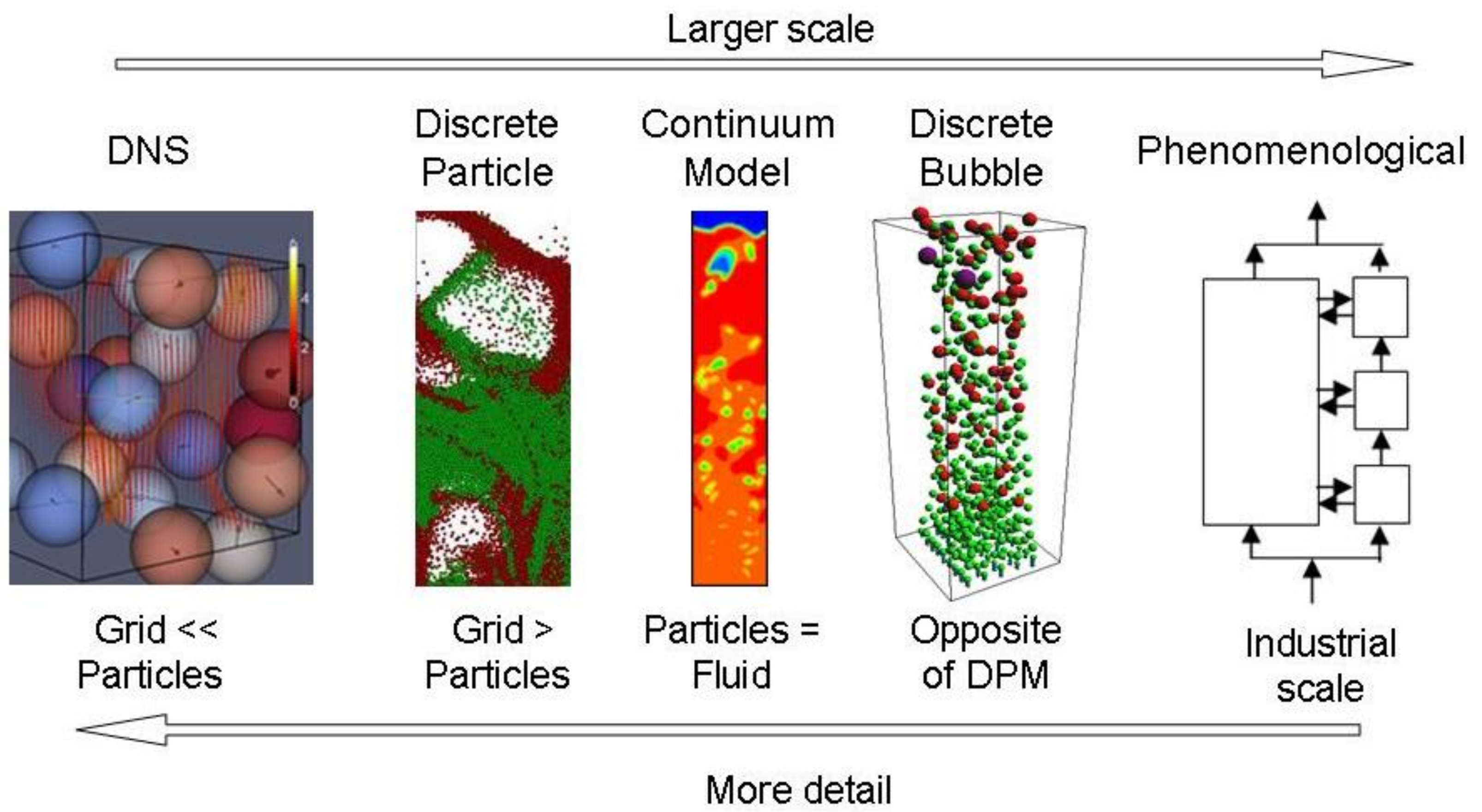


Recent Developments

- CFD models versus GLOBAL SYSTEM MODELS

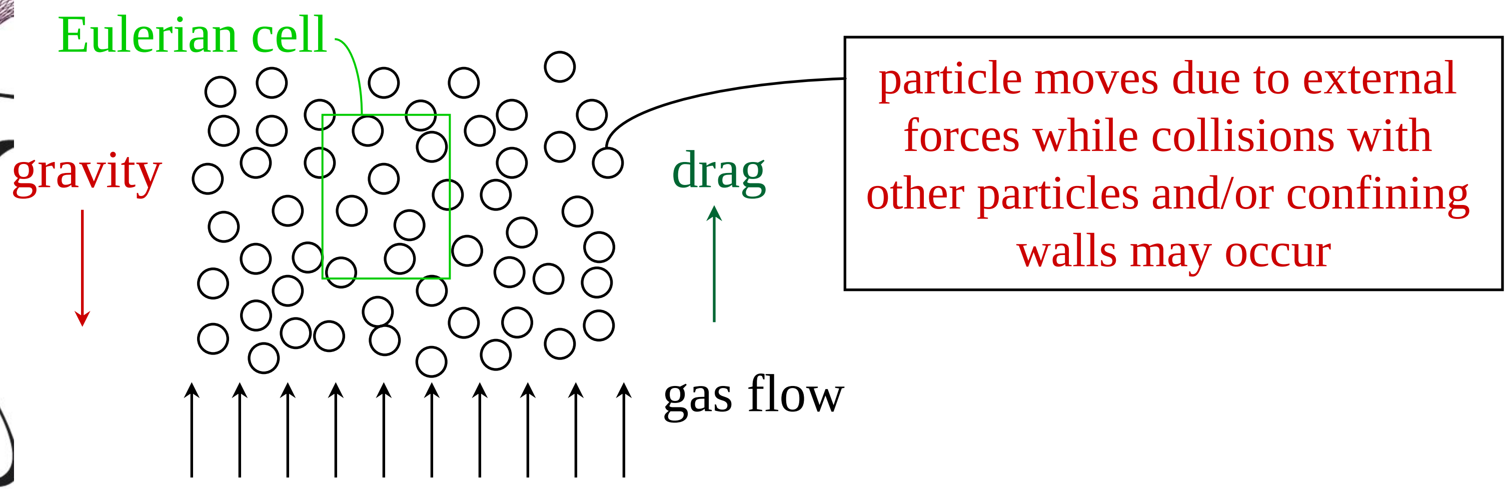
CFD Models	***Global System Models***
Advantage	Advantage
More exact solution available	Simple models and simple solutions facilitate understanding
Phenomena follow from calculation a priori	After adjustment of parameters accurate macro scale behavior prediction
Disadvantage	Disadvantage
Detailed knowledge required about the elementary processes	Experimental validation and adjustment of parameters necessary
Macroscopic behavior not always accurately predicted	Meaning of parameters sometimes unclear due to lumping

Recent Developments
Computational Fluid Dynamics (CFD) approach



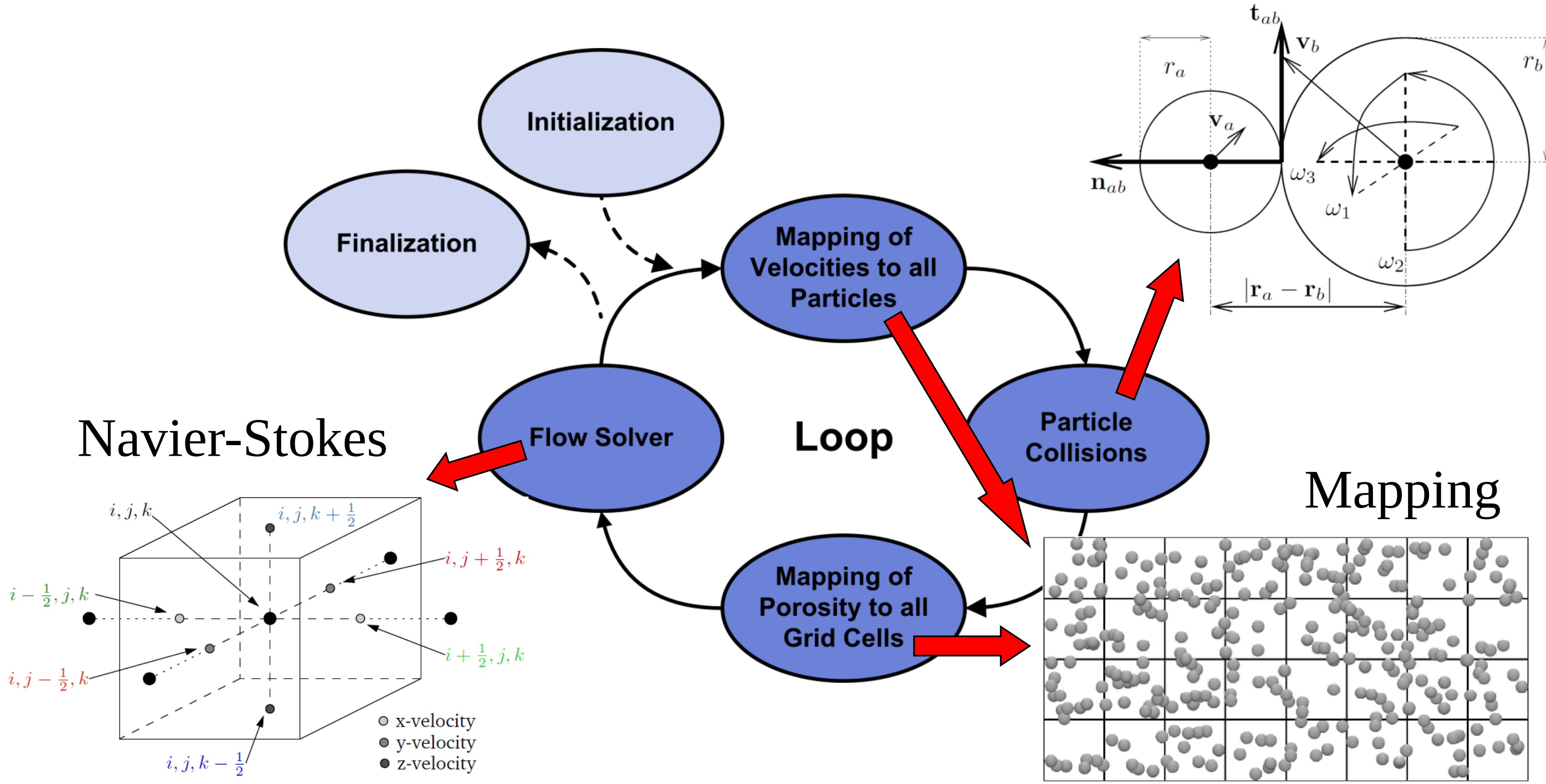
DPM

- Basic idea in discrete particle models

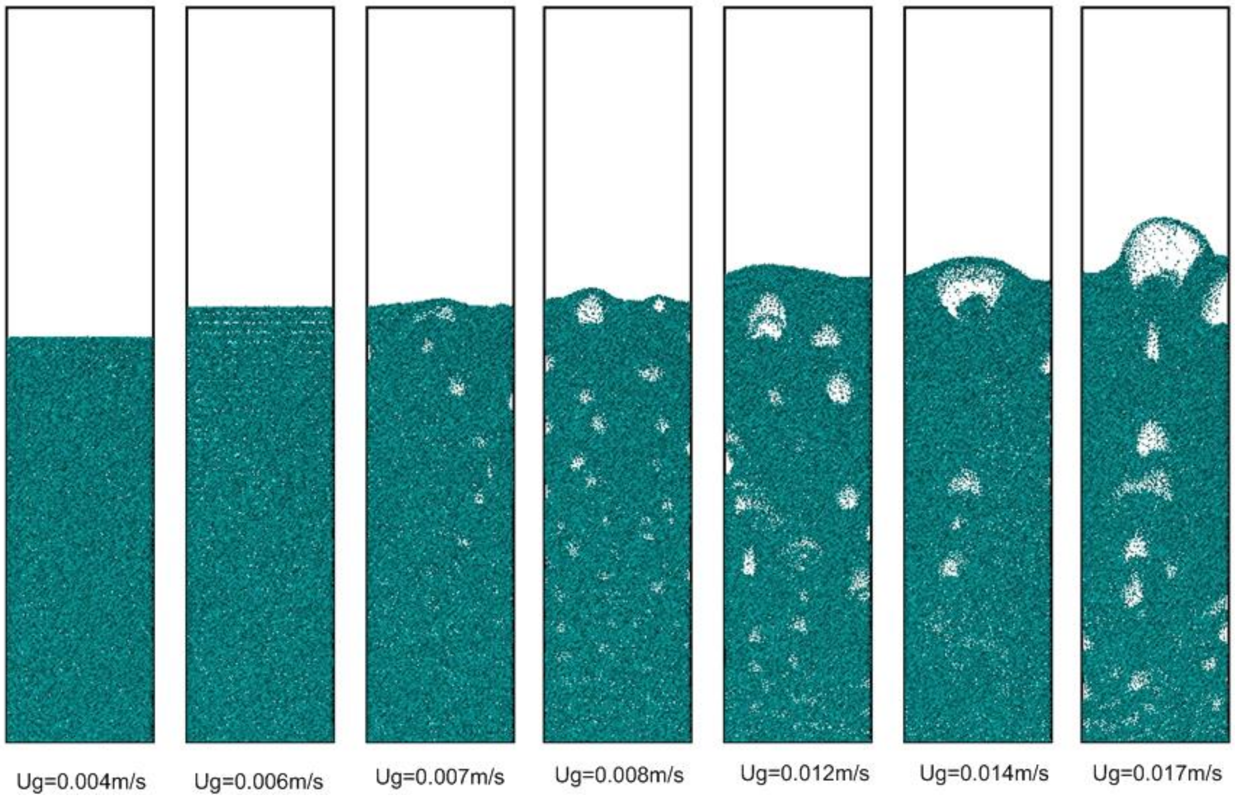


+ Eulerian grid size is larger than the particle size
+ empirical correlation for inter-phase interaction force is needed

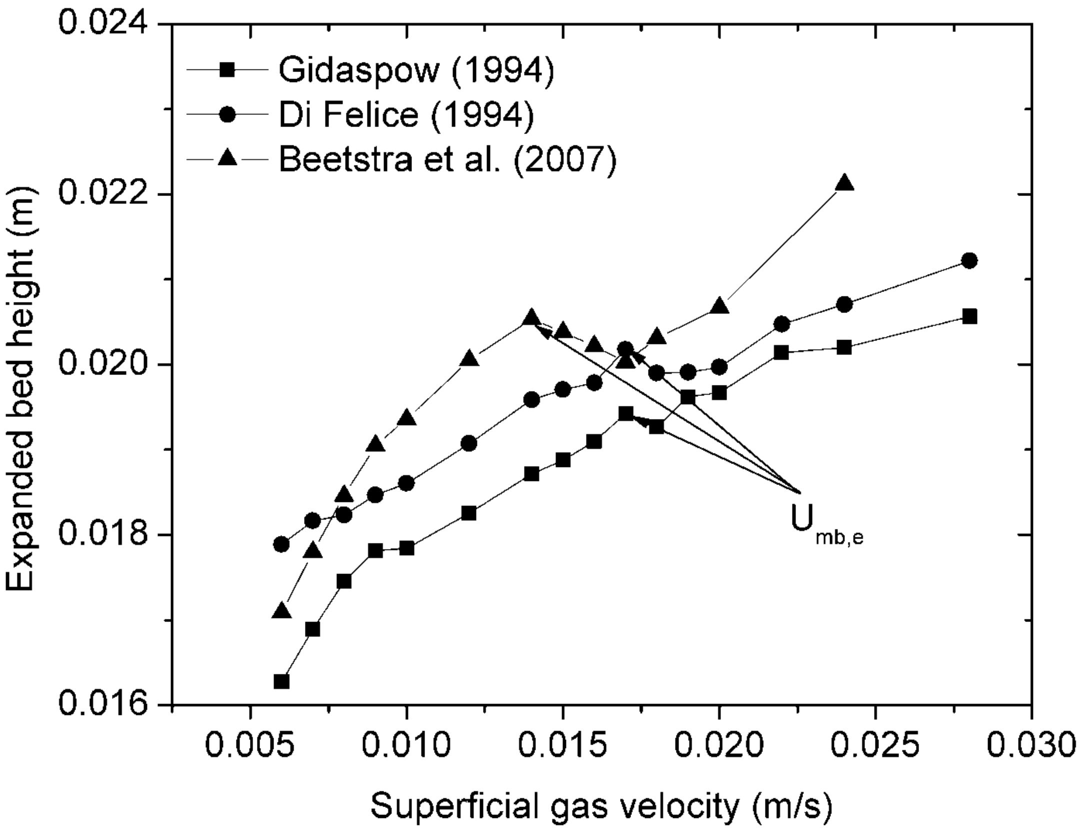
DPM


Initialization
Finalization
Mapping of Velocities to all Particles
Flow Solver
Loop
Particle Collisions
Mapping of Porosity to all Grid Cells
Navier-Stokes
Mapping
x-velocity
y-velocity
z-velocity

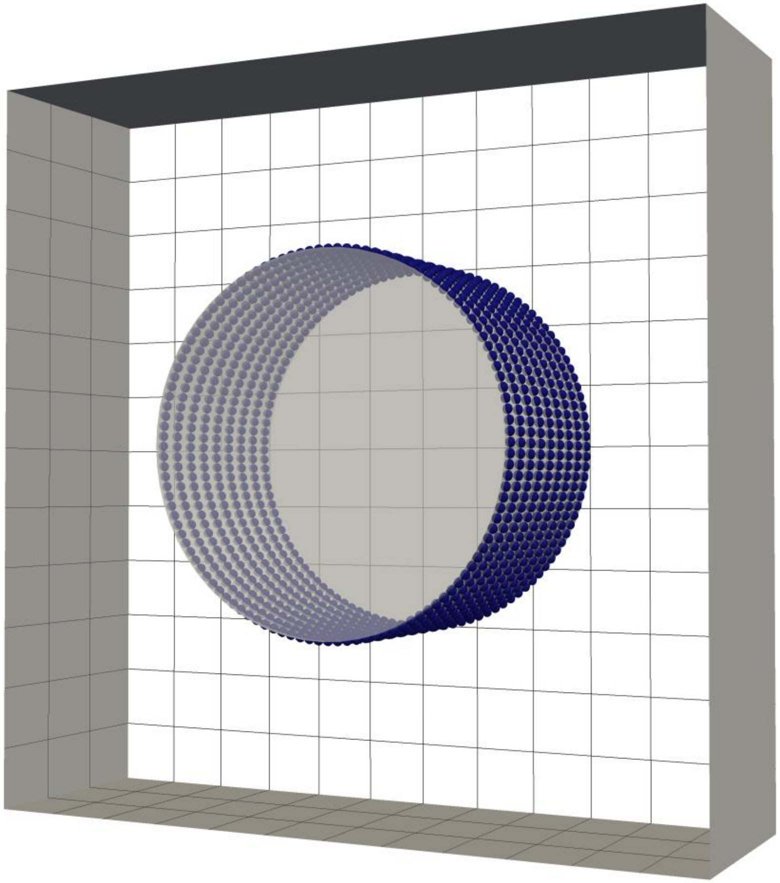
Bed contraction phenomenon

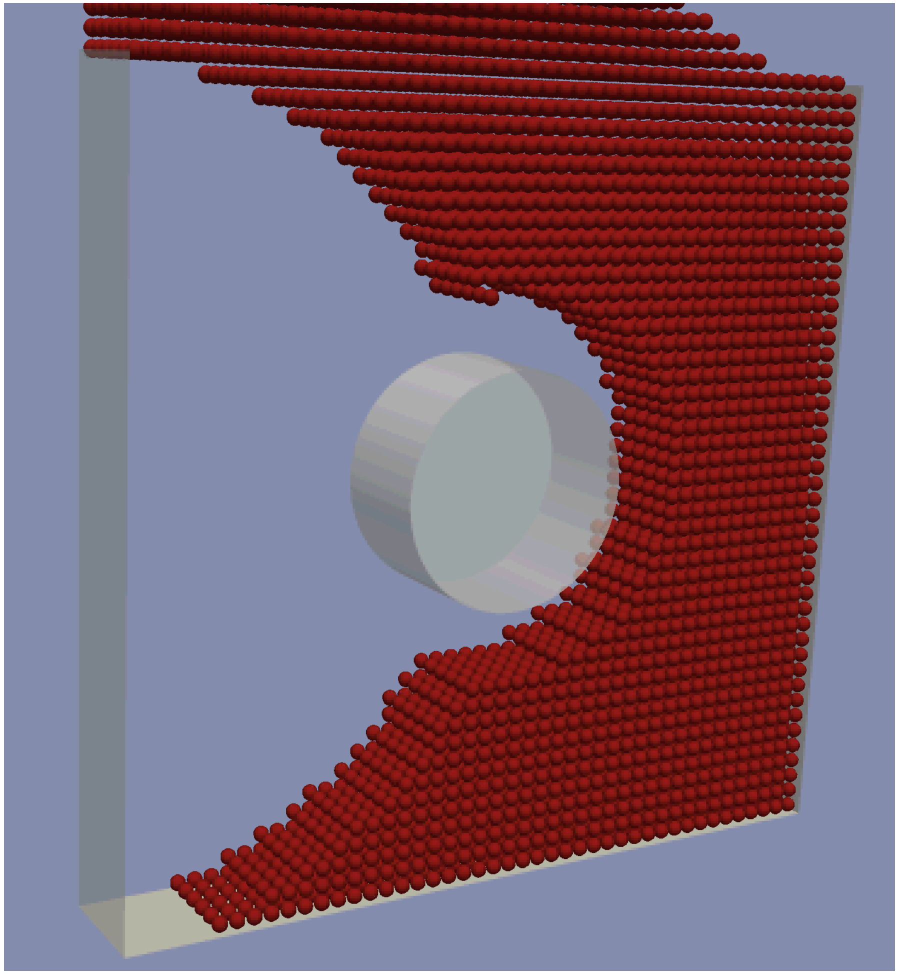


Bed contraction phenomenon



Immersed Boundary Method





Resume

- What is a fluidized bed
- How to define a particle (dimension, classification)
- Particles (and fluid) determine u_{mf}, u_t
- Different types of distributors
- Distributor determines the initial bubble size. Solids and fluid determine bubble velocity/dimensions at a given position
- Bubbles have wakes which are responsible of solid movements/mixing, gas exchange, solid segregation
- A model can be easily solved providing that Kf can be defined (and it depends on all the exchanges already discussed)