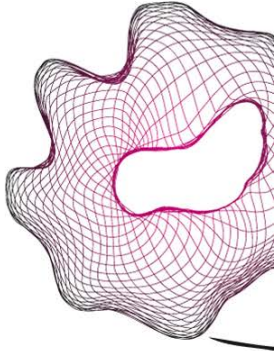


Advanced Chemical Reaction Engineering

Slurry Reactors

Wim Brilman

Contents



✓ Introduction

- ✓ Introduction, examples
- ✓ classification of G/L/S systems
- ✓ slurry properties: density, viscosity, specific heat

✓ Hydrodynamics

- ✓ Suspension of particles
- ✓ Gas-liquid contacting; regimes, gas holdup, mixing

✓ Mass transfer

- ✓ Mass transfer and reaction in series
- ✓ Mass transfer parameters
- ✓ Enhancement of mass transfer

✓ Heat transfer

- ✓ Assignment workshop Slurry Reactors





Introduction

For further reading

“Gas-Liquid-Solid Reactor Design”, Y.T. Shah (McGraw-Hill, 1979)

“Multiphase chemical reactors”, A. Gianetto and P.L. Silveston

“Three Phase Sparged Reactors”, K. Nigam, A. Schumpe (1996)

Beenackers, van Swaaij, Chem.Eng.Sci., 48, 3109-3139, 1993

PhD thesis Vandu (UvA, 2005)

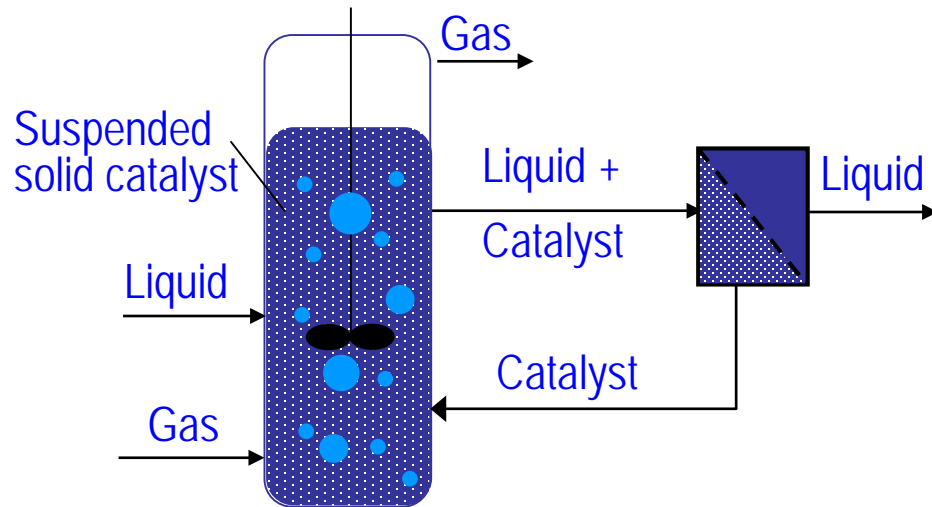
PhD theses Brilman, Huizenga, Cents (UT, 1998, 1998, 2003)

Introduction

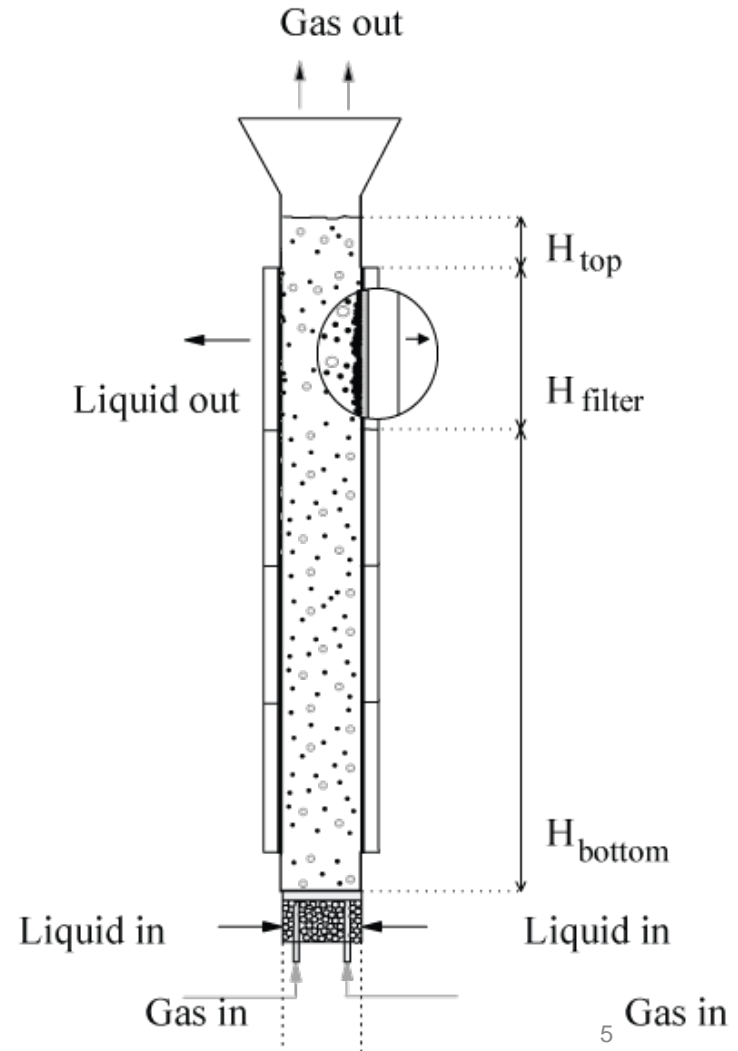
- Slurry reactor = Gas – Liquid – Solid System
- Solid : a product, reactant and/or (often) a heterogeneous catalyst.
- **Advantages** of slurry processes:
 - Good temperature control (*due to backmixing, high c_p , HTC, small d_p*)
 - Application of very fine solids is possible
 - high specific solids contact area $a_p = 6\varepsilon_p / d_p$
 - No concentration or temperature profiles within the particle
 - Continuous and batch operation
 - Continuous solids replacement possible
- **Disadvantages**
 - Backmixing of liquid phase (sometimes)
 - Liquid solid separation required (in most cases)

Liquid-Solid separation: external or in-situ

(more conventional approach)



In-situ filtration (UT, 1998)



- Several possible liquid solid separation techniques
 - Sedimentation
 - Filtration
 - Cyclone
 - Centrifuge
 - Magnetical



Introduction – some applications

TABLE 12-1. APPLICATIONS OF THREE-PHASE REACTORS

I. *Slurry Reactor*

A. Hydrogenation

1. of fatty acids over a supported nickel catalyst.
2. of 2-butyne-1,4-diol over a Pd-CaCO_3 catalyst.
3. of glucose over a Raney nickel catalyst

B. Oxidation

1. of C_2H_4 in an inert liquid over a PdCl_2 -carbon catalyst
2. of SO_2 in inert water over an activated carbon catalyst

C. Hydroformylation

of CO with high-molecular-weight olefins on either a cobalt or ruthenium complex bound to polymers

D. Ethynylation

Reaction of acetylene with formaldehyde over a CaCl_2 -supported catalyst

Also: Fischer-Tropsch, Methanol synthesis, Bioreactors

Introduction

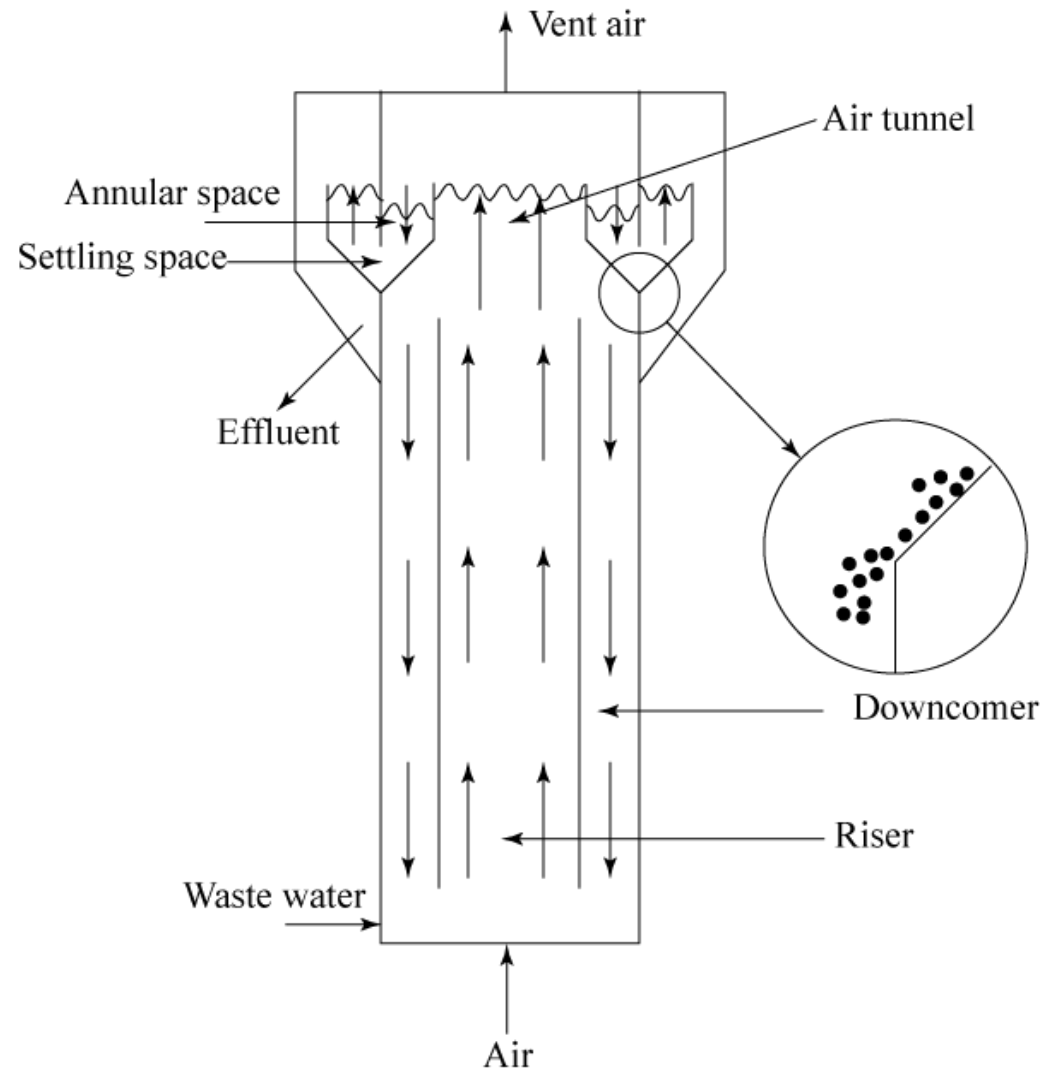


(a) biothane biofilm airlift suspension and expanded granular sludge blanket



(b) Paques CIRCOX reactors at a brewery in Brazil (foreground; 140 m³) and with internal circulation (background; 385 m³)

Introduction



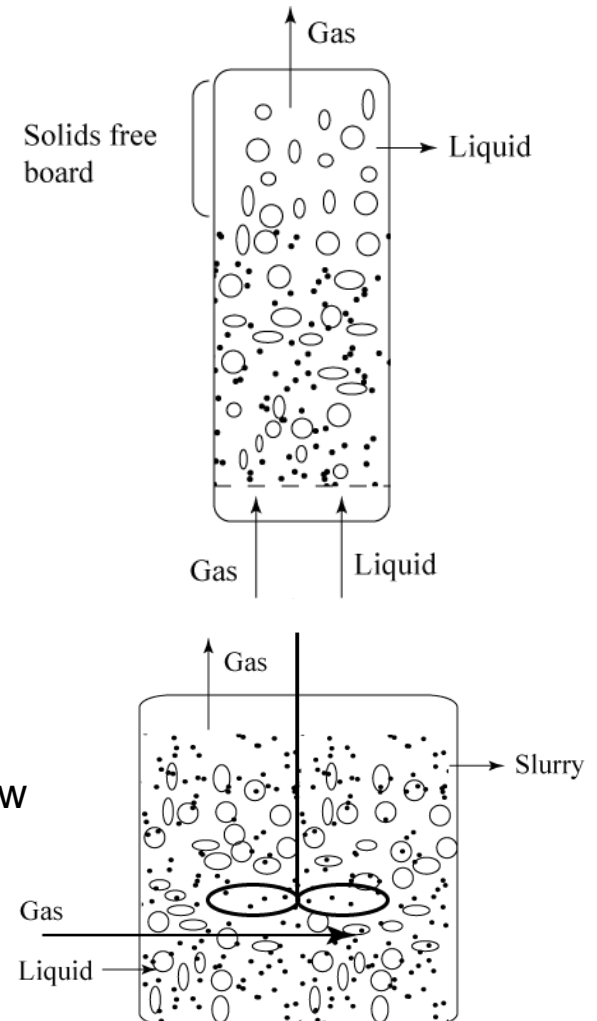
Introduction

- Particles are kept in suspension by liquid eddies, induced by gas flows and/or mechanical agitation

Characteristic parameter values for a slurry reactor:

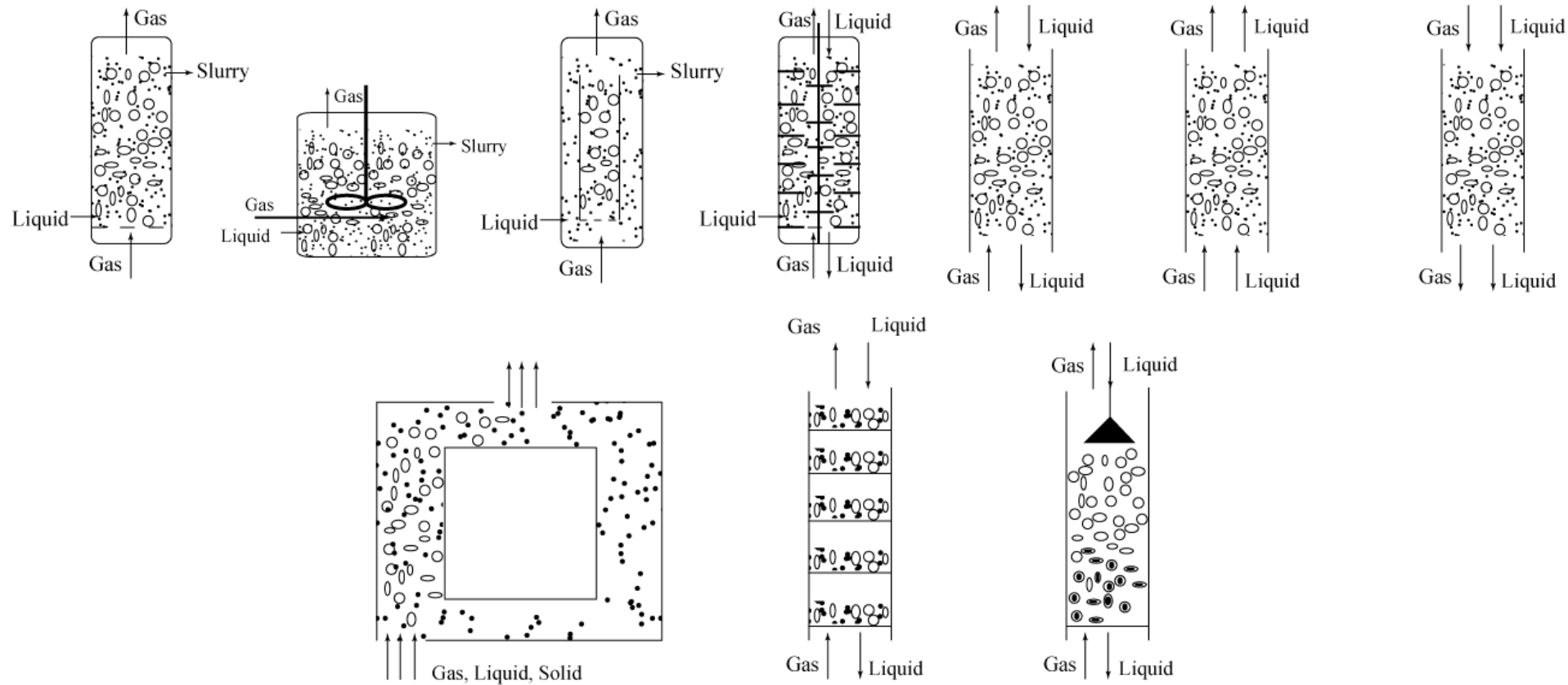
Particle size	10 - 200	μm
Solids fraction	1 - 20	vol.%
Gas hold-up	10 - 20	vol.%
G/L interfacial area	100 - 400	m^2/m^3

At high density differences $|\rho_s - \rho_L|$ and large particle sizes (i.e. high settling velocity) a net upward(or downward) liquid flow is required. This regime, called **3 phase fluidization**, is **not** considered here.



Classification of slurry reactors

Classification by contact pattern and mechanical devices



Classification of slurry reactors

■ Classification by reaction system

Type	Reactant	Example
I	Gas, Liquid and Solid	CO ₂ adsorption in lime suspension
II	Gas and Solid, Inert Liquid	Hydride formation
III	Gas phase reactant, solid catalyst, inert liquid	<u>Fischer Tropsch</u> , Methanol
IV	Gas and liquid phase reactants, solid catalyst	Hydrogenation of edible oil

Fischer-Tropsch slurry reactors



Oryx I - 34000 bbl/day in Qatar



To be scaled up from pilot plant by factor 10^2 - 10^3

Scale-up strategy ?

Which parameters to consider ?



*Fischer-Tropsch reactors shipping
underway*

Weight = 2200 tons

Height = 60m

Outer Diameter = 10m nominal

On the addition of solid particles ...

- **Physical properties change...**

- Density $= \varepsilon_L \cdot \rho_L + \varepsilon_S \cdot \rho_S$
with ε the volume fraction of liquid resp. solid phase
- Heat capacity $= w_L \cdot c_{p,L} + w_S \cdot c_{p,S}$
with w the weight fraction of liquid / solid phase and c_p the specific heat (J/kg K)
- Viscosity *(see next slide)*

- **Effect on heat- and mass transfer processes**

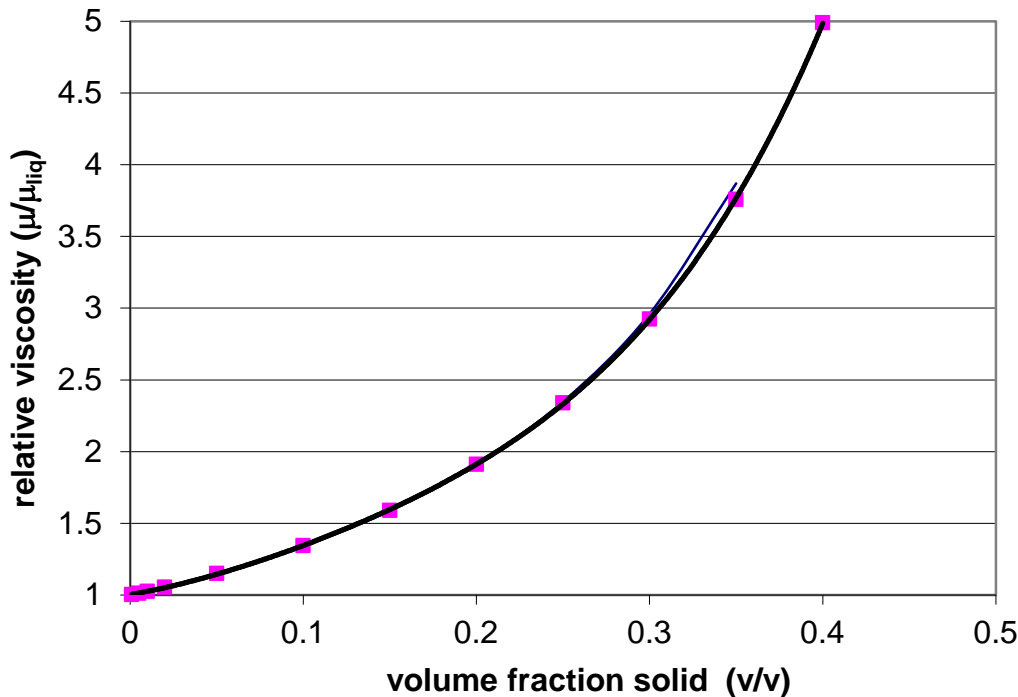
- **Operational issues**

- Blockage (due to local settling)
- Erosion (of particles and equipment)

Slurry viscosity - estimation

Thomas eqn.: $\mu_{\text{slurry}}/\mu_{\text{Liq.}} = 1 + 2.5 \cdot \varepsilon_s + 10.05 \cdot \varepsilon_s^2 + 0.00273 \cdot e^{(16.6 \cdot \varepsilon_s)}$

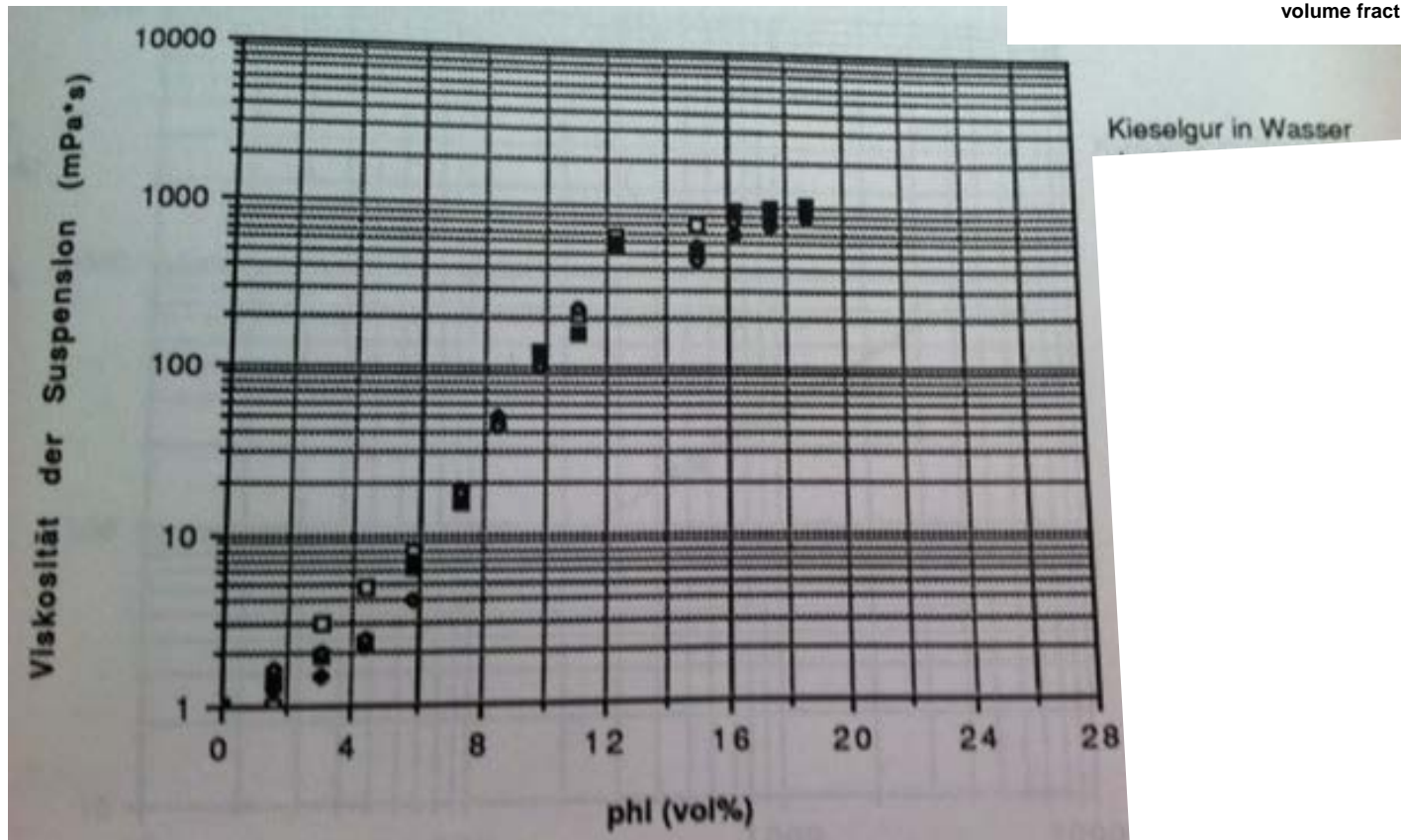
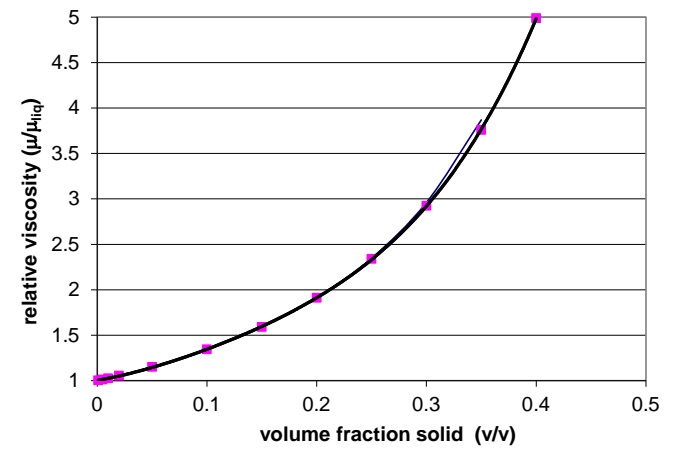
Graham corrl.:
$$\mu_r = \left(\frac{9}{4}\right) \cdot \left[1 + \frac{h}{2a}\right]^{-1} \cdot \left[\frac{1}{\left(\frac{h}{a}\right)} - \frac{1}{1 + \left(\frac{h}{a}\right)} - \frac{1}{\left[1 + \left(\frac{h}{a}\right)\right]^2} \right] + \left(1 + \left(\frac{5}{2}\right)\phi\right) \left(\frac{h}{a}\right) = 2 \cdot \left[\frac{1 - \left(\frac{\phi}{\phi_m}\right)^{1/3}}{\left(\frac{\phi}{\phi_m}\right)^{1/3}} \right]$$



Max. packing density, via experiment or estimation ($\approx 0.6-0.65$)

Viscosity kieselguhr in water

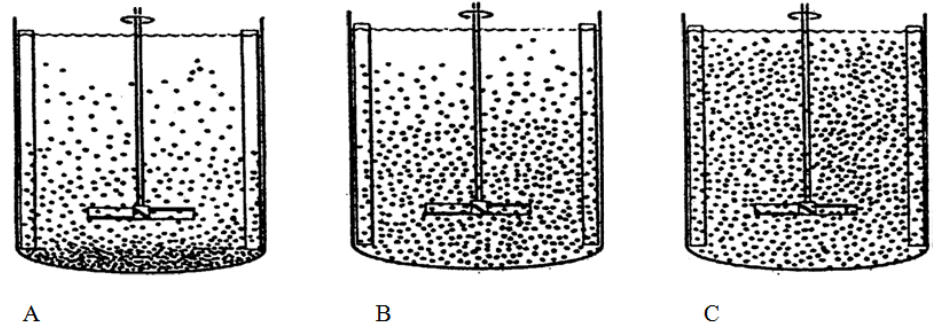
exp. data.



Hydrodynamics; Suspension of the solid particles

- Suspension of solid particles

⇒ Particle suspension criteria



$$N_{js} = \frac{S(g\Delta\rho/\rho_L)^{0.45} (d_p)^{0.2} X^{0.13} \gamma^{0.1}}{D^{0.85}}.$$

1. Homogeneous suspension: particles are homogeneously distributed in liquid
2. Complete suspension: all particles are just suspended

To reach a homogeneous suspension much more energy is required

Generally, suspension conditions are chosen close to complete suspension to reduce the power input required

This implies that a solids concentration gradient over the height may exist

Hydrodynamics – power input per unit of mass

The concept of power input per unit of mass (or volume) as a key parameter has found wide application in correlating other data as e.g. mass transfer parameters, interfacial areas, dispersion coefficients

Mechanical agitators: the average power input is given by:

$$\bar{\varepsilon} = \frac{C \rho_L n^3 d_s^5}{\rho_L V_R}$$

However, locally this value can be very different

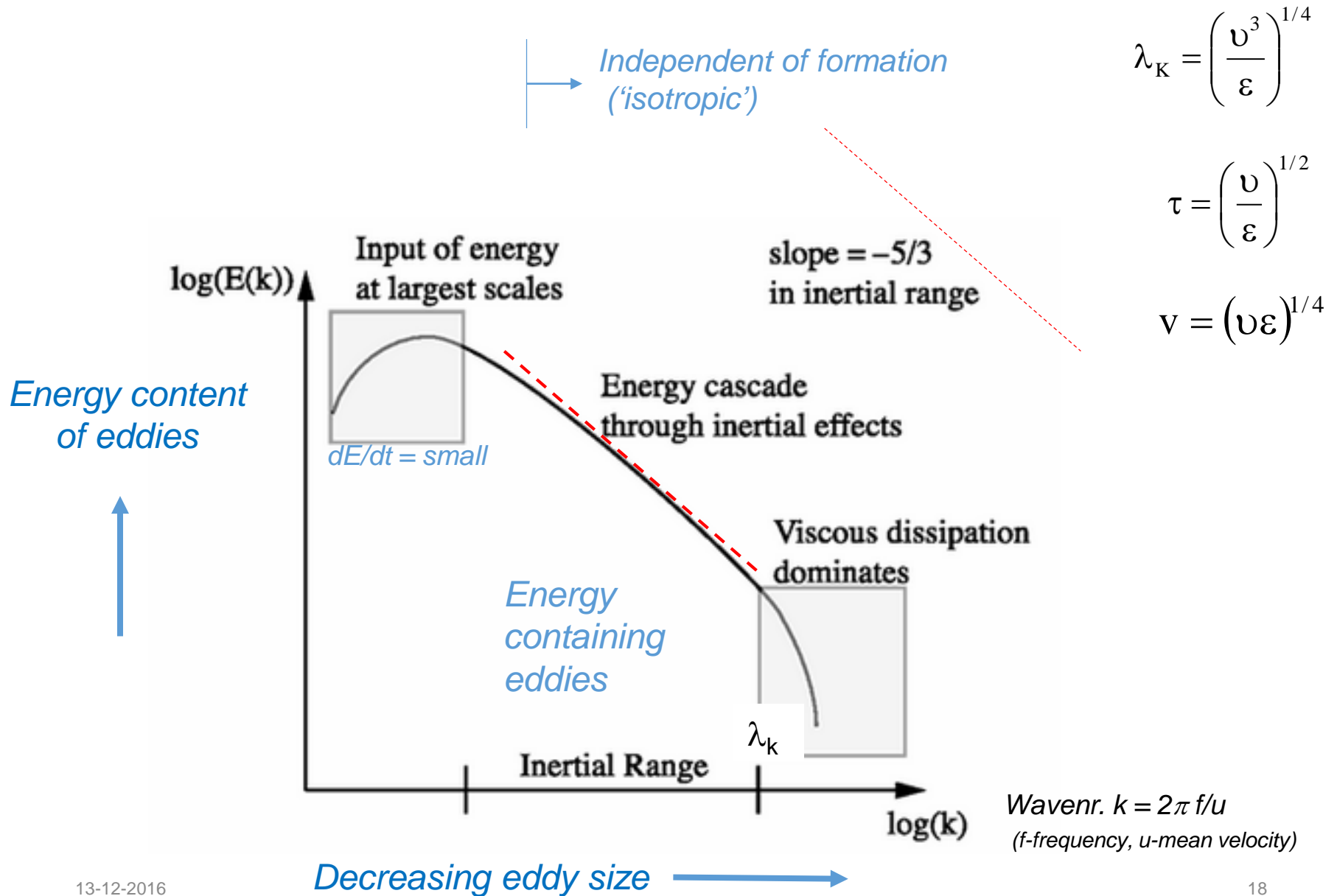
In sparged columns the power input by bubble agitation can be estimated using the liquid head above the gas sparger

$$\bar{\varepsilon} = \frac{\Delta P \Phi_G}{M} = \frac{\rho_L g \varepsilon_L H \cdot A u_G}{\rho_L \varepsilon_L H A} = g u_G$$

The concept of the isotropic turbulence theory has, also for bubble columns, proved to be very useful in correlating data

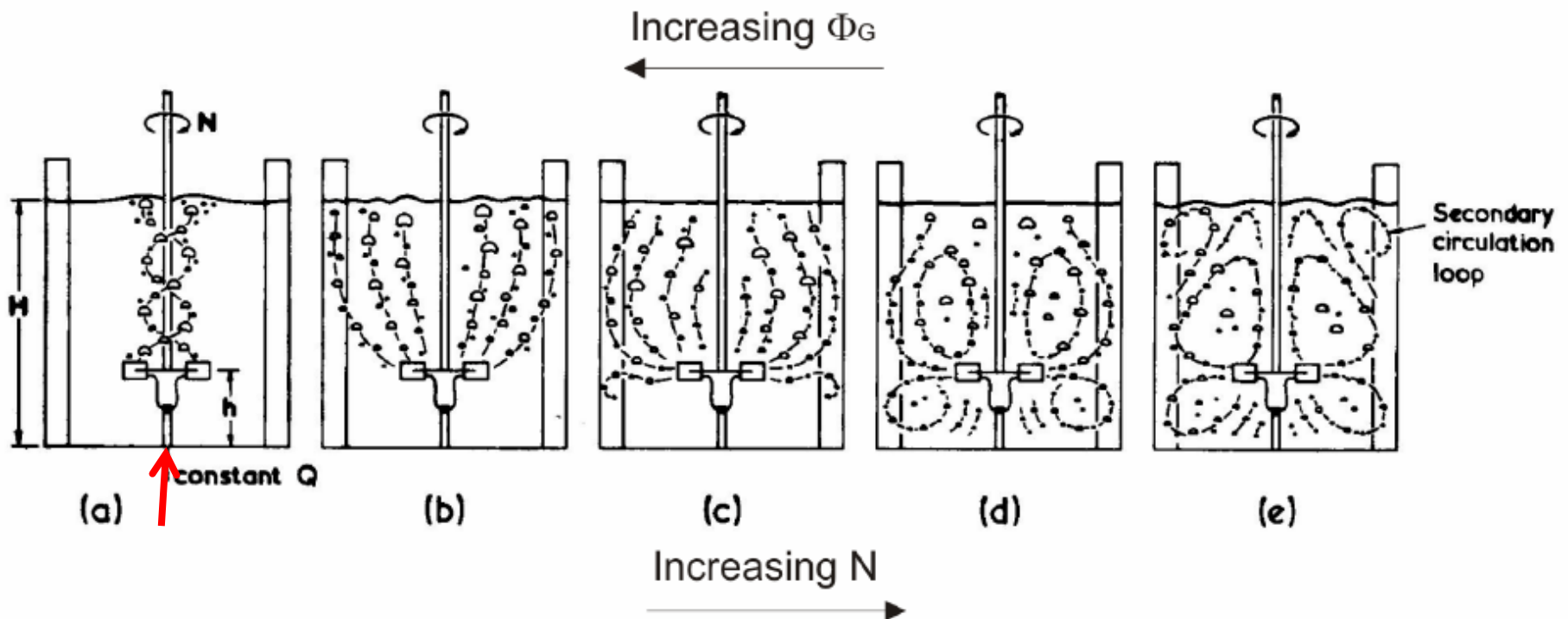
Experiments showed that also in bubble columns the power input is dissipated in the small eddies-regime

Energy spectrum of (turbulent) eddies



Hydrodynamics: Gas-dispersion in stirred tank reactors

$$\bar{\varepsilon} = \frac{C \rho N^3 d_s^5}{\rho_L V_R}$$



Hydrodynamics – Effect of gas sparging on power input (stirred slurry)

- For gassed, stirred liquids the following relationship is recommended

$$\frac{P_G}{P_0} = 0.1 \left(\frac{nV}{F_G} \right)^{0.25} \left(\frac{n^2 d_s^4}{g w_s V^{\frac{2}{3}}} \right)$$

P_G is power input for gassed slurry,
 P_0 for liquid-solid system

- The presence of gas lowers the liquid density and with that the power input at equal impeller speed. Therefore it is no surprise that the critical impeller speed n_c for suspension of particles is higher when gas is present.
- Limiting gas flows for keeping
 - Complete suspension: $F_G < 5 \cdot 10^{-3} (n - n_c) d_s^3$
 - Partial suspension: $5 \cdot 10^{-3} < \frac{F_G}{(n - n_c) d_s^3} < 5 \cdot 10^{-2}$
 - Total sedimentation: $F_G > 5 \cdot 10^{-2} (n - n_c) d_s^3$

(for propeller type of stirrer, [Subbaro, Taneja, 1979])

Hydrodynamics: Solid Dispersion

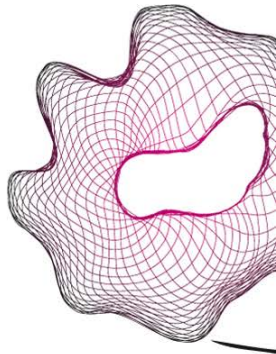
- The average solids concentration simply follows from an overall mass balance over the reactor. However, the local solids concentration depends on the state of suspension

- Especially in bubble columns a solids concentration profile along the height h can exist:

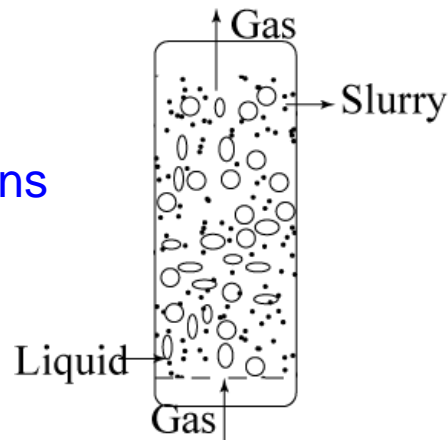
$$\frac{w_s}{w_{s,0}} = \exp \left[-\frac{v_p h}{E_{LS}} \right]$$

- The uniformity of the solids concentration is enhanced by higher gas and liquid velocities (higher power input) and higher viscosities

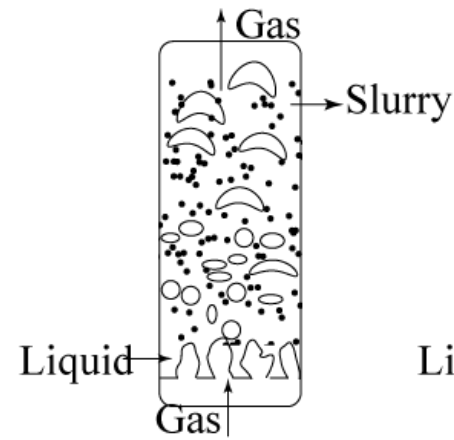
Hydrodynamics: Gas-liquid flow regimes



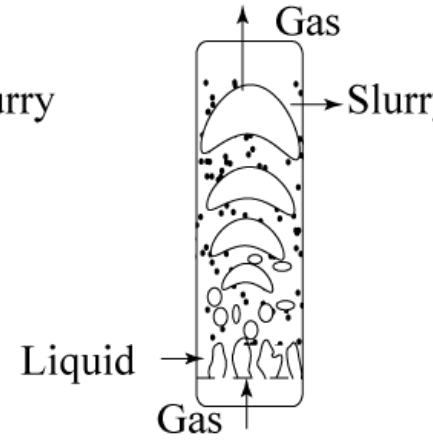
Flow patterns



Homogeneous flow



Churn turbulent flow

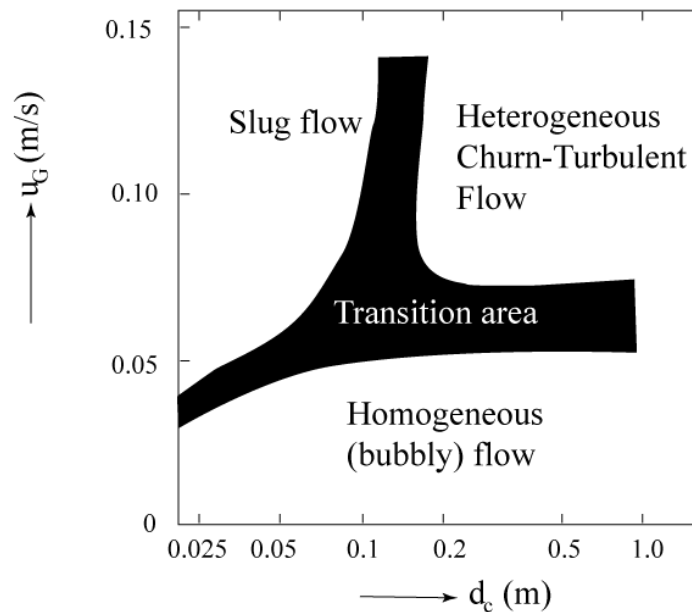


Slug flow

Regime map

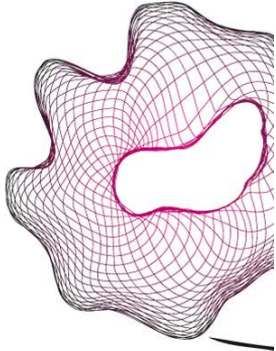


Without solids

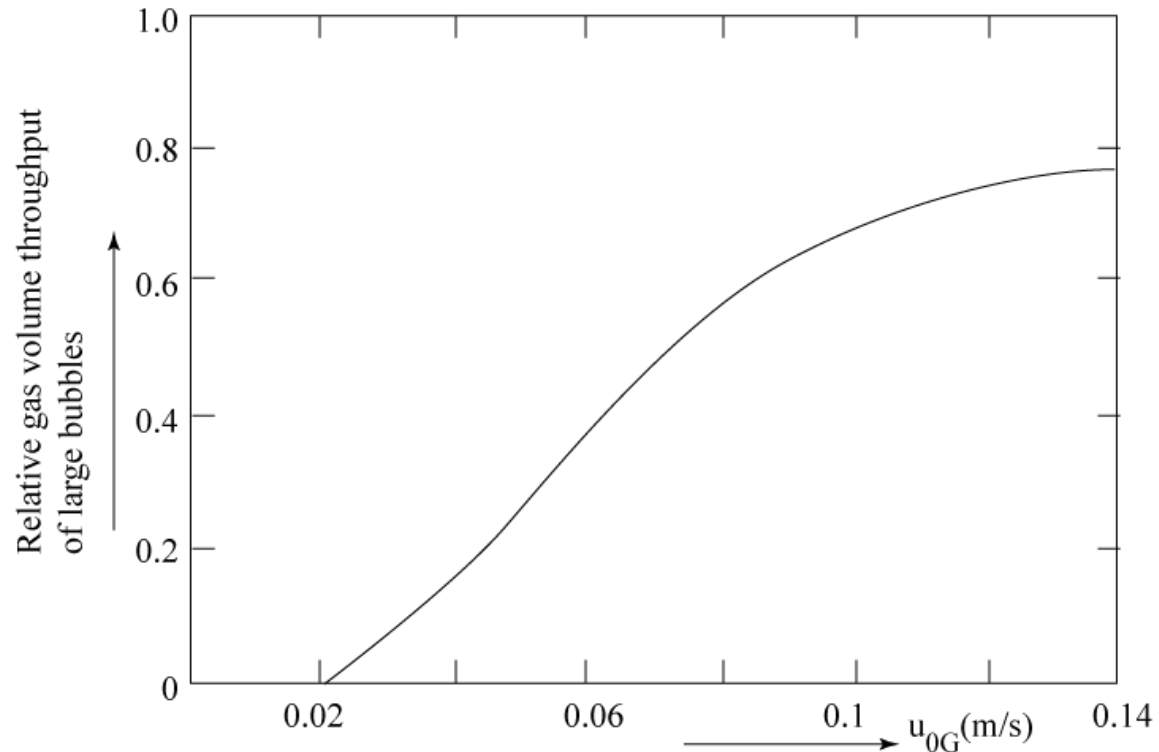


With solids

Hydrodynamics: Gas-liquid flow regimes



Fraction of gas phase in large bubbles (heterogeneous regime)



The figures should be adjusted for the effects of the particles present. At this time, for slurry systems no such flow maps are available in literature

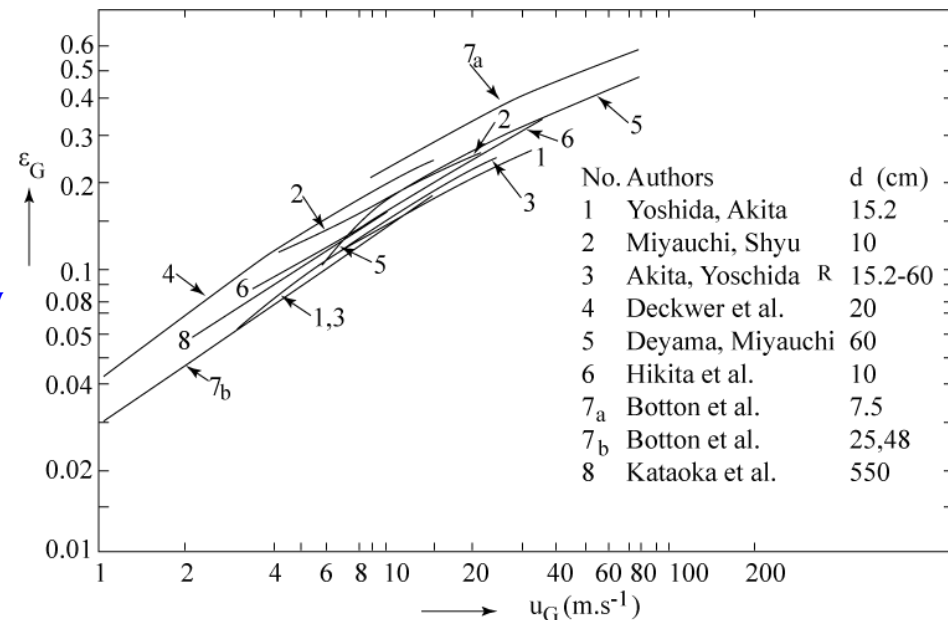
Hydrodynamics: Gas velocity and gas holdup

- The superficial gas velocity in sparged reactors (bubble columns) is very important since it determines not only the gas hold-up ε_G and (via $(1 - \varepsilon_G)$) the reaction phase volume, but it also determines the heat and mass transfer characteristics and mixing
- As a first approximation correlations for gas-liquid systems can be used

For bubble columns:

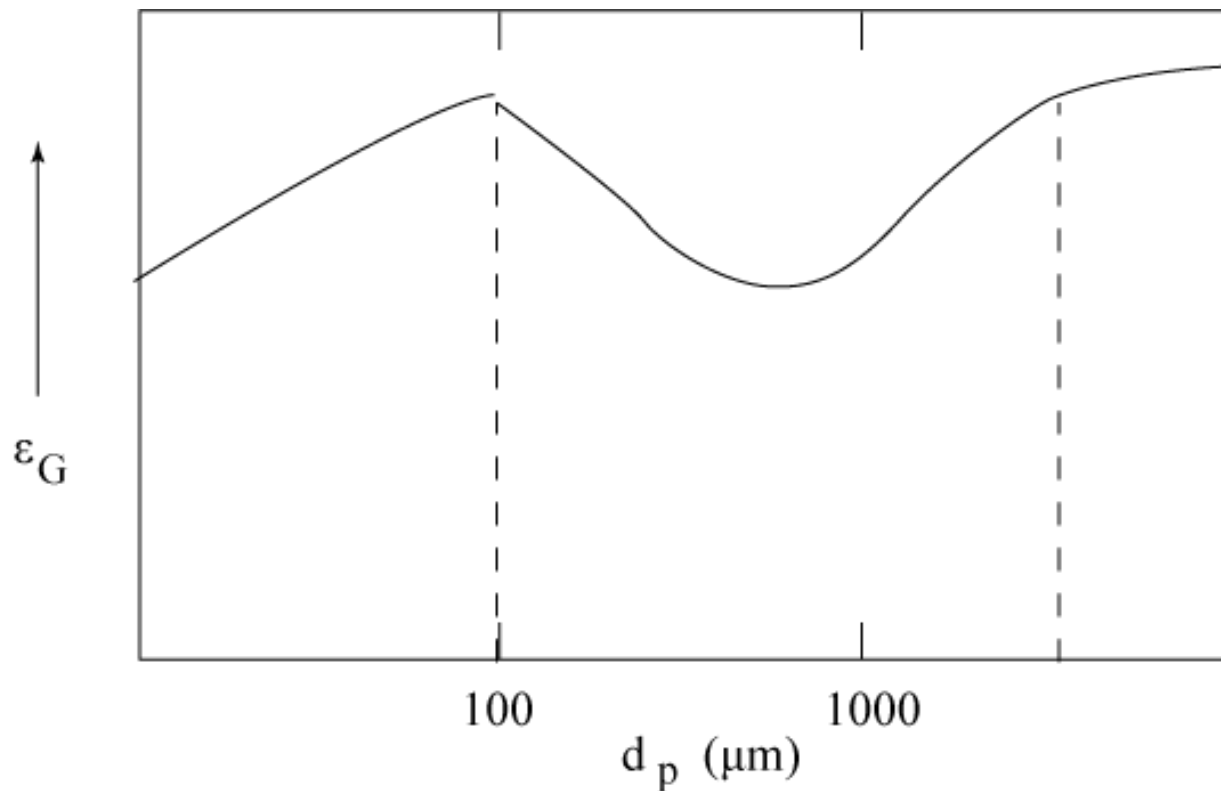
$$\varepsilon_G \equiv u_G^n \quad \begin{array}{l} 0.7 < n < 1.2 \text{ homogeneous flow} \\ 0.4 < n < 0.7 \text{ heterogeneous flow} \end{array}$$

$$\varepsilon_G = 0.0322 u_G^{-0.674} \quad \text{Slug flow}$$



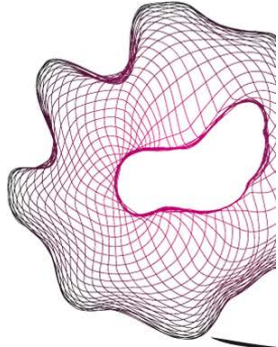
Hydrodynamics: Gas hold-up at low solids concentration

- A typical variation of ϵ_G with d_p given below



- However, not only the particle size but esp. the solids concentration is important

Hydrodynamics: Gas hold-up

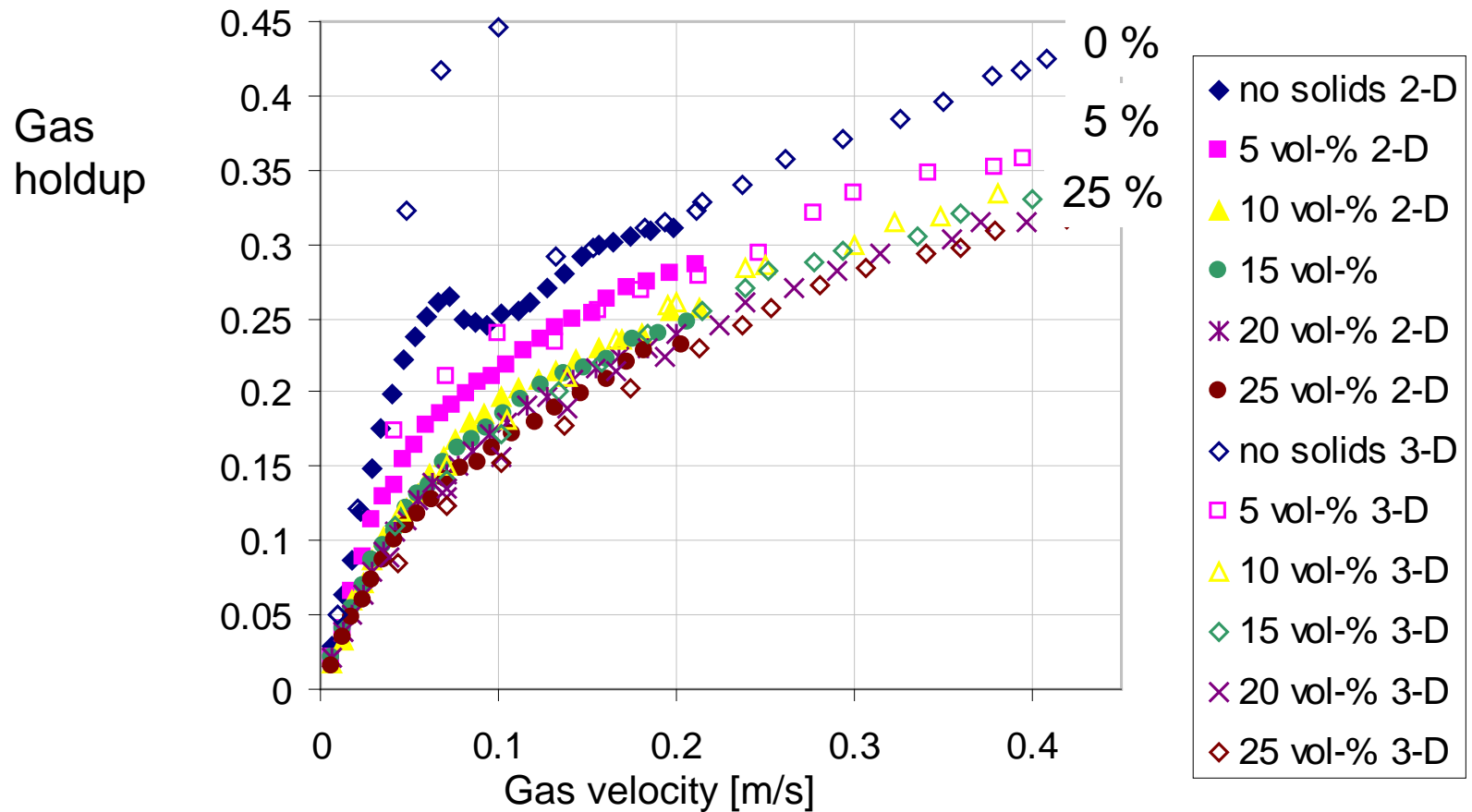


- Generally, the gas hold-up decreases with increasing solids content. For particles $< 100 \mu\text{m}$ however, ϵ_G first increases with solid concentration, until solid concentration reaches $\approx 0.6 \text{ wt\%}$. At higher solids loads ϵ_G decreases again
- A similar effect is found for ϵ_G vs. liquid viscosity in G/L systems
 - Two counteracting effect account for the maximum ϵ_G :
 - Higher viscosity promotes bubble coalescence ($\epsilon_G \downarrow$)
 - Drag force on bubble increases, u_{bubble} decreases ($\epsilon_G \uparrow$)
- At higher superficial velocities ($u_G > 0.1 \text{ m/s}$) the effect becomes insignificant

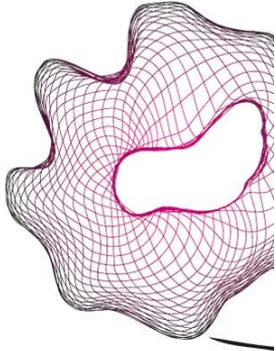


Example of ...

Effect of u_G and solids concentration on gas holdup



Hydrodynamics: State of mixing



- **Agitated vessels**

- 1st approximation: all phases (G,L,S) completely backmixed
- Empirical correlations

- Liquid phase
$$E_L = 0.116 \left(\frac{(d_R + H_{cL})^2 n}{\left(\frac{d_r}{d_s}\right)^2} \right) \quad [\text{Joshi, 1980}]$$

- Gas phase: 1-2 CISTRs in series model

- **Sparged columns**

- 1st approximation:
 - gas phase: plug flow
 - liquid phase: ideally mixed
 - solids phase: ideally mixed



Hydrodynamics: Liquid phase mixing

- Correlations for (axial) dispersion coefficient in G-L systems

- Liquid phase
$$E_L = 0.76 d_R^{\frac{4}{3}} (g u_G)^{\frac{1}{3}}$$

- Solid phase
$$E_{LS} = E_L = 0.76 d_R^{\frac{4}{3}} (g u_G)^{\frac{1}{3}}$$

- Gas phase
$$E_G = 56.4 d_R^{1.33} \left(\frac{u_g}{e_g} \right)^{3.56}$$

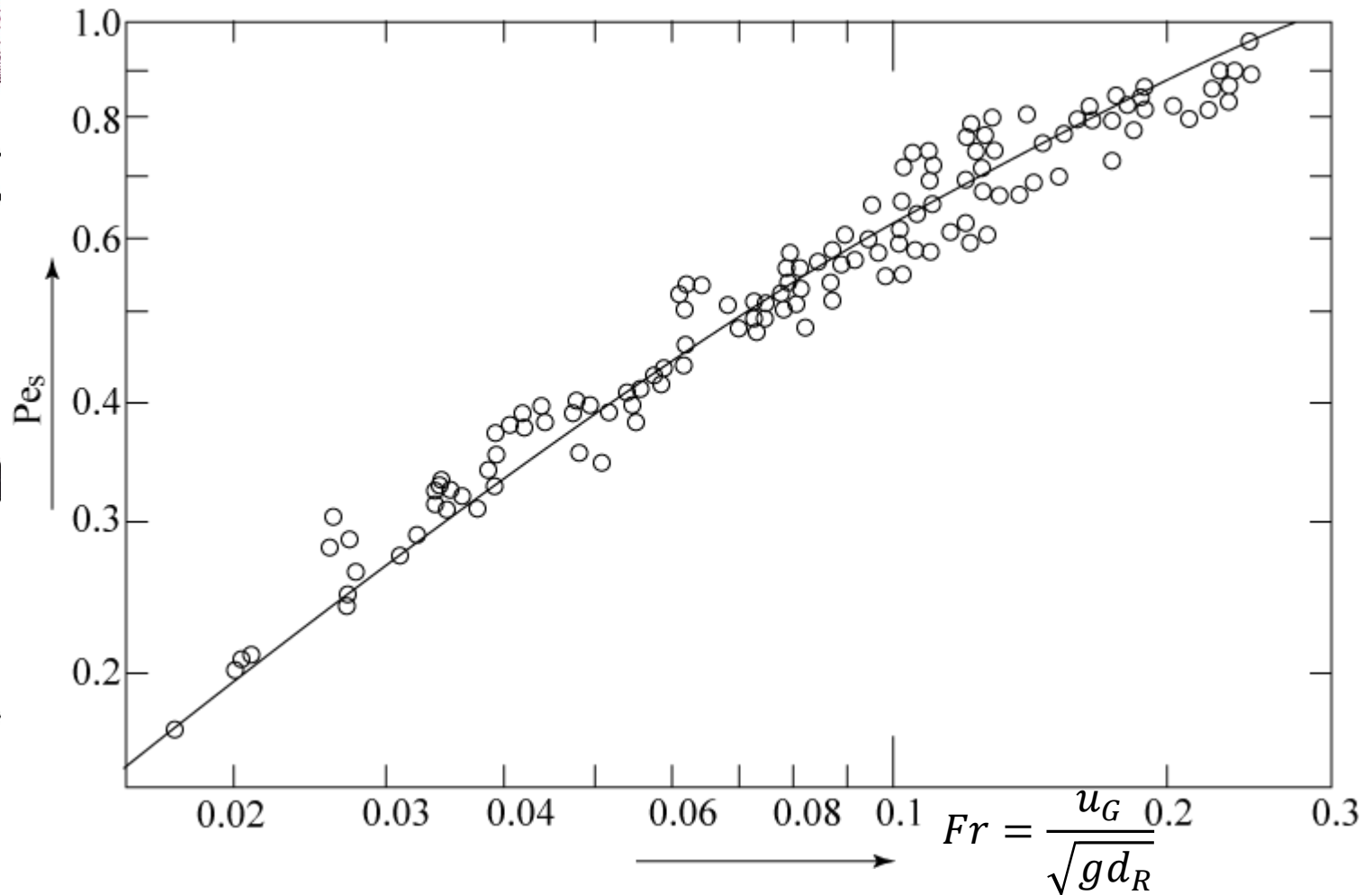
- Kato[1972] found:

- Gas-Liquid-Solid systems
$$Pe_L = \frac{u_g d_R}{E_L} = \frac{13 Fr^{\frac{1}{2}}}{1 + 8 Fr^{0.425}} \quad Fr = \frac{u_g^2}{g d_R}$$

$Pe_s = Pe_L$ for $70 < d_p < 180 \mu\text{m}$, solid load $\leq 2000 \text{kg/m}^3$

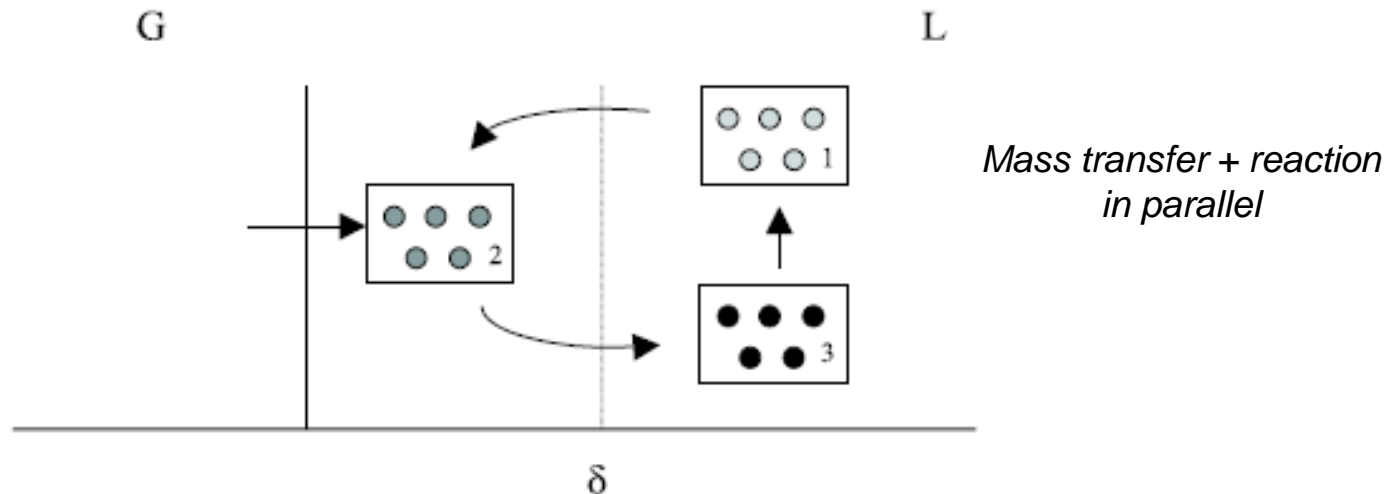
- Gas-Liquid
$$Pe_L = \frac{u_g d_R}{E_L} = \frac{13 Fr^{\frac{1}{2}}}{1 + 6.5 Fr^{0.4}}$$

Hydrodynamics: State of mixing solids



Mass Transfer in G-L-S (and G-L-L) systems

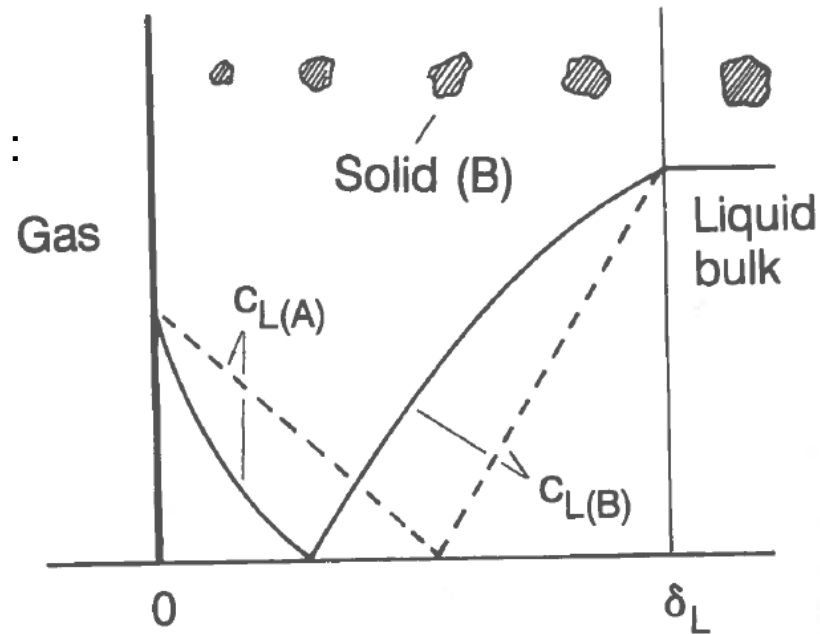
- For small particles (droplets) mass transfer near interface can be influenced



- Most often, mass transfer and reaction can be assumed to occur in series
- Effect of presence of fine particles in the liquid film near gas-liquid interface is a special situation and is not discussed in detail here.

SPECIAL CASE: REACTIVE PARTICLES INSIDE FILM

DISSOLVING PARTICLES
serving as **reactant supply** :

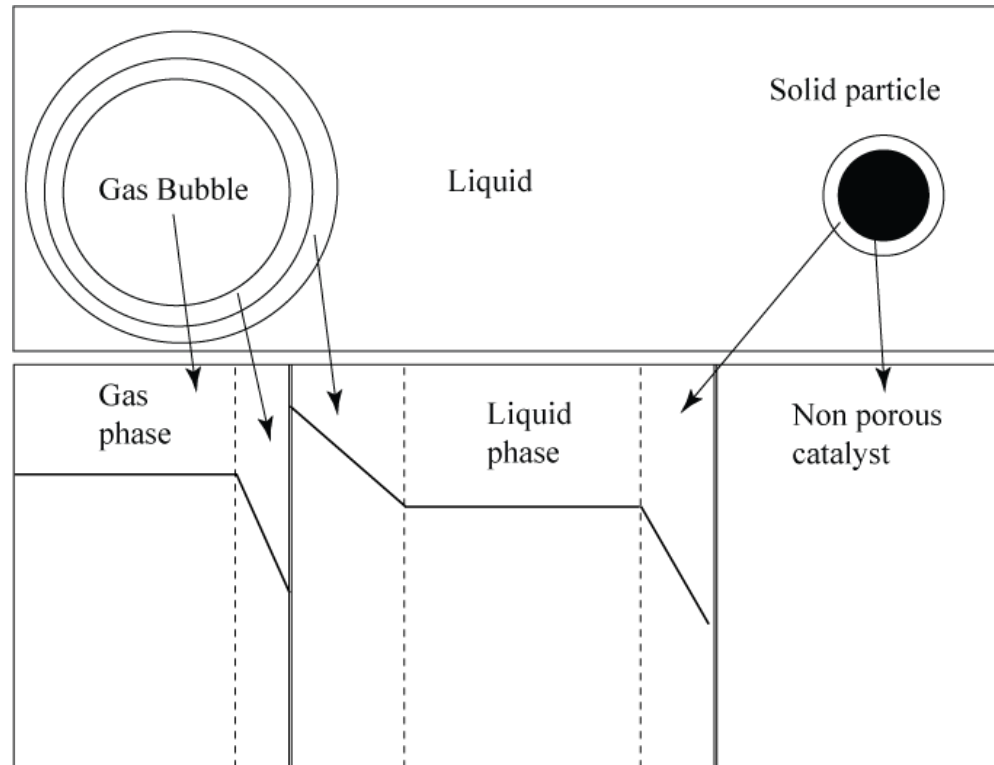


OR **CATALYST** PARTICLES ($d_p \ll \delta_L$) or **microbes /yeast** or ...

Mass transfer

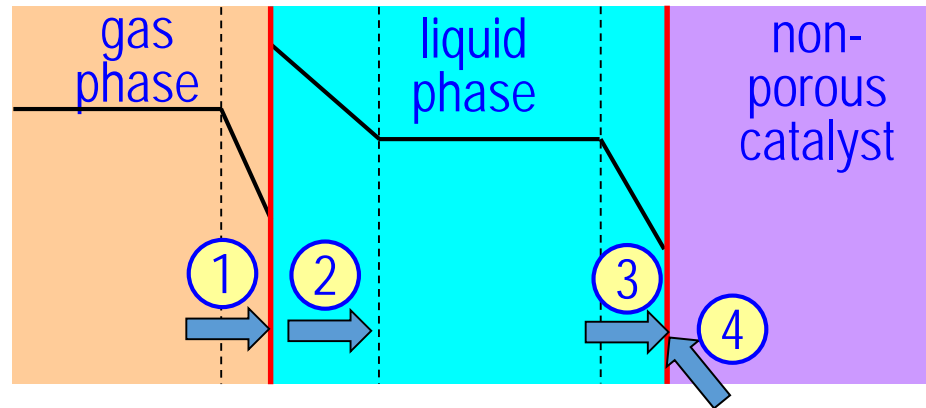
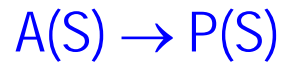
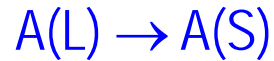
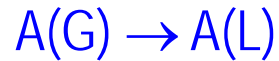
Mass transfer and chemical reaction in series

In case of a non porous particle and a surface reaction



$$N_A = \frac{c_{A,G}}{\frac{1}{k_G a_{GL}} + \frac{1}{mk_L a_{GL}} + \frac{1}{mk_{LS} a_S} + \frac{1}{mk_1'' a_S}}$$

Reaction system:



1) Gas bulk to gas-liquid interface:

$$N_{A,G} = J_A a_{GL} = k_G a_{GL} (c_{A,G} - c_{A,i,G})$$

2) Gas-liquid interface to liquid bulk:

$$N_{A,L} = J_A a_{GL} = k_L a_{GL} (c_{A,i,L} - c_{A,L})$$

3) Liquid bulk to catalyst surface:

$$N_{A,S} = J_A a_S = k_{LS} a_S (c_{A,L} - c_{A,S})$$

4) Reaction at catalyst surface:
(1st order irreversible reaction assumed)

$$R_A = -k_1'' a_S c_{A,S}$$

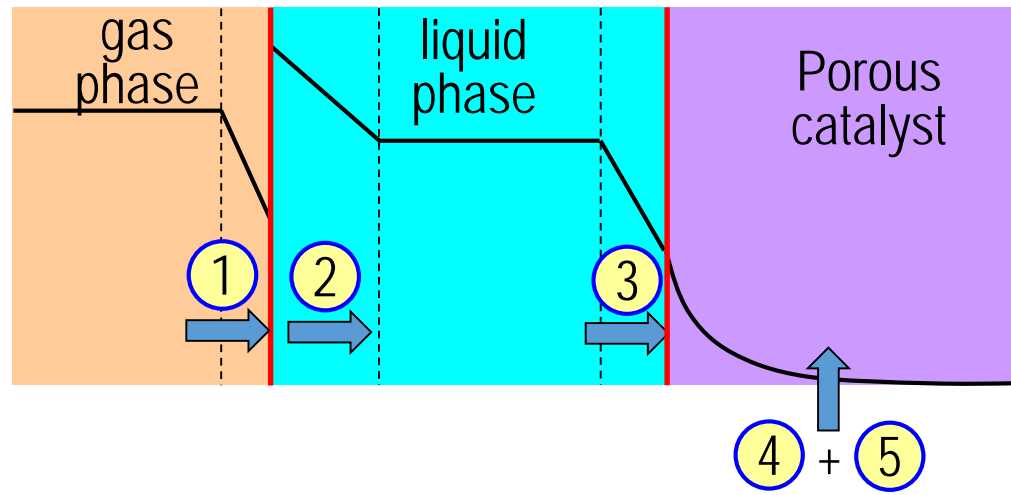
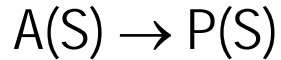
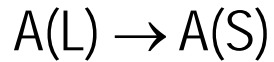
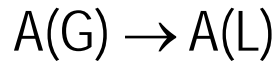
Thermodynamic equilibrium at gas-liquid interface: $m = c_{A,i,L} / c_{A,i,G}$

Note that flows (N, R) are in mole/m³_{Reactor}s!!!

In steady state all flows are equal:

$$N_A = \frac{c_{A,G}}{\frac{1}{k_G a_{GL}} + \frac{1}{m k_L a_{GL}} + \frac{1}{m k_{LS} a_S} + \frac{1}{m k_1'' a_S}}$$

Reaction system:



Mass transfer resistances:

- in series {
- 1) Mass transfer from gas bulk to gas-liquid interface
 - 2) Mass transfer from gas-liquid interface to liquid bulk
 - 3) Mass transfer from liquid bulk to catalyst surface
- parallel {
- 4) Diffusion inside the catalyst particle
 - 5) Reaction at catalyst surface

$$N_A = \frac{C_{A,G}}{\frac{1}{k_G a_{GL}} + \frac{1}{mk_L a_{GL}} + \frac{1}{mk_{LS} a_s} + \frac{1}{mk_1 \epsilon_s \eta}} \longrightarrow \eta \approx \frac{\tanh(\phi)}{\phi} \longrightarrow \phi = \frac{V}{A} \sqrt{\frac{k_n'' a_{int}}{D_{A,eff}}} \quad (\text{for 1st order irr. kinetics})$$

Mass transfer

- Reactor modeling requires information about
 - Reaction kinetics
 - Reaction rate expression
 - Kinetic parameter values
 - Thermodynamics
 - Mass transfer parameters
 - k_G, a or $k_G \cdot a$
 - k_L, a or $k_L \cdot a$
 - k_{LS}, a_P
 - D_{ip}
 - Physical properties

Mass transfer

- **Determination of ($k_L \cdot a$)**

- For low solid concentrations the gas-liquid relations can be used:

- Agitated vessel
$$k_L a = b_0 \left(\frac{P_G}{V_R} \right)^m u_G^n$$
 [van 't Riet, 1979]

$$500 < P_G/V_R < 10000 \text{ W/m}^3; 0.002 < V_R < 4.4 \text{ m}^3$$

$$\text{Pure water: } b_0 = 0.026; m = 0.4; n = 0.5$$

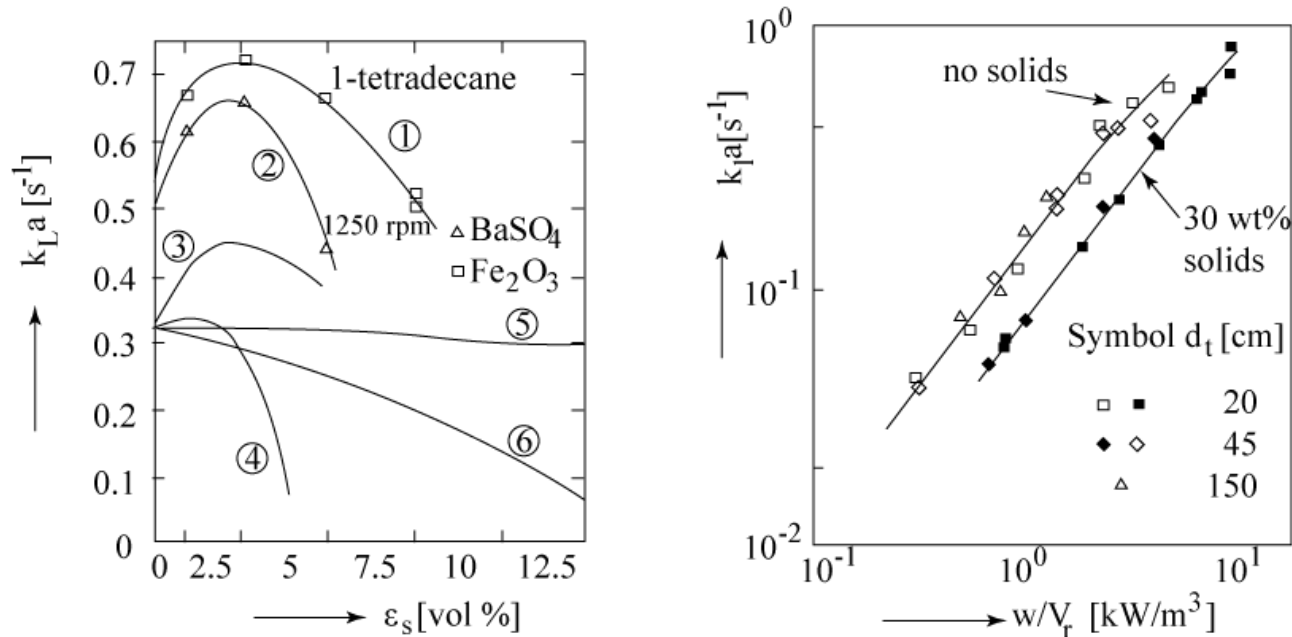
$$\text{Ionic soln: } b_0 = 0.002; m = 0.7; n = 0.2$$

- Bubble column
$$\frac{k_L a d_R^2}{D} = 0.6 \left(\frac{v_l}{D} \right)^{0.5} \left(\frac{g d_R^2 \rho_L}{s} \right)^{0.62} \left(\frac{g d_R^3}{v_L^2} \right)^{0.31} \varepsilon_G^{0.31}$$
 [Akita, Yoshida]

When the slurry can be considered pseudo homogeneous, a correlation with the effective suspension viscosity can be applied :

$$k_L a = (k_L a)_0 \left(\frac{\mu_{eff}}{\mu} \right)^{0.42}$$

Experimental data stirred slurry reactors



The following equation was determined from this data (for $\mu_{eff} > 1.3 \mu_0$) [Oguz, 1987]

$$\frac{k_L a d_R^2}{D} = 0.162 \left(\frac{n d_s^2 \rho_{slurry}}{\mu_{eff}} \right)^{1.5} \left(\frac{\mu_{eff}}{\rho_{slurry} D} \right)^{0.5} \left(\frac{n^2 d_s}{g} \right)^{0.19} \left(\frac{\sigma_L}{\mu_{eff} \mu_G} \right)^{0.6} \left(\frac{d_s}{n_s u_G} \right)^{0.09} \sqrt{\frac{\sigma_{H_2O}}{\sigma_L}}$$

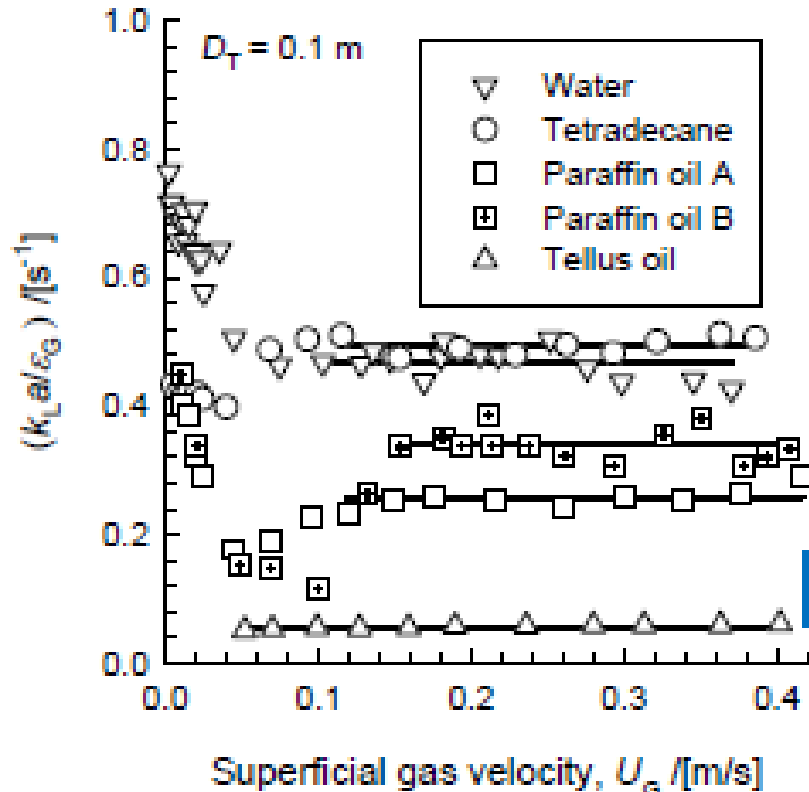
Experimental determination

Ideally stirred tank $J_a = k_L a m C_{A,G} = -\Phi_A / V_r$;

with R_2 (G/L – L-side) $\gg R_1, R_2, R_3, R_4$ and $C_{A,bulk} = 0$

Also for sparged slurry reactors :

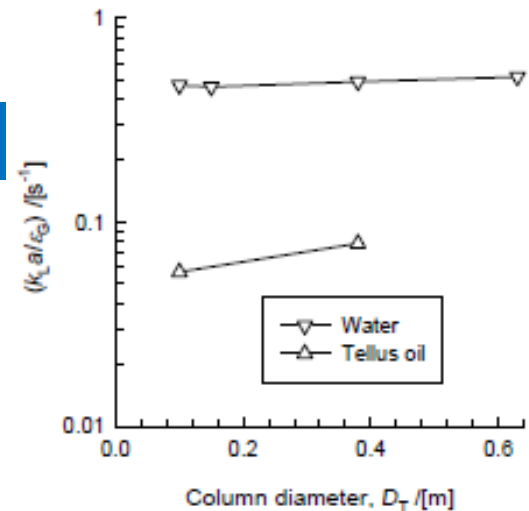
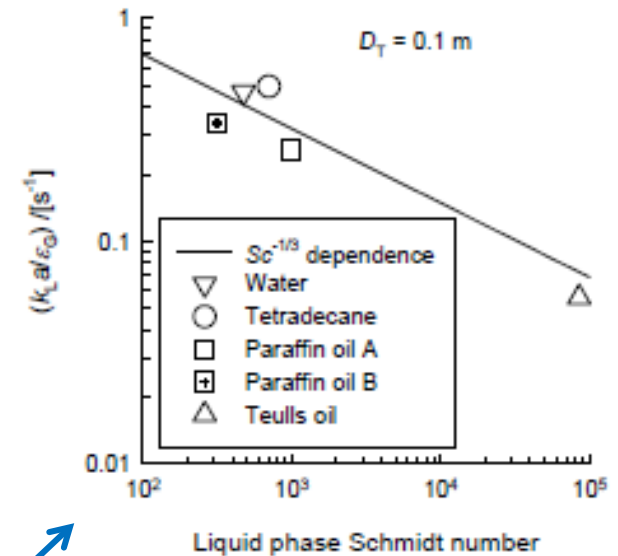
$k_L a$ in heterogeneous regime



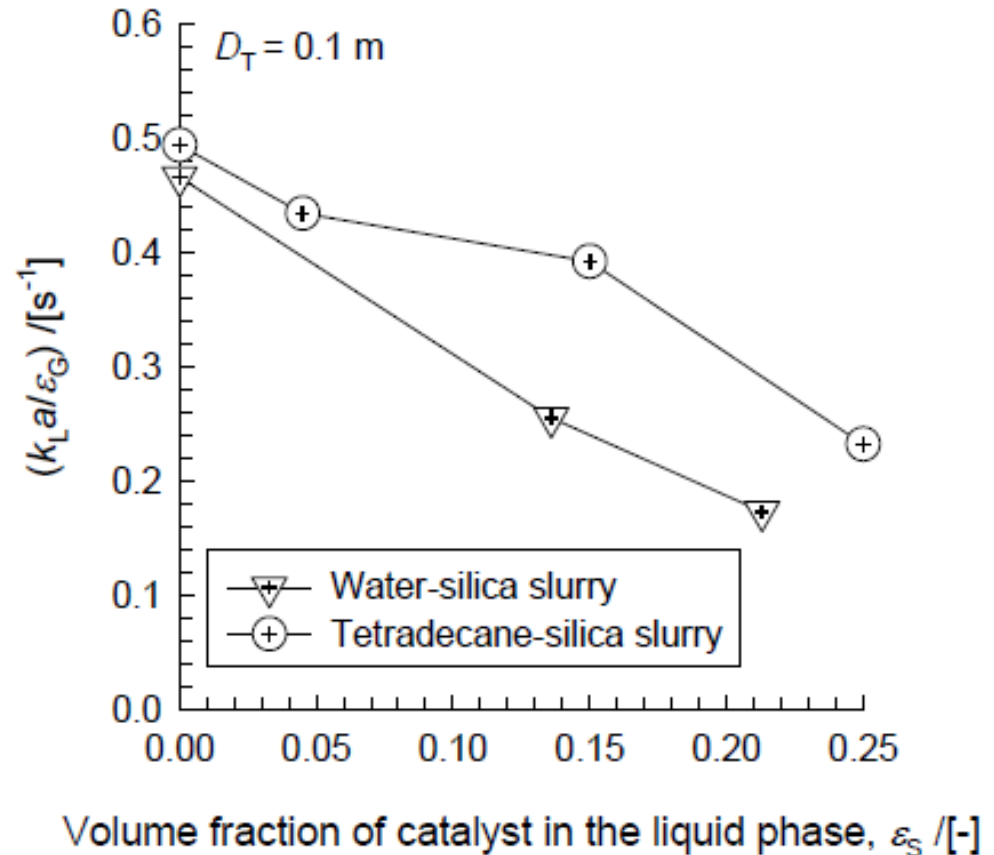
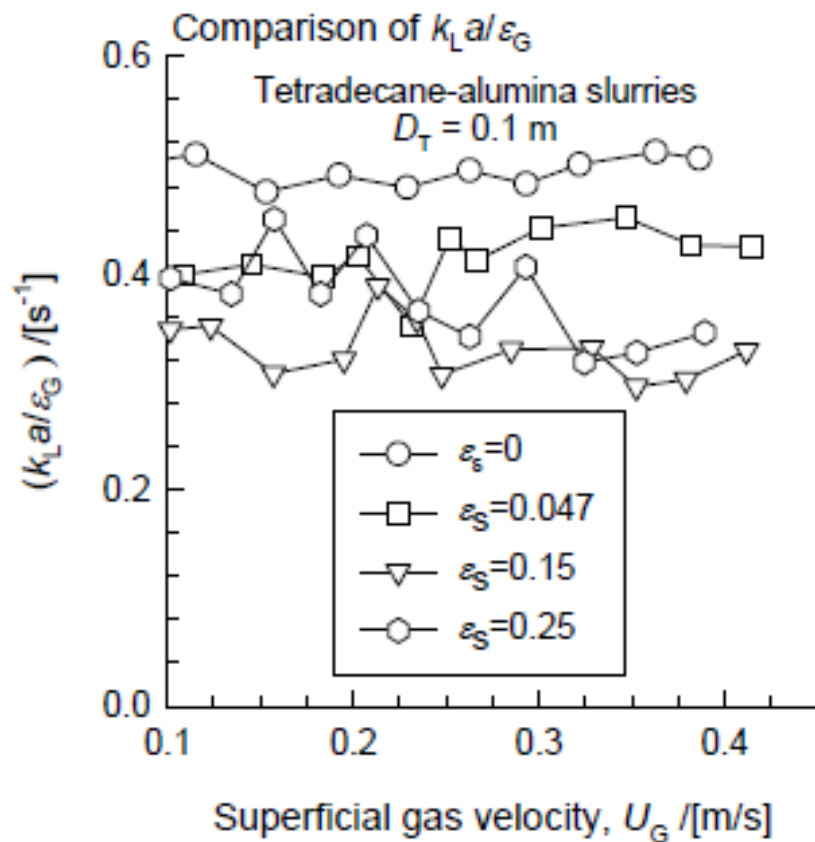
$k_L a / \epsilon_G \approx \text{constant}$

Effect liq. phase

Effect D_R



$k_L a / \varepsilon_G$ in the heterogeneous regime vs. ε_S



Determination of the specific contact area a

- Using $a=6\varepsilon_G/d_{b,\max}$ and the isotropic turbulence theory, the following correlation is found

$$a \cong \frac{\rho_L^2 \epsilon^{0.4}}{\sigma^{0.6}} \varepsilon_G \quad \epsilon \text{ is the energy input (W/kg)}$$

- This structure is found in many semi-empirical correlations:

- Agitated vessel
(Shridhar, Potter)

$$a = 1.44 \left(\frac{\left(\frac{P_{mech}}{V_R} \right) \rho_L^{0.2}}{\sigma_L^{0.6}} \right) \left(\frac{u_G}{u_b} \right)^{0.5} \left(\frac{P_{total}}{P_{mech}} \right) \left(\frac{\rho_G}{\rho_{air}} \right)^{0.16}$$

- Bubble column $a=6\varepsilon_G/d_{b,\max}$ and $d_b, \varepsilon_G = f(\varepsilon, \mu, u_G, \sigma, \dots)$

Literature correlations exist for ε_G and d_b , see e.g. Akita&Yoshida, van Dierendonk,...

MASS TRANSFER - Determination of the specific contact area a

- Experimental determination

- Measure d_b , ε_G :

$$a = 6 \varepsilon_G / d_b$$

- Chemical methods with $E_a = Ha$

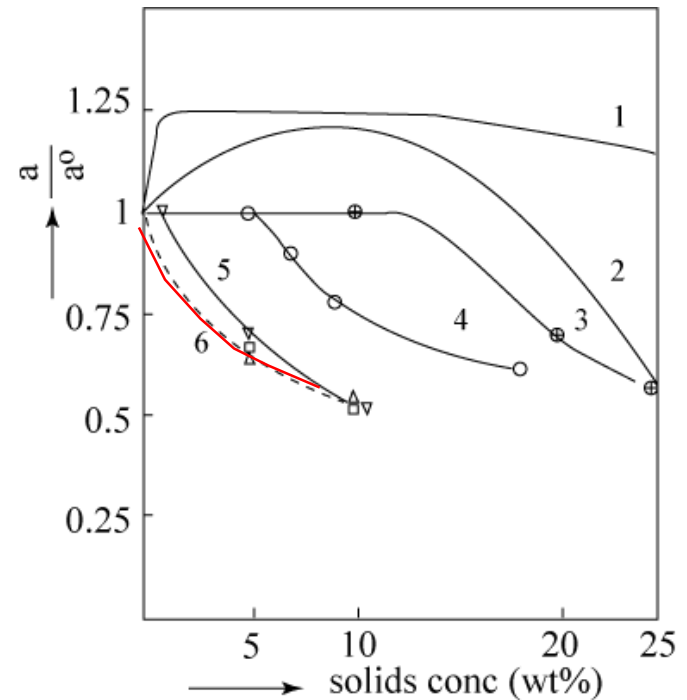
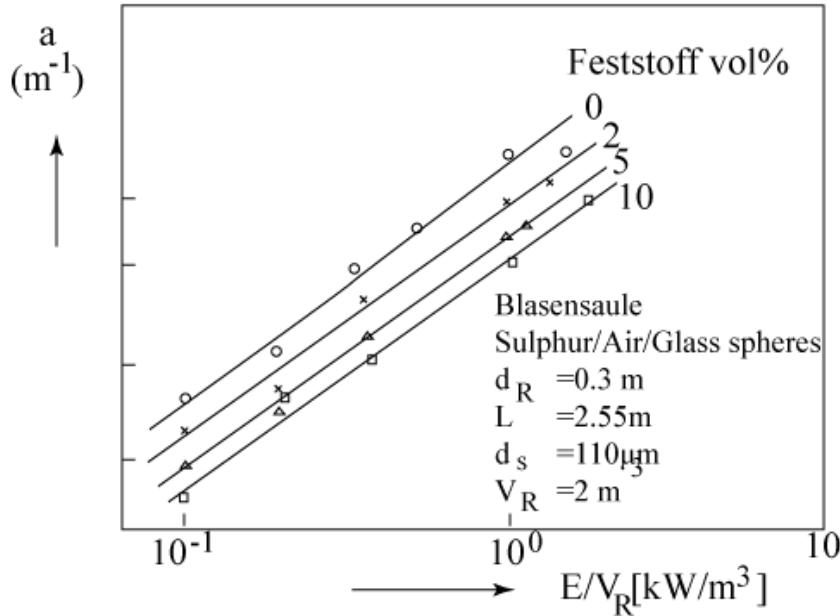
$$JaV = m \sqrt{(k_1 D)} C_G a V$$

- Using the chemical method at a high gas conversion level can lead to serious error in the estimation of a .
 - At high gas conversion level the small bubbles may be depleted in the gas phase reactant. The contribution of these bubbles to the overall interfacial area will then be underestimated

- Influence of solid particles

- Generally, addition of particles decreases ε_G and, with that, area a
However a small amount (<0.6 wt%) of fine particles ($d_p < 100 \mu m$) may increase a up to 50%

Determination of the specific contact area a



$$a = (a)_0 \left(\frac{\mu_{eff}}{\mu} \right)^{-0.24}$$

Est. Method (Schumpe, 1987)

Relative change of gas-liquid interfacial area as a function of solids concentration

- 1 7 μm Ca(OH)_2 particles in BC [46]
- 2 2 μm Mg(OH)_2 particles in BC [46]
- 3 0-10 μm activated carbon in BC; $u = 8 \text{ cm/s}$
- 4 0-10 μm Kieselguhr in BC; $u = 8 \text{ cm/s}$
- 5 0-10 μm Al_2O_3 in BC; $u = 8 \text{ cm/s}$
- 6 100 μm glass beds in STR
 $W/V_R = 0.1 \text{ kW/m}^3$
 $W/V_R = 1 \text{ kW/m}^3$

BC = bubble column

STR= stirred tank reactor

MASS TRANSFER: Experimental determination of k_L

- Using $k_L = (k_L \cdot a)/a$
- Danckwerts plot:

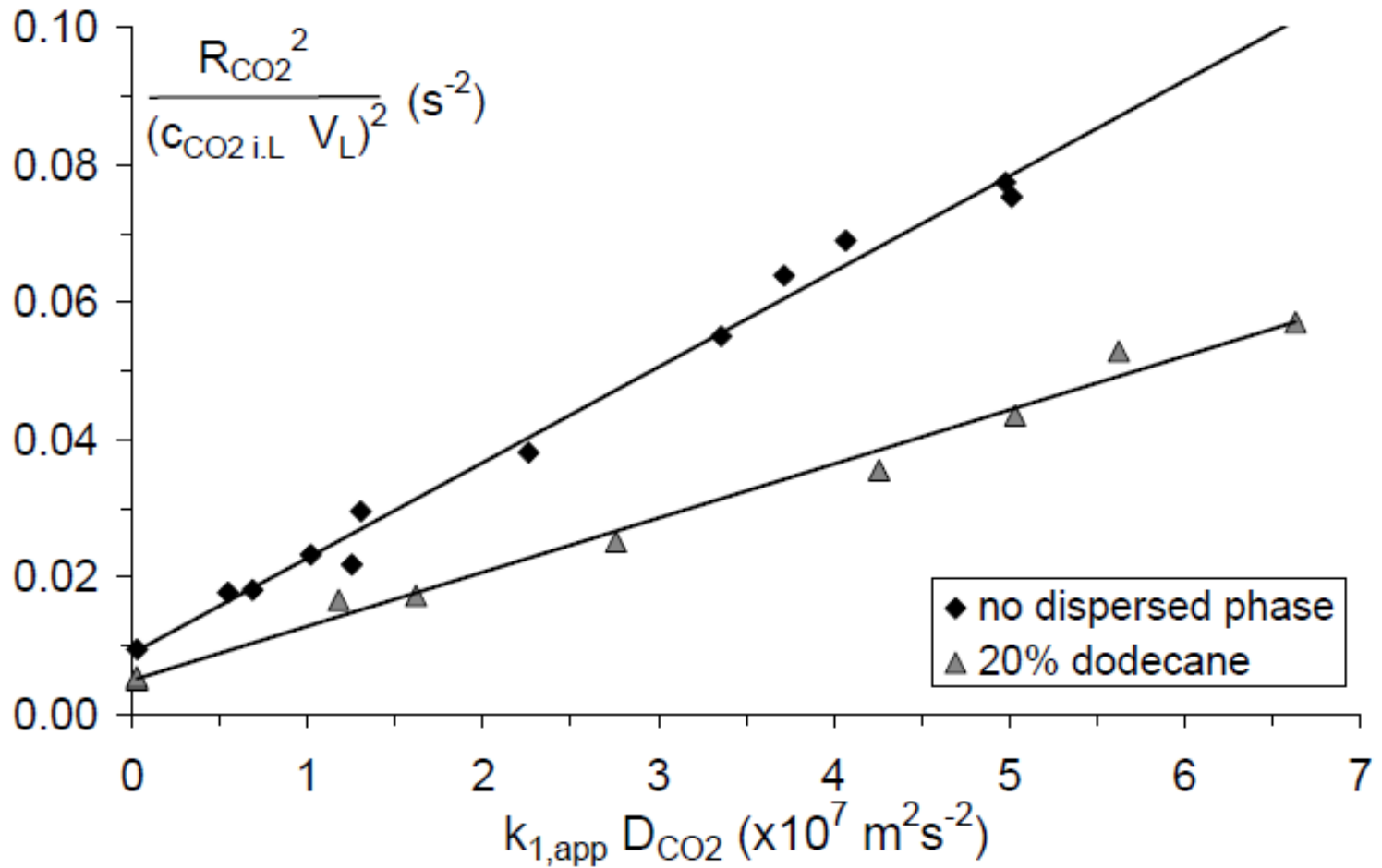
for (1,1) irreversible reaction kinetics, no diffusion limitation for comp. B and $C_{A,L}=0$:

flux of component A:
$$J_a = mC_{A,G} \sqrt{Dk_{1,1}C_{B,L} + k_L^2}$$

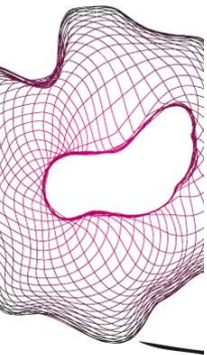
From mass balance for a stirred tank reactor:
$$\frac{-F_A}{mC_{A,G}V_R} = a \sqrt{Dk_{1,1}C_{B,L} + k_L^2}$$

by changing e.g. $k_{1,1}$ or $C_{B,L}$ and plotting of $\left(\frac{-F_A}{mC_{A,G}V_R}\right)^2$ vs. $k_{1,1} \cdot C_{B,L} \cdot D$ values for $k_L a$ and k_L are obtained

Example: Danckwerts-plot result



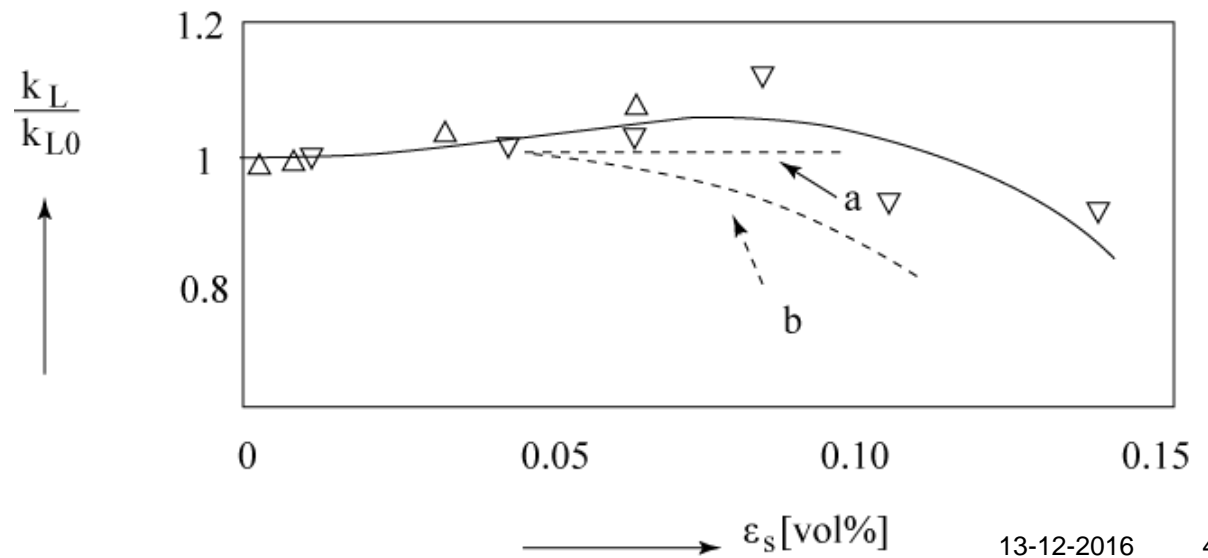
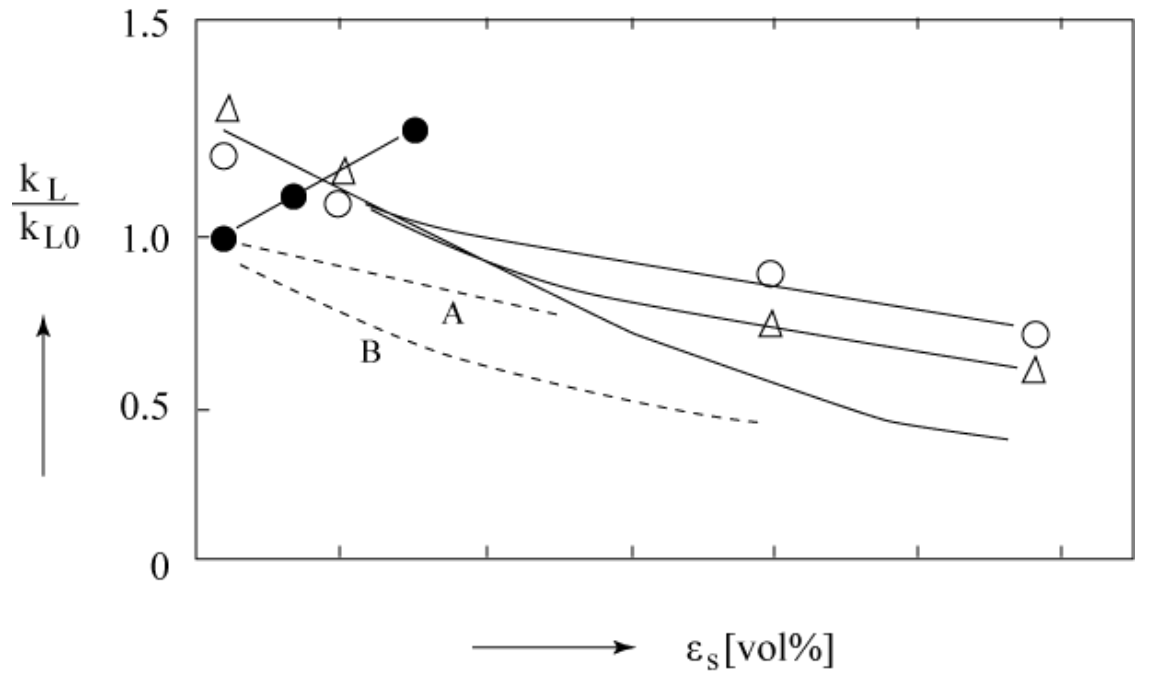
Experimental determination of k_L



Bubble column



Stirred cell



Experimental determination of k_L

- Experiments in a stirred cell showed that fine, inert particles hardly effect k_L at solid loads < 10 wt%
- At higher loads the increasing effective viscosity is likely to cause the observed decrease in k_L value
- For $\rho_s \approx \rho_L$ the slurry may be considered pseudo homogeneous.
for this situation it was found:

$$k_L = 0.3 \left(\frac{\epsilon \rho_L}{\mu} \right)^{\frac{1}{4}}$$

(Newtonian fluid) [Kawase& Moo Young]

Experimental determination of k_L

- Semi-empirical correlations for gas-liquid systems:

- Sparged columns
$$k_L = C_1 \left(\frac{(\rho_L \rho_g) \mu_L g}{\rho_L^2} \right)^{\frac{1}{3}} \left(\frac{\mu_L}{\rho_L D} \right)^{C_2}$$

- Small bubbles ($d_b < 2.5$ mm): $C_1=0.0031$ $C_2=-1/3$
- Large bubbles ($d_b > 2.5$ mm): $C_1=0.0042$ $C_2=-1/2$
[Calderbank, Moo-Young]

- Stirred tank

$$k_L = 0.592 D_L^{0.5} \left(\frac{\epsilon \rho_L}{\nu} \right)^{0.25} \quad [\text{Prasher, Willis}]$$

Mass transfer at liquid solid interface

$$\lambda_K = \left(\frac{\nu^3}{\varepsilon} \right)^{1/4}$$

- Sphere particle in infinite field $k_s = \frac{2D}{d_p}$

$$\tau = \left(\frac{\nu}{\varepsilon} \right)^{1/2}$$

$$\nu = (\nu \varepsilon)^{1/4}$$

- For flow around particles: $Sh = 2 + C Re^n Sc^m$
e.g. Ranz and Marshall: $C = 0.6$, $n = 1/2$, $m = 1/3$

- Re is usually based on particle slip velocity. This causes complications when $\rho_s \approx \rho_L$

Using $Re_p = \frac{u d_p}{\nu_L}$

& the isotropic turbulence theory it follows: $Sh_p = 2 + C \left[\frac{d_p^{4/3} \varepsilon^{1/3}}{\nu_L} \right] Sc^m$

Mass transfer at liquid-solid interface

- The group $Re' = \left[\frac{d_p^{4/3} \epsilon^{1/3}}{v_L} \right]$

is the new definition of Re to be applied in the Sh-relations

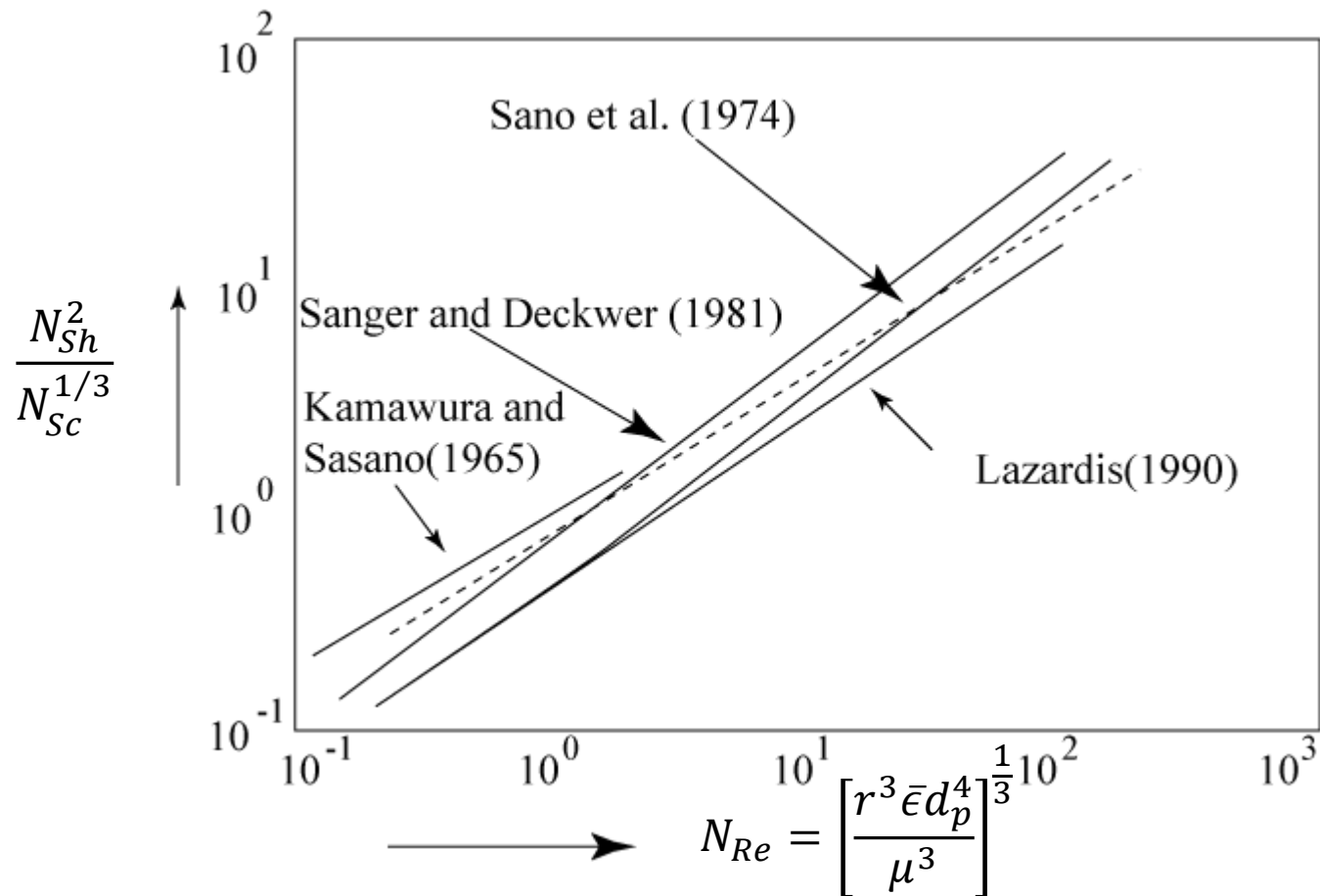
- Correlations from literature

▪ Sparged columns		$d_p[\text{m}]$	$u_G[\text{cm/s}]$
[Sano]	$Sh = 2 + 0.4Re'^{0.25}Sc^{1/3}$	60-2900	0.5-17
[Sänger]	$Sh = 2 + 0.545Re'^{0.264}Sc^{1/3}$	1100-8700	1.0-10
[Jadhav]	$Sh = 2 + 0.0914Re'^{0.265}Sc^{0.45}$	500-1100	2.0-10

▪ Stirred slurry reactors		$d_p[\text{m}]$
[Sano]	$Sh = 2 + 0.4Re'^{0.75}Sc^{1/3}$	60-1500
[Marrone]	$Sh = 2 + 0.36Re'^{0.75}Sc^{1/3}$	17-90
[Asia]	$Sh = [2^{5.8} + (0.61Re'^{0.58}Sc^{0.45})^{5.8}] / 5.8$	5-970

Literature correlations for k_s

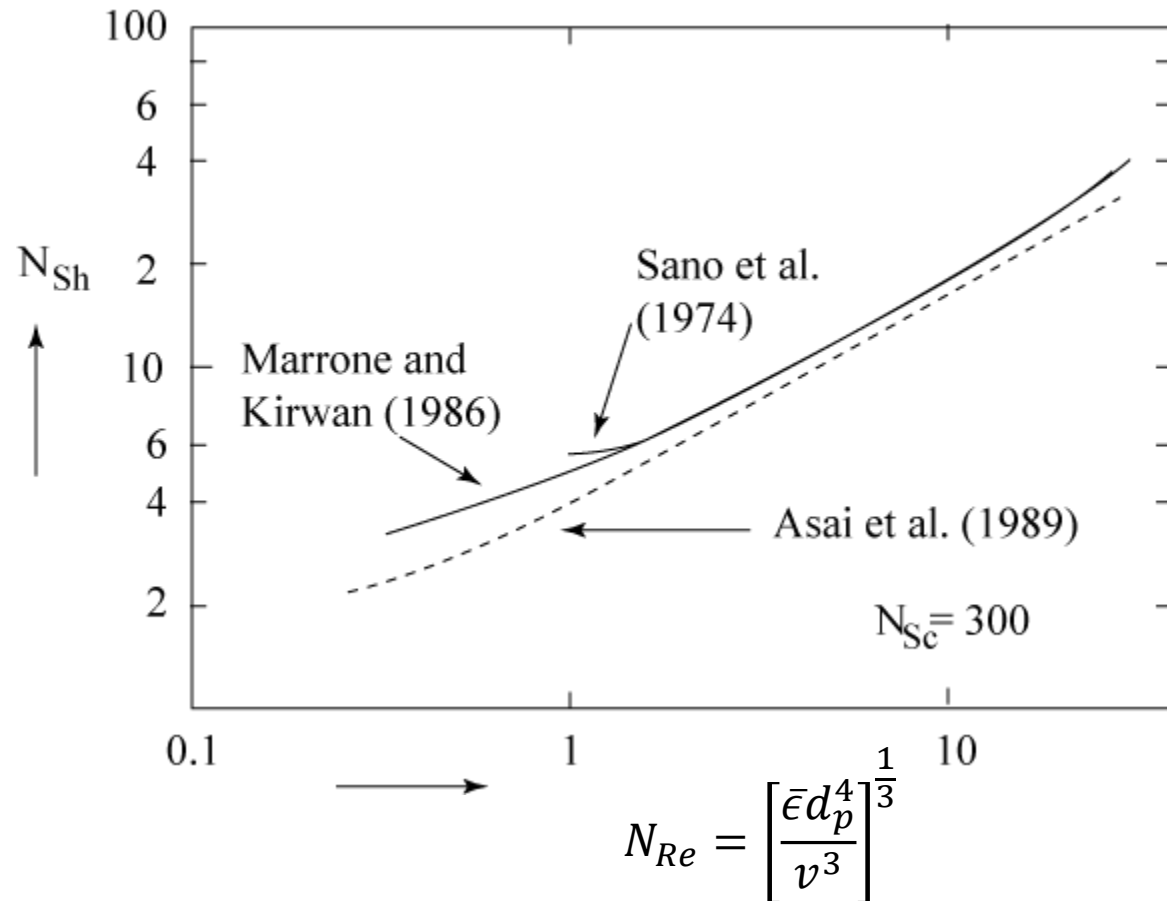
The literature correlations for k_s match well
Sparged colums



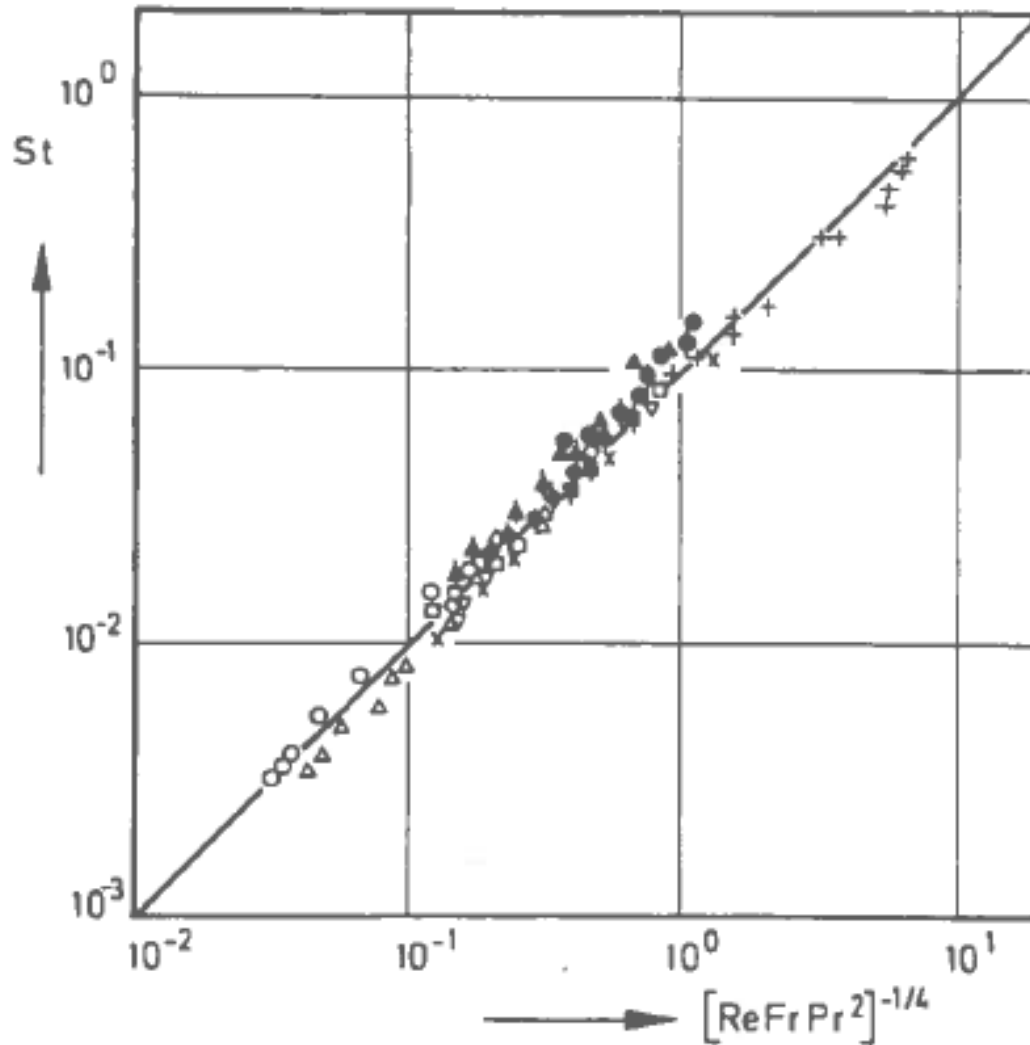
Literature correlations for k_s

agitated slurry reactors

k_s is a relatively well known parameter and can be estimated using literature correlations (always check the validity range)



Heat transfer in Bubble Columns



$$h = 0.1 \left[C_P \rho \lambda \left(\frac{u_G g}{\eta_L} \right)^{0.5} \right]^{0.5}$$

Deckwer-corr.

Heat transfer

Heat transfer is affected by solids holdup

- Solids type and solids concentration is important

These effects are attributed to intrinsic properties of the particles and their effect on slurry viscosity

NB: kieselgur = High porous, diatomaceous earth
10-200 μm particles,
Composition: 80-90% silica; rest = Al and Fe oxides

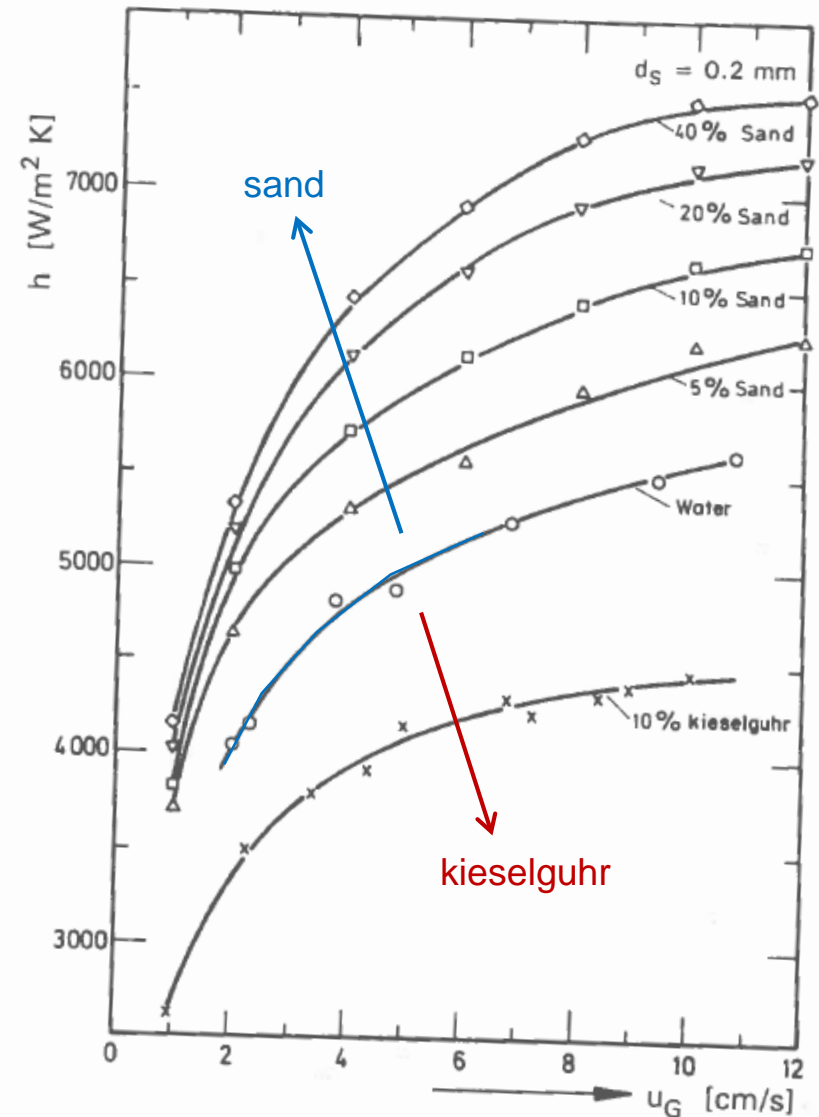
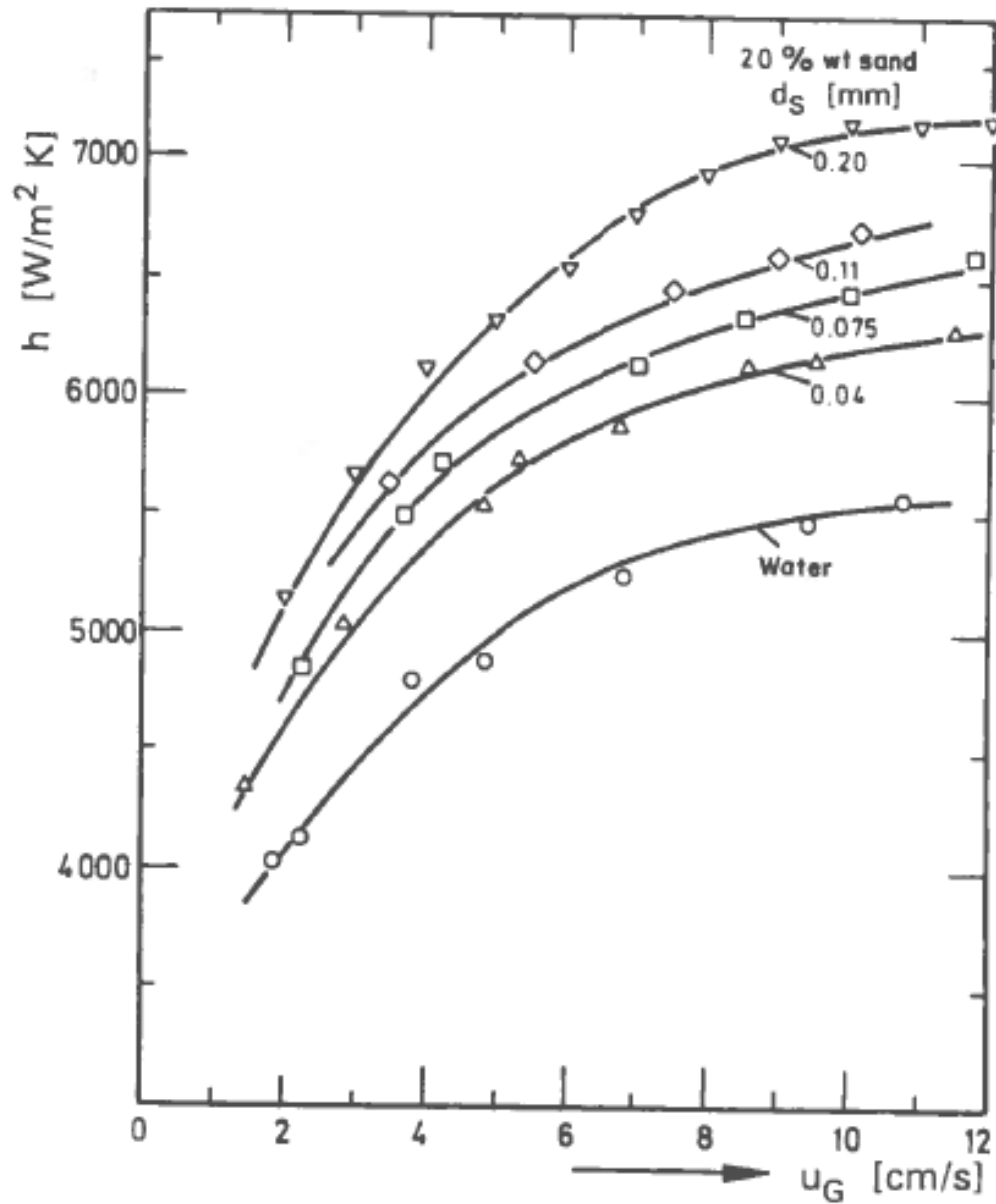


FIGURE 2.5 Effect of solid content on heat transfer coefficients in slurry bubble columns (Kölbel *et al.*, 1958b and 1960).

Heat transfer: Effect of particle size



Heat transfer correlation

$$h = 0.1 \cdot \left[c_P \rho \lambda \cdot \left(\frac{u_G g}{\mu} \right)^{0.5} \right]^{0.5}$$

Heat transfer coefficient h

Apply slurry properties for slurry systems, liquid properties for G-L systems)

- Density $\rho = \varepsilon_L \cdot \rho_L + \varepsilon_S \cdot \rho_S$
- Heat capacity $c_P = w_L \cdot c_{P,L} + w_S \cdot c_{P,S}$
- Viscosity $\mu = \mu_L \cdot (1 + 4.5 \cdot \varepsilon_S) \quad (\varepsilon_S < 0.25)$
- Conductivity $\lambda_{sl} = \lambda_L \cdot \frac{2\lambda_L + \lambda_S - 2\varepsilon_S(\lambda_L - \lambda_S)}{2\lambda_L + \lambda_S + \varepsilon_S(\lambda_L - \lambda_S)}$

Resume

- ✓ Slurry systems; stirred and sparged
 - ✓ Reactor types
 - ✓ Liquid-solid separation
- ✓ Effect solids content on
 - ✓ Slurry density, viscosity
 - ✓ Gas holdup
 - ✓ Regime transition
 - ✓ Mass transfer parameters (k_L , a , kLa)
- ✓ Heat transfer
- ✓ Slurry Reactor design using Matlab: F-T (see hand out)