UNIVERSITEIT TWENTE.

Multiphase Reactor Technology
HC 3: Fluidization

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Resume Fixed Bed

- What is a fixed bed
- Operational options
- Limitations in using fixed bed
- Micro kinetics
- H&M Transfer coefficients
- Different models for design

Contents Fluidized Bed

- Learning objectives
  - ✓ Be able to describe what a fluidized bed is
  - ✓ Be able to evaluate the characteristics of solids to be used in fluidized bed reactors
  - ✓ Be able to evaluate the characteristics (FD) of a fluidized bed
  - ✓ Be able to design a fluidized bed reactor for a given process

Contents

- Introduction
  - ✓ Phenomenon of fluidization
  - ✓ Applications of fluidization in physical and chemical technology

- Fluidization regimes
  - ✓ Characterization of particulate solids (PSD)
  - ✓ Geldart's classification
  - ✓ Minimum fluidization velocity u_{mf} and terminal velocity u_t
  - ✓ Pressure- and temperature dependence of fluidization
  - ✓ High velocity fluidization

Contents

- Dense fluidized beds

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

  - ✓ 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

Contents

- Recent developments

  ✓ Applications of fluidized bed chemical reactors

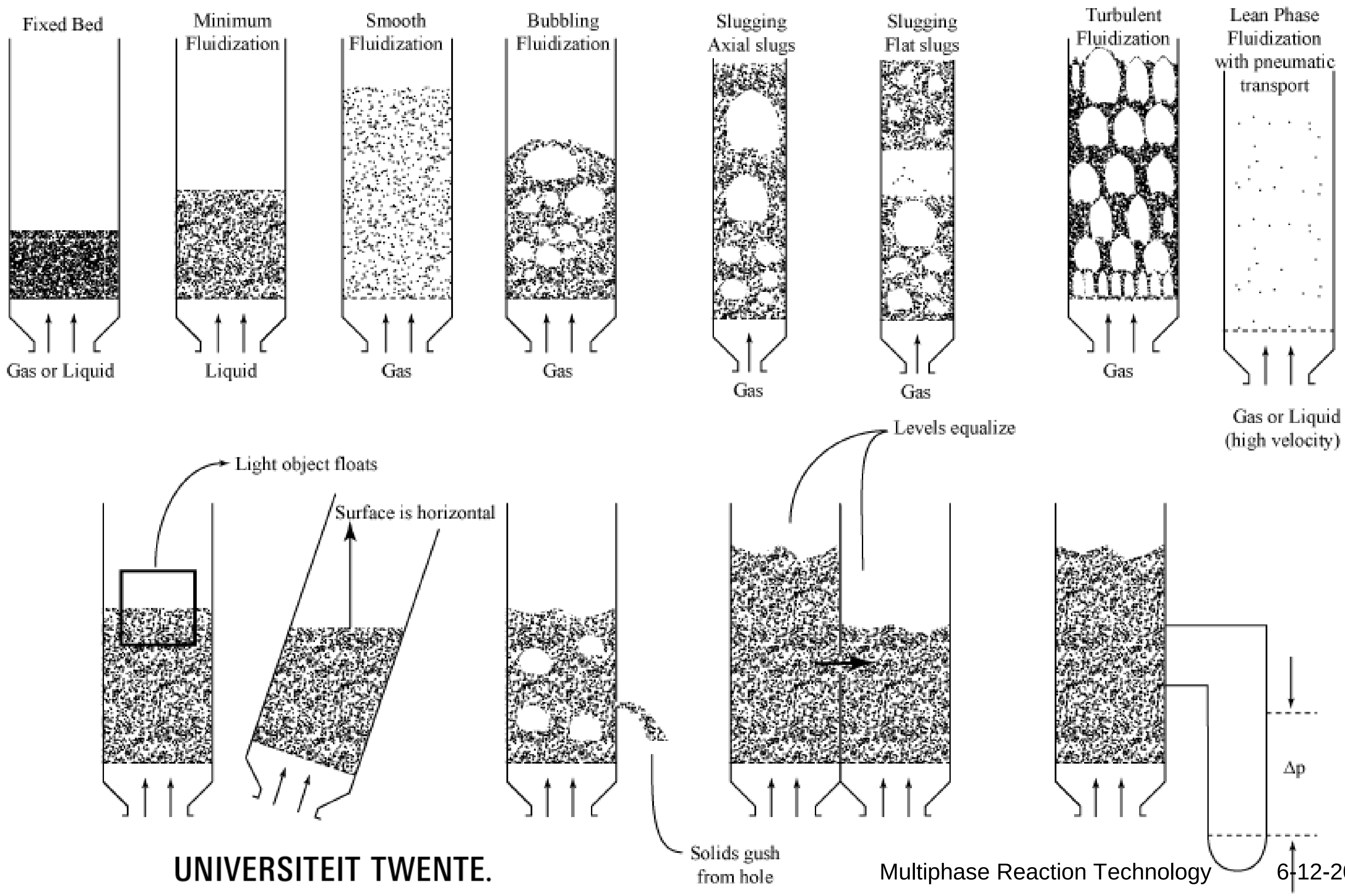
  ✓ Computational fluid dynamics (CFD) based modelling

- Further reading

  ‘Fluidization engineering”, Daizo Kunii and Octave Levenspiel

  Publisher: Butterworth Heineman series in chemical engineering(1991)

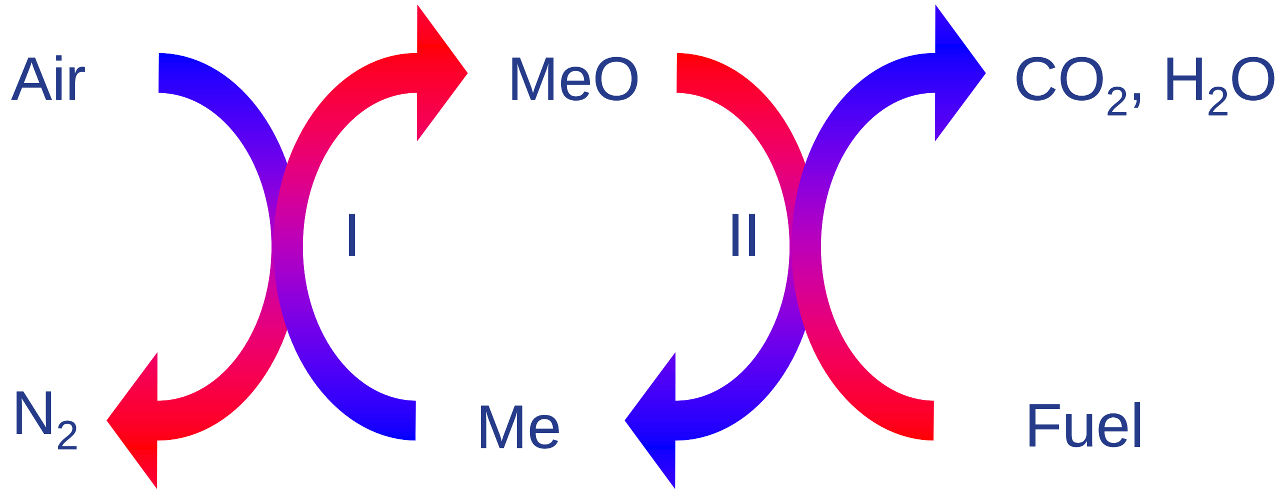
Introduction
Phenomenon of fluidization



Introduction

- Applications of fluidized systems
  - ✓ Heat exchange and drying
  - ✓ Coating and granulation
  - ✓ Gas purification via adsorption
  - ✓ Chemical synthesis(acrylonitrile, maleic and phtalic anhydride)
  - ✓ Polymerization of lower olefines(propylene)
  - ✓ Fischer-Tropsch synthesis
  - ✓ Fluid coking and Flexi-Coking
  - ✓ Combustion and incineration
  - ✓ Fluid Catalytic Cracking (FCC)
  - ✓ Chemical looping combustion (CLC)

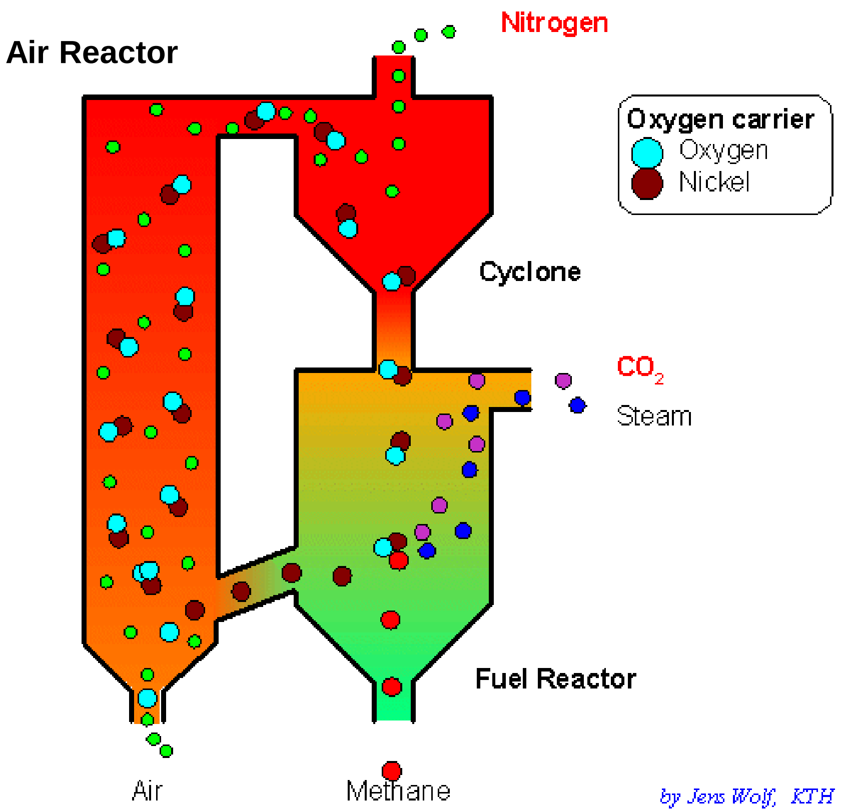
Introduction – CLC



I: $4\ Me + 2\ O_2 \rightarrow 4\ MeO\ (\Delta H < 0)$

II: $4\ MeO + CH_4 \rightarrow 4\ Me + CO_2 + 2\ H_2O\ (\Delta H > 0)$

Introduction – CLC



(T_{ex} = 1200 ^{0}C, p = 25 bar)

Circulating fluidized bed:

+ Continuous power production
+ Proven technology
− Recirculation of particles
− Gas-solid separation

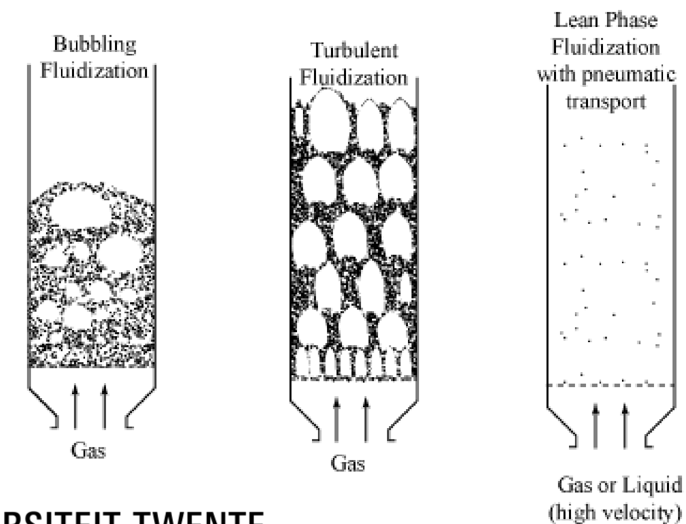
When we select a fluidized bed reactor

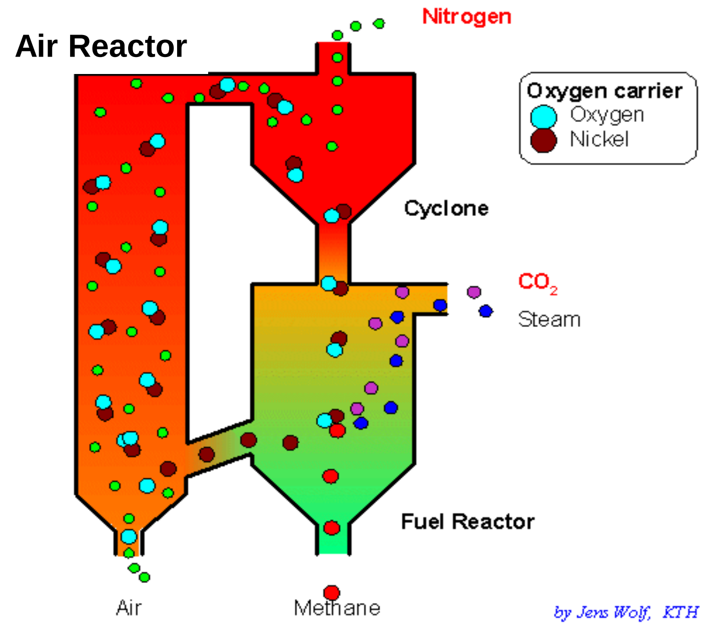
- We can use a fluidized bed basically as:

  ✓ Bubbling fluidized bed

  ✓ Turbulent fluidized bed

  ✓ Pneumatic transport (fast fluidization)





(T_{ex} = 1200 ^{0}C, p = 25 bar)

When we select a fluidized bed reactor
Bubbling and turbulent

Solid-catalyzed gas phase reaction	For small granular or powdery nonfriable catalyst. Can handle rapid deactivation of solids. Excellent temperature control allows large scale operations
Gas solid reaction	Can use wide range of solids with much fines. Large scale operations at uniform temperature possible. Excellent for continuous operations, yielding a uniform product
Temperature distribution in the bed	Temperature is almost constant throughout. This is controlled by heat exchange or by proper continuous feed and removal of solids

When we select a fluidized bed reactor
Bubbling and turbulent

Particles	Wide size distribution and much fines possible. Erosion of vessel and pipelines. Attrition of particles and their entrainment may be serious
Pressure drop	For deep beds pressure drop is high, resulting in large power consumption
Heat exchange and heat transport	Efficient heat exchange and large heat transport by circulating solids so that heat problems are seldom limiting in scale-up
Conversion	For continuous operations mixing of solids and gas bypassing result in poorer performance than other reactor types. For high conversion, staging or other special design is necessary

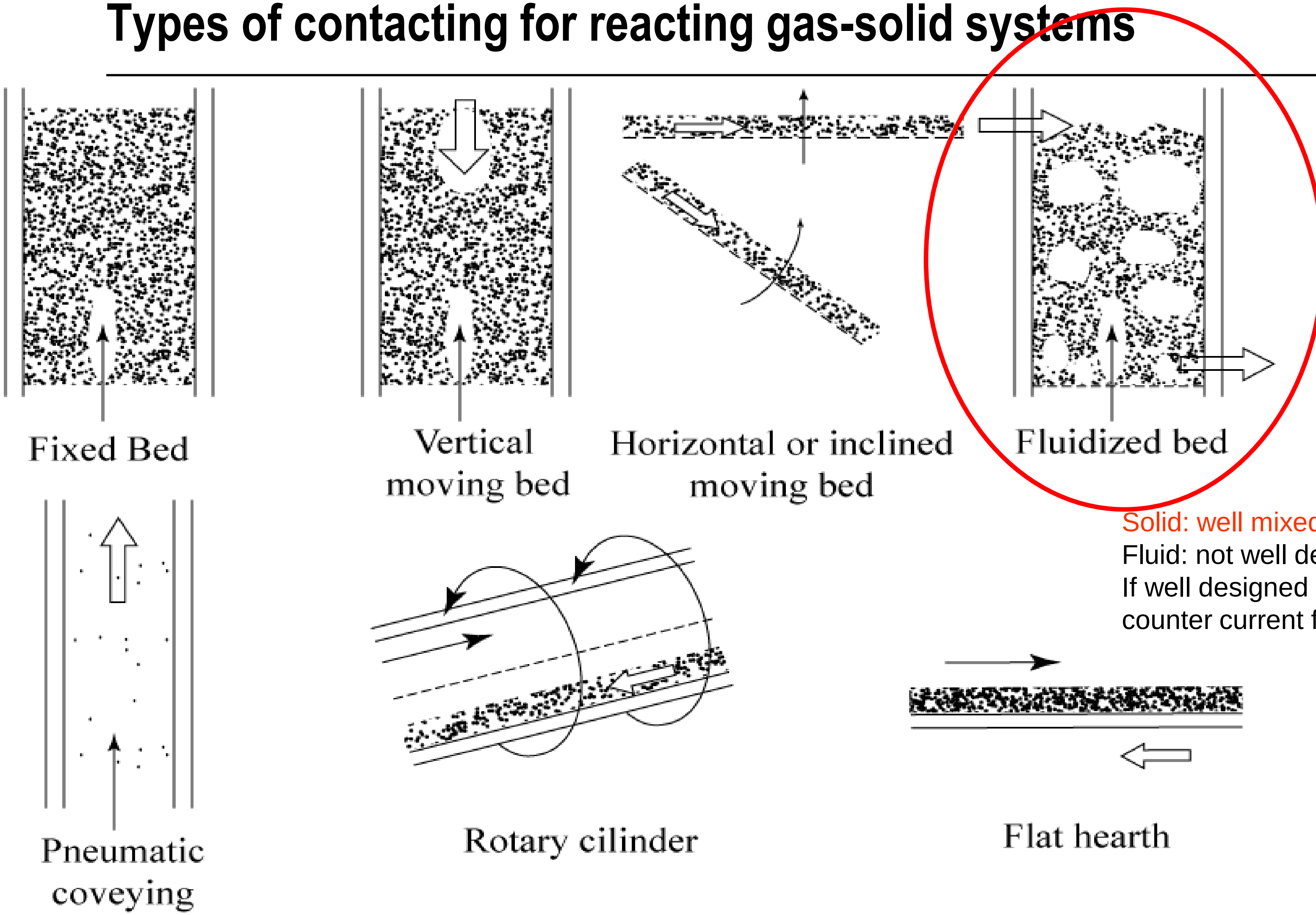
When we select a fluidized bed reactor
Pneumatic transport

Solid-catalyzed gas phase reaction	Suitable for rapid reactions. Attrition of catalyst is serious
Gas solid reaction	Suitable for rapid reactions. Recirculation of fines crucial
Temperature distribution in the bed	Temperature gradients in direction of solids flow can be minimized by sufficient circulation of solid

When we select a fluidized bed reactor
Pneumatic transport

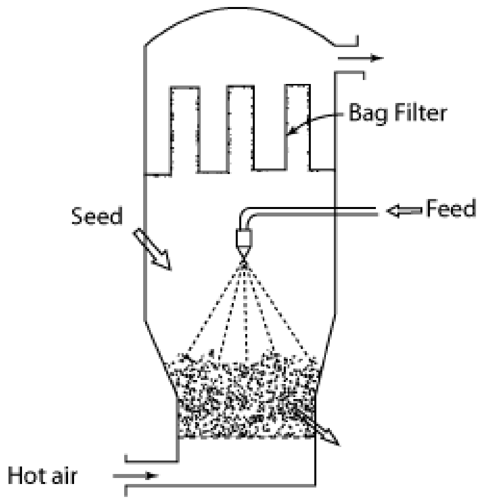
Particles	Fine solids, top size governed by minimum transport velocity. Severe equipment erosion and particle attrition.
Pressure drop	Low for fine particles, but can be considerable for larger particles
Heat exchange and heat transport	Intermediate between fluidized and moving beds
Conversion	Flow of gas and solid both close to cocurrent plug flow, hence high conversion possible

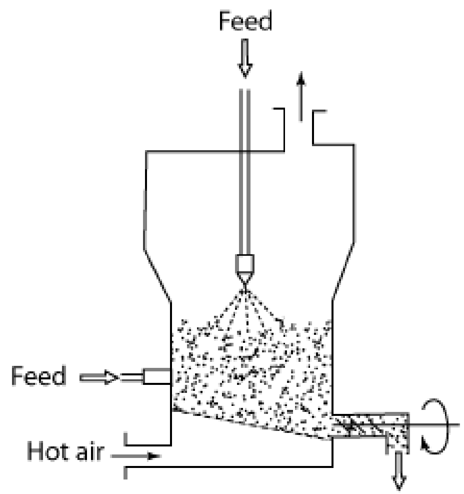
Introduction
Types of contacting for reacting gas-solid systems

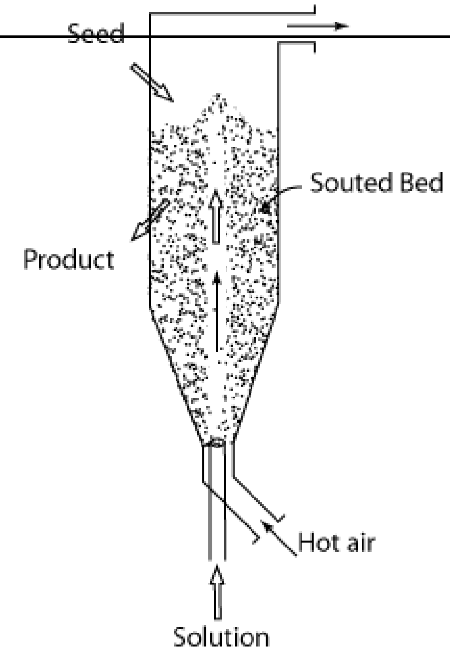


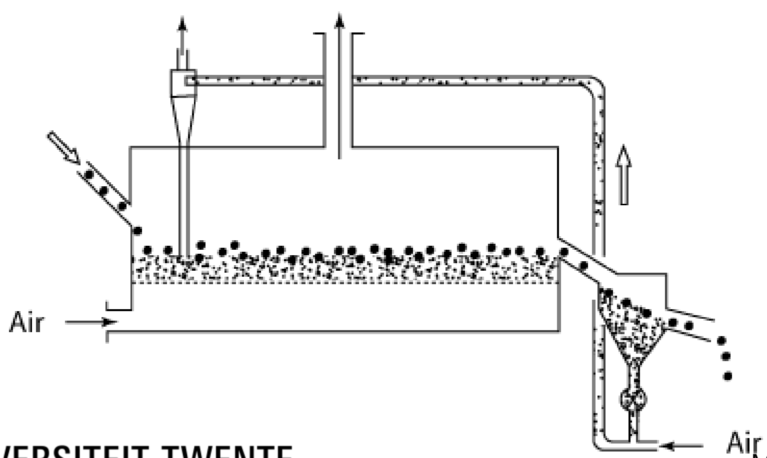
Solid: well mixed flow
Fluid: not well defined
If well designed can be counter current flow

Introduction
Granulation and coating

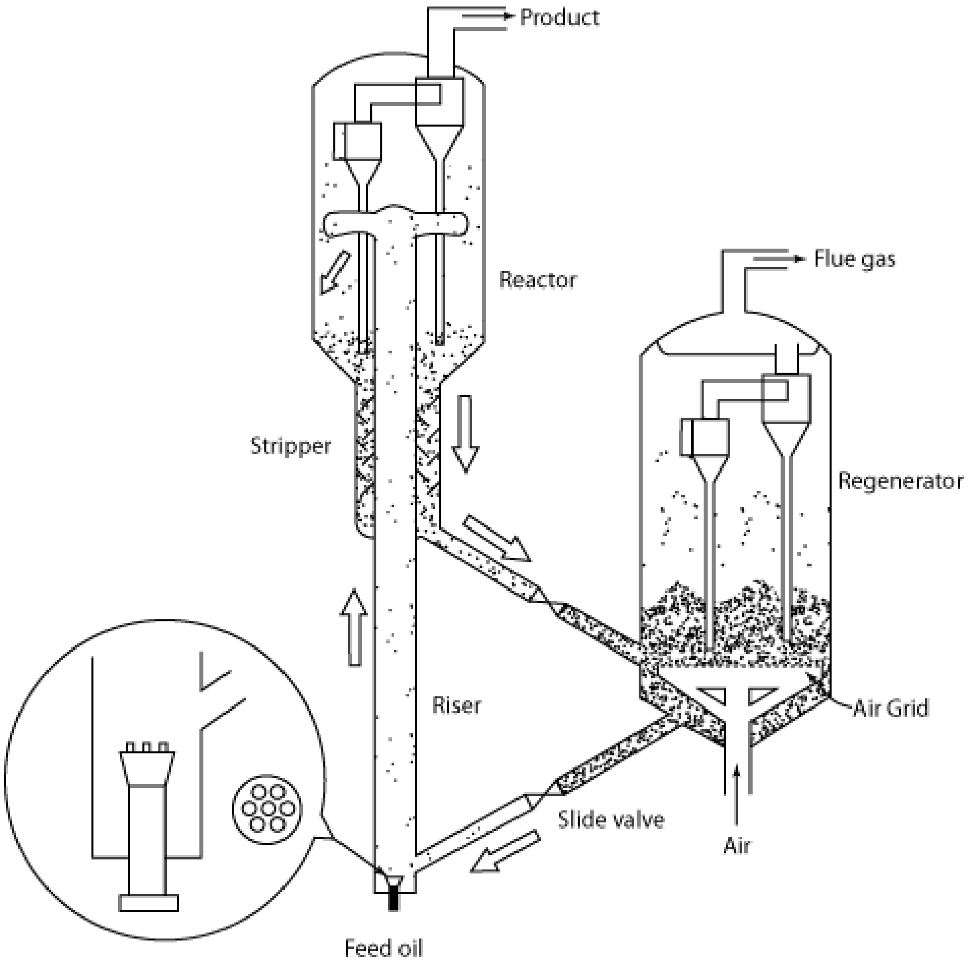








Introduction
Fluid Catalytic Cracking (FCC)



Zeolite catalyst (very active)

Reaction takes place in the riser

High conversion in short contact time

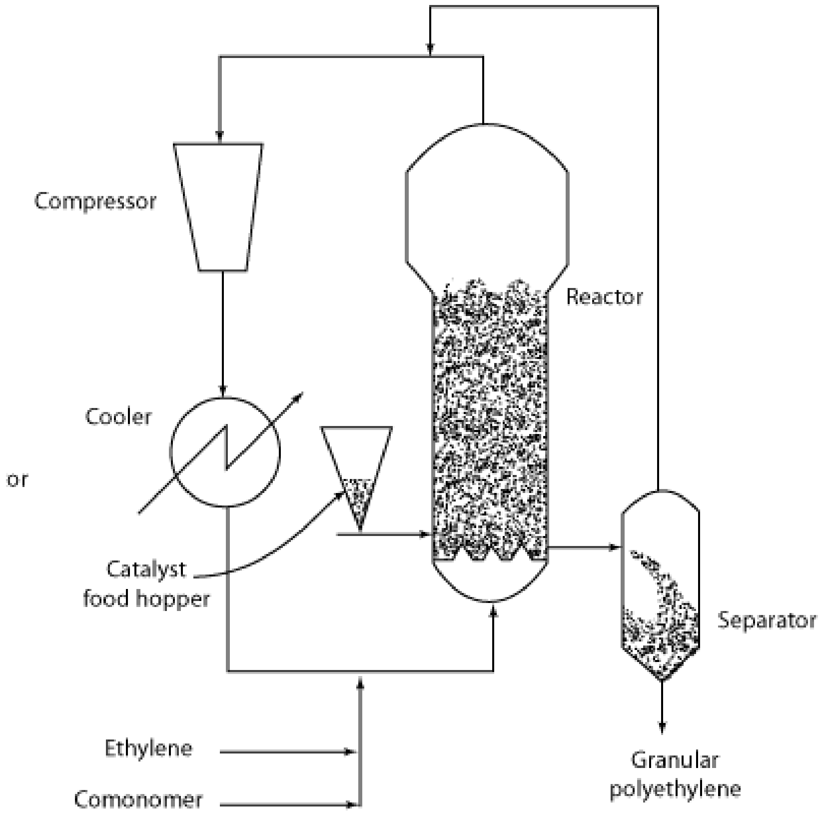
Plug flow ➔ high yield of gasoline
(no overcracking)

Reactor at 470-550°C up to 3.5 bar
Regenerator at 580-700°C up to 4 bar

Reactor 5m ID
Riser 1.5 m ID
Regenerator 8 m ID

Catalyst circulation rate 15-30tons/min

Introduction
Olefine polymerization (Polyethylene)



Feed at 3-6 times minimum fluidization velocity

75-100°C at 20 bar

Particles grow up to 300-1000 μm

Height is 2.5-5 times the diameter

High amount of heat involved 3300 kJ/kg of ethylene converted

Fluidization

- ✓ Characterization of particulate solids (PSD)
- ✓ Geldart's classification
- ✓ Minimum fluidization velocity u_{mf} and terminal velocity u_t
- ✓ Pressure- and temperature dependence of fluidization
- ✓ High velocity fluidization

Fluidization Regimes

Characterisation of particulate solids

- Equivalent spherical diameter d_{sph}

$$\frac{\pi}{6} d_{sph}^3 = V_{particle}$$

- Sphericity ϕ_s (particle and sphere have same volume):

$$\phi_s = \frac{A_{sphere}}{A_{particle}}$$

- Specific surface of particle a'

$$a' = \frac{A_{particle}}{V_{particle}} = \frac{6}{\phi_s d_{sph}}$$

- Specific bed surface a:

$$a = \frac{6\,(1-\varepsilon_b)}{\phi_s d_{sph}}$$

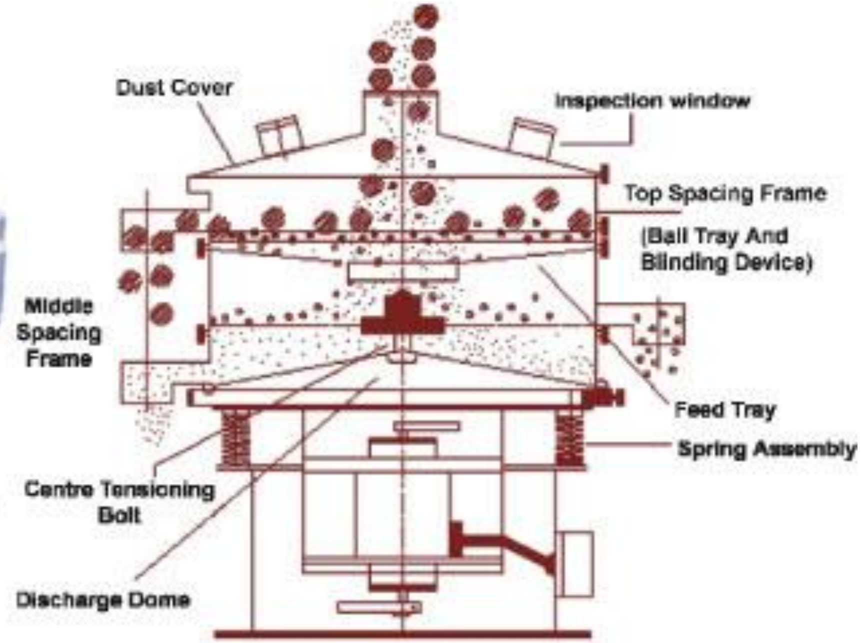
Fluidization Regimes
Characterisation of particulate solids

- Sphericity of particles

Type of particle		Sphericity Φ_S	Source
Sphere		1	(a)
Cube		0,81	(a)
Cylinder			
	h=d	0,87	(a)
	h=5d	0,7	(a)
	h=10d	0,58	(a)
Disks			
	h=d/3	0,76	(a)
	h=d/6	0,6	(a)
	h=d/10	0,47	(a)
Activated carbon and silica gels		0,7-0,9	(b)
Broken solids		0,63	(c)
Coal			
	anthracite	0,63	(e)
	bituminous	0,63	(e)
	natural dust	0,65	(d)
	pulverized	0,73	(d)
Cork		0,69	(d)
Glass, crushed, jagged		0,65	(d)
Magnetite, Fischer-Tropsch catalyst		0,58	(e)
Mica flakes		0,28	(d)
Sand			
	round	0,86	(e)
	sharp	0,66	(e)
	old beach	as high as 0,86	(f)
	young river	as low as 0,53	(f)
Tungsten powder		0,89	(d)
Wheat		0,85	

Fluidization Regimes
Characterisation of particulate solids

Particle diameter from screen analysis



Fluidization Regimes
Characterisation of particulate solids

Particle diameter from screen analysis

Mesh Number[a]	Aperature[b] (in)	Aperature[b] (μm)		Mesh Number[a]	Aperature[b] (in)	Aperature[b] (μm)
3	0,263	6680		35	0,0165	417
4	0,185	4699		48	0,0116	295
6	0,131	3327		65	0,0082	208
8	0,093	2362		100	0,0058	147
10	0,065	1651		150	0,0041	104
14	0,046	1168		200	0,0029	74
20	0,0328	833		270	0,0021	53
28	0,0232	589		400	0,0015	38

[a]Number of wires per inch

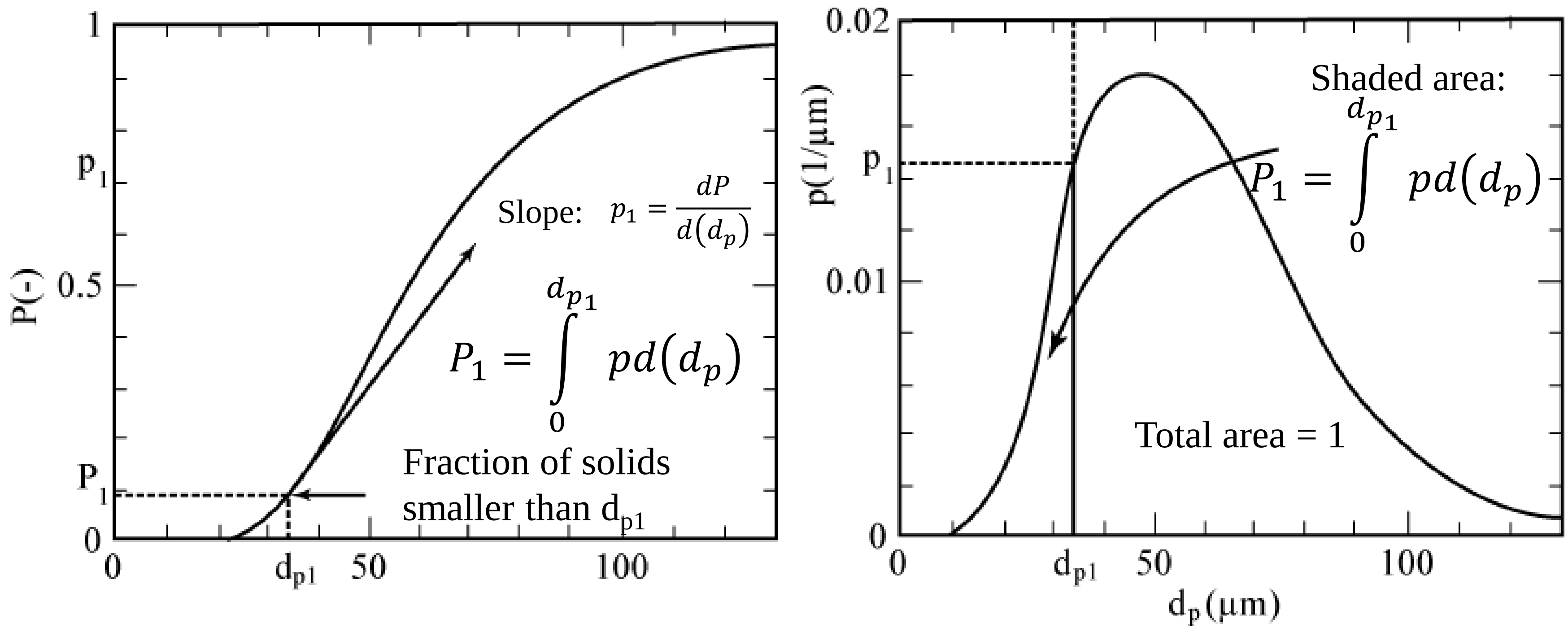
[b]Opening between adjacent wires

-150 +200 mesh particles have by definition the following screen size

$$d_p = \frac{104 + 74}{2} = 89\ (\mu m)$$

Fluidization Regimes
Characterisation of particulate solids

Differential and integral particle size distribution functions p and P



Fluidization Regimes
Characterisation of particulate solids

- Mean specific particle surface for continuous PSD:

$$\bar{a}' = \frac{6}{\phi_s \bar{d}_p} = \int_0^{d_{p,\max}} a' p d(d_p) = \int_0^{d_{p,\max}} \frac{6}{\phi_s d_p} p d(d_p)$$

continuous PSD

- Mean specific particle surface for discrete PSD:

$$\bar{a}' = \frac{6}{\phi_s \bar{d}_p} = \sum_i \frac{6}{\phi_s} \left(\frac{x}{d_p}\right)_i$$

discrete PSD

- Mean diameter of mixture based on particle surface (relevant for frictional pressure drop in fixed and fluidized particles):

$$\bar{d}_p = \frac{1}{\int_0^{d_{p,\max}} \frac{p}{d_p} d(d_p)} \qquad \bar{d}_p = \frac{1}{\sum_i \left(\frac{x}{d_p}\right)_i}$$

Fluidization Regimes
Characterisation of particulate solids

- Example for a practical case (polydisperse particles)

Calculate the mean diameter d_p of material of the following size distibution:

Cumulative weight of a representative 300-g sample...		… having a diameter smaller than d_p (µm)
0		50
35		80
86		110
135		140
210		170
300		200

Exercise

Solution:

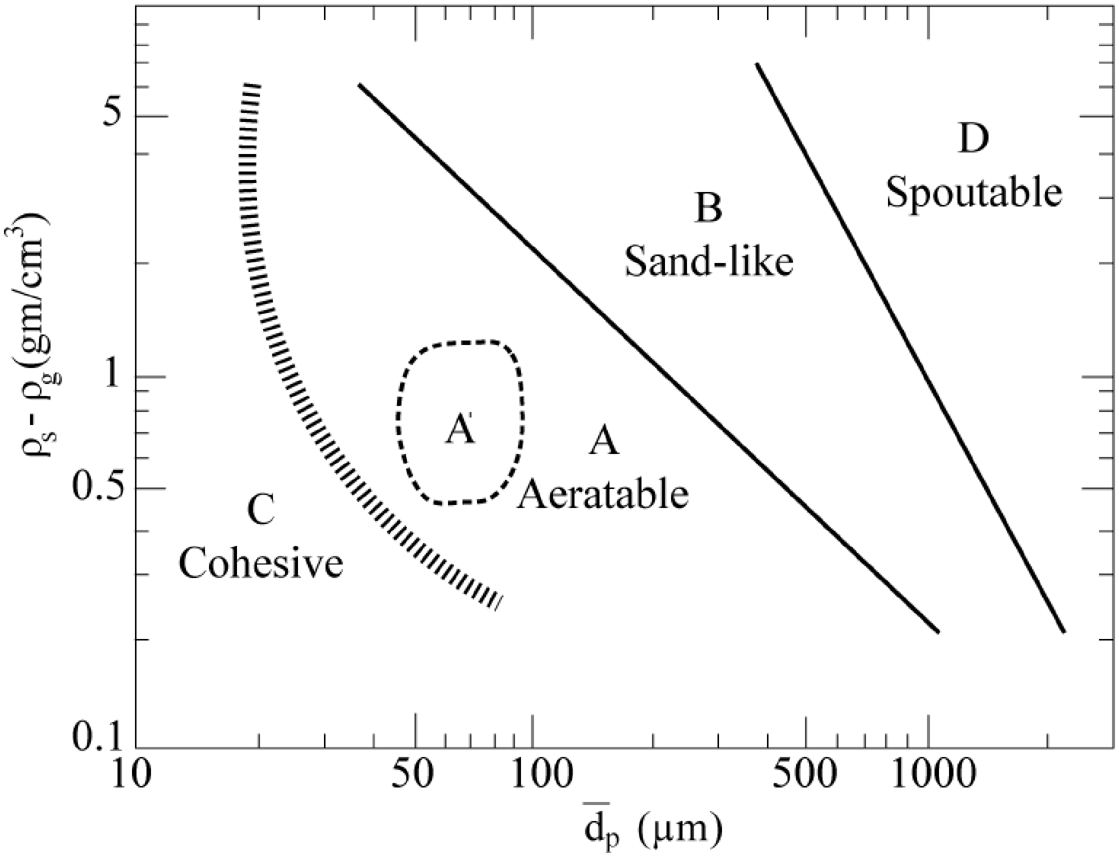
Diameter range (μm)	d_{pi}	Weight fraction in interval $(\mathbf{p}\Delta d_p)_i = x_i$	$(x/d_p)_i$
50-80	65	(35-0)/300 =0,117	0,117/65=0,0018
80-110	95	(86-35)/300 =0,170	0,170/95=0,0018
110-140	125	0.163	0.0013
140-170	155	0.250	0.0016
170-200	185	0.300	0.0016
			$\Sigma(x/d_p)_i$=0,080

The mean diameter is:

$$\overline{d_p} = \frac{1}{\sum_{all\ i}\left(\frac{x}{d_p}\right)_i} = \frac{1}{0.0080} = 124.5\ \mu m$$

Fluidization Regimes
Geldart's classification

- Four types of fluidization behaviour (A, B, C and D)



Fluidization Regimes
Geldart's classification

- Group A: Aeratable, or particles with a small size and/or density (<1.4 g/cm^3). Smooth fluidization at low u and controlled bubbling (small bubbles). FCC is a typical example

- Group B: Sandlike, most particles of size 40 μm$<d_p<$500 μm and density $1.4<\rho_s<4$ g/cm^3. These particles fluidize well with vigorous bubbling action and bubbles that grow large (coalescence)

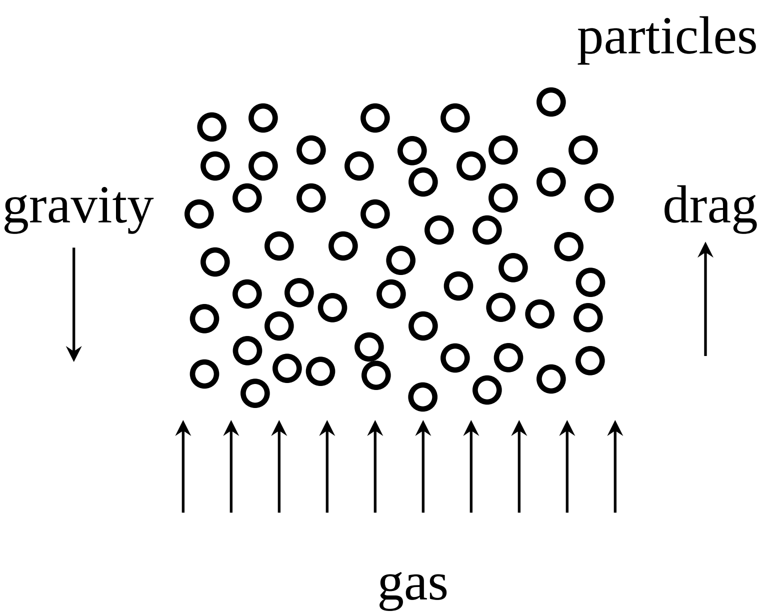
- Group C: Cohesive, or very fine powders. Normal fluidization is extremely difficult due to strong interparticle forces which exceed the gas drag. Examples: flour and starch

- Group D: Spoutable, or large and/or dense particles. Deep beds of these particles are difficult to fluidize due to "explosive" bubble growth (coalesence). Examples: peas and coffee beans

note: Geldart's classification effectively includes only ρ_s and d_p !!!

Fluidization Regimes
Minimum fluidization velocity u_{mf} and terminal velocity u_t

- Prevailing regime depends on (effective) forces acting on particles: drag in (dense) swarm differs from drag for isolated particles in an unbounded fluid



$F_d > F_g$: entrainment

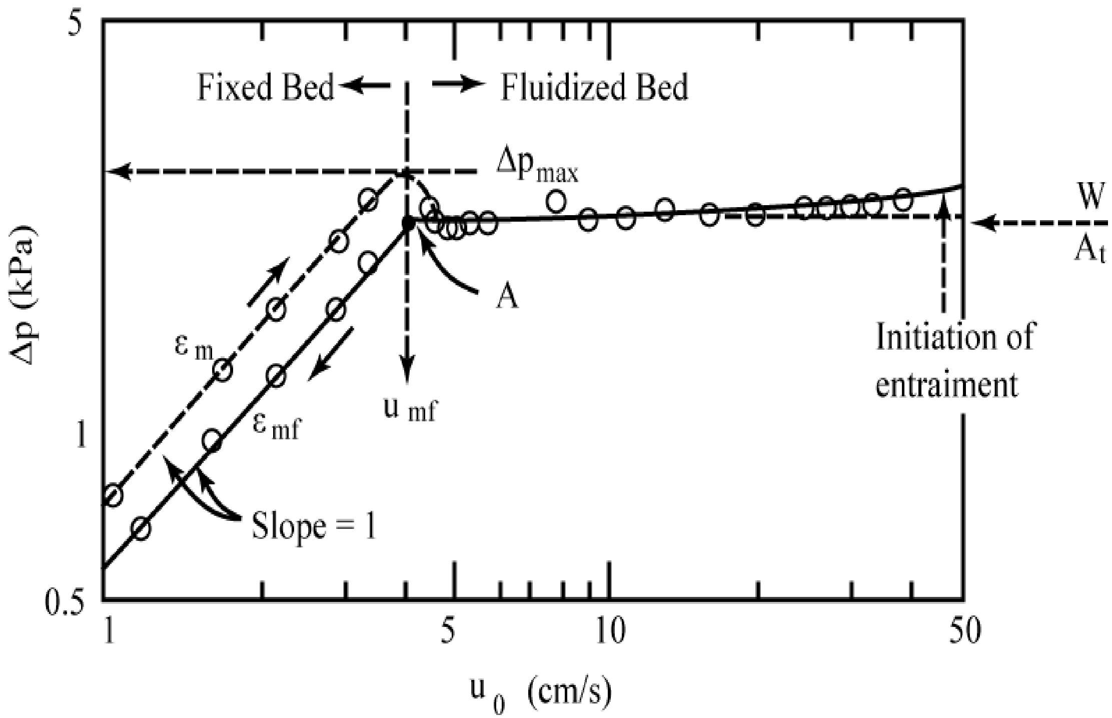
$F_d = F_g$: fluidization

$F_d < F_g$: defluidization

Richardson-Zaki drag correction: $C_{d,swarm} = C_{d,isolated} f(\varepsilon) = C_{d,isolated} \varepsilon^{-2.65}$

Fluidization Regimes
Minimum fluidization velocity u_{mf}

- Pressure drop Δp versus fluidization velocity u_0



Uniform size distribution

- Incipient or minimum fluidization: weight of particles is equal to the drag force

Minimum fluidization velocity (u_{mf})

drag force due to gas flow = weight of particles

Or

(pressure drop) (cross sectional area) = (volume of bed) (fraction of solid) (specific weight of solid)

$$\Delta p_b A_t = A_t L_{mf} (1 - \varepsilon_{mf}) [(\rho_s - \rho_g) g]$$

Fluidization Regimes
Minimum fluidization velocity u_{mf} and terminal velocity u_t

- Expression for u_{mf}:

$$\frac{1.75}{\varepsilon_{mf}^3 \phi_s}\left(\frac{d_p u_{mf} \rho_g}{\mu}\right)^2 + 150\frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3 \phi_s^2}\left(\frac{d_p u_{mf} \rho_g}{\mu}\right) = \frac{d_p^3 \rho_g (\rho_s - \rho_g) g}{\mu^2}$$

- Dimensionless form:

Archimedes number

$$\frac{1.75}{\varepsilon_{mf}^3 \phi_s}\left(Re_{p,mf}\right)^2 + 150\frac{(1-\varepsilon_{mf})}{\varepsilon_{mf}^3 \phi_s^2}\left(Re_{p,mf}\right) = Ar$$

Particle Reynolds number

Fluidization Regimes

Minimum fluidization velocity u_{mf} and terminal velocity u_t

- Voidage ε_{mf} at incipient fluidization conditions:

			Size, d_p (mm)						
Particles			0,02	0,05	0,07	0,1	0,2	0,3	0,4
Sharp sand, Φ_s=0,67			-	0,6	0,59	0,58	0,54	0,5	0,49
Round sand, Φ_s=0,86			-	0,56	0,52	0,48	0,44	0,42	-
Mixed round sand			-	-	0,42	0,42	0,41	-	-
Coal and glass powder			0,72	0,67	0,64	0,62	0,57	0,56	-
Anthracite coal, Φ_s=0,63			-	0,62	0,61	0,6	0,56	0,53	0,51
Absorption carbon			0,74	0,72	0,71	0,69	-	-	-
Fischer-Tropsch catalyst, Φ_s=0,58			-	-	-	0,58	0,56	0,55	-
Carborundum			-	0,61	0,59	0,56	0,48	-	-

Fluidization Regimes
Pressure and temperature dependence of fluidization

- Summary of experimental findings for pressure dependence

  - ε_{mf} increases slightly (1-4%) with rise in operating pressure

  - u_{mf} decreases with rise in operating pressure. This decrease is negligible for fine particles (d_p<100 μm) but is significant for larger particles (d_p>360 μm). This effect is consistent with predictions from the Ergun equation.

  - u_{mb}/u_{mf} increases up to 30% for coarse alumina (d_p= 450 μm). Thus an increase in operating pressure widens range of particulate or homogeneous fluidization in gas-solid systems.

⇨ "smoother fluidization at elevated pressure"

Fluidization Regimes
Pressure and temperature dependence of fluidization

- Summary of experimental findings for temperature dependence

  - ε_{mf} increases with temperature for fine particles

(up to 8% for temperatures up to 500 °C)

  - ε_{mf} unaffected by temperature for coarse particles (B and D type of particles)

  - u_{mf} can be predicted reasonably well by Ergun equation provided that correct values for incipient fluidization porosity e_{mf} and physical properties (density r and viscosity m) are used

Exercise

- Calculate the minimum fluidization velocity for the following system

Bed: $\varepsilon_{mf} = 0.55$

Solid: sharp sand dp = 160 μm, $\phi_s = 0.67$ $\rho_s = 2.6 g/cm^3$

Fluid phase: ambient air $\mu = 0.00018$ g/(cm s)

Exercise solution

Fluid phase: ambient air (T = 25 °C and 1 atm)

Mw = 0.21*32+0.79*28 = 28.8 g/mol

PV = nRT ➔ PV = gRT/Mw ➔ ρ_g = PMw/(RT) = 1*28.8/(0.0821*298.15) = 1.2 g/l = 0.0012 g/cm^3

$$\frac{1.75}{\varepsilon_{mf}^3 \phi_s}\left(Re_{p,mf}\right)^2 + 150\frac{\left(1-\varepsilon_{mf}\right)}{\varepsilon_{mf}^3 \phi_s^2}\left(Re_{p,mf}\right) = Ar$$

$$15.7\left(Re_{p,mf}\right)^2 + 903.79\left(Re_{p,mf}\right) = 416.52$$

$$Re_{p,mf} = \frac{d_p u_{mf} \rho_g}{\mu} = 0.457 \qquad u_{mf} = 4.29\ cm/s$$

Fluidization Regimes
Terminal velocity u_t

- Terminal velocity of particle of size dp falling through a fluid :

$$u_t = \left[\frac{4d_p(\rho_s - \rho_g)g}{3\rho_g C_d}\right]^{\frac{1}{2}}$$

- Drag coefficient C_d given by: Powder Technology 58 (1989) 63

$$C_d = \frac{24}{Re_p}\left[1 + \left(8.1716e^{-4.0655\phi_s}\right)Re_p^{0.0964+0.5565\phi_s}\right] + \frac{73.69Re_p e^{-5.0748\phi_s}}{Re_p + 5.378e^{6.2122\phi_s}}$$

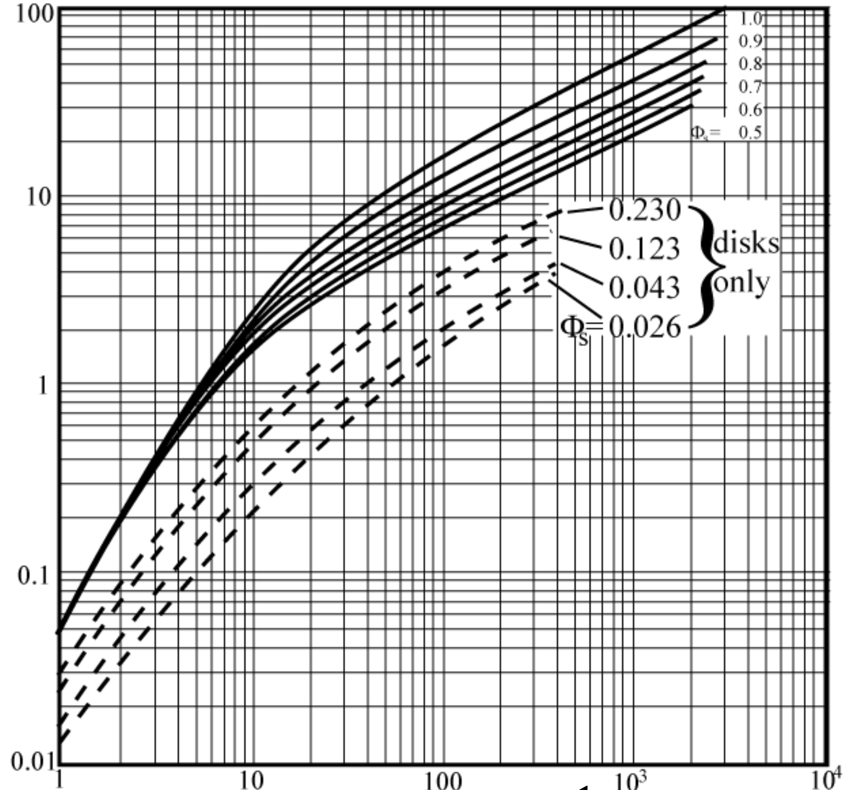
- Drag coefficient for spherical particles (ϕ_s=1):

$$C_d = \frac{24}{Re_p} + 3.3643Re_p^{0.3471} + \frac{0.4607Re_p}{Re_p + 2682.5}$$

Fluidization Regimes
Terminal velocity u_t

- Graphical determination of terminal velocity

$$u_t^* = u_t \left[\frac{\rho_g^2}{\mu(\rho_s - \rho_g)g}\right]^{\frac{1}{3}}$$

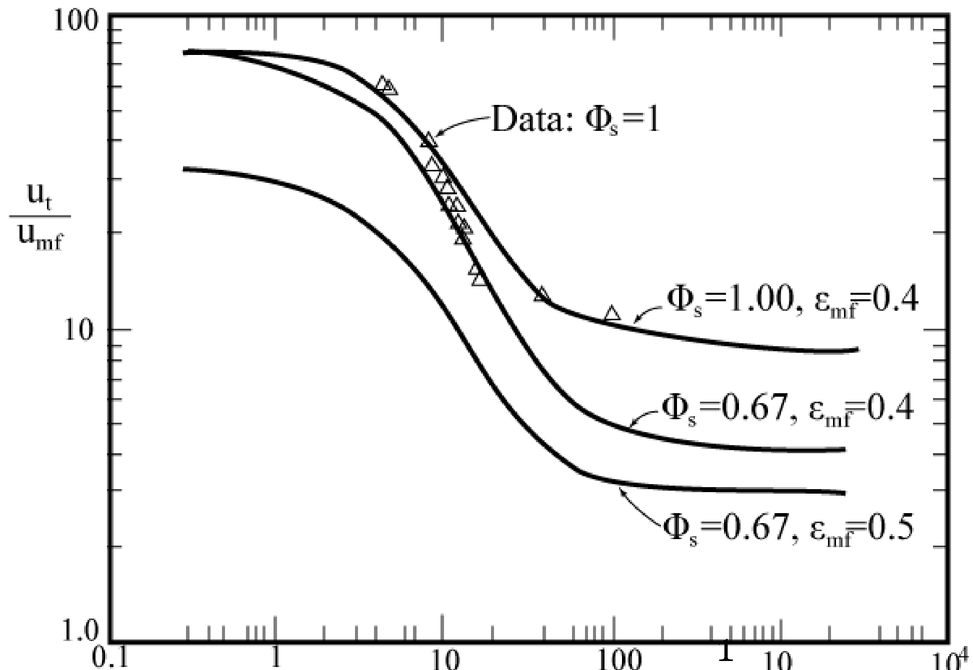


$$d_p^* = d_p \left[\frac{\rho_g(\rho_s - \rho_g)g}{\mu^2}\right]^{\frac{1}{3}}$$

To avoid carryover of particles the velocity should be between u_{mf} and u_t

Fluidization Regimes

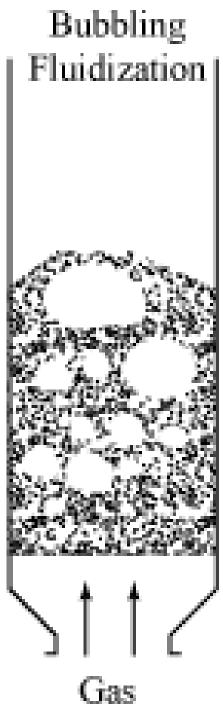
Ratio of terminal velocity u_t and minimum fluidization velocity u_{mf}



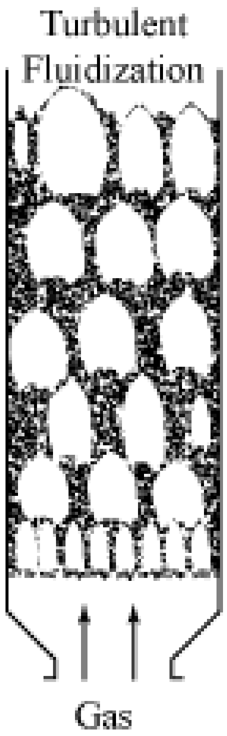
$$d_{sph}^{*} = d_{sph}\left[\frac{\rho_g\left(\rho_s - \rho_g\right)g}{\mu^2}\right]^{\frac{1}{3}}$$

- Useful velocity range for large particles is much smaller than that for small particles (reflected in applications)

Increase of fluidization velocity



Increasing of fluidization velocity

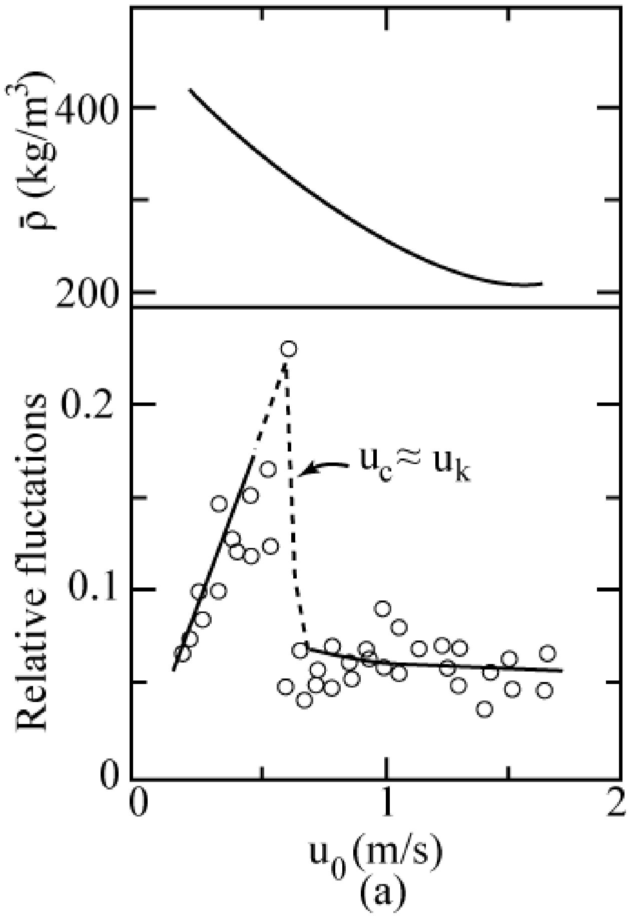


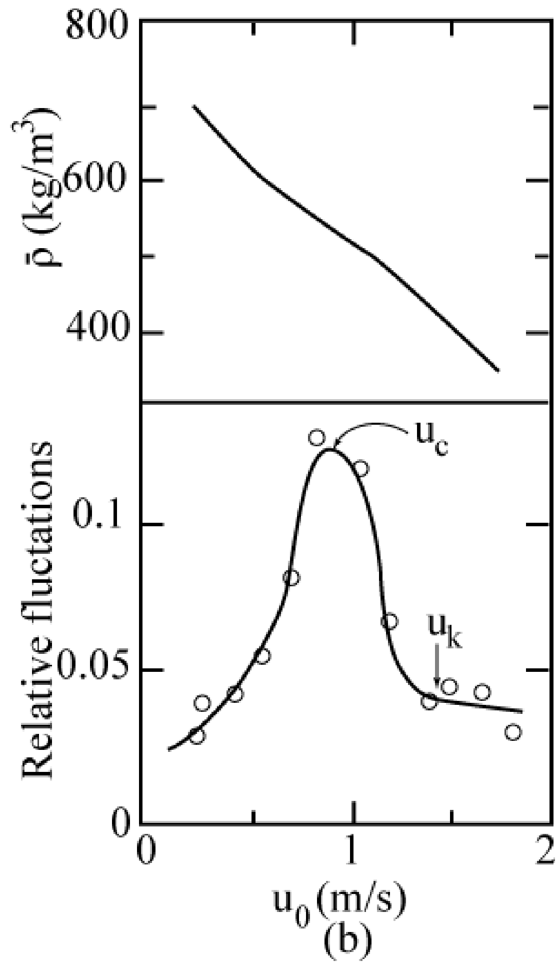
An increase of pressure fluctuations is observed

Pressure fluctuations level off when the transition is obtained

Fluidization Regimes
High velocity fluidization

- Turbulent fluidization





Pressure fluctuation and mean bed density in 15.2 cm bed for two solids d_p=49 μm;

(a) FCC catalyst: ρ_s=1070 kg/m^3, u_t=7.78 cm/s, u_c=61 cm/s, u_k=61 cm/s;

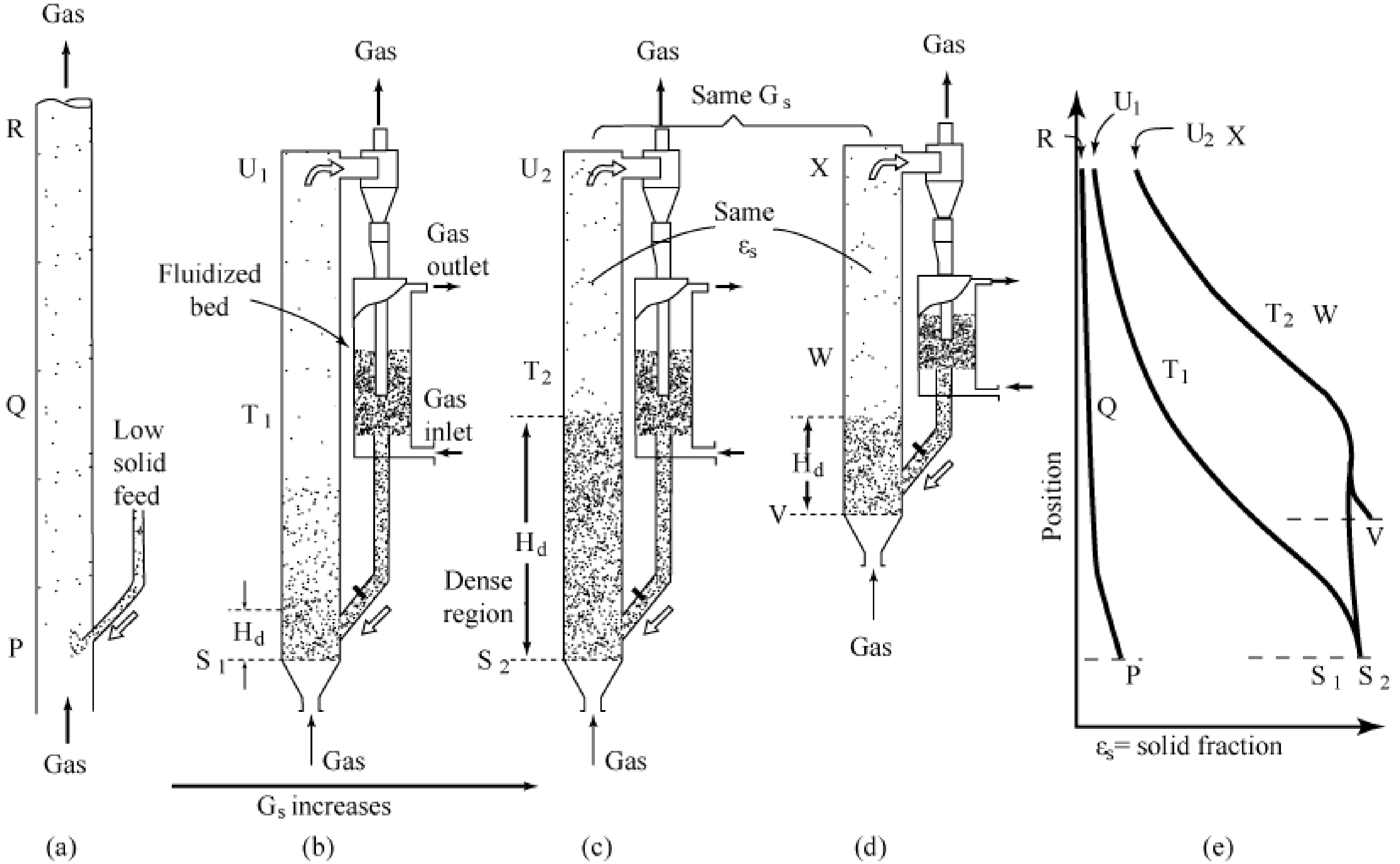
(b) silica alumina catalyst: ρ_s=1450 kg/m^3, u_t=10.6 cm/s, u_c=91 cm/s, u_k=137 cm/s.

Fluidization Regimes
High velocity fluidization

- Pneumatic conveying [a] and fast fluidization [b], [c] and [d]

$u=20u_t$

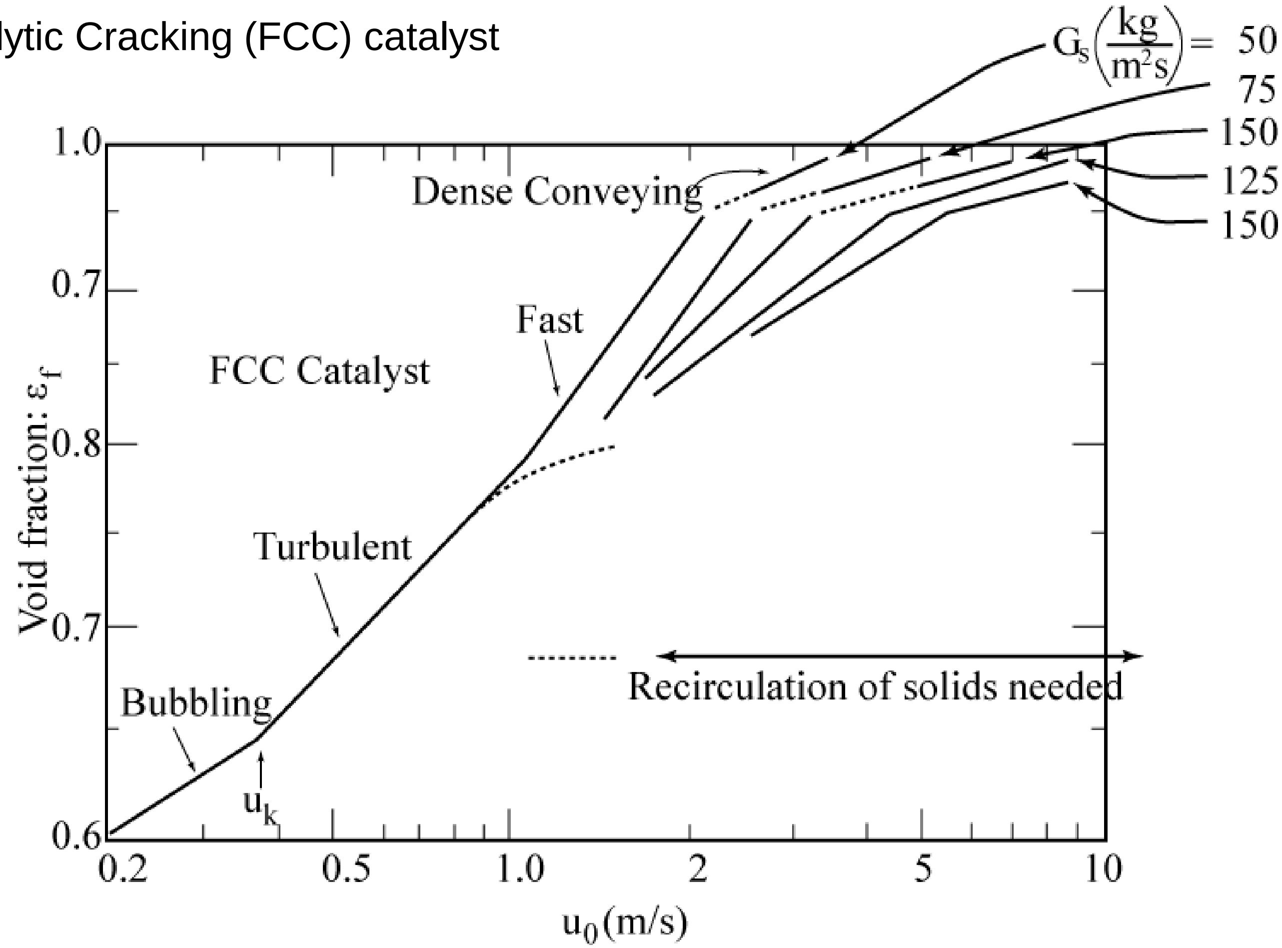
If decrease u and increase feed rate of solid a *choking condition* is reached



Fluidization Regimes

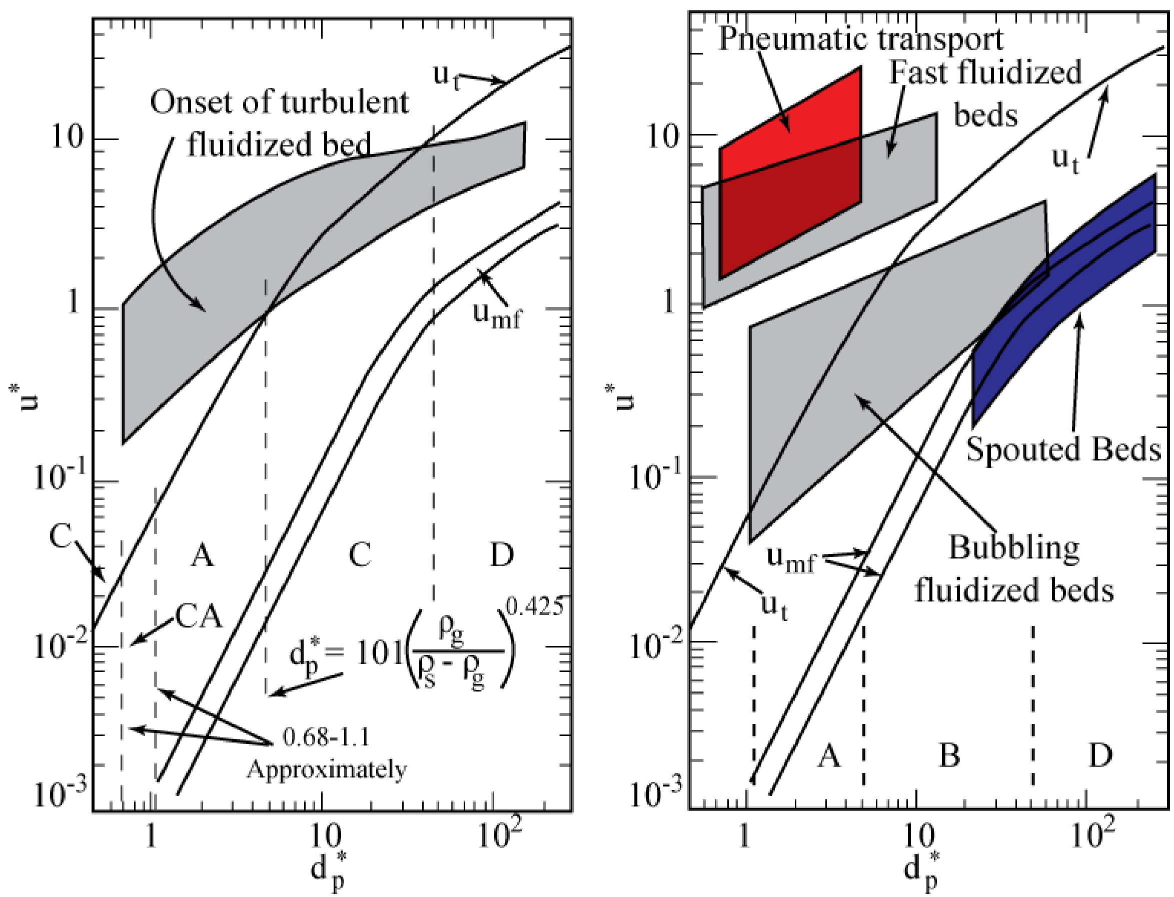
High velocity fluidization

- Void fraction ε_f and fluidization regimes as a function of the superficial gas velocity u_0 for Fluid Catalytic Cracking (FCC) catalyst



Dense Fluidized Beds
High velocity fluidization

- General flow regime for gas-solid contacting



dimensionless particle diameter

$$d_p^* = d_p\left[\frac{\rho_g(\rho_s - \rho_g)g}{\mu^2}\right]^{\frac{1}{3}}$$

dimensionless superficial velocity

$$u^* = u\left[\frac{\rho_g^2}{\mu(\rho_s - \rho_g)g}\right]^{\frac{1}{3}}$$

Resume

- What is a fluidized bed
- Which kind of fluidization regimes we can have (and we can use)
- Which kind of particles we have (and we can use)
- Characterization of particles
- Minimum fluidization velocity
- Terminal velocity