Scaling and particle size optimization of mass transfer in gas fluidized beds

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SCALING AND PARTICLE SIZE OPTIMIZATION OF MASS TRANSFER IN GAS FLUIDIZED BEDS.

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ABSTRACT.

Conversion was measured in a gas fluidized bed, using the decomposition of ozone as a model reaction. The results were expressed in terms of the height of a mass transfer unit.

Mass transfer data from our own experiments and a large number of literature data were analyzed. We found a Scaling parameter S , that could describe all experiments. This Scaling parameter can be helpful in scale up. For A type powders, the height of a mass transfer unit was shown to rise with increasing particle size, using S as reference.

KEYWORDS

Scale up; Fluidization; Particle size; Mass transfer; Chemical Reaction; Residence time distribution; Scaling parameter.

1. INTRODUCTION

Scale up is essentially the target for all investigations concerning gas fluidized beds: a method has to be found for predicting the conversion of a gas fluidized bed, under given circumstances. A very important factor that determines the conversion, is the mass transfer from the bubble phase to the dense phase.

The influence of the average particle size d_p on the mass transfer is an important, but still not sufficiently investigated factor. Based on the theory of Sit and Grace (1978), we estimated the height of a mass transfer unit H_k with increasing particle size and constant bed height, bed diameter and U/U_{mf} -ratio. H_k was found to increase for A and small B powders (according to the classification of Geldart (1973)) and found to decrease for the large B and D type powders, with a maximum in between. This was confirmed by measurements of Borodulya et al. (1981). They performed experiments with a single bubble bed. However, hydrodynamics, and therefore mass transfer, are completely different in a freely bubbling bed.

Particle size influences a lot of other parameters. This means that it is impossible to simply determine the influence of d_p on the height of a mass unit H_k , by measuring H_k with several particle sizes. It is necessary to find a factor that holds for all fluidized beds. Then it also should be possible to use this parameter as a tool in scale up. This work describes experimental work and analyses, concerning a lot of literature data, to obtain this parameter. For A/B powders the influence of d_p on H_k was determined.

2. MODEL DESCRIPTION

The model of Van Deemter (1961) was used to analyze the data. The fluidized bed is divided into two phases: the bubble phase in ideal plug flow and the dense phase in plug flow with axial dispersion.

Based on calculations and work of Van Swaaij and Zuiderweg (1972) we neglected the dispersion term for the dense phase. For A and B powders most of the gas enters the bed in the bubble phase, because $U \gg U_{mf}$. Then it can easily be derived that the following holds (Van Swaaij and Zuiderweg, 1972):

$$\frac{C_e}{C_f} = \exp(-N_t) \quad , \quad \text{with} \quad \frac{1}{N_t} = \frac{1}{N_k} + \frac{1}{N_r} \quad (1)$$

From equation (1), known number of reaction units N_r and measured conversion, the total number of units N_t and the number of mass transfer units N_k can be calculated.

3. EXPERIMENTAL

For our experiments the decomposition of ozone was chosen as a model reaction, because it is a first order reaction and process control is relatively simple.

The catalyst was quartz sand, coated with iron oxide. The properties of the catalyst are listed in table 1.

Table 1. Properties of catalyst used with experiments.

mean sieve particle size	d_{32}	67 μm
particle density	ρ_p	2590 kg/m^3
minimum fluidization velocity	U_{mf}	0.6 cm/s
bed porosity	ϵ_{mf}	0.55
wt % Fe		0.33

Reaction rates were determined in a fixed bed reactor. The reaction was found to be first order with an activation energy of about 109 kJ/mole. This is of the same order as was found in previous investigations. Conversion experiments were performed in a fluid bed reactor with varying bed temperature and superficial velocity. Bed diameter was 10 cm and bed height was about 40 cm. A porous plate distributor was used. More detailed information can be found elsewhere (van Lare, 1990).

At low k_r values the system is reaction rate controlled (I), which means that $N_t \approx N_r$. This implies that large N_k and small H_k values will be found. In this region it is also difficult to obtain accurate H_k values, due to experimental error. Once the mass transfer controlled region is reached (II), H_k will become constant, until accelerated mass transfer occurs (III). This is completely comparable with fixed beds. Indeed we found this trend (fig 1), but we did not reach region (III).

fig. 1 H_k values as a function of k_r (67 μm catalyst, experimental)

From literature it is known that region III indeed occurs. Figure 1 also shows that the influence of superficial velocity is rather small for these particles, probably due to the small U_{mf} value. A H_k value of about 18 cm was found.

4. DATA ANALYSIS

The process parameters, that were taken into account, were divided into three main groups:

- a) particle properties: d_p , ρ_p , U_{mf} , ϵ_{mf} .
- b) gas properties: μ_g , D_g , ρ_g .
- c) reactor dimensions and external influences: U , H , D , H_k , N_r , internals, distributor etc.

Data taken from the papers listed in table 2, were used for the analysis.

The values for H_k were calculated from the given data and from the figures, presented in the papers. If not given, μ_g , D_g and ρ_g were estimated from Perry's Handbook (1973). Occasionally some particle properties had to be estimated also. More detailed information can be found elsewhere (van Lare, 1990).

Table 2. List of papers used for data analysis.

Author	reaction	D [m]	H [m]	d_p [μ m]	particle type	N_r
-Shen/Johnstone (1955)	decomposition NOx	0.114	0.26-0.32	103	alumina	0.08-0.35
-Lewis et al. (1959)	dehydrogen. ethylene	0.052	0.25	122	microspherical	6-48
-Massimilla/Johnstone (1961)	ox. ammonia	0.114	0.19-0.60	105	alumina	0.19-2.4
-Orcutt et al. (1962)	decomposition ozone	0.102, 0.153	0.295-0.71	39	alumina/ Fe_2O_3	0.2-60
-Kobayashi/Arai (1965) *	decomposition ozone	0.20	0.34	194	silica	0.89-7.5
-Calderbank et al. (1967)	decomp. ozone	0.15, 0.457	0.5 and 1.0	192, 100 and 83	alumina/ Fe_2O_3	0.2-20
-de Groot (1967)	RTD (hydrogen-tracer)	0.1-1.5	1.0-4.9	96	silica	0
-van Swaaij/Zuiderweg (1972)	decomp. ozone	0.1-0.3	1.0-1.22	147	quartz/ Fe_2O_3	1-133
	RTD (He-tracer)	0.1-0.6	0.9-2.35	147 - ?	quartz	0
-Fryer/Potter (1976)	decomposition ozone	0.229	0.115-0.66	117	quartz/ Fe_2O_3	0.2-110
-Bauer (1980)	decomposition ozone	0.2, 1.0	0.6	108	quartz/ Fe_2O_3	0.1-2.6
-Boonstra (1983)	RTD (methane-tracer)	0.457	0.41, 0.82	69	spent crack-cat.	0

* Data taken from Kunii and Levenspiel (1969)

The chemical reaction experiments were all in steady state. H_k was calculated according to equation (1). The H_k values from the non steady state (residence time distribution (RTD) experiments) were taken directly from the papers.

Statistical analyses did not give a practical relationship for H_k as a function of all the parameters, but some trends did show. Based on these statistical results we defined a Scaling parameter S (with dimensions of length):

$$S = \sqrt{H \cdot D \cdot (U/U_{mf}) / \sqrt{N_r}} \quad \text{for } N_r \geq 1 \quad (2a)$$

$$S = \sqrt{H \cdot D \cdot (U/U_{mf})} \quad \text{for } N_r = 0 \quad (2b)$$

As was shown earlier too low a N_r value gave some inaccuracies in calculating H_k . Based on the data

fig. 2 A-H. Examples of H_k versus Scaling parameter S .

the minimum value $N_r = 1$ was estimated. For the residence time distribution (RTD) measurements ($N_r = 0$), the N_r value was simply left out. This implies that the S values for reaction data are smaller than those calculated from RTD data.

5. RESULTS AND DISCUSSION

Plots of H_k versus S for the several experiments are given in fig. 2. They show that, in all cases, H_k could be correlated by:

$$H_k = c_1 + c_2 \cdot S \quad (3)$$

In equation (3) S and the constants c_1 and c_2 are dependent on the physical properties of the solid and gas phases and of the bed geometry. Equation (3) holds for all experiments, even for those where internals or perforated plates were used. Even for experiments with a wide range in H , D and U/U_{mf} (table 2), H_k could be correlated with S by equation (3), indicating the wide applicability of the equation. Van Swaaij and Zuiderweg (1972) found that $H_k \sim H^{0.5} \cdot D^{0.42}$. We found virtually the same dependency by equations (2) and (3). Furthermore this factor makes it possible to compare experiments, whenever reactor geometry, reaction type, particle type and gas properties are the same.

Distributor (porous plate), catalyst (quartz sand) and reaction type (ozone decomposition) were the same for the experiments of Van Swaaij/Zuiderweg (1972), Bauer (1980) and ourselves. Plotting H_k , from these experiments, as a function of d_p with constant Scaling parameter S , confirmed our theory (see introduction) for A and small B powders: from fig. (3) it can be seen that H_k indeed rises with increasing particle size. A relatively small S value corresponds to a small reactor. Therefore figure (3) also shows that very small scale experiments are not suited for investigating the influence of d_p on H_k . This was also found by de Groot (1967).

fig. 3. Height of a mass transfer unit versus mean particle size, with constant Scaling parameter (from various authors, see text).

6. CONCLUSIONS

H_k values obtained from our experiments and a large number of literature data were analyzed. All these data could be correlated with the Scaling parameter S .

The constants c_1 and c_2 can be obtained from small scale experiments, as long as the scale is not too small. Therefore we believe that equations (2) and (3) can be useful in scale up of fluidized beds.

It was shown that, for A/B type powders, H_k rises with increasing particle size. This was also found from theoretical considerations. This shows that particle size selection is an important factor for improving the conversion in a fluidized bed.

Further experiments will be carried out to determine the influence of d_p on H_k for large particle systems.

LIST OF SYMBOLS

c_1	constant in eq. (3).
c_2	constant in eq. (3).
C_e	exit concentration.
C_f	feed concentration.
D	bed diameter, m.
d_{32}	mean sieve particle size, μm .
d_p	particle size, μm .
D_g	diffusion coefficient of (tracer)gas, m^2/s .
H	bed height, m.
H_k	height of a mass transfer unit, m.
k_r	reaction rate constant, $\text{m}^3/(\text{kg}\cdot\text{s})$.
N_r	number of reaction units.
N_k	number of mass transfer units.
N_t	total number of units.
S	Scaling parameter, m.
U	superficial gas velocity, m/s .
U_{mf}	minimum fluidization velocity, m/s .

Greek

ϵ_{mf}	bed porosity at minimum fluidization.
μ_g	gas viscosity, $\text{N}\cdot\text{s}/\text{m}^2$.
ρ_g	gas density, kg/m^3 .
ρ_p	particle density, kg/m^3 .

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