

EFFECT OF MIXING OF GRANULAR MATERIAL ON CONTINUOUS PROCESS IN FLUIDIZED BEDS

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The effect of ideal and non-ideal mixing of granular material on continuous process in fluidized beds is studied. The study is based on distribution of residence times of particles in the equipment. For description of non-ideal mixing the Gilliland-Mason's equation is used and subjected to an analysis. Application of the derived procedure is demonstrated on the kinetic relation of such type which is well representing the kinetics of different chemical reactions and of the fluidized bed drying of crystal sugar and of other materials. In addition to theoretical solution a survey of experimental data together with their discussion is given.

This study is related to the paper¹ in which we have restricted ourselves to scaling-up of results of fluidized bed drying from a batch process to a continuous one for ideal mixing of granular particles in the fluidized bed. Here, we have concentrated ourselves primarily to the process in general and to the non-ideal mixing. The continuous processes are described on basis of the statistics of residence times^{2,3} which was in general also applied by other authors^{4,5}. Theoretical considerations are applied to the kinetic dependence of a relatively general type.

THEORETICAL

Let us consider the system of solid particles of the same properties which are subjected to a certain process inside a continuous fluidized bed reactor. The process takes place only inside the fluidized bed in the reactor and it is suddenly interrupted by a step change at the exit from the reactor. Let us assume that the particles are not affecting mutually each other at their passage through the reactor. The initial concentration of reacting (active) matter in each particle at its inlet into the reactor is expressed as its relative mass fraction c_0 . The state in the equipment is considered to be steady. Concentration of the reacting component in the particle c leaving the reactor thus depends under the given conditions only on its initial concentration c_0 and on its residence time τ in the fluidized bed.

Actually, the processes in the system are at the most quasistationary: Due to the effect of inhomogeneities in the gas-fluidized bed pulsations take place, local values

of state quantities vary in the bed and thus the local concentrations of reactants in the granular material vary as well. The over-all holdup of the solid phase in the reactor and thus its outlet concentration also varies. But if the amplitude and the length of the period of mean pulse are small we can base our description on mean values and on the assumption that the process is steady.

Individual particles of the solid phase will reside in the fluidized bed for a generally different time intervals. The residence time distribution of particles in the fluidized bed depends on a number of circumstances, first of all on the dynamics of the fluidized bed and on the design of the reactor.

Residence time distribution of particles in the fluidized bed is expressed by the frequency function (density of probability distribution)² $E(\tau; \bar{\tau}, \{P\})$, where the argument is the residence time τ of individual particles in the fluidized bed. Parameters of the frequency function are the mean residence time $\bar{\tau}$ of particles in the active space defined by the relation

$$\bar{\tau} \equiv m_s / \dot{m}_s \quad (1)$$

and the set of other mutually independent parameters $\{P\}$. The number of these parameters depends on the type of frequency function. With regard to the considerable number of solid particles in the fluidized bed it is admissible to consider the frequency function of residence times and thus of the outlet concentration c as well to be continuous.

The mean concentration \bar{c} in particles leaving the fluidized bed is considered to be the weighed mean from the function $c(\tau)$, *i.e.* from the dependence of concentration of the reacting substance in granular material on time in a discontinuous (batch) operation. The weighing function is the frequency function $E(\tau; \bar{\tau}, \{P\})$. The mean concentration of the active component in the particles leaving the reactor is then given by the relation

$$\bar{c}(\bar{\tau}, \{P\}) = \int_0^{\infty} c(\tau) E(\tau; \bar{\tau}, \{P\}) d\tau, \quad (2)$$

where the parameters of the frequency function E are the arguments of the averaged quantity \bar{c} .

As long as the analytical form of functions $E(\tau; \bar{\tau}, \{P\})$, and $c(\tau)$ are known and the function $c(\tau) E(\tau; \bar{\tau}, \{P\})$ is directly integrable or at least as long as it can be expressed by use of tabulated integrals, the solution $\bar{c}(\bar{\tau}, \{P\})$ can be obtained in analytical or other form which simplifies the calculation. In other cases it is first of all necessary to integrate numerically or graphically. Let us first to present the analytical solution of Eq. (2) for special cases of the frequency function of residence times for the plug flow and ideal mixing of particles and then for the non-ideal mixing:

Plug Flow

The relations hold

$$E(\tau; \bar{\tau}) \begin{cases} \rightarrow \infty, & [\tau/\bar{\tau} = 1] \\ = 0, & [\tau/\bar{\tau} \neq 1]. \end{cases}$$

Thus we obtain

$$\bar{c}(\bar{\tau}) = \int_0^{\infty} c(\tau) E(\tau; \bar{\tau}) d\tau = c(\bar{\tau}), \quad (2a)$$

where $c(\tau)$ is the concentration of active component in granular material for a batch process in time $\tau = \bar{\tau}$ where $\bar{\tau}$ is the characteristics of the continuous process according to Eq. (1).

Ideal Mixing

In this case the frequency function is given by the equation²

$$E(\tau; \bar{\tau}) = (1/\bar{\tau}) \exp(-\tau/\bar{\tau}). \quad (3)$$

By use of this relation in Eq. (2) the mean exit concentration for ideal mixing of particles in the bed is obtained in the form

$$\bar{c}^*(\bar{\tau}) = \int_0^{\infty} c(\tau) (1/\bar{\tau}) \exp(-\tau/\bar{\tau}) d\tau. \quad (2b)$$

If the concrete dependence of concentration c on time τ is known, the integration according to Eq. (2b) can be carried out. The needed function $c(\tau)$ can be determined by the known technique with the use of the batch reactor.

As a concrete case let us now consider the dependence $c(\tau)$ given in study⁶ which is satisfactorily describing the batch drying of crystal sugar and of other soluble matters^{7,8}, where $c(\tau)$ is the moisture of material in time τ from the beginning of drying:*

$$c(\tau) = c_0 - \tau/(k_1 + k_2\tau). \quad (4)$$

The quantities k_1 and k_2 are constants whose physical meaning results from Eqs

$$k_1 = \lim_{\tau \rightarrow 0} \left[-\frac{1}{dc(\tau)/d\tau} \right], \quad k_2 = \frac{1}{c_0 - c_r}, \quad (4a,b)$$

* This equation is for the considered case simultaneously the concrete expression of relation (2a) for plug flow of particles if the substitution $\tau = \bar{\tau}$ is made.

where

$$c_r = \lim_{\tau \rightarrow \infty} c(\tau) \quad (4c)$$

is the equilibrium moisture of dried material under the conditions of drying. The constants k_1 and k_2 are dependent on the initial moisture c_0 . As it has already been demonstrated^{7,8} it is advisable to define a new constant

$$K_u \equiv k_2^2/k_1, \quad (5)$$

which is independent of the initial moisture of material. For drying is $K_u > 0$.

Eq. (4) is also expressing the behaviour of other operations inclusive the chemical reactions especially those taking place in surface layers. The constants k_1 and k_2 have a similar physical meaning as those given by Eqs (4a) and (4b), or they may have a purely empirical character.

By combining Eqs (2b) and (4) the relation is obtained

$$\bar{c}^*(\bar{\tau}) = c_r - (k_1/k_2^2 \bar{\tau}) \exp(k_1/k_2 \bar{\tau}) \text{Ei}(-k_1/k_2 \bar{\tau}), \quad (6)$$

where the integral exponential function $\text{Ei}(z)$ is determined by the relation

$$\text{Ei}(z) = \int_{-\infty}^z (1/y) \exp(y) dy, \quad [z < 0]. \quad (6a)$$

If from Eqs (4b) and (5) the substitutions are made the relation is obtained

$$\bar{c}^*(\bar{\tau}) = c_r - (1/K_u \bar{\tau}) \exp[1/K_u(c_0 - c) \bar{\tau}] \text{Ei}[-1/K_u(c_0 - c_r) \bar{\tau}]. \quad (7)$$

Eqs (6) and (7) are satisfying the known properties of the course of the continuous process:

$$\lim_{\bar{\tau} \rightarrow 0+} \bar{c}^*(\bar{\tau}) = c_0, \quad \lim_{\bar{\tau} \rightarrow \infty} \bar{c}^*(\bar{\tau}) = c_r.$$

The integral exponential function $\text{Ei}(z)$ for $z < 0$ is tabulated⁹ or it can be evaluated by use of series¹⁰.

Nonideal Mixing

Nonideal mixing can be studied by use of various models: The purely empirical expression of nonideal mixing is based on comparison of the empirically found distribution function with its shape for ideal mixing of particles. The size of nonideality is then quantitatively expressed *e.g.*¹¹ by segregation SG. Segregation SG has the values in the range $(-1, 1/e)$. The value $\text{SG} = -1$ corresponds to one limiting case

when the whole fluidized bed represents the inefficient space ("dead water") and the value $SG = 1/e$ corresponds to the second limiting case which is the plug flow of particles. Of course for ideal mixing $SG = 0$.

Semiempirical expression of nonideal mixing is based on the use of more or less complex models. The simpler one is the set of ideal mixers with the possible inclusion of backmixing¹². The formally simple model has been proposed by Gilliland and Mason³ which is used in this study in spite of its shortcomings which are discussed in the following part. It is expressed by Eq.

$$\ln [1 - F(\tau; \bar{\tau}, S)] = -S[\tau/\bar{\tau} - (S - 1)/S], \quad (8)$$

where $F(\tau; \bar{\tau}, S)$ is the distribution function of residence times τ with parameters $\bar{\tau}$ and S . The distribution and the above given frequency functions are related by

$$E(\tau; \bar{\tau}, \{P\}) = \partial F(\tau; \bar{\tau}, \{P\})/\partial \tau. \quad (9)$$

Let us introduce

$$J \equiv (S - 1)/S \quad (10)$$

where for ideal mixing: $S = 1, J = 0$

and for plug flow: $S \rightarrow \infty, J = 1$.

The case for which $S < 1$ represents the unsuitably situated inlet and outlet of the material from the equipment⁴.

The frequency function is obtained from Eq. (8) by derivation according to Eq. (9) and by arrangement in the form

$$E(\tau; \bar{\tau}, S) = (S/\bar{\tau}) \exp(S - 1) \exp(-S\tau/\bar{\tau}). \quad (11)$$

But application of this Eq. for evaluation of experimental data is restricted. This is because the function $F(\tau; \bar{\tau}, S)$ appearing in Eq. (8), for the case $S \neq 1$ which is here studied, is actually not the distribution function as

$$\lim_{\tau \rightarrow 0} F(\tau; \bar{\tau}, S) \neq 0, \quad [S \neq 1].$$

But from Eq. (8) results

$$\lim_{\tau \rightarrow 0} F(\tau; \bar{\tau}, S) = 1 - \exp(S - 1) \quad (8a)$$

and

$$F(\tau; \bar{\tau}, S) = 0, \quad (8b)$$

when

$$\tau/\bar{\tau} = (S - 1)/S.$$

Thus the function $F(\tau; \bar{\tau}, S)$ has no physical meaning for $S < 1$ in the interval $\langle 0; 1 - \exp[S - 1] \rangle$ when $\tau < 0$ and for $S > 1$ for $\tau/\bar{\tau} \in \langle 0; S - 1/S \rangle$ when $F(\tau; \bar{\tau}, S) < 0$. Eq. (8) fits the plug flow only formally ($S \rightarrow \infty$) as it does not satisfy the condition of the step change which is required from the distribution function $F(\tau; \bar{\tau}, \{P\})$ describing the plug flow. Eq. (8) is practically suitable for the dimensionless time $\tau/\bar{\tau}$ greater than some definite time Θ which must be determined experimentally for the selected accuracy and for the given operating conditions. The function $F(\tau; \bar{\tau}, S)$ according to Eq. (8) is in a solid line plotted in Fig. 1 for $S = 0.8$ and $S = 1.2$. For comparison there is also plotted the function $F(\tau; \bar{\tau})$ for the case of ideal mixing *i.e.* for $S = 1$. In Fig. 1 is moreover plotted as the dashed line the actual (experimental) shape of the distribution function for non-ideal mixing with the denoted time Θ . By the limited validity of Eq. (8) it can be explained, beside other effects, the irreproducibility of the constant S for small values of $\tau/\bar{\tau}$.

With regard to the mentioned disadvantage, the Gilliland-Mason's equation can be used for evaluation of combined Eqs (2) and (11) only for processes with behaviour very close to ideal mixing ($S \approx 1$). This disadvantage can be overcome if it is possible to obtain the distribution or frequency function experimentally with a sufficient accuracy in the range of values $\tau/\bar{\tau} \in \langle 0; \Theta \rangle$.

It is usually very difficult to obtain the distribution and frequency functions especially for narrow time intervals. But if we are satisfied by determination of the upper and lower boundaries of the interval into which the outlet concentration c , is situated we can apply the following procedure:

Let us consider the following two cases:

$0 < S < 1$: This case can be described in a simplified manner so that one part of material leaves without a change in concentration of the active component in the time $\tau \rightarrow 0+$ and the remaining part reacts at mixing according to Eq. (11). $S > 1$:

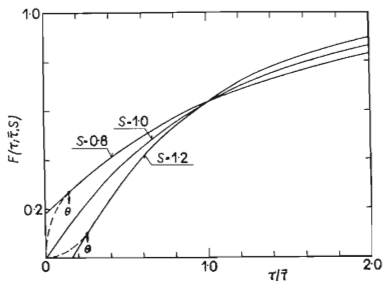


FIG. 1
Experimental (---) and Calculated (—) Distribution Functions F

Calculation is based on the Gilliland-Mason's equation for different values of S .

This case can be described so that no part of the "freshly" entering material is leaving the equipment up to the time $\tau = \bar{\tau} (S - 1)/S$ and that from this time the material reacts at mixing according to Eq. (11).

In both cases the actual mean concentration of the active component in the outlet granular material is situated in the range which is given on the one side by Eq. (2b) (corresponding to ideal mixing) and on the other side by one of both just mentioned simplified concepts. Solution of both alternatives is further on demonstrated on a concrete situation of fluidized bed drying of crystal sugar whose kinetics results from Eq. (4):

$0 < S < 1$: In this case the actual mean moisture of the leaving material is $\bar{c}(\bar{\tau}, \{P\})$ greater than the value $\bar{c}^*(\bar{\tau})$ calculated for the given physical conditions at ideal mixing according to Eqs (6) or (7). But the actual moisture is smaller than the value $\bar{c}^+(\bar{\tau}, S)$ calculated from the relation derived for non-ideal mixing according to the Gilliland-Mason's *i.e.*

$$\bar{c}^+(\bar{\tau}, S) = [1 - \exp(S - 1)] c_0 + \exp(S - 1) \int_0^{\infty} [c_0 - \tau/(k_1 + k_2\tau)] \times \\ \times (S/\bar{\tau}) \exp(S - 1) \exp(-S\tau/\bar{\tau}) d\tau. \quad (12)$$

The first term in Eq. (12) represents the amount of water per unit amount of dry material which left the dryer in the time $\tau \rightarrow 0+$ with the granular material without a change in its moisture:

$$\lim_{\tau \rightarrow 0+} F(\tau; \bar{\tau}, S) c(\tau) = [1 - \exp(S - 1)] c_0. \quad (12a)$$

The second term in Eq. (12) represents the relative amount of water which remained after the outlet from the dryer in the fraction of material $\exp(S - 1)$ which took part in drying.

By partial integration in Eq. (12) with regard to Eq. (5) the relation is obtained

$$\bar{c}^+(\bar{\tau}, S) = c_0[1 - \exp(S - 1)] + c_r \exp[2(S - 1)] - \\ - \{S \exp[2(S - 1) + k_1 S/k_2 \bar{\tau}]/K_u \bar{\tau}\} \text{Ei}(-k_1 S/k_2 \bar{\tau}). \quad (13)$$

If $\bar{c}(\bar{\tau}, \{P\})$ is the actual mean moisture of material at the outlet from the dryer then it holds $\bar{c}^+(\bar{\tau}, S) \geq \bar{c}(\bar{\tau}, \{P\}) \geq \bar{c}^*(\bar{\tau})$.

The function $\bar{c}^+(\bar{\tau}, S)$ according to Eq. (13) has the following properties

$$\lim_{S \rightarrow 1} \bar{c}^+(\bar{\tau}, S) = \bar{c}^*(\bar{\tau});$$

$$\lim_{S \rightarrow 0} \bar{c}^+(\bar{\tau}, S) = c_0[1 - \exp(S - 1)] + c_r e^{-2};$$

$$\lim_{\bar{\tau} \rightarrow 0+} \bar{c}^+(\bar{\tau}, S) = c_0 \{1 - \exp(S - 1) + \exp[2(S - 1)]\} < c_0 ;$$

$$\lim_{\bar{\tau} \rightarrow \infty} \bar{c}^+(\bar{\tau}, S) = c_0 [1 - \exp(S - 1)] + c_r \exp[2(S - 1)] ;$$

$$c_r < \lim_{\bar{\tau} \rightarrow \infty} \bar{c}^+(\bar{\tau}, S) < c_0 .$$

At the mean residence time $\bar{\tau} \rightarrow \infty$ and value $S < 1$ the material must have a greater moisture than it corresponds to equilibrium value c_r , as part of material has passed through the dryer in the time $\tau \rightarrow 0+$ without a changed moisture.

$S > 1$: In this case the actual moisture $\bar{c}(\bar{\tau}, \{P\})$ is smaller than the value calculated at the assumption of ideal mixing according to Eqs (6) or (7). But the actual moisture is greater than the value $\bar{c}^+(\bar{\tau}, S)$ determined according to the equation

$$\bar{c}^+(\bar{\tau}, S) = \int_{[(S-1)/S]\bar{\tau}}^{\infty} [c_0 - \tau/(k_1 + k_2\tau)] (S/\tau) \exp(S - 1) \exp(-S\tau/\bar{\tau}) d\tau . \quad (14)$$

The lower limit of integral $\bar{\tau}(S - 1)/S$ represents the time in which the function $F(\tau; \bar{\tau}, S)$ has the zero value (Fig. 1). By integration in Eq. (14) the relation is obtained

$$\bar{c}^+(\bar{\tau}, S) = c_r - \{S \exp[S - 1 + (k_1 S/k_2 \bar{\tau})]/K_u \bar{\tau}\} \text{Ei}[-(k_1 S/k_2 \bar{\tau}) - S + 1] . \quad (15)$$

The inequality holds

$$\bar{c}^+(\bar{\tau}, S) \leq \bar{c}(\bar{\tau}, \{P\}) \leq \bar{c}^*(\bar{\tau}) .$$

The function $\bar{c}^+(\bar{\tau}, S)$ according to Eq. (15) has the following properties

$$\lim_{S \rightarrow 1} \bar{c}^+(\bar{\tau}, S) = \bar{c}^*(\bar{\tau}) ;$$

$$\lim_{S \rightarrow \infty} \bar{c}^+(\bar{\tau}, S) = c_0 - \bar{\tau}/(k_1 + k_2 \bar{\tau}) ;$$

$$\lim_{\bar{\tau} \rightarrow 0} \bar{c}^+(\bar{\tau}, S) = c_0 ;$$

$$\lim_{\bar{\tau} \rightarrow \infty} \bar{c}^+(\bar{\tau}, S) = c_r .$$

EXPERIMENTAL

Experimental apparatus and operating conditions. The apparatus is in detail described in our recent publication⁷. Its most important part is the metal fluidized bed column with the diameter 97.5 mm equipped by the multiorifice grid, movable weir, hopper for introduction of the tracer (KCl), storage vessel for granular material with the vibrating doser and rotary sampler of fractions. The important independent variables were fixed in the experiments in intervals: Mean

equivalent diameter of sugar crystals d [mm] $\in \langle 0.315; 1 \rangle$, mean residence time $\bar{\tau}$ [s] $\in \langle 37.7; 255 \rangle$, mass of sugar m_s [g] $\in \langle 105; 499 \rangle$, ratio of the fluidized bed height to the diameter of equipment $H/D \in \langle 0.5; 1 \rangle$, relative velocity of air $W \in \langle 2; 4.47 \rangle$ and the ratio $w_{p,KCl}/w_{p,S} \in \langle 0.908; 2.162 \rangle$. Summary of experimental data has been given in detail in our earlier publication¹.

Determination of the frequency function: Crystals of potassium chloride were used as the tracer for determination of the frequency function. Their minimum fluidization velocity was close to that of fluidized sugar. Potassium chloride was added at once (its concentration was equal to 1 up to 1.5% of the sugar holdup in the column) into the bed in steady state (realisation of the impulse function). Concentration KCl in the outlet material then was measured conductometrically.

RESULTS AND DISCUSSION

The constant S in the Gilliland–Mason's equation was determined by evaluation of experimental data¹. The values S were also recalculated to the segregation SG according to the approximate relations¹

$$SG \approx (1/e)(S - 1)/S, \quad [S > 1] \quad (16)$$

and

$$SG \approx -\{1 - (1/e) - (1/eS)[\exp(S) - 1]\}, \quad [S < 1]. \quad (17)$$

By the analysis of the results we have determined that in the range of experimental errors and in the range of our experimental conditions the ratio $w_{p,KCl}/w_{p,S}$ has no effect on the segregation SG. It can be stated according to the experimental data that, in agreement with the literature^{4,13}, segregation SG is independent of the mean residence time τ .

The analysis of variance of data¹⁴ has pointed to the fact that on the 5% level of statistical significance there has appeared only the effect of interaction of the ratio H/D and of the relative air velocity W on segregation SG. The effect of individual factors, d , H/D and W has not been – with respect to the given experimental error and to the range of these variables – statistically significant.

By evaluation of experimental data the relation has been obtained

$$SG = -0.65 + 0.22W + 0.88H/D - 0.29W(H/D). \quad (18)$$

From 33 experimental data not affected by the blunders data 75% of them are in the range $S \in \langle 0.8; 1.2 \rangle$. It can be concluded on basis of these results that the found distribution functions of residence times of particles in the fluidized bed under the given experimental conditions are mostly very close to ideal mixing. For such cases the application of the Gilliland–Mason's Eq. according to the proposed procedure is acceptable.

The proposed procedure of calculation of the mean outlet moisture for non-ideal mixing is illustrated in Table I. The results are calculated from Eqs (7), (13) and (15).

The subscripts max and min are denoting the interval in which the mean moisture of material at the outlet from the fluidized bed drier can be expected, for the mean residence time $\bar{\tau}$ [s] = 60, 120, ∞ and for the value of the constant in the Gilliland-Mason's Eq. $S = 0.8; 0.9; 0.95; 1; 1.1; 1.2; \infty$. The dried material is crystal sugar with the initial and equilibrium moisture $c_0 = 0.01$ and $c_r = 2 \cdot 10^{-4}$. The rate constant determined at the batchwise operation is $K_u = 176 \text{ s}^{-1}$.

Calculation of moisture at the assumption of ideal mixing can give – as is obvious from the given Table – a considerably smaller outlet moisture than in the case of the proposed procedure considering the non-ideal mixing at $S < 1$. However, for technical purposes it is necessary to take into consideration such a moisture which guarantees the result and thus a reliable operation of the equipment. If it is *e.g.* required to obtain the mean outlet moisture $0.50 \cdot 10^{-3}$ at $S = 0.95$ at the conditions for which the Table was calculated we can see that this moisture would safely be reached neither at a very long mean residence time of material in the drier. But according to the calculation by use of Eq. (7) for ideal mixing the material will leave the drier with the required moisture if the mean time of drying is kept in between 60 and 120 s (Table I).

If necessary for reasons of reliability to apply the results giving larger outlet moisture it is recommended for cases when $S < 1$ to use Eq. (13) and in cases when $S \geq 1$ Eqs (6) or (7). In this way we secure the outlet moisture of material to be at least equal to the calculated value.

It can be seen that to each value of S in our example there exists some limiting smallest outlet moisture which is safely reached. This safe limiting value is reached

TABLE I

Example of Mean Outlet Moisture of Crystal Sugar \bar{c} Calculated from Eqs (7), (13) and (15) for Various Mean Drying Times $\bar{\tau}$ (s) and Various Mixing S in the Fluidized Bed Drier

S	$\tau = 60$		$\bar{c}(\bar{\tau}, S) \cdot 10^3$ $\bar{\tau} = 120$		$\bar{\tau} = \infty$	
	max	min	max	min	max	min
0.8	2.12	0.56	2.09	0.44	1.94	0.20
0.9	1.41	0.56	1.30	0.44	1.11	0.20
0.95	1.01	0.56	0.86	0.44	0.68	0.20
1.00	0.56		0.44		0.20	
1.1	0.56	0.38	0.44	0.29	0.20	
1.2	0.56	0.33	0.44	0.27	0.20	
∞	0.56	0.29	0.44	0.25	0.20	

at the mean residence time $\bar{\tau} \rightarrow \infty$. If $S \geq 1$ the limiting moisture equals to the equilibrium value c_r (in our case $c_r = 0.20 \cdot 10^{-3}$).

It is also obvious from the presented Table that the effect of the mean drying time on the outlet moisture is relatively smaller when $S < 1$ than when $S > 1$. The necessary large mean residence times resulting from the outlet concentration of the active substance as $S < 1$ are caused by the material passing through the equipment without taking part in the process, *i.e.* as the by pass. This disadvantage can be solved by arrangement of the active space of the equipment which prevents by passing and removes the inactive spaces (suitable choice of the inlet and outlet of material, built in baffles and intensive mixing).

CONCLUSIONS

Though this study is partially oriented to problems of fluidized bed drying it has a wider validity *e.g.* for solution of heterogeneous chemical reactors with a fluidized bed of granular material. These main conclusions have been obtained: 1. The Gilliland-Mason's equation³ used in literature for description of non-ideal mixing of particles is not suitable for this process in its initial stages. The deviations are the greater the more the mixing differs from ideal mixing. In this way it is explained why the Gilliland-Mason's equation does not fit the experimental data at the beginning of the operation (*e.g.*^{13,15}). 2. The relations are given for calculation of the outlet concentration of the active component in the solid phase in the continuous fluidized bed operation which includes the effect of ideal and non-ideal mixing of granular material. These relations are applied to one type of curve for discontinuous drying given by Eq. (4) which is describing well the drying of crystal sugar and of other especially soluble substances. This equation is also expressing the kinetics of other operations inclusive some chemical reactions. 3. The procedure is proposed for estimation of the outlet concentration of the active component in the solid phase in the continuous fluidized bed process through which is secured the reliable design of fluidized bed units.

LIST OF SYMBOLS

c	concentration (relative mass fraction) of active component in the solid phase (moisture of granular material)
\bar{c}	mean concentration of active component in solid phase
\bar{c}^*	mean concentration of active component in solid phase at ideal mixing
\bar{c}^+	mean concentration of active component in solid phase at non-ideal mixing
c_0	initial concentration of active component in solid phase
c_r	equilibrium concentration of active component in solid phase
d	mean equivalent diameter of granular material
D	inner diameter of fluidization column

$E(\tau)$	frequency function of residence times
$Ei(z)$	integral exponential function
$F(\tau)$	distribution function of residence times
H	height of fluidized bed
J	constant defined by Eq. (10)
k_1	constant in Eq. (4)
k_2	constant in Eq. (4)
K_u	rate constant defined by Eq. (5)
m_s	holdup of granular material inside the column
\dot{m}_s	mass flow rate of granular material through the column
S	constant in Eq. (8) characterizing the non-ideality of mixing
SG	segregation characterizing the non-ideality of mixing
w	superficial fluid velocity
w_p	minimum fluidization velocity
W	relative fluid velocity, $W \equiv w/w_p$
y	auxiliary variable
z	upper limit of integral exponential function
Θ	dimensionless time, at which the distribution function of residence times merges with the curve according to the Gilliland-Mason's equation
τ	time, residence time of individual particles of material inside the equipment
$\bar{\tau}$	mean residue time of granular material in the equipment

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