Solid-Solid and Gas-Gas Mixing Properties of a New Two-Impinging-Streams Mixer

A new device for solid-solid and gas-gas continuous mixing, based on the impingement of two streams, was developed and successfully tested. It was found that in comparison to other available solid-solid pneumatic mixers, such as the fluidized bed, spouted bed, air mixer, and mixing silo, the new device is very convenient and easy to operate and more energy-efficient. This is also true for gas-gas mixing when compared with a fluidized bed. A model for the gas phase behavior in the mixing of two gases was successfully tested, which supplements the picture of the solid particles behavior predicted from a model previously derived by the authors.

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SCOPE

Solid-solid and gas-gas mixing processes find wide application in many technological processes. The present study concerns a continuous mixer of a new type which was developed and successfully tested. The new mixer for solids is based on two streams of air, each containing a different kind of solid, which are forced to impinge at a certain point. The oscillatory behavior of the particles at the impingement zone causes intensive mixing to form a homogeneous solid mixture. The same mixer can be applied also for mixing of gases when the two streams contain different gases. An evaluation of the performance of the mixer was made by comparing it with fluidized-bed and spouted-bed mixers, which are well-known in the field of solid mixing.

CONCLUSIONS AND SIGNIFICANCE

The major result of the present work is the development of a new mixer for solid-solid and gas-gas continuous mixing based on the principle of two impinging streams. Experimental tests of the new mixer proved it to be convenient and easy to operate, as well as superior with respect to energy consumption in comparison with other available pneumatic mixers such as the fluidized bed, spouted bed, and mixing silo. The specific energy for mixing of solids in the present mixer varied between 300 to 450 J/kg which is less by a factor of 2 to 180 than the other devices, according to published data and to values measured by us. Similar conclusions regarding the energy demands and convenience in operation of the new mixer were also drawn for gas-gas mixing. Other important characteristics of the new mixer are as follows:

a) The pressure drop across the mixer in solid mixing is constant in time, while in fluidized or spouted-bed batch mixers the pressure needed for incipience of fluidization is 2 to 4 times higher than the pressure needed to maintain fluidization conditions during steady state operation.

b) Significant pulverization of alumina particles was observed both in the fluidized-bed and spouted-bed mixers. In contrast, attrition of alumina in our mixer could hardly be detected.

c) Unlike the case in the batch mixers, segregation of the solid particles cannot occur in this new mixer.

d) For gas-gas mixing the reactor may be looked upon as composed in principle of two sections: a well-mixed zone, and a zone in which the gas flows in plug flow. Thus, when mixing of gases takes place the output of this reactor is a perfectly mixed solution. In contrast to the oscillatory behavior of solid particles in the reactor, in the mixing of gases the gaseous phase does not reveal any oscillatory behavior.

INTRODUCTION

Solid-solid and gas-gas mixing processes are important operations in chemical engineering and in other fields of technology. In general, such operations can be carried out either in a batch mode or continuously. Continuous mixing is widely accepted as an economical and practical operation and it has two functions: first, to intermingle two or more streams to produce an intimate mixture, and second, to smooth out inhomogeneities in a product stream.

The objectives of the present investigation are: 1) to describe the characteristics of a novel continuous mixer for solid-solid and gas-gas systems; 2) to convince the potential user that the new device is superior to available devices directed to the same purpose; 3) to provide means for scale-up and estimation of the energy consumption in mixing solids; and 4) to establish a model describing the gas phase behavior, which supplements the model suggested by the authors in the past (Luzzatto et al., 1984) for the solid phase behavior. Detailed reviews of available mixing equipment and processes appear in Weidenbaum (1958), Ries (1969), Perry (1973), and Mueller (1981). The present mixer for solids may be included

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in the class of mixers designated as pneumatic mixers, such as the fluidized bed (Nicholson and Smith, 1966), spouted bed (Mathur and Epstein, 1974), air mixer, and mixing silo (Mueller, 1981). Among the devices used for gas-gas mixing such as the injector, baffle-column, propeller, turbine, or fluidized bed, the new mixer may be included in the class of jet mixers (Perry, 1950).

In all cases of solid mixing, air is employed as the mixing medium. The air is fed to the bottom of a vessel containing two or more kinds of solids, and while it flows through the particles or powder, mixing takes place. Usually, this operation is carried out batchwise because this is an easier way to operate. As will be shown later, the major drawback of the available pneumatic mixers is their significantly higher energy consumption in comparison to the new device, for the same degree of mixing of solids.

The performance characteristics of solid mixing equipment which should be studied before selecting or comparing solid-mixing devices are mixing time and power. Mixing time usually should be less than 15 minutes per batch. In addition, segregation ("unmixing") can occur if the batch mixing time is too long. Sufficient power must be supplied to overcome the maximal resistance when the mixer starts from rest. For this reason, air-supply equipment is much more powerful and expensive than is required during the actual mixing. The additional factors of uniformity of mixing, charging and discharging, cleaning, agglomerate breakdown, attrition, dust formation, electrostatic charge, equipment wear, contamination of product, heating or cooling, flexibility, and others should be taken into account. In the case of gas-gas mixing, the most important factors which should be considered before selecting the mixer are: mixing time, power, uniformity of mixing, and "unmixing" which might occur in batch systems in cases of gases with significant differences in density.

The new mixer for solids consists of two streams of air, each containing a different kind of solid, which are forced to impinge at a certain location. The oscillatory behavior of the particles at the impingement zone causes an intensive mixing of the particles, and consequently a homogeneously mixed material is obtained. For the mixing of gases in the new mixer, separate streams of the two gases are forced to impinge at a certain zone in which intensive mixing takes place.

It should also be noted that the two-impinging-streams mixer is an additional application of a new multipurpose reactor for heterogeneous systems developed by Elperin and Tamir (1982). Additional data on the performance of this reactor were reported by Elperin et al. (1982), Tamir et al. (1984), and Luzzatto et al. (1984).

In order to evaluate the performance of the new mixer for solids, it was compared with fluidized-bed and spouted-bed batch mixers, which are well-known devices in the field of solid mixing. The evaluation of its performance with respect to the gas-phase mixing properties was made by carrying out pulse experiments.

THE TWO-IMPINGING-STREAMS MIXER

In the two-impinging-streams mixer for mixing solids, granular particles are caused to oscillate with a motion similar to the oscillation of the pendulum in a clock. This behavior was demonstrated by Luzzatto et al. (1984). Consequently, if the particles are of two or more different kinds, they will mix and form a homogeneous mixture. As shown in Figure 1, this idea is realized by tangentially feeding two separate identical streams of air containing granular material or powder to a cylindrical reactor, and allowing them to meet at the upper portion thereof. This zone, designated by *12* in Figure 1, is symmetrically positioned with respect to the entrance point *4* of the streams. Particles in the two countercurrent streams may experience direct collision, as a result of which they fall immediately downward. If direct collision does not take place, or if the particles only slightly touch one another, the particles penetrate from one gas stream into the other; they first lose their velocity and then are accelerated by the other stream in a direction opposite to their original one. It is readily understood that the harmonic motion accomplished in the impingement zone produces very efficient

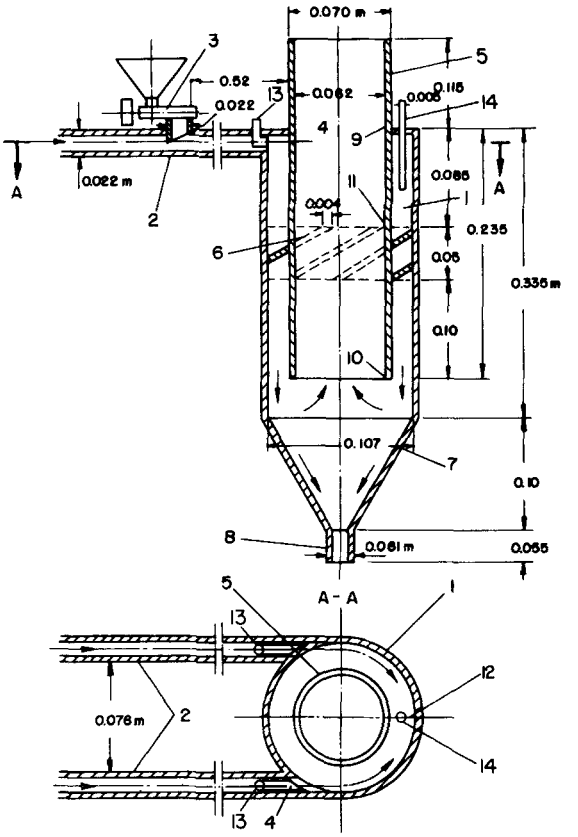

Figure 1. Two-impinging-streams mixer.

1. Annular space
2. Air feed pipes
3. Particle/powder feeder
4. Mixer inlets
5. Air outlet pipe
6. Swirling blades
7. Conical section
8. Mixture discharge port
9, 10, 11, 12. Zone reference points, see text.
13. Tracer inlet
14. Tracer sampling points

mixing. After several collisions, either between the particles themselves or between the particles and the walls of the mixer, the particles lose their speed and are eventually carried away from the active zone by the gas stream flowing out. In the case of the mixing of two gases, two situations may occur: If the volumetric flow rates of the gases are the same, each gas may be introduced through the inlet tubes *4*. If the gas flow rates are different, one stream may be introduced through the inlet tubes *4* and the other one introduced through inlets designated by *13* in Figure 1. Mixing takes place in the annular space *1* which is the active zone. Blades for air swirling, designated by *6*, cause the particles to move radially toward the walls of the mixer and hence to be discharged through opening *8*. The region between points *11* and *10* is designed as a separation stage for air-particle separation. A tracer inlet for pulse experiments is designated *13;* tracer sampling points are designated *14*.

EXPERIMENTAL

Mixing of Solids

The experimental system is shown in Figure 2 and operates as follows: Air is supplied by an appropriate blower through rotameter *3*. By means of regulator *2* the air flow is adjusted so that the flow after the rotameter is split between the two pipes in equal portions. The two kinds of particles are fed to the air streams by means of screw feeders 3 at a point positioned about 0.5 m from the inlet to the mixer. This distance is required in order to allow the particles to accelerate to a velocity sufficient to prevent them from falling downward and be immediately discharged. It should be noted that the particle feed pipes penetrate into the air pipes in a particular arrangement, as shown in Figure 1. This was done in order to create a low-pressure zone at the particle entrance due to the formation of a vena con-

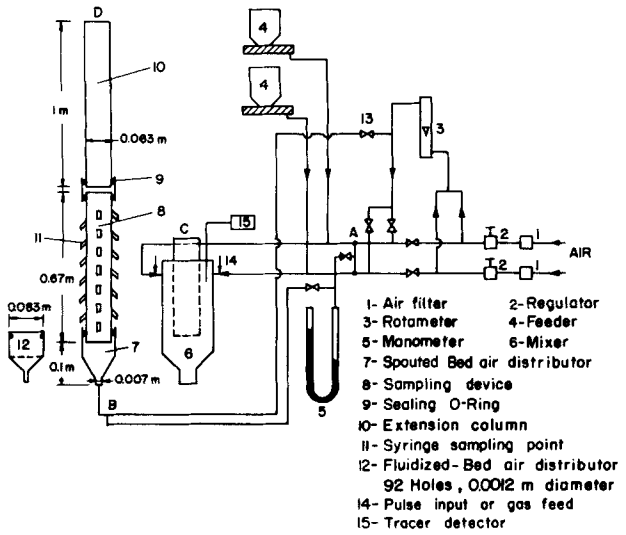

Figure 2. Experimental system.

tracta in the air pipe, for the purpose of preventing back-pressure effects in the particle feed pipe, which might cause nonconstant input mass flow rates. The variables for an experimental run were: (1) the mass flow rates of the two kinds of particles, which were adjusted by varying the speed of rotation of the screw-feeders, and (2) the air flow rate. The mixture leaving the mixer was collected during a certain interval of time (on the order of 60 s) in a special sampling device (8 in Figure 2) which was also utilized as a spouted bed or a fluidized bed as explained later. The sampling device was a Perspex tube 0.67 m high with a diameter of 0.063 m. Thirty-five 0.01 m dia. openings were drilled at intervals of 0.07 m along its height and circumference. Short tubes, 0.026 m long and inclined at 45° with respect to the vertical, were connected to the holes.

A syringe was inserted in each tube by means of which samples of various sizes, (1–2 cm^3), could be withdrawn, depending on the size of the particles which were mixed. The variation in composition of the mixture along the sampling device served as an indication of the performance of the mixer. The composition of the mixture in each sample was determined by counting the number of particles of each kind and expressing it in terms of particles-fraction.

The effectiveness of the new mixer was compared with fluidized-bed and spouted-bed mixers. This was done by modifying the above sampling device as follows. In the case of the spouted bed, a cone-shaped device was connected to the bottom of the sampler as shown in Figure 2. The ratio between the bed diameter and the air inlet pipe was about 10, which is an accepted value (Mathur and Epstein, 1974). In order to use the sampler as a fluidized bed, item *12* of Figure 2 was used. As shown in the figure, pipe *10* was added in order to provide ample space for the expansion of the bed. A typical comparison run was carried out as follows: A mixture of two kinds of particles was prepared in the proportion tested in the new mixer and charged into the spouted or fluidized bed, in order to obtain approximate values of the minimal fluidization velocity and the pressure drops which develop across the bed. Air flow rates in the batch mixers were chosen so as to be above the minimal fluidization conditions and so that the bed would be fully expanded. Efforts were made to employ air flow rates as near as possible to those employed in the new device, for comparison purposes.

After this stage, known quantities of the two types of particles were separately charged into the column so that two layers were obtained. By means of valve *13* (Figure 2) air was introduced into the column at flow rates higher than those for which minimal fluidization took place. After a certain operation time, the air flow was stopped and samples (1–2 cm^3) were collected. After each sampling, air was again introduced for a few seconds and stopped for sample withdrawal. Such a procedure was repeated 4 to 5 times until complete mixing was observed. In each experiment, the pressure drop across the bed and the air flow rate were recorded in order to calculate the energy consumption for particle mixing.

TABLE 1. SPHERICAL PARTICLES PROPERTIES

	$D_1 \times 10^3$ m	ρ_b kg/m^3	ρ_1 kg/m^3	n_1 Particles/g
Yellow Millet Seeds	1.90	730	1,158	206
Brown Seeds	1.52	662	1,085	330
Alumina	1.76	524	1,588	450
Acetal	2.82	780	1,278	59

Mixing of Gases

The characteristics of the reactor with respect to the mixing properties of the gas phase were tested by performing pulse experiments and recording the response at the exit of the reactor and other points. Such a test run consisted of introducing air into the reactor at a constant flow rate through the inlet pipes *2* in Figure 1. At a certain moment a pulse of CO_2 was introduced into the reactor through the tubes designated by *13*. The response of the reactor was determined by continuous suction of samples of air and CO_2 through a Packard CO_2 analyzer and recording the output. Similar experiments were also conducted in the presence of particle flow. In order to explore also the local mixing characteristics of the reactor, samples were withdrawn from several points inside the reactor by introducing the mixture suction probe into the hole designated by *14* in Figure 1. This helped us also in establishing a model for the behavior of the gas phase in the reactor.

RESULTS AND DISCUSSION

Mixing of Solids

The properties of the particles which were employed in the experiments are summarized in Table 1. Typical distributions of particle-fraction X along the sampling device are given in Table 2 for three mixtures: brown seeds-acetal, alumina-acetal, and brown seeds-yellow seeds. For each mixture, distributions were obtained in three mixers, namely, the new continuous mixer, the batch fluidized bed, and the batch spouted bed.

The results correspond to mixing times in the batch mixers identical to the time required for output collection in the contin-

TABLE 2. TYPICAL DISTRIBUTIONS OF PARTICLE-FRACTIONS X_1 ALONG THE SAMPLING DEVICE

	Brown Seeds (1)—Acetal (2)			Alumina (1)—Acetal (2)			Brown Seeds (1)—Yellow Seeds (2)		
	New Mixer	Fluidized Bed	Spouted Bed	New Mixer	Fluidized Bed	Spouted Bed	New Mixer	Fluidized Bed	Spouted Bed
t (s)	93	92.8	91.2	100	104	94	43	43	41
	0.963	0.989	0.938	0.893	0.990	0.989	0.528	0.526	0.348
	0.962	0.966	0.919	0.906	0.973	0.958	0.577	0.438	0.538
	0.960	0.952	0.939	0.936	0.930	0.949	0.522	0.523	0.570
	0.980	0.943	0.908	0.915	0.937	0.893	0.666	0.565	0.562
	0.945	0.963	0.978	0.973	0.891	0.912	0.612	0.532	0.604
	0.948	0.944	0.969	0.946	0.927	0.931	0.611	0.534	0.487
	0.966	0.950	0.942	0.984	0.904	0.820	0.486	0.508	0.494
	0.951	0.941	0.973	0.937	0.829	0.878	0.549	0.543	0.638
	0.953	0.918	0.911	9.926	0.907	0.732	0.596	0.582	0.570
	0.963	0.940	0.994	0.915		0.769	0.556	0.592	0.474
$\bar{X}_1$	0.958	0.951	0.947	0.933	0.921	0.833	0.570	0.534	0.533
σ_{N-1}	0.010	0.019	0.029	0.029	0.047	0.085	0.053	0.043	0.084

TABLE 3. MAIN DATA AND COMPARISON OF PERFORMANCE OF VARIOUS MIXERS IN MIXING BROWN SEEDS (1) AND ACETAL PARTICLES (2)

	New Mixer	Fluidized-Bed Mixer				Spouted-Bed Mixer			
t (s)	9.3	16.8	30.5	47.7	92.8	10.4	31.0	60.8	91.2
N			15.0				10.0		
$Q(m^3/s) \times 10^3$	6.7		6.7				7.5		
$\Delta P(Pa) \times 10^{-3}$	0.78		13.33				53.33		
W_1 (kg)	1.377		0.602				0.602		
W_2 (kg)	0.572		0.250				0.250		
$\overline{X}_1$	0.964	0.935	0.937	0.954	0.951	0.956	0.949	0.951	0.947
σ_{N-1}	0.014	0.029	0.019	0.027	0.019	0.026	0.032	0.028	0.030
$E(J/kg) \times 10^{-3}$	0.298	1.76	3.12	5.00	9.73	4.90	14.55	28.56	42.80
E/E(new mixer)	1.0	5.90	10.5	16.8	32.7	16.4	48.8	95.8	143.6

TABLE 4. MAIN DATA AND COMPARISON OF PERFORMANCE OF VARIOUS MIXERS IN MIXING BROWN SEEDS (1) AND YELLOW SEEDS (2)

	New Mixer	Fluidized-Bed Mixer					Spouted-Bed Mixer					
t (s)	43.1	10.6	18	21	33	43.1	6.6	10.9	14	20.4	30.6	40.8
N	10.0			10.0					10.0			
$Q(m^3/s) \times 10^3$	7.5			5.2					5.2			
$\Delta P(Pa) \times 10^{-3}$	1.08			10.0					13.33			
W_1 (kg)	0.336			0.316					0.316			
W_2 (kg)	0.534			0.502					0.502			
$\overline{X}_1$	0.570	0.538	—	0.516	0.548	0.534	0.546	0.533	—	0.538	0.521	0.533
σ_{N-1}	0.053	0.072	0.053*	0.046	0.020	0.043	0.124	0.072	0.053*	0.036	0.047	0.084
$E(J/kg) \times 10^{-3}$	0.401	0.673	1.15	1.34	2.10	2.74	0.560	0.931	1.19	1.74	2.61	3.47
E/E(new mixer)	1.0	1.68	2.86	3.34	5.24	6.83	1.39	2.32	2.97	4.34	6.51	8.65

* Interpolation from σ_{N-1} vs. t curve.

TABLE 5. MAIN DATA AND COMPARISON OF PERFORMANCE OF VARIOUS MIXERS IN MIXING ALUMINA (1) AND ACETAL (2) PARTICLES

	New Mixer	Fluidized-Bed Mixer				Spouted-Bed Mixer			
t (s)	100	15.5	33.8	67.2	104.4	15.3	33.5	63.5	94
N	15.0		10.0				10.0		
$Q(m^3/s) \times 10^3$	7.5		7.5				6.75		
$\Delta P(Pa) \times 10^{-3}$	1.13		20.53				33.33		
W_1 (kg)	0.559		0.400				0.400		
W_2 (kg)	0.729		0.521				0.521		
$\overline{X}_1$	0.935	0.850	0.917	0.867	0.921	0.882	0.899	0.872	0.883
σ_{N-1}	0.024	0.092	0.071	0.064	0.047	0.080	0.078	0.076	0.085
$E(J/kg) \times 10^{-3}$	0.456	2.59	5.65	11.24	17.5	3.74	8.19	15.52	22.97
E/E(new mixer)	1.0	5.68	12.4	24.7	38.4	8.20	17.9	34.0	50.4

uous mixer. The degree of mixing in all examples can be appreciated by comparing the standard deviations, σ_{N-1}, appearing at the bottom of Table 2. On the basis of this criterion it is concluded that the new mixer is in most cases superior to the other devices.

Tables 3–5 summarize the main values of the operating conditions as well as the variations of the standard deviation σ_{N-1} with time for the batch mixers. σ_{N-1} is also given for the new mixer for total time of output collection nearly equal to the maximal time of mixing for the batch mixers. It should be noted, however, that the time has no influence whatsoever on the value of σ_{N-1} (i.e., on the degree of mixing) in our mixer since, when it works at steady state conditions, the quality of the resulting mixture is dependent only on the air flow rate.

Another important quantity which is reported is the specific energy E in J/kg. This is the energy required for mixing a quantity of W_1 kg of one component with W_2 kg of another component; it is given by:

$$E(J/kg) = \frac{Q\Delta Pt}{(W_1 + W_2)} \quad (1)$$

where Q is the air flow rate in m^3/s; ΔP is the pressure drop in pascals across the mixing device (between points AC in Figure 2 for the new mixer, and BD for the fluidized-bed or spouted-bed mixers); and t is the output collection time in seconds for the new mixer, or the mixing time for the batch mixers. $W_1 + W_2$ is the total amount in kg of particles mixed. The additional quantity appearing at the bottom of the tables is the ratio between the energy consumed in the fluidized-bed or spouted-bed mixers and the energy required by the new mixer. This quantity is indicative of the efficiency of the new mixer.

The general conclusions which may be drawn from Tables 3–5 are as follows:

1. For nearly identical values of the degree of mixing (Table 4), based on the standard deviation σ_{N-1}, the ratio E(fluidized or spouted bed)/E(new mixer) is always significantly greater than unity. In other words, the new mixer is more energy-efficient.

2. When compared on the basis of minimal values of σ_{N-1} (best degree of mixing) in spouted or fluidized-bed mixers, the above ratio is significantly greater than unity, as shown in Tables 3–5. In other words, the new device is again the less energy-consuming one.

3. The greater the difference in density between the mixed particles, the greater is the energy consumption of the batch mixers in comparison to the new continuous mixer. For example, E(fluidized bed)/E(new mixer) = 6.83 for σ_{N-1} = 0.043 for mixing of brown seeds with yellow seeds (Table 4) where the difference in

densities is 6.73%. On the other hand, the above ratio is equal to 38.4 for $\sigma_{N-1} = 0.047$ for mixing of alumina and acetal particles (Table 5) where the difference in densities is 24.3%.

It is also instructive to compare the new device with other available pneumatic mixers, with respect to energy consumption. Mueller (1981) described a pneumatic mixer based on the mixing silo developed by Bayer. This device is similar to a spouted-bed mixer, and for granular material with $D_p = 3 \times 10^{-3}$ m and $\rho =$ 1,000 kg/m^3, the specific energy is of the order of magnitude of 1,800–5,400 J/kg. According to Ries (1978), blending of plastic refractory clay (which is a natural mixture of sand and clay in which the clay is usually present in rather large agglomerates) was tested in different mixing systems, including a pan mixer, a muller mixer, and an intensive counterflow mixer. It was found that regardless of the mixing system, 43,200 to 54,000 J/kg are consumed to achieve a certain degree of homogeneity. Ries (1969) mentioned a work of mixing of 540 to 2,160 J/kg for pneumatic mixers. In comparison to the above figures, it can be seen from Tables 3–5 that in our new mixer the specific energy varies between 298 to 456 J/kg. The major conclusion which can be drawn from the above figures is that the new device is more energy-efficient by a factor of 2 to 180.

Finally, we observed that in both the fluidized-bed and the spouted-bed mixers the initial pressure needed to obtain incipient fluidization is about 2 to 4 times higher than the pressure needed to maintain fluidization. This may be explained as follows: Below the incipience of fluidization, air flows through the void spaces of a dense fixed bed, whereas in a fully-expanded bed the flow of air is through big spaces between the particles. In contrast to this behavior, in our continuous system the pressure drop across the mixer is constant throughout. In mixing alumina with acetal, in both the fluidized and spouted beds, a significant pulverization of the alumina particles was observed. Conversely, attrition of alumina in our mixer could hardly be detected. Moreover, a trend for segregation was observed in the above systems because of the relatively large difference in densities between acetal and alumina particles. In contrast to the batch systems, the segregation phenomenon cannot happen in the new continuous mixer. The segregation phenomenon was easily observed by operating as follows: Air was introduced through the mixed materials in the fluidized bed, its flow being continuously increased. At a certain air flow rate we observed a creeping motion of the alumina particles upward through vacant spaces of the bed, as well as a formation of new spaces when the bigger particles were pushed by the smaller ones moving upward. Eventually, two layers of two kinds of particles were obtained, indicating a complete separation. It should be noted that the solid-solid separation in a fluidized bed is based on the fact that the superficial fluid velocity at minimum fluidizing conditions is dependent on particle size D_p and density ρ. Consequently, the greater the difference in size, the better is the separation. On the other hand, fluidized or spouted beds are less effective for solid mixing under such conditions, while the new mixer is practically unaffected by these factors. The segregation phenomenon was also observed by Wen and Yu (1966) and Horio and Wen (1978).

For practical purposes it would be useful to establish on the basis of the design of the present reactor, shown in Figure 1, a scale-up procedure for the energy consumption needed for the mixing of two solids. In other words, we seek an answer to the following question: "What is the energy needed to prepare a mixture of two solid particles containing W_1 kg of component 1 and W_2 kg of component 2, where the particles have diameters D_1 and D_2 and densities ρ_1 and ρ_2, respectively?". This can be done on the basis of the work by Tamir et al. (1985) as follows: Assume that the mixture particles mean diameter D and density ρ are known, as explained later. Thus, from the equation

$$\alpha^\circ = 5.3\rho D - 2.0 \tag{2}$$

it is possible to calculate an angle α in degrees, where ρ is in kg/m^3 and D is in m. The angle α is a key value characterizing a certain kind of solid particle and is the slope between the so-called η-μ lines where $\eta = Eu_p/Eu_a$ and $\mu = (G_1 + G_2)/G_a$. Once α is known, we

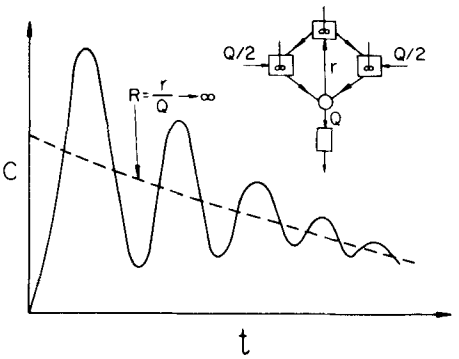

Figure 3. Theoretical response of reactor with recycle ratio approaching infinity.

draw a straight line on a plot of η versus μ which starts at $\eta = 1$ and $\mu = 0$ and has a slope α. By fixing a desired ratio for μ we obtain from this plot the value of η. From a plot of Eu_a vs. Re of our laboratory-scale reactor in Figure 2 of the work by Luzzatto et al. (1984), we obtain the value of Eu_a and thus Eu_p from η.

$Eu_p = \Delta P/(\rho_a U_a^2)$, gives ΔP, and hence from Eq. 1 the desired value of the energy E. It should be noted that all of the foregoing is based on the assumption that we maintain geometrical similitude between the laboratory-scale apparatus of Figure 1 and the large-scale reactor, and that we operate in the range of $Eu_a =$ constant in Figure 2 of Luzzatto et al. (1984). If we wish to operate in the zone in which the Eu_a number varies with the Re number, constraints must also be put on the projected value of Re in order to maintain hydrodynamic similitude. Finally, we propose the following equations for estimating the values of ρ and D needed in Eq. 2. On the basis of the conservation of the total mass $W_1 + W_2$, and that ρ_1, ρ_2 and D_1, D_2 of the particles are known, the density of the mixture is given by

$$\rho = \frac{W_1 + W_2}{W_1/\rho_1 + W_2/\rho_2} \tag{3}$$

An expression for the particle mean diameter of the mixture is derived as follows. As seen, α in Eq. 1 depends solely on the particle diameter and density, and it was found that the shape of the solid particles has practically no relevance to the pressure-drop estimation. Bearing in mind this fact and assuming that the volume of a particle of kind i is given by $V_i = KD_i^3$ and that the mean diameter D is defined by $(n_1 + n_2)KD^3 = n_1KD_1^3 + n_2KD_2^3$ where n_1, n_2 are the number of particles of kind 1 and 2 and K is a shape factor, yields:

$$D = (X_1D_1^3 + X_2D_2^3)^{1/3} \tag{4}$$

The particle-fraction $X_i (i = 1,2)$ in terms of known quantities is given by

$$X_i = \frac{W_i/(\rho_iD_i^3)}{W_1/(\rho_1D_1^3) + W_2/(\rho_2D_2^3)} \tag{5}$$

Optimal Mixing Conditions of Solids

It is well known that optimal mixing of two different solids (i.e., uniform mixture in a continuous mixing operation) is obtained if a mixer exhibits characteristics of a perfectly mixed vessel. It has already been shown by Luzzatto et al. (1984) that this mixer may present an oscillatory behavior which can vary from slight to intense, depending on a recycle ratio R, which represents the extent to which solid particles oscillate in the active zone I of Figure 1, and which is dependent on the gas flow rate.

Whenever a highly oscillatory behavior is obtained—namely, R approaches infinity—the response of the reactor approaches that of a perfectly mixed reactor, as shown in Figure 3. The operating conditions (the gas flow rates) for which such a behavior is obtained are of course dependent on the characteristics of the solid particles employed and should be determined for each case, as explained elsewhere by Luzzatto et al. (1984). Therefore, the optimal oper-

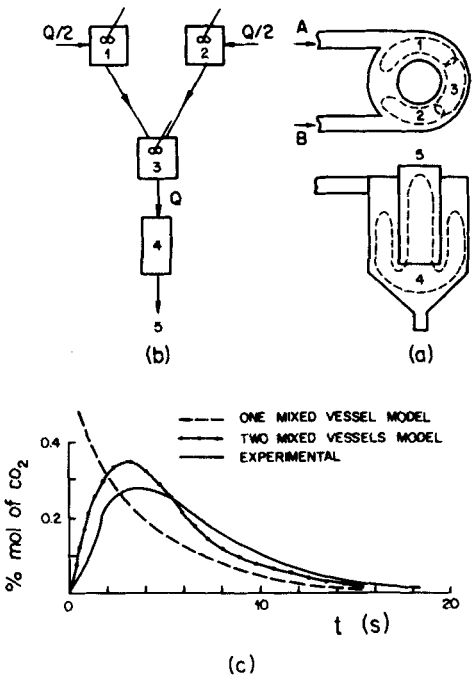

Figure 4. Model of reactor for gas mixing.

ating conditions for the mixing of two (or more) solid particles of different densities will be achieved by working at the conditions which result in a recycle ratio which approaches infinity ($R > 20$) for the particles having the highest density. Since above a value of about $R = 20$ no significant change occurs in the behavior of the solid phase, all solid particles will behave similarly (even if the less dense particles will, ideally, have a higher recycle ratio) and optimal mixing will be achieved.

Mixing of Gases

The mixing properties of the reactor for gases were tested by performing pulse experiments. Air was passed through the reactor at flow rates of the same magnitude employed in the solid mixing experiments. At a certain time, a pulse of CO_2 was introduced at point *13* (Figure 1) and the system response at the exit pipe 5 was recorded. It was found that the response curve conformed with the behavior of two identical perfectly mixed reactors in series; hence it may be concluded that the reactor is a good mixer for gases. Moreover, by comparing the pressure drop in the reactor with a fluidized-bed reactor at identical flow rates and in the absence of particles, it was found that it is 4 to 5 times smaller and hence the energy consumption is less by the same factor. Practically, the fluidized bed is not operated empty and hence the energy needed for mixing two gases will be much higher. In addition it should be noted that the major shortcoming of a fluidized bed according to Perry (1973) is the poor horizontal mixing, which requires effective distributors for good mixing, causing an increase of the pressure drop in the fluidized bed.

REACTOR MODEL FOR THE GAS PHASE IN GAS-GAS MIXING

In order to gain more insight into the mixing behavior of the gas phase, a more thorough analysis of the system response to a CO_2 pulse input at point *13* in Figure 1 was done. The response was measured at the impingement zone by a probe inserted at point *14*, in the reactor exit pipe (point 5 in Figure 1), and also at point *10*. It was found that all curves were practically identical, regardless of the sampling position, and were similar to the experimental curve shown in Figure 4c. This characteristic result led to a reactor model for the gas behavior which is composed of the following zones:

1. The impingement zone (approximately, the annular space *1* between points *9* and *11* in Figure 1) which is perfectly mixed. A more accurate picture, which is in accordance with the experimental results, visualizes this zone as composed of three perfectly mixed vessels designated by *1*, *2*, and *3* in Figure 4a,b.
2. The rest of the reactor including the exit pipe designated by *4* in Figure 4a behaves like a plug flow reactor in which no backmixing takes place. In order to verify that indeed a pulse input introduced simultaneously at both inlets *A* and *B* in Figure 4a "passes" two mixed vessels in series (*2* and *3* or *1* and *3*), the prediction of the experimental response curve in Figure 4c was made as follows: From the experimental curve the mean residence time τ was calculated and found equal to 6 s. Assuming for the sake of simplicity that vessels *2* and *3* are identical, the mean residence time in each vessel is 3 s. For two identical well-mixed reactors in series, it is well-known that the response curve to a pulse input is given by:

$$C(t) = \frac{t}{\tau^2} e^{-t/\tau} \tag{6}$$

which yields the curve corresponding to a two-mixed vessels model shown in Figure 4c. As can be seen, the agreement between the experimental curve and the predicted curve according to Eq. 6 is good enough. It should be noted that the agreement could be improved by assuming unequal vessels with slightly different mean residence times τ_1 and τ_2 such as $\tau_1 + \tau_2 = 6$. On the other hand, in a single mixed vessel, namely, *1* or *2*, the response curve to a pulse input:

$$C(t) = \frac{1}{\tau} e^{-t/\tau} \tag{7}$$

will result in the dotted curve in Figure 4c for $\tau = 6$ s. This curve is completely different from the model predicted by Eq. 6 which is described in Figure 4a,b.

It is interesting to note that the above gas behavior is different from the solid particles, which exhibit an oscillatory behavior, namely, a particle entering at point *A* in Figure 4a may reach entrance *B* and vice versa. A stochastic model based on Markov processes which takes into account this behavior was developed by Luzzatto et al. (1984). Finally, it should be noted that oscillation might indeed occur in gas-solid systems due to the significant difference in densities between the phases, as a result of which inertia of the particles is maintained. On the other hand, the difference in densities between two gases to be mixed is usually negligible and thus inertia forces as well as oscillatory behavior are absent, as demonstrated by the results shown in Figure 4c.

ACKNOWLEDGMENT

This paper is dedicated to our colleague, the late Professor Isaak Elperin, whose contribution permitted the realization of this work.

NOTATION

D_1, D_2	= particle mean diameter of substances 1 and 2, m
D	= mean diameter of a mixture (Eq. 4), m
E	= specific energy (Eq. 1), J/kg
Eu	= Euler number
Eu_a, Eu_p	= Euler number of the gas, and in the presence of particle flow, respectively
G_1, G_2, G_a	= mass flow rate in kg/s of solid particles 1,2 and of the gas phase, respectively
n_1, n_2	= number of particles per g of substances 1 and 2
N	= number of samples
Q	= volumetric air flow rate, m^3/s
ΔP	= pressure drop across the mixing device, Pa

t = collection time of the quantity $W_1 + W_2$ in the continuous system (new device), or time of mixing quantity $W_1 + W_2$ in fluidized-bed or spouted-bed mixers

U_a = velocity of gas phase at inlet pipe 2, Figure 1, m/s

W_1, W_2 = weight of substances 1 and 2 collected in a continuous system (new device) during a time interval t, or initial weight to be mixed in fluidized-bed or spouted-bed mixers

X_1, X_2 = particle-fraction of substances 1 and 2, $X_i = n_i/(n_1 + n_2)^2$

$\overline{X}$ = mean particle-fraction, $(1/N) \sum_{i=1}^{N} X_i$

Greek Letters

ρ_1, ρ_2, ρ_a = density of substances 1, 2 and of the gas phase respectively, kg/m^3

ρ = density of solid particles mixture (Eq. 3), kg/m^3

ρ_b = bulk density of particles, kg/m^3

σ_{N-1} = standard deviation defined by $\sum_{i=1}^{N} (X_i - \overline{X})^2/N - 1$

τ = mean residence time, s

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Manuscript received July 7, 1983; and accepted Apr. 6, 1984.