Pressure Drop and Power Requirements in a Stirred Fluidized Bed

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The paper reports fundamental observations on the effect which stirring of fluidized solids has on the pressure drop. Typical power requirements are also given.

The experimental pilot unit was of 6-in. diameter and 12 in. high. A special blade or paddle type of stirrer had been built which permitted the evaluation of the effects of blade positioning, blade spacing relative to sense of rotation, as well as the effect of stirrer immersion and rotational speed.

The observations disclosed that blade positioning relative to sense of rotation had a very profound effect on pressure drop as well as on power requirements. When the solids circulation pattern as normally induced by the stirrer was of the same general character as that ordinarily encountered in a dense phase fluidized charge, the resulting pressure drop was always considerably lower than the conventional unstirred fluidized bed pressure drop. Furthermore for this type of stirrer the power requirements were also lowest. For other stirrer designs and other induced solids convection patterns the pressure drop was less influenced and the power requirements were much greater.

The solids used for the study were a relatively coarse silica sand for which both pressure drop as well as power requirement data were reported. Power requirement data were also obtained for a much finer alumina powder. When one emphasized the effect which particle properties, especially surface characteristics may have, the level of the power data was much lower, although the data were of the same character as the characteristic values pertaining to the sand.

It has long been known that beds of aerated solids may under certain conditions offer greatly reduced resistance to mechanical stirring. In ordinary solids beds with a normal state of compaction the individual particles are in such close proximity to each other that the interparticle friction is quite high when mechanical stirring is attempted. When the bed is however aerated, owing to expansion of the column, the particles may then become sufficiently disengaged from each other to reduce the internal friction to a sufficiently low level so that mechanical agitation becomes feasible. It is under such conditions of operation that a fluidized column may become an attractive device for solids blending.

Although stirred beds have been employed occasionally in pilot-plant studies for the purpose of improving fluidization and simulating large-scale behavior, only one single study describing the effect of stirring on pressure drop and other fluidization parameters has so far been disclosed (3). It employed a rectangular test chamber that carried a central shaft that could be fitted with either an 8-mesh screen agitator element or another similar element, comprising ½-in. diameter holes. These agitator elements, also rectangular in shape, could be coordinated at certain intervals above each other and were parallel to the perforated bottom of the fluidizing chamber. Oscillations were imparted to the shaft by way of a revolving cam, and these oscillations were transmitted to the agitator elements, which in turn imparted the oscillations to the bed.

The agitation data given in this paper are wholly different. Whereas the vibrations of Reed and Fenske were only in a vertical direction, the agitation presently described was achieved by insertion of centrally located stirrers, rotated horizontally. In essence therefore the work described in this paper appears rather related to fluid-mixing studies in upright tanks, with central impellers. Hence the present paper will report on the effect of centrally located stirrers on pressure drop through small fluidized beds, as well as power requirements demanded by certain stirrer designs and modes of operation.

EXPERIMENTAL UNIT

A description of the experimental apparatus is given in Figure 1. Major parts were a cylindrical Pyrex glass section of 6-in. I. D. and 12-in. high. This rested on a course-grade porous plate ½ in. thick. Below was joined a conical air inlet adapter with the air inlet and pressure drop connections as indicated. The porous plate extended well beyond the periphery of the glass pipe; however air leaks were prevented since the plate had been peripherally impregnated and coated by paraffin. The air rate to the unit was controlled and measured by the valve and rotameter preceding the unit.

A central shaft stirrer could be lowered to various positions into the glass pipe. There were three separate pulleys which could be communicated with the motor pulley. With the largest pulley (14-in. O.D.) 3.4 rotations/sec. resulted; with the next pulley (10-in. O.D.) this increased to 4.7, and with the smallest (6-in. O.D.) 7.8 rotations/sec. were achieved. The motor was a 1/3 hp., 1,725 rev./min., ball-bearing unit.

Details of the stirrer are shown in Figures 2*a* to 2*d*. It was principally a ½-in. diameter steel shaft with 3/16-in. holes drilled as indicated in Figure 2*a*; individual blades were as indicated in Figure 2*d*. The blades were simply mounted by passing the threaded short-blade shaft through a hole in the stirrer and fastening its position by nut and lock washer. This design and arrangement permitted considerable flexibility as the blades could be arranged at any angle in any desired position.

For the purpose of properly referring to a particular stirrer design the holes in the shaft were numbered from the base up, as shown. In a mounted position the stirrer appears as indicated in Figure 2*b*. For the purpose of defining further the characteristics of a certain design the angle between the plane of the blade and the respective horizontal plane as indicated may be used. In defining this angle is was decided to look at the stirrer with blade nearest to the eye. The four basic positions which the blades can thus occupy are indicated in Figure 2*c*. Finally in order to fully define the stirrer characteristic the sense of rotation must be known. Thus the arrangement was such that the stirrer always turned counterclockwise when looked at from above.

OPERATION

For a typical run the glass pipe was charged with a definite weight of a granular solid. Next the charge was slightly expanded and the stirrer, arranged in the desired manner, was lowered into the bed to a definite position. The bed was then subject to certain air rates, and while stirred at certain speeds, pressure-drop readings were made. The top of the bed was always open to the atmosphere; hence the static inlet pressure, as indicated by the manometer, was then also the bed pressure drop.

Prior to investigating stirred beds the apparatus was examined for its general functioning as a fluidization unit. This involved the measurement of the pressure drop across the porous plate, the pressure drop across the fixed and then fluidized bed, without containing a stirrer, and the experimental evaluation of the point of initial fluidization and its checking against a calculated value.

Pressure drops across the porous plate are given in Figure 3. Order of magnitude of the data is in line with values reported for porous plates in general, and hence the peripheral impregnation with paraffin had not influenced the flow characteristics.

For the purpose of examining the general functioning in fluidization the pipe was charged with 12.34 lb. of a silica sand. At the visual onset of fluidization the pressure drop was 12.25 in. of water column. On the basis of the weight of the bed and an internal column diameter of 6 in. a value of 12.05 in. is obtained. The small excess observed by experiment is primarily due to the resistance of the porous plate. Hence the observed and calculated values are in satisfactory agreement.

Upon plotting the column pressure drop vs. mass velocity the data were observed to flatten at an air rate of 2.0 cu. ft./min. This, when checked against a standard correlation for estimating the point of initial fluidization, was within 5% of the calculated value (*1*).

The sand used for the preceding as well as all subsequent experiments with stirrers had a density ρ_s = 165 lb./cu. ft. and the following size distribution:

Mesh	d_p (in.)	Percent by weight (X)
42	—	0.30
42-60	0.0116	27.00
60-80	0.0082	62.50
80-115	0.0058	9.25
115-170	0.0041	0.50
170	—	0.45

Its average particle diameter as calculated according to

$$D_p = \frac{1}{\sum \frac{X}{d_p}}$$

was 0.0086 in.

For all subsequent stirring tests the glass pipe was charged with 10.0 lb. of the above sand. All pressure drops were corrected for resistance through the porous plate.

A few experiments were also made with an alumina powder. Its density was ρ_s = 208 lb./cu. ft., and the size distribution is given below:

Mesh	d_p (in.)	Percent by weight, X
80	—	3.0
80-115	0.0058	22.5
115-170	0.0041	32.0
170-250	0.0029	30.5
250-325	0.0021	7.0
325	—	5.0

Power measurements for stirring were made by winding a string around a pulley, passing the loose end of the string over another pulley, and attached various weights. The stirrer and the weights were then entirely wound up and the bed subject to a range of air rates. The weights were then allowed to drop, thus revolving the stirrer in the aerated charge. By measuring the time required for the fall power requirements were calculated according to

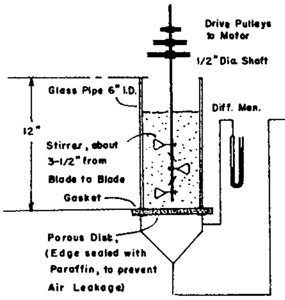

Fig. 1. Experimental unit.

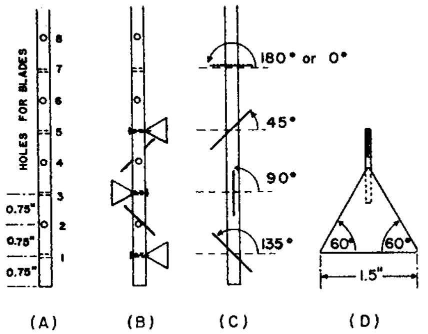

Fig. 2. Stirrer detail.

$$HP = \frac{2\pi R W n}{33{,}000}$$

where R is the radius of the pulley, W is the pulling weight, and n is the number of revolutions made by the pulley per minute.

The reproducibility of the pressure-drop data was in all instances excellent. The power requirement data reproducibility was generally quite good. Check runs indicated that with the method of operation adopted the individual data did not vary more than ±15% from each other.

RESULTS AND CONCLUSIONS

An orientation of the various runs and details pertaining to stirrer arrangements are given in Table 1.

Effect of Positioning of Blades

A comparison of Figures 4, 5, 6, and 7 emphasizes the effect which the positioning of the blades has on the pressure drop. In all instances the stirrer contained eight blades, built up in succession from the base. A sharp effect on pressure drop is noted when the blades are at an angle of 135 and 90 deg. The distinct effects which these blade angles produce in combination with counterclockwise rotation is readily understood and explained from a consideration of the natural solids circulation pattern in a dense phase fluidized charge. Thus in the nonstirred dense phase fluidized bed of the type discussed the solids will normally be lifted in the center, whereas in the outside annular ring the solids will descend. This is indicated in Figure 8. This pattern does by no means preclude any lateral mixing of solids, as in fact such movement of solids is definitely known to exist, especially in equipment of appreciable diameter. However in columns of relatively small diameter and with the air distributor extending virtually across the entire bed the solids circulation pattern may thus be simplified. Since with the above blade angle of 135 deg. and the counterclockwise rotation the tendency is to lift the solids in the center; thus the effective bed density is actually lessened in the center. To an appreciable extent thus reduction of bed density occurs at the expense of an increase in bed density in the annular ring. In fact the increase in bed density in the ring may be so considerable that the solids in the ring are no longer fluidized. This seems to be indeed supported by visual observation, as the solids in the ring were found to descend almost like a solid mass of matter toward the bottom of the column. The familiar and frequently observed oscillating motion of the particles, found usually in nonstirred beds, was entirely absent. Obviously then under such conditions a much greater than usual portion of the gas will rise in the center of the bed. In a sense this phenomena is akin to channeling, where a disproportionately large fluid quantity passes up through the center. The phenomenon is also similar to spouting (*2*), where nonhomogeneous lateral fluid distributions are also prevalent. With the blades at an angle of 90 deg. the situation is similar, but less pronounced.

As a consequence of this lesser resistance in the center of the bed the operating pressure drop is always lower than the pressure drop calculated from the weight of the bed, if the angle is 135 deg. and the sense of rotation counterclockwise, as seen from above. Moreover the pressure drop-flow curves for these types of stirred beds do no longer exhibit a discontinuity at the

Table 1. Orientation of Data

Figure number	Position of blades	Angle (deg.)
4	1·2·3·4·5·6·7·8	135
9	1·2·3·4	135
11	5·6·7·8	135
12	1·2·3·4 – – – 8	135
10	1·2·3·4·5·6·7·8·9·10·11	135
5	1·2·3·4·5·6·7·8	90
6	1·2·3·4·5·6·7·8	45
7	1·2·3·4·5·6·7·8	0

point of initial fluidization; that characterizes the unstirred bed. This would then indicate that the sudden transition from the fixed to the expanded bed does no longer occur in this type of stirred bed. Another point of interest is that the slopes of the pressure drop-flow data that pertain to this mode of stirring are all smaller than unity, an unusual result indeed. Hence it follows that this type of stirred bed exhibits a greater capacity for passing gas than does a non-stirred bed, composed of the same particles and of the same geometric dimensions. Of course the faster the rotational speed of the stirrer the greater will be the deviation of the stirred bed pressure drop from that calculated on the basis of the weight of the bed.

When the blade angle is 0 deg., that is with a horizontal blade arrangement, there is virtually no effect on the pressure drop. This is of course expected and is supported by the data shown in Figure 7.

Considering now the action of a stirrer with the blades at an angle of 45 deg., from an analogy of the action already described for a stirrer with a 135-deg. arrangement, one would expect that for the 45-deg. arrangement the pressure drop should be in excess of that calculated on the basis of the weight of the bed. While it is doubtlessly so that with this arrangement the tendency is to force the solids downward in the center and thus create a bed core of greater density than the annular ring, it is precisely owing to this greater bed density that a disproportionately greater flow may now occur in the annulus. In effect then a stirrer with blades at a 90-deg. angle may tend to divert the flow actually away from the center. However there is another complicating feature observed with this as well as with a blade angle of 90 deg., namely the appearance of a depression in the center of the bed manifests itself as the stirrer rotates. Hence whereas the bed may have become more dense in the center when subjected to the action of a stirrer with blades at 45 deg., due to the lesser height in the center, owing to the resulting central depression the final resulting flow resistance may nevertheless remain essentially unaltered. This is virtually indicated by the data of Figure 6, where the pressure drop is seen to be only little influenced by this type of stirrer.

Effect of Stirrer Height

As expected the height of the stirrer in relation to the bed height will be an important factor. This is apparent from a consideration of Figures 9, 4, and 10, where data for stirrers composed of four, eight, and eleven blades are given.

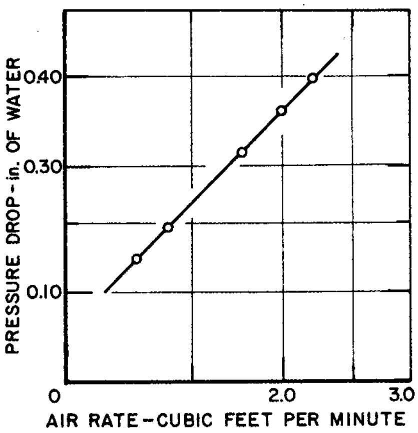

Fig. 3. Pressure drop through porous plate.

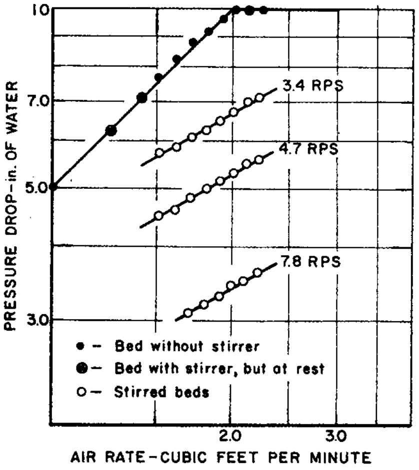

Fig. 4. Stirrer—eight blades at 135-deg. angle.

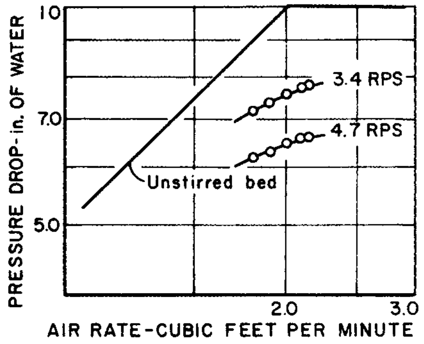

Fig 5. Stirrer—eight blades at 90-deg. angle.

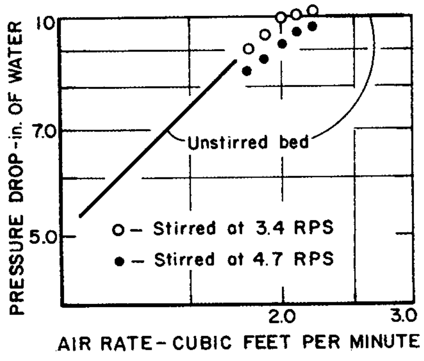

Fig. 6. Stirrer-eight blades at 45-deg. angle.

In all cases the static bed height was 7 in. Compared with this, stirrer heights were as follows:

4 blades	3.50 in.
8 blades	6.50 in.
11 blades	8.75 in.

Thus the four-blade stirrer was essentially only active in the base of the column, whereas the top of the bed remained relatively unaffected. This is indicated by the considerably lessened effect which is observed on the pressure drop. With the eight-blade stirrer the effects became much more pronounced, especially at high speeds. With eleven-blades the effects became even greater. Whereas the eleven-blade stirrer was higher than the unexpended bed, the bed height increased under the combined influence of gas flow and stirring sufficiently to cover all the blades. With a speed of 7.8 rotations/sec. the eleven-blade stirrer expanded the bed considerably and caused a severe depression in solids level in the center due to centrifugal forces imparted to the particles. This phenomenon was probably responsible for the very great pressure-drop decrease observed in this instance. Rate of solids circulation, as judged by visual observations, was extremely rapid when the eleven-blade stirrer was operated at elevated speeds. Solids losses occurred at the highest speed, owing to splashing.

Effect of Stirrer Position

Since gas-fluidized solids beds are known to be nonhomogeneous in density and gas-solids dispersion it is to be expected that the position of a stirrer in a charge should be a significant factor. This is supported by the data of Figures 9, 11, and 12. In Figures 9 and 11 a four-blade unit was active in the bottom and the top of the charge, respectively, whereas in Figure 12 the data pertain to a stirrer that carried four blades in the bottom with one single blade near the top of the bed. Considering first the two four-blade units the greater effect is observed when the stirrer is in the bottom. This may be explained readily since it is to be anticipated that the main contribution toward fluidized bed pressure drop originates from the bottom portion of the bed, where the gas-solids dispersion is better than in the top. Therefore the solids-lifting action of a short stirrer will be felt more significantly when placed into the lower half of the bed.

The curves exhibited by the data of Figure 11 are of some interest. They appear somewhat different from the remainder of the stirred-bed data and there is, especially at the high speed, a definite flattening of the pressure drop beyond the point of initial fluidization. This trend, constituting an approach to

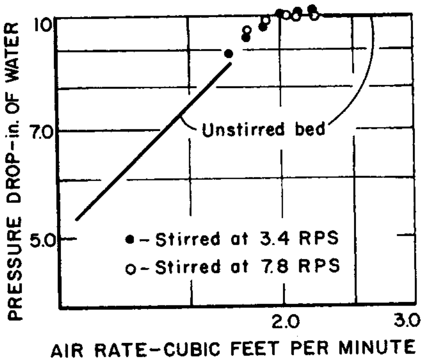

Fig. 7. Stirrer—eight blades at 0-deg. angle.

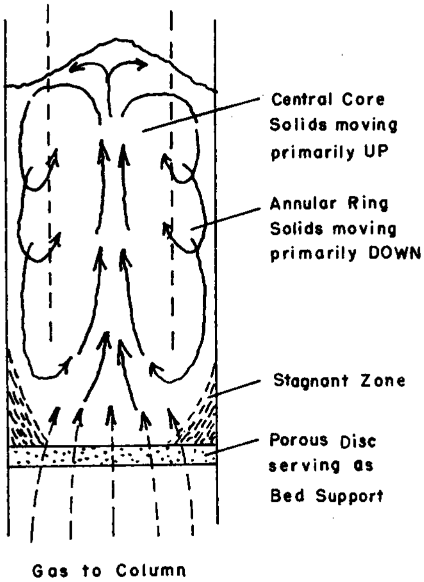

Fig 8. Greatly simplified solids flow pattern in gas-fluidized small diameter bed

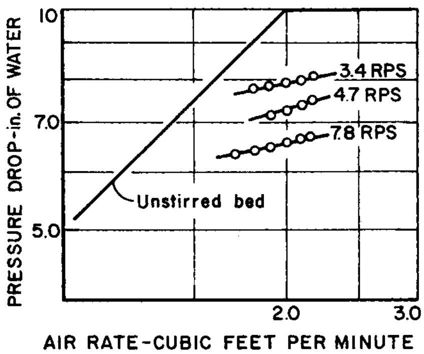

Fig. 9. Stirrer—four blades in bottom of bed at 135-deg. angle.

the normal pressure drop-flow data of an unstirred bed, is readily understood if it is assumed that normal unstirred fluidization is probably still proceeding in the lower half of the bed. The curves of Figure 11 are therefore the result of a superimposition of the lower unstirred portion of the bed and the upper stirred section.

The addition of the single blade in position 8, as indicated by the data of Figure 12, will of course depress the pressure drop further. More important however was the general behavior of the column when acted on by this type of stirrer. Whereas with the stirrer as referred to in Figure 4 the expanded column was relatively steady in height, wide fluctuations were now noted with the stirrer of Figure 12. These fluctuations were however not reflected in the steadiness of the pressure drop. The solids circulation pattern was extremely rapid.

Effect of Stirrer Speed

It has already been noted from the preceding data presentation that the speed of the stirrer has a significant effect on the magnitude of the pressure drop. In order to show this more systematically Figure 13 has been composed, showing the relative pressure drop in relation to stirrer speed. The relative pressure drop is defined as the ratio of the stirred-bed pressure drop at the minimum fluidization mass velocity to the pressure drop calculated from the weight of the bed.

Examination indicates that all the data when extrapolated to zero stirrer speed coincide and terminate at a relative pressure drop of unity. This is of course not unexpected, since for the unstirred bed the pressure drop should be that demanded by the weight of the bed. The good agreement of the data in this respect emphasize that the accuracy with this simple equipment was quite satisfactory.

The effect of the number of blades composing the stirrer is again readily apparent. Thus comparing curves 2, 1, and 5 with each other it is learned that the effect of stirring speed on pressure drop becomes more pronounced as the number of blades in the stirrer increases. The very rapid decrease experienced with the stirrer carrying eleven blades is of course to a part again due to the centrifugal forces imparted to the particles in the bed and hence to the induced central depression in the top of the bed, resulting from this mode of stirring.

Comparison of stirrers pertaining to curves 1 and 6 is of some interest. Both stirrers contained eight blades, but with curve 1 these were arranged at an angle of 135 deg. whereas for the other stirrer the angle was 90 deg. Examination of the course of the curves indicates that the relative pressure drop is less when the angle is 135 deg. This is as expected, since with this angle a substantial solids lifting action in the center of the bed is accomplished. The relative pressure-drop decrease accomplished with the stirrer blades at 90 deg. stems primarily from the application of centrifugal acceleration to the particles and a resulting depression of the top surface of the bed. This becomes of course more pronounced as the stirrer speed is increased, and for this reason the rate of decrease of relative pressure drop with the increasing stirrer speed is found to be more pronounced here than when the blades are arranged at an angle of 135 deg.

Power Requirements

With the various stirrers having as pronounced an effect on pressure drop as learned it should not be a surprise to discover that the power requirements

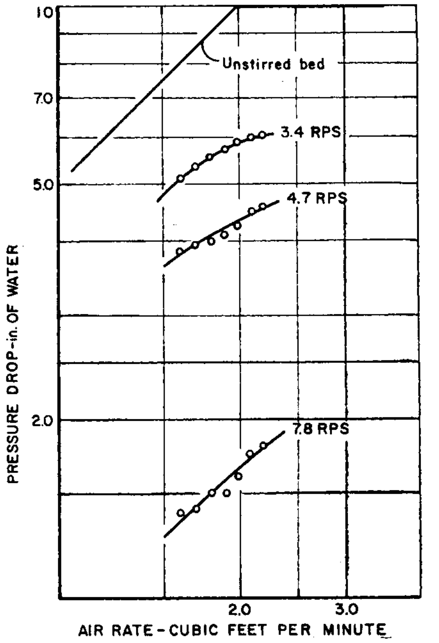

Fig. 10. Stirrer—eleven blades at 135-deg. angle.

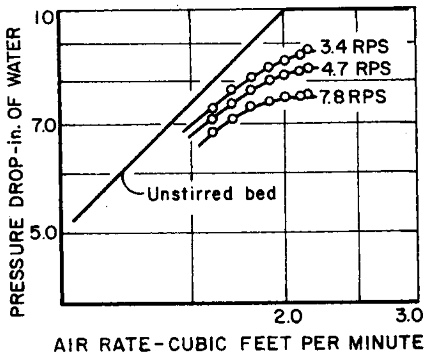

Fig. 11. Stirrer—four blades in top section of bed at 135-deg. angle.

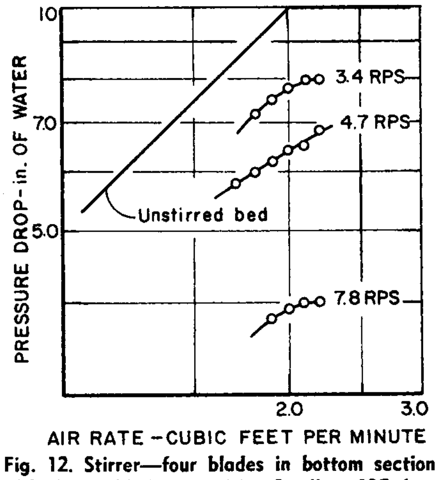

Fig. 12. Stirrer—four blades in bottom section of bed, one blade in position 8, all at 135-deg. angle.

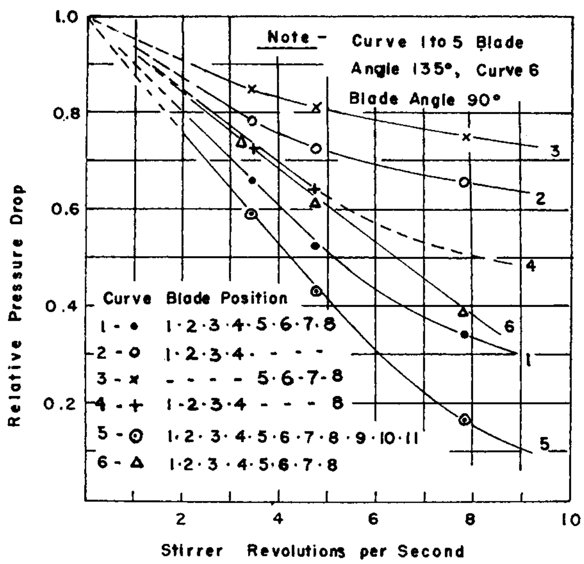

Fig. 13. Effect of stirrer speed on pressure drop at point of incipient fluidization.

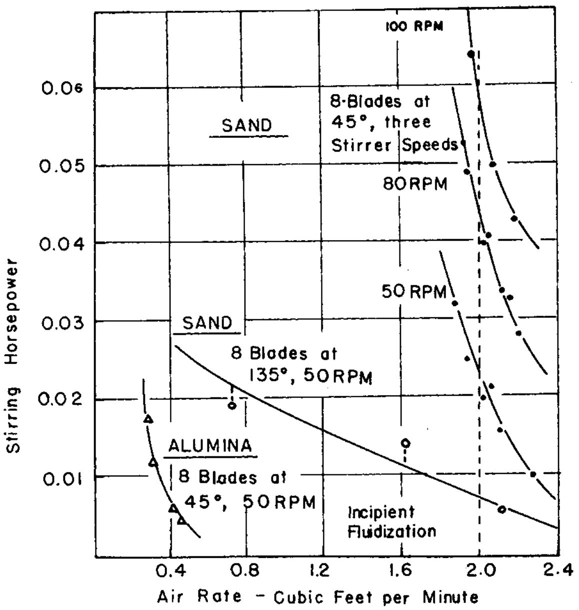

Fig. 14. Power requirements for various stirrers and rotational speeds.

for stirring an aerated solids bed will also depend markedly on the design of the stirrer and the sense of operation. This is already shown by the data of Figure 14, where some typical horsepower values are given for two designs and a range of stirrer speeds, in relation to aeration rate.

A general observation valid for all instances is of course that the required stirring horsepower increases as the aeration rate decreases. Furthermore the rather surprising result is observed that in the horsepower-aeration curves no discontinuity is observed in the region of the normal onset of fluidization of the unstirred bed. This indicates clearly that in the stirred bed the original point of initial fluidization has lost its identity. The lack of discontinuity of the power data seems to be strongly related to the lack of discontinuity of the pressure drop-flow data for stirred beds, when the region of incipient fluidization of the unstirred bed is considered.

For 50 rev./min. two sets of data are available. The data pertaining to the stirrer with blades at 45 deg. are considerably higher than the data for the stirrer with blades at 135 deg. Moreover this is true for the entire aeration rate indicated; however, it is principally true for such air rates which would not normally be sufficient to fluidize an unstirred bed. At air rates substantially in excess of those required for minimum fluidization the data clearly approach each other. This is an interesting observation which may perhaps permit some insight into the nature of the power expenditure associated with stirred beds. Since the two beds are wholly different as far as solids flow currents are concerned, it may at this stage be inferred that the fact that the stirring horsepower requirements approach each other at elevated gas flow points to the possibility that interparticle friction forces must now be about equal in both beds. Plausible as this sounds it cannot be accepted at this time without giving simultaneous consideration to the state of bed expansion or, in other words, the average bed voidages for the two beds. It must be born in mind that the magnitude of the frictional forces do somehow depend on this quantity. Since in the present measurements bed heights and hence average voidages were unfortunately not recorded, the reason for the approach of 50 rev./min. stirrer data cannot be resolved.

Stirrer speed is of course a major variable as indicated by the data pertaining to the stirrer with the 45-deg. blade angle. Similarly solids properties are an important factor in the horsepower requirement. The alumina particles, though of considerably higher material density, required much less power for stirring, although the general trend of the curve is strictly analogous to the respective sand data. Furthermore the particle size was considerably smaller. With the total particle surface area much larger in the alumina bed than in the sand bed it is indeed surprising to find the stirring-power requirements so much lower for the alumina. However it was noted that the alumina particles were of more regular shape than the sand particles. Furthermore the alumina surface characteristics were such as to induce flow of the solids more freely. From these observations it may be concluded that surface condition of the bed particles are a most important factor in determining stirring horsepower requirements. From this it must follow that interparticle frictional forces play an important role in determining agitation power requirements of aerated solids beds.

The average level of the power data shown in Figure 14 ranges from approximately 0.01 to 0.07 horsepower. If the data are extrapolated to a rotational speed of 3.4 rev./sec. a power requirement of about 0.15 horsepower would result. The power requirement for fluidizing the sand bed initially is readily calculated from the 10-in. pressure drop and the fact that about 2.0 cu. ft./min. of air are required. The calculation indicates that only about 0.003 horsepower is needed for normal fluidization, very considerably less than is indicated by the data of Figure 14. With a normal fluidizing bed at the state of minimum fluidization the column of solids is just slightly expanded and the particles are virtually almost without motion. With the mechanical stirrers applied to the bed the particle movement was many times faster, giving rise to much greater frictional forces and kinetic effects and hence calling for much larger power requirements.

ACKNOWLEDGMENT

The author wishes to thank the Colonial Iron Works Company in Cleveland for supporting this research and their permission to publish these findings.

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Manuscript received March 3, 1958; revision received March 15, 1960; paper accepted March 17, 1960. Paper presented at A.I.Ch.E. Chicago meeting.