# ASSESSMENT OF PARTICLE SUSPENSION CONDITIONS IN STIRRED VESSELS BY MEANS OF PRESSURE GAUGE TECHNIQUE

Giorgio Micale[*], Franco Grisafi[**] and Alberto Brucato[**]
[*]Department of Chemical Engineering, University College London, Torrington Place, WC1E 7JE London, United Kingdom
[**]Dipartimento di Ingegneria Chimica dei Processi e dei Materiali, Università di Palermo, Viale delle Scienze, 90128 Palermo, Italy

In this work the quantitative assessment of the mass of suspended solid particles in stirred vessels is performed using the Pressure Gauge Technique. This is based on the measurements of the pressure increase on the tank bottom due to the presence of suspended solid particles at any agitation speed. The method has the advantages of not utilising visual observations and of easy and inexpensive application to both laboratory and industrial equipment. Very few data are available in literature and the experimental results collected using the present PGT technique and the correlations here proposed are of considerable academic and industrial interest.

Keywords: particles suspension, stirred vessels, Pressure Gauge Technique

## INTRODUCTION

The suspension of solid particles in stirred tanks is involved in many industrial operations such as multiphase catalytic reactions, adsorption, crystallisation, dissolution, leaching etc. Due to its industrial importance the subject has been extensively studied in the past but, in spite of all efforts, the results are far from being conclusive and they can even be misleading to the practitioner (Chudaceck, 1986). As a matter of fact, solid-liquid operations have been claimed to be the source of major economic losses, being distinctly more troublesome than gas-liquid operations, during the design and start-up stages of new plant implementations (Smith, 1990).

Most of the research efforts have been concentrated on the determination of the minimum impeller speed "$N_{js}$" required to "just suspend" all the particles on the basis that this condition corresponds to the lowest specific power input at which all particles surface area is made available for effective processing. Since its proposal (Zwietering, 1958), the well-known visual *one-second criterion* for the $N_{js}$ determination has been widely used to ascertain the achievement of complete suspension conditions. As for all visual methods, it is necessary to build transparent vessels in order to observe the bottom of the tank: this can be easily done at laboratory scale but is impracticable for large scale installations. The main advantage of this method is its intrinsic simplicity; but it must be pointed out that it is affected by subjective evaluations, so that only with a careful and skilled observation it is possible to get about ±5% reproducibility at best in the case of dilute suspensions, while no data can be reliably reproduced when solids concentration is above 8% w/w (Oldshue and Sharma, 1993). Other methods either visual (Rieger and Ditl, 1994) or instrumental (Rao et al, 1988; Buurman et al, 1986) have been proposed, that provide a more or less accurate determination of values for $N_{js}$.

None of the above mentioned methods allow quantitative information on partial suspension conditions to be obtained. As a matter of fact, these conditions have hardly been investigated in the past and very few data are available in the scientific literature. This is rather surprising if one considers that many industrial installations are usually operated under such conditions (Oldshue, 1983). Information on partial suspension conditions in conventional mixing tanks can be found in a work by Bourne and Sharma (1974). These authors, by means of a withdrawal sampling technique, assessed the existence, at low agitation speeds, of a "trapped" mass of particles that gradually decreases while increasing impeller speed. It should be pointed-out however that it is not clear how these data should be interpreted as they were obtained in a system in which the total amount of solid phase varied extensively during each run. Some more information on partial suspension conditions is given by Chudacek (1982) who employed a visual inspection based estimate of the unsuspended solids mass. The only author that devised instrumental ways to quantitatively assess the suspension degree at various impeller speeds is Biddulph (1990), though its work only deals with an unconventional geometry (a large rectangular tank stirred by a submersible mixer). Two measuring methods were used: (i) a scale plate, completely surrounded by the liquid phase, whose top surface carries the weight of unsuspended particles, whose mass could therefore be directly assessed from the scale readings, and (ii) a pressure transducer in the proximity of the bottom, from whose readings the suspended mass of solids could be deduced. The two methods gave rise to consistent results, so that each one validated the other.

Aim of the present work is the quantitative assessment of fractional suspension in stirred tanks using a suitable technique based on pressure measurements (i.e. the Pressure Gauge Technique, recently proposed by some of the authors, Brucato et al., 1997), following the work by Biddulph (1990) in order to make it applicable to conventional stirred tanks. The resulting method was found to be simple and reliable for the quantitative evaluation of the fraction of suspended particles at any agitation speed.

## EXPERIMENTAL

The experimental set-up used for the present work is shown in Fig.1. A flat bottomed, fully baffled, transparent vessel (T=0.19m, Volume=5lt) was stirred either by a six blade Rushton turbine (D=T/2) or by a 45 pitched blade turbine (D=T/2) set at a distance from the vessel bottom equal to T/3. A measured mass of solid particles was formerly added to the vessel and then the liquid phase was added up to an height H=T=0.19m. The impeller shaft was driven by a DC electric motor provided with a speed control loop, while an optical tachometer was used to independently measure the impeller speed.

A pressure gauge connected to a point of the vessel bottom allowed pressure readings to be taken. The point selected for this purpose was located on a diametral line at 45° between subsequent baffles, at a radial location midway between the axis and the side wall. A hole in the vessel bottom transmitted the pressure at this point to a dead volume to which the pressure gauge (a simple inclined manometer) was actually connected.

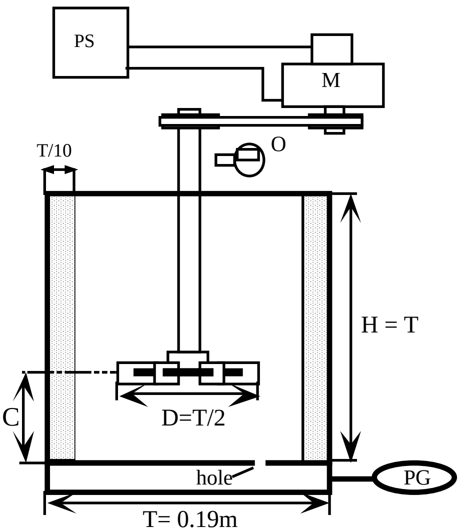


**Fig.1**-Experimental set-up

Two types of solid particles were used, glass ballotini and silica particles, both with a measured density $\rho_s$ of 2500kg/m$^3$. Five particle size classes of sand (180-212, 350-425, 425-500, 500-600, 850-1000 µm) and two of glass ballotini (212-250, 500-600 µm), were obtained by sieving. The concentration of solid particles was varied from 3.77 to 33.8 per cent by weight. The liquid phase was deionised water in all the experimental runs and ethylene glycol was also

employed in a few runs to assess viscosity effects.
A geometrically similar, larger vessel (T=0.51m, Volume=100 lt) was also employed in order to investigate the scale-up effects. In this case the pressure reading was taken by means of a small glass tubing ended by a cylindrical frit glass that was placed on the front corner of one of the baffles.

## THE PRESSURE GAUGE METHOD

The technique is based on the measurement of the pressure increase on the tank bottom due to the suspension solid particles. In order to understand how pressure measurements can give information on the amount of suspended solids, one may consider that when solid particles lie on the vessel bottom their apparent weight is borne by the bottom itself by direct mechanical interaction with the particles. On the other hand, when particles are lifted into suspension, their apparent weight is borne by the liquid, which eventually discharges it on the vessel bottom as a pressure increase with respect to the case where no particles are suspended. Another way of looking at the same phenomenon, is by considering that when particles are suspended, the apparent density of the agitated fluid phase increases, resulting in a greater hydrostatic pressure gradient and hence again in a greater hydrostatic head on the tank bottom.

In the following, the relationship between the pressure increase and the mass of suspended particles will be derived by simply performing an macroscopic force balance on the system, which is defined as the volume occupied by the liquid and particles introduced in the vessel. With reference to Fig.2, where a stirred tank containing a mass of solids $M_t$ and a mass of liquid $M_l$ is shown, a force balance in the vertical direction performed on the system leads to:

$$F_{tot} = (M_l + M_t) \cdot g \qquad (1)$$

where $F_{tot}$ is the total force acting on the vessel bottom.

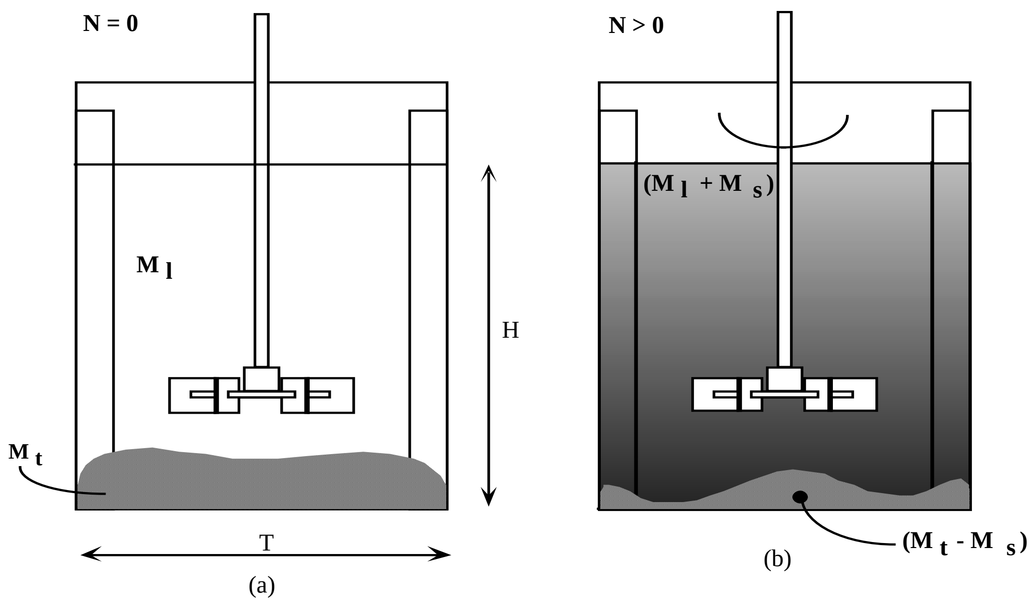


**Fig.2** – Particle distribution at (a) N=0 and (b) N>0

This total force may be split into two contributions. The first one is due to the direct contact of solid particles with the vessel bottom. At no agitation conditions (N=0, Fig.2a), this is equal to the *apparent* weight of the total mass of solids introduced in the vessel, as in such conditions all particles lay on the vessel bottom. In formula:

$$F_{direct\ contact} = M_t \cdot \left(1 - \frac{\rho_l}{\rho_s}\right) \cdot g \qquad \text{(for N=0)} \qquad (2)$$

In still conditions (N=0), the second contribution is due to the pressure exerted by an height of liquid column equal to the total height H reached by the amounts of liquid and solid phases added into the vessel. Its expression can be obtained by simply subtracting the $F_{directcontact}$ contribution from the total force $F_{tot}$.

$$F_{pressure} = \left(M_l + M_t \cdot \frac{\rho_l}{\rho_s}\right) \cdot g \qquad \text{(for N=0)} \qquad (3)$$

With reference to Fig.2b, when the agitator speed is greater than zero the total force acting on the vessel bottom does not change, but an amount of solid particles $M_s$ will be suspended: therefore the contribution due to particle direct contact with the vessel bottom will be correspondingly reduced:

$$F_{direct\ contact} = (M_t - M_s) \cdot \left(1 - \frac{\rho_l}{\rho_s}\right) \cdot g \qquad \text{(for N>0)} \qquad (4)$$

and the contribution due to pressure becomes:

$$F_{pressure} = F_{tot} - F_{directcontact} = (M_l + M_t) \cdot g - (M_t - M_s) \cdot \left(1 - \frac{\rho_l}{\rho_s}\right) \cdot g \qquad \text{(for N>0)} \qquad (5)$$

Hence the increase of $F_{pressure}$ at any agitator speed (N>0) with respect to the still stirrer condition (N=0), can be related to the mass of suspended solids:

$$\Delta F_{pressure} = F_{pressure\ (N>0)} - F_{pressure\ (N=0)} = M_s \cdot \left(1 - \frac{\rho_l}{\rho_s}\right) \cdot g \qquad (6)$$

and the related average ΔP on the whole vessel bottom is simply given by:

$$\Delta P = P_{(N>0)} - P_{(N=0)} = \frac{M_s \cdot \left(1 - \frac{\rho_l}{\rho_s}\right) \cdot g}{A_b} \qquad (7)$$

Eqn.7 shows that simple pressure measurements on the vessel bottom allow to assess the mass of suspended solid particles at any agitation speed.

It is important to point out here that the above considerations hold true for the *average* pressure on the vessel bottom. If the pressure gauge is placed at any given point of the bottom, a local pressure is actually read. As fluid motion in the vessel gives rise to pressure gradients on the tank bottom, the local pressure increase actually measured at any single point of tank bottom includes other contributions due to the fluid motion (*dynamic head effects*).

Also, there are other vertical forces that arise in agitation conditions and that have been neglected so far. For instance, a significant axial force may be exchanged between the impeller and the fluid, especially in the case of axial impellers. Other vertical forces that are not transmitted to the tank bottom through a pressure increase are the vertical friction forces on the vessel lateral wall and baffles. All these forces only arise in agitation conditions and contribute to average bottom pressure variations. Their contribution can however be included in the above mentioned dynamic head effects, that clearly have to be compensated somehow before applying eqn.7.

As a matter of fact, if the observed pressure increase with respect to no agitation conditions was solely due to the particle suspension phenomenon, a graph of the observed $\Delta P$ *vs* N should settle on an horizontal line for values of N greater than $N_{js}$. In Fig.3 a typical experimental data set has been reported (empty circles) and it can be observed that the measured value of $\Delta P$ keeps increasing indefinitely. It is therefore necessary to account for the dynamic head effects in order to extract the pressure increase solely due to particle suspension from the observed total pressure increase.

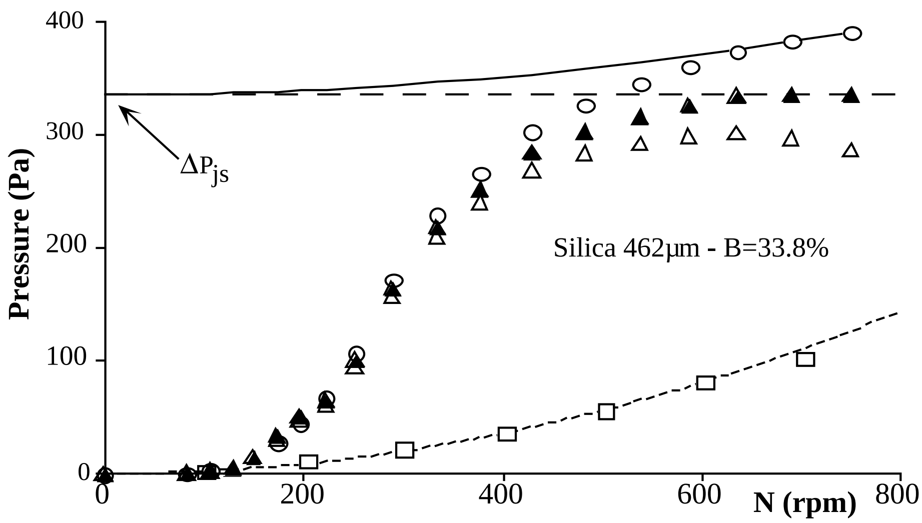


**Fig.3 –** Dynamic Head Effects

On the basis of dimensional analysis considerations it can be predicted that, for fully turbulent conditions (i.e. for Re>10000), the dimensionless reduced-pressure field is invariant with respect to the agitation speed. As pressure is made dimensionless by dividing its value by the "kinetic term", which contains the square of the agitation speed, it can be predicted that at any point inside the vessel, in the absence of solid phase, the deviations with respect to the hydrostatic pressure distribution should be proportional to the square of agitation speed. As a matter of fact, the experimental readings obtained with the apparatus shown in Fig.1 in absence of solid particles, do actually lie on a centred parabola as shown in Fig.3 (empty squares). In the same figure it can be observed that at the highest agitation speeds the experimental data points begin to deviate from the simple dependence on the square of agitation speed. This is clearly a consequence of the air entrapping phenomenon becoming important and significantly affecting the flow field. Such deviations however became significant only at fairly high agitation speeds (N>700 rpm) where the suspension phenomenon is already complete for most of the experimental data of this work. Moreover it should be observed that when solid particles are present, their damping effect on turbulence results in a further delayed air entrapping phenomenon, as visually observed during the experimentation.

In order to estimate the pressure increase solely due to solids suspension an attempt was made by simply subtracting the single-phase dynamic pressure effects (dashed line in Fig.3) from the total pressure increase (empty circles in Fig.3) but the resulting data points (empty triangles in Fig.3) tended to inexplicably fall down at the highest speeds. Moreover, the pressure increase expected at complete suspension $\Delta P_{js}$, computed substituting the total mass of solid particles introduced in the system $M_t$ for $M_s$ into eqn.7, was never attained. These considerations led to the conclusion that this way of accounting for the dynamic effects is not satisfactory. The reason for that lies probably in the fact that the flow field at a given agitation speed is significantly affected by the presence of solid particles in suspension and therefore the relevant dynamic effects are different from those measured when no solid particles are present.

On the basis of these considerations, a different data treatment was devised, that had to account for the fact that the evolution of the suspension phenomenon inhibits the collection of meaningful dynamic effect data for the two-phase systems. It was considered that above complete suspension the fluid-dynamic properties of the two-phase mixture become independent of agitation speed, as solids concentration does not change any more. Experimental data points at sufficiently high speed should therefore lie on a parabola that intercepts the ordinate axis at $\Delta P_{js}$ which corresponds to the hydrostatic head of the fully suspended two-phase fluid. This was actually found to be the case, as shown in Fig.3. The dynamic effects for fully suspended conditions are therefore given by the difference between the fitted parabola (solid line) and $\Delta P_{js}$. By subtracting from the raw data these dynamic effects, a smooth "S" shaped curve was obtained (solid triangles) which represents the $\Delta P_s$ net values (i.e. proportional to the suspended mass of solids only) looked for.

At lower speeds, where only part of the solid phase is suspended, the actual value of the dynamic effect should lie between the one obtained in this way and the one obtained when no solids are present (dashed line in Fig.3). It can be observed however that the difference between the two is not great, and that it becomes smaller and smaller while decreasing agitation speed. Moreover there is a greater interest in the precision of data at almost complete suspension conditions rather than at smaller suspension levels. On the basis of these considerations the above outlined way to obtain an estimate of $\Delta P_{js}$ was adopted. The lack of precision at the lower suspension levels introduced in this way is believed to be small, and in any case there are other effects that contribute to the uncertainty of data in this region, as for instance the shaping of the bottom due to the presence of the fillets, which affect the flow patterns and may therefore change the dynamic effects. It is worth noting that some uncertainty may also affect the data at almost complete suspension, as in principle the variation with agitation speed of solids concentration distribution might give rise to deviations of dynamic pressure effects from the square dependence on N.

It is finally worth noting that by dividing the $\Delta P_s$ values obtained in this way by the $\Delta P_{js}$ value in Fig.3, the mass fraction of suspended particles out of the total particle mass introduced in the system, hereafter referred to as the "fractional suspension", is obtained.

## RESULTS AND DISCUSSION

Fig.4 shows typical experimental results as fractional suspension X versus impeller speed. As can be seen, the "S" shaped curves change when changing particle size or concentration. In order to fit each data set, the commonest "S" shaped mathematical functions were tried, and it was found that a Weibull function with an exponent of 2 was particularly successful. For the present case the selected function can be written as:

$$X = \begin{cases} 0 & \text{for } N < N_{min} \\ 1 - \exp\left[-\left(\frac{N - N_{min}}{N_{span}}\right)^2\right] & \text{for } N \geq N_{min} \end{cases} \qquad (8)$$

where N is the impeller speed while $N_{min}$ and $N_{span}$ are parameters which control the shape of the "S".

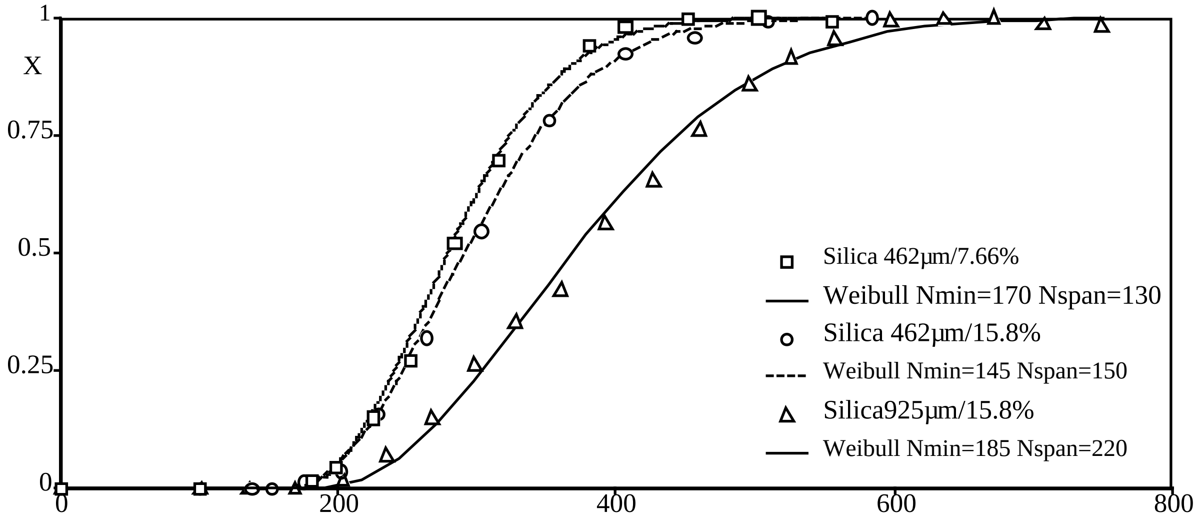


**Fig.4** – Fractional suspension data

In Fig.4 the Weibull functions (eqn.8) best fitted to the experimental data sets are reported as solid lines and the good matching with the experimental data can be observed.

One of the advantages of using the Weibull function in this case is that the value of the two parameters has an immediate physical meaning, as $N_{min}$ is the value at which the suspension phenomenon starts while $N_{span}$ is such that twice its value gives the range of N in which most of the suspension takes place, as can be immediately recognised observing that at $(N_{ss}-N_{min})=2N_{span}$ the value of X is 0.982, so that at this agitator speed less than 2% of the particles is still lying on the bottom. It is worth observing that these few particles are likely to be subject to random motion from time to time so that the consequent renewal of the liquid phase in their proximity allows them to participate in the exchange processes occurring in the vessel. Even if this was not the case, the lack of their participation to the processes could hardly be considered significant. These considerations suggest the definition of a "sufficient suspension" condition as the state attained at the agitation speed $N_{ss}$ given by

$$N_{ss} = N_{min} + 2\,N_{span} \qquad (9)$$

It is clear that the sufficient suspension speed $N_{ss}$ here defined is somewhat similar to the $N_{js}$ defined by Zwietering.

Best fitting the Weibull function (eqn.8) to the experimental data obtained resulted into a couple of values for the two parameters $N_{min}$ and $N_{span}$. for each data set. The results obtained with silica particles are reported in Fig.5 a-c. It can be observed there that both $N_{min}$ and $N_{span}$ increase when increasing the particle size (Fig.5 a and b). As regards the dependence of $N_{span}$ on total solids concentration (not shown) it increases when increasing B while $N_{min}$ decreases, as shown in Fig.5c. This last behaviour is not surprising if one considers that the higher the total solids concentration the higher the thickness of the layer of solids resting on the bottom at lowest agitation speeds, where particle suspension is still to be started-up; the correspondingly shorter distance between the fillets and the stirrer may well result into an anticipated start-up of the suspension phenomenon.

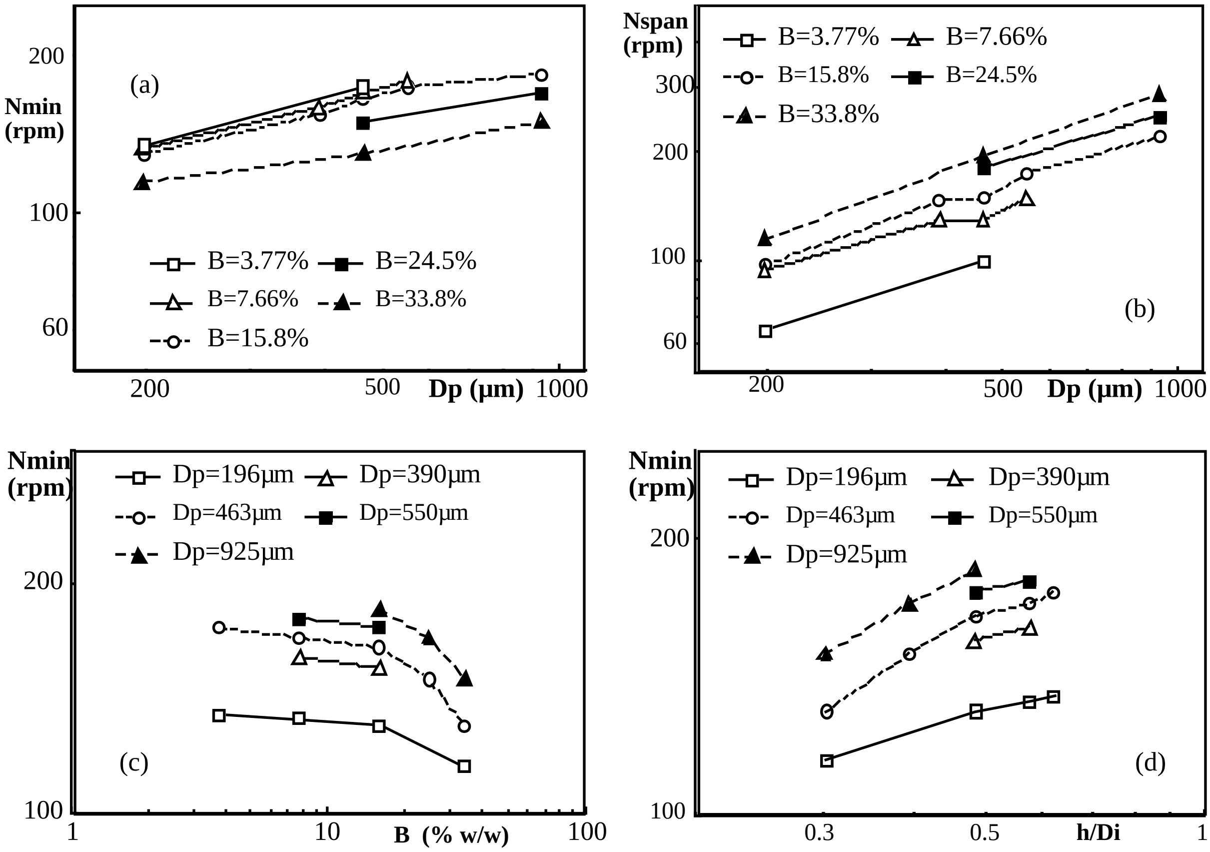


**Fig.5** – Plots of $N_{min}$ *vs* $D_p$ (a), $N_{span}$ *vs* $D_p$ (b), $N_{min}$ *vs* B (c), $N_{min}$ *vs* $h/D_i$ (d).

A thorough inspection of Fig.5 a,b shows that the available data line-up reasonably well onto straight and parallel lines, as is also the case of the plot of $N_{span}$ *vs* B (not shown). This feature indicates that simple power-law equations may successfully correlate the data. On the contrary, in the case of the $N_{min}$ *vs* B graph (Fig.5c) an increasingly steeper dependence at the highest concentrations is found. As this is likely to depend on the above mentioned shortening of the distance between the impeller and the particle fillets, an attempt was made to correlate $N_{min}$ to the ratio h/D, where h is the average distance between the stirrer and the top surface of the fillets when no particles are suspended. The average distance h was computed as:

$$h = C - \frac{4M_{tot}}{\rho_s(1-\varepsilon_s)\pi T^2} \quad (10)$$

The result is shown in Fig.5d where it can be observed that a better linearity is obtained. Therefore a power law correlation with h/D appear to be viable, although some scatter can be observed in Fig.5d. In this respect it should not be forgotten that, at the present stage of development of the experimental technique, the determination of $N_{min}$ is affected by some uncertainty, as discussed in the experimental section. Therefore the matter was not further addressed in this work and a power-law correlation with h/D was adopted for $N_{min}$. A least-squares regression analysis of the whole set of data led to the following dependences:

$$N_{min} = 49.1\ D_p^{0.236}\ (h/D)^{0.377} \qquad (11)$$

$$N_{span} = 3.10\ D_p^{0.522}\ B^{0.265} \qquad (12)$$

The $N_{min}$ and $N_{span}$ values computed by means of eqns.11 and 12 are compared with the experimental data in Figs 6a and 6b respectively, and a good fitting can be observed.

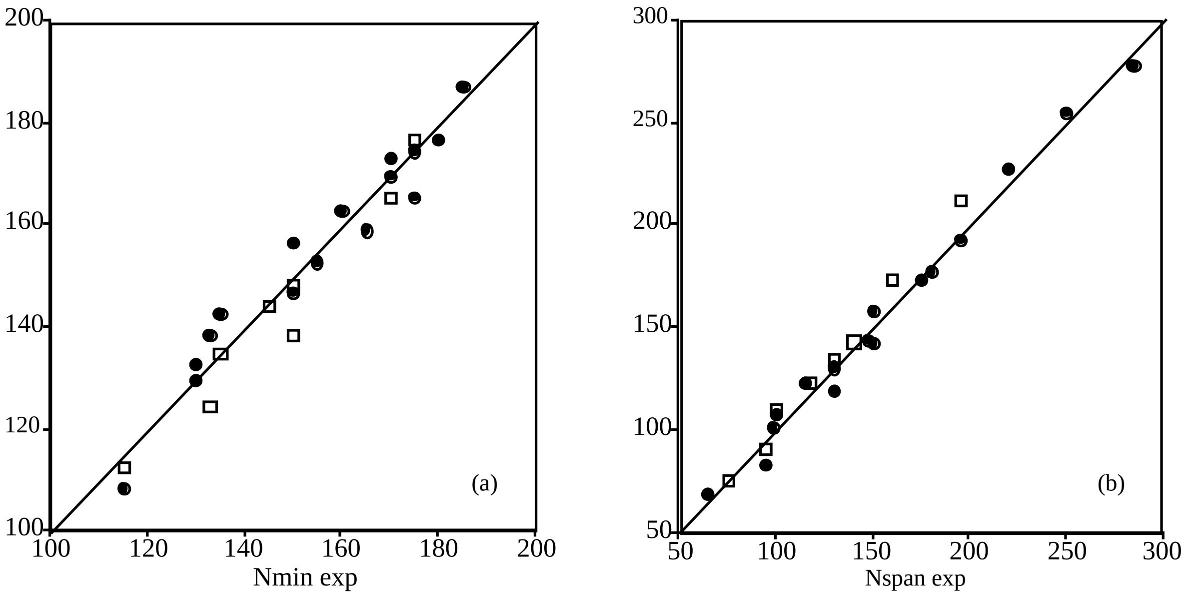


**Fig.6** – Plots of computed $N_{min}$ *vs* exp $N_{min}$ (a); computed $N_{span}$ *vs* exp $N_{span}$ (b). Solid circles indicate silica particles, empty squares glass ballotini.

Eqns. 11 and 12 deserve a few considerations. It can be noted that the dependence on $D_p$ is much more pronounced for $N_{span}$ than for $N_{min}$, a fact clearly observable in Fig.6a,b. Obviously, if the interest is only on complete (or almost-complete) suspension conditions, eqns 11 and 12 may be combined, by means of eqn.9, into a correlation for $N_{ss}$. The resulting correlation is binomial and cannot be reduced to a monomial form, because of both the difference on the exponents for $D_p$ and the diverse correlating parameter for the total particle concentration. This consideration would suggest that also $N_{js}$ data might be better interpreted by means of binomial correlations instead of the monomial correlations usually adopted. Interestingly, this might be one of the reasons for the lack of agreement of these last, especially in the scale-up to industrial sizes (Harnby et al, 1992). If this had been the case, then attempting a monomial correlation for the experimental $N_{ss}$ data should result in a significantly scattered graph. A least squares regression of the experimental $N_{ss}$ data leads to the following correlation:

$$N_{ss} = 24.1\ D_p^{0.428}\ B^{0.13} \qquad (13)$$

which is compared with experimental data in Fig.7. The remarkable agreement obtained is practically identical to that observable with the binomial correlation (not shown here for the sake of space limitations). Therefore it has to be concluded that at the present stage there is no firm indication that a binomial equation should be adopted for the correlation of complete suspension data. Obviously in this situation the simpler monomial form is to be preferred. On the other hand, two separate correlations like eqns 11 and 12 are to be obtained if the interest is on the whole suspension phenomenon rather than on complete suspension conditions only. It is worth noting that the limited data scatter, which is as good as that obtained with the binomial correlation, is smaller than that observable in Figs.6a and 6b; this probably depends on the fact that, due to the particular dynamic effects compensation adopted, data on almost complete suspension conditions are affected by a smaller uncertainty than those on $N_{min}$. In these conditions, an overestimation of this last parameter results in an underestimation of $N_{span}$, but their combination into $N_{ss}$ is practically unaffected.

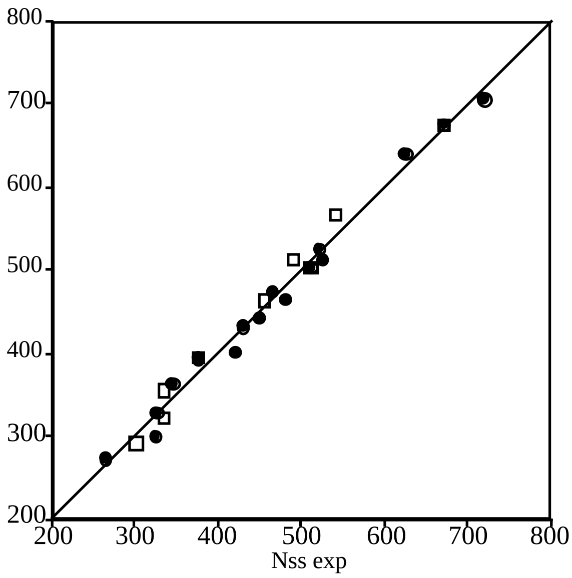


**Fig. 7 -** Plot of Nss computed vs Nss experimental. Symbols as in Fig.6.

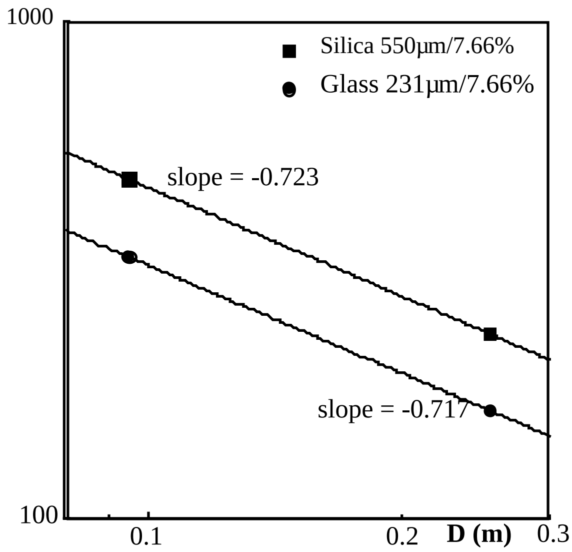


**Fig. 8 -** Plots of Nss experimental vs vessel size.

Liquid viscosity effects were also assessed by using ethylene glycol (whose viscosity is about 20 times higher than that of the water used so far) for a number of runs at various concentrations and particle sizes. The relevant results (not reported for the sake of space) indicate apparently that viscosity doesn't practically affect the $N_{ss}$ values.

As already mentioned, in order to study scale up effects, a limited number of experiments was carried out in a larger vessel. The results obtained are shown in Fig.8 where it can be observed that $N_{ss}$ depends on the scale of the tank raised to an exponent of -0.72. This result falls between that obtained by Buurman, (1986) who suggested an exponent of -0.667, and the value of -0.85 suggested by Zwietering (1958) and is in practical agreement with the value of (-0.76) found by Chapman et al, (1983). Interestingly it was observed that the dynamic head effects were much less important in the larger tank as compared with the smaller tank. This is due to the fact that, for a given particle type and concentration, combining the scale-up of dynamic head effects (which are proportional to $N^2D^2$), that of $N_{ss}$ (which depends on $D^{-0.72}$) and on $\Delta P_{js}$ (which is proportional to D) results in a ratio $\Delta P_{dh}/\Delta P_{js}$ proportional to $D^{-0.44}$. This implies that the proposed technique should be more reliable the larger the agitated tank, confirming the good applicability to industrial scale tanks.

So far all the results presented refer to the case of a Rushton turbine. As mentioned earlier, for the purpose of this work a 45-angled pitched blade turbine (PBT downward pumping) was also employed, in order to further apply the technique to the important case of axial impellers. Typical results are reported in Fig.9. It can be observed once again that fractional suspension data *vs* agitation speed take the form of "S" shaped curves. The data collected allowed for the assessment of $N_{ss}$ as a function of particle concentration B, particle diameter $D_p$ and scale-up effects. The relevant results are shown in Figs. 10, 11 and 12.

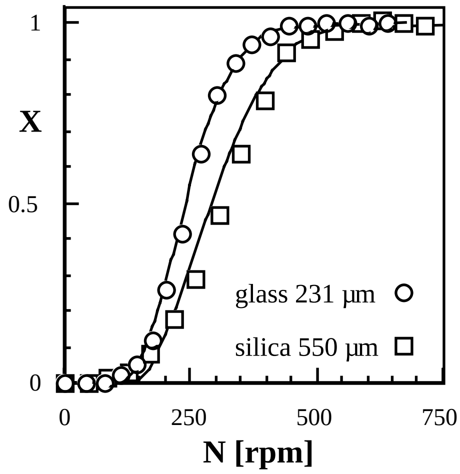


**Fig.9**–(PBT) Fractional suspension *vs* N; B=15.8%

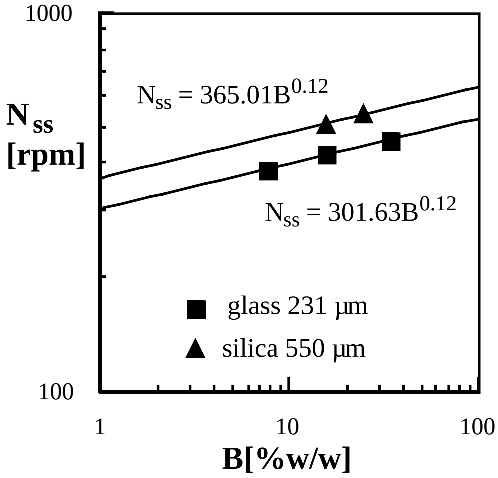


**Fig.10**–(PBT) $N_{ss}$ *vs* B; $D_T$=0.19m

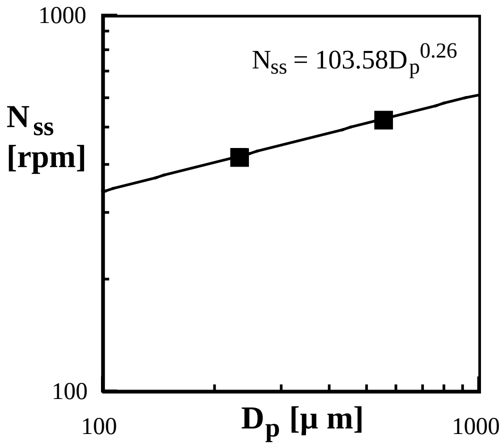


**Fig.11**–(PBT) $N_{ss}$ *vs* $D_p$; B=15.8%

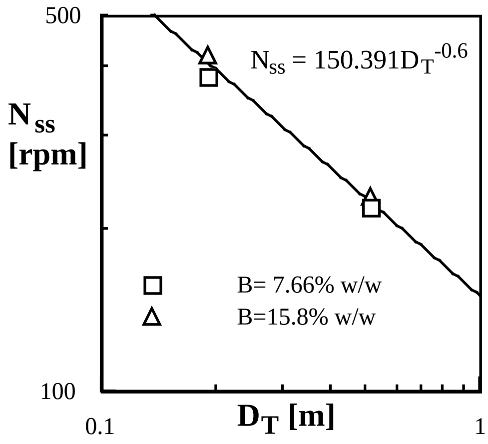


**Fig.12**–(PBT) $N_{ss}$ *vs* $D_T$; glass 231µm

The above results can be summarized in the following monomial correlation for $N_{ss}$:

$$N_{ss} = 73.5\ D_p^{0.26}\ B^{0.12} (D_T/0.19)^{-0.6} \qquad (14)$$

As it can be observed, the exponent for B is almost identical to the one found for the case of the Rushton turbine; conversely the exponent for $D_p$ is found to be smaller and definitely closer to the one adopted in Zwietering's correlation (i.e. 0.20). One final comment for the scale-up exponent which was found to be equal to –0.6. This finding apparently indicates that the power input per unit volume should be required to increase (i.e. $P/V \propto D^{+0.2}$) rather than to decrease as in the case of the Rushton turbine (i.e. $P/V \propto D^{-0.16}$).

## CONCLUSIONS

A simple technique has been developed for measuring the fraction of suspended solids in stirred tanks operating below the just-complete suspension speed. As the technique is based on simple pressure measurements on the vessel bottom, it does not require expensive instrumentation and is very suitable for the application to industrial scale agitated tanks.

All the fractional suspension data collected showed a typical "S" shaped dependence on agitation speed. The experimental curves were well fitted by Weibull functions with an exponent of 2. The two parameters ($N_{min}$ and $N_{span}$) there appearing have an immediate physical meaning and could be correlated to the physical parameters varied in the experimentation (particle size and concentration, scale of operation) by simple power-law functions.The correlations so far proposed allow, at least for the investigated geometries and physical properties of the two phases, the estimation of the amount of suspended (or unsuspended) particles in an agitated tank operated at partial suspension conditions. Since a number of industrial units are actually run under such conditions, it is believed that the present work may have an immediate industrial interest.

## NOTATION:

| | | |
|---|---|---|
| $A_b$ | = | vessel bottom area [$m^2$] |
| B | = | total solids concentration [% w/w] |
| C | = | impeller clearance from tank bottom [m] |
| D | = | impeller diameter [m] |
| $\varepsilon_s$ | = | fractional voidage of bed of particles [-] |
| $\Delta P$ | = | total pressure increase measured at the tank bottom [Pa] |
| $\Delta P_{dh}$ | = | pressure increase at the tank bottom due to dynamic head effects [Pa] |
| $\Delta P_{js}$ | = | hydrostatic pressure increase at complete suspension [Pa] |
| $\Delta P_s$ | = | pressure increase at the tank bottom due to particles suspension [Pa] |
| h | = | average distance between the impeller and the surface of fillets [m] |
| g | = | gravitational acceleration [$m/s^2$] |
| $M_s$ | = | mass of suspended solid particles [kg] |
| $M_t$ | = | total mass of solid particles in the vessel [kg] |
| $\rho_s$, $\rho_l$ | = | densities of solid and liquid phases [$kg/m^3$] |
| N | = | agitation speed (rpm); |
| $N_{min}$ | = | agitation speed at which the suspension phenomenon starts (rpm); |
| $N_{span}$ | = | parameter that determines the width of the "S" (rpm); |
| T | = | vessel diameter [m] |
| $V_s$ | = | volume of suspended solid particles [m] |
| X | = | fractional suspension ($M_s/M_t$); |

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