LETTERS TO THE EDITOR

To the Editor:

In a recent work titled "Hydrodynamics of a Semibatch Slurry Bubble Column with a Foaming Liquid," Pino et al. (November 1990, p. 1758) mentioned, based on results of Smith and Ruether (1985), that under the most common operating conditions, the volume of gas and liquid phases in a slurry bubble column are approximately uniform throughout the column. They also indicated that despite the concentration of the solid phase in the slurry might vary with the axial position, the solid volume fraction is so small that the axial variation of the solid concentration does not result in gas or liquid holdup changes throughout the column. We would like to address some concerns regarding these statements.

Pino et al. (1990, Figure 4b) reported for a given operating condition ($C_{s\,\mathrm{avg}} = 337.5$ kg/m^3) a solid concentration of approximately 700 kg/m^3 in the bottom of the column and 192 kg/m^3 at the top of the column. This corresponds to a solid volumetric fraction of 32.3% in the bottom of the column and of 8.9% at the top of the slurry column, using the apparent solid density given by the authors (2,170 kg/m^3). This represents a variation of the solid holdup from the bottom of the column to the top of the column of approximately 72.4%. The axial variation of the solid holdup for the other operating condition reported (Figure 4b, Pino et al., 1990, $C_{s\,\mathrm{avg}} = 86.0$ kg/m^3) was estimated to be even more important (74.2%). This significant axial variation of the solid holdup must be necessarily compensated by a proportional variation of the liquid holdup and/or the gas holdup according to the following equation:

$$\epsilon_g + \epsilon_l + \epsilon_s = 1 \tag{1}$$

In fact, when considering the axial profile of the gas holdup constant, which is consistent with the findings of several authors (Smith and Ruether, 1985; Tang and Fan, 1989), it was estimated that the mean liquid fraction in the slurry ($\psi_{l,\mathrm{avg}}$) and the actual local liquid fraction in the slurry ($\psi_{l,\mathrm{loc}}$) could differ by up to 19%. The assumption of constant axial liquid fraction in the slurry suggested by Pino et al. (1990) was based on results obtained by Smith and Ruether (1985) who reported differences between $\psi_{l,\mathrm{avg}}$ and $\psi_{l,\mathrm{loc}}$ almost twice as small. Consequently, the assumption of constant axial liquid fraction in the slurry (ψ_l) suggested by Pino et al. (1990) might be unrealistic for some of the conditions reported by the authors. Actually, important axial variation of the liquid fraction in the slurry (ψ_l) could directly affect the value of the Bodenstein number as shown in the following equation:

$$Bo = \psi_l U_p H / E_s \tag{2}$$

In addition, important axial variation of ψ_l could also affect indirectly the value of the Bodenstein number, knowing that the particle sedimentation velocity U_p could be a strong function of the liquid fraction in the slurry (ψ_l), as illustrated in Eqs. 3, 4, and 5, and assuming that the solid dispersion coefficient (E_s) is not significantly dependent on the liquid fraction in the slurry as it is generally observed (Smith and Ruether, 1985; Smith et al., 1986; O'Dowd et al., 1987).

$$U_p = 1.91\, U_g^{0.26}\, U_t^{0.8}\, \psi_l^{3.5} \quad \text{[Smith and Ruether, 1985]} \tag{3}$$

$$U_p = 1.33\, U_t\, (U_g/U_t)^{0.25}\, \psi_l^{2.5} \quad \text{[Kato et al., 1972]} \tag{4}$$

$$U_p = 1.69\, U_g^{0.23}\, U_t^{0.80}\, \psi_l^{1.28} \quad \text{[O'Dowd et al., 1987]} \tag{5}$$

It can consequently be speculated that the Bodenstein number could be affected strongly by axial variations of the liquid fraction in the slurry ψ_l. The sharp changes in the solid concentrations profiles near the bed upper interface indeed imply the requirement of a global Bodenstein number varying with the axial position. The axial variation of the global Bodenstein number is also suggested, as reported by Pino et al. (1990), by the nonlinear behavior of the natural logarithm of the total concentration of the solid vs. the axial position (nonconstant slope: $-Bo/H$) and represented in Figure 4b (Pino et al., 1990). In addition, the nonlinear behavior of the natural logarithm of the solid concentration vs. the axial position or, in other words, the changes of the Bodenstein number with the axial position might be different for each individual particle size ranges considered, being negligible for the smaller particles and more important for the larger particles. Consequently, the imprecision in the evaluation of the Bodenstein number could be dependent on the particle size range investigated. Actually, this might be at the origin of the somewhat scattered behavior of the Bodenstein number reported in Figures 2 and 3a (Pino et al., 1990) for the larger particles ranges.

So, the axial variation of the liquid fraction in the slurry ψ_l might be of importance in some cases. Actually, Matsumoto et al. (1989) succeeded to fit the one-dimensional sedimentation-dispersion model in systems where important axial solid concentration variations were present by including in the model the dependency of the sedimentation term on the axial variation of the solid concentration.

Notation

C_s = solids concentration, kg/m^3
$C_{s\,\mathrm{avg}}$ = average solids concentration, kg/m^3
E_s = solid dispersion coefficient, m^2/s
H = total gas-liquid-solid suspension height, m
U_g = superficial gas velocity, m/s
U_p = particle sedimentation velocity, m/s
U_t = terminal settling velocity of a single particle, m/s
ϵ_g = gas holdup
ϵ_l = liquid holdup
ϵ_s = solid holdup
ψ_l = volume fraction of interparticle liquid
$\psi_{l,\mathrm{avg}}\, \psi_{l,\mathrm{loc}}$ = mean and local volume fraction of interparticle liquid

Literature cited

Kato, Y., A. Nishiwaki, T. Fukuda, and S. Tanaka, "The Behavior of Suspended Solid Particles and Liquid in Bubble Columns," *J. Chem. Eng. Jap.*, 5(2), 112 (1972).

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Reply:

We would like to thank Drs. Chabot and de Lasa for their comments on and interest in our recent article (Pino et al., 1990). They raise their concern on the effects that axial variations of the interparticle volume fraction of liquid might have on our analysis. We would like to elaborate on this point.

Working within the frame of the actual knowledge in the field, the only parameter related to phase distributions in the bubble column that affects the value of the Bodenstein number seems to be the interparticle volume fraction of liquid. As Chabot and de Lasa point out, a change of this parameter with axial position could imply a change of the local Bodenstein number with z, in which case the semilog plot C_{si} vs. z would not be a straight line. In some of the data presented in our work, there are appreciable variations of ψ_l with z. To quantify these variations, we have calculated the maximum percent deviation of the local values of ψ_l in the data with respect to the average value of this parameter in the bubble column. The results corresponding to all the experiments reported in our work are presented in Table 1. In this table, b and f represent bubbling and foaming regions, respectively, for which the sedimentation-dispersion model was applied separately. The asterisk indicates the data point to which Chabot and de Lasa refer explicitly in their comments.

Table 1. Variations of Interparticle Volume Fraction of Liquid

V_g (cm/s)	$\bar{C}_s$ (kg/m^3)	$\Delta\psi_l$ (%)
3	22.7	0.8 (b), 0.1 (f)
3	111.0	2.3 (b), 0.8 (f)
3	260.5	9.7 (b), 1.8 (f)
3	297.3	5.1 (b), 0.8 (f)
6	42.6	1.3 (b), 0.3 (f)
6	122.4	10.1
6	210.3	12.6
6	302.2	16.4
12.5	86.0	4.0
12.5	178.9	8.0
12.5	277.4	9.0
12.5	337.5	14.1 (*)

Even though ψ_l changes noticeably in some of the experiments, the plots C_{si} vs. z were always straight lines (within the confidence limits imposed by experimental error). We can speculate that this might imply that the Bodenstein number corresponding to each particle size range is not as sensitive to the interparticle volume fraction of liquid as previous empirical correlations might lead to think. The correlations developed by Kato et al. (1972), Smith and Ruether (1985), and O'Dowd et al. (1987) predict dependencies of the Bodenstein number on ψ_l that range between $\psi_l^{2.3}$ to $\psi_l^{4.5}$. It, however, should be pointed out that these correlations were developed in systems where the liquid phase was aqueous, whereas our liquid phase was an organic foaming liquid.

The experimental data in Figures 4a and 4b of our article correspond to total solids concentration. The data show a definite curvature in the semilog scale, which is caused, as we pointed out, by the superposition of exponential functions of all the particle size ranges, not by changes in the local Bodenstein number. Note that the curve corresponding to an average solids concentration of 86.0 kg/m^3 in Figure 4b, which is not a straight line, corresponds to an almost negligible variation of ψ_l with the axial position (Table 1).

Notation

$\bar{C}_s$ = average solids concentration
C_{si} = solids concentration of ith particle size range
V_g = superficial gas velocity
z = axial position
ψ_l = interparticle volume fraction of liquid
$\Delta\psi_l$ = maximum deviation of ψ_l with respect to the average value

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To the Editor:

In the note titled "Gas-Liquid Mass Transfer in Stirred Tanks," Arrua et al. (November 1990, p. 1768) presented a method for determining physical mass-transfer coefficient (k_La_B) between gas bubbles and liquid in stirred vessel. The method for determining physical mass-transfer coefficient (k_La_B) between gas bubbles and liquid in stirred vessel. The sparged with helium. They used the following rate equation to describe mass transfer of CO_2 from liquid to gas bubbles:

$$r = k_L a_B \left\{ [CO_2]_{aq} - \frac{[CO_2]_g}{H} \right\} \quad (1)$$

This is a correct form of Eq. 4 given by Arrua et al. (1990), in which $[CO_2]_{aq}/H$ was used, instead of $[CO_2]_{aq}$. This mistake did not proliferate to their key relation (Eq. 12). The concentration of CO_2 in the liquid at the interface is assumed to be in equilibrium with that in the bulk gas bubble, and CO_2 concentration in the solution, $[CO_2]_{aq}$, is the equilibrium value determined from the reversible reactions:

$$H_2O + CO_2 \rightleftharpoons HCO_3^- + H^+ \qquad (2)$$

$$HCO_3^- \rightleftharpoons CO_3^= + H^+ \qquad (3)$$

They assumed that the reactions are fast with respect to mass transfer, and from this assumption they concluded that the rate of evolution from the carbonate solution is controlled by mass transfer from liquid to gas bubble described by Eq. 1. This equation, however, entirely ignores the effect of the accompanying reaction on the mass-transfer coefficient as it assumes a *physical* desorption of CO_2 with driving force given by the difference of free CO_2 concentration in the liquid bulk and at the interface. If we assume very rapid reactions (Eqs. 2 and 3), in agreement with the authors, it is necessary to take into account the chemical enhancement caused by these accompanying rapid reversible reactions.

The analysis of the effect of very rapid reversible reactions on the measured value of $k_L a_B$ is absent in their work. Therefore, their conclusions lack substantiation and, according to our opinion, are misleading: "The procedure should be applicable to other chemical systems where a gas is evolved from the liquid by a chemical reaction of very rapid or known kinetics." This recommendation does not specify exactly the type of reactions suitable for this purpose. Only reactions, which do not affect the value of the coefficient of physical mass transfer, are proper ones. This is the case where the accompanying chemical reaction occurs mainly in the bulk of liquid and only to negligible extent in the liquid film at the interface. Very rapid reversible reaction used by the authors in their example is the least suitable one for this purpose, because it predominantly takes place in the liquid film, as a part of the equilibrium CO_2 leaves the liquid film to enter the gas phase and is replaced by very rapid decomposition of HCO_3^- by the equilibrium reaction (Eq. 2). On the contrary, proper sort of reactions are rapid *irreversible* reactions that occur in the bulk of liquid without any significant effect on the concentration profiles of the components in the liquid film. A good example of this reaction is the decomposition of hydrogen peroxide to water and oxygen catalyzed by the enzyme catalase. The oxygen produced is then desorbed to the air stream (Hickman, 1988).

We suppose here that the method presented by Arrua et al. (1990) concerns the measurement of the coefficient of *physical* mass transfer, even though it is not explicitly stated in their work. This suspicion stems from the fact that as alternative methods of $k_L a_B$ measurements the authors mentioned the dynamic methods reviewed by van't Riet (1979), all of them concerning the physical mass-transfer coefficient.

If Arrua et al. did not verify that the effect of the very rapid reversible reaction in the liquid film could be neglected, then the comparison of their results with physical mass-transfer coefficients obtained by others seems pointless.

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Reply:

We regret the typographical error in Eq. 4 of our note (Arrua et al., 1990) and are pleased to have this noted. This misprint, which also appears in Eqs. 10 and 11, does not affect subsequent equations or the results and conclusions of the note.

Other comments by Linke and Sinkule appear to be based on a misunderstanding of the objective of our work. Their comments are concerned with the *interpretation* of the resultant mass transfer, that is, whether it is affected by rapid reactions in the liquid film around the gas bubble. This was not the goal of our study. Our objective was to describe and illustrate a new *experimental method* for evaluating an *observed* mass transfer.

Linke and Sinkule also misinterpreted the statement in the note: "The procedure should be applicable to other chemical systems where a gas is evolved from the liquid by a chemical reaction of very rapid or known kinetics." What is meant is in our example that the reactions were rapid enough to assume equilibrium (essentially zero intercept in Figure 2), but that the experimental method was also applicable if the kinetics of the liquid-phase reactions were known. The note did not address the question of how reactions in the liquid film affect $k_L a_B$, which appears to be the theme of the Linke-Sinkule comments.

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To the Editor:

I would like to make some comments on a recent note by Nikov and Karamenev (May 1991, p. 781) titled "Liquid-Solid Mass Transfer in Inverse Fluidized Beds."

The authors made some contradictory statements: while, referring to their Figure 3, they say that "the mass transfer remains *almost* unaffected by the velocity," they later state that "there is *no* effect of the liquid superficial velocity on the mass-transfer rate." Their Figure 3, as well as their Figures 2 and 4, indicates that there is in fact a detectable effect of liquid velocity. Referring to their Figure 4, the authors stated that "there is a *small* effect of the particle diameter, which is within 11%," but subsequently added that "there is also *no* effect of the particle size."

The authors cited seven empirical equations that had been recommended for liquid-solid mass transfer in fluidized beds, including their own for inversely fluidized beds. These equations, presumably all for spherical particles, all have the simple exponential form:

$$Sh = C\, Re^a\, Ga^b\, Mv^c\, Sc^d \qquad (A)$$

Since $GaMv = Ar$, where the Archimedes number $Ar = d_p^3 |\rho_s - \rho_L| g \rho_L / \mu_L^2$, it follows that Eq. A can be rewritten as:

$$Sh = C\, Re^a\, Ar^b\, Mv^{c-b}\, Sc^d \qquad (B)$$

The Archimedes number, in which the term d_p^3, proportional to particle volume,

Table 1. Values of Exponents in Eq. B

	a	b	$c-b$	d
Nikov and Karamenev (1991)	0	+0.33	0	+0.33
Nikov and Delmas (1987)	0	+0.33	0	+0.33
Arters and Fan (1986)	0	+0.323	−0.023	+0.4
Riba and Couderc (1980)	−0.07	+0.36	+0.01	+0.33
Tournié et al. (1977)	0	+0.323	−0.023	+0.4
Tournié et al. (1977)	+0.004	+0.319	−0.020	+0.4
Calderbank (1967); Lee et al. (1974)	0	+0.333	0	+0.333

is multiplied by $|\rho_s-\rho_L|g$, the buoyed specific weight of the particles, to represent the net gravitational force on a particle, is more obviously related to the physics of particle fluidization than the Galileo number, in which $|\rho_s-\rho_L|$ is replaced by ρ_L. Thus, equations for Re_{mf} (Grace, 1982) and for ϵ in the case of particulate fluidization (Wen and Yu, 1966) commonly contain *Ar* as a crucial dimensionless variable which characterizes the relevant fluid and particle properties (note that Wen and Yu mislabeled their *Ar* as *Ga*). The values of exponents in Eq. B for the seven equations cited by Nikov and Karamanov are listed in Table 1.

As Table 1 shows, there is no unanimity about the influence of *Re*, but data such as those of Nikov & Karamenov's Figures 2-4 imply a real, but minor, effect of that dimensionless variable on mass transfer. This table also shows that the exponent on *Mv* is in three cases zero, and its absolute value is never larger than 0.023. These values of $c-b$ are too small to be taken seriously, even if one accepts the simple exponential form of Eq. B, given the experimental error associated with a measurement of fluidized-bed mass transfer. After all these years of experiments on the subject, one should thus conclude that $c-b$ is essentially zero and that for particle-fluid mass transfer in liquid fluidized beds at Reynolds numbers sufficiently high that pure molecular diffusion can be neglected relative to convective mass transfer ($Sh \gg 2$),

$$Sh = C\,Re^a\,Ar^b\,Sc^d \qquad \text{(C)}$$

Notation

a = exponent on particle Reynolds number
Ar = Archimedes number $= d_p^3|\rho_s-\rho_L|g\rho_L/\mu_L^2$
b = exponents on Galileo and Archimedes numbers
c = exponent on density number in Eq. A
C = coefficient in Eqs. A, B and C
d = exponent on Schmidt number
d_p = particle diameter, m
D = molecular diffusivity, m^2/s
g = acceleration of gravity, m/s^2
Ga = Galileo number = $d_p^3 g\rho_L^2/\mu_L^2$
K = liquid-solid mass transfer coefficient, m/s
Mv = density number = $|\rho_s-\rho_L|/\rho_L$
Re = particle Reynolds number $= d_pU_L\rho_L/\mu_L$
Re_{mf} = particle Reynolds number at minimum fluidization
Sc = Schmidt number = $\mu_L/\rho_L D$
Sh = Sherwood number = Kd_p/D
U_L = superficial liquid velocity, m/s
ϵ = voidage
μ_L = liquid viscosity, Pa·s
ρ_L = liquid density, kg/m^3
ρ_s = solid particle density, kg/m^3

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Reply:

We would like to thank Dr. Epstein for his interest in our work. I would like to respond to two main points he raised:

1) Is there any detectable effect of the Reynolds number (liquid velocity) on the Sherwood number (mass-transfer coefficient) in liquid-solid fluidized beds.

2) Is it rational to replace *Ar* by *Ga* and *Mv* in the empirical equation (Eq. C)?

The first question can be divided into two issues: 1a) effect of *Re* on *Sh* in classic (upflow) fluidization; 1b) in inverse fluidization. Table 1 of his letter answers question 1a. It shows that the exponent on *Re* is in four cases equal to zero, and in the remaining two cases of classic fluidized beds is very close to this value. We can refer also to the work of Fukuma et al. (1988). The effect of liquid velocity on the mass-transfer coefficient in the case of two-phase inverse fluidization (issue 1b) has been studied in our previous article (Nikov and Karamanev, 1991). Figure 1, reproduced from this work, shows that the slope of this dependence can be both negative and positive, and the scatter is within the experimental error (5.4%). We plotted all the data used in the above cited article (34 points) as the *Re* function of $Sh\cdot(GaMvSc)^{-0.33}$ (Figure 2). It shows that the slope of the line obtained by linear regression is close to 0. Besides, all these data points have been processed by nonlinear regression to obtain the constants in the following empirical equation:

$$Sh = C\cdot Re^a Ga^b Mv^c Sc^d \qquad \text{(A)}$$

The results were: $c=0.28$; $b=c=d=0.33$; $a=-0.0057$. The exponent *a*, therefore, can be assumed to be zero.

It can be generally concluded that liquid velocity (and *Re*) has no detectable effect on the mass transfer in both upflow and inverse fluidization.

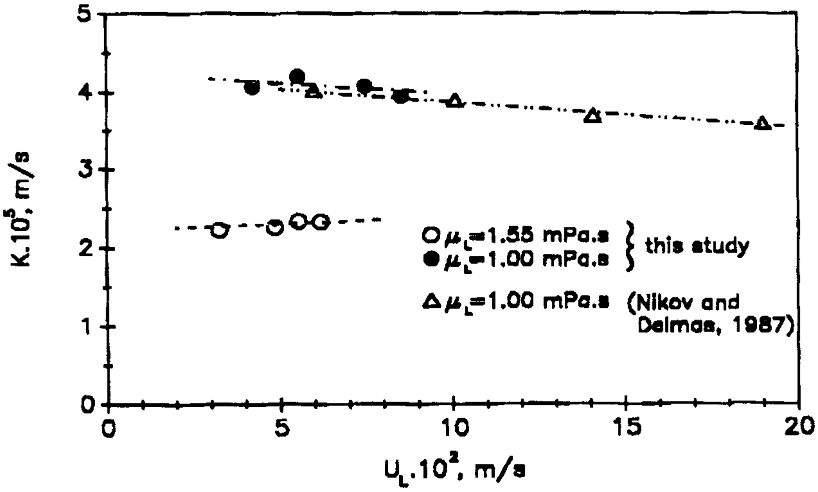

Figure 1. Mass-transfer coefficient vs. liquid velocity.

We used the dimensionless density number *Mv* in Eq. A because of the following reasons. Undoubtedly, the Archimedes number accounts for the effect of both gravitational and Archimedes forces on fluidized particles. The division of *Ar* into two dimensionless groups, *Ga* and *Mv*, has been made by many researchers studying mass transfer in fluidized beds (Tournie et al., 1977; Riba and Couderc, 1980; Nikov and Delmas, 1987; Arters and Fan, 1986; Lee et al., 1974) to underline the importance of the effect of density difference on the mass transfer. Dr. Epstein is correct in saying that in the cases of upflow and inverse fluidization, the exponents on both *Ga* and *Mv* (*b* and *c*, respectively) are similar or equal to each other. Nevertheless, we used *Ga* and *Mv*, instead of *Ar*, to compare the coefficients and exponents in both types of fluidization. Therefore, we accept Dr. Epstein's proposal that Eq. C is more correct from physical point of view. However, it is not contradictory to our Eq. A.

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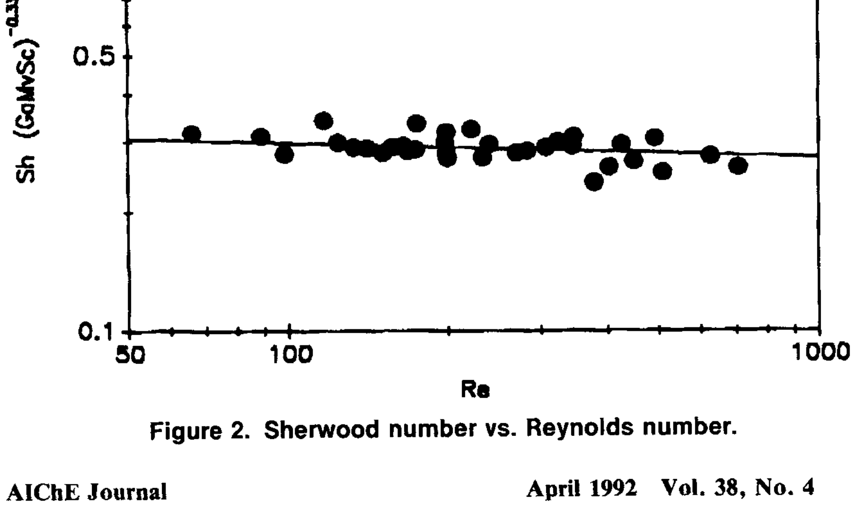

Figure 2. Sherwood number vs. Reynolds number.

To the Editor:

In a recent article titled "Prediction of the Three-Dimensional Turbulent Flow in Stirred Tanks," Kresta and Wood (March 1991, p. 448) correctly highlighted the importance of impeller boundary conditions and should be commended for extending swirling radial jet model to generate these boundary conditions. The authors, however, missed some points in their comparison between their model predictions and experimental data and previous modeling results. We would like to point out the following:

1. Ranade and Joshi (1990) previously presented detailed experimental data and predictions of three-dimensional turbulent flow field for flow generated by disc turbine in baffled vessels. Costes and Couderc (1988) also presented detailed data for the bulk region of the tank.

2. Authors describe some of the previous modeling attempts to incorporate the influence of baffles as inadequate and go on describing their attempt of including baffles in the 3-D calculation domain as the first attempt. However, previous models of Ranade and Joshi (1990) and Ranade et al. (1989) successfully included baffles in the 3-D calculation domain. Ranade and Joshi (1990) even presented the comparisons of predicted and experimental axial and tangential mean velocities in near wall region. Since the authors presented their predictions in

the form of vector plots, we cannot present direct comparison of their predictions with our data/predictions here.

3. The comparison section of the mean velocities with the predicted results lacks information on mean tangential velocity. It is essential to predict the decay of tangential velocity along the radius in impeller center plane, which is much faster than the radial velocity, for the correct prediction of flow field in the bulk region.

4. While comparing their model predictions of turbulent kinetic energy with experimental data, the authors state that such an agreement has not been previously reported. In our work, we presented detailed comparison of predicted results with experimental data, not only in impeller center plane but also in the bulk region of the tank. In fact, our model correctly predicts a slight increase in the turbulent kinetic energy after the impeller [as data of Cutter (1966) and Wu and Patterson (1989) suggest from Figure 9 of the authors' article], which was not predicted by the authors' model.

5. The success of the swirling radial jet model can be properly judged only by comparing the predicted flow characteristics in the bulk region with the experiments. In addition, if the authors can compare axial profiles of energy dissipation in impeller stream with the data of Wu and Patterson (1989) (which show double peaks), it would increase the confidence in swirling jet model.

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