MASS TRANSFER EFFICIENCY IN SX MIXERS

By

R. Sheinman, Y. Kokotov, L. Braginsky, J. Riordan, **M. Vancas**

Turbulent Technologies Ltd. Israel
Tenova Bateman Advanced Technologies Ltd.
Israel, Australia

ABSTRACT

Mass transfer is a key issue in solvent extraction processes and may actually be a bottleneck in some plants. To achieve the necessary stage efficiency an optimum droplet size must be created. Those droplets must be maintained in a turbulent environment to ensure intensive mass transfer. Low intensity mixing does not provide sufficient turbulence or mass transfer. Mixing that is too aggressive will develop fractions that cannot be separated in the settler. Previous studies have shown that the primary reason for separation problems is the extremely non-uniform energy distribution in the mixer which leads to extremely non-uniform distribution of droplet sizes.

New types of impellers have been developed in recent years which produce a more uniform energy distribution and more uniform droplets which reduce the entrainment of the dispersed phase in the continuous phase; even for high power inputs.

This study examines the influence of distribution of energy on mass transfer efficiency.

ENERGY DISTRIBUTION IN THE MIXING TANK

Studies have shown that the volume of the highly turbulent zone around typical hydro-foil impellers is small. Consequently, only a fraction of the extraction takes place in this zone and the majority of the extraction actually takes place in the, less efficient, bulk of the mixer volume. For this reason, an increase in turbulence in the bulk of the mixer has a positive impact on mass transfer efficiency. Studies have also shown that mass transfer is increased the most under conditions of low percentage of extractant.

Distribution of energy in mixing tanks is normally highly non-uniform. Because of a complete lack of systematic experimental data collection on the energy distribution, the models and experimental correlations related to drop or bubble breaking and mass transfer in disperse systems have been based on average energy dissipation values. The average energy dissipation is also known as the specific power – mixing power per unit of mass (W/kg) in the tank. The specific power has been used for many years as the basic parameter for scaling-up, especially – for the processes connected with mass transfer. The use of the average parameter instead of data on real values of turbulent dissipation places strict limits on ones ability to compare and select optimum mixing devices. The knowledge of energy distribution provides a useful tool for development of mixing equipment.

A method has been developed (1) that allows definition of not only the location of zones with different intensity of turbulence, but also a quantitative evaluation of the volume and energy dissipation in these zones as a function of tank and impeller geometry. The intent of this paper is to show how application of these methods corresponds to solvent extraction of metals.

As mentioned previously, in the typical design of mixers for liquid-liquid metallurgical extraction processes there exists a very small zone of high energy dissipation in the vicinity of the impeller blades and extremely low energy dissipation in the bulk volume of the mixing tank. The special mixers, newly developed for these applications by Turbulent Technologies (TT) and Tenova Bateman, provide much more uniform distribution of energy.

With any mixing impeller and tank design, the degree of non uniformity depends not only on the impeller type but also on many other parameters. Figures 1 and 2 illustrate the influence of impeller diameter on the ratio of maximum to average energy dissipation rates and on the relative volume of zone of maximum energy dissipation. The calculations were performed for an industrial scale (80 m^3) mixing system with a standard hydrofoil impeller and the same system with a special TT impeller.

According to these results, the maximum energy dissipation rate (W/kg) decreases while the volume of the zone of maximum dissipation increases with increase of impeller diameter.

The width of the baffles is another important parameter that affects turbulence distribution in the tank. Studies by others have shown (2) that the installation of baffles in mixing tanks creates a resistance to the tangential flow and thus dramatically decreases tangential velocity of liquid, transforming tangential flow to axial circulation. As a result, the difference between the velocities of the blade and the liquid increases, thereby increasing the energy dissipation.
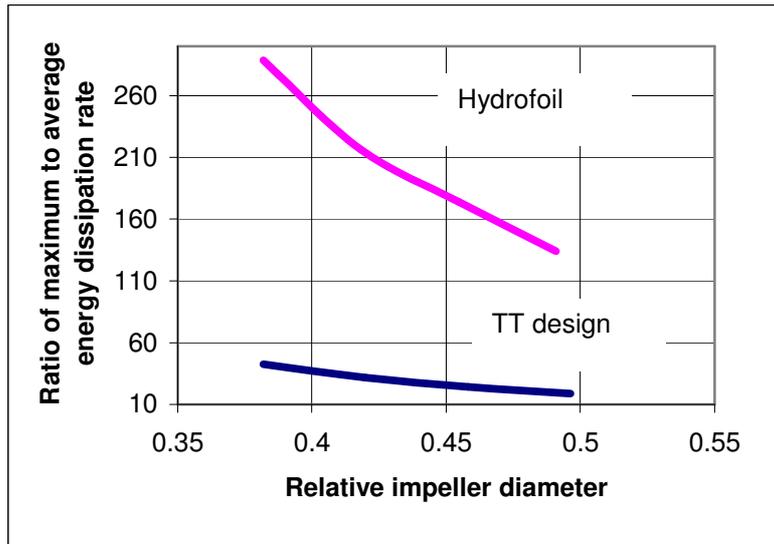
2

Figure 1: $\varepsilon_{max}/\varepsilon_{avg}$ vs. relative impeller diameter in production scale mixer.
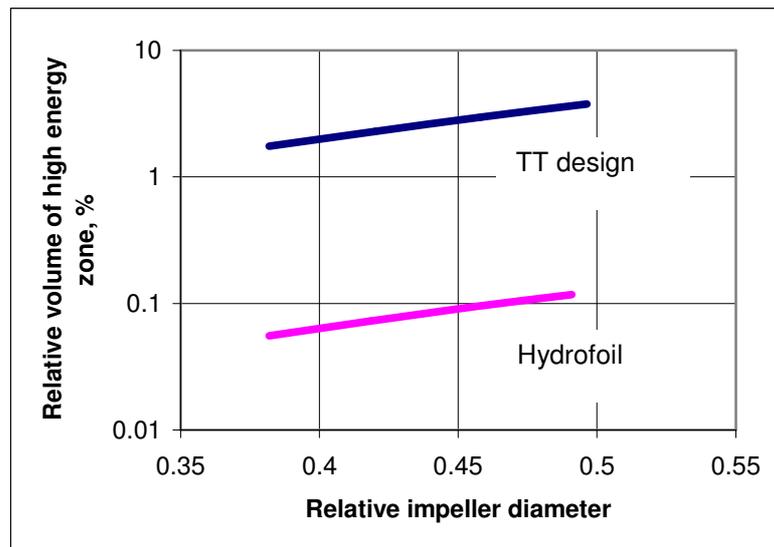

Figure 2: Relative volume of high shear zone vs. relative impeller diameter in production scale mixer.

As expected an increase of the blade width W_{blade} causes an increase of the average and maximum energy dissipation rates. This is correct for the mixer with hydro-foil impeller. As shown in Figure 3, the average energy dissipation decreases monotonically with decreasing baffle width, and is at a minimum at a baffle width of about 0.03 times the tank diameter.

Mixing systems with TT impellers have much bigger blades and create higher torques, causing more complicated dependence (Figure 4).
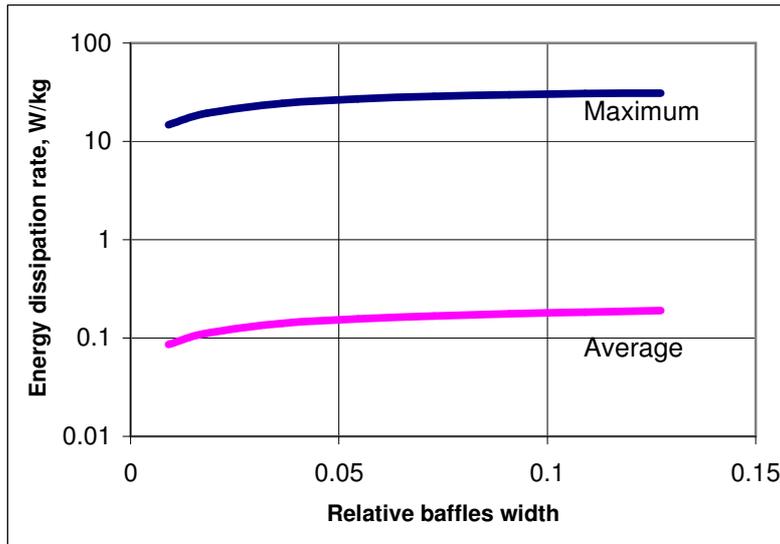
Figure 3: Maximum and average energy dissipation vs. relative blade width in typical mixing system with hydrofoil impeller.

This effect can be explained by the transition of the maximum energy zone from the zone behind the impeller blades to the zone behind the baffles. Following Levich (3), the specific energy dissipation in this area depends on the velocity of the liquid near the baffles and baffles' width:

$$\varepsilon_{max,baffle} \sim \frac{v_{t,baffle}^3}{W_{baffle}}$$

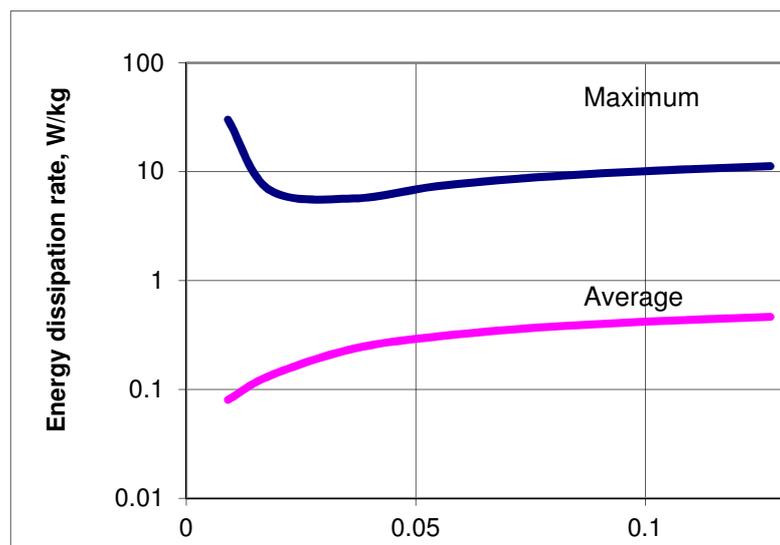The velocity of the liquid, approaching the baffles ($v_{t,baffle}$) increases as the width of the baffles decreases.

Figure 4: Maximum and average energy dissipation vs. relative blade width in typical mixer with Special TT impeller.

4

ENERGY REDISTRIBUTION AND REDUCTION OF AFTER-SEPARATION ENTRAINMENT

The possibility of controlling the turbulence distribution in the mixing tank enables one to optimize different technological processes dependent on mixing. In particular, based on mathematical models of droplet breakage and droplet coalescence in turbulent flow (which were developed by the authors) (4,5,6), results have shown that with newly developed impellers a more uniform energy distribution is achieved. This has proven true over a range of maximum specific powers of 4-8 W/kg and a minimum of 0.3-0.8 W/kg. As a result, the quantity of very small and practically inseparable droplets decreases with time, and this leads to the dramatic reduction of entrainment.

This was proved by pilot and industrial tests in a number of Cu SX plants in Australia, Chile and USA, where reduction in entrainment values by a factor of 2 - 7 times was achieved (7).

ENERGY REDISTRIBUTION AND MASS TRANSFER EFFICIENCY

The theoretical analysis presented in this section is intended to evaluate the possible influence of turbulent energy distribution on the mass transfer rate.Specific mass transfer area

Mass transfer rate, and thus stage extraction efficiency for SX in hydrometallurgy depends on (a) residence time, (b) concentration of dispersed phase, φ, (c) specific mass transfer area, A and (d) mass transfer coefficient, k.

The mass transfer coefficient (k) is a function of average drop diameter $d_{32} = 6\varphi / A$, the local energy dissipation rate, the physical properties of the aqueous and organic phases, the frequencies of droplet breakage and droplet coalescence, equilibrium at the interface and the physical - chemical properties of the metal and the extractant.

Due to significantly higher levels of energy dissipation, one can expect that a typical mixing system will create smaller drops and the mass transfer area will be larger, compared to a mixer with uniform energy distribution (Figures 3 and 4). However, kinetic curves for both cases (Figures 5, 6) show that the process of droplet breakage in a typical mixer configuration may take hours. The reason for such slow process is the extremely small volume of the zone where droplet breakage takes place. In the newly developed system the process of droplet breakage takes only about 2 minutes. Taking into account that the normal residence time in a SX mixer in a hydrometallurgical plant is 1 to 3minutes, the average drop size in the newly developed TT system is smaller than in a typical mixing system.
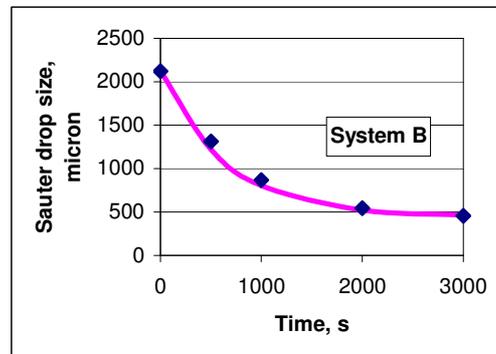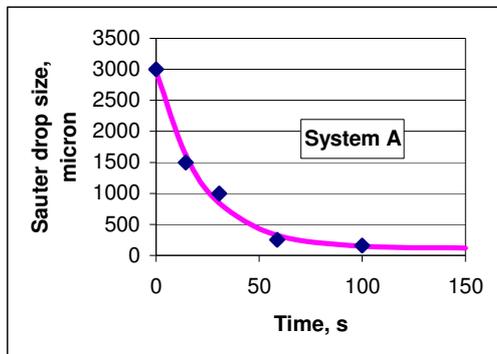
Figures 5, 6:   Typical kinetic curves of drop breaking (average drop diameter vs. time) for solvent extractor with newly developed TT mixing system with uniform energy distribution – A, and for solvent extractor with typical mixing system-B

Another important argument in favor of the new TT system is the more uniform distribution of the dispersed phase along the tank height due to much more intensive mixing in the bulk. The O:A ratio near the overflow of the mixing tank is equal to target value and dictated by the organic and aqueous flow rates. If the mixing in the bulk is not intensive enough, the heavier aqueous phase can have significantly higher concentration near the bottom of the vessel (aqueous lock-up), that may change φ and thus A.

Mass transfer coefficient

A comparative analysis of the effect of energy distribution on the mass transfer was performed for the same average drop diameter of 1 mm for both systems discussed above.

The following assumptions and simplifications were used:

1. In continuous phase the following two main mechanisms of mass transfer work: Danckwerts mechanism of drop surface renewal (mechanism A) and turbulent diffusivity provided by small scale eddy vortices (mechanism B).

2. Following Higbie (8), mass transfer coefficient due to the mechanism A can be estimated as

$$k_{cont,1} = C_1 \sqrt{D_m^{cont} / t_e}, \quad C_1 \approx 1$$

where $t_e = d / u$ is a mean surface renewal time.

3. Following (2), u is a drop slip velocity. However, for the case of liquid – liquid system with comparable viscosities this is an average velocity of the interface that can be derived from the equality of shear stresses from both sides of the interface that in the first approximation gives

$$u \approx d^2 \left(\rho_c - \rho_d \right) g / \left(\mu_c + 2\mu_d \right).$$

4. Following Lamont et al. (9) and Kawase et al. (10) the mass transfer coefficient for the mechanism B can be estimated as

$$k_{cont,2} \approx \sqrt{D_m^{cont}} \left(\varepsilon / V_{cont} \right)^{1/4}$$

5. In the disperse phase the following two mechanisms of mass transfer are dominant: Molecular diffusivity inside the drop enhanced by internal circulation (11) (C) and Danckwerts mechanism (12) of drop surface renewal (D) caused by both drop breaking and drop coalescence acts.
6. Mass transfer coefficient according to the mechanism C can be estimated as

$$k_{disp,1} = C_2 \cdot D_m^{disp} / d, \quad C_2 \approx 10$$

7. Mass transfer coefficient according to the mechanism D can be estimated as

$$k_{disp,2} \approx \sqrt{D_m^{cont} / \tau},$$

where τ is the average time of droplet breakage or coalescence, that was estimated for the TT mixing system with uniform energy distribution as 10s and for the standard mixing system as 100 s, using models of drop breaking and coalescence presented in (4,5).

8. If for two mechanisms coefficients of mass transfer are k_1 and k_2 mass transfer coefficient k in first approximation can be estimated as geometric mean of the average and the maximum

$$k \approx \sqrt{(k_1 + k_2) \max(k_1, k_2)},$$

6

9. In the case of a large excess of extractant in the organic phase, the concentration overall mass transfer coefficient is close to its value in the aqueous phase. To evaluate the effect of the limited excess of extractant we supposed that metal concentrations in organic c_{org} and aqueous c_{aq} phases on the interface satisfy the equation

$$c_{aq} \approx \beta \cdot c_{org},$$

where β is equilibrium coefficient.

This simplification for effective mass transfer coefficient from the aqueous to the organic phase gives the following equations:

$$k_{eff} = \frac{k_{disp}k_{cont}}{k_{disp} + \beta k_{cont}} \quad \text{for aqueous continuous system and} \quad k_{eff} = \frac{\beta k_{disp}k_{cont}}{k_{disp} + \beta k_{cont}} \quad \text{for organic}$$

continuous system.

Results of estimation of volumetric mass transfer coefficients and of stage extraction efficiency using assumptions 1 – 9 are presented in the Table 1. The following properties of liquids and process characteristics were used for the calculations:

- Molecular diffusivities for organic and aqueous phases – $5 \cdot 10^{-9}$ and $1 \cdot 10^{-10}$ m²/s correspondingly
- Kinematic viscosities for organic and aqueous phases – $3 \cdot 10^{-6}$ and $2 \cdot 10^{-6}$ m²/s correspondingly
- Equilibrium coefficient $\beta = 100$ for high and 10 for limited excess of extractant
- Mean residence time in the stage – 30 s
- Volume concentration of disperse phase – $\varphi = 40\%$

Volumetric mass transfer coefficient was calculated as $k_{vol} = k_{eff}\dfrac{6\varphi}{d_{32}}$ and stage extraction

efficiency in continuous flow apparatus (11) as $\dfrac{k_{vol} \cdot T}{k_{vol} \cdot T + 1}$.

Table 1. Estimated values of mass transfer coefficient and stage efficiency for an industrial extraction stage mixer in a Cu SX plant.

Mixing system	Process definition	Volumetric mass transfer coefficient, 1/s	Stage extraction efficiency, %
TT mixing system with uniform energy distribution: $\varepsilon_{max} = 10$ W/kg, $\varepsilon_{avg} = 0.5$ W/kg	Organic continuous, high excess of extractant	0.0554	62.4
	Organic continuous, limited excess of extractant	0.055	62.3
	Aqueous continuous, high excess of extractant	1.05	96.9
	Aqueous continuous, limited excess of extractant	0.294	89.8
Standard mixing system: $\varepsilon_{max} = 40$ W/kg, $\varepsilon_{avg} = 0.1$ W/kg	Organic continuous, high excess of extractant	0.036	52.0
	Organic continuous, limited excess of extractant	0.0359	51.8
	Aqueous continuous, high excess of extractant	0.656	89.8
	Aqueous continuous, limited excess of extractant	0.161	82.8

DISCUSSION

The data presented in Table 1 shows that an increase in turbulence in the main volume in the TT mixing system, and the corresponding acceleration of the coalescence and breakage of droplets, has a positive impact on mass transfer efficiency both in aqueous and organic continuous systems. Note that while in aqueous continuous, improvement of mass transfer is shown to be even more significant. Also note that the described effect of mass transfer acceleration is the most pronounced under conditions of low extractant concentration.
The examples shown above are related to specific process conditions for specific substances with known values of physical properties. In practice the actual values of mass transfer coefficients depend on all of these parameters and also on other factors, like the homogeneity of mixing, the nature and the concentration of the extractant and the interfacial properties of the liquids. Thus the exact values of mass transfer coefficients for both organic continuous and aqueous continuous regimes are dependent on a multitude of factors, however, our conclusions below on the influence of the turbulence will still hold.

CONCLUSIONS

Quantitative evaluation of energy dissipation and energy dissipation rates in different tank zones as a function of tank and impeller geometry as well as models of droplet breakage and coalescence has enabled development of new mixing systems with a more uniform energy distribution.
These new mixing tank and impeller designs significantly improve mass transfer and thus stage efficiency while, at the same time, improving separation and reducing entrainment losses.
Non-uniformity of the turbulent energy distribution is an important factor influencing liquid-liquid mass transfer deserving of further theoretical and practical investigation.

NOMENCLATURE

A – specific mass transfer area, 1/m;
d – drop diameter, m;
d_{32} – average (Sauter) drop diameter, m;
$D_{impeller}$ – impeller diameter, m;
D_{tank} – tank diameter, m;
D_m – molecular diffusivity, m^2/s;
$D_{m,cont}$, $D_{m,disp}$– molecular diffusivities of continuous and disperse phases correspondingly, m^2/s;
g – gravity acceleration, m/s^2;
k – mass transfer coefficient, m/s;
k_{cont} , k_{disp}– mass transfer coefficients, from the side of continuous and disperse phases correspondingly m/s;
k_{vol} – volumetric mass transfer coefficients, 1/s;
t_e – a surface renewal time
T – residence time, s;
$v_{t,baffles}$ – tangential velocity near the baffle; m/s;
$W_{baffles}$ – baffle width, m;
μ_{cont}, μ_{disp} – dynamic viscosities of continuous and disperse phases correspondingly, Pa·s;
ρ_{cont}, ρ_{disp} – dencities of continuous and disperse phases correspondingly, kg/m^3;
ε_{avg} – average energy dissipation rate, w/kg;
ε_{max} – maximum energy dissipation rate, w/kg;

8

φ – volume concentration of disperse phase.

REFFERENCES

1. Braginsky L. N. and Belevitskaya M. A. Chapter 1. Kinetics of drops break-up in agitated vessels. In Liquid—Liquid Systems, Ed. by N.N. Kulov. New York: Nova Science Publishers Inc., 1996. pp. 1–36. ISBN 978-1560721895.

2. Paul E. L., Atiemo-Obeng V. A, Kresta S. M. (Eds.), Handbook of Industrial Mixing: Science and Practice (2004), Print ISBN: 9780471269199.

3. Levich V. G., Physicochemical Hydrodynamics. Prentice Hall,1962. ISBN 978-0136744405.

4. Braginsky, L. N., Kokotov, Y. V. , Influence of Turbulence and Viscosities on The Kinetics of Drop Breaking, Journal of Dispersion Science and Technology, Vol. 14(3), pp. 373-394, 1993.

5. Y. Kokotov, L. Braginsky, Kinetics of Break-Up and Coalescence of Drops in Mixing Vessels. - International Symposium on Liquid-Liquid Two Phase Flow and Transport Phenomena, Antalya, Turkey, pp.567-574, 1997.

6. Braginsky, L. N., Kokotov, Y. V., The Effect of Coalescence on Drop Size in Turbulent Flow, World Congress On Particle Technology, Vol. 3, 1998

7. Kokotov Y., Braginsky L., Shteinman D., Slonim E., Barfield V., Grinbaum B., Turbulent Settling (Ts) Technology For Solvent Extraction, ALTA 2010

8. Higbie, R., The Rate of Absorption of a Pure Gas Into a Still Liquid During Short Periods of Exposure. Trans. Am. Inst. Chem. Eng. Vol. 35, pp. 36-60, 1935

9. Lamont J.C., Scott, D.S., An Eddy Cell Model of Mass Transfer Into Surface of a Turbulent Liquid. AIChE J., Vol. 16. pp. 513-519, 1970

10. Kawase, Y., Halard, B., Moo-Yang, M., Theoretical Prediction of Volumetric Mass Transfer Coefficient in Bubble Columns for Newtonian and Non – Newtonian Fluids. Chemical Engineering Science, Vol. 42, pp. 1609-1617, 1987

11. Treybal, R. E., Estimation of the Stage Efficiency of Simple, Agitated Vessels Used in Mixer-Settler Extractors. AIChE J., Vol. 4(2), pp. 202-207, 1958

12. Danckwerts P.V., Significance of Liquid Film Coefficients in Gas Absorption, Ind. & Engng. Chem., Vol. 43(6), pp. 1460-1467, 1951