

Surfactant effects on alpha factors in full-scale wastewater aeration systems

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Abstract Aeration is an essential process in the majority of wastewater treatment processes, and accounts for the largest fraction of plant energy costs. Aeration systems can achieve wastewater oxygenation by shearing the surface (surface aerators) or releasing bubbles at the bottom of the tank (coarse- or fine-bubble aerators). Surfactants accumulate on gas–liquid interfaces and reduce mass transfer rates. This reduction in general is larger for fine-bubble aerators. This study was conducted to evaluate mass transfer effects on the characterization and specification of aeration systems in clean and process water conditions. Tests at different interfacial turbulence regimes were analysed, showing higher gas transfer depression for lower turbulence regimes. Higher turbulence regimes can offset contamination effects, at the expense of operating efficiency. This phenomenon is characteristic of surface aerators and coarse bubble diffusers and is here discussed. The results explain the variability of α factors measured at small scale, due to uncontrolled energy density. Results are also reported in dimensionless empirical correlations that describe mass transfer as a function of physiochemical and geometrical characteristics of the aeration process.

Keywords Aeration; alpha factor; efficiency; gas–liquid interface; oxygen transfer; surfactant

Introduction

Aeration is an essential process in the majority of wastewater treatment processes, namely in the aerobic biological treatment. The aeration process accounts for the largest fraction of plant energy costs, ranging from 45 to 75% of the plant energy expenditure on a daily basis (Reardon, 1995; Rosso and Stenstrom, 2005). Aeration systems transfer oxygen into a liquid medium by either diffusing gas through a gas–liquid interface, or dissolving gas into the liquid solution using a semi-permeable membrane. Environmental technologies usually rely on interfacial gas transfer, with a gas–liquid interface created by either shearing the liquid surface with a mixer or turbine, or by releasing air through spargers or porous materials. Surface aerators shear the wastewater surface producing a spray of fine droplets that land on the wastewater surface within a few seconds and over a radius of a few metres. Diffusers are nozzles or porous surfaces of various geometry placed on the tank bottom that release bubbles that travel towards the tank surface. In general, bubbles are considered coarse when their diameter exceeds 50 mm, and fine when diameters are less than 5 mm. Falling droplets and rising coarse bubbles have large interfacial gas–liquid velocity gradients and can be grouped as high flow regime interfaces, whereas fine bubbles have low interfacial velocity gradients and can be grouped as low flow regime interfaces.

Fine pore diffusers have become the most common aeration technology in wastewater treatment in the OECD (Organisation for Economic Co-operation and Development) Countries. They have higher efficiencies per unit energy consumed (Standard aeration efficiency or SAE, $\text{kgO}_2 \cdot \text{kWh}^{-1}$). The impact of contamination on aeration performance is usually quantified by the α factor (ratio of process water to clean water mass transfer coefficients, or $K_{La_{pw}}/K_{La_{cw}}$). Lower flow regime gas–liquid interfaces (such as the ones

produced by fine-pore diffusers) generally have lower α factors than higher flow regime interfaces (such as the ones produced by coarse bubble diffusers or surface aerators) for similar conditions (Stenstrom and Gilbert, 1981). Differences in α factors amongst aeration systems were noted as early as in the 1930s (Kessener and Ribbius, 1935), but were generally forgotten until the energy crisis of the 1970s increased the awareness for energy efficient technologies. Prior to the 1980s, many plants were designed with α factors of 0.8, which was considered as a “universal” α factor for all types of aeration systems. It has been shown that different aeration methods have different α factors, and for fine-pore diffusers the initial α factor decreases over time in operation due to fouling or scaling (Rosso and Stenstrom, 2006). Furthermore, for fine-bubble systems the α factor is a function of process conditions such as the mean cell retention time (MCRT) or the airflow rate (Rosso *et al.*, 2005).

Mass transfer depression caused by contaminants has long been observed (Mancy and Okun, 1960). Eckenfelder and Barnhart (1961) reported the effects of organic substances on mass transfer, showing that contamination as low as 15 mg/l of sodium lauryl sulphate can reduce mass transfer coefficients to 0.5 times the value in clean water. The effects of wastewater contamination on mass transfer can be related to the decrease in dynamic surface tension (McKeown and Okun, 1961; Masutani and Stenstrom, 1991). The interfacial accumulation of surfactants causes an increase in interfacial rigidity (hence in the drag coefficient), the reduction of internal gas circulation, and the reduction of interfacial renewal rates. These gas–liquid interfacial phenomena have been recently discussed by the authors (Rosso *et al.*, 2006 submitted).

There exists a continuing debate over the reasons for different α factors for different aeration methods and in some cases there are claims that fine pore diffusers from different manufacturers have different α factors. Part of the difficulty is the reliance on different experimental methods to measure α factors, and the variations in the range of interfacial flow regimes.

The goal of this paper is to analyse the effects of contamination on gas–liquid interfaces due to surfactants, and compare it for the two cases of high and low interfacial velocities. In wastewater treatment, these regimes correspond respectively to: 1. coarse-bubble aerators and surface aerators; 2. fine-pore diffusers. This paper summarizes previously observed phenomena, and explains the microscopic phenomena that cause different contamination effects at different flow regimes. The results of two laboratory studies using dynamic surface tension measurements and several other laboratory and full scale studies are used to verify the conclusions.

Methods

Aeration tests were performed in a well-mixed vessel, according to the standard procedure by the American Society of Civil Engineers (ASCE, 1984, 1991). In these tests, measurements of dissolved oxygen (DO) are recorded over time in a solution where oxygen is first sequestered with sodium sulphite, using cobalt chloride as a catalyzer. DO measurements were taken with a YSI Model 58 DO meter (Yellow Springs Instruments, Yellow Springs, OH), and recorded *via* a NI DAQ-6015 (National Instruments, Austin, TX) acquisition board connected to a computer. For each measurement, airflow rate and dynamic wet pressure (i.e. the headloss across the aerator) were also collected. Photographs of the process at high (1/500 s) and low (1/60 s, 1/125 s) shutter speed were taken. These were used to calculate bubble mean diameters and rising velocities, as well as to visually observe the process. The mass transfer coefficient can be calculated by integrating over time the differential mass balance on the time-dependent DO concentration c with upper integration boundary the equilibrium concentration

at saturation c_{sat} . The ASCE DO Parameter Estimation Program (DO_PAR 1.08; ASCE 1997; downloaded from: <http://fields.seas.ucla.edu/research/dopar/>) was utilized to calculate volumetric mass transfer coefficients ($k_L a$, time^{-1}), oxygen transfer rates in standard conditions (SOTR, $\text{mass}_{\text{O}_2} \cdot \text{time}^{-1}$), oxygen transfer efficiency in standard conditions for clean water (SOTE, %), and oxygen transfer efficiency for surfactant solutions (α SOTE, %). The α factor can be calculated as the ratio $\alpha\text{SOTE}/\text{SOTE}$. Additional, previous investigations from our laboratory were also included (Masutani and Stenstrom, 1991; Rosso *et al.*, 2006 submitted).

A dimensional analysis was performed on the datasets, using the method based on the procedure described in detail by Zlokarnik (2002). The physical quantities included in the analysis were: bubble life t_B , mass transfer coefficient $k_L a(t_B)$, bubble diameter $d_B(t_B)$, orifice diameter d_o , dynamic surface tension $\gamma(t_B)$, weight (i.e. buoyancy) of the bubble $g\Delta\rho$, surfactant bulk concentration $c_B(t_B)$, oxygen interfacial diffusivity $D_{s,\text{O}_2}(t_B)$, surfactant bulk diffusivity $D_{B,\text{SAA}}$, and airflow rate Q . The gas-side (oxygen) interfacial diffusivity was calculated by using Higbie's penetration theory:

$$k_L = 1.13 \cdot \sqrt{\frac{u_B}{d_B}} \cdot D_{s,\text{O}_2} \quad (1)$$

where k_L , u_B , and d_B are the velocity of adsorption, the interfacial gas–liquid velocity (or bubble velocity), and the bubble diameter, respectively. For verification, the liquid-side (surfactant) interfacial diffusivity was calculated by integrating a Ward–Tordai interfacial accumulation equation at short-time limits, as fully described elsewhere (Rosso *et al.*, 2006 submitted). Due to the low shear rate of this process, the liquid viscosity has no influence. Highly viscous systems need a different dimensional analysis, because the transport phenomena occurring in those systems are due to different transport mechanisms.

From the dimensional matrix we generated six independent dimensional numbers, and by taking their combinations we can fully define the problem with the pi-set:

$$(\text{Sh}) = f\{(\text{Pe}), (\text{Ro})\} \quad (2)$$

where:

$$(\text{Sh}) = \frac{k_L a \cdot d_B^2}{D_{s,\text{O}_2}} = \text{Sherwood number}$$

$$(\text{Pe}) = \frac{Q}{d \cdot D_{s,\text{O}_2}} = \text{Péclet number} = (\text{Re}) \cdot \frac{\nu}{D_{s,\text{O}_2}}$$

$$(\text{Ro}) = \frac{c_B \cdot t_B \cdot D_{B,\text{SAA}}^2}{\gamma \cdot d_B^3} = \text{time-dependent surface contamination number}$$

The physical significance of the dimensionless numbers is: (Sh) = ratio of mass diffusion forces to molecular diffusion forces; (Pe) = ratio of advective transport forces to molecular diffusion forces; (Ro) = ratio of surface contamination to surface tension forces. (Re) is the Reynolds number, and ν the viscosity. The datasets were analysed with nonlinear regression methods using SYSTAT 10 (SPSS Corporation, Chicago, IL).

Results and discussion

Figure 1 shows the evolution of mass transfer coefficients with increasing time. As time elapses, surfactant molecules migrate towards the surface, land, and hinder the interfacial renewal process, resulting in a steep and rapid decrease in value for the volumetric mass

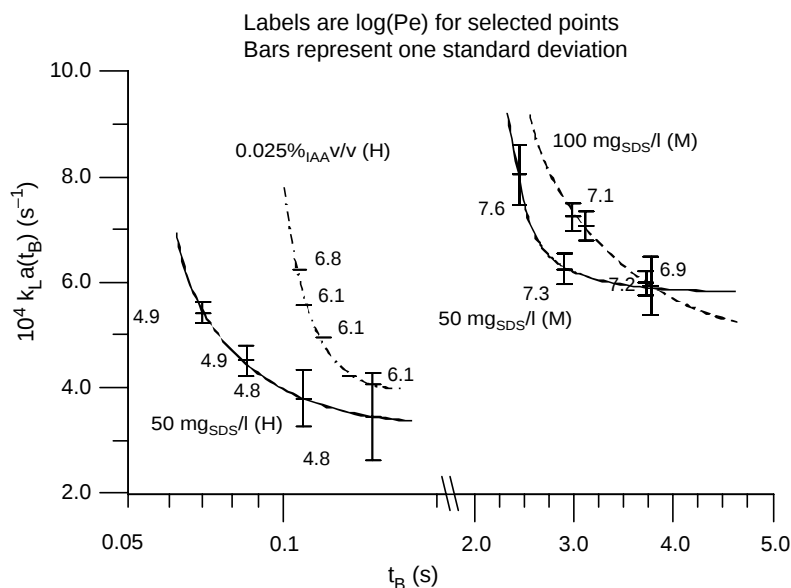


Figure 1 Evolution of mass transfer coefficients over increasing bubble surface age t_B . Labels are interfacial $\log(\text{Pe})$ for selected points. (H: data from Rosso *et al.*, submitted; M: data from Masutani and Stenstrom, 1991)

transfer coefficient $k_L a$. The volumetric mass transfer coefficient is the product of the velocity of reaction k_L and the interfacial area (a). Since bubbles are stable at smaller diameters in contaminated waters, the interfacial area is higher. Also, with increasing time, bubbles are losing mass by transferring oxygen to the liquid, hence their volume decreases and their specific area increases. Yet, the volumetric mass transfer coefficient $k_L a$ decreases, therefore k_L must decrease more rapidly than the interfacial area (a) increases. Eckenfelder and Barnhart (1961) observed that k_L decreases with increasing surfactant concentrations, but $k_L a$ is minimum at low surfactant concentrations, and partially recovers at higher concentrations. This is the effect of a larger interfacial area (a), but reduced velocity of reaction k_L .

In the case of moving interfaces, which is the case for diffused and surface aerators, turbulence exists behind the interfacial laminar films. At a given interfacial flow regime, hence at a given interfacial (Re), the film renewal rate is decreased with increasing contamination. This is observed in lower internal gas circulation rates in bubbles (Garner and Hammerton, 1954). Roy and Duke (2004) photographed two-dimensional DO gradients near surfaces contaminated with Triton X-100, using a laser-induced fluorescence technique. Their photographs show reduced circulation outside contaminated bubbles, with higher interfacial O_2 concentration gradients for higher contaminations.

The intuitive concept of “molecular obstruction” is usually considered the cause of mass transfer depression (Mancy and Barlage, 1968). This phenomenon is dominant for stagnant gas–liquid interfaces, where the interfacial fluid velocity is zero. In this case, molecular diffusion through the stagnant film is the only transport mechanism. In case of moving interfaces, turbulent transport towards the interface is the driving force for mass transfer, for two reasons: interfacial renewal rates and actual area covered by the surfactant molecules.

Surface tension is inversely proportional to the boundary layer thickness (Azbel, 1981). Higher boundary layer thicknesses create lower probability for turbulent eddies to reach the interface and carry “fresh” gas packets from the bulk, i.e. resulting in lower

renewal rates. Figure 1 shows different mass transfer coefficients at the same contaminations ($50 \text{ mg}_{\text{SDS}}/\text{l}$). Higher mass transfer coefficients result from higher interfacial velocities [characterized by higher (Pe)]. The same contaminations can yield different mass transfer coefficients in different flow regimes; therefore, we must conclude that the molecular obstruction phenomenon has a negligible effect on this mass transfer process.

Another consideration about the molecular obstruction theory is the effective diffusional area available for O_2 molecules to travel across the interface. Oxygen is present within the bubble at 20.95% v/v concentrations, while the surfactant accumulates on the bubble surface in much lower quantities. Furthermore, due to the nature of the surfactant, its polar head is anchored to the interface, and the hydrophobic tail is fluttering into the bubble volume. The molecular diameter for an oxygen molecule does not exceed 0.3 nm, while the diameter of the surfactant head is on the order of 10 nm or more. The frontal diameter to account for the surfactant is the diameter of the polar head, as the hydrophobic tails are fluttering inside the bubble, where gas molecules are free to move by virtue of molecular diffusion. If the molecular obstruction phenomena were dominant, reduced transfer should also occur for salts and certain aliphatic alcohols, which show mass transfer enhancement (i.e. $\alpha > 1$), instead of depression (Zlokarnik, 1980a). For the range of surfactant accumulation of our tests, the number of O_2 molecules far exceeds the number of surfactant molecules. The interfacial accumulation of surfactant molecules is different at different angles from the front stagnation point (i.e. the highest point of the rising bubble), and the ratio between fore and aft accumulation was recorded to vary between 10 to 100 times at $(\text{Pe}) = 10^5$ (Ramirez and Davis, 1999). The interfacial ratio of surfactant to oxygen molecules for our tests was calculated and exceeded 1/100 on an average around the bubble and 1/1,000 on the upper cap of the bubble. If molecular obstruction is assumed as dominant, it should be the same for both high and low interfacial velocities. Nevertheless, mass transfer depression has a higher magnitude at lower interfacial velocities, confirming that molecular obstruction is negligible for flowing systems.

Figure 2 shows a photograph at $1/500''$ of coarse and fine bubbles in a $50 \text{ mg}_{\text{SDS}}/\text{l}$ solution, and the inner fluid dynamic patterns. These photographs were used to calculate the bubble mean diameter. It is visible that the large majority of bubbles have a diameter lower than 1 mm. Clean water tests without surfactant in analogous conditions yielded bubble mean diameters of 4 mm, larger than any bubble in the left half of Figure 2. One fine and one coarse bubble are highlighted and their interior circulation pattern is sketched to the side. The accumulation of surfactant at the fine-bubble interface occurs to

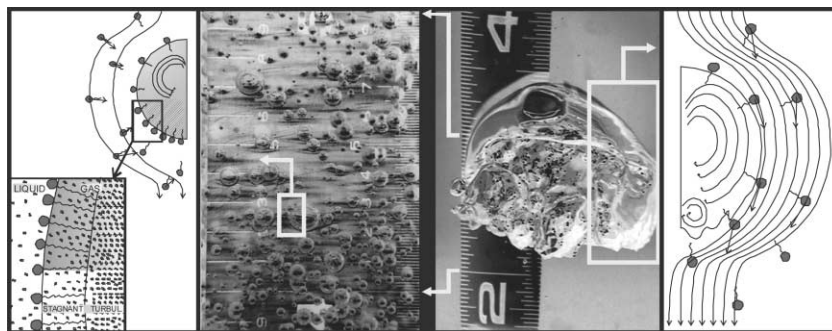


Figure 2 Comparison of fine (left photograph) and coarse (right photograph) bubbles generated by two different aerators operating at same airflow rate in the same surfactant solution ($50 \text{ mg}_{\text{SDS}}/\text{l}$). The fine-pore aerator is a $150 \times 150 \text{ mm}$ ($6 \times 6 \text{ in}$) panel mounting a plastic sintered membrane with mean porosity of $9 \mu\text{m}$. The coarse bubble aerator is a 9.53 mm ($3/8 \text{ in}$) air nozzle. Scale units are inches ($1 \text{ in} = 25.4 \text{ mm}$)

a larger extent than for coarse bubbles, as the hydraulic residence time is higher, and surfactant molecules have a longer time available for migration towards the interface. Also, smaller bubbles have a much lower interfacial velocity, and once the surfactants have attached to the surface, their hydrophobic tails inside the bubble reduce the internal gas circulation, acting like a baffle in a stirred reactor. Surface active agents tend to accumulate at the bottom of the bubble, creating a stagnation zone inside the bubble, referred to as a stagnant cap. Evidence supports the existence of fore-and-aft symmetry in the interfacial concentration distribution (Clift *et al.*, 1978; Ramirez and Davis, 1999).

Additional photographs were taken at 1/60 s and used to calculate the bubble mean rising velocity, which is approximately 0.2 m/s for fine bubbles and 1.5 m/s for coarse bubbles in Figure 2. With large interfacial area and large mass transfer time, fine bubbles should have a very high $k_L a$. Yet, mass transfer coefficients in surfactant solutions are smaller than in clean water, although bubble diameters are larger in clean water (i.e. smaller specific interfacial area). Therefore, mass transfer times in clean water are reduced due to greater rise velocities. For coarse bubbles the differences between clean and process water transfer rates are reduced, due to higher interfacial velocity and higher rate of turbulence.

Figure 3 shows the decline in α factors with increasing time for selected measurements. The horizontal axis on this graph shows the mean bubble age (t_B). The rate of decline in α factor is greater for surfactants with greater molecular diffusivity. Conversely, higher molecular weight surfactants, which generally have lower diffusivity, require greater time to migrate towards the bubble interface, resulting in less depression of the α factors for the same mean bubble age. To avoid flow effects on α , the measurements in Figure 3 were performed at the same interfacial velocities.

Figure 4 shows the results for selected time-dependent tests in dimensionless form. The surfactant solution for this case was SDS in concentration of 50 mg/l, and the interfacial flow velocity was in the high range for fine bubbles in our datasets ($u_B \sim 1.68 \cdot 10^{-1}$ to $2.58 \cdot 10^{-1}$ m/s). The dimensionless time scale t^+ represents short bubble lives, which in dimensional terms range from 2.46 to 3.78 s. The evolution of mass transfer with increasing time is plotted as $(Sh) = f(t^+)$. After a steep decrease, the value of (Sh) tends to a minimum,

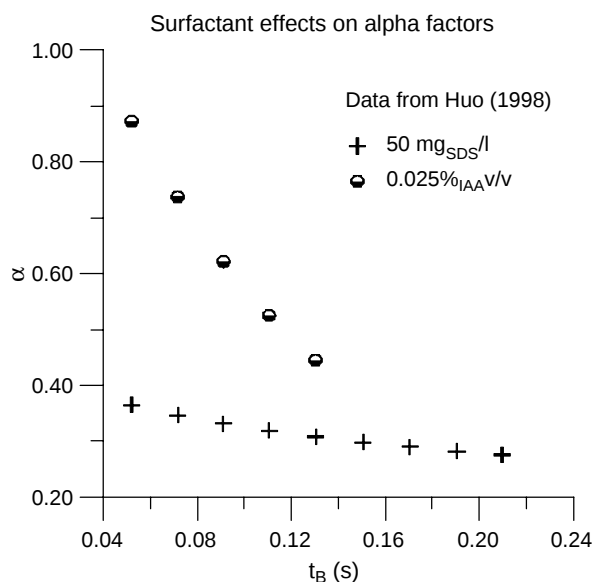


Figure 3 Evolution of α factors with increasing bubble surface age t_B

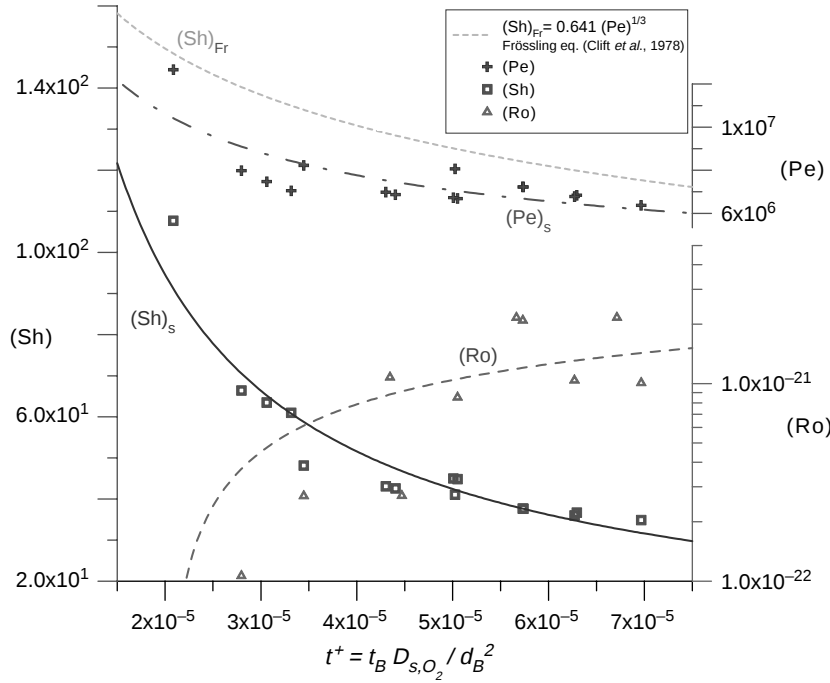


Figure 4 Dimensionless characterization of mass-transfer phenomena at a fine bubble interface at short bubble lives. (Sh), (Pe), and (Ro) are the interfacial Sherwood, Péclet, and time-dependent contamination numbers, respectively. (Sh)_{Fr} is calculated with the Frössling equation for clean systems, which predicts well at high interfacial velocities. Trendlines are power best-fits

which is reached at surface equilibrium. Concurrently, (Pe) decreases rapidly at first, and reaches a plateau earlier than (Sh). This is because while (Pe) is fairly constant ($t^+ > 3 \cdot 10^{-5}$), the contamination number (Ro) is still increasing. (Ro) represents surface contamination, which for this specific case is not yet completed at $t \sim 3$ s. In dimensionless terms, (Ro) reaches a plateau for $t^+ > 5 \cdot 10^{-5}$ (for sake of clarity (Ro) is plotted in log-linear scale). In this upper range, (Sh) will also reach a plateau, since (Pe) is already quasi-constant. Note that the values of (Pe) may seem higher than expected because they refer to the interface, and not the bulk of the liquid, where diffusivities are constant and at their equilibrium values.

The equation resulting from the dimensional analysis is:

$$(Sh) = \frac{0.382 \cdot (Pe)^{1/3}}{1 + \log [1 + 10^{19} \cdot (Ro)]} \quad (3)$$

Equation 3 is statistically significant ($R^2 > 0.8$; $t > 10$; $P < 10^{-3}$; unbiased residuals), and represents a modification for the case of contaminated solutions of the well known Frössling equation (Frössling, 1938). The bubbles in this study behave as solid spheres, as the exponent of (Pe) is 1/3 (Clift et al., 1978).

Figure 5 shows the evolution of α factors with increasing Reynolds numbers (Re) and compares them to other types of aeration. For fine-pore diffusers, (Re) was calculated using as equivalent diameter the diameter of the bubble column, accounting for the gas hold-up fraction. Coarse-bubble experiments were performed with a single nozzle, and the equivalent diameter for the water column was calculated using the coarse bubble diameter as length-scale.

Surfactant effects on alpha factors

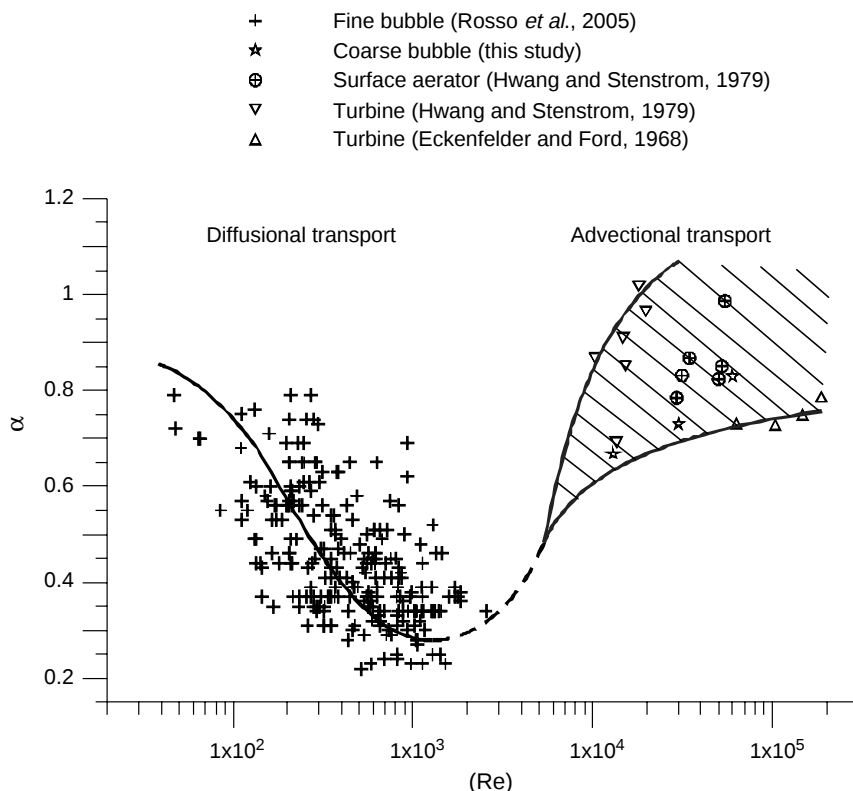


Figure 5 α factors at different interfacial flow regimes. The low (Re) zone is the domain of fine bubble aerators, while higher (Re) correspond to coarse-bubble, jets, surface mixers, and turbines. The shaded area represents a family of trendlines parametric in the energy density

Data from Eckenfelder and Ford (1968) are included in Figure 5. These data show α factors for turbine aerators at different rotation regimes. For these data, (Re) were calculated as the Reynolds number of a droplet 5 mm in diameter travelling in the air at a distance of 5 metres from the turbine shaft. The trendline in Figure 5 confirms the behaviour predicted by Eckenfelder and Ford (1968). The dashed portion of the trendline shows an interfacial flow regime that is usually not encountered in aeration systems. Viscous-flow domains [with (Re) < 10] are usually not encountered in wastewater practice, with the possible exception of membrane bioreactors (Wagner *et al.*, 2002). In this region, there is practically no effect of flow on α factors, as the resistance to gas-transfer is due primarily to the transport within the bulk of the liquid. In the transitional-flow domain, where fine-bubble aerators operate, an increase in (Re) results in a decline in α . This occurs because the gas-transfer process is controlled by the surfactant interfacial migration, which has higher weight at higher (Re) within this region [50 < (Re) < 5,000]. The data plotted in Figure 4 belongs to this region.

Coarse-bubble diffusers, jets, and surface aerators have a flow regime in the inertial-flow domain. For these gas-liquid interfaces, the higher range of (Re) results in higher interfacial renewal rates, consequently in lower interfacial surfactant accumulation. The trendline shows that at very high (Re), α can be restored to clean water values and, if the energy density is sufficient, values of α larger than 1.0 are possible (Zlokarnik, 1980a, b). Values of α above 1.0 mean higher mass transfer coefficients in process water than in

clean water. The reasons for the higher α factors include ionic interfacial effects due to the presence of inorganic salts, such as in seawater, or of some concentrated aliphatic alcohols (Zlokarnik, 1980a). Another reason is the entrainment of foam created by surfactants. Such foams are either ephemeral or do not exist without surfactants. These foams can be stable and composed of fine bubble aggregates with high interfacial areas (Fan and Tsuchiya, 1990). Operations at these conditions have low aeration efficiency, since values of α approaching or higher than 1.0 may be reached at the sole expense of additional energy supplied to the system.

At extremely high energy densities, corresponding to $(Re) > 10^5$, the higher interfacial shear rate reduces the thickness of the laminar film. At the upper limit, the interfacial resistant film will reduce its thickness to a few layers of water molecules, and the phenomenon of liquid vaporisation will be dominant. This occurs when turbines or pumps cavitate at high rotational speeds. The cross-hatched area in Figure 5 corresponds to a region where mass transfer is affected by the local energy density. In this region, (Re) and the energy density may be independent and produce different mass transfer rates. For example, the results from Hwang and Stenstrom (1979) were collected in a small vessel with higher energy density while the results from Eckenfelder and Ford (1968) were collected in a large vessel. This phenomenon creates a range of α measurements using experimental setups where (Re) and energy density are not controlled (Bass and Shell, 1977; Stukenberg et al., 1977).

Recommendations for scale-up

Aeration systems for biological environmental processes are traditionally designed relying upon data collected from re-aeration tests in pure water. The scale-up factors are commonly left to the designer's experience. However, this has resulted in frequent over-estimation of performance. Pure water re-aeration tests are characterized by bubbles belonging to the fluid-sphere regime, while process water tests can be characterized by both the fluid- and solid-sphere regimes. Bubble diameter, gas flowrate, and contaminant quantitative and qualitative characteristics determine the transition between the two regimes.

Our aeration experiment, as shown in Figure 3, was performed using a full-scale diffuser and confirms that the similitude between the laboratory- and full-scale systems is maintained.

Figures 3 and 4 show the celerity of gas transfer depression operated by surfactants. In Figure 5 the departure between clean and process water behaviour begins at bubble formation and reaches a steady value before bubble detachment. In aerated biological processes such as sequencing batch reactors, PFR, or CSTR reactors, the average hydraulic retention time for a fine bubble is in the order of 20–80 s. This reduces the time of bubble formation to a negligible fraction of the total bubble residence time. Therefore, the gas transferred in the bubble formation process is a negligible fraction of the total gas transferred.

When sizing aeration systems for contaminated liquids it is necessary to take the corrected $k_L a$ value that can be calculated from Equation 3, and not the Frössling solution. Design engineers typically utilize α as a scaling factor to account for process conditions. Our results can be expressed in terms of α :

$$\alpha = \frac{(Pe)^{2/3}}{1 + \log [1 + 10^{19} \cdot (Ro)]} \quad (4)$$

Equation 4 is the ratio of Equation 3 and the interfacial Frössling equation for clean water (see legend in Figure 4). Note that the flow effects do not cancel, as in our flow

regime the bubbles behave as solid spheres but the Frössling prediction is for fluid spheres. For fine-pore aerators, α factors typically range from 0.2 to 0.8, depending on the contaminant characteristics. This implies a design over sizing of 1.25 to 5 times the aeration capacity in pure water. The choice of α factors for biological process design is crucial for overall efficiency and economic viability of the aeration process (Rosso *et al.*, 2005).

Conclusions

This paper has shown that the turbulence regime controls the depression of gas transfer caused by surface active agents, which is often quantified by an empirical ratio, called the α factor. The turbulence regime can be characterized by the interfacial Reynolds or Péclet numbers and the mass transfer by the interfacial Sherwood number. The Sherwood number can be correlated with the Péclet number to account for turbulence and to a dimensionless number, called the interfacial contamination number, to account for interfacial contaminant accumulation. The final correlation agrees with observed transfer rates for both time-dependent and time-integrated mass transfer measurements.

The results explain why fine bubble diffusers have greater mass transfer depression than coarse bubble or surface aerators. The turbulence associated with coarse bubble aerators makes them behave more like surface aerators than fine bubble aerators. High turbulence regime aerators may achieve better transfer rates, but at the expense of greater energy density, and lower aeration efficiency (mass transfer per unit of power). The results also describe the variability of α factors measured at small scale, due to uncontrolled energy density.

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