

Gas-liquid Transfer

14.1 GAS TRANSFER

By gas transfer is meant the mass movement of gas to or from solution, which occurs when a liquid surface is in contact with a gas phase with which it is not in equilibrium. In chemical engineering practice the uptake of gas by a liquid is generally termed 'absorption' while the removal of gas from a liquid is usually termed 'desorption' or 'stripping'.

In water and wastewater engineering practice the gas phase is usually air and hence gas transfer is generally termed 'aeration'. The gases of interest to the water and wastewater engineer are O_2 , CO_2 , CH_4 , H_2S , NH_3 , and Cl_2 . The driving force that activates the transfer of these gases to and from water is the difference between their concentration in solution and their solubility under the prevailing conditions.

14.2 GAS SOLUBILITY

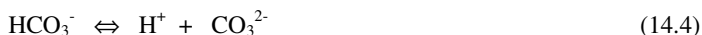
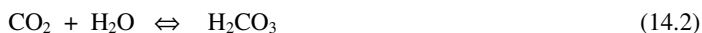
The main factors that influence gas solubility in water are: the water temperature, the partial pressure of the gas in the gas phase, the dissolved solids concentration in the water phase and the chemical nature of the gas.

The solubility of gases, unlike that of solids, decreases with increase in temperature. For partial pressures up to 1 atm and for gases that do not react with water to any great extent, the equilibrium concentration of gas in solution at a given temperature is proportional to the partial pressure of the gas in contact with the water in accordance with Henry's Law:

$$c_s = Hp \quad (14.1)$$

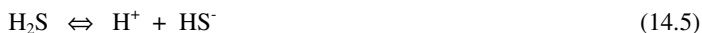
where c_s ($mg\ l^{-1}$) is the saturation or equilibrium concentration of the gas in solution, p (atm) is the partial pressure of the gas in the gas phase in contact with the water and H is the solubility coefficient. Henry's Law applies to most of the gases of interest in water and wastewater engineering including O_2 , CH_4 , CO_2 , H_2S . The latter two gases undergo reaction in solution.

Dissolved CO_2 reacts with water as follows:



For the conditions normally encountered in water engineering the concentration of H_2CO_3 will not exceed 1% of the CO_2 concentration (refer Chapter 1 for discussion of the role of the carbonate system in water).

Hydrogen sulphide reacts in solution as follows:



It is apparent from equations (14.5) and (14.6) that the dissolved form of H_2S depends on the pH of the solution.

Ammonia (NH₃) and chlorine (Cl₂) are highly soluble gases that readily react with water. Their pressure-solubility relationships deviate from Henry's Law.

Water solubility data for the above gases are presented in Tables 1.7, 1.8 and 1.9 in Chapter 1. It is clear from the tabulated data that the gases which react with water, including CO₂, H₂S, NH₃ and Cl₂, are considerably more soluble than the remaining non-reactive gases.

14.3 MECHANISM AND RATE OF GAS TRANSFER

When a water surface is exposed to a poorly soluble gas with which it is not in equilibrium, it is assumed that the water at the interface becomes instantaneously saturated with the gas and that the gas is transported into the body of the liquid by the process of molecular diffusion:

$$\frac{\partial m}{\partial t} = -D \frac{\partial c}{\partial x} \quad (14.7)$$

where $\partial m/\partial t$ is the rate of gas transport across unit area of interface, D is the coefficient of molecular diffusion and $\partial c/\partial x$ is the concentration gradient normal to the interface. The simplest physical model of interface conditions is that proposed by Whitman and Lewis (1924), shown in Fig 14.1. They assumed that the resistance to gas transfer resided in fixed gas and liquid films at the gas-liquid interface. The movement of gas across the gas film implies the existence of a pressure gradient in the gas film and hence the gas pressure P_i at the interface is less than the bulk gas pressure P_g . However, for gas of low solubility the rate of transfer of gas to the liquid phase is slow and hence the pressure gradient in the gas phase is negligibly small.

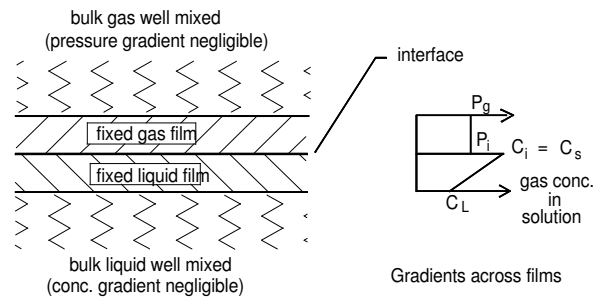


Fig 14.1 Two-film gas transfer model

Whitman and Lewis assumed a linear concentration gradient across the liquid film:

$$\frac{\partial c}{\partial x} = \frac{C_L - C_s}{h} \quad (14.8)$$

where C_s is the saturation or interface concentration of the gas in solution, and C_L is the gas concentration in the bulk liquid (assumed to be well-mixed). The rate of transfer across a gas-liquid interface of area A is calculated from equations (14.7) and (14.8) to be

$$A \frac{\partial m}{\partial t} = \frac{AD}{h} (C_L - C_s) \quad (14.9)$$

Dividing by the liquid volume V :

$$\frac{A}{V} \frac{\partial m}{\partial t} = -\frac{AD}{Vh} (C_L - C_s)$$

Hence

$$\frac{dc}{dt} = -a \frac{D}{h} (C_L - C_s) \quad (14.10)$$

or

$$\frac{dc}{dt} = -k_L a (C_L - C_s) \quad (14.11)$$

where $a = A/V$, k_L is called the 'liquid film coefficient' (also called the 'absorption coefficient') and the product $k_L a$ is the 'overall transfer coefficient'. Although it can be verified experimentally that an equation such as (14.11) accurately represents gas-transfer kinetics in systems of the O_2 /water and CO_2 /water types, it is unlikely that the simple liquid film model, on which the equation is based, correctly represents actual conditions at the interface. In the aeration systems used in water and wastewater treatment processes, large areas of air-water interface are created and continually renewed. When an element of interface is created, the initial rate of gas transfer to or from a non-equilibrated liquid is high, but is reduced as the concentration gradient reduces (equation 14.7). Thus the more rapidly an element of interface is renewed the higher will be the mean rate of gas transfer through it. Also, since gas concentration gradients in the bulk liquid phase are generally very small, indicating little resistance to gas transport within the bulk liquid phase, it is apparent that resistance to gas transfer resides in the organised water at the interface. It is probable that the effective thickness of this high-resistance film is influenced by the level of turbulence in the bulk liquid. Thus the overall transfer coefficient, which is a constant for a given system, is dependent on the total area of air-water interface created, the rate of interface renewal, the level of turbulence in the liquid and temperature.

14.4 OXYGEN TRANSFER

The main function of aeration in wastewater treatment is the solution of oxygen in the mixed liquor of the activated sludge process. In the activated sludge process, as in other similar biochemical processes, the rate of oxygen input to the system must balance the rate of oxygen uptake, since, owing to its low solubility, the storage of oxygen in the system is negligible. Aerators are also sometimes used in polluted rivers and lakes to prevent excessive oxygen depletion.

Aerators for oxygen transfer are rated in terms of oxygenation capacity (OC), which is defined as the rate at which an aerator transfers oxygen to clean water under standard conditions, i.e. at a temperature of 20 °C (or 10 °C), an atmospheric pressure of 1 atm and a zero concentration of oxygen in solution. Under these conditions, equation (14.11) becomes:

$$\frac{dc}{dt} = -k_L a C_{*(20)} \quad (14.12)$$

where $C_{*(20)}$ is the equilibrium saturation concentration of oxygen in the water being aerated, at a temperature of 20 °C. It should be noted that $C_{*(20)}$ may differ from $C_{s(20)}$, the saturation concentration for a water in equilibrium with atmospheric air. This is particularly the case for dispersed air systems with submerged diffusers, where $C_{*(20)}$ typically exceeds $C_{s(20)}$. With surface aeration systems, $C_{*(20)}$ may be taken as equal to $C_{s(20)}$ for practical computational purposes. Since

$$OC = V \frac{dc}{dt} \quad (14.13)$$

then

$$OC = k_L a C_{*(20)} V \quad (14.14)$$

where V is the liquid volume.

The conditions under which aerators operate in the activated sludge process are quite different from the standard conditions to which their OC-values relate. The more significant differences are the presence

of solids in suspension, the presence of surfactants in solution and the temperature difference. These factors may influence both the values of $k_L a$ and C_s . Their effects on these parameters are taken into account by the introduction of the empirical coefficients α and β , which are defined as follows:

$$\alpha = \frac{k_L a \text{ in suspension}}{k_L a \text{ in tapwater}} \quad (14.15)$$

$$\beta = \frac{C_s \text{ in suspension}}{C_s \text{ in tapwater}} \quad (14.16)$$

The concentration of suspended solids in the mixed liquor of activated sludge plants is usually within the range 1000-6000 mg l⁻¹. Since suspended solids may alter the bulk and surface properties of the suspending medium, they are likely to affect gas transfer kinetics (Karmo, 1972). Suspended solids may effect an increase in viscosity, which tends to reduce the level of turbulence and rate of surface renewal. Particles that have a preferential affinity to adsorb on the water surface may influence surface hydrodynamics. On the credit side, air bubbles may become attached to the solids in suspension and be carried down into the bulk liquid.

Surface-active agents are present in most wastewaters treated by the activated sludge process and for this reason their influence on oxygen transfer has been widely studied. By reducing surface tension, they effect an increase in the diameter of spherical bubbles and a corresponding increase in the air-water interfacial area per unit volume of air. On the negative side, however, they tend to attach to the bubble surface, where they constitute a barrier to oxygen transfer. The latter effect is more marked in diffused air systems, where the rising bubbles have a sufficiently long residence time for this layer to become established.

Test results show that the operational α -value in activated sludge processes is typically less than unity. It is well established (Barnhart, 1969) that surface-active agents in solution reduce the oxygen transfer rate in diffused air aeration systems, resulting in a typical operational α -value range of 0.4-1.0. The corresponding typical α -value range for surface aerators is 0.8-1.2. It should be noted that an α -value in excess of unity infers an enhanced rate of oxygen transfer under process conditions. This may arise where there is a rapid renewal of the air-water interface and hence insufficient time for the formation of a surfactant interfacial barrier to be formed.

The value of the β -factor under process conditions is also typically less than unity and clearly depends on the dissolved solids concentration of the aqueous suspension being aerated. In activated sludge processes, the value of β is generally between 0.9 and 1.0.

14.5 INFLUENCE OF TEMPERATURE ON OXYGEN TRANSFER

The oxygen absorption coefficient k_L increases with increasing temperature, due mainly to the temperature-dependent nature of those water properties that affect gas transfer, namely diffusivity, surface tension and viscosity. Oxygen diffusivity in water increases with increasing temperature. Surface tension and viscosity, both of which affect the surface renewal rate, decrease with increasing temperature. These latter properties also influence the energy required to create air-water interfaces. The total area of air-water interface created per unit energy input should increase with increasing temperature, as surface tension and viscosity correspondingly decrease. This infers that the overall transfer coefficient $k_L a$ should be affected to a greater extent by temperature than is the absorption coefficient k_L .

In rating aeration systems, it is usual practice to relate performance to a standard temperature of 20 °C. The value at any temperature T may be related to its value at 20 °C by a temperature coefficient f, as follows:

$$(k_L a)_{20} = (k_L a)_T f^{(20-T)} \quad (14.17)$$

where T is the temperature ($^{\circ}\text{C}$) at which $k_L a$ is measured. Barnhart (1969) has reviewed studies on the temperature dependence of oxygen transfer in diffused air aeration systems, for which he reported an average value of 1.02 for f . The ASCE Standard for the Measurement of Oxygen Transfer in Clean Water (1992) recommends an f -value of 1.024.

It is also worthy of note that the rate of oxygen transfer by aeration systems, at zero oxygen concentration, and within the temperature range generally encountered in water and wastewater engineering practice, is not greatly affected by temperature since the increase in $k_L a$ with rising temperature is offset by a corresponding decrease in the solubility of oxygen. Using the oxygen solubility values given in Table 1.7 and an f -value of 1.02, calculations show that, while there is an increase of 48.6% in the value of $k_L a$ due a temperature rise from 0°C to 20°C , the actual rate of oxygen transfer, at zero oxygen concentration, by an aeration system at 20°C is less than its value at 0°C , but only by about 8%.

14.5.1 AERATION SYSTEMS USED IN THE ACTIVATED SLUDGE PROCESS

The three main types of aeration system used in the activated sludge (AS) process are as follows:

- (a) dispersed air systems
- (b) mechanical surface aeration systems
- (c) combinations of (a) and (b)

14.6.1 Dispersed air systems

In dispersed air systems, air is introduced to the mixed liquor through submerged 'diffusers'. Dispersed air systems are broadly classified according to the bubble size generated by the diffusers, which may range from fine to coarse. Fine bubble diffusers typically produce bubbles in the 2-5mm size range in clean water, while coarse bubble diffusers produce bubbles in the size range 6-10mm.

Coarse bubble diffusers typically consist of drilled holes or slots in a submerged air distribution system. The produced bubbles may be smaller than the orifice size through being sheared off by the water pressure. The oxygen transfer efficiency (OTE) of coarse bubble systems is generally lower than fine bubble systems (OTE is defined as the percent of oxygen transferred per metre submergence, at zero dissolved oxygen concentration).

Fine bubble diffusers (USEPA, 1989) are constructed from a range of materials including ceramics, porous plastics and perforated membranes. Ceramics and porous plastics have interconnected pore structures through which the air flows to be discharged through the top surface as a bubble stream. Their pore structure is rigid and hence subject to fouling either due to impurities in the air or precipitation from the water side. To prevent clogging of porous diffusers, it has been found necessary (Pasveer and Sweeris, 1965) to reduce the dust content of the air used to a level not greater than 0.036 mg m^{-3} . This is accomplished by filtration.

The perforated membranes are made from either thermoplastic or elastomeric sheet materials (plasticized PVC, EPDM rubber, neoprene rubber) of thickness 1-2mm. They are mechanically perforated with a pattern of small holes or slits, which expand as the air flow increases, hence they are commonly known as flexible membrane diffusers.

Oxygen transfer from an individual air bubble may be considered to take place in three consecutive stages: bubble formation, bubble ascent and bubble escape at the surface. The oxygen transfer rate is high during the bubble formation phase, since a new air-water interface area is being created and the interfacial oxygen concentration gradient is therefore high. The extent of oxygen transfer during the ascent depends on the rate of interface renewal, bubble surface area and bubble contact time, factors that are dependent on bubble size and subsequent depth of the diffuser. Barnhart (1969) found that for given volumes of air and water, the overall transfer coefficient had a maximum value at a bubble size

of 2.2mm. Downing (1960) found that the average value of the overall transfer coefficient decreased with increasing depth of the diffuser. The total contact time, which effectively equals the time of ascent, is greatly reduced if the water, into which the air is released, has itself an upward velocity, as in the case of spiral-flow tanks, where diffusers are located on one side of the tank only. The final phase of oxygen transfer is due to surface disturbance generated by bubble escape. Experimental evidence suggests that in most cases this is the least significant of the three phases.

While the overall oxygen transfer coefficient $k_L a$ for diffused air aeration systems is readily determined experimentally (refer section 14.7), the determination of the absorption coefficient k_L is more problematic. Reported values of k_L vary from 3.67 mm min⁻¹ for a single large bubble in a narrow tube (Adeney and Becker, 1918, 1919, 1920) to 41.67 mm min⁻¹ for bubbles in a highly turbulent aeration tank (Holroyd, 1952).

Fig 14.2 shows a schematic arrangement of a typical AS process diffused air aeration system.

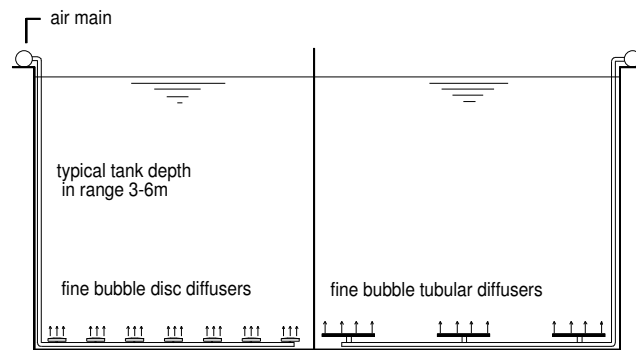


Fig 14.2 AS process diffused air aeration system layout

14.6.2 Mechanical surface aeration systems

Mechanical surface aeration systems effect oxygen transfer by creating and rapidly renewing large areas of air-water interface. Such interfaces result from a combination of entrainment of air, spraying of water into the air and the generation of a high level of surface turbulence.

Mechanical surface aerators may be broadly classified in the following categories:

- (a) Impeller-type devices, mounted on a vertical shaft
- (b) Rotor-type devices, mounted on a horizontal shaft.

Vertical shaft devices are either of the plate or turbine type.

Plate aerators (Downing et al., 1960) consist of a horizontal circular plate, mounted on a vertical shaft, to the underside of which is attached a set of vertical radial or curved blades. In operation, the top of the plate is located at or slightly below the still water surface level. The rotation of the disc causes liquid to be discharged radially, creating a circular hydraulic jump in which air is entrained. Air may also be sucked down through openings in the plate to the low-pressure zones created behind each vane as the plate rotates. A plate aerator on its own may not provide adequate mixing in an aeration tank. Supplemental mixing can be provided by extending the plate shaft downwards and attaching a mixing impeller at a lower level.

Turbine aerators effect oxygen transfer primarily by a pumping action. The impeller, which is located at the liquid surface, is designed to pump large quantities of liquid at a low head. Because the impeller is located at the liquid surface a large amount of air is entrained in the pumped liquid, which is thrown upwards and outwards in a low trajectory, creating considerable turbulence as it strikes the liquid surface. In deep tanks, turbine aerators may be used in conjunction with a draft tube, which extends

downwards to within a short distance of the tank bottom, as shown in Fig 14.3. This arrangement ensures good mixing of the tank contents.

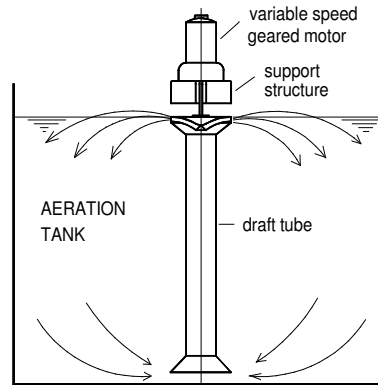


Fig 14.3 Typical arrangement of a vertical shaft surface aeration system

Rotor-type or ‘brush’ aerators consist of a horizontal revolving shaft, to which are attached flat or angular blades, projecting radially outwards. The overall diameter of brush rotors is usually between 0.6-1.0m; their immersion depth may vary between 100mm and 300mm, depending on diameter.

Brush aerators (Casey, 1971) effect oxygen transfer by entraining air in the liquid spray which is thrown upwards and outwards by the revolving blades. Considerable surface turbulence is also generated. Brush aerators are generally used in closed-loop aeration tanks of the oxidation ditch type, in which the mixing requirement is met by generating a horizontal flow velocity of sufficient magnitude to keep the mixed liquor solids in suspension.

It should be noted that AS process aeration systems are required to provide adequate mixing as well as transfer the required amount of oxygen into solution. Hence, the geometry of the aeration tank must be matched to the mixing characteristics of the aeration device. In some situations it may be necessary to supplement a diffused air system by a mechanical mixing device to ensure an adequate distribution of mixing energy throughout the aeration tank.

14.7 EXPERIMENTAL DETERMINATION OF OXYGENATION CAPACITY

The standard oxygen transfer rate (SOTR) of an aeration system (oxygenation capacity) is defined as the rate at which the aeration system transfers oxygen into solution in clean water at a temperature of 20 °C, a zero oxygen concentration in solution and a prevailing atmospheric pressure of 1 atm. In general, the prevailing test conditions will not be standard and hence temperature and pressure corrections have to be applied to the test results. Combining equations (14.14) and (14.17), the resultant expression for SOTR becomes:

$$\text{SOTR} = k_L a_{(T)} f^{(20-T)} C_{*(20)} V \frac{P_s}{P_b} \quad (14.18)$$

where $C_{*(20)}$ is the saturation or equilibrium concentration of oxygen in solution (kg m^{-3}) under test conditions at a water temperature of 20 °C, V is the aerated water volume (m^3), P_b is the prevailing atmospheric pressure (atm) at the test site and P_s is the standard barometric pressure of 1 atm.

For surface aeration systems, the value of $C_{*(20)}$ can be taken as being equal to $C_{s(20)}$, the saturation concentration corresponding to an air pressure of 1 atm, which, as given in Table 1.7, is 9.08 g m^{-3} . For diffused air systems, however, the value of $C_{*(20)}$ exceeds $C_{s(20)}$ by an amount which depends on the diffuser submergence and, hence, has to be determined experimentally.

As shown in section 14.3, the rate of oxygen transfer during an aeration process is represented by the expression:

$$\frac{dc}{dt} = k_L a (C_* - C)$$

or

$$\frac{dc}{(C_* - C)} = k_L a (dt)$$

Integrating this equation between an initial concentration of oxygen in solution C_i at time zero and a concentration C_t after an aeration interval of t :

$$\int_{C_i}^{C_t} \frac{dc}{(C_* - C)} = \int_0^t k_L a (dt)$$

This gives

$$\ln \left[\frac{(C_* - C_i)}{(C_* - C_t)} \right] = k_L a \cdot t \quad (14.19)$$

or

$$\ln(C_* - C_t) = \ln(C_* - C_i) - k_L a \cdot t \quad (14.20)$$

If the natural log of oxygen deficit ($C_* - C_t$) is plotted as a function of aeration time t , then a straight-line graph is obtained in accordance with equation (14.20), as illustrated in Fig 14.5. The overall transfer coefficient $k_L a$ is the slope of this line and has the units time^{-1} .

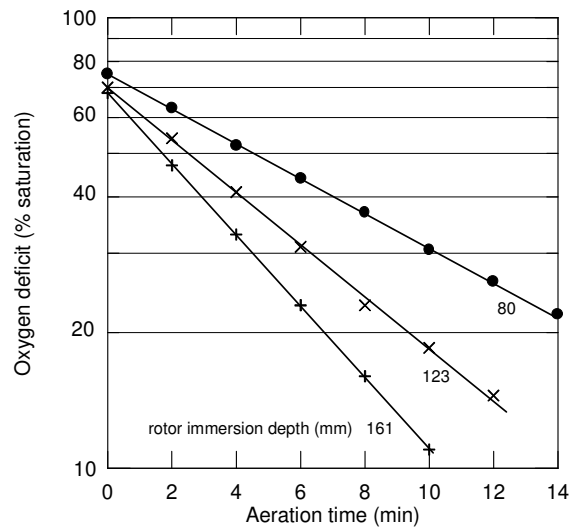


Fig 14.5

**Results of an oxygenation capacity test
on a bladed rotor; rotor diameter 700 mm.**

The determination of the oxygenation capacity of an aeration system requires measurement of the overall oxygen transfer coefficient $k_L a$ (equation (14.18)). The unsteady-state re-aeration method is the usual experimental method for $k_L a$ determination (ASCE, 1992). The water is first de-oxygenated by gas stripping or by the addition of sodium sulphite (Na_2SO_3). To accelerate the chemical reduction of oxygen, a catalyst such as Co^{2+} is added at a concentration of about 0.1 mg l^{-1} . If an independent mixing system is not available it may be necessary to leave the aeration system in operation during the chemical de-oxygenation process to ensure dispersion of the sulphite throughout the test tank volume. Where the latter applies an amount of sulphite in excess of the stoichiometric requirement should be added to allow for the excess consumption by the aeration process. The rate of re-oxygenation is then monitored, usually by DO sensors at a number of locations in the test tank. For reliable $k_L a$ determination, it is desirable that the re-oxygenation measurement should extend over a wide value range, e.g. 10%-90% saturation. Since aeration tanks are normally well mixed, the spatial variation in oxygen concentration at a given instant in time tends to be small. The measured re-oxygenation rate data are used to compute the value of $k_L a$ for the system, using the semi-log graphical method or the three-parameter fit method as described in the following paragraphs.

The simplest method of $k_L a$ determination is the semi-log graphical method, in which the oxygen saturation deficit is plotted on semi-log paper as a function of time, resulting in a straight line plot, as illustrated in Fig 14.5. The value of $k_L a$ is calculated as the slope of the semi-log plot and is corrected for temperature in accordance with equation (14.170). The SOTR value for the aeration system is computed by insertion of the appropriate parameter values in equation (14.8). This method is suitable for aeration systems where the saturation concentration can be assumed to be the same as the concentration in equilibrium with atmospheric air.

To illustrate the three-parameter fit method, equation (14.20) is written in the form:

$$C_t = C_* - (C_* - C_i) \exp(-k_L a \cdot t) \quad (14.21)$$

Non-linear regression is used to find the best-fitting values of the three parameters C_* , C_i and $k_L a$ (ASCE, 1992). An example of the output from such a computation is presented in Fig 14.6. Details of the computational procedure, together with an example are given in Appendix C. This is the preferred method of $k_L a$ determination as it is based on a best-fit model of the experimental measurements. It has the advantage of not requiring an assumed value for the saturation concentration C_* , which is of particular significance for diffused air systems where the saturation concentration is a function of the immersion depth of the diffusers.

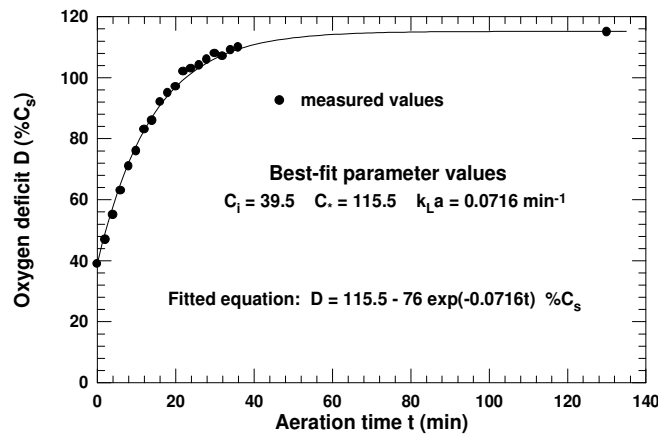


Fig 14.6

Experimental data for SOTR test on diffused air aeration system
Diffuser submergence 4m; DO sensors calibrated to read 100% in air-equilibrated water

The steady-state oxygen transfer rate (OTR) by an activated sludge aeration system, operating at a fixed DO concentration C (mg l^{-1}) and at a temperature T ($^{\circ}\text{C}$), is related to the system SOTR value as follows:

$$\text{OTR} = \alpha \cdot \text{SOTR} \cdot f^{(T-20)} \frac{(\beta C_{*(T)} - C)}{C_{*(20)}} \frac{P_b}{P_s} \quad (14.22)$$

The OTR value for an activated sludge diffused air aeration system can also be measured under field operating conditions by monitoring the oxygen concentration in the off-gas from the aeration tank and thus quantifying the drop in the oxygen content of the air flow. Ideally, carbon dioxide and water vapour should be stripped from both streams to eliminate partial pressure effects. A set of guidelines for carrying out off-gas tests has been drawn up by the American Society of Civil Engineers (ASCE, 1996). The oxygen content of the influent and emergent air streams is conveniently done using an air-calibrated DO meters. The procedure can be used as a steady-state method for measuring oxygen transfer in activated sludge processes operating at a constant loading rate and hence at fixed DO level.

14.7 ENERGY CONSUMPTION AND OXYGENATION EFFICIENCY

The standard aeration energy utilisation efficiency (SAE) is usually defined in terms of the energy expended per unit mass of oxygen transferred under SOTR measurement conditions:

$$\text{SAE} = \frac{\text{SOTR}}{P} \text{ kg O}_2 \text{ kWh}^{-1} \quad (14.23)$$

where P is the 'wire-to-water' power input (kW). The power consumption of air blower units used to supply air to diffused air aeration systems is related to air flow and delivery pressure as follows:

$$P = \frac{1.014 Q_{a(m)} T_1}{\eta} \left[\left(\frac{p_2}{p_1} \right)^{0.283} - 1 \right] \text{ kW} \quad (14.24)$$

where $Q_{a(m)}$ is the intake air flow (kg s^{-1}), T_1 is the absolute temperature (K) of the intake air, η is the combined motor/blower efficiency, p_1 and p_2 are the absolute pressures of the intake and compressed air streams, respectively.

The SAE of diffused air systems is a function of the combined efficiencies of the air delivery and oxygen transfer processes. The SAE value for fine bubble diffused air systems typically falls within the range $2.5\text{--}4.0 \text{ kg O}_2 \text{ kWh}^{-1}$. The SAE of surface aeration systems is typically in the range $1.5\text{--}2.0 \text{ kg O}_2 \text{ kWh}^{-1}$.

The energy utilisation efficiency of aeration systems under field operating conditions generally fall within a much narrower band than given by the foregoing SAE value ranges. This is due to the fact that the performance of diffused air systems is more adversely affected (reduced α - and/or β -factors) by contaminants such as surfactants than is the performance of surface aeration systems. While fine bubble aeration systems are generally more efficient than surface aeration systems, they have the disadvantage of being susceptible to clogging and reduction in performance over time.

The standard oxygen transfer efficiency (SOTE) of diffused air aeration systems is the fraction of oxygen in the compressed air flow that is taken into solution under SOTR conditions:

$$\text{SOTE} = \frac{\text{SOTR}}{0.228 Q_{a(m)}} \quad (14.25)$$

where $Q_{a(m)}$ is the air mass flow rate with an oxygen mass fraction of 0.228.

14.8 GAS STRIPPING

14.9.1 Removal of carbon dioxide

As may be seen from the data in Table 1.9 relating to the solubility of atmospheric gases in water, the saturation concentration of carbon dioxide (CO₂) in a water body in contact with free air is less than 1 mg l⁻¹ at ambient temperature. Waters become supersaturated with CO₂ in a variety of circumstances. For example, many groundwaters are found to have high concentrations of CO₂, taken into solution as rainwater percolates through soils in which CO₂ is produced by microbial respiration. High CO₂ concentrations are produced in the treatment of high-alkalinity waters through the addition of trivalent aluminium or iron salts in the chemical coagulation process used for the removal of colour and turbidity in the production of drinking water from surface waters. As discussed in Chapter 9, waters that are supersaturated with CO₂ may be chemically unstable and corrosive.

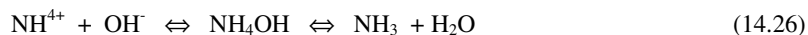
The basic gas transfer principles already outlined for the dissolution of oxygen in water are also applicable to the transfer of CO₂ out of solution. For gas removal it is important to keep the partial pressure of the gas in the gas phase as low as possible - this can be achieved by good ventilation with a high ratio of air/water. Spray nozzles, cascades, forced draft multiple tray aerators and packed columns are used for CO₂ stripping.

14.9.2 Removal of hydrogen sulphide

Hydrogen sulphide (H₂S) is found in some groundwaters and in effluents from anaerobic treatments. Even at low concentrations it gives rise to odour and corrosion problems and has a high chlorine demand. Its removal by aeration is complicated by its dissociation in solution (see equations (14.5) and (14.6)). To control the extent of formation of the ionic species HS⁻ and S²⁻ it may be necessary to reduce the pH by acid addition. Added oxygen may also react with H₂S to form free sulphur. While the latter reaction reduces the H₂S content it may lead to further problems in chlorination and distribution (growth of sulphur bacteria in the distribution system). Although the general aeration principles already outlined are valid, because of the additional complicating factors involved in H₂S removal, pilot plant studies are advisable prior to plant design.

14.9.3 Removal of ammonia by aeration

Ammonia is produced in most wastewaters, including domestic sewage, through the microbial degradation of organic compounds containing nitrogen. Since it is an important nutrient, its removal from effluents is often necessary to control eutrophication in receiving waters. This can be achieved biochemically by nitrification (conversion to NO₃⁻) followed by subsequent denitrification to N₂, in which form it is no longer an available source of nitrogen for most organisms. Ammonia can also be removed by aeration. For efficient air-stripping a high pH is required to convert ammonium ion to NH₃:



Because ammonia is a highly soluble gas (see Table 1.8), large volumes of air are required to effect its removal. The efficiency of the process increases with increasing temperature owing to the resulting reduction in gas solubility.

14.10 GAS TRANSFER IN PACKED COLUMNS

Inter-phase gas transfer is achieved in packed columns by the creation and renewal of the gas-water interface area as water trickles down through a bed of discrete packings of the type shown in Fig 14.7. The downward cascading water is brought into close contact with the gas contained in the voids of the packing. Depending on the requirements of the application the gas phase may be maintained at a fixed pressure (as in saturators used in conjunction with the flotation process) or may be in the form of a counter-current flow (as in gas-stripping applications).

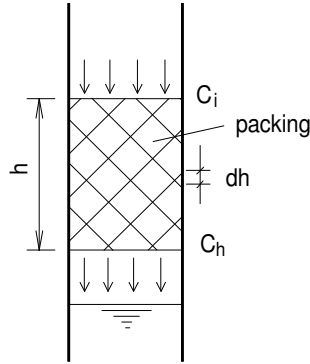


Fig 14.7 Schematic representation of packed column

Using Fig 14.7 as a process definition diagram, the gas transfer characteristics of packed columns can be derived by writing equation (14.7) in the form:

$$dc = k_L a \, dt (C_s - C)$$

Applied to a packed column, this relationship can be written as follows:

$$\begin{aligned} \frac{dc}{dh} &= k_L a (C_s - C) \frac{dt}{dh} \\ \frac{dc}{dh} &= k_L a (C_s - C) \frac{t}{h} \end{aligned} \quad (14.27)$$

where t is the 'hold-up time' in the column and h is the column height. It is assumed that $dt/dh = t/h$, i.e. there is no variation in the liquid residence time over the column height.

The ratio $h/(k_L a \cdot t)$ has the dimensions of length and has been found to be approximately constant for a given gas transfer process. This packed column characteristic is the 'transfer unit height' or HTU. Thus equation (14.27) can be written as

$$\frac{dc}{dh} = \frac{1}{\text{HTU}} (C_s - C) \quad (14.28)$$

Integration yields:

$$\ln \left(\frac{(C_s - C_i)}{(C_s - C_h)} \right) = \frac{h}{\text{HTU}} \quad (14.29)$$

where h is the packing height, C_i is the inflow dissolved gas concentration and C_h is the outflow dissolved gas concentration.

Typical HTU values (air to water transfer)

	HTU (mm)
25 mm and 38 mm spheres	285
46 mm tellerettes	405
70 mm tellerettes	375
25 mm Raschig rings	500

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