CHARACTERISTIC EVALUATIONS OF ICI AIR-LIFT TYPE DEEP SHAFT AERATOR

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An aeration system employing a long vertical shaft of 100 m or more depth has recently been developed by ICI. The interior of the shaft is divided into two parts. Air bubbles sparged in one side of the interior flow down to the bottom and then flow up through the other side of the shaft to the surface, as liquid circulates through both sides. This liquid circulation is maintained by the gas lift action of the sparged air. In this aeration system, the flow path of the air bubbles introduced is long enough to yield high oxygen utilization.

In the present paper, the authors discuss operating problems of ICI's deep shaft aerator. The air-lift type deep shaft was simulated by a mathematical model. The momentum balance equation for the fluid flow, the theoretical and empirical relations between gas velocity, liquid velocity, gas slip velocity and gas holdup, and the mass balance equation for oxygen transferred through the gas-liquid interface, were all simultaneously solved to obtain the distributions of gas holdup and oxygen utilization along the shaft length.

The calculated results give valuable information, such as the relations between the gas velocity, liquid circulation velocity, depth of air introduced and oxygen utilization. The power economy of the deep shaft is discussed in connection with its design and operating problems.

Introduction

Treatment of high-concentration organic wastewater or production of single-cell protein requires a high uptake rate of oxygen in the liquid phase. It is important, therefore, to reduce the power consumption for requisite oxygen transfer through the gas-liquid interface. Various gas-liquid contacting devices have been investigated and developed for this purpose.

Recently an aerator consisting of a long vertical shaft, reaching 100-300 m depth beneath the ground, has been developed by ICI Limited. As shown schematically in Fig. 1, liquid circulation maintained by the gas lift action of injected air or by a liquid pump carries air bubbles deep into the bottom of the shaft. Effective oxygen transfer from gas phase to liquid phase is achieved at elevated pressure at a deep water level. It seems to offer a high oxygen transfer rate per unit power consumption in comparison with other aerators.

The present study is aimed at obtaining quantitative information about operating characteristics of the air-lift deep shaft aerator (Fig. 1 (a)), the type more commonly used at present.

1. Mathematical Model for Air-Lift Aerator

1.1 Principle of liquid circulation by air-lift

In the air-lift deep shaft, the static head difference, caused by the difference of gas holdup distributions between the downward flow path and the upward flow path in the shaft, is the driving force of liquid circulation. Under steady-state liquid circulation, the pressure drop due to hydraulic friction of the flow circuit along the flow path is balanced with the static head difference. Therefore, the condition of air injection into the shaft to maintain necessary liquid circulation velocity can be obtained by solving this balance equation.

In the following, steady-state liquid circulation in the shaft is assumed.

1.2 Static head difference ΔP, between the downcomer and the riser

The section used as the liquid downward flowpath is called the 'downcomer' and that used as the upward flowpath is called the 'riser'. A static head in the downcomer, \( P_{\text{down}} \), and that in the riser, \( P_{\text{up}} \), at a loca-
tion \( l \) \( m \) in depth from the top liquid surface are expressed by
\[
P_{\text{st}} = \rho_l (g/g_c) \int_0^l (1 - \varphi_l) dl + P^* \tag{1}
\]
\[
P_{\text{st}} = \rho_l (g/g_c) \int_0^l (1 - \varphi_l) dl + P^b \tag{2}
\]
From Eqs. (1) and (2), the static head difference at the shaft bottom must satisfy the equation
\[
\Delta P_s = -(P_{\text{st}})_{z=H} - (P_{\text{st}})_{z=H'} = \rho_l (g/g_c) \int_{H'}^H \varphi_l dl - \int_0^H \varphi_l dl \tag{3}
\]
1.3 Pressure drop \( \Delta P_f \) along the liquid flow circuit in the shaft
The pressure drops \( \Delta P_{f1} \) and \( \Delta P_{f2} \), caused by wall friction in the downcomer and in the riser, respectively, are
\[
\Delta P_{f1} = 4 \left( f_i h + f'_{i} (H-h) \right) \rho_l U^2_{sl}/2g_sD_1 \tag{4}
\]
\[
\Delta P_{f2} = 4 f_i \rho_l U^2_{sl}/2g_sD_2 \tag{5}
\]
where \( f_i \) denotes the friction factor for a homogeneous liquid flow without gas bubbles, and \( f'_{i} \) and \( f'_{i} \) are those with the effect of gas bubbles.
The pressure drop at the bottom and that at the top of the shaft, \( \Delta P_{f3} \) and \( \Delta P_{f4} \), respectively, which arise from changes in flow direction and in cross-sectional area of the flow path, are denoted by
\[
\Delta P_{f3} = 4 f'_{i} L \rho_l U^2_{sl}/2g_sD_1 \tag{6}
\]
\[
\Delta P_{f4} = 4 f'_{i} L \rho_l U^2_{sl}/2g_sD_2 \tag{7}
\]
The total pressure drop along the liquid flow circuit is
\[
\Delta P_f = \sum_{i=1}^{4} \Delta P_{f_i} \tag{8}
\]
Under steady-state liquid circulation, the following equation should hold:
\[
\Delta P_s = \Delta P_f \tag{9}
\]
1.4 Bubble slip velocity \( U_s \) and gas holdup \( \varphi \)
The following relation must hold among superficial gas velocity \( U_g \), gas holdup \( \varphi \), superficial liquid velocity \( U_l \) and apparent slip velocity of gas bubble \( U_s \) to the liquid phase:
\[
U_s = U_g/\varphi - U_l/1(1-\varphi) \tag{10}
\]
In the deep shaft aerator, the effect of pressure change with depth on \( U_s \) must be taken into account, and thus
\[
U_s = U^*_g (1 - \delta Z_{g0})(P^*/P) \tag{11}
\]
where \( \delta = 1 - \beta \), and \( \beta \) is moles of gas produced per mole of oxygen consumed. In biochemical reactions nearly a mole of carbon dioxide is produced per mole of oxygen uptake in liquid.
The gas slip velocity \( U_s \) in tubes is in general a function of gas holdup \( \varphi \), tube diameter \( D \) and liquid velocity \( U_l \). For comparatively large bubbles, so-called coarse bubbles, which are generated from a single nozzle or a perforated plate, and under conditions of low gas holdup values, it is known empirically that \( U_s \) in water is almost constant at 0.3 m/sec.
With increasing gas holdup \( \varphi \) and tube diameter \( D \), however, \( U_s \) increases. According to the authors' experience, the following correction of gas holdup \( \varphi \) and tube diameter \( D \) on \( U_s \) will apply.
\[
U_s/U_{\varphi 0} = 1 + 1.35(\sqrt{gD}/U_{\varphi 0})\varphi \tag{12}
\]
where
\[
U_{\varphi 0} = 0.3 \text{ m/sec}
\]
1.5 Oxygen transfer through gas-liquid interface
The air injected into the downcomer of the deep shaft aerator flows down with liquid to the bottom and then rises in the riser. If there is oxygen demand in the liquid, oxygen must transfer from the gas phase to the liquid phase through the interface between them. The oxygen transfer rate is important not only to predict the value of oxygen utilization, which indicates how effectively the oxygen injected is utilized, but also to obtain correct values of superficial gas velocity and of gas holdup, which are affected by oxygen consumption in the gas phase. When a biochemical reaction takes place in the liquid, the influence of generating carbon dioxide must also be taken into account.
In a more strict analysis, the effect of solution and discharge of nitrogen gas of the injected air in the liquid under various pressures must also be considered, but this effect is neglected for simplicity in the present calculations.
Assuming plug flow of the gas phase along the flow paths, the differential mass balance for oxygen which transfers through the gas-liquid interface gives
\[
G_m (d\varphi/dl) = K_{r,a}(C_m - C_l)S_a \tag{13}
\]
The relation between oxygen utilization \( \eta \) and mole fraction of oxygen in the gas phase is given by
\[
\eta = (Z_m - Z_m)/Z_m/1(1-\delta Z) \tag{14}
\]
2. Embodiment of Design Calculations
2.1 Hypothesis for illustrative calculations
There are two ways of dividing a shaft section as shown in Fig. 2, that is, the concentric type and the divider type. The former is equipped with an air sparger in the inside pipe. Therefore, the inside pipe is employed as the downcomer and the annular part as the riser. This type is usually suited for a smaller diameter shaft. In the present illustrative calculations, it was assumed that the two cross-sectional areas are equal for both the concentric and the divider types. Physical properties of the fluids needed for calculations are summarized in Table 1.
2.2 Information and assumptions for calculations of pressure drops

The friction factor \( f \) in Eq. (4) is given by Eq. (15), because illustrative calculations are made for operations ranging in Reynolds number from \( 3 \times 10^6 \) to \( 1.6 \times 10^7 \).

\[
1/ \sqrt{f} = 3.2 \log (Re \sqrt{f}) + 1.2 \quad (15)
\]

where

\[
3 \times 10^6 < Re < 10^7
\]

The friction factor for the gas-liquid two-phase flow system \( f' \) in Eqs. (4) and (5), is corrected by the foregoing \( f \) as

\[
f' = \xi f \quad (16)
\]

According to the study of Koide and Kubota, it will be presumed that \( \xi = 1 \) under conditions of practical application of the deep shaft.

The prediction of the pressure drop at the bottom end, caused by the change of flow path, is somehow erroneous at the present stage, because correct information has not yet been reported. Nishimura and Fukumuro proposed the following equation for concentric-pipe equipment, in which the direction of liquid flow was opposite to the present case. Here, however, it was assumed that their equation could still be applied.

\[
\Delta P_{f,3} = 9.8 \left( \frac{S_1}{S_1 + S_2} \right) \left( \rho_1 U_1^2 / 2g_s \right) \quad (17)
\]

where \( S_1, S_2 \) are the cross-sectional areas of the inside pipe and of the outside pipe, respectively.

For the divider-type shaft, as \( \Delta P_{f,3} \) the value of the pressure drop for a 180° bend tube was used:

\[
L_a = 7.5 D_1 \quad (18)
\]

The values estimated above may include a large error. At the present stage, however, there is no way for correct estimation. Besides, the contribution of \( \Delta P_{f,3} \) is not important in comparison with \( \Delta P_{f,1} \) and \( \Delta P_{f,2} \).

At the top of the shaft, for the open liquid surface the pressure drop can usually be neglected, and thus

\[
\Delta P_{f,1} = 0 \quad (19)
\]

2.3 Mass transfer capacity coefficient and oxygen utilization

There is another difficulty in estimating the mass transfer capacity coefficient \( K_{La} \) in Eq. (13) under the condition of elevated pressure in the deep shaft. One of the present authors proposed the following equation for \( K_{La} \) in the deep tank aeration up to 20 m in depth.

\[
K_{La} = \alpha U_s \quad (20)
\]

where as \( U_s \) the corrected value under pressure should be used. The coefficient \( \alpha \) is characteristic of the properties of the gas sparger and has a value ranging from 0.5 to 1.0 1/m, if the dimension of \( K_{La} \) is 1/hr and that of \( U_s \) is m/hr.

Since Eq. (20) is based on data taken under the conditions where liquid flow was absent, the following was assumed.

\[
U_s = U_s \cdot \psi \quad (21)
\]

This is directly obtained from Eq. (10), if one takes \( U_i = 0 \). Thus Eq. (20) will be generalized to the case with liquid flow as

\[
K_{La} = \alpha U_s \cdot \psi \quad (22)
\]

By considering the case where the oxygen demand in the liquid phase is large enough, it was assumed that the dissolved oxygen concentration \( C_i \) is negligible everywhere in the liquid flow paths. Under this assumption the maximum value of oxygen utilization in the shaft was obtained. By substituting Eq. (14) and Eq. (22) into Eq. (13), the following equation to calculate the maximum oxygen utilization is derived:

\[
G_o \frac{d \gamma}{dt} = a \cdot U_s \cdot \psi \cdot C_o \cdot S_s \cdot \frac{p_0(1-\gamma)}{1-\delta \gamma \cdot \eta} \quad (23)
\]

In the present calculations, the constant \( a \) is employed as 0.5, which corresponds to be the case where comparatively coarse bubbles are sparged. Beside, \( \beta \) is assumed to be unity, that is, a mole of CO\(_2\) is generated by a mole of O\(_2\) uptaken. Therefore, the calculation is carried out as \( \delta = 1 - \beta = 0 \).
2.4 Calculation

Figure 3 shows the flow chart of the design calculations according to the mathematical model described above. Specifying a superficial liquid velocity $U_l$ and a location of the air injection measured as a distance from the liquid surface $h$, the supplying gas velocity required to maintain the liquid circulation $U*o$ was determined. As $U*o$, the value converted under atmospheric pressure was employed. As necessary liquid circulation velocity $U_l$, values were selected ranging from 1 m/sec to 2 m/sec. The distributions of gas holdup $\varphi$ and oxygen utilization $\gamma$ were also determined simultaneously.

To proceed with calculation, the shaft length $H$ was fractionated in equal distances $\Delta l$ and then $\varphi$ and $\gamma$ in their increments were determined in turn from the inlet.

Determining $\varphi_i$ from the value of $\varphi_{i-1}$, a trial-and-error procedure was used, where as the average gas holdup from the inlet to the $i$-th section, $\varphi_i=(1/i) \sum_{\ell=1}^{i} \varphi_{\ell}$ was used. $\gamma$ from Eq. (23) was similarly determined by employing numerical integration.

After completion of the calculations for the downcomer, calculation shifts to the riser, where $U_l=-U_t$ and $U_r=-U_g$ were set. The consecutive calculation proceeded from the top to the bottom in a manner similar to that for the downcomer. Since $\gamma$ at the top is unknown, its trial-and-error determination was required. Both values of $\varphi$ at the bottom, which are obtained in the downcomer and the riser, must of course agree.

2.5 Oxygen transfer rate per unit power consumption

As previously described, a postulated characteristic of the deep shaft is that the oxygen transfer rate per unit power consumption is quite large compared with the others. For the air-lift deep shaft, the power serves to compress air to the pressure of gas injection. Assuming adiabatic compression, the power required to compress air with molal flow rate of $G_o$ is expressed by

$$P_w=G_o \frac{RT}{\gamma-1} \left[ \left( \frac{P}{P_0} \right)^{\frac{(\gamma-1)/\gamma}{\gamma-1}} - 1 \right]$$

where $\gamma$ is the ratio of the constant-volume specific heat to the constant-pressure specific heat of air and is given as 1.4. The mechanical efficiency of compression $\varepsilon$ was taken as 0.6 in the present calculations. The resistances of air movement in pipes are included in this figure. Oxygen transfer rate per unit power consumption is thus calculated as $G_o/P_w$.

3. Calculated Results and Discussion

3.1 Reported data of deep shafts

Table 2 shows the data reported by ICI. As seen there, relatively high values of oxygen utilization and oxygen transfer rate per unit power consumption were given. The design calculations were carried out in reference to these figures.

3.2 Relations between $U*o$ and $U_l$

For the execution of the design calculations, the relation between the slip velocity $U_s$ of gas bubbles against the liquid phase and the gas holdup $\varphi$ was an important factor. As previously described, the calculations proceeded first by assuming the constant $U_s$ to be 0.3 m/sec, called here the first-order approximation. Then $U*o$ was calculated from Eq. (12) as a function of $\varphi$, called the second-order approximation. The pressure drop $\Delta P_f$, caused by the circulating liquid flow described in 2.2, is given as a function of the superficial liquid velocity $U_l$ only. Figure 4 shows the relations between $\Delta P_f$ and $U_l$ for both the concentric type and the divider type shafts. $\Delta P_f$ for the concentric type is larger than that for the divider type, because the former has a larger wall surface area than the latter. Since the pressure drop at the bottom end was not too large compared to that at the wall, an error introduced in the calculation of the
pressure drop at the bottom end would little affect the design calculations.

**Figure 5** shows the relation between static head difference $\Delta P_s$ and gas flow velocity for various gas sparger locations $h$, where liquid flow velocity is constant at 1 m/sec and the shaft depth is 100 m. The values of $\Delta P_s$ resulted from the calculated distributions of gas holdup $\varphi$ as described in the foregoing sections. Since the liquid velocity is fixed, the pressure drop $\Delta P_f$ is almost constant. The crossing points between the curves of $\Delta P_s$ and $\Delta P_f$ give the required gas velocities $U_g^*\varphi$ to maintain steady liquid circulation of 1 m/sec. Here $U_g^*$ denotes the values of $U_g$ converted under atmospheric pressure for the introduced air. When the first-order approximation was used to calculate $U_s$, crossing points to give $\Delta P_s = \Delta P_f$ exist for widely varied values of $h$. On the other hand, when the second-order approximation was used, at small values of $h$ the curve of $\Delta P_s$ did not cross that of $\Delta P_f$ and therefore no solution giving a stationary liquid circulation was obtained.

In **Fig. 6** and **Fig. 7**, the calculated relations among $U_g^*\varphi$, $U_l$ and $h/H$ are illustrated for the first-order and the second-order approximations, respectively. One can see that two different values of $U_l$ are given for the same values of $U_g^*\varphi$ and $h$. Around a $U_l$ value, $U_l$ increases with increase of $U_g^*\varphi$ and decreases with decrease of $U_g^*\varphi$. On the other hand, around another $U_l$ value, the reverse is observed. When the second-order approximation is applied, the region of no solution widely covers the left half of the diagram, in which $U_l$ increases with decrease of $U_g^*\varphi$. It should be mentioned that in order to maintain liquid circulation at a constant flow velocity for a fixed location of air...
injection, a minimum value of $U_{sp}^*$ exists. Beyond this minimum stable liquid circulation cannot be obtained and operation near this minimum will thus be impractical. In conclusion, the region of practical stable operation is limited to the right half of the diagram.

3.3 Gas holdup $\varphi$ and oxygen utilization $\eta$

In the deep shaft aerator, the hydraulic pressure at the bottom becomes more than 10 atm, because the depth of the shaft exceeds 100 m, and the gas holdup distribution is widely varied. Illustrative calculations for the distributions of $\varphi$ among the flow paths in the concentric-type shaft under various conditions are shown in Fig. 8. The fact that $\varphi$ in the riser is considerably different from that in the downcomer is worth noting. The reason is that the direction of slip velocity $U_*$ in the downcomer is different from that in the riser.

Figure 9 shows illustrative calculations for distributions of $\eta$ for the same concentric-type deep shaft. In Fig. 10, the outlet values of the oxygen utilizations $\eta$ are shown against $U_*$ at varied positions of the gas sparger. Figure 10 also shows the effect of the value of mass transfer capacity coefficient $K_{L,a}$, which is characterized by the coefficient $a$ in Eq. (20). As values of $a$, 1.0 for fine bubbles and 0.5 for coarse bubbles were employed. In these calculated results, attention should be paid to the fact that the value of $\eta$ is limited, because there is a region of unstable operation.

3.4 Oxygen transfer rate per unit power consumption and oxygen demand

In practical operation of aeration devices the power
economy, defined as the maximum oxygen transfer rate per unit power consumption, is an important factor. This is obtained easily in the diagram shown in Fig. 11, in which calculated values of $G_o/D / P_w$ for $C_i=0$ through the liquid flow path under varied operating conditions are plotted against oxygen demand, given by $G_o/V$.

The solid curve (D) in Fig. 11 shows the power economy for the case where the depth of air injection is 50 m in a shaft of 100 m in depth. The operation, when $U_i=1.0$ m/sec and $C_i=0$, corresponds to a point A. Even when the oxygen demand in liquid decreases from that shown by point A, however, in order to maintain stable liquid circulation, set as $U_i=1$ m/sec here, the supplied air velocity cannot be reduced. And thus a large part of the air supply power is unnecessarily spent to increase the dissolved oxygen concentration in liquid $C_i$, while the power consumption $P_w$ is kept almost constant. Therefore, the power economy $G_o/D / P_w$ decreases linearly along the AO line. On the other hand, if the oxygen demand in liquid increases more than that of point A, the air supply rate must be increased to supply dissolved oxygen in the shaft. In this case, the oxygen utilization decreases according to the increased $U_i$, and consequently the power economy decreases along the curve AB. In the case of high oxygen demand, the location of air injection $h$ must be decreased to obtain a high value of $G_o/D / P_w$. This is illustratively shown in the curve (2). It is concluded that in applications of the deep shaft, in order to improve the power economy, the selection of the air injection location is an important design factor. With this in view, it must be deduced also that high power economy will occur only where high oxygen demand exists in the liquid.

**Conclusion**

1) A mathematical model to simulate the operation of ICI's air-lift type deep shaft aerator was presented.

2) Based upon the model, the condition of air supply to maintain a stable liquid circulation in the shaft was obtained and discussed.

3) It was pointed out that the maximum oxygen transfer rate per unit power consumption for the deep shaft is considerably affected by the oxygen demand in liquid, and for the rational design of shaft operation selection of air sparger location is important.

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**Nomenclature**

- $a$: characteristic coefficient of gas sparger for oxygen transfer [1/m]
- $C_i$: dissolved oxygen concentration in liquid [kg/m³]
- $C_o$: dissolved oxygen concentration saturated with pure oxygen [kg/m³]
- $D$: equivalent diameter of stream path [m]
- $f$: friction factor [--]
- $G_o$: supplied oxygen flow rate [kg-mol/hr] or [kg-O₂/hr]
- $g$: acceleration of gravity [m/sec²]
- $g_0$: gravitational conversion factor [kg·m/Kg·sec²]
- $H$: length of shaft [m]
- $h$: depth of air injection [m]
- $K_{La}$: mass transfer capacity coefficient [1/hr]
- $l$: distance from liquid surface [m]
- $P$: pressure [Kg/cm²]
- $P_s$: static head [Kg/cm³]
- $\Delta P_f$: pressure drop due to liquid circulation [Kg/cm²]
- $\Delta P_s$: static head difference between riser and downcomer [Kg/cm³]
- $P_w$: power consumption [kcal/hr] or [kW]
- $R$: gas constant [k-cal/kg-mol °K]
- $Re$: Reynolds number [--]
- $S_a$: sectional area of stream path [m²]
- $T$: temperature of supplied air [°K]
- $U_f$: superficial gas velocity [m/sec]
- $U_i$: superficial liquid velocity [m/sec]
- $U_s$: superficial slip velocity of gas bubbles against liquid phase [m/sec]
- $V$: effective volume of shaft [m³]
- $z$: mole fraction of oxygen in gas [--]
- $\beta$: moles of produced gas per mole of consumed oxygen [--]
- $\epsilon$: power efficiency [--]
- $\eta$: oxygen utilization; ratio of oxygen consumed per oxygen supplied [--]
- $\rho$: liquid density [kg/m³]
- $\phi$: gas holdup [--]
- $\phi$: average gas holdup [--]

**Literature Cited**


