Aeration in Bernoulli Types of Devices

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Many industrial processes, including waste treatment and fermentation to produce alcohol and acids, require oxygen in substantial quantities to promote biological growth. Oxygen is only a very slightly soluble gas of the order of 10 parts per million in water from air. Even the use of pure oxygen will increase the solubility to only about five times this value. Thus, the amount of oxygen transferred in a single pass through an aeration device can only be very small. An ordinary waste treatment process may require an amount of oxygen equivalent to saturation of the liquid from 100 to 200 times. The low solubility leads to relatively high energy requirements to provide the oxygen needed. An ideal aeration device would be one in which the gas was not compressed and liquid turbulence was minimized.

It was observed that saturation of water by air was approached in an ordinary laboratory aspirator (*8*). This nozzle type of device, as well as an orifice or Venturi, operate on the Bernoulli principle of the conservation of energy; that is an increase of velocity (kinetic energy) at a point in the flow stream results in a decrease of pressure (pressure energy). The velocity can be increased such that the pressure at the restriction falls below that of the atmosphere and air is forced into the liquid stream through a suitable opening.

Preliminary work (*10*) indicated that there was at least some evidence that devices of this type could be used as economical mass transfer equipment, and that even though residence times were small, transfer rates were relatively large. There was also an indication that the energy consumption expended in flow of the liquid through the device might be less than that for other methods of aeration. The air-water system afforded observation of mass transfer phenomena. The effects of dissolved substances and simulated industrial wastes on the transfer process were also investigated.

THEORETICAL CONSIDERATIONS

It has been reported (*6, 14*) that the bulk of the mass transfer from a gas to a liquid (bubbles) occurs when interfacial area is first formed. This observation results from unsteady state transfer to liquid elements which predominates during the initial period of contact of the phases (*17*). This behavior is significant for transfer in Bernoulli types of devices in which the residence time in the high velocity section is quite small and flow is highly turbulent. In these devices the gas is brought into contact with the liquid at the point of low pressure. Subsequent movement of the two-phase stream through the expansion section, a region of decreasing velocity (increasing pressure), results in a reduction of bubble size and a smaller interfacial area of transfer. Because of high turbulence, the liquid elements of the interface are rapidly replaced and a new transfer area is formed.

The partial pressure of oxygen on the gas side increases with increasing Venturi pressure, and the liquid concentration increases downstream from the throat; the driving forces involved for mass transfer are thus somewhat speculative. Unless the liquid is saturated with oxygen in this short period of time, additional transfer will occur in the downstream piping as long as the two phases remain in contact. Downstream transfer beyond the device is small because of approach to saturation and coalescence of gas bubbles which reduces the interfacial area.

The transfer of oxygen in the air-water system represents a case of a slightly soluble gas. Under these conditions it is considered that the gas-phase resistance is negligible and the resistance to transfer lies within the liquid phase.

Transfer Coefficients by an Unsteady State Method

Use of a closed system with recycle of the liquid, with observation of the increase of solute concentration in a storage tank, eliminates some sampling difficulties. A large volume of liquid at constant composition is not necessary, and simultaneous sampling upstream and downstream from the device is unnecessary. A closed system introduces the requirement of instantaneous mixing in the tank and minimum transfer of solute in the piping between transfer device and storage tank. The unsteady state procedure was selected as minimizing experimental effort for small scale observations.

The rate of transfer of mass by a device is given by an equation of general form

$$r = k_L A\,(C_i - C)_{\text{mean}} \tag{1}$$

which is in terms of liquid side transfer. The gas-phase resistance is considered negligible, and the interfacial concentration on the liquid side becomes the saturation value in equilibrium with the bulk gas phase $C_i = C_S$. Thus, the overall and liquid side coefficients become the same $K_L = k_L$, and

$$r = K_L A\,(C_S - C)_{\text{mean}} \tag{2}$$

The rate of transfer of solute by the device can also be expressed by the change in concentration of the liquid in passing through the device:

$$r = (w/\rho)\,(C_2 - C_1) \tag{3}$$

Further, the rate at which solute is being transferred is given by the rate of change of solute in the entire liquid volume

$$r = V\,(dC_1/d\theta) \tag{4}$$

where C_1 is the uniform tank composition and also the concentration of the liquid entering the device.

For convenience the mean driving force is taken as an arithmetic average (error introduced estimated at 5 to 10% and justifiable in view of the variation in C_S):

$$(C_S - C)_{\text{mean}} = C_S - (C_1 + C_2)/2 \tag{5}$$

The downstream concentration C_2 which may be difficult of accurate measurement is eliminated from Equations (2), (3), and (5) and the resulting rate is equated to that of Equation (4) to give

$$\frac{dC_1}{(C_S - C_1)} = \frac{(2\,w\,K_L A)\,d\theta}{V(2\,w + \rho\,K_L A)} \tag{6}$$

The equation is integrated to relate tank liquid concentration to time by imposing lower limits of $\theta = 0$ at $C_1 = C_0$:

$$\ln\frac{(C_S - C_0)}{(C_S - C_1)} = \left[\frac{(2\,w\,K_LA)}{V(2w + \rho\,K_LA)}\right]\theta = m\theta \qquad (7)$$

This is the equation of a straight line, if the left side of the equation is plotted against time, passing through the origin and having a slope equal to the terms multiplying θ. The slope m permits calculation of the transfer factor by the equation

$$(K_LA) = \frac{(V\,m)}{1 - (\rho\,V\,m)/(2\,w)} \qquad (8)$$

The derivation requires the assumption of a constant value for C_S at the conditions existing in the device and is justified by circumstances described later. The transfer coefficient and the interfacial area are combined as (K_LA), termed the *transfer factor*, which is necessary because of the absence of any knowledge of interfacial area. This is not a serious limitation because the transfer factor is the property which characterizes the process in a Bernoulli type of device. The interfacial area created varies with the type of device, the flow rates, and the interfacial tension.

Certain conditions impose limiting values for (K_LA) and the slope of the line described by Equation (7). At high liquid flow rates the change across the device becomes small, and C_2 approaches C_1. At this limit Equation (2) reduces to $r = (K_LA)\ (C_S - C_1)$ which, on integration with Equation (4), gives an equation of the form of Equation (7) but with a slope such that $(K_LA) = (V\,m)$:

$$\ln\frac{(C_S - C_0)}{(C_S - C_1)} = \frac{(K_LA)}{V}\,\theta \qquad (9)$$

Another limiting case occurs for $C_2 = C_S$ which results for low flow rates or high transfer rates. Equation (2) reduces to $r = (K_LA)(C_S - C_1)/2$ and integration gives Equation (10). In this case $(K_LA) = 2(Vm)$; further, (Vm) cannot exceed a value of (w/ρ):

$$\ln\frac{(C_S - C_0)}{(C_S - C_1)} = \frac{2\,(K_LA)}{V}\,\theta \qquad (10)$$

The limiting condition $C_2 = C_S$ is encountered for some situations of transfer in Bernoulli types of devices, and usually C_2 is an appreciable fraction of the saturation value. This represents a rather unusual situation because in most mass transfer equipment it is not usually feasible to operate under such conditions. Equation (7) and the limiting situation Equation (10) have not previously been reported.

Equation (9) has been commonly employed to evaluate performance of diffusers located in tanks and equipment of this type. Under such conditions C_1 is not distinguished from C_2, and C_1 is taken as the bulk liquid concentration. Many investigators [*(5)*, *(7)*, among many] have employed Equation (9) which leads to low values for the transfer factors. For example, a bubble rising through a liquid drags elements of liquid along at the interface. The concentration in the elements is greater than that of the bulk liquid, and the driving force is reduced accordingly. The actual transfer factor is higher than that indicated by Equation (9) up to a factor of 2 as indicated by Equation (10) if the liquid elements become saturated. The situation can be easily visualized as it would correspond to the present case for the Bernoulli device being submerged in a tank of liquid. The driving force obviously cannot be based upon bulk liquid concentration only.

It is sometimes the practice to define the transfer factor in terms of K_La *(5, 7)* which can lead to an incorrect performance evaluation if improperly determined. The term a is the interfacial area per volume of transfer device and is commonly applied to packed columns. (K_La) is not necessarily equal to (K_LA/V), and the volume of the system must be distinguished from the volume of the transfer equipment. For an external transfer device it is obvious that the interfacial area of transfer is independent of system volume V and that it would be erroneous to define $a = A/V$ for the device submerged in a tank of liquid. It is believed that use of K_La, even for diffusers in tanks, is incorrect unless gas bubbles are uniformly distributed throughout the entire system volume. It would be more correct, and also more informative, to report (K_LA) and V values independently. Any effect of volume on design could be easily evaluated, and separate values are quite as readily obtainable from the same data.

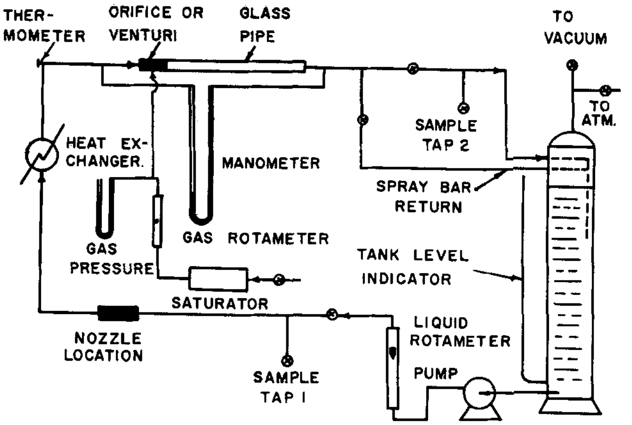

Fig. 1. Schematic diagram of apparatus.

Lack of knowledge of interfacial area, the fact that this area is changing because of pressure changes and coalescence, and that C_S and K_L may vary make any attempt at a strict analytical analysis exceedingly difficult. It was found experimentally that Equation (7) would give a straight line if the value of C_S employed was the limiting value obtained by operating the system until the liquid became saturated. Also, straight lines were obtained *(12)* by plotting ΔC across the device against C_1 and C_2 giving two lines which intersected at $\Delta C = 0$. This indicated a constant value for C_S near but below that computed from handbook values, and also that $\Delta C/(C_S - C_1/2 - C_2/2)$ was constant. This apparent constancy of C_S and (K_LA) must be somewhat fortuitous and involve compensating effects. It can be explained qualitatively by the fact that most of the pressure is recovered rapidly, immediately downstream from the throat, before much mixing of gas and liquid has occurred (velocity varies inversely as the square of the diameter).

Observations from a clear plastic Venturi (Figure 3) confirm the fact that mixing of gas and liquid does not begin until some distance downstream from the throat and that the gas does not entirely mix with the liquid until at least two-thirds of the downstream section has been traversed. At this point, the pressure is quite close to the discharge pressure. Because of the short times involved (up to 5 ft./sec. pipe velocity and 45 ft./sec. throat velocity) some transfer occurs immediately beyond the end of the device. Some error in C_S tends to be compensated because of the ratio indicated in Equation (7).

Transfer Coefficients by Change Across the Device

The transfer coefficient can also be obtained by simultaneous determinations of the upstream and downstream concentrations. Equations (2) and (3) are equated, and the driving force is expressed as a log mean to give

$$(K_LA) = (w/\rho)\ \ln\frac{(C_S - C_1)}{(C_S - C_2)} \qquad (11)$$

The value of (K_LA) given by Equation (11) should agree

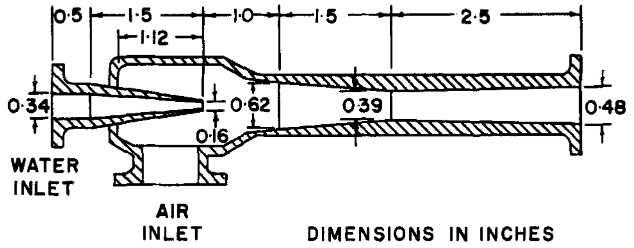

Fig. 2. Approximate dimensions of nozzle No. 2N.

with that obtained from Equation (7). The use of Equation (7) is preferred as minimizing experimental difficulties and as providing more accurate results.

Transfer factors for use of sulfite solutions are obtained from Equation (2) for $C = 0$ with the transfer rate determined from the change in sulfite normality in the system over a long interval of time.

System Characteristics

If Henry's law applies, a simple relationship exists between the liquid and gas film coefficients:

$$1/K_L = 1/k_L + 1/(H\ k_G) \tag{12}$$

When the same concentration units are used for both gas and liquid phases, H is dimensionless and for oxygen $H \cong 33$. Thus, the oxygen-water system represents the case of a slightly soluble gas where the gas-phase resistance is small or negligible. The saturation concentration of oxygen in water is sensitive to temperature; at 1 atm. pressure and 20°C. the solubility from air is 9.2 parts per million, and at 30°C. it is 7.6 (*1*).

Dilute water solutions of sodium sulfite have frequently been used to observe the transfer of oxygen from air. It is stated that use of this system permits the comparison of performance of various devices on the basis of constant tranfer rates ($C_1 = C_2 = 0$ at all times). It is necessary to use cupric or cobaltic ions as a catalyst in order to obtain fast reaction rates which are independent of sulfite ion concentration in the range 0.035 to 1.0 normal (*3*). It has been reported (*19*) that K_L values are the same for the air-water and air-sulfite systems, that sulfite ions lead to larger interfacial transfer areas, and that the presence of sulfate ions have an effect similar to that of the sulfite ions.

Energy Requirements

An important item in the evaluation of transfer devices is the quantity of solute transferred per unit of energy expended. For oxygen transfer to water the amount of oxygen (pounds) absorbed per unit of energy (horsepower-hour) will vary considerably with the driving force and becomes zero when the liquid is saturated; the fluid mechanics and the friction losses remain independent of concentration changes. To provide a consistent basis of comparison, performance is sometimes given as that which would obtain for transfer to water having zero oxygen content at an air pressure of 1 atm. and a temperature of 20°C. (*18*). The energy consumed by the device must be distinguished from the larger amount of energy supplied as the motive force to operate the device. Disregard of these factors makes difficult the accurate comparison of various devices reported in the literature.

In the present work the minimum energy requirement at standard conditions will be developed in accordance with Equation (13), where $C_S' = 9.2$ parts per million, the saturation value of oxygen from air in water at 20°C. and 1 atm. pressure:

$$(\text{lb. O}_2)/(\text{h.p.-hr.}) = 108.6\ (K_LA\ C_S')/(\Delta P\ w) \tag{13}$$

Table 1. Characteristic Dimensions of Transfer Devices Employed

Type of device	Throat or minimum dimension, in.	Descriptive information
Nozzle ejector		
1 N	1/8	Penberthy 1A-XL-96
2 N	5/32	Penberthy 62A-XL-96
3 N	3/32	Penberthy 81A-XL-96
4 N	1/8	Penberthy 182A-XL-96
Orifice	1/4 3/8	1½-in. diameter plate inserted into a modified ¾-in. brass union, one air inlet 0.18-in. diameter.
Venturi	1/4 3/8	6.0-deg. contraction section, 22.6-deg. expansion section; throat length equal to diameter; four air inlet holes, ⅛-in. diameter.

EXPERIMENTAL PROCEDURES

It was believed that energy requirements would be minimized if the devices operated without use of compressed air. The devices should perform such that the pressure developed at the throat is below ambient pressure with the result that air is aspirated into the device. Smaller throat diameters were required than necessary if a compressed gas had been employed. Diameters were also smaller than those necessary when such devices are used for one-phase flow rate measurements.

Three types of devices were employed: nozzle ejectors, orifice plates, and Venturi tubes. In each, air entered at the point of maximum liquid velocity and mixed with the liquid stream. Operation of orifice and Venturi devices was observed singly and with two of a given type in series.

Figure 1 is a schematic diagram of the apparatus which involved a closed (recycle) system with respect to the liquid; ¾-in. I.D. copper piping was used throughout. Oxygen was removed from the liquid prior to a run by vacuum (liquid flow through the spray bar) or by sweeping with nitrogen. During transfer the gas space at the top of the tank was maintained at atmospheric pressure.

Characteristic dimensions of the various devices are indicated in Table 1 and a nozzle ejector is shown in Figure 2. Figure 3 is a sketch of the Venturi. A 1-in. glass pipe section, 36 in. long, located immediately downstream from the orifice and Venturi positions, permitted visual observation of gas-liquid flow.

The liquid phases employed were distilled water, sodium sulfite solutions, water with a wetting agent, and a simulated industrial waste (caustic potato peel waste). The liquid volume of the system was maintained at 5.1 to 5.5 cu. ft. The air was saturated with water vapor prior to entering the transfer devices.

Analysis for oxygen was by polarograph [dropping mercury electrode, 3 sec. drop time, saturated calomel reference electrode, 0.01 N potassium chloride supporting electrolyte, 15 parts per million methyl red maximum suppressor, (*13*), (*16*),

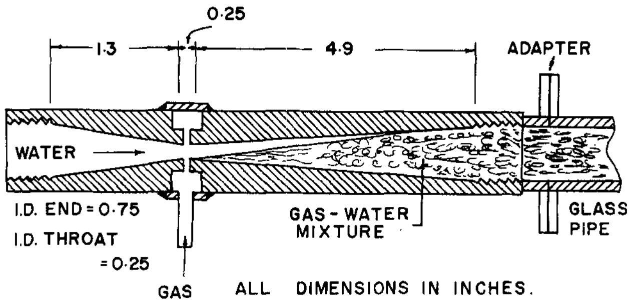

Fig. 3. Venturi device for mass transfer.

among many]. Calibrations and checks of the polarographic method were made with the standard Winkler procedure (*1*).

Analysis for sulfite was by an iodometric procedure (*1*). The sodium sulfite solution was approximately 0.2 normal with cupric sulfate added as a catalyst (0.001 molar). At this concentration the reaction rate is considered very rapid (*3*).

The manufacturers calibration for the gas flow rotameter was checked against a wet test meter with the various gases and found satisfactory. The liquid flow rotameter was calibrated for the conditions of use.

A first objective was to observe the performance of the nozzle type of device for several designs, a second was to compare the three different types of devices, and a third was to further evaluate the most promising type of device for systems other than water.

NOZZLE DEVICES

Hauxwell (*10*) compared several designs of nozzle ejectors as to transfer factors and energy consumption and concluded that the No. 2 N device gave the best performance. It was observed that in no case did the values for ($V\,m$) exceed the limiting value of (w/ρ). K_LA values calculated from Equations (7) and (11) showed good agreement, although for some runs the downstream values of concentration were unreliable.

Of the four nozzles tested, No. 2 N had the largest throat diameter, provided for the free flow of air, and had contraction and expansion sections both upstream and downstream. These factors resulted in the best design as judged by mass transfer requirements. This nozzle was selected for later comparison with the orifice and Venturi devices. In a larger size, this device has been employed in waste treatment installations (*15*). Performance claims are usually based on the use of supply of compressed air supply to the device, possibly because of the submergence effect. Compressed air may be necessary if the device operates under a considerable head of water at the discharge.

ORIFICE AERATORS

Ho (*11*) compared two sizes of orifice with the air introduced just downstream from the orifice plate at the vena contracta. Because the 3/8-in. orifice would not create a sufficiently low throat pressure for air to enter the device most runs employed compressed air. The 1/4-in. orifice would aspirate air satisfactorily.

One effect observed was that at a given air rate water above a certain flow rate passed through the orifice as a jet stream, and intimate mixing of the liquid and gas phases did not occcr. Figure 4 indicates the air and water rates at which phase separation occurred. These represent limits of operation for the orifice device because the transfer rates become minimal (small interfacial areas result). Below this limiting condition the effect of increasing air or water rates is to increase the transfer coefficient, although the effect of air rate becomes relatively less as the separation region is approached. At higher air flow rates the liquid downstream closely approached saturation.

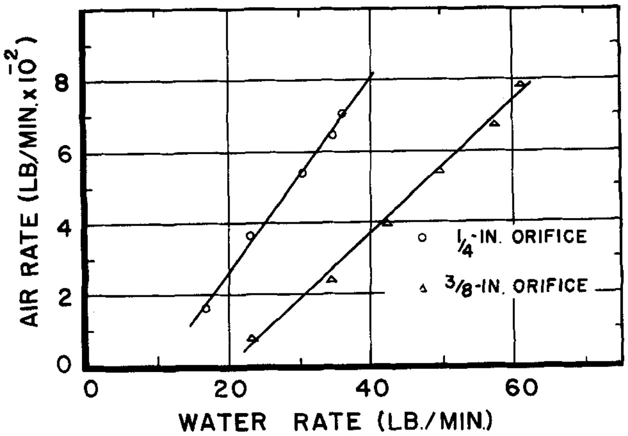

Fig. 4. Flow rates for phase separation at orifice.

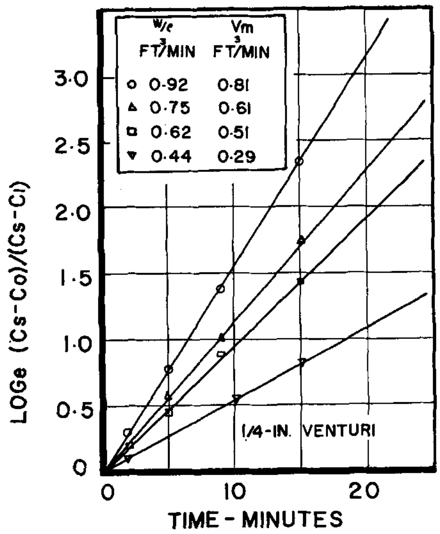

Fig. 5. Typical plot to determine transfer factors by Equation (7).

The 3/8-in. diameter orifice had a lower energy requirement as judged by the pressure drop than that for the 1/4-in. device. However it was apparent at certain conditions that a considerable part of the energy consumed in the process was suppled by the compressed air stream. Two orifices in series, with air being supplied only to the first, did not show any significant increase in performance over the use of a single device. The additional orifice, located 18 in. downstream, would have the effect of expanding the air and reforming interfacial area presumably leading to additional transfer. Passage through the first orifice nearly saturated the stream, and the second device offered no advantage from increased mass transfer. An additional pressure loss was introduced by the second orifice. The use of such an arrangement might be advantageous at lower liquid and air flow rates.

VENTURI AERATOR

A Venturi made of clear plastic, similar to that of Figure 3, was employed to observe flow patterns. The flow pattern with three of the four holes blocked is indicated in Figure 3. The air enters as a thin layer apparently at very high velocity. Mixing proceeds in almost a flat plane across the downstream section, and the gas-liquid mixture does not completely fill the entire section until a short distance from the end of the expansion section. Rapidly fluctuating flow is apparent. The air-water mixture fills the entire cross section for about 6 in. of the glass pipe; separation is complete at about 18 in. Near saturation occurs in this short distance and short exposure time.

Gimpel (9) investigated the Venturi as an aeration device under a variety of flow conditions and compared performance with that of the nozzle and orifice for similar methods of analysis and operation. No compressed air was employed, and all data (except one run) were for air aspiration at a variety of water flow rates. The 3/8-in. Venturi would not develop throat pressures sufficient for satisfactory air aspiration.

Figure 5 indicates the method employed to obtain (K_LA) values for the unsteady state analysis in accordance with Equation (7) (a similar procedure was applied to

TABLE 2. PERFORMANCE OF A ¼-IN. VENTURI AERATOR; COMPARISON WITH ORIFICE AND NOZZLE AERATORS (air aspirated)

Device	Water rate, lb./min.	Air rate, lb./min.	C_S, p.p.m.	ΔP, in. Hg	cu. ft./min. (w/ρ),	$(V\,m)$, cu. ft./min.	(K_LA), cu. ft./min.	Standard lb. O_2/hp.-hr.
Air-water system (26 to 30°C., 689 to 695 mm. Hg barometer)								
¼-in. Venturi	58	0.0104	7.90	17.3	0.92	0.81	1.45	1.5
	47	0.0097	7.51	10.4	0.75	0.61	1.02	2.1
	39	0.0091	7.34	6.9	0.62	0.51	0.86	3.2
	27	0.0067	7.24	3.2	0.44	0.29	0.51	5.9
¼-in. orifice	39	0.0100	6.92	30.6	0.62	0.54	0.96	0.8
	37	0.0100	7.10	31.2	0.60	0.53	0.95	0.8
	27	0.0040	7.24	15.4	0.44	0.22	0.29	0.7
Nozzle 2 N	27	0.0170	7.40	39.8	0.44	0.47	0.88	0.8
¼-in. Venturi	58	0.0074	6.71	—	0.92	0.61	0.91	
+ ⅜-in. Venturi	58	0.0104	6.71	—	0.92	0.83	1.51	(compressed air)
Air-sulfite solution (30°C., 689 mm. Hg barometer, $C_S = 692^+$ calculated)								
¼-in. Venturi	58	0.0104	+	18.7	—	—	2.11	2.0
	47	0.0097	+	11.2	—	—	1.38	2.6
	39	0.0091	+	7.5	—	—	1.03	3.6
¼-in. orifice	39	0.0110	+	33.0	—	—	1.36	1.0

the data for the nozzles and orifices). In all cases a straight line resulted which passed through the origin. Data scatter occurred for longer times corresponding to approaching equilibrium and calculations disregarded this region. Table 2 gives some results for the ¼-in. Venturi. Transfer factors increased almost linearly with liquid flow rate, and operation of the Venturi was much more efficient, with respect to energy requirements, at lower liquid rates.

The ⅜-in. Venturi, placed immediately downstream from the ¼-in. Venturi with the latter aspirating air, offered no transfer advantage and pressure losses were larger. The aspiration rate was lowered, and when compressed air was supplied to raise the air flow rate to the normal values, the transfer factor was similar to that for only one Venturi. Operation at lower liquid rates might have shown improved performance for the double Venturi procedure.

With sulfite added to the water $C_1 = C_2 = 0$ and the transfer process is not limited by downstream saturation. In Table 2 (K_LA) values are seen to be 20 to 45% higher for the sulfite solutions as compared with water. Bubbles tended to persist longer in the downstream glass pipe section which added to the transfer area.

The amount of oxygen transferred per unit of energy expended is more favorable for the use of sulfite, although pressure losses were a little higher because of the more persistent downstream bubble-water mixture. Higher transfer areas should result for liquids containing surface active materials such as for industrial wastes. Foam formation, if not too persistent, would prolong bubble life and lead to a closer approach to saturation. This effect would be especially significant at lower liquid rates where energy losses are much reduced. No difficulty with foaming occurred for the liquid containing sulfite even at the highest flow rates.

A surface active agent was employed at 0.01% concentration in water at flow rates corresponding to those for water. At all flow rates the downstream liquid was saturated giving $(K_LA) = 2(w/\rho)$. Transfer factors were 25 to 75% higher than the corresponding values for water resulting in better energy utilization. Reductions of 25% in transfer coefficients upon the addition of a surfactant are reported (*4*), where the area of transfer remained unchanged.

A simulated caustic potato peel liquid waste was prepared and employed with the ¼-in. Venturi. This liquid had a pH of 11, a soluble solids content of 4%, and an interfacial tension of 17 dynes/cm. below that of water. (K_LA) values at the two highest liquid flow rates corresponded to saturation and at the two lowest liquid rates were only a little below saturation. This confirmed the expectation that performance of the Venturi device would be better for the aeration of industrial waste containing surface active material (proteins in this case). Foaming of the potato waste became quite pronounced at the highest flow rate, and fine bubbles tended to be entrained in the liquid pumped from the bottom of the tank. The foam did not persist and did not overflow the top of the tank.

PERFORMANCE COMPARISON

The Venturi and orifice may be compared directly as to performance. The minimum flow area for the nozzle was smaller, and it is not known exactly how the performance of this device varies with diameter. A nozzle with a ¼-inch throat might have shown somewhat lower energy consumption, and any comparison should be qualified by this difference in design. However, it is apparent that the nozzle could not surpass the Venturi in performance from the standpoint of energy consumption.

Table 2 indicates the distinct advantage offered by the Venturi in the amount of oxygen transferred per unit of energy expended which exceeds that of either the orifice or nozzle by factors of 4 to 8. The value of 5.9 (lb. oxygen)/(hp-hr.) surpasses the performance of most aeration devices in so far as this can be judged (*7*). Venturi performance is better at lower flow rates and for liquids of lower interfacial tension. Liquid flow capacity is also greatest for the Venturi for a fixed pumping pressure. The performance of the Venturi in a larger size has been considered (*12*).

The permanent losses in pressure across the devices were substantially different as indicated in Figure 6 for flow of water only and for water-air flow. The air admitted was that aspirated by the device. The pressure drop was more than doubled when air was admitted to the Venturi device. The orifice arrangement showed almost no difference with or without air aspiration, and the nozzle showed a small increase over the already large pressure drop when air was admitted. The apparent advantage of the Venturi as a mass transfer device, over the orifice and the nozzle, is due mostly to the behavior apparent from Figure 6. The Venturi design minimizes pressure loss for single-phase flow, and this is also the case for two-phase flow of

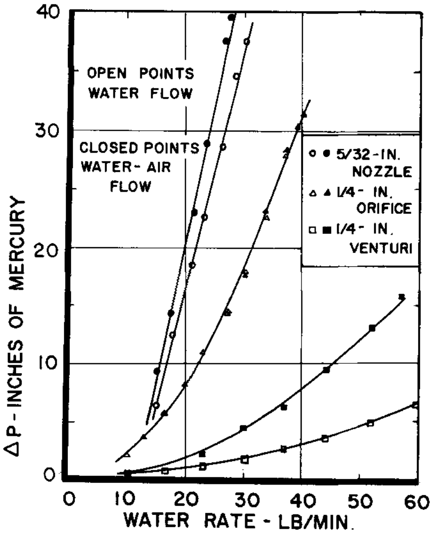

Fig. 6. Pressure losses with and without air input.

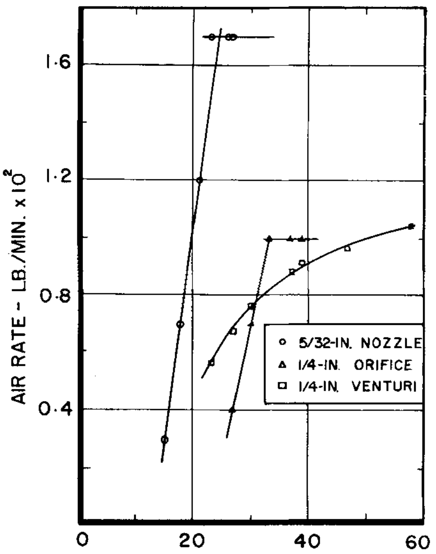

Fig. 7. Aspirated air flow rates.

a liquid and gas. The loss characteristics for the orifice, being the same for liquid or liquid-gas flow, are probably the result of high downstream turbulence for single-phase flow which remains relatively unchanged for two-phase flow. The high-pressure loss for the nozzle results from the design to produce high liquid velocities for use as an ejector.

Figure 7 shows aspirated air rate as a function of water rate, where a limiting air flow rate appears to be indicated for each device. Calculations give approxmate throat pressures approachng the vapor pressure of the water at the higher flow rates. This is a limiting condition leading to incipient cavitation which could be observed with the plastic Venturi. Desorption of the solute at the throat could also be observed for the saturated liquid. Throat pressures could not be calculated precisely because the presence of the air speeds up the water flow over that for single phase flow. For example, at the highest flow condition for the ¼-in. Venturi the water rate is approximately 1 cu. ft./min.; the air rate is about 0.1 cu. ft./min. at standard conditions and is larger than this by an unknown factor at throat conditions because of expansion to local pressure.

The higher air aspiration rates of the nozzle result from less restriction to air flow in the lines leading to the device and in the device itself (Figure 2). Although air flow rates in the lines and entrances to the orifice and Venturi did not approach sonic conditions, it is conceivable that at the air inlets to the throat (Figure 3) and immediately downstream, points of very low pressure, sonic flow conditions might occur. This would contribute to a limiting air flow condition.

A recent paper (2) provides data on the operation of a Venturi with an upstream diameter similar to that of the present work but having nearly twice the throat diameter. Carbon dioxide gas was supplied under pressure and absorbed in water. Sample and pressure taps were provided in the expansion section of the Venturi. Results were reported as a percentage of saturation at local points and are not directly comparable to present results. Photographs are provided of the mixing in the downstream section at various flow rates. The behavior agrees qualitatively, at least, with the results reported herein.

SUMMARY AND CONCLUSIONS

Nozzle, orifice, and Venturi devices were employed to effect oxygen transfer from air to liquid phases. A recycle liquid system was employed for the unsteady state analysis of the mass transfer process. Equations are presented to properly interpret the unsteady state procedure, and limiting factors for the transfer process are defined. The minimum energy to effect transfer was determined for the devices which provided a basis for the comparison of performance.

Four types of nozzles were investigated, and the design showing the best performance was selected for later comparison with the orifice and Venturi. High transfer factors were observed for the nozzle type of device, but capacity was lower, and more energy was required, than for the Venturi to transfer a given amount of oxygen.

Orifice aerators, supplied with compressed air, were used to observe the effect of various air rates. Higher air rates, for a fixed liquid flow rate, resulted in higher transfer factors of the order of 25%. The air rates employed were high relative to normal air aspiration rates, and the exit water stream was at or near saturation for the conditions observed.

The ¼-in. Venturi was investigated under conditions of air aspiration. Pressure drops across this device were relatively low, and performance in terms of amount of oxygen transferred per unit of energy expended was very favorable, up to 5.9 lb. oxygen/hp.-hr. Compared at the same water flow rate but at air aspiration rates peculiar to the device, the performance of the Venturi was at least seven times better than that of either the nozzle or orifice. The amount of oxygen transferred per unit of energy supplied was much more favorable at low liquid flow rates.

The use of sulfite solution, with a catalyst added, resulted in higher transfer factors and better energy utilization in the Venturi. The addition of a wetting agent to water achieved higher transfer factors at the higher liquid rates with saturation of the exit liquid. A similar behavior occurred for transfer to caustic potato peel liquid wastes selected to simulate an industrial waste. The higher transfer factors resulted from reduced interfacial tensions and correspondingly higher interfacial transfer areas. The use of sulfite solutions served as an approximate check of the unsteady state method.

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NOTATION

a = interfacial area per volume of transfer device, sq. ft./cu. ft.
A = interfacial transfer area, sq. ft.
C = solute concentration in liquid, lb./cu. ft.; C_i, at the interface; C_S, in equilibrium with bulk gas phase; C_0, at time zero; C_1, upstream from the device and uniform tank concentration; C_2, downstream; (C' refers to parts per million $= C \times 10^6/\rho$)
H = Henry's law constant for the same concentration units in gas and liquid phases, dimensionless
k_L, k_G = liquid and gas film transfer coefficients, ft./min.
K_L = overall liquid transfer coefficient, ft./min.
m = slope of the line resulting from a plot of Equation (7), liter/min.
ΔP = pressure drop across device, in. Hg
r = rate of transfer, lb. mass of solute/min.
V = liquid volume in system, cu. ft.
w = liquid flow rate, lb. mass/min.
θ = time, min.
ρ = liquid density, lb. mass/cu. ft.

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Mass Transfer in a Fluidized Bed Crystallizer

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Continuous crystallizers represent a wide range of mass transfer equipment types, from stirred tanks to tubular reactors. Except for stirred tanks however there are no analyses of operating characteristics which can be used for design or scale-up calculations. In other types of crystallizers the variables are inter-related in a complicated manner, and few of the basic design data are available. Even for the case of a stirred tank, where there are fewest variables, the basic kinetic data are scarce. Crystallizing velocities have been measured for many materials (*1*), but few are in the form suitable for plant design and few have been measured under conditions even approximating plant operation.

It was the purpose in the present experiments to examine the characteristics of the so-called *classifying* type of crystallizer. The extent of size classification in a fluidized bed is small however, and it has been shown (*2*) that there is an apparent advantage to be gained by increasing the stirring action in this type of machine. The complex pattern of the supersaturated liquor distribution under such conditions makes it practically impossible to analyze or predict performance. If a fluidized bed crystallizer is designed and operated to approximate to plug flow of the liquor, then an analysis of the mass transfer processes in terms of the diffusional effects and the surface reaction would be simplified. The important secondary processes such as the generation of new nuclei, the effects of suspension density, etc., cannot be correctly interpreted until the most fundamental processes are understood.

It was shown in an earlier paper (*3*) that if a fluidized bed crystallizer is regarded as a tubular reactor with a low dispersion number, that is essentially plug flow, certain characteristics can be calculated from the basic rate equation for crystallizing velocity expressed as

$$dL/dt = k \cdot (N'_{Re})^b \cdot S^n \qquad (1)$$

$(N'_{Re})^b$ takes the effect of eddy diffusivity of the liquor into account in an empirical manner, and when the liquor density and viscosity can be taken as constant, this term can be conveniently abbreviated to L^b. While plug flow of the liquor was assumed for each case, it was shown that the behavior of the suspended crystals affects the characteristics of the process. Tentative solutions were proposed for the cases of complete mixing of the crystals in the bed and for exact size classification. Bennett (*4*) has shown that the size distributions of crystals, taken from industrial machines, differed from those predicted for the simplified models. This is to be expected since the ma-