Biotech Advs vol.1,pp 17-30, 1983 0734-9750/83 $0.00.+.50

TRANSPORT PHENOMENA, REACTOR DESIGN AND SCALE-UP

JAMES Y. OLDSHUE

Vice President, Mixing Technology, Mixing Equipment Co., Inc., A Unit of General Signal Corp., Rochester, New York, USA

ABSTRACT

This review will cover the area of impeller-mixed stirred-tank reactors. In addition, it will consider bubble columns, in which air or gas is passed up a liquid filled column through distribution plates covering the full area of the column, and also airlift reactors, in which the air is confined in a channel by means of a loop or draft tube designed to impart a certain type of overall circulatory pattern to the entire tank.

There is considerable interest in the kinetics inside the solid part of various kinds of immobilized solid pellet type of enzymes and catalysts. The use of these particles in fixed bed reactors is also covered.

KEYWORDS

Mass transfer, heat transfer, mixing, aeration, biological reactors, blending, scale-up, fermenters, air-lift, enzyme reactors, aerobic, anaerobic, bubble columns, loop reactors.

IMPELLER-MIXED REACTORS

One of the basic design parameters is the power consumption of an impeller under a wide range of fluid and gas flow conditions. Work by Gray (11) discusses a new correlation of the power drawn by an impeller and the various parameters of speed, diameter, liquid flow and gas flow rate. The correlation involves the total power which is the sum of the impeller power and the gas expansion power. The correlation is said to be accurate to within ± 20%, but it includes a quantity, the hold-up of gas in the liquid, which is normally not known in advance.

Actually, the power of an industrial mixer needs to be calculated more accurately than the estimated 20%, so that considerable adjustability in the impeller configuration may be required if close control of power applied to the impeller is needed over a variety of operating gas flow rates.

A review article by Van't Riet (37) summarizes many articles giving data on mass transfer rates in gas liquid systems for batch and continuous steady-state systems. The mass of gas entering and mass of gas leaving give an unequivocal measure of the gas absorption rate. For correlation, a calculation of the mass transfer coefficient, $K_G a$ and $K_L a$ is normally needed. This involves knowledge of the concentration of solute from the gas in the liquid. If this is oxygen, then there are several methods available of various degrees of accuracy for measuring the dissolved oxygen level in liquid. Depending upon the blend time in the tank, questions of the instantaneous and average dissolved oxygen concentration are appropriate. However, in order to use the correct units, either the gas phase partial pressure must be converted to a liquid phase equilibrium concentration by means of Henry's Law, giving the C^*, or the liquid phase concentration, DO, must be converted to an equilibrium partial pressure, P^*.

There is often discussion as to whether to use $K_G a$ or $K_L a$ depending on where the major resistance lies. However, since we only know bulk concentration in the gas and liquid phase, it does not make any difference which driving force we use, ΔC or ΔP, and the use of $K_G a$ or $K_L a$ is arbitrary. Studies of gas phase and liquid phase variables will affect either of the $K_G a$ or $K_L a$ appropriate to their effect on the system, and it does not do much good to speculate as to where the resistance lies as far as the use of $K_G a$ or $K_L a$ is concerned.

Using sodium sulfite solutions to measure mass transfer rates give numbers that are only applicable to that particular reaction and concentration of ions. Having the data can be of value in sorting out the effect of major variables, such as mixer and tank geometry, and sparge ring geometry. Mass transfer rates with fast chemical reactions give a normal situation in that the p^* is close to zero. This means that there must be an empirical or experimental correlation to translate from sulfite data to another type of system.

The dynamic gassing out method as discussed by Van't Riet has problems of the response time with the system to sudden changes in gas rate, gas concentration, fluid shear rates, mixer variables, and blend time, which affect the mechanism of the whole operation. This method is of limited applicability, and a considerable number of tests must be made to make sure that the mass transfer rate measured is appropriate to the mixing and mass transfer dynamics of the whole system.

The measurement of interfacial area gives one component of the $K_G a$, but unless it is coupled with the K_G, it does not give the overall correlation. Physical methods of measuring "a", such as photographs or chemical methods, yield widely different results. In addition, a procedure must be established to relate instantaneous "a" values in the tank with the overall average value of the "a" throughout the tank.

It is pointed out by Van't Riet (37), that $K_G a$ results can be correlated by the expression, $K_G a \propto (P/V)^{\alpha} (F)^{\beta}$.

Another factor which is mentioned by Oldshue (26) is that the ratio of the energy put in by the gas and the energy put in by the mixer affect the mass transfer correlation. For example, if the energy put in by the mixer and the gas were equal, exponents alpha and beta are usually lower than if mixer energy were three times higher than the gas energy. This means that correlations from the literature which bracket these areas will encompass these different ranges of the ratio of mixers to gas energy, and will have a different exponent alpha or beta depending upon the experimental range used, and the technique used for giving an average alpha and beta.

Another problem discussed by Oldshue (26) is that when working with small scale experiments, if the impeller blade gets physically out of proportion to the gas bubble, comparisons of small and large scale systems are skewed inappropriately compared to the scale up correlation in which the impeller blade is more than two or three times bigger than the gas bubbles. A comparison is "to hit a baseball with a baseball bat compared to hitting a basketball with a baseball bat". The effect of ion concentration is very remarkable as discussed thoroughly by Van't Riet (37). Of the major impeller geometry variables, the effect of D/T ratio and the ratio of the sparge ring diameter to the impeller diameter ratio has turned out to be of major importance. In general, power per unit volume and superficial

gas velocity are the major variables needed to correlate the $K_G a$ term assuming other chemical and gas phase relationships are appropriate to the system under study.

Another method of study involves the use of batch liquid, and the measurement of absorption of oxygen or other solute with time in the batch liquid, the progress being measured by suitable probes, or the removal of the solute by suitable stripping gas. These methods have a host of operating problems. The uniformity of the solute in the liquid depends upon relative mass transfer and the blending rate, and therefore changes during the run. The gas phase off gas is continually changing during the run, and the instantaneous relationship of the off gas concentration to the solute measured depends upon many different kinds of fluid mechanics and mixing dynamics. In addition, in the case of dissolving oxygen in water, when saturation is reached, the dissolved oxygen level, if blending is rapid, will be a mean of the concentrations at the bottom and the top of the tank. Therefore, the tank will be absorbing in the bottom part of the tank and stripping the top part, which is not at all the usual situation of the tank in steady state use in the process.

In addition, the velocity head generated by the impeller complicates the equilibrium solubility of the solute, and as other variables, it must be correlated. The chemicals used in the oxygen removal step prior to the run also build up during a series of experiments and cause many problems with their effect on the mass transfer coefficient. In addition, the current practice of using the uptake rate at zero solute concentration means that data are always being extrapolated back into a range prior to the first several experimental data points, thus resulting in a variety of problems. It is equivalent to trying to measure the acceleration of a racing car at zero time, from velocity measurements made during the acceleration up to full speed.

In terms of some practical methods of controlling oxygen uptake with computers, Spriet (36) presents a study called "Static Method'. It shows that the computer controlled accuracy depends upon the precision of the oxygen analyzer.

The actual measurement of oxygen mass transfer coefficient has been treated by several investigators. An article by Vardar (38) used a frequency response technique by controlling the air flow rate by a thermal mass flow

controller, and the pressurization of the vessel was controlled by allowing valves connected to four orifice tubes with different inner diameters to open at certain times. The opening and closing times of the orifices were determined by a cam drive which could also be adjusted to give different cycling periods. The diameters of the orifice tubes were chosen so that the pressure cycle would possess a sinusoidal characteristic.

A paper by Linek (20) looked at the role of inter-phase nitrogen transport, in the measurement of the overall volumetric mass transfer coefficient. Experiments were conducted in two different ways. In one method, the interchange of oxygen and nitrogen in air were performed without either interrupting the aeration or agitation of the charge. The second method was to remove the dissolved oxygen, and begin the aeration-agitation at the same time. It was found that the technique of accounting for the oxygen transport gave values that were independent to some degree, of the nitrogen transfer mechanism for the first method, while the second method gave large differences between accounting or not accounting for the nitrogen mass transfer.

A report by Ruchti (33) looked at six different models of the dynamic oxygen electrode method for measuring $K_L a$. In general, the found that $K_L a$ should be less than the inverse electrode response time. They present a method which accounts for gas, film and electrode dynamic effects, and requires only a simple semilog plot of response time. In viscous gas liquid systems, there is a fraction of very tiny bubbles, less than one millimeter, and it is expected that the oxygen tension in these bubbles will be in equilibrium with that in the liquid within seconds. This "liquid-small bubble dispersion", may be considered a homogenous phase, according to Heijnen (14), and the use of dynamic $K_L a$ method in viscous gas liquid systems can be quite problematical. Andre (2) looked at the problems when the substrate in, for example, a cellulosic waste, is insoluble. He made a change by means of a step input of CO_2 in the inlet gas stream and found that by taking into account the difference in diffusivity of oxygen and carbon dioxide, preliminary results indicating good mass transfer data could be obtained.

For use of mass transfer data in an actual microbial process, there are many variables that must be considered. Kappelli (15) used a yeast system as a means of measuring the maximum possible oxygen uptake rate in a

reactor and felt that this gave a typical representation of the reactor for other fermentation calculation purposes. Linek (21) used glucose oxidase to measure the oxygen absorption in fermenters. They used techniques employing both the dynamic and steady state method, and found that in certain areas, the dynamic method gave erroneously lower K_La values if the K_La value was higher that 0.03 s^{-1}. Other complications arose at K_La values around 0.08 s^{-1} and simultaneous interfacial transfer of nitrogen and oxygen had to be taken into account in some of these cases.

Einsele (9) looked at a tank completely filled with liquid, which had essentially a marine-type propeller and a draft tube. A gas liquid separator was at the top of the vessel. They found that the gas liquid separator acted as though it were another mixer. The blend time in this reactor was 50% of that for a traditional turbine-stirred gas-liquid unit at the same power level.

One of the important parameters in a fermentation study is the oxygen solubility in the fermentation medium. Quicker (30) was able to develop a solubility model where the solubility reduction is log additive with respect to various compounds, mainly sugar electrolytes.

Firevod (10) illustrated the exceptional accuracy of a galvanic probe measuring low oxygen concentrations for certain types of fermentations on yeast and other facultative anaerobes which require oxygen for lipid synthesis in order to grow and ferment.

There has always been a speculation as to the availability of oxygen by direct gas solid uptake into living organisms. Sobotka (35) gave experimental data which looked at this phenomena and showed that a two-phase model was effective in predicting mass transfer coefficients.

Wick (42) gave some calculations on the direct liquid-liquid heat exchange in continuous bio-reactors for very low microbial heats of activity.

Brown (4) discussed the changes in variables on scale-up. Mixing variables were treated in detail. Geometric similarity causes changes in important process ratios on scale-up, and non-geometric design may be needed to control selected parameters.

The additional pressure in large tanks over that in pilot scale caused a decrease in the antibiotic productivity of asparagine and neomycin accor-

with superficial gas velocity. Initially the $K_L a$ decreased to a minimu value when the liquid velocity was in the order of 7.5 cm/s, and then increased at higher liquid superficial velocities. The article (1) pub lished an extensive series of profiles in these tanks and concluded tha a two-zone model should be used in which the $K_L a$ data is split between *grid zone and the bulk zone. Shah (34) gives a very extensive review o* design parameters for a bubble column reactor. He presents a series of illustrations of the various kinds of reactors and modifications, and g an extensive list of some 20 processes that have been published concern the application of these columns industrially.

Shah (34) also mentions the homogeneous (bubbly flow) regime, and the t heterogeneous areas, churn turbulence, and plug flow. He shows the area where homogeneous or heterogeneous churn turbulent flow occurs on large size columns. He shows that columns wider than about 15 cm in diameter needed to obtain data which are relevant to large size units. Curves a included for gas hold-up, gas liquid interfacial areas, and extensive d on mass transfer coefficients. In general, $K_L a$ seems to increase in proportion to the gas phase velocity to the exponent 0.8. Data on the liquid/solid mass transfer coefficient and on some columns having gas/ liquid/solid phase present, and heat transfer relationships in bubble columns, are also given.

There have been a different group of findings on the volumetric mass transfer coefficients in CMC solutions in bubble columns, and Deckwer (went back and did some new experiments in a 14-cm diameter, 270-cm high bubble column. The $K_L a$ values were determined as well as dispersion coefficients by fitting the predictions of the axial disperse plug flow model with experimental oxygen concentration found in the liquid phase. His correlation described $K_L a$ values measured in fermentation broth, *Pe cillium chrysogenum*, with excellent agreement. Roels (32) presented a non-equilibrium thermodynamic approach to the power dissipation and the heat production of bubble columns.

SOLID-PHASE BIOCATALYST REACTORS

A group of papers treat the diffusion and kinetics inside the particles of immobilized enzymes either on the exterior or interior portions of a solid support. These papers are listed in the bibliographies primarily

source of additional information:Parke (28), Verhoff (39), Lee (19), ma (27), Kulkarni(18), Do (7), and Webster (41)

ies of articles describe the performance of packed bed reactors: ff (40), Patwaidhan (29), Karanth (16). These are given for reference ses.

o-reactors which have a solid substrate attached to a surface, fluid rate has an effect on the thickness and diffusion in these slimes. er by Duddridge (8) looked at a radial flow growth chamber to study nitial phases of bacterial adhesion to surfaces under flowing con- ns. He found the maximum levels of adhesion occurred in general of lower surface shear rate, particularly less that 6 to 8 pascals. dhesion was still noticeable up to a shear stress of 130 pascals. attached under static conditions could be detached at surface shear of about 10 to 12 pascals.

er paper by Rittmann (31) showed the effect of shear stress on the of product on the bio-film reactor. A study by Chen (5) looked at n transfer in filter slimes by means of a microelectrode.

ANAEROBIC PROCESSES

moto (12) studied the effect of mixing duration and vacuum on methane ction rates from anaerobically fermented beef cattle waste. Continuous- xed fermenters seem to produce higher methane production rates than nters mixed only two hours per day. However, the continuously mixed nters were only 8 to 11% higher compared to intermittently mixed fermen- at 6 and 4 days in hydraulic retention time. They concluded there was e potential for increasing the anaerobic fermentation rates of live- waste by increased mixing or vacuum.

obic systems have a requirement for suspension and nutrient supply.

eau (3) found that the activity of the anaerobic organism was much r at 700 rpm than 400 rpm, but a further increase to 940 rpm caused uction in the rate of growth in enzyme production. His study d at the growth of *Citrobacter intermedius*. Heertjes (12) looked uid flow patterns in up-flow reactors used for the anaerobic treat- of beet sugar factory waste water. He worked in a $30m^3$ pilot

ding to Kaszab (17).

AIRLIFT AND LOOP FERMENTERS

Airlift fermemters are characterized usually by a central draft tube, in ich gas is admitted to a central tube, and forms a circulation pattern throughout the vessel. These have an appeal in eliminating the maintenance associated with the conventional stirred tank mixers. An unpublished June, 1983 presentation by Robinson and Moo-Young at the ACHEMA in Germany, indicated the obvious truism that you can always get higher oxygen mass transfer coefficients with mechanical mixing added to whatever gas rate is used. If the organism can use this uptake rate, and if it is desired to maximize the productivity of a given volume of reactor, then this is the direction that fermentations have traditionally taken in the past. However, if the organism does not need this high uptake rate, or if optimization of amount of oxygen tranfer per unit of capital and operating costs is desired, then airlift fermenters need to be evaluated carefully to see what the economics of the operation are. The oxygen uptake rate per reactor volume will be lower than with mechanical stirring.

Mechanically stirred fermenters are normally available with some process and mechanical design know-how as supplied by both the equipment supplier and the user. Airlift fermenters are normally based on purchasing a compressor of a certain known volumetric output, and no consideration of the process and mechanical characteristic in the reactor are available from the compressor supplier.

Margaritis (24) studied the effect of draft tube geometry with four jets at e bottom. He used various single and double draft tubes. They found that the air bubble formation characteristics were different with the various draft tubes. This explained differences observed in mass transfer and mixing characteristics. Their power levels range from about 0.02 to 0.25 kW/m^3. Typical stirred fermenters go up to the range of 4 kW/m^3 when oxygen requirements are suitable for these power levels.

Merchuk (25) considered a previously presented model and extended it by considering the range of pressure along the tubes. His new model allows the prediction of oxygen concentration at different points of airlift fermenters and how to determine the best value for the gas flow rate. Luttmann (22) set up a distributive parameter model for the simulation of single cell

protein production using reactors with an outer loop. He considered variations of the substrate concentrations. CO_2 concentrations in the liquid, and O_2 and CO_2 phase concentrations in the gas phase. He took into account variations of dissolved oxygen concentration and pressure and $K_L a$ along the column. He used his model to describe the cultivation of *Hansenula polymorpha* in a tower loop reactor 275 cm high and 150 cm in diameter.

Luttmann (23) also took the data from the smaller reactor and used it to simulate the cultivation process in a 40-m high production reactor. This model was simplified somewhat to examine variables in a 20-m high pilot plant airlift loop reactor. Depending on the economics, they determined that the maximum profit was attained at the boundary between substrate and oxygen transfer limited growth.

Ziegler (43) used a 22-m long, 20-l tubular fermenter for oxygen transfer characteristic tests as a reactor for mycelial growth. $K_L a$ values were correlated for power consumption and aeration rates. They used a variety of cultures, and show that the product spectrum on some of these were dependent upon the type of reactor used. They used power consumptions up to 8 kW/m^3 in the tubular reactor, which did not appear to harm the mycelia.

BUBBLE COLUMNS

Bubble columns involve a sparger or distributor covering the entire area of the column. They have been used for many years in various types of chemical processing, but their use in the bio-reactors is relatively new. Studies of their process characteristics is quite different than the approach to stirred tank fermenters, since the user buys primarily an air supply, and no mixing or mass transfer experience comes with this. In an article published in 1981, Alvarez-Cuenca (1) presented data on a consideration of three different models for mass transfer and mixing. He looked at the axial dispersion model, a plug flow model, and a two-zone model which included a grid zone around the distributor, and a bulk zone in the rest of the column. Most investigators have concluded that columns must be wider than about 15 cm in diameter to give meaningful scale-up relationships and it is difficult to obtain these large flow rates on experimental equipment.

Most studies indicate that $K_L a$ increased, in some cases almost linearly,

plant and a 200 m^3 plant reactor.

NOMENCLATURE

a - *Interfacial area*; *DO* - *Dissolved Oxygen*; C^* - *Equilibrium Liquid*-phase Concentration; p^* - Equilibrium Gas-phase Partial Pressure; $K_G a$ - Overall Mass Transfer Coefficient based on ΔP; $K_L a$ - Overall Mass Transfer Coefficient based on ΔC.

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