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DESIGN CONSIDERATIONS FOR A THREE-PHASE FLUIDIZED-BED BIOREACTOR

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ABSTRACT

The design of a three-phase fluidized bed bioreactor using novel biomass support for industrial wastewater treatment is reported in this paper. A column of 0.20 m diameter with variable height up to 6.60 m equipped with a foam breaker tank was designed, commissioned and operated. A novel biomass support, KMT^R support, has been used. Separate gas and liquid feeding systems were designed to facilitate proper bioreactor operation. A flat distributor plate with 4 mm holes-diameter was desined. A study on the hydrodynamic properties of the three-phase fluidized bed bioreactor revealed high values of bed expansion and air hold-ups. Purification efficiencies of 84% and 40% are reported for brewery and refinery wastewater samples, respectively. The values of the volumetric oxygen transfer coefficients reached up to 120 h^{-1}, depending on the biorector model used. All available informationhave been used in the design work, details are given for each component and design step. A compromise has been reached between many design aspects. It has been concluded that the support should be used in industrial wastewater treatment, and that bioaugmentation techniques can improve the oxygenation and wastewater purification efficiencies of the bioreactor.

INTRODUCTION

In recent years, there has been a growing interest in the use of three-phase fluidized beds for contacting a gas, a liquid and solid. In some applications such as the hydrogenation of liquid hydrocarbons of petroleum products, the solid acts as a catalyst. In the fluidized-bed biological reactors (or bioreactors), the solid particles act as a support for the microbial film (biomass) which adheres either on the external surface or within some kind of porous structure. Thus, proper selection of the solid support is vital in the design of the three-phase fluidized bed bioreactors. Treatment of industrial wastewater is well established and the methods used therein are changing. Moreover, has been the advent of the three-phase fluidized bed bioreactor for treatment of industrial wastewater, first in UK [1], but now with

ever-increasing adoption in USA and Canada. Use of such bioreactors has therefore come to stay, and it is essential to discover ways of improving the efficiency of such industrial wastewater treatment plants. In all aerobic applications, an important aspect of the three-phase contacting is the dispersion of air or oxygen within the fluidized bed. This necessitates intensive design considerations on the gas distributor, which governs oxygen transfer rate in the bioreactor. Fluidized bed biological treatment technology is attracting considerable interest as an alternative to the conventional suspended growth and fixed-film processes in a wide variety of wastewater treatment applications. The process has been proved to be feasible for carbonaceous oxidation of domestic and high strength industrial wastewater, for nitrification and denitrification [1].

The use of the three-phase fluidized bed bioreactors in treatment of industrial wastewater is advancing due to the following advantages: that such plants are resistant to toxic loads, operates at low retention time, due to the high biomass concentration that can be achieved, thus offering high purification efficiency. The three-phase fluidized bed bioreactors require less land and low capital cost, and that, high biomass concentration can be maintained in a fluidized bed bioreactor without need for secondary clarification [3]. The importance of biological treatment of wastewater lies on the fact that these plants have a wide range of application, that is, they can be used in treating different types of wastewater. Treatment of industrial wastewater require a great deal of space when using systems based on activated sludge or aerated lagoons in which the retention time is many days. Investigation of the potential of fluidized bed reactors for BOD removal carried out by Vanderborght and Gilliard [4], have shown the ability of this technique for treating concentrated effluent at high loading. The common problems faced in the design of the wastewater treatment plants include: uncertainty associated with current biological process models; biological foaming (a mechanism of which is not fully developed); excessive biomass growth (that is, unsteady biomass hold-up), and excessive power consumption for the aeration process. Other uncertainties lie in the control and dissolution of oxygen and in the regulation of up-flow velocities in biological fluidized beds treating varying flow and loading.

The objectives of this work was to design, construct and assemble the experimental rig for industrial wastewater treatment. In particular, the following has been accomplished: design and assembling of the air and liquid distributors; selection of the biomass support for wastewater treatment; selection and assembling of the bioreactor accessories; and finally, assembling the experimental rig. Designing a three-phase fluidized bed bioreactor involves a wider

consideration of many factors. In the work presented in this paper, the following factors were considered: selection of the biomass support particles, gas distributor design, distributor plate design parameters, foam breaking equipment and gas/liquid distribution systems. The approach used during the design was to evaluate different alternatives for each design step from which the best options were selected.

DESIGN CONSIDERATIONS

Selection of the Biomass Support Particles

The importance of the support particles is to accommodate the biomass within the bioreactor so as to maintain a high biomass concentration compared to the dispersed growth systems, and if possible to control the biomass size. The particles also, play a role of increasing the rate of oxygen mass transfer and utilization by breaking larger air bubbles to smaller size [5]. Several support (or packing) materials are available like: carbon, ceramic, glass, paraffin, polyvinylchloride, and steel. Highly corrosive resistant polyethylene particles were selected in this work aimed at handling acids and many corrosive substances found in industrial wastewater. The following factors were considered during selection of the biomass support particles:

(a) steady or accumulating biomass hold-up?

Steady particle biomass hold-up has an advantage that all aspects of performance of the bioreactor becomes steady and that both particle and bed biomass hold-ups become pre-selected [5]. If the accumulating or increasing biomass hold-up per particle is a feature of the bioreactor, then, one needs to establish the timing of the backwash or particle removal (determined by the rate of the biomass accumulation). In this work, the particles selected were capable of offering a steady biomass hold-up, due to their shape.

(b) density of the biomass support particles

The spectrum of solid/fluid contacting covers a wide range of solid particle sizes from sub-microns to several millimeters, and specific gravity slightly above 1.0 up to 2.9 [6-11]. From Table 1, it is clear that the KMT^R support is lighter and larger than other particles. When the density is higher than that of the wastewater to be treated, then fluidization can be achieved by liquid up-flow, aided by low gas flow. The gas flow helps to enhance fluidization and turbulence in the fluidized bed. In this case, operating the bioreactor with heavy particles, may result into poor aeration of the bioreactor. Low density support particles were selected in this

work to allow gas fluidization only, since the particles were floating on the liquid surface. This necessitated operation of the bioreactor at high

Table 1 Survey of size and density of solid particles used in three-phase fluidised beds

Researcher	Equivalent size d_p (mm)	Density ρ_s[kg/m^3]	Material
This Work(1996)	8.3	950	Polyethylene
Lee et al [6]	6 4	2540 2560	Polyethylene Polyurethane
Karamanev et al [7]	3	1030	Polyurethane
Novakovick et al [9]	1-2 1-2	1040 1040-1500	Alginite Alginite+Magnetite
Nitta et al [10]	1.444 2.66	2874 2804	Glass Glass
Nore et al [11]	3.1 2.1 2.1 2.1	1700 1350 1290 1130	Polypropylene with inclusion of mica

gas rates so as to lower the bulk density of the liquid, until the solids sunk and moved freely in the fluidization column. Thus, low density polyethylene particles, the KMT^R support, with a density of 950 kgm^{-3} were selected and used in this work.

(c) problems caused by excessive biomass growth

Support particles can be placed in fixed, expanded or fluidized beds or in agitated tanks. In all cases, it is clearly necessary to prevent excessive accumulation of the biomass associated with the particles, so as to avoid blockage caused by the growing together of the individual particles, and control the overall size of the particles within the prescribed limits. Operation of the three-phase fluidized bed

bioreactors is also affected by a problem of excessive biomass growth on the support particles, making the later less dense as the process proceeds, leading to easy carryover of the bed, and sometimes leads to bed clogging [12]. Means of overcoming such problems suggested by other researchers, that is, replacing the heavily coated particles with biomass free charge, are more expensive and time consuming, because, solids replacement involves either removal, washing and then re-use, or complete renewal of support. Tables 2 and 3 give more information on plastic support and their properties when used in fluidized beds [5]. The problem of excessive biomass growth complicates the operation of the wastewater treatment plants, necessitating use of extra equipment such as vibrating screens or incinerators.

Table 2 Characteristics of reactor containing biomass support particles [5]

Property	Fixed/Expanded beds	Fluidised or agitated beds KMT	
Steady/variable particle biomass hold-up	Variable	Variable	Steady
Biomass control	Backwashing	Periodic particle removal	Strong attrition caused by air agitation
Biomass recovery	Washings collection	External to reactor	Conventional clarifier
Bed biomass hold-up	Variable	variable	Steady
Reactor Performance	Variable	Steady	Steady
Design	Predict biomass variation with time	Predict biomass variation with time	All aspects of performance are pre-selected

A new technique of solving the problem of excessive biomass growth is presented in this paper, that is, selecting the shape of the support which can enhance biomass growth and allow scrapping of the excess biomass. The novelity of this

research lies on the fact that the support used in this work is new in the field of industrial wastewater treatment. The support were cylindrical in shape with fins and internal matrix, which enhance biomass growth on the space between fins and maintain a constant biomass concentration in the long run, because the excess biomass growing on the surface of the support is scrapped off by the action of the solid-solid and solid-wall collisions. The advantage of such support is that the biomass growth take place on both inside and outside surfaces and that the flow of feed and air through the internal matrix gives easy access of the microorganisms to oxygen and substrate (carbon source). By allowing only air fluidization, the particles had an advantage of increasing the relative velocity between gas and liquid phases, thus increasing the mass transfer rate and that, the supply of oxygen was in excess of the uptake by microorganisms.

Table 3 Particle Biomass hold-up data [5]

Particle	Size and Shape (mm)	Biomass hold-up (mg/l)
KMT^R -Polyethylene	dp=8.3mm perforated cylinder	?
Stainless Steel Cylinders	d = 6 mm	3.1
Polypropylene Toroid	overall diameter =53mm Toroid diameter=20mm	5.0
Rectangular Polyester foam	Cuboid: 10x10x2	5.0
Reticulate propylene sheets	Cuboids: 25x25x6	5.0

(d) details of the support particles

The particles had a porosity of 0.84; height, h_p = 7 mm; outer diameter: cylinder = 8 mm, including fins = 11 mm. Total surface area = 866.2 mm^2; surface diameter, d_s = 16.6 mm; mass per particle = 3.1 x 10^{-4} kg; overall specific area = 2610 m^2/m^{-3}; volume per particle = 3.32 x 10^{-7} m^3. The sphericity, ι, was calculated to be 0.5, giving an equivalent size, $d_p = \iota d_s$ = 8.3 mm. Figure 1 shows the cross-section of the support particle. Comparing the size of the KMT^R support particles used in this work with those from other workers, Table 1, it can be seen that the KMT^R particles were comparably larger.

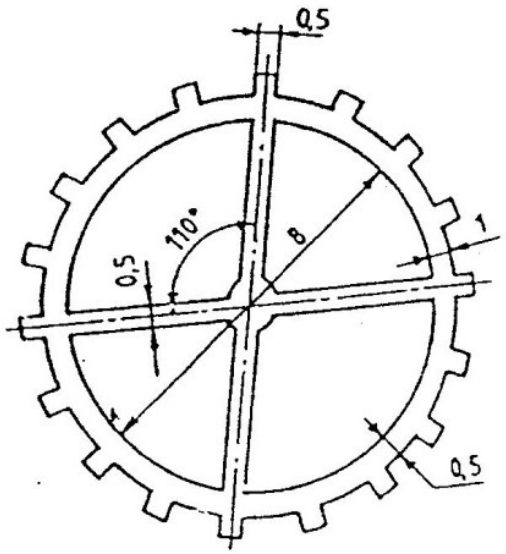


Fig. 1 The KMTR support, cross-sectional view

Gas Distributor Design

The major purpose of the distributor was to inject air bubbles into the bioreactor. The design of the distributor was of paramount importance since it affects the hydrodynamics of the three-phase fluidized beds. Non-ideal flow behavior in the distributor region can cause significant distortion of the bed flow dynamics [13]. Options available were bubble and mechanical aerators, for which a comparison

Table 4 Comparison between bubble and mechanical aerators [14]

Comparison Factor	Buble Aerator	Mechanical Aerator
Oxygenation capacity kg O_2/kWh	1.5-3.6	0.8-2.4
Allowable liquid height	8 m	4 m
Allow able Air flow rate, m^3 air/m^3 WW	7-10	4
Running costs	Pumping air	1-Pumping air 2-Rotating the system

is given in Table 4 [14]. The bubble aerator was selected because of the following advantages: it offers high oxygenation capacity; it allows high liquid height in the column; the aerator itself is of small size that it can be cheaply and easily fabricated; that such aerators do not involve moving parts, hence, energy is only required in pumping air while the other alternatives require energy in both pumping air and rotating the system, leading to high running costs.

Alternative plate-shapes are concave upwards or downwards, and flat plates. For the purpose of increasing the gas flow at the center of the column, and hence promote a circular motion of the solids in the fluidized bed, concave downwards plates are recommended. In this work, the same effect was achieved by having a flat plate, with high holes density at the central part [15]. The following are major parameters considered during distributor plate design: the fractional open area, f_{or}; number of holes, N_{or}; holes density, D_{or}; and holes diameter, d_{or}. The fractional open area, f_{or}, affects the volumetric oxygen mass transfer coefficient, k_La, gas orifice velocity, U_{or}, and the distributor pressure drop. When f_{or} is kept constant, the pressure drop then depends only on the gas flow rate across the distributor [15]. Dead zones in the distributor were avoided by having the holes density as large as possible. Small holes were avoided to eliminate any tendency of holes clogging by the biomass which can result into dead zones. The hole diameter, d_{or}, was made smaller than the smallest dimension of the support particle, that is, less than h_p = 7 mm, to prevent the particles from entering the wind box, and thus eliminating the need for wire mesh on the distributor plate. Alternative pitches are equilateral triangular and square pitch, Table 5. As strongly recommended in mass and heat transfer applications, the equilateral triangular pitch was used in this work.

Foam Separation Tank

Foaming occurs when bubbles rise to the surface of a liquid and persist without breaking. Presence of impurities increases the viscosity of the liquid film which becomes rigid. The rate of foaming in the three-phase fluidized bed bioreactor is high due to relatively high degree of liquid agitation by the air, compared to packed-beds which favour liquids tending to foam. Treatment of brewery wastewater proceeds with excessive foam generation [16]. In general liquids tend to foam due to presence of impurities. In treatment of industrial wastewater, the impurities are normally proteins of the following categories: either extracellular proteins or cell like products. Foaming can be advantageous to mass transfer but causes very serious operating problems, and therefore has to be nullified. Mechanical foam destruction is often proposed but consumes too much energy and is not fully reliable.

Table 5 Selected values for plate design parameters

Parameter	Range	Selected values
Fractional open area	$f_{or} = 0.10$	0.10
Holes density	$D_{or} = 200\text{-}13300/m^2$	8000
Column diameter	$D_c = 0.10, 0.20$ m	0.20
Number of holes	$N_{or} = f(A_c, D_{or})$	250
Pitch of holes	$s = f(d_{or}, f_{or})$	12mm
Pitch type	Equilateral triangular or square	Equilateral triangular
Hole diameter	d_{or}	4 mm

Chemical antifoam agents are also necessary. These are strong surface-tension-lowering substances. When they are added to wastewater, faominess is strongly retarded. The antifoam liquids are usually composed of oils, fatty acids, esters, polyglycols and siloxanes which destabilize protein films by: hydrophobic bridges between two surfaces; displacement of the absorbed proteins, and, rapid spreading on the surface of the film. The second mechanism being often relevant. Preferred antifoams are the low moleculr weight compound, which give a lower surface tension than proteins. The effect of antifoam agent lies on the mass transfer. The volumetric mass transfer coefficient, k_La, depends on the rigidity of the bubble surface. In tap water, bubbles of radius less than 1 mm have a rigid surface, and bubbles with radius greater than 2 mm have a mobile interface, which is advantageous for the mass-transfer rate. The conditions which cause breakage of bubbles in foam, also favour the coalescence of bubbles within the liquid phase, resulting in larger bubbles with reduced surface to volume ratios and hence reduced rate of oxygen transfer [17].

To overcome this problem, the following alternatives were evaluated: firstly, foam breaker tank, mounted at the top of the fluidization column, to accommodate a foam when it arises, allowing the gas to separate from the recycled liquid and foam, after the exhaust gas flow pressure has been reduced in the tank. Second alternative was a stirrer-motor unit, to be mounted at the top of the column, which could destroy any arising foam. Application of antifoam agents, that could stop or

minimize foam, was the third alternative, which require compromise between minimizing the foam height and maximizing mass transfer. The first alternative was selected due to the fact that only initial cost of manufacturing could be incurred, and no more running costs like the other two alternatives. Figure 2 shows the dimensions and shape of the tank. That is, the tank was mounted at the top of the column, as shown in Fig. 2. The tank was fabricated from stainless steel sheet, 4 mm thick.

Liquid Feeding System

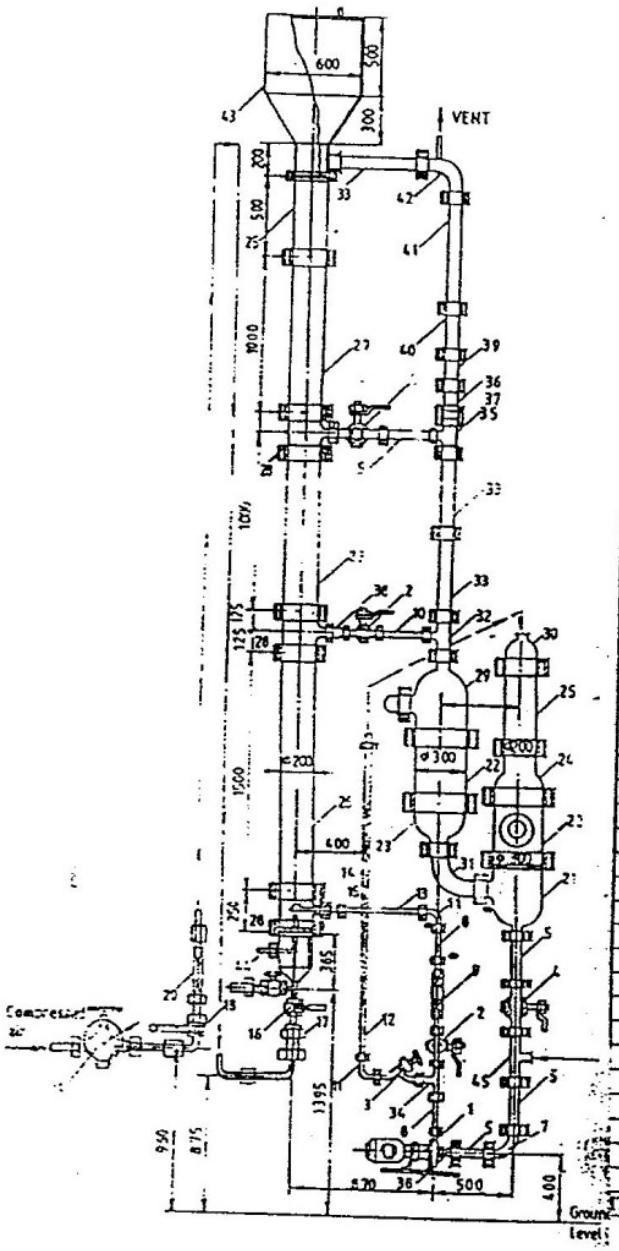
Wastewater from the feed tank was pumped into the column via a horizontally mounted sprinkler through an unequal tee-piece (major diameter, 200 mm, and a minor diameter of 50 mm) connected above the distributor. A firing end of the sprinkler was positioned such that it was firing liquid jets upwards at the center of the column, as shown in Fig. 2. Originally, a glass made T-piece was used, but due to limitations upon insertion of dissolved oxygen probes, an aluminium made T-piece of the same size was fabricated. As part of modification, two extra holes were made to allow insertion of pressure taps.

Power Consumption

The designed pilot plant utilizes electricity in the three-phase pump, 0.55 kW (at 2900 rpm, 240 V supply), having a maximum discharge rate of 6 m^3/h and 4.00 kW in the centrifugal air compressor (with a capacity of 500 litres at 300 K with a maximum operating pressure of 10 bar). Thus a total of 4.55 kW is expended, with aeration energy accounting for 88% of the total energy budget, which is comparable to 80% reported by Trevan, [18]. The *COD* removal efficiency was tested for five days, while the mean operating time of the compressor was 118.6 s for a mean interval of 1131.5 s, which gives a fraction of 0.105 for each time interval. The energy consumption for a treatment time of 5 days was calculated to be 206 MJ.

Materials of Construction

The pilot plant was constructed mainly from borosilicate glass, with a trade name of *DURAN*® , DIN standards: 28801 - 28815 (column and liquid pipelines). Properties of the *DURAN* glass are: specific gravity 2300 kg/m^3; thermal shock resistance - 250°C; tensile stress - 3.92 MPa; compressive stress - 98.1 MPa. The nature of the material was borosilicate glass group. The use of *DURAN* glass had the following advantages: that the material is resistant to highly corrosive media; glass provides a smooth surface without porosity and therefore offers a reduced solid deposits on walls of the column; the material is physiologically harmless; the glass is transparent and hence allows a good visual observation. The other



ITEM	NAME	No.	MATERIAL
45	Unequal T-Piece	1	DURAN
44	Sampling Pipe	1	Copper alloy
43	Foam Breaker	1	S.S
42	90 degree bend	1	DURAN
41	Pipe section	1	DURAN
40	Pipe section	1	DURAN
39	Pipe section	1	DURAN
38	Pipe reducer	2	DURAN
37	Universal bellow	1	PTFE
36	Pipe section	1	DURAN
35	Unequal T-piece	1	DURAN
34	Equal T-piece	1	DURAN
33	Pipe section	3	DURAN
32	Unequal T-piece	1	DURAN
31	90 degree bend	1	DURAN
30	Pipe reducer	1	DURAN
29	Column adaptor	1	DURAN
28	Unequal T-piece	3	DURAN
27	Column section	2	DURAN
26	Column section	1	DURAN
25	Column section	2	DURAN
24	Pipe reducer	1	DURAN
23	Pipe reducer	1	DURAN
22	Unequal T-piece	2	DURAN
21	Column adaptor	1	DURAN
20	Air rotameter	1	GG/Glass
19	Pressure regulator	1	Copper alloy
18	Ball valve	1	GG/DURAN
17	Nonreturn valve	1	GG
16	Ball valve	1	GG/DURAN
15	Distributor plate	1	Aluminium
14	Sprinkler	1	DURAN
13	Pipe section	1	DURAN
12	Pipe section	1	DURAN
11	90 degree bend	2	DURAN
10	Pipe section	1	DURAN
9	Rotameter	1	DURAN
8	Pipe section	2	DURAN
7	90 degree bend	1	DURAN
6	Pipe section	1	DURAN
5	Pipe section	3	DURAN
4	Ball valve	2	GG/DURAN
3	Valve	1	DURAN/PTFE
2	Ball valve	1	GG/DURAN
1	Centrifugalpump	1	GG/DURAN

Fig. 2 The three-phase fluidised bed bioreactor, complete setup

advantage is that the units gave flexibility in building a modular block system, which simplifies installation. Cylinders of 200 mm were used for construction of the column. The column height, 6.6 m, was achieved by using cylinders of 0.25, 0.50, 1.00 and 1.50 m high as shown in Fig. 2.

Aerating Medium and Operating Conditions

Two alternatives were evaluated, that is high purity oxygen and compressed air. Use of high purity oxygen increases the saturation oxygen concentration in the bulk liquid and allows operation of the bed at low k_La values, which implies low power consumption [18]. From literature, it is claimed that when pure oxygen is used, there is a possibility of treating wastewater at higher organic loading with high purification efficiency. A survey made on the methods utilizing pure oxygen, Table 6, shows that the *DOC* values are not very high and the loading is low compared with other treatment works which utilize compressed air [14]. The target in this work was not to increase C_L, but to fluidise the bed, and since pure oxygen is denser than air, pure oxygen would fluidise the bed at higher flow rates, which would consume more oxygen and more power. The use of pure

Table 6 Survey of wastewater treatment plants utilizing pure oxygen [14]

System	Properties/Capacity
UNOX	Food to Mass ratio (F/M)=0.4-1.0kg BOD/kg.day DOC: 4 -8mg/l
OASES	Loading: 0.7 kg BOD/kg.day; Retention time 2 h; DOC = 10 mg/l
VITOX	Loading: 0.3-0.8 kg BOD/kg.day; aeration capacity: 400 Wh/tone oxygen dissoved; Pressure: 2-3 bar
MEGOX	F/M=0.3-0.8 kg BOD/kg.day; E=86% for COD_o range 3800-4400 mg/l
MAROX	Loading: 0.4-1.0 kg BOD/kg.day; Retention time = 1- 2.5 h; Operating pressure range: 1.3-2.0 bar

oxygen favours oxygen dissolution, but not oxygen utilisation which is a constant for *DOC* above 0.6 mg/l. This requires that the oxygen be limited just below the critical oxygen concentration. Use of pure oxygen is allowed only when the bioreactor loading rises rapidly. Under well controlled design, injection of pure

oxygen in an air stream is encouraged. High purity oxygen was not used in this work. In industrial practice, if high purity oxygen is to be used, its supply should be installed together with the compressed air supply, but should be used only when there is overshoots in loading. Under well controlled process conditions, injection of pure oxygen into the air stream is encouraged [19]. It can be seen from Table 6 and Table 10 that the dissolved oxygen concentration values are not very high, and that the loading to plants utilising pure oxygen are comparable to the pilot plant when fed with wastewater samples from TIPER, the most loaded wastewater, that is 0.64 kg of COD/kg wastewater per day. The efficiency of these oxygenic reactors are nearly equal to the values observed in this work that is 86% and 40%. This signifies that use of pure oxygen is not feasible.

Operating Temperature

For optimum growth and metabolism of microorganisms, the temperature is subdivided into three ranges. The first range is below and up to 10°C, where psychophilic growth takes place. At temperatures ranging from 15 - 40°C, (covering the room temperature), mesophilic growth takes place, while thermophilic growth takes place at temperatures above 45°C [20]. The mesophilic range was selected because most heterotrophs (microorganisms utilizing organic matter as source of energy) operate in the mesophilic temperature range. Also, this temperature range is low enough to offer high momentum transfer. Low temperature, below 10°C, will render the transitional Reynold's number high, leading to low turbulence in the bed. The pH range depends on the type of microorganisms to be used in treatment works. For instance, 8.5 - 8.8 is favourable to the nitrosomonas, while nitrobactor survive in a wider pH range of 8.5 - 9.3. The pH range selected in this work, was from 6.5 to 8.5 to suit the mixed culture [20].

The Bioreactor Column and other Accessories

There is no limitations with regard to configuration and column shape for bioreactors. However, due to the fact that glass cylinders are the most available, a cylindrical column was chosen. While finding the height of the column, it was important to consider that a sufficient height is required for both the expansion of the bed and entrained materials. Further more a large height is advantageous for dissolution of oxygen in a liquid. On the other hand there was a limitation with regard to the air pressure obtained from a compressor. Finally, a space available in the laboratory had to be considered. When assembling, the rig was built in such a way that dimensions of modular parts fitted to a grid. Dimensions were either equal to, or corresponding to, a multiple of this length. This enabled fitting the pipes using standard lengths. It was therefore possible to exchange when

necessary, for example, tee-pieces by elbows, elbows by valves, and valves by straight pipes. The possibility of interchanging was advantageous because it enabled any combination required. DURAN glass pipes of nominal diameters 25, 50, and 80 mm and lengths equal to 100, 200, 400, and 800 mm were used for enabling various lengths of the pipelines, as depicted in Fig. 2.

METHODOLOGY

In this research, wastewater samples were collected from two industries, Tanzania Breweries Ltd., *TBL*, and Tanzania-Italy Petroleum Refinery Co., *TIPER*. Artificial wastewater was made by dissolving selected compounds (phenol in this case) in tap water. The major tasks performed are summarized as: pilot plant design, selection of biomass support and establishment of measurement techniques, and experimentation. The study of hydrodynamics of the three-phase fluidized bed bioreactor, investigation of oxygen transfer rate and purification efficiency started with complete design, erection and testing of the pilot plant. Liquid from the feed tank, was pumped into the column to a required height of slurry (solid and liquid). While keeping the liquid flow rate constant, compressed air was introduced at the bottom to fluidise the bed. Liquid fluidization was not possible, due to low density of particles, which were floating at the liquid surface for liquids with density above 950 kg/m^3, except light liquids, for instance very concentrated refinery wastes. Throughout treatment time liquid was recycled while keeping the fluidized bed height constant. The bioreactor was operated in the range of 3.7 to 42.5 mm/s for air velocity; while the liquid velocity ranged from 0 to 6.6 mm/s. Compressed air pressure was varied between 2 to 5 bar, while the liquid height was varied between 1.0 to 3.0 m. Solids loading was varied from 0.5 to 4.5 kg. The total fluidized bed height ranged from 1.0 to 6.0 m. All experiments were carried out at room temperature.

Two dissolved oxygen probes (model *RL 425*, supplied by Russell Laboratory Equipment Co., UK). The electrodes were fixed into the column 1.70 m apart (the lower probe being 10 cm above the distributor plate). The probes read directly the dissolved oxygen concentration (mg/l), temperature (°C), and percent saturation in the bulk liquid.

The $(k_La)_h$ was determined after aerating the bed for 3 to 5 minutes (to attain equilibrium in the bed) at known values of u_g and u_L. Initial and final bed heights were recorded for each run, at the end of which the mass of the liquid present in the bed was measured after stopping the gas and liquid flows abruptly (fast-close-valve method). Equations utilizing the hydrodynamic parameters of the bioreactor,

were used to predict the $(k_La)_h$ values as shown in equation (1) [21]

$$(k_L a)_h = 1.154\left(\frac{P_O}{V_R}\right)^{0.79} \quad (1)$$

whereby the power per unit volume was determined experimentally as per quation (2) [22]

$$\left(\frac{P_O}{V_R}\right) = \frac{g}{\varepsilon_g}[\rho_s \varepsilon_s (u_L + u_g) - \rho_{Lu_L}(1 - \varepsilon_L) + \rho_L u_g \varepsilon_L] \quad (2)$$

The air hold-up, ε_g, that is, the fraction of the bed volume occupied with air, was calculated from the bed expansion measurements. The liquid phase hold-up, ε_L, was determined from the measurements of mass and density of liquid in the bed (after measuring M_L, ρ_L, and H_c). The solid phase hold-up, ε_s, was calculated as the difference (ε_s=1-ε_g-ε_L).

Two equations utilising different bioreactor models were used to calculate the k_La values from the dissolved oxygen concentration values, $(k_La)_c$: the static continuously stirred tank bioreactor model, CSTR, leading to $(k_La)_{CT}$, and the plug flow model, PFR, which led to $(k_La)_{PF}$. The employed equations were (3) [21]

$$(k_L a)_{CT} = \frac{u_L}{\Delta Z}\left(\frac{C_2 - C_1}{C^*_g - C_2}\right) \quad and \quad (k_L a)_{PF} = \frac{u_L}{\Delta Z}\left(\frac{C^*_g - C_1}{C_g{}^* - C_2}\right) \quad (3)$$

To investigate the effect of k_La on purification efficiency, it was necessary to determine the *COD* values at different air velocities (at which the corresponding k_La values were known). The purification efficiencywas calculated at each stage. Throughout the treatment period, a constant setting of u_g, H_c, u_L and M_s was maintained. The *COD* values were measured using a digester-photometer unit. The digester was used to heat test tubes containing wastewater samples from the bioreactor at a temperature of 148°C for 2 hours. The photometer, operated in the wavelength range of 400-800 nm, with two standard filters of 445 and 485 nm, was used to read directly the *COD* value in ppm after digestion and cooling to room temperature.

DISCUSSION

A closer examination of the fluidized bed revealed that particles close to the walls

were almost moving downward. This was attributed to the fact that since the distributor plate had many holes at the centre than towards its rim, then many bubbles were initiated at the centre of the column. Thus the few bubbles on the wall are attracted towards the other bubbles near the centre, while themselves attract only a limited number. Hence, the probability of a bubble moving towards the walls was low while that of the bubble moving into the centre of the bed was high. As a consequence of few bubbles closer to the wall, there were a predominantly downward flow of particles close to the wall. At the top of the fluidized bed, solids were emerging at the centre and moving towards the walls. This suggested that the motion of the solids in the fluidized bed was circular in nature, and that solids were being carried up by the gas bubbles. It was further observed that the vertical mixing (or vertical motion) of the particles in the fluidized bed was many times faster than the lateral motion. The observed motion of the solids can be explained as being induced by bubbles, whereby, particles were carried upwards in the wake of the bubble (i.e. solids occupying the bottom of the completed sphere), and the drift (defined as the region behind the completed sphere of the bubble). The bed expansion and air hold-ups increased monotonically at low air velocities, reaching a maximum of 22% and 30% respectively, at u_g = 42.5 mm/s. For larger values of u_g, E_b values remained practically constant. The air hold-ups reported in this work are comparable to those found in literature, showing that the bioreactor was capable of promoting aerobic conditions. The oxygen transfer rate expressed in terms of the volumetric oxygen transfer coefficient, k_La, was found to range between 3 and 120 h^{-1}. The values of k_La were found to increase with air velocity, mass of solids charged into the column, and with increasing pressure of the compressed air while they decreased with increasing liquid height, and liquid viscosity. The pilot plant was fed with highest *COD* loads of 26000 and 48000 mg/l for brewery and petroleum refinery wastewater, respectively. The highest *COD* removal efficiency was found to be 84% and 40% for brewery and refinery wastewater respectively. These results suggest potential advantages for treatment of brewery wastewater in a three-phase fluidized bed bioreactor using KMT^R support. Low values of purification efficiency for refinery wastewater was attributed to the fact that it contains a big number of components some of which could inhibit the growth of micro-organisms especially phenolics [22].

CONCLUSIONS

From the above findings it can be concluded that:
Due to high purification efficiency obtained in this work, the KMT^R support can be has proved to be a potential biomass support for use in the three-phase fluidized

bed bioreactor when the later is used in industrial wastewater treatment. The purification efficiency in the three-phase fluidized bed bioreactor depends on the biodegradability of the components of the wastewater. The low purification efficiency for refinery wastewater reveals that further work need to be done to raise its biodegradability. The oxygen transfer rate in the three-phase fluidized bed bioreactor using the KMT^R support, the air hold-ups and bed expansion reported in this work, are high enough to maintain aerobic conditions in the bioreactor. Also the k_La values are higher compared to those obtained in the activated sludge and aeration tanks at the same air flow rates. The *DOC* level in the bioreactor, was well above the critical dissolved oxygen concentration for a mixed culture of microorganisms, that is 0.01 mg/l. With reference to the results obtained in this work, the design procedure adopted proves to be a key to the design of industrial scale bioreactors.

Novelty of this research and my contribution to the wastewater treatment engineering can be summarised as follows: that such support particles are new in wastewater treatment practice being lighter and larger. By the virtual of being light, the KMT^R support was capable of being fluidized with air upflow only, encouraging high oxygen content in the bioreactor, and avoiding use of pure oxygen. The bed expansion observed in this work is an advantageous finding, this allows high air hold-ups, large interfacial area, a, giving rise to high oxygen transfer rate. Solutions to the problems associated with the three-phase fluidized bed bioreactors, that is power consumption and excessive biomass growth, has been accounted for in this work.

Improvement of the oxygenation efficiency of the bioreactor involves seeking a number of new approaches to enhance the treatment capabilities of biological processes (the so called, bioaugmentation). This include optimizing the operational parameters, which can be done after performing many experiments. It is further recommended that, the three-phase fluidized bed bioreactor with KMT^R support should be used in treatment of industrial wastewater. The design for industrial scale plants should aim at attaining high values of k_La and a good mixing of phases. Further studies are encouraged on the optimization of the operating parameters and control of the parameters especially the oxygen concentration.

NOTATION

a - interfacial surface area
A_c - cross-sectinal area of the column, m^2

C_1, C_2 - dissolved oxygen concentration at bottom and top of bioreactor column, mg/l
C_g^* - saturation oxygen concentration, mg/l
C_L - dissolved oxygen concentration in the bulk liquid, mg/l
COD_o - initial COD, mg/l
COD_t - final COD after time, t
d_{or} - hole diameter, m
D_{or} - holes density, holes/m^2
d_p - equivalent particle size, mm
E - COD removal or wastewater purification efficiency, %
E_b - bed expansion during aeration, %
f_{or} - fractional open area
g - acceleration due to gravity, m/s^2
H_c - fluidized bed height, m
H_L - liquid height in the column, m
H_{sL} - bed height before aeration, m.
ΔH - change in bed height due to aeration, m.
h_p - height of the particle, mm
k_La - volumetric oxygen mass transfer coefficient, h^{-1}
$(k_La)_h$ - k_La determined from the bioreactor hydrodynamics
$(k_La)_{PF}$, $(k_La)_{CT}$ - k_La determined from oxygen concentration measurements, using the plug flow model and CSTR model respectively.
(P_o/V_R) - power consumed in aerating the bioreactor, W/m^3.
p_t - distributor plate thickness, mm
s - pitch of holes, mm
u_g, u_L - gas and liquid velocities, mm/s

Greek letters

ε_g, ε_L, ε_s - air, liquid and solid hold-up,
ρ_s - density of biomass-free support, kg/m^3
ρ_L - density of liquid, kgm^{-3}
ι - sphericity of solids particles.

Abbreviations

BOD - Biological Oxygen Demand
COD - Chemical oxygen Demand
CSTR - continuously stirred tank reactor
PFR - plug flow reactor
KMT - Kaldnes Miljø Teknologi Co., Norway.
TBL - Tanzania Breweries Ltd.

TIPER - Tanzania-Italy Petroleum Refinery Co.

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