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Brewery wastewater treatment in a fluidised bed bioreactor

Aoyi Ochieng [a,*], Tom Ogada [b], William Sisenda [c], Paul Wambua [d]

[a] *Chemical Engineering Department, University of Cape Town, Rondebosch 7701, Cape Town, South Africa*
[b] *Production Engineering Department, Moi University, Box 3900, Eldoret, Kenya*
[c] *Chemical and Process Engineering Department, Moi University, Box 3900, Eldoret, Kenya*
[d] *Textile Engineering Department, Moi University, Box 3900, Eldoret, Kenya*

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Abstract

A hydrodynamic characteristic performance of a three phase fluidised bed bioreactor has been studied with brewery wastewater. The influence of operating parameters, such as phase hold up, phase mixing, aspect ratio and superficial gas velocity, on an aerobic biodegradation in a bioreactor of 0.16 m i.d. and 2.7 m in height, was analysed. A low-density (960 kg/m^3) support particle with an internal interstice was employed. The particle and liquid loading were varied in order to determine the effect of phase hold up on bed homogeneity. The ranges in which particle loading and bed height affect fluidisation, and consequently chemical oxygen demand (COD) reduction, were determined. The distributor used in this work was designed such that fluid flow pattern similar to that of a draft tube was induced in the reactor. The low-density particles enabled cost effective operation at a relatively low gas superficial velocity (2.5 cm/s). Aspect ratio significantly influenced the overall bed homogeneity, and the optimum aspect ratio was 10, with volume of the support particles being 21% of the reactor volume.

Keywords: Brewery; Bioreactor; Aspect ratio; Hydrodynamics; Phase hold up; Biodegradation.

1. Introduction

Three phase fluidised beds have gained a considerable application in chemical, wastewater treatment and in biochemical industries. Some of the reasons for their extensive use are simplicity in construction, low maintenance due to lack of moving parts, high effective interfacial areas and therefore, a high heat and mass transfer per unit volume [1].

* Corresponding author. Tel.: +27-216-5040-35; fax: +27-216-8975-79.
E-mail address: aoyio@yahoo.com (A. Ochieng).

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Nomenclature

C	constant in Eq. (3)
Hx	height of fluidised bed (m)
H/D_c	ratio of bed height to column diameter
U_L	superficial liquid velocity (cm/s)
U_g	superficial gas velocity (cm/s)
U_{mf}	minimum fluidisation velocity (cm/s)
V_L	liquid volume (m^3)
V_r	reactor volume (m^3)
V_s	support particle volume (m^3)

Greek letters

ε_g	gas hold up
ε_L	Liquid hold up
ε_S	solid hold up
ρ_L	density of the liquid (kg/m^3)
ρ_p	density of the support particles (kg/m^3)

Subscripts

L	liquid
g	gas
mf	minimum fluidisation
p	particle
r	reactor
s	solid

Superscript

n	constant in Eq. (3)

In particular, there is an increasing use of three phase fluidised bed bioreactors and bubble columns in biological treatment of wastewater. This is as a result of a need for a more specialised method of treatment for industrial wastewaters, which occasionally contain volatile toxic components. Treatment of these toxic substances in open lagoons could be a health hazard, therefore, the fluidised bed bioreactor becomes a system of choice. The geometry of the reactor is dictated by the characteristics of the waste to be treated. Success of the design and the operation of a three phase fluidised bed bioreactor (TFBB) depends a great deal on the accurate prediction of the fundamental properties of the system, especially, the hydrodynamics in the reactor as well as the physical and chemical properties of the feed [2]. Although, TFBB is simple to construct and easy to operate, its scale up poses a significant challenge due to the complex interaction among many parameters. The complexity of the interdependence of the hydrodynamic aspects and the intrinsic kinetic ones has resulted in the development of different mathematical models. However, none of the models explicitly describes the bed characteristics to enable accurate scale up [3]. Some attempts have been made to identify and quantify the design parameters, however, wide scale application of a fluidised bed bioreactor still poses a challenge due to the fact that the parameters are waste specific [4].

One of the principal requirements for a bioreactor is the ability of the micro-organisms to grow on the support particle and remain embedded for significant biodegradation to take place. The microbial growth rate is so crucial for bioreactor performance that a lot of attention has been given to this, owing to the fact that bed turbulence and the nature of support particle appreciably affect microbial attachment. Considering the nature of the waste, it has been found necessary to adjust nutrient content and pH to enable initial microbial growth [5,6]. Gas flow rate affects phase mixing and gas hold ups, which are important parameters that influence oxygen mass transfer [7]. The survival as well as the activity of the aerobic micro-organisms depends on the amount of oxygen dissolved. A well mixed medium will give rise to a homogeneous bed that promotes interfacial mass transfer, and this will make oxygen available to the micro-organisms.

A correlation of gas velocity, phase hold ups and pressure drop can reveal the influence of the hydrodynamic characteristics on biodegradation. Pressure drop across the bed is an economic aspect, which must be considered since it determines the amount of power consumption for fluidisation. Pressure drop per unit height and phase hold ups can be calculated as follows [2]

$$-\frac{\mathrm{d}p}{\mathrm{d}z} = (\varepsilon_{\mathrm{s}}\rho_{\mathrm{s}} + \varepsilon_{\mathrm{L}}\rho_{\mathrm{L}} + \varepsilon_{\mathrm{g}}\rho_{\mathrm{g}})g \tag{1}$$

where ε_{g}, ε_{L}, and ε_{s} are gas, liquid and solid hold ups, respectively; ρ_{L}, ρ_{g}, and ρ_{s} similarly represent densities; g is acceleration due to gravity.

$$\varepsilon_{\mathrm{s}} + \varepsilon_{\mathrm{g}} + \varepsilon_{\mathrm{L}} = 1 \tag{2}$$

Eq. (1) shows that the use of dense particles leads to an increase in pressure drop, which consequently causes an increase in power consumption. Low density particles reduce this tendency, however, they are prone to elutriation. A careful design of a gas distributor and a bioreactor internals, like draft tube, can minimise this problem.

Gas flow rate contributes significantly to the cost of operating a bioreactor and there is a correlation between gas hold up and the flow rate, given by

$$\varepsilon_{\mathrm{g}} = CU_{\mathrm{g}}^{n} \tag{3}$$

where the value of the constant n varies slightly from unity, depending on the system employed and flow regime of the operation. A lot of work has been done on hydrodynamics of two phase and three phase fluidised bed, however, the data concerning the effects of the hydrodynamic factors on biodegradation is still limited. There is, therefore, a need to correlate the hydrodynamic operating parameters with the reactor design parameters in order to achieve a more efficient biodegradation.

This study focuses on the influence of the hydrodynamic characteristics on aerobic biodegradation of a brewery wastewater. The aim of this work was to study the influence of the reactor design and hydrodynamic parameters on the chemical oxygen demand (COD) and biochemical oxygen demand (BOD) reduction.

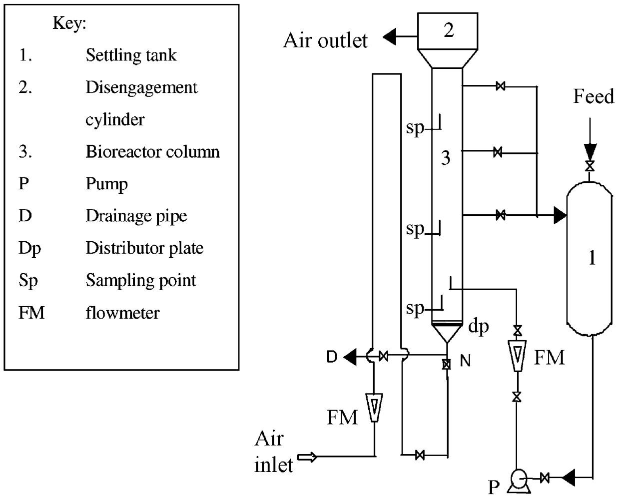

Fig. 1. Schematic diagram of the apparatus.

2. The equipment set-up

The major features of the rig are bioreactor column (5.5 m high and 0.16 m i.d.), disengagement cylinder, and settling tank. A flow sheet for the rig is shown in Fig. 1, where compressed air was injected into the bioreactor through a U-turn pipe, which was necessary to prevent back flow of liquid into the air rotameter, in case of power failure. A non-return valve (N) was used, however, it was observed that some types of wastewater react with internal components of the valve rendering it ineffective [8]. In such an event, the U-turn pipe would prevent liquid back flow into air rotameter and subsequently into the cylinder of the compressor. The disengagement cylinder (2) acted as a foam breaker, while any entrained bubbles would escape from the settling tank (1) without getting into the pump. The tank with a capacity of 80 l also provided the necessary head for the pump. The bioreactor (3) and the settling tank were made of Duran glass, while the tank and the cylinder were made of stainless steel. Sieves were put in place to prevent particles from getting into tank (1).

The design of the distributor plate was such that, there were more orifices at the centre of the plate. This was necessary to improve phase mixing with the low-density support particles by simulating a draft tube fluid flow pattern required for low density particles.

3. Experiments

Initial experiments were carried out with tap water and air at ambient conditions in order to analyse the fluid hydrodynamic aspects like phase mixing, aspect ratio, phase hold up and minimum fluidisation velocity (U_{mf}) with an 11 mm diameter support particle. Given that the particles used were large, phase mixing was studied by visual observation, such a method of observation has been reported in literature [2]. The minimum fluidisation velocity was taken as the minimum gas velocity that effected uniform phase mixing. Aspect ratio was altered by changing bed height, at constant column diameter. The volume of feed in the bioreactor was dependent on the required operation bed height. The loading was such that the real volume of the particle was in the range of 4–30% of that of the liquid mixture.

Biodegradation experiments were carried out with a brewery wastewater, which was collected from a brewery industry. Before commencement of the full operation, a fresh brewery wastewater feed was introduced into the reactor through which air was passed for a period of 50 days. At this point stability was attained after a COD reduction from 16,000 to 140 ppm. Beyond this period, there was a gradual decrease in microbial concentration, and very little change in COD. After attaining a well mixed and adapted culture, further experiments were carried out batchwise with a hydraulic retention time of 4 h. Microbial concentration was measured by the method used by Lenas et al. [9], whilst the determination of COD was done by the standard methods of water analysis [10]. Verification of the accuracy of the measurements was done with potassium hydrogen phthalate for which the theoretical COD value is known. The error in COD determination was found to be about 7%. Dry mass concentration, which was determined by measuring the optimal density of the sample with a spectrophotometer [9], was found to be about 165 g/m^3.

4. Results and discussion

Preliminary experiments with tap water revealed that the movement of the particles was influenced by the nature of the fluid flow pattern. Quite significantly, the fluid hydrodynamics was affected by the nature of the design of the gas distributor. Fluid flow pattern was studied by comparing the performance of two distributors: one with uniform orifice distribution and another one with many orifices at the centre. The latter, which was later adopted, exhibited flow characteristics similar to a draft tube, in which the part of the fluid at the centre of the reactor moves upwards, and the movement was downwards in the annular. Limited radial mixing was observed and the entire flow characteristic was akin to that of a continuous stirred tank reactor. The degree of bed turbulence was governed by the gas flow rate and the particle hold up, and to a remarkable extent, the minimum fluidisation velocity depended on these two parameters.

4.1. Phase hold ups and bed height

Fig. 2 shows a sudden increase in U_{mf}, when the ratio of the volume of the biomass-free solid support to that of the liquid mixture (V_s/V_L) was greater than 20%. At higher support particle loadings ($V_s/V_L > 20\%$), the particles were at the top part of the liquid. As the gas velocity increased, most particles got stuck at the top part of the bed, and the value of the minimum fluidisation velocity tended to infinity.

The particles used here being less dense than water, initially floated on the surface and the liquid circulation alone could not fluidise the bed. Fluidisation was, therefore, effected by the gas phase. It has been reported that in such an operation, with porous plate distributor, the resultant density of gas–liquid mixture is lower than the density of the liquid phase alone. Thus, this resultant low density enables particles to sink, hence fluidisation starts [11].

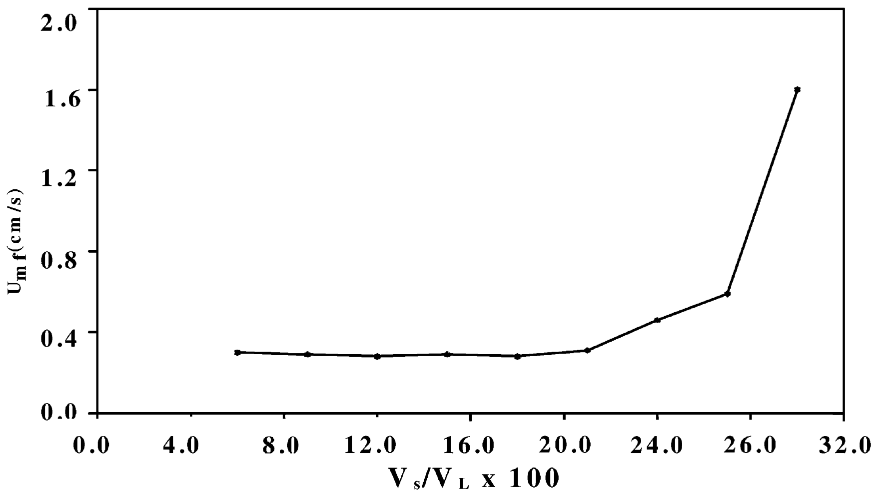

Fig. 2. Variation of U_{mf} with V_s/V_L.

However, in this work, it was apparent that fluidisation occurred not due to the reduced apparent density but rather due to the flow pattern in the column.

There was no indication that the fluidisation occurred due to reduced density because the bubbles were relatively big, with the diameter almost equal to that of the particles. Further, since the bubbles were moving downwards, it was evident they were doing so due to the induced liquid current flow pattern and not due to reduced density of the mixture. Best fit curves were drawn though the data point and Fig. 3 shows that there was an increase in gas hold up (ε_g) with an increase in static bed height and increase in superficial gas velocity (U_g).

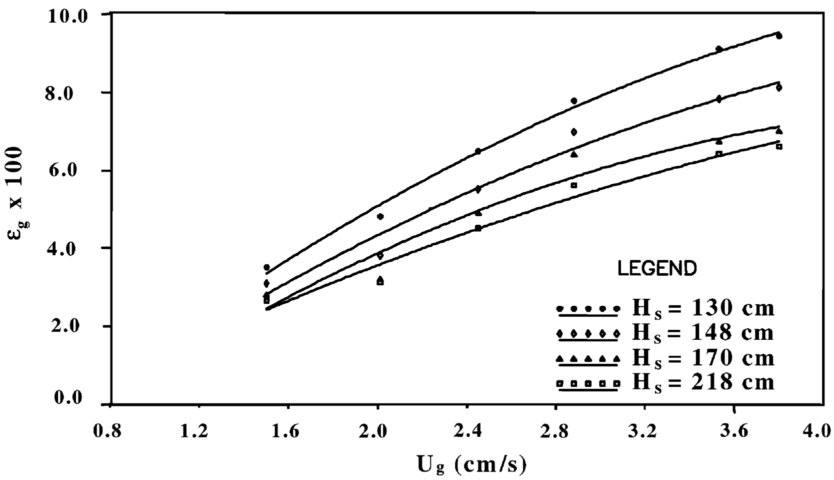

Fig. 3. Variation of gas hold up with superficial gas velocity.

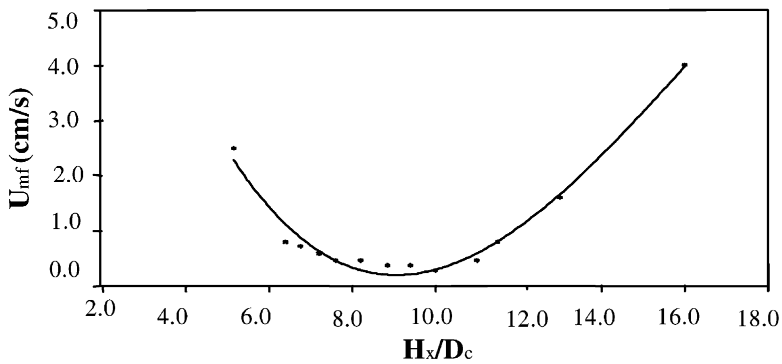

Fig. 4. Variation of aspect ratio with U_{mf}.

It appeared that there was almost a direct proportionality between ε_g and U_g, and the tendency was more pronounced at a relatively low U_g (e.g $U_g < 3.5$ cm/s). This observation is consistent with the results obtained by Wen and Fan [12], in which it was observed that the relationship between gas flow rate and gas hold up is linear at low gas velocity but tends to level off at high gas flow rates. There was a change from homogeneous to churn-turbulent regime and plug flow characteristics were exhibited at high U_g.

The concept of bed height, per se, is not sufficient to precisely characterise the performance of a bioreactor. Bed height to column diameter ratio (aspect ratio) influences the fluid hydrodynamics in such a way that when the ratio is high, bubbles have to travel long distance up the column and as a result, there is more increased chances of coalescence. Bigger bubbles move faster resulting in short residence time and consequently low gas hold up. This fact can explain the decrease in gas hold up with the increase in bed height as observed in Fig. 3. The decrease in bubble size with the increase in column height is due to the decrease in pressure with height. Li et al. [13] used a similar reactor with a perforated gas distributor, and found that bubble size decreases with an increase in pressure.

From Fig. 4, it can be observed that there was a minimum value of U_{mf} at which the value of Hx/D_c (Hx $=$ 1.9 m; $D_c = 0.2$ m) was about 9.5. Uniform phase mixing took place at this point. The left hand side of the graph represents a situation in which liquid was added to a constant mass (3 kg) of support particles. At the initial stage, the reactor behaved like a fixed bed, and the value of U_{mf} was very high. This situation persisted as long as the liquid hold up was still low. As liquid hold up increased, a point (valley) was reached when the particles could be suspended at a very low gas flow rate, and this was the minimum U_{mf} (about 0.3 cm/s). The right hand side was due to the increase in bed height owing to the increasing liquid loading, this resulted in high pressure drop. Higher gas flow rate was therefore, required to fluidise the bed.

Joshi et al. [14] gave a correlation for gas hold up which showed that D_c (and not H) affects the gas hold up, however, the correlation contained the bubble velocity. This, however, does not contradict the present result, since bubble velocity increases with bed height, it follows therefore that there is an implicit correlation between bed height and gas hold up.

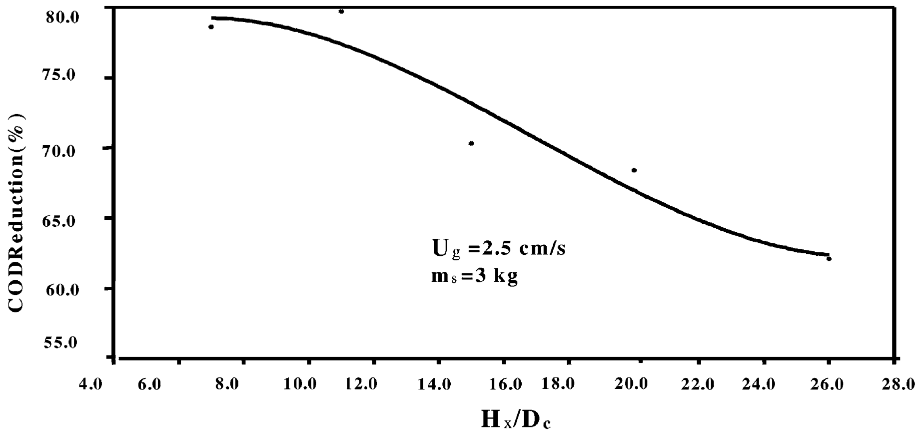

Fig. 5. Variation of COD reduction aspect ratio.

The decrease in biodegradation with increase in aspect ratio as shown in Fig. 5 confirms the fact that there was poor phase mixing at higher aspect ratios. Fig. 6 shows that there was an initial increase in biodegradation with increasing V_s/V_L. This increase could be due to the fact that, as more particles were added, there was an increase in surface area for micro-organism attachment.

However, as the particle loading increased, there reached a point when the solids inhibited phase mixing, and this led to a decrease in mass transfer. Under such conditions, aerobic micro-organisms are likely to die due to oxygen deficiency leading to decrease in biodegradation as depicted by the last part of the curve. Optimum V_s/V_L was about 21%, and this result compares quite well with result obtained in Fig. 2, and is consistent with the

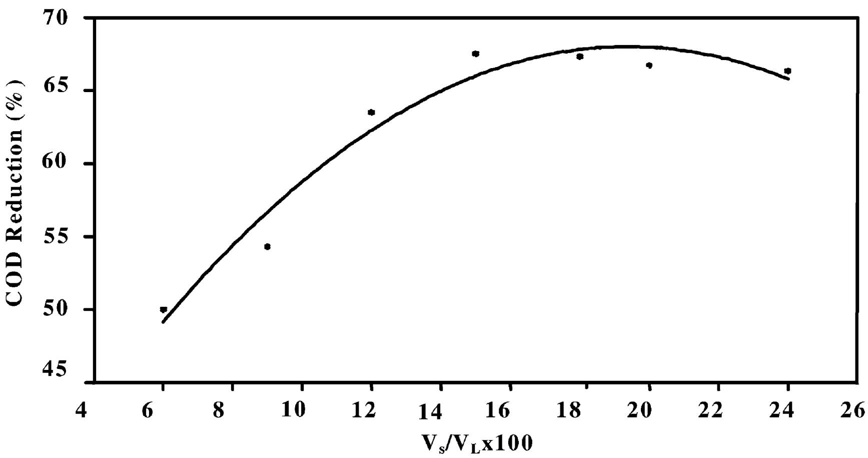

Fig. 6. Variation of COD reduction with V_s/V_L.

result obtained by [8]. With 9 vol.% of support particles, it was observed that the optimum substrate (liquid waste) loading required was at 2.5 m bed height. This was a trade off between the rate of biodegradation and the substrate loading. The pressure drop across the bed does increase with increase in bed height. Increase in bed pressure drop results in decrease in the stirring effect, it is therefore expected that there should be a decrease in phase mixing with an increased bed height.

4.2. Superficial gas velocity and gas hold up

Fig. 7 shows that there is a coincidence in the maximum percentage biodegradation for gas velocity and gas hold up. The optimum gas flow rate was about 2.7 cm/s and the ratio of U_{mf}/U_g was about 0.6. COD reduction was very low at $U_g > 4$ cm/s, this observation is consistent with the results of Chen et al. [15], where the column used had an aspect ratio of 11.8, with particles of density 1250 kg/m^3, which are comparable to the present work. Similarity of the response of biodegradation to U_g and ε_g confirms the result in Fig. 3, which indicates that there is almost a linear relation between gas hold up and gas velocity at low gas flow rate. It was unlikely that further increase in the gas hold up could cause the decrease in biodegradation as might be portrayed in Fig. 7, rather, the decrease in biodegradation was due to the increase in the gas velocity, which resulted in bubble coalescence. As expected, bubble coalescence reduces mass transfer and this is what caused the decrease in biodegradation at higher gas flow rates.

4.3. Superficial liquid velocity

Apparently, superficial liquid velocity did not significantly affect COD reduction, however, there was some increase in COD reduction with increase in U_L as shown in Fig. 8. This tendency could be attributed to an increase in mass transfer caused by increased liquid turbulence. However, the effect on biodegradation was not significant due to the fact that the liquid turbulence caused by high fluid flow rate was dampened at high bed height. It

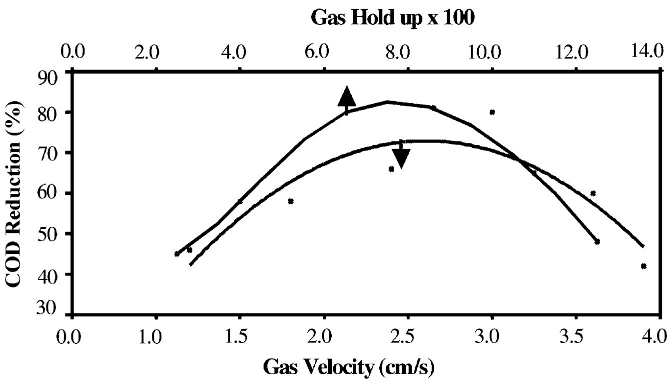

Fig. 7. Effect of gas velocity and gas hold up on biodegradation.

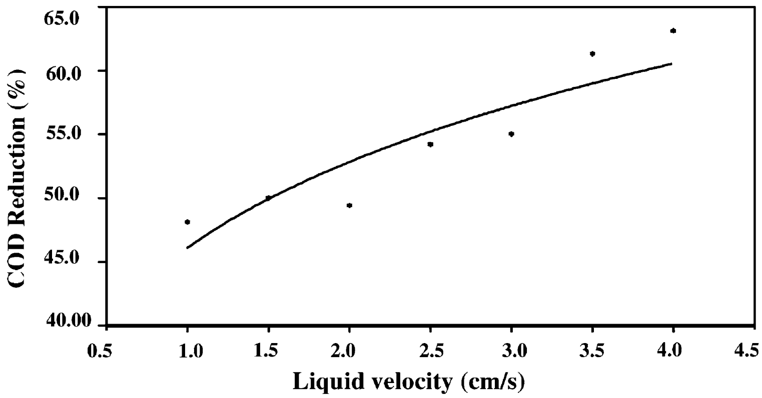

Fig. 8. Variation of liquid velocity with COD reduction.

was also observed that turbulence caused by liquid flow was not very significant beyond 1.5 m above the liquid injection point.

5. Conclusions

High aspect ratio results in low biodegradation, and the optimum bioreactor geometry was such that the aspect ratio was about 10 and V_S/V_L was 20%. The results indicate that under similar conditions as of the present work, it would be more economical to operate a biodegradation process in a wide column diameter with the aspect ratio not more than 10. The optimum superficial gas and liquid velocities were 2.7 and 0.4 cm/s, respectively. The results revealed that biodegradation increased with particle loading up to a maximum and decreased thereafter.

The result obtained showed that the high cost of operating a fluidised bed bioreactor, as a result of pumping liquid and gas, can be reduced by using low-density particles. However, if the low-density support particles are used, the aspect ratio should be low in order to achieve bed homogeneity at low gas flow rates. These results can be useful for a bioreactor design and the operation.

Acknowledgements

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