Chemical Engineering Journal 150 (2009) 63–68

Contents lists available at ScienceDirect

Chemical Engineering Journal

journal homepage: www.elsevier.com/locate/cej

Biological wastewater treatment in the inverse fluidised bed reactor

Włodzimierz Sokół[a,b,*], Alemayehu Ambaw[b], Belay Woldeyes[b]

[a] Department of Chemical and Bioprocess Engineering, University of Technology and Life Sciences, 3 Seminaryjna Street, 85-326 Bydgoszcz, Poland
[b] Department of Chemical Engineering, Addis Ababa University, P.O. Box 385, Addis Ababa, Ethiopia

ARTICLE INFO

Article history:
Received 20 October 2007
Received in revised form 4 May 2008
Accepted 3 December 2008

Keywords:
Inverse fluidised bed reactor
Biological wastewater treatment
Three-phase fluidised bed biological reactor
Aerobic wastewater treatment
Low-density biomass support

ABSTRACT

The biological wastewater treatment was investigated in the inverse fluidised bed reactor (IFBR) in which polypropylene particles of density 910 kg/m^3 were fluidised by an upward flow of gas. Measurements of chemical oxygen demand (COD) versus residence time t were performed for various ratios of settled bed volume to reactor volume (V_b/V_R) and air velocities u_g.

The largest COD removal was attained when the reactor was operated at the ratio $(V_b/V_R)_m = 0.55$ and an air velocity $u_{gm} = 0.024$ m/s. Under these conditions, the value of COD was practically at steady state for times greater than 30 h. Thus, these values of $(V_b/V_R)_m$, u_{gm} and t can be considered as the optimal operating parameters for a reactor when used in treatment of industrial wastewaters. A decrease in COD from 36,650 to 1950 mg/l, i.e. a 95% COD reduction, was achieved when the reactor was optimally controlled at $(V_b/V_R)_m = 0.55$, $u_{gm} = 0.024$ m/s and $t = 30$ h. The pH was controlled in the range 6.5–7.0 and the temperature was maintained at 28–30 °C.

The biomass loading was successfully controlled in an IFBR with support particles whose matrix particle density was smaller than that of liquid. The steady-state biomass loading depended on the ratio (V_b/V_R) and an air velocity u_g. In the culture conducted after switching from the batch to the continuous operation, the steady-state biomass loading was attained after approximately 2-week operation. In the cultures conducted after change in (V_b/V_R) at a set u_g, the steady-state mass of cells grown on the particles was achieved after about 6-day operation. For a set ratio (V_b/V_R), the biomass loading depended on u_g. With change in u_g at a set (V_b/V_R), the new steady-state biomass loading occurred after the culturing for about 2 days.

1. Introduction

The fluidised bed biofilm reactors, in which the biomass is fixed on particles of a bed fluidised by liquid upflow, are amongst the most effective in the class of spontaneously fixed biofilm reactors [1–7]. The large biofilm–liquid interfacial area, high-interfacial velocities and good mass transfer characteristics are the main advantages of this type of bioreactors [8–14].

A three-phase (gas–liquid–solid) fluidised bed reactor (TPFBR) has been successfully applied to aerobic biological treatment of industrial and municipal wastewaters [1–4]. The use of biomass support allows the partial replenishment of the fluidised bed without interrupting the operation in order to maintain high-microbial activity [4,9]. The vast growth surface afforded by the media results in a biomass concentration approximately an order of magnitude greater than that maintained in a suspended growth system [4,8].

However, the excessive growth of biomass on support media can lead to washout of bioparticles (particles covered by biofilm) from a reactor since the biomass loading can increase to such an extent that the bioparticles began to be carried over from the apparatus [2,9]. The application of the so-called *inverse fluidised bed*, in which low-density particles (matrix particle density smaller than that of liquid) are fluidised either by a downflow of liquid or an upflow of gas through a bed, allows the control of biomass loading in a reactor [1,10,11].

In a reactor containing low-density particles, fluidisation can be conducted either by an upward cocurrent flow of gas and liquid through a bed (Fig. 1) or by a downward flow of liquid and countercurrent upward flow of gas [1,12]. In the former, fluidisation is achieved by an upward flow of gas whereby the gas bubbles make the bed expanding downwards into the less dense mixture of gas and liquid [1,8]. In the latter, the bed is fluidised by a downward flow of a liquid counter to the net buoyancy force of the particles [9,12]. Such type of fluidisation, where fluidised bed expands downwards, is termed the *inverse fluidisation*.

Abbreviations: FBR, fluidised bed reactor; IFBR, inverse fluidised bed reactor; TPFBR, three-phase fluidised bed reactor.

* Corresponding author at: Department of Chemical Engineering, Addis Ababa University, P.O. Box 385, Addis Ababa, Ethiopia. Tel.: +251 11 1232417; fax: +251 11 1239480.
E-mail address: sokolwlodzimierz@yahoo.com (W. Sokół).

1385-8947/$ – see front matter © 2008 Elsevier B.V. All rights reserved.
doi:10.1016/j.cej.2008.12.021

Nomenclature

COD	chemical oxygen demand (mg/l)
D	dilution rate (h^{-1})
H_d	length of draught tube (m)
Q_g	aeration rate (m^3/s)
t	mean residence time (h)
u_g	superficial upflow air velocity (m/s)
u_L	superficial upflow liquid velocity (m/s)
V_b	volume of settled bed (m^3)
V_R	reactor volume (m^3)
w_s	solids loading (kg/m^3)

Greek symbols

ε_g	air hold-up
ε_L	liquid hold-up
ε_s	solid hold-up

Subscripts

cr	refers to critical
f	refers to minimum fluidisation velocity
m	denotes values giving the greatest COD reduction
o	refers to overall

Sokół and Halfani [8] have reported that a steady-state biomass loading was obtained in an inverse fluidised bed reactor (IFBR), in which polypropylene particles of density 910 kg/m^3 were fluidised by an upward co-current flow of gas and liquid. Nikolov and Karamanev [12] have achieved a constant biomass loading in an IFBR, in which low-density particles were fluidised by downflow of the liquid.

The aim of this work was to investigate the biological treatment of industrial wastewater in an IFBR, in which polypropylene particles of density 910 kg/m^3 were fluidised by an upward flow of gas through a bed. Experiments on COD reduction were performed for various ratios of settled bed volume to reactor volume (V_b/V_R), air velocities u_g and residence times t.

2. Characteristics of an inverse fluidised bed biological reactor

The application of a fluidised bed technique to biological wastewater treatment has brought a remark breakthrough [1,4]. A fluidised bed biological treatment technology owes its high-rate success to much higher surface area and biomass concentration than those that can be achieved in the conventional treatment processes [1–5,15,16]. A fluidised bed reactor (FBR) outperforms other

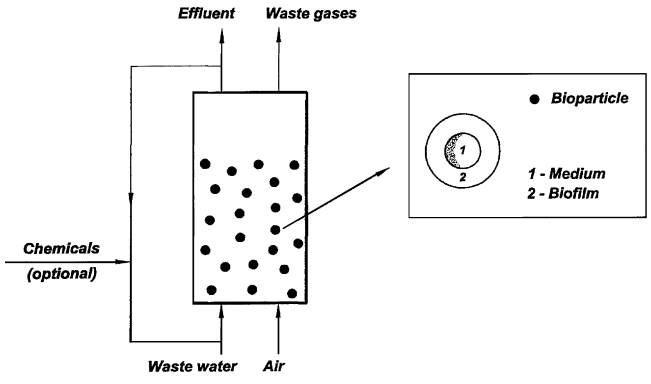

Fig. 1. Scheme of the inverse fluidised bed reactor (IFBR). Low density particles are fluidised by upward flow of gas (air).

reactor configurations used in biological wastewater treatment such as the activated sludge system (continuous stirred tank reactor) and packed-bed (or trickling-filter) reactor [1,7,8]. The superior performance of the FBR stems from the very high-biomass concentration (30–40 kg/m^3) that can be achieved due to immobilisation of cells onto and/or into the solid particles [1–4,15].

The application of low-density particles in a FBR allows the control of biomass loading in the apparatus and provides the high-oxygen concentration in the reacting liquid media [1–3,12]. One of the characteristics of a gas–liquid–solid fluidised bed of low-density particles which most distinguishes it from that of high-density particles (matrix particle density higher than that of liquid) is the axial non-homogeneity of the hold-up of the phases [1,17]. The non-homogeneity in the individual phase hold-ups affects the mass transfer properties, and hence the reactant conversion.

Tang and Fan [17] have studied the hydrodynamics of a gas–liquid–solid FBR containing spheres made of polystyrene, acetate, acryl or nylon, with sizes ranging from 1 to 6.35 μm in diameter. It has been found that the gas hold-up ε_g in the fluidised beds of 4.8 vol.% of polystyrene, acrylic or acetate particles was smaller than that achieved in a gas–liquid system. The liquid velocity u_L played a crucial role in the uniformity of the solids distribution in the bed. The higher velocity u_L the more uniform solids distribution in the bed was observed. The gas velocity u_g had only a slight effect on the distribution of axial solids hold-up ε_s.

Lu et al. [18] have studied the effects of operating parameters on a hold-up ε_g and liquid velocity u_L in a three-phase internal loop airlift reactor containing calcium alginate beads of different particle size. The parameters examined were the top clearance, aeration rate Q_g, draught tube length H_d and solids loading w_s. The average diameters of the particles were 1, 2 and 3.6 mm. Air and water were used as the gas and liquid phase, respectively. It has been found that at low-aeration rate ($Q_g < 9 \times 10^{-4}$ m^3/s) values of ε_g increased linearly with the Q_g. However, at high-aeration rate ($Q_g > 9 \times 10^{-4}$ m^3/s) values of ε_g increased only slightly with increase in Q_g due to bubble coalescence. Moreover, values of ε_g decreased with an increase in solids loading w_s. It has been established that the velocity u_L increased as the length of the draught tube H_d was increased. On the other hand, the velocity u_L decreased with an increase in the particle size and particles loading w_s.

Nikolov and Karamanev [12] have studied the aerobic degradation of glucose in an IFBR in which low-density particles were fluidised by downflow of the liquid. An aqueous glucose solution (0.4–12 g of glucose/l) and a mixed culture of aerobic heterotrophic microorganisms were used. Polyethylene granules with diameter 2–3 mm were used for fixation of the bacteria. It has been found that the reaction rate at a 90% conversion was 80 mg glucose consumed per litre reactor volume per hour. The maximum rate achieved was 1000 mg glucose/(l h). The results were compared with those obtained in the basic draft tube airlift reactor, that is, in the same reactor but without polyethylene particles. The IFBR was 3.6 times more effective at a glucose conversion of 90%. The ratio between the maximal glucose uptake rates was even greater, viz. 14.8 times.

The researchers [12] have also studied the ferrous ion (Fe^{2+}) oxidation by *Thiobacillus ferrooxidants*. Aqueous solutions of $FeSO_4$, containing 3–4 g Fe^{2+}/l and sulphuric acid for control of the pH value in the range of 2.0–2.5, were used as substrate. Polystyrene spheres with diameter 0.8–1 mm and density 330 kg/m were used as biomass support. It has been found that at a value of 70% conversion, the oxidation rate was 0.7 g Fe^{2+}/(l h). Compared with the data obtained in the reactor under identical conditions with only suspended bacteria, these values were 4.4 and 10 times larger, respectively.

The hydrodynamics of a three-phase internal loop airlift reactors containing soft polyurethane foam particles has been studied by several investigators [5,19,20]. It has been found that the gas

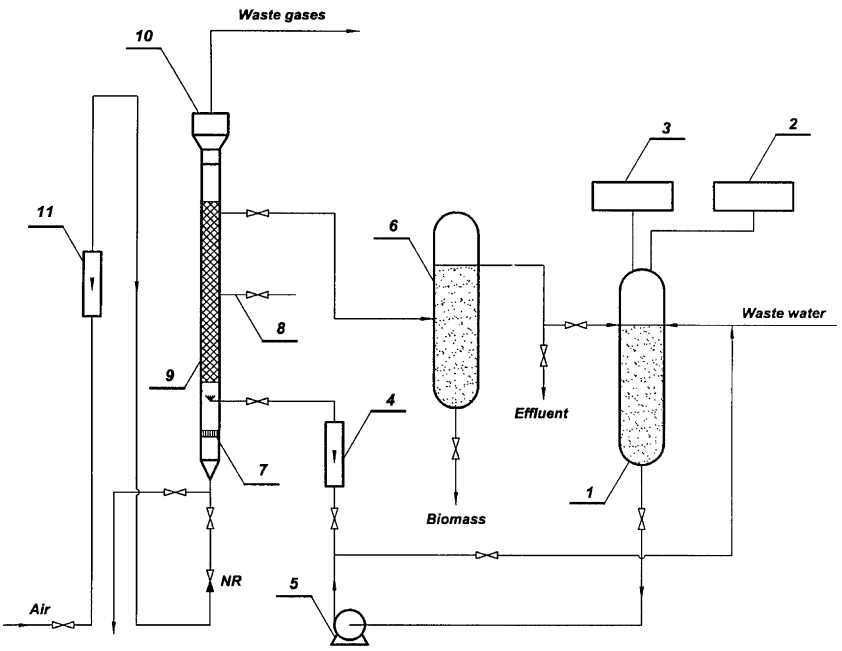

Fig. 2. Schematic diagram of the experimental apparatus: 1, reservoir; 2, temperature control system; 3, pH control system; 4, liquid rotameter; 5, pump; 6, intermediate reservoir; 7, air distributor; 8, sampling; 9, fluidised section; 10, disengaging section; 11, air rotameter.

hold-up ε_g decreased significantly with increasing solids loading w_s and was proportional to ${u_g}^{1.2}$ [19,20]. Miyahara and Kawate [19] have reported that values of ε_g in a three-phase internal loop airlift reactor containing low-density particles decreased significantly when the solid hold-up ε_s was larger than 0.2.

Rusten et al. [14] have applied particles made of ethylene in an IFBR used for treatment of wastewaters from dairy and food industry. With this support, a considerably higher treatment efficiency was achieved than that obtained in a bioreactor with a sand support. The biomass loading was practically constant over the entire period of a 3-month operation, and yet, the reactor performed very well with a 95% removal of total *COD*.

Sokół [21] has reported a 95% COD removal in aerobic treatment of brewery wastewater in a TPFBR containing particles made of polypropylene. In the same reactor, Sokół [6] has achieved a 90% COD reduction in aerobic treatment of highly toxic refinery wastewater. Sokół and Korpal [2] have reported a 98% COD reduction in treatment of phenolic wastewaters in a TPFBR with light particles.

3. Experimentation

3.1. Experimental setup

Experiments were performed in the reactor shown in Fig. 2. A growing medium, stored in a reservoir (1), was pumped into the bottom of the reactor by a centrifugal pump (5). Before entering the bed, the liquid was mixed with air by means of a sparger. The air was introduced into the bed through a distributor (7) whose plate had 200 mm × 4 mm diameter holes on a triangular pitch. The fluidised bed section (9) had a 20 cm internal diameter and was 6 m high. It was ended by a disengaging cap (10) with a 60 cm internal diameter and a height of 80 cm. The biomass sloughed off from the particles was separated from the effluent in a vessel (6) and removed from the system. The flow rate of the liquid was measured by a rotameter (4) and controlled by a ball valve. The air flow rate was measured by a rotameter (11) and controlled by a needle valve. The pH was adjusted by a control system (3), consisting of a pH-meter and micropumps supplying base or acid; as required. The temperature control system (2) consisted of a coil with cold water and an electric heater coupled with a contact thermometer.

The biomass support was the polypropylene particles of density 910 kg/m^3 whose dimensions are given in work [1]. The specific surface area of the biomass support could have been controlled up to 400 m^2/m^3 by introducing the corresponding volume of the particles into the reactor.

3.2. Feed and microorganisms

The growing medium was the wastewater whose composition is given in Table 1. The wastewater was enriched in mineral salts by adding the following (mg/l): $(NH_4)_2SO_4$, 500; KH_2PO_4, 200; $MgCl_2$, 30; NaCl, 30; $CaCl_2$, 20; and $FeCl_3$; 7 as recommended by Sokół [22], and Sokół and Migiro [23].

The inoculum was the activated sludge taken from the biological treatment unit operated at the chemical process plant from which the wastewater was used in this research.

3.3. Methodology

Sokół and Halfani [8] have reported that the largest values of air hold-up ε_g were obtained when a TFBBR with polypropylene particles was controlled at ratio (V_b/V_R) in the range 0.50–0.60. Sokół [6] has established that the optimal ratio (V_b/V_R) for a TFBBR when

Table 1
Composition of a wastewater (COD = 36,650 mg/l) and effluent from a reactor optimally controlled at $(V_b/V_R)_m = 0.55$, $u_{gm} = 0.024$ m/s and $t = 30$ h.

Constituent	Concentration (×10^3 mg/l)	
	Wastewater	Effluent
Phenol	2562	3.46
o-Cresol	6448	1.03
m-Cresol	3919	1.39
Isopropylphenol	1052	1.91
2,4-Dimethylphenol	1979	0.74
3,4-Dimethylphenol	914	0.38
3,5-Dimethylphenol	2741	0.96
Benzene	1457	4.27
Toluene	1398	4.15

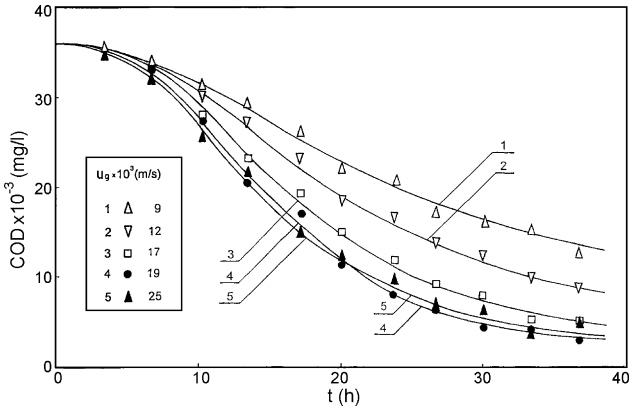

Fig. 3. Dependence of COD values on time (t) for ratio (V_b/V_R) = 0.50 and various air velocities (u_g).

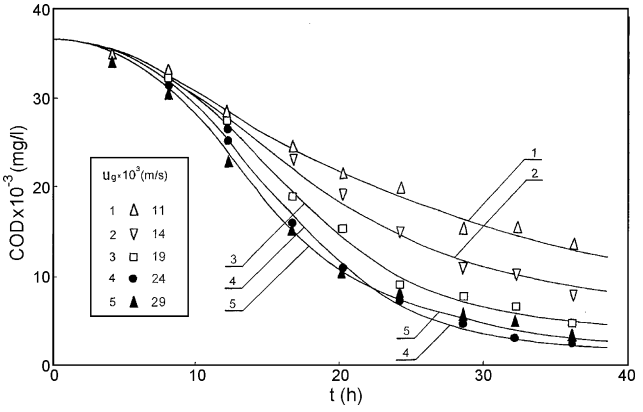

Fig. 4. Relationship between COD values and time (t) for ratio (V_b/V_R) = 0.55 and various air velocities (u_g).

used in treatment of refinery wastewater was equal to 0.55. Therefore, in this study experiments were performed for the ratios (V_b/V_R) equal to 0.50, 0.55 and 0.60. This was to cover the searched range of (V_b/V_R) from 0.50 to 0.60 in step 0.05 which is sufficient accuracy for industrial practice.

The air velocities u_g applied in experiments are given in Figs. 3–6.

3.3.1. Biomass culturing

The particles and the growing medium were introduced into the reactor to give a ratio (V_b/V_R) = 0.50. To start growth of the

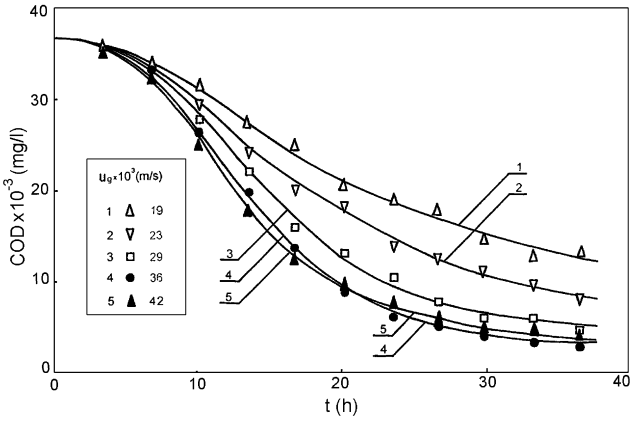

Fig. 5. Dependence of COD values on time (t) for ratio (V_b/V_R) = 0.60 and various air velocities (u_g).

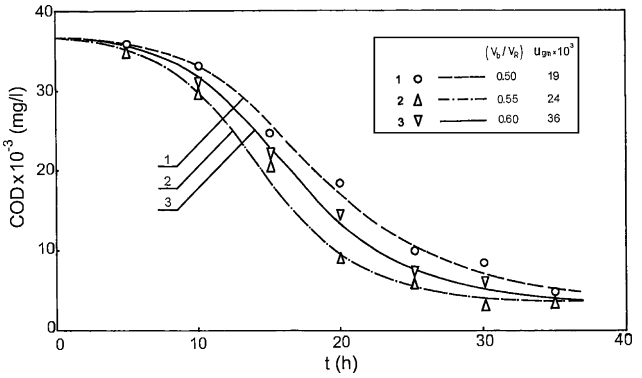

Fig. 6. Relationship between COD values and time (t) for wastewater treatment conducted in a reactor controlled at the values of $(V_b/V_R)_m$ and u_{gm} for which the greatest COD removals were achieved in runs shown in Figs. 3–5.

microorganisms on the particles, a batch culture was first initiated by introducing about 15 l of the inoculum into the reactor. Next, the culture was incubated for approximately 48 h to encourage cell growth and the adhesion of freely suspended biomass on the particles. The air was supplied at the flow rate of 0.025 m^3/s and this was found to be sufficient for biomass growth [1–3]. The pH was controlled in the range 6.5–7.0 and the temperature was maintained at 28–30 °C.

When the biofilm had begun to grow on the particles, the growing medium was started to be pumped into the reactor at a dilution rate $D = 0.30\,h^{-1}$. This value of D corresponded to the smallest time t applied for the ratio (V_b/V_R) = 0.50 ($t = 1/D = 3.33$ h in Fig. 3). Next, the air velocity was set at the smallest value applied for the (V_b/V_R) = 0.50 (u_g = 0.009 m/s in Fig. 3) and the cultivation was continued until the constant biomass loading was achieved in a reactor. The occurrence of the steady-state biomass loading was established by weighting the mass of cells grown on the support. The biomass was scraped from sample particles and dried at temperature 105 °C for 1 h. It was considered that the steady state occurred when the weight of biomass in two consecutive samples differed less than 5%. The constant biomass loading was attained in a reactor after the cultivation for approximately 2 weeks.

3.3.2. Wastewater treatment

When the steady-state biomass loading was achieved in a reactor, a sample liquid was withdrawn from the apparatus and *COD* was measured by the procedure recommended by Verstraete and van Vaerenbergh [24]. It was established that once the constant biomass loading occurred in a reactor, the value of *COD* was practically at steady state.

Next, the air velocity was increased stepwise to its next value of u_g applied for (V_b/V_R) = 0.50 (u_g = 0.012 m/s in Fig. 3) and the cultivation was continued until the new steady-state biomass loading was achieved. When this was attained, *COD* was measured by the method mentioned earlier [24]. These experiments for (V_b/V_R) = 0.50 were conducted for all values of u_g shown in Fig. 3.

Then the dilution rate was decreased stepwise to its next value applied for (V_b/V_R) = 0.50 ($t = 1/D = 6.67$ h in Fig. 3) and the air velocity was re-set to its smallest value applied for the (V_b/V_R) = 0.50 (u_g = 0.009 m/s in Fig. 3). The cultivation was continued until the steady-state biomass loading was achieved. When this occurred, *COD* was measured following the procedure mentioned earlier [24]. These experiments were conducted for all air velocities u_g and times t shown in Fig. 3. The results are given in Fig. 3.

The above experiments were also performed for the ratios (V_b/V_R) equal to 0.55 and 0.60. In order to get the ratio (V_b/V_R) = 0.55, an adequate volume of biomass-free particles was added to a reactor at the end of experimentation for (V_b/V_R) = 0.50. Similarly, the

ratio $(V_b/V_R) = 0.60$ was obtained by the addition of fresh particles to a reactor at the end of experimentation for $(V_b/V_R) = 0.55$. The results of the experiments are shown in Figs. 4 and 5.

In order to establish time t for which the value of *COD* was practically at steady state, experiments on *COD* reduction were performed for these values of u_{gm} for which the largest *COD* removals were achieved in runs shown in Figs. 3–5. The results of experiments are shown in Fig. 6.

It should be pointed out that the air velocities u_g applied in the experiments were several times larger than the minimum fluidisation velocity u_{gf}. This was possible because the reactor was operated at the ratios (V_b/V_R) smaller than the critical values of $(V_b/V_R)_{cr}$ [1,8]. At the ratios (V_b/V_R) equal to, or larger than, the $(V_b/V_R)_{cr}$ movement of the whole bed was impossible: the particles either remained at the top of the reactor or they settled at its bottom. On the other hand, the air velocities u_g were smaller than the critical velocity u_{gcr} at which the entire bed settled at the reactor bottom.

Stratification of the particles coated with the biomass led to their movement to the base of the bed where concentrations of constituents of the wastewater were the highest. This was desirable since the constituents could penetrate far into the biofilm so that most of the biomass was active [1,5,8].

4. Results and discussion

It can be seen in Figs. 3–5 that, for a set time t and ratio (V_b/V_R), a concentration of *COD* in effluent depended on the air velocity u_g (Figs. 3–5). As can be noted in Fig. 3, for a set t values of *COD* were decreasing with an increase in u_g up to 0.020 m/s. The smallest value of *COD* was attained for $u_{gm} = 0.020$ m/s. For velocities u_g larger than 0.020 m/s, values of COD were increasing with an increase in u_g. This can be explained by the fact that with an increase in u_g up to 0.020 m/s, an interfacial (air–liquid) area increased [25], and consequently the amount of the oxygen supplied for biomass growth increased [1,4,6]. Thus, for the u_g smaller than 0.020 m/s, oxygen was the limiting factor for biomass growth. On the other hand, for the air velocities greater than 0.020 m/s, the degradation rate of the constituents of the wastewaters was the controlling factor of the treatment process [1,11].

The velocity u_{gm}, for which the smallest value of *COD* was obtained for a set t, depended on the ratio (V_b/V_R), and hence on volume V_b of the particles applied in the reactor (Figs. 3–5). With the V_b increasing, the value of u_{gm} increased. Thus, a large volume of the particles leads to an increase in the amount of the air required for biomass growth, and consequently to an increase in the resulting energy cost [26].

However, an increase in the volume of particles results in more biomass grown on the media, which, in turn, improves *COD* removal. It can be noticed in Figs. 3–5 that the smallest value of *COD*, and hence the largest *COD* removal, was attained at $(V_b/V_R) = 0.55$. An increase in *COD* removal with an increase in the (V_b/V_R) from 0.50 to 0.55 can be attributed to the fact that for increasing (V_b/V_R), more biomass grown on the particles participated in degradation of the constituents of the wastewater [1,27–29]. On the other hand, a decrease in *COD* removal observed with an increase in (V_b/V_R) from 0.55 to 0.60 was due to the fact that in this case, a significant volume of the reactor was occupied by the particles, and consequently the aeration characteristics of the bed has worsened [6,30,31].

The value of *COD* was practically at steady state for times t greater than 30 h (Fig. 6). The largest *COD* removal occurred when the reactor was operated at $(V_b/V_R)_m = 0.55$ and $u_{gm} = 0.024$ m/s. A decrease in *COD* from 36,650 to 1950 mg/l, i.e. a 95% *COD* reduction, was achieved when a reactor was optimally controlled at $(V_b/V_R)_m = 0.55$, $u_{gm} = 0.024$ m/s and $t = 30$ h.

The biomass loading was successfully controlled in a reactor containing low-density particles used as biomass support. This was due to particle geometry and particularly availability of the internal surface and the grooves on external surface of the particles for biomass growth. With such geometry of the particles, shear forces occurring between the particles and the liquid sloughed off excess of biomass mainly from the external, and to less extend from the internal, surface of the particles. Furthermore, attrition, associated with particle–particle and particle–wall collisions, of biomass grown in the grooves and on the internal surface was less abrupt than the cells grown on the external surface of the particles.

The steady-state biomass loading in a reactor depended on the ratio (V_b/V_R) and an air velocity u_g. In the cultures conducted after change in (V_b/V_R) at a set u_g, the steady-state mass of cells grown on the support media was achieved after approximately 4 days of operation. With change in u_g at a set (V_b/V_R), the new steady-state biomass loading occurred after the culturing for about 2 days.

5. Conclusions

From analysis of the experimental results obtained in this research, the following conclusion can be drawn:

1. A concentration of *COD* in effluent depended on mean residence time t, ratio of bed volume to bioreactor volume (V_b/V_R) and an air velocity u_g. For set t and (V_b/V_R), *COD* removal was increasing with an increase in u_g, attaining the largest value at u_{gm}. The values of u_{gm}, at which the lowest *COD* values were obtained, depended on the ratio (V_b/V_R) and they increased with an increase in (V_b/V_R). Similarly, for set t and u_g, *COD* removal was increasing with an increase in (V_b/V_R), attaining the largest value at $(V_b/V_R)_m = 0.55$.
2. The largest *COD* removal was achieved when the reactor was controlled at the ratio $(V_b/V_R)_m = 0.55$, air velocity $u_{gm} = 0.024$ m/s and residence time $t = 30$ h. Thus, these values of $(V_b/V_R)_m$, u_{gm} and t can be considered as the optimal operating parameters for the inverse fluidised bed reactor (IFBR) when used in industrial wastewater treatment.
3. A decrease in *COD* from 36,650 to 1950 mg/l, i.e. a 95% *COD* reduction, was achieved in wastewater treatment conducted in a reactor optimally controlled at $(V_b/V_R)_m = 0.55$, $u_{gm} = 0.024$ m/s and $t = 30$ h. The pH was controlled in the range 6.5–7.0 and the temperature was maintained at 28–30 °C.
4. The biomass loading was successfully controlled in an IFBR with low-density particles used as biomass support. The steady-state biomass loading depended on the ratio (V_b/V_R) and an air velocity u_g. In the cultures conducted after change in (V_b/V_R) at a set u_g, the constant mass of cells grown on the particles was achieved after 6 days of operation. With change in u_g at a set (V_b/V_R), the new steady-state biomass loading occurred after the culturing for about 2 days.

Acknowledgement

A financial support received from Deutscher Akademischer Austauch Dienst (DAAD) by one of the authors (WS) during his stay at Addis Ababa University is gratefully acknowledged.

References

[1] W. Sokół, W. Korpal, Chem. Eng. J. 118 (2006) 199.
[2] W. Sokół, W. Korpal, J. Chem. Technol. Biotechnol. 80 (2005) 884.
[3] W. Sokół, W. Korpal, Biochem. Eng. J. 20 (2004) 49.
[4] W.K. Shieh, D.K. Keenan, Fluidized bed biofilm reactor for wastewater treatment, in: A. Fiechter (Ed.), Advances in Biochemical Engineering/Biotechnology, vol. 33, Springer, Berlin, 1986, pp. 132–168.
[5] D.G. Karamanev, T. Nagamune, I. Endo, Chem. Eng. Sci. 47 (1992) 3581.

[6] W. Sokół, Biochem. Eng. J. 15 (2003) 1.
[7] P. Hüppe, H. Hoke, D.C. Hempel, Chem. Eng. Technol. 13 (1990) 73.
[8] W. Sokół, M.R. Halfani, Biochem. Eng. J. 3 (1999) 185.
[9] A.G. Livingston, H.A. Chase, Chem. Eng. J. 45 (1991) B35.
[10] O. Nore, C. Briens, A. Margaritis, G. Wild, Chem. Eng. Sci. 47 (1992) 3573.
[11] W.J. Wu, S.J. Hwang, C.M. Chang, Chem. Eng. Sci. 49 (1994) 1465.
[12] L. Nikolov, D. Karamanev, Can. J. Chem. Eng. 65 (1987) 214.
[13] W. Sokół, M.R. Halfani, Hydrodynamic characteristics of a three-phase fluidised bed bioreactor with the KMT® biomass support, in: Proceedings of the Third International Meeting on Chemical Engineering and Biotechnology, ACHEMASIA'94, Beijing, PR China, 1994, pp. 139–145.
[14] B. Rusten, H. Odegaard, A. Lundar, Water Sci. Technol. 26 (1992) 703.
[15] D.F. Walker, E.P. Austin, The use of plastic and porous biomass supports in a pseudo-fluidised bed for effluent tratment, in: P.F. Cooper, B. Atkinson (Eds.), Biological Fluidised Bed Treatment of Water and Wastewater, Ellis Horwood, Chichester, UK, 1981, pp. 272–284.
[16] A.K.Y. Wu, K.D. Wisecarver, Advances in Fluidization Engineering, AIChE Symposium Series No. 276, 86 (1990) 113.
[17] W.T. Tang, L.S. Fan, AIChE J. 35 (1989) 355.
[18] W.J. Lu, S.J. Hwang, Chem. Eng. Sci. 50 (1995) 1301.
[19] T. Miyahara, O. Kawate, Chem. Eng. Sci. 48 (1993) 127.
[20] M. Kennard, M. Janekeh, Biotechnol. Bioeng. 38 (1991) 1261.
[21] W. Sokół, Biochem. Eng. J. 8 (2001) 203.
[22] W. Sokół, Chem. Eng. J. 55 (1994) B47.
[23] W. Sokół, C.L.C. Migiro, J. Chem. Tech. Biotechnol. 54 (1992) 223.
[24] W. Verstraete, E.E. van Vaerenbergh, Aerobic activated sludge, in: H.-J. Rehm, G. Reed (Eds.), Biotechnology, vol. 8, VCH Verlagessellschaft mbH, Weinheim, 1986, pp. 51.
[25] J.C. Lee, P.S. Buckley, Fluid mechanics and aeration characteristics of fluidised beds, in: P.F. Cooper, B. Atkinson (Eds.), Biological Fluidised Bed Treatment of Water and Wastewater, Ellis Horwood, Chichester, UK, 1981, pp. 62–74.
[26] D.H. Wheeldon, R.W. Bayley, Economic studies of biological fluidised-beds for wastewater treatment, in: P.F. Cooper, B. Atkinson (Eds.), Biological Fluidised Bed Treatment of Water and Wastewater, Ellis Horwood, Chichester, UK, 1981, pp. 306–328.
[27] M.A. Young, R.G. Carbonell, D.F. Ollis, AIChE J. 37 (1991) 403.
[28] K. Schügerl, Can. J. Chem. Eng. 67 (1991) 178.
[29] I. Coelhoso, R. Boaventura, A. Rodrigues, Biotechnol. Bioeng. 40 (1992) 625.
[30] K. Fujie, H.Y. Hu, Y. Ikeda, K. Urano, Chem. Eng. Sci. 47 (1992) 3745.
[31] E.T. Yoong, P.A. Lant, P.F. Greenfield, Water Sci. Technol. 36 (1997) 75.