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Evaluation of the Inverse Fluidized Bed Biological Reactor for Treating High-Strength Industrial Wastewaters

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Abstract

The aim of this work was to investigate the aerobic degradation of high-strength industrial (refinery) wastewaters in the inverse fluidized bed biological reactor, in which polypropylene particles of density 910 kg/m^3 were fluidized by an upward flow of gas through a bed. 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*. The largest COD reduction, namely, from 54,840 to 2190 mg/l, *i.e.* a 96% COD decrease, was achieved when the reactor was operated at the ratio (V_b/V_R) = 0.55, air velocity u = 0.046 m/s and t = 65 h. Thus, these values of (V_b/V_R), u and t can be considered as the optimal operating parameters for a reactor when used in treatment of high-strength refinery wastewaters. In the treatment operation conducted in a reactor optimally controlled at (V_b/V_R) = 0.55, u = 0.046 m/s and t = 65 h, the conversions obtained for all phenolic constituents of the wastewater were larger than 95%. The conversions of about 90% were attained for other hydrocarbons.

1. Introduction

The application of a fluidized bed technology to biological wastewater treatment has brought a remarkable breaktrough. The 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. A fluidized bed biological reactor (FBBR) has attracted considerable interest as an alternative to the conventional suspended growth and fixed-film wastewater treatment processes due to its high efficiency performance.

Treatment of industrial wastewaters requires a great deal of space when using systems based on activated sludge in which the retention time is many days [1]. On the other hand, a FBBR is capable of achieving treatment in low retention time because of the high biomass concentrations that can be achieved in the apparatus [2,3].

Among the compounds of refinery wastewaters that are the greatest biohazards to the environment are phenols (monohydric and polyhydric), derivatized phenols, polycyclic organics (polynuclear aromatic hydrocarbons), thiocyanates, ammonia and cyanides [4].

A three-phase (gas-liquid-solid) FBBR has been successfully applied in aerobic biological treatment of industrial and municipal wastewaters [5-8]. The reactor outperforms other reactor configurations used in wastewater treatment such as the activated sludge system and packed-bed (or trickling-filter) reactor [9-12]. The superior performance of the FBBR stems from the very high biomass concentration (up to 30 - 40 kg/m^3) that can be achieved due to immobilisation of cells onto or into the solid particles.

However, the excessive growth of biomass on support media can lead to the channelling of bioparticles (support particles covered by biomass) in fluidized bed since the biomass loading can increase to such an extent that the bioparticles began to be carried over from a reactor. The application of a low density (matrix particle density smaller than that of liquid) biomass support in a reactor allows the control of biomass loading and provides the high oxygen concentration in the reacting liquid media [2,13].

In a FBBR containing low-density particles, fluidiza-

tion can be conducted either by an upward co-current flow of gas and liquid through a bed (**Figure 1**) [2,3] or by a downward flow of liquid and countercurrent upward flow of gas [5,14]. In the former, fluidization 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. In the latter, the bed is fluidized by a downward flow of a liquid counter to the net buoyancy force of the particles. Such type of fluidization is termed the *inverse fluidization*.

Sokół and Halfani [13] have reported that a steady-state biomass loading was obtained in a FBBR, in which polypropylene particles of density 910 kg/m^3 were fluidized by an upward co-current flow of gas and liquid. Karamanev *et al*. [14] have achieved a constant biomass loading in a FFBR, in which low density particles were fluidized by downflow of the liquid. Rusten *et al*. [15] have demonstrated that a steady-state biomass loading was attained in a FFBR containing low-density particles made of polyethylene.

The aim of this work was to investigate the aerobic degradation of high strength wastewaters in a FFBR, in which polypropylene particles of density 910 kg/m^3 were fluidized 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 and residence times t.
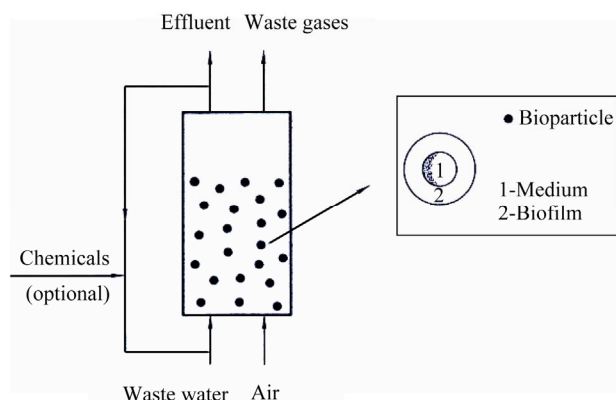

Figure 1. Scheme of the inverse fluidized bed biological reactor.

2. Experimentation

2.1. Experimental Set-up

Experiments were conducted in the reactor shown in **Figure 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 to the bed through a distributor 7 whose plate had 200 holes of 4 mm diameter on a triangular pitch.
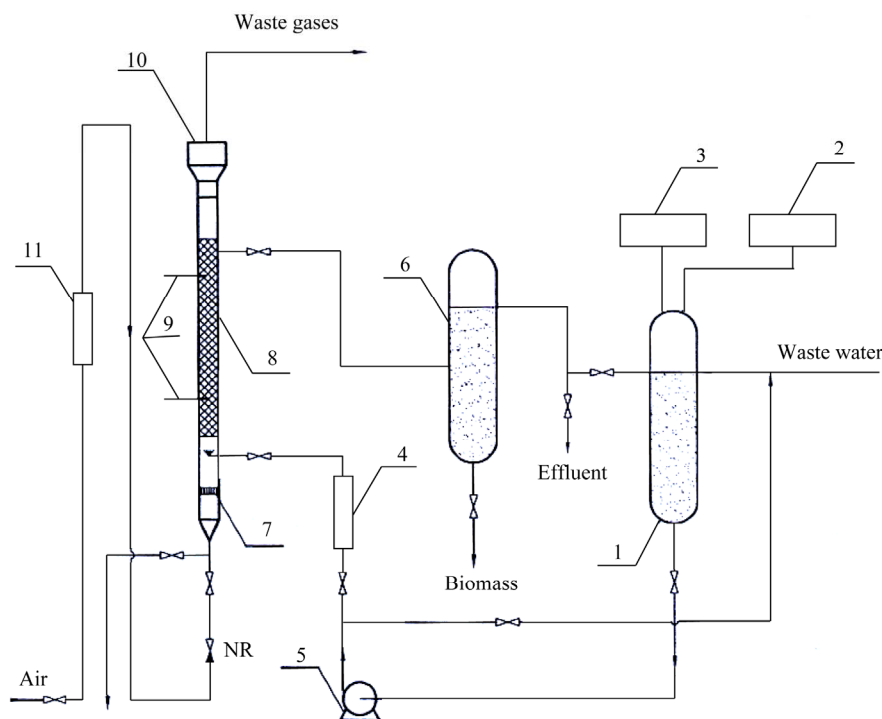

Figure 2. Schematic diagram of the experimental apparatus: 1, reservoir; 2, temperature control system; 3, pH control system; 4, liquid rotameter; 5, pump; 6, intermediate reservoir; 7, air distributor; 8, sampling; 9, fluidized section; 10, disengaging section; 11, air rotameter.

The fluidised bed section 9, made of Duran glass, 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 using a rotameter 11 and controlled by a needle valve. The pH was adjusted by a control system 3, consisting of a pH-meter and micro-pumps 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 **Figure 3**.

2.2. 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 [1].

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

2.3. Methodology

Sokół and Korpal [1] have established that the optimal ratio (V_b/V_R) for a FBBR when used in biological wastewater treatment 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 applied in experiments are given in **Figures 4-6**.

2.3.1. 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 microorganisms on the particles, a batch culture was first initiated by introducing about 15 l of the inoculum into the reactor. Then 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 a flow rate of 0.025 m^3/s and this was found to be sufficient for biomass growth [1,2]. The pH was controlled in the range 6.5 - 7.0 and the temperature was maintained at 28°C - 30°C.
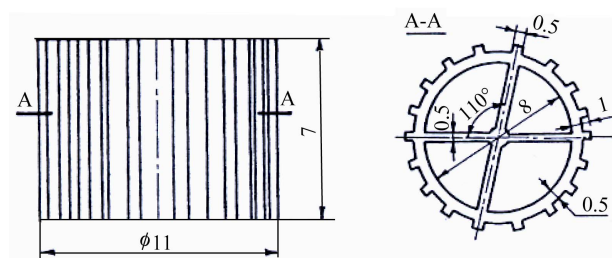

Figure 3. Dimensions (in mm) of biomass support.

Table 1. Composition of feed (wastewater) and effluent from a reactor optimally controlled at $(V_b/V_R)_m = 0.55$, $u_m = 0.046$ m/s and $t = 65$ h.

| Constituent	Concentration ×10^3 mg/l		Fractional conversion
	Feed	Effluent	
o-Cresol	15250	597.8	0.96
m-Cresol	8240	326.2	0.96
3,5 Dimethylphenol	7050	421.4	0.94
Phenol	5740	174.6	0.97
2,4 Dimethylphenol	4760	286.2	0.94
Benzene	3450	102.5	0.97
Toluene	2972	88.6	0.97
3,4-Dimethylphenol	2190	107.2	0.95
Isopropylphenol	2180	87.4	0.96
o-Xylene	1130	57.9	0.95
2,6-Dimethylphenol	980	67.3	0.93
C$_3$-Phenyl	240	16.4	0.93
Ethylphenol	190	5.6	0.97
C$_4$-Phenyl	70	6.3	0.91
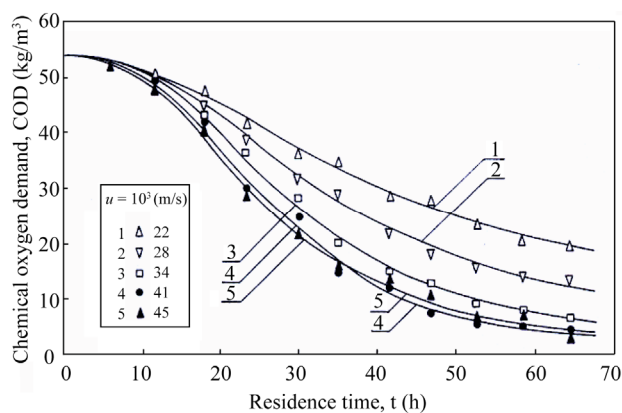
Figure 4. Relationship between chemical oxygen demand COD and residence time t for ratio $(V_b/V_R) = 0.50$ and various air velocities u.

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 $D = 0.15$ h^{-1}. This value of D corresponded to the smallest time t applied for the ratio $(V_b/V_R) = 0.50$ ($t = 1/D = 6.67$ h in **Figure 4**). Next, the air velocity u was set at the smallest value applied for the $(V_b/V_R) = 0.50$ ($u = 0.022$ m/s in **Figure 4**) 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 the 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 steady-state biomass loading was attained in a reactor after the cultivation for approximately two weeks.

2.3.2 Treatment Operation

When the steady-state biomass loading was achieved, a sample liquid was withdrawn from the reactor and COD was measured by the procedure recommended by Verstraete and van Vaerenbergh[16]. 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 u was increased stepwise to its next value applied for $(V_b/V_R) = 0.50$ ($u = 0.028$ m/s in **Figure 4**) and the cultivation was continued until the new steady-state biomass loading was obtained. When this was attained, a value of COD was measured by the method mentioned earlier [16]. These experiments for $(V_b/V_R) = 0.50$ were conducted for all values of u shown in **Figure 4**.

Then the dilution rate D was decreased stepwise to its next value applied for $(V_b/V_R) = 0.50$ ($t = 1/D = 13.34$ h in **Figure 4**) and the air velocity u was re-set to its smallest value applied for the $(V_b/V_R) = 0.50$ ($u = 0.022$ m/s in **Figure 4**). The cultivation was continued until the steady-state biomass loading was achieved. When this occurred, a value of COD was measured following the procedure mentioned earlier [16]. These experiments were conducted for all air velocities u and residence times t shown in **Figure 4**. The results of the experiments are given in **Figure 4**.

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 experimenttation 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 **Figures 5** and **6**.

In order to determine the largest COD removal, experiments on COD reduction were performed for those values of u_m for which the largest COD reductions were achieved in runs shown in **Figures 4-6**. The results of the experiments are given in **Figure 7**.

It should be pointed out that the air velocities u applied in the experiments were several times larger than the minimum fluidization velocity u_f. This was possible
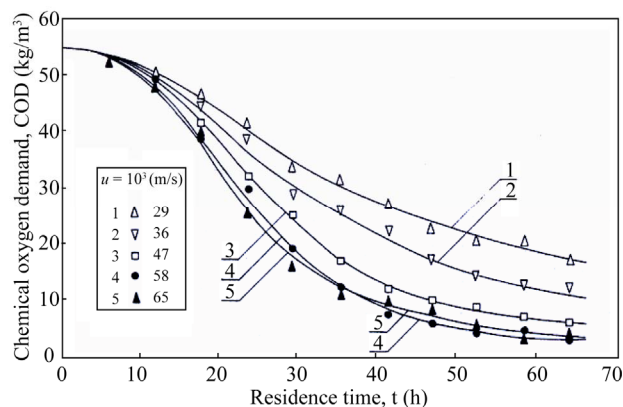

Figure 5. Dependence of chemical oxygen demand COD on residence time t for ratio $(V_b/V_R) = 0.55$ and various air velocities u.
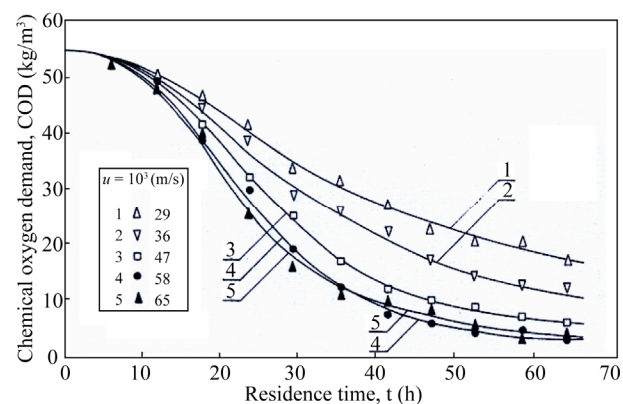

Figure 6. Relationship between chemical oxygen demand COD and residence time t for ratio $(V_b/V_R) = 0.60$ and various air velocities u.
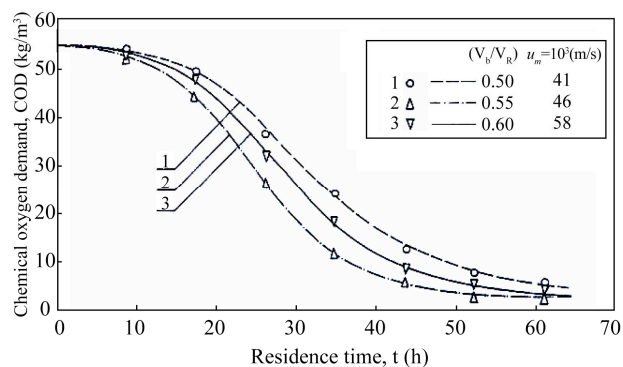

Figure 7. Dependence of chemical oxygen demand COD on residence time t for treatment operation conducted in a reactor controlled at values of $(V_b/V_R)_m$ and u_m for which the greatest COD removals were obtained in runs shown in Figures 4-6.

because the reactor was operated at the ratios (V_b/V_R) smaller than the critical values of $(V_b/V_R)_{cr}$ [13]. At the ratios (V_b/V_R) equal to, or larger than, $(V_b/V_R)_{cr}$, the

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 were smaller than the critical velocity u_{cr} 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 [1,2].

3. Results and Discussion

It can be seen in **Figures 4-6** that, for a set time t and ratio (V_b/V_R), a concentration of COD in effluent depended on the air velocity u. A reduction in COD initially increased, and then decreased with an increase in u. It can be noticed, for example, in **Figure 4** that for a set t the values of COD were decreasing with an increase in u up to 0.041 m/s. The smallest value of COD was attained for $u = 0.041$ m/s. For the air velocities u larger than 0.041 m/s, the values of COD were increasing with an increase in u. This can be explained by the fact that with an increase in u up to 0.041 m/s, an interfacial (air-liquid) area increased, and consequently the amount of the oxygen supplied for biomass growth increased [1]. Thus, for the u smaller than 0.041 m/s, oxygen was the limiting factor for biomass growth. On the other hand, for the air velocities greater than 0.041 m/s, the degradation rate of the constituents of the wastewaters was the controlling factor of the treatment process [3].

The value of u_m for which the largest decrease in 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 (**Figures 4-6**). With increasing V_b, the value of u_m increased. Thus, a large volume of the particles can lead to an increase in the amount of the air required for biomass growth, and consequently to an increase in the resulting energy cost [13].

It can be noted in **Figures 4-6** that for set t and u, a decrease in COD values depended on the ratio (V_b/V_R). The largest reduction in COD 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. 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 had worsened [2].

It can be noticed in **Figure 7** that the values of COD were practically at steady state for times t greater than 65 h. The largest COD removal occurred when the reactor was operated at $(V_b/V_R)_m = 0.55$ and $u_m = 0.046$ m/s. A decrease in COD from 54,840 to 2190 mg/l, that is, a 96% COD removal, was achieved when a reactor was optimally controlled at $(V_b/V_R)_m = 0.55$, $u_m = 0.046$ m/s and $t = 65$ 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, the attrition, associated with particle-particle and particle-wall collisions, sloughed off biomass grown in the grooves and on the internal surface was less abrupt than the cells grown on the external surface of the particles.

In the first continuous culture, conducted at $(V_b/V_R) = 0.50$ after switching from batch to continuous culture, the steady-state biomass loading was achieved after culturing for about 12 days. In the continuous cultures performed after change in (V_b/V_R) at a set u, the constant mass of cells grown on the support media was achieved after approximately 6 days of operation. With change in u at a set (V_b/V_R), the new steady-state biomass loading occurred after the culturing for about 4 days.

4. Conclusions

The largest COD decrease, namely, from 54,840 to 2190 mg/l, *i.e.* a 96% COD reduction, was achieved when the reactor was operated at the ratio $(V_b/V_R) = 0.55$, air velocity $u = 0.046$ m/s and $t = 65$ h. Thus, these values of (V_b/V_R), u and t can be considered as the optimal operating parameters for a reactor when used in treatment of high-strength refinery wastewaters.

In the treatment operation conducted in a reactor optimally controlled at $(V_b/V_R) = 0.55$, $u = 0.046$ m/s and $t = 65$ h, the conversions obtained for all phenolic constituents of the wastewater were larger than 95%. The conversions of about 90% were attained for other hydrocarbons.

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Notation

COD   Chemical oxygen demand (kg/m^3)
D   Dilution rate (h^{-1})
t Mean residence time (h)
u Superficial upflow air velocity (m/s)
u_f Minimum fluidization air velocity (m/s)
u_{cr} Critical air velocity (m/s)
V_b Volume of settled bed (m^3)
V_R Reactor volume (m^3)

Subscript

m denotes values giving the greatest COD reduction