Effect of aeration rate on the performance of a novel non woven flat plate bioreactor

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Abstract
The aim of this work was to evaluate the effect of aeration rate (U_g) on the performance of a novel non woven bioreactor. Immobilized microorganisms were in horizontal non woven flat plates with large biomass retention capacity (1.44 g VS/g support, 61.3% active) settled to form a zig-zag path for the air-water flow. Increases in U_g, ranging from 0.080 to 0.129 m/s, resulted in increases on the apparent substrate consumption rate (8.38 to 11.86 mg/L h) and decreases the mixing time (from 13 to 7 min), thereby positively affecting the oxygen (k_La) and external (k_c) mass transfer. Biofilm detachment was less than 1% despite of the shear stress value was 1.12 Pa at the highest airflow rate. The system could treat a superficial organic loading from 13.54 until 100 g phenol /m^2 d with almost 100% phenol removal.

Keywords
Biofilm detachment, mass transfer, nonwoven fibrous support

Abbreviation

APHA	American Public Health Association
BET	Brunauer, Emmett and Teller
CSTR	Continuous stirred-tank reactor
G	Bulk gas
L	Bulk liquid
PET	Poly(ethylene terephthalate)
PFR	Plug flow reactor
PVC	Poly(vinyl chloride)
S	Bulk solid
UNAM	Universidad Nacional Autónoma de México
WEF	Water Environment Federation

Variables

C_L	Dissolved oxygen concentration in the bulk liquid (mg_{O_2}/L)
C^*	Oxygen saturation concentration in the bulk liquid (mg_{O_2}/L)
d_p	Bubble diameter (mm)
D	Diffusion coefficient of phenol in water (m^2/s)
$D_{biofilm}$	Diffusion coefficient phenol in the biofilm (m^2/h)
$\frac{dC}{dt}$	Accumulation the oxygen rate ($mg_{O_2}/L\ s$)
D/uL	Dispersion coefficient
E	Normalized exit age distribution
HRT	Hydraulic residence time (min)
g	Gravitational acceleration (m/s^2)
k_c	External mass transfer coefficient (m/s)
k_La	Volumetric oxygen mass transfer coefficient (min^{-1})

K_s	Half-saturation coefficient *(mg/L)*
LL	Boundary layer thickness (μm)
OUR	Oxygen uptake rate (mg_{O_2}/L min)
RTD	Residence time distribution (s^{-1})
TSS	Total suspended solids (mg/L)
U_g	Aeration rate (m/s)
VS	Volatile solids *(g)*
VSS	Volatile suspended solids (mg/L)

Greek letters

γ	Shear stress (Pa)
θ	Mixing time (min)
μ_s	Kinematic viscosity (m^2/s)
μ_{max}	Maximum *growth rate (h^{-1})*
μ_L	Liquid viscosity (kg m/s)
ρ_L	Liquid density (kg/m^3)

INTRODUCTION

Currently, the international trend on designing municipal and industrial wastewater treatment technology involves the construction of compact and low energy-consuming systems (WEF, 2010). According to scientific research, biofilm systems, have the advantages of compactness and low energy consumption over others approaches (Nicolella et al., 2000; González-Brambila and López-Isunza, 2007). However, biofilm systems also have disadvantages, such as operational problems due to slow mass transfer of contaminants and nutrients to biofilm. Moreover, when biofilm reactors have treated high organic loading, they have presented a limitation in the oxygen transfer.

Najafpour (2007) reported that increasing the air flow allows the improvement of the oxygen transfer coefficient (k_La). Furthermore, González-Brambila and López-Isunza (2007) demonstrated that in a biofilm system (permeable membrane), when the superficial liquid velocity was increased, the external mass transfer was reduced and the diffusion into the biofilm was improved. Nevertheless, these researchers determined that at a superficial velocity above 0.0107 m/s, the biofilm was detached from the support due to shear stress. Moreover, Arrojo et al. (2008) demonstrated that increases in the up flow velocity of the air, which involves shear stress, decreased the specific activity of Anammox granules by approximately 85%.

One of the most active research areas regarding flow aerobic biological systems is the analysis of the biomass supported on nonwoven materials (Jajuee et al., 2007). These materials are porous fabrics that are composed of a random array of polymeric fibers or filaments in a tridimensional structure in several layers. The more frequently used polymers for biofilm systems are polyester (PET), polypropylene, polyethylene and nylon (Hutten, 2007). In addition, these polymeric fibers can used in several arrays allowing the design of novel bioreactors, being sponge cubes one of the most preferred immobilization systems (Lin and Hsien, 2009; Lin, 2010). However, other unexplored configurations, such as zig-zag flow paths can be considered. The principal advantages of these materials are that they

provide a high specific surface area for cell attachment, high and constant surface-to-volume ratio, high mechanical strength, high permeability, low cost and lower mass transfer resistance compared to micro-carrier particles Kilonzo et al., 2010.

The novel details of the present fixed biofilm reactor are a horizontal zig-zag air-water flow through the reactor, a relatively reduced air flow cross section, microorganisms attached to a nonwoven fibrous support and a relatively high separation between fiber dishes. As a consequence of these design improvements, it was expected that the aeration rate, the liquid agitation, the contaminant removal rate and mass transfer coefficients were increased. Thus, the effects of aeration rate (U_g) on hydrodynamics, mass transfer, and biological degradation were evaluated to obtain the necessary data to describe the operation of this type of reactor.

MATERIAL AND METHODS

Experimental device

The non woven flat plate bioreactor consisted of an acrylic column (8.2 L, 65-cm height and 15-cm internal diameter) with eight synthetic fiber-scouring pads as support matrix (15 cm x 0.5 cm thickness, Light Duty Scrubbing Pad 9030-Scotch Brite™, 3M Company). Each tray was separated at 5-cm distance, kept in form by a 14.5-cm outer diameter PVC ring and with an alternated 30% empty space intentionally left between each to induce a zig-zag flow pattern in the bioreactor (Fig. 1).

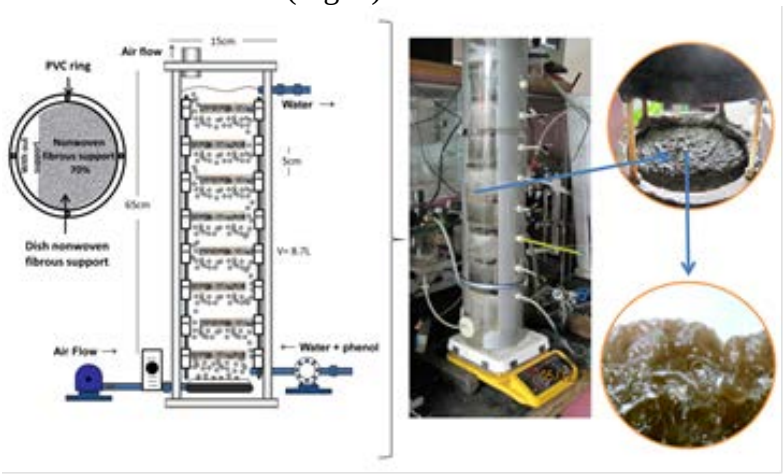

Figure 1 Schematic diagram of the non woven flat plate bioreactor

The fiber specific surface area (0.20 m^2/g) was measured by nitrogen physical adsorption (BET isotherm), and the apparent porosity (94.3%) was determined according to Hutten (Hutten, 2007 and Russell, 2007). A peristaltic pump (Ismatec™ Ecoline CP 78022-40) was used to supply a constant wastewater flow through the multi-tray reactor. Because the pressure drop was relatively low, the liquid flow could be considered constant along the reactor. A micro-pore diffuser provided the free-oil airflow in the reactor bottom. The bioreactor was operated at constant temperature (21°C ± 0.1) and pH was kept at 7.4±0.1.

Inoculation and quantification of the physicochemical parameters

For inoculation, mixed liquor samples were collected from an activated sludge wastewater treatment plant at the UNAM campus. Sludge samples were grown in gradually enriched phenol media, using the methodology of (Mamma et al; 2004) and El-Naas et al., 2009), until a reaching a concentration of 0.8 g/L. Finally, the bioreactor was inoculated with 8.7 L of adapted phenol-degrading microorganisms, considering a biomass concentration of 1450 mg VSS/L. Tests of the total suspended solids (TSS) and the volatile suspended solids (VSS) were performed according to the Standard Methods Guidelines (APHA,WWA and WEF 1999). The phenol concentration was evaluated by UV spectrophotometry at 270 nm according to the methodology Mkandawire et al. (2009), and pH and turbidity were measured using a Orion™ 2-Star Benchtop pH meter (Thermo Orion, USA) and a HI 98703 turbidity meter (HANNA Instruments), respectively. Biofilm cell growth was evaluated through the Monod model (WEF, 2010, Moser, 1988), which allows the kinetic constants to be determined.

Hydrodynamics

The flow regime inside of the reactor was measured by methylene blue dye pulse injection (Levenspiel, 2004). The concentration was measured each minute by spectrophotometry at 665 nm (Varian´s Cary™ 50) until the dye complete disappearance in the reactor. The theoretical hydraulic residence time (HRT) used for this experiment was 2.4 h. The residence time distribution (RTD) and the flow type were calculated using the dye concentration values in the biological reactor. The mixing times (θ) were evaluated at three different Ug values (0.064, 0.080 and 0.096 m/s).

Mass transfer (G/L)

The oxygen transfer from bulk gas (G) into the bioreactor bulk liquid (L) was determined by the dynamic method (Ramalho, 2003, Garcia-Ochoa, Gomez, 2009). Additionally, once the biofilm was formed after a three-week operation period, the volumetric oxygen mass transfer coefficients (k_La) were calculated by Eq. (1).

$$\frac{dC}{dt} = k_L a(C^* - C_L) - OUR \quad (1)$$

Where dC/dt is the accumulation oxygen rate, C_L is the dissolved oxygen concentration in the bulk liquid, C* is the oxygen saturation concentration in the bulk liquid and OUR is the oxygen uptake rate. C_L in the liquid phase batch reactor was measured using a HANNA HI 9143 dissolved oxygen electrode. In the experiments, the aerator was stopped, and the agitation of the system was turned on at a speed of 350 rpm, and the dissolved oxygen was subsequently measured until it decreased its value to 2.5 mg/L. Later the agitation system was stopped, and then the aeration system started. During the aeration time, the dissolved oxygen was measured every three seconds, and the values of k_La were calculated at different U_g values: 0.009, 0.021, 0.050, 0.064, 0.080, 0.096, 0.112 and 0.129 m/s.

Biofilm detachment

The biofilm detachment caused by shear stress is an instantaneous process in comparison with the biofilm growth time; the hydrodynamic (shear stress) characteristic times in a biofilm system are much smaller, by six orders of magnitude, than the characteristic times

for biomass growth (IWA Task Group on Biofilm, 2006). The shear stress was calculated using Eq. (2) (Chisti; 2001), considering three different U_g values (0.080, 0.096 and 0.112 m/s). For each shear stress value, the biofilm detachment was evaluated by TSS in the bulk liquid.

$$\gamma = \left(\frac{\rho_L g U_g}{\mu_L}\right)^{0.5} \quad (2)$$

Mass transfer (L/S) and diffusion in the biofilm

The external mass transfer coefficient (k_c) between the outer bulk liquid (L) and the bulk solid (S) was calculated, after four weeks of continuous reactor operation. The system was operated in batch, considering four U_g different values (0.080, 0.096, 0.112 and 0.129 m/s), with 100 mg phenol/L as a model contaminant, to evaluate the air flow effect in the k_c values. Additionally, the apparent substrate consumption rates at different air flows were calculated. Based on this information, k_c was calculated using the Aquasim model (Reichert, 1998[a,b],1995)

Continuous operation

The reactor was operated during two months at different superficial organic loadings of phenol, ranging from 13 to 100 g/m^2 d. The amounts of dissolved oxygen and phenol removal were measured to evaluate the operation of the reactor.

RESULTS AND DISCUSSION

Acclimatization of microorganisms to phenol

In the first 50 days of the trial, a quick adaptation of microorganisms to the phenol was observed, as confirmed by a close to 100% pollutant removal efficiency. After this acclimation period, the phenol-adapted biomass was used to perform different batch experiments to calculate the kinetic coefficients. Care was taken to prevent dissolved oxygen limitation or substrate inhibition. The results obtained from the Aquasim model for the half-saturation coefficient (K_s), 15.47 mg/L, and the maximum growth rate (μ_{max}), 0.1158 h^{-1}, were consistent with values reported in literature (Bajaj, et al., 2009).

Hydrodynamics

The hydraulic residence time **(**HRT) was measured to determine the system hydrodynamics. The results are shown in the Figs. 2a and 2b. In the absence of airflow, the reactor behavior followed the plug flow reactor (PFR) model. The HRT in the biological reactor was 142 min, which was slightly inferior to the theoretical residence time (144 min). The experimental residence time distribution (RTD) data were fitted to the normalized exit age distribution (E), which was calculated using a dispersion coefficient (D/uL) value of 0.0463 (Levenspiel, 2004), (Fig 2b). The RTD experimental curve is out of phase when superimposed to E, due to the stagnant zones in the system. A value of 9.2% for these zones was obtained using the method of Pérez and Torres, (2008).

With an aeration rate (U_g) value of 0.064 m/s in the system, the reactor behaved as a completely mixed flow; this trend was obtained from the comparison of these results with the continuous stirred-tank reactor (CSTR) mathematical model. Based on an Origin

software simulation, a correlation coefficient of 0.99936 was found between the experimental data and the CSTR model, as shown in Fig. 2c.

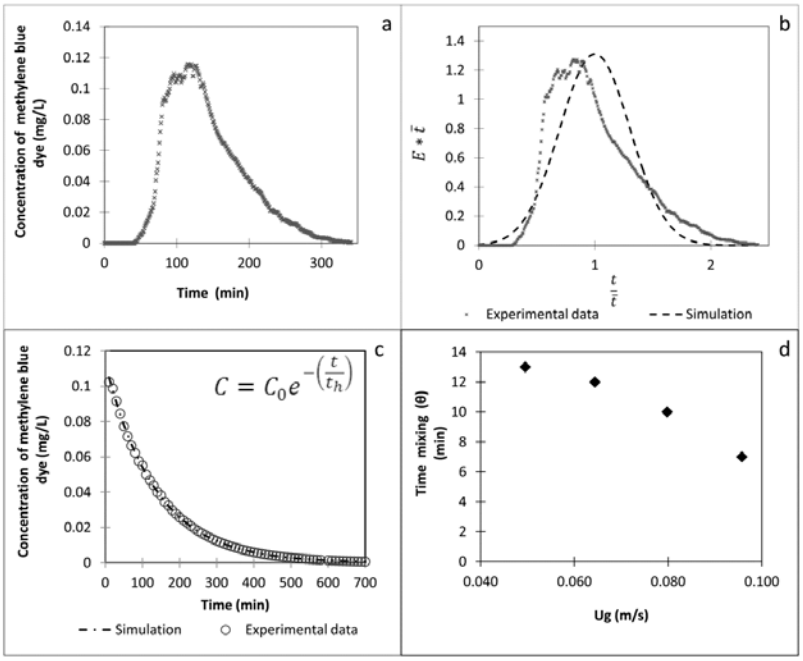

Figure 2 a) Determination of the RTD curve without airflow, b) Modeling the RTD curve without airflow, c) Determination of the RTD curve with airflow, d) Measurement of the mixing time (θ) in the biological reactor at different aeration rate (U_g) values.

Mixing time

A few minutes after the injection, a complete mixing system was achieved, with the resulting concentrations of methylene blue dye shown in Fig. 2c. The point, when the system reaches the complete mixture is known as the mixing time (θ).), which indicates the degree of agitation in the liquid. The experiments were conducted at various air flows, increasing the aeration rate from 0.050 to 0.096 m/s, while θ was reduced from 13 to 7 min, as is shown in Fig. 2d.

Mass transfer (G/L)

Fig. 3a shows the experimental data of the biological reactor dissolved oxygen at different U_g values. When the U_g values are increased above 0.096 m/s, the dissolved oxygen curves were found to be very similar. The calculated coefficients $k_L a$ are shown in Fig. 3b, where increasing U_g was found to also increase $k_L a$ increases. However, when the U_g value was higher than 0.096 m/s, it did not have a significant effect over $k_L a$, which varied only from 1.02 to 1.10 min^{-1}. It can be established that an increase in the air flow above the previously mentioned value does not promote the oxygen transfer into the system. At different U_g values, from 0.009 to 0.129 m/s, the $k_L a$ value increased from 0.12 to 1.10 min^{-1}; this demonstrates that $k_L a$ is function of U_g. This trend was previously reported by Ozbek and Gayik (2009) and Gourich et al. (2008).

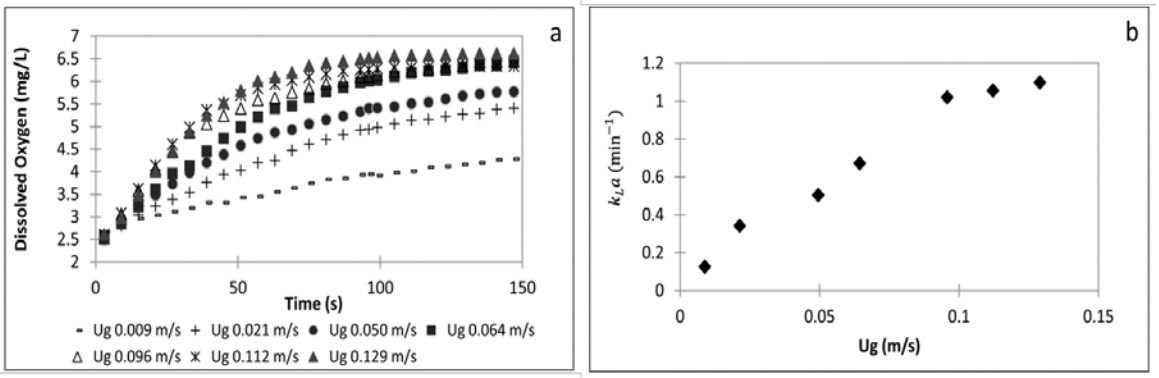

Figure 3 a) Dissolved oxygen measurements at different aeration rate (U_g) values, b) Values of the volumetric oxygen mass transfer coefficients ($k_L a$) considering different aeration rate (U_g) values.

When the condition of oxygen saturation was obtained, the aeration system was stopped to determine the OUR value in the biological reactor. The oxygen concentration in the reactor decreased linearly after 50 min, resulting on a OUR value of 0.0179 mg_{O_2}/L min, demonstrating that the biomass in the biofilm was constant.

Biofilm detachment

When the reactor operated in batch mode, the detached solids in the reactor were measured as TSS at different shear rate values (0.88, 0.97 and 1.05 Pa). The VSS in the reactor never exceeded 20 mg/L (Fig. 4a) so that, all the biomass can be considered to be attached to the nonwoven fibrous support, and a minimal fraction was suspended in the reactor. The detached solids were less than 1% of the total biofilm in the support despite the shear stress value of 1.12 Pa at the highest air flow rate; this is very important because the support could immobilize a large amount of active microorganisms (1.44 g VS/$g_{support}$) to degrade the phenol. The total active biomass in the nonwoven fibrous support was 61.3% according to the catalase activity test (Guwy, 1999).

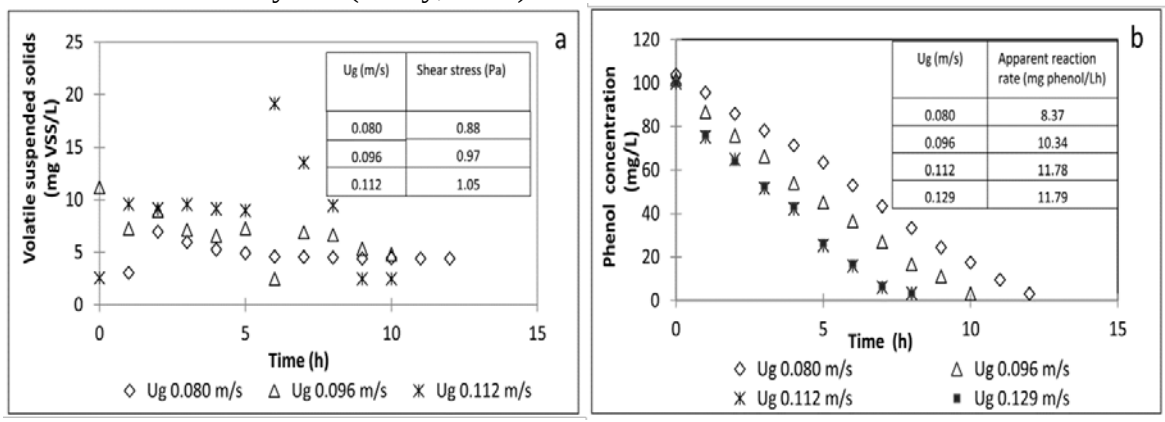

Figure 4 a) Biofilm detachment in the reactor at different aeration rate (U_g) values, b) Apparent reaction rate at different aeration rate (U_g) values.

The nonwoven fibrous support furthermore immobilized biofilm similar to the traditional supports; in addition, it protected the cells of shear stress from the aeration rates

Operation in the batch biofilm reactor to evaluate the mass transfer L/S

Several experiments were performed at different airflow conditions in the bioreactor. Fig. 4b shows the phenol concentration at different U_g values, ranging from 0.080 to 0.129 m/s. In all the experiments, the initial phenol concentrations in the reactor were 100 mg/L and the concentrations were measured until the total disappearance of the phenol. Fig. 4b shows that reaction time decreased from 12.1 to 8.34 h when the U_g values increased. González-Brambila and López-Isunza (2008) demonstrated that if the Reynolds number increases in a biofilm reactor, then the boundary layer thickness decreases; therefore, the reaction rates in the reactor increases, mainly due to the increase in k_c increases. Recalling that $k_c = D/LL$, in this case, D is the diffusion coefficient of phenol in water and LL is the boundary layer thickness, as a result of the decrease in the thickness, the mass transfer and the reaction rate also increased. The reaction rate into the reactor increased 50% (from 8.37 to 11.79 mg_{phenol}/L h) when the reactor was operated at the highest aeration rate (0.1290 m/s). Note that the measurements of dissolved oxygen in the reactor were similar and always greater than 7.4 mg/L, indicating that reactions were not limited by the concentration of dissolved oxygen in all the experiments.

Mass transfer (L/S), modeling of the batch biofilm reactor using the Aquasim model

For modeling the system using the Aquasim model, it was considered that: 1) the values of K_s and μ_{max} were obtained in a suspension system; 2) the hydrodynamics of the reactor was considered as a biofilm system with L completely mixed; 3) the diffusion coefficient of contaminant in the biofilm ($D_{biofilm}$) was estimated from the degradation profiles that indicated the phenol concentration in Fig. (4b), when the reactor operated in a batch mode. With the U_g value of 0.1290 m/s, the system did not register any significant change in the reaction rate in comparison with the system with the U_g value of 0.112 m/s, and then it was supposed that the boundary layer thickness was at least 0.1 μm. The value of the $D_{biofilm}$ was 2.36 × 10^{-8} m^2/h, which was used to simulate other degradation profiles, with an external mass transfer opposed to the transport of the contaminants.

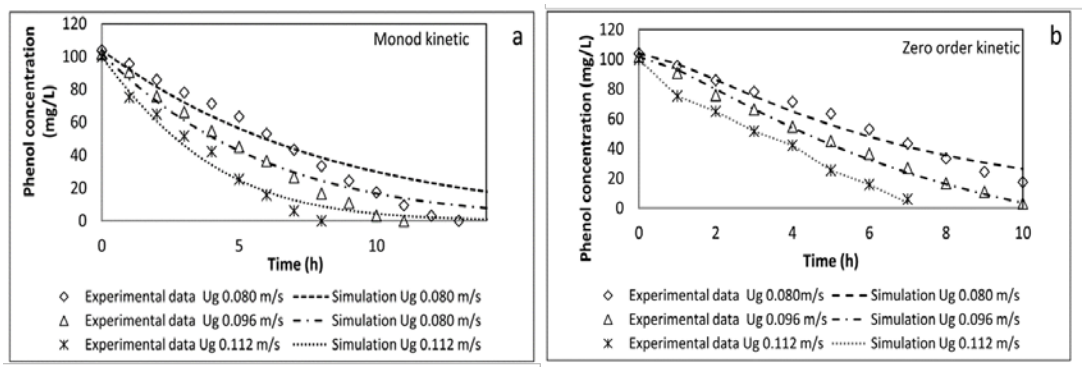

Figure 5 a) Modeling at different aeration rate (U_g) values with the Monod kinetic, b) Modeling at different aeration rate (U_g) values with the zero-order kinetic.

The simulation was performed based on the Aquasim model using two different kinetic types: Monod and zero order. As shown in Fig. (5a), the Monod kinetic does not

appropriately represent the experimental data: the mean relative deviation was 82%, which was calculated according to the method of Kablan et al. (2008). As a result, the diminution of the phenol concentration at the time was similar to a straight line, which was simulated using a zero order kinetic, as shown in Fig. (5b), where the mean relative deviation was 6.2%; thus, the second simulation is more representative for the experimental data.
Finally, the external mass transfer coefficients (k_c) changed from 3.67E^{-05} to 2.68E^{-04} m/s.

Continuous operation biofilm reactor

The biofilm type of system can treat high superficial organic loading of contaminants in continuous operation. In this experimental stage, organic loading of phenol (from 13 to 100 g/m^2 d) were tested, and the system behavior was stable. Fig. (6) presents the contaminant removal and the dissolved oxygen concentration during the experiments. The removal of the contaminant was greater than 99%, and the reactor behavior was stable, with a superficial organic loading of phenol smaller than 50 g/m^2 d. However, at higher organic loading (from 50 to 100 g/m^2 d), there was an oxygen deficit in the reactor.

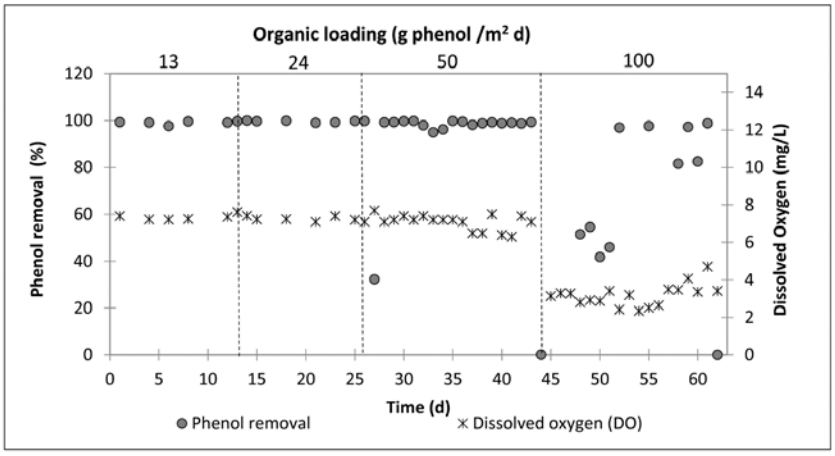

Figure 6 Dissolved oxygen and phenol removal in a continuous reactor.

Conclusions

The innovative design of the multi-tray bioreactor studied increased the aeration rate by reducing θ. With air flow, the reactor is CSTR, and the values above 0.096 m/s of U_g did not exert a significant effect on $k_L a$. The amount of solids detachment was less than 1% of the total biofilm, and the support immobilized microorganisms to degrade phenol (1.44 g $_{biomass}$/g$_{support}$), where 61.3% were active. When the aeration rate increased (0.080 to 0.129 m/s), the reaction rate increased by 50% as a consequence of the values of $k_c$values increasing. Therefore, the Aquasim model better represents the process because it takes parameters, such as diffusion-reaction and mass transfer, into account.

Acknowledgments

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