# A Study on Using Fluidized Bed Reactor for Treating Sanitary Sewage

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***Abstract—Fluidized bed reactor (FBR) is an attached growth system used mainly for biological treatment of industrial wastewater of high organic content. These wastewaters are usually resulted from refineries and milk, starch, and olive oil industries. The objective of this study is to investigate the use of fluidized bed reactor for treating sanitary sewage. The study was accomplished using a pilot plant of the FBR. The pilot plant was constructed and installed in Hamdan Sewage Treatment Plant in Basrah governorate. That was to maintain continuous source of settled sewage which is the influent to the FBR. The period of plant operation was nine weeks. During, this period, the plant was operated at three phases of different conditions (up flow velocity and recirculation ratio). To study the performance of FBR, the main measured parameters were; BOD, DO, VSS, pH, and temperature. The most important conclusions of this study are; (1) the maximum efficiency of BOD removal is 78.6% which was obtained for hydraulic retention time (HRT) of 24min and upflow velocity of 1.59m/min, (2) the effluent BOD values during phases-1 and 2 of plant operation match that of stabilization ponds and trickling filters and during phase-3 matches that activated sludge process, (3) during all operation phases, the values of effluent pH are within the limits specified in national standards of secondary effluents, (4) as F/M increases, the efficiency of BOD removal decreases and the maximum efficiency of BOD removal (78.6%) was obtained at F/M ratio equals 23.47 day$^{-1}$, and (5) the HRT of fluidized bed reactor is on order of minutes, while, the values of HRT of activated sludge systems and stabilization ponds are on order of hours and days, respectively.***

*Index Terms*— **Biological uidized beds, sanitary sewage, pilot plant**

## I. INTRODUCTION

The function of sewage treatment plant is to speed up the natural process by which wastewater cleans itself. This process does that by reducing the concentrations of solids, organic matter, nutrients, pathogens, and other pollutants in sewage. The heart of sewage treatment plant is the biological treatment system. There are three basic categories of biological treatment systems; aerobic, anaerobic and anoxic. Aerobic biological treatment involves contacting wastewater with microbes and oxygen in a reactor to optimize the growth and efficiency of the biomass. The microorganisms act to catalyze the oxidation of biodegradable organics and other contaminants such as ammonia, generating innocuous by-products such as carbon dioxide, water, and excess biomass (sludge). Anaerobic (without oxygen) and anoxic (oxygen deficient) treatments are similar to aerobic treatment, but use microorganisms that do not require the addition of oxygen. These micro-organisms use the compounds other than oxygen to catalyze the oxidation of biodegradable organics and other contaminants, resulting innocuous by-products. Regardless of the type of system selected, one of the keys to effective biological treatment is to develop and maintain an acclimated, healthy biomass, sufficient in quantity to handle maximum flows and the organic loads to be treated.

Maintaining the required population of “workers” in a bioreactor is accomplished in one of two general ways [1]:

- Attached growth processes which include; trickling filters, rotating biological contactors (RBC), and fluidized bed reactors (FBR).
- Suspended growth processes such as activated sludge systems, oxidation ditches, and stabilization ponds.

In fluidized bed reactors (FBRs), the microorganisms grow attached to small carrier particles, such as sand grains, which are remain suspended in the fluid and maintained in a fluidized state by the drag force associated with upward flow velocity of waste-water undergoing treatment. The effluent from such bioreactor generally contains little suspended biomass but particles must be continually removed and cleaned to maintain constant mass of microorganisms in the system. The cleaned particles are continually returned to bioreactor while the wasted biomass is sent to an appropriate treatment process. Recirculation of effluent around the bioreactor is usually needed to achieve the required fluidization velocity and thus the system tends to behave as if it were completely mixed [2].

The application of FBRs as a biological wastewater treatment system was started in 1970’s. This can be credited to the efforts of researchers at Manhattan College in New York, at EPA Municipal Environmental Research Laboratory in Cincinnati, OH, and at the Water Research Center in Medmenham, England [3]. The previous studies showed the importance of fluidized bed reactor in removing organic materials, but, most of them were conducted on industrial wastewaters such as; olive mill wastewater [4], red wine distillery wastewater [5], [6], [7], phenolic wastewater [8], [9], [10], milk wastewater [11], wastewater containing volatile organic compounds [12], refinery wastewater [13], [14], starch industry wastewater [15] ,[16], [17], dairy wastewater [18], sago industry wastewater [19], textile wastewater [20], pulp and paper-mill effluent.[21], palm oil mill effluent [22], and coking wastewater [23]. In treating of sanitary sewage, the application of FBR is mainly as denitrification system such as the work of Naik and Setty [24] who introduced an experimental study on biological denitrification carried out in a fluidized bed bioreactor using synthetic wastewater. All the previous studies highlighted the main advantage of using FBR for biological treatment of wastewater which is the low hydraulic retention time and small size of equipment required. This advantage raises the necessity for performing a study on the use of FBR for sanitary sewage treatment using a pilot plant and real sewage.

The objectives of this study include; (1) examining the feasibility of using aerobic FBR for treating sanitary sewage via the use of pilot plant, (2) investigating the performance of the pilot plant under the effect of different operating conditions, and (3) studying the factors affecting the performance of FBR such as food/ microorganisms ratio and hydraulic retention time.

## II. Materials and Methods

### *A. Pilot Plant Location*

The pilot plant of FBR was constructed and installed in Hamdan Sewage Treatment Plant (HSTP), which is the main sewage treatment plant in Basrah governorate. The plant was installed in HSTP because the last treats sanitary sewage of Basrsh city and the proposed study concerning the feasibility of using FBR for treating sanitary sewage. Also, because FBR is continuous flow system, thus, it is necessary to maintain sewage flow during plant operation which needs continuous source of real sewage.

The pilot plant was installed near the distribution/collection chamber of primary settling tanks which distributes/collects the raw/settled sewage to/ from primary settling tanks. The feed to the pilot plant was drawn from the settled sewage channel.

### *B. Pilot Plant Description*

A schematic diagram and photo of FBR pilot plant are presented in Figs.1 and 2. The plant is composed of the following components:

1- PVC column of 0.2 m diameter and 3.75 m total height. It composes of three sections; inlet, reactor, and outlet. *The inlet section* is of conical bottom and 0.5 m height. It is separated from the reactor section using perforated plate and gravel layer to uniformly distribute the feed into the reactor. The perforations of plate are of 0.4 mm diameter and the thickness of gravel layer is 0.2 m. *The reactor section* is of 3.75 m height. It is supplied with five sampling cocks at spacing of 0.65 m center to center. It contains sand with an effective size of 0.6 mm as supporting growth media. This effective size is commonly used in fluidized bed reactors because it offers good stability of bed height against changes in superficial velocity [1]. The thickness of sand layer (before bed operation) is of 1.8 m. *The outlet section* is of 0.5 m diameter and 0.5 m height. It is made of steel and fixed at the top of the reactor. In this section, a rotary mixer of 0.78 kW power is installed. The function of mixer is separating the sand from biomass by shearing the biofilm covering the support particles.

2- Final settling tank which is made of steel and has a surface area of 1.25×1.25 $m^2$ and a side water depth of 0.8 m. The tank is supplied with hopper at its bottom for sludge collection.

3- Effluent collection tank of 500 liter capacity.

4- Air blower of 1500 l/min. capacity, 1.7 bar pressure and 2.2 kW power. It is used for supplying the air to the reactor through air pipe of 37.5 mm diameter ending with ceramic air diffuser, of 100 mm diameter fixed in the inlet section of the reactor.

5- Air compressor, of 250 l/min. capacity and 2.2 kW power with air tank of 200 liter capacity and 0.8 mPa pressure. It is used as additional source for air supply.

6- Recycle pump of 100 l/min. capacity, 30 m head, and 1.1 kW power. It is used for circulating the activated sludge from the bottom of the final settling tank to the inlet section.

7- Submersible feed pump for continuously pumping of sewage from the settling channel of HSTP to the inlet section. It has a maximum capacity of 90 l/min, power of 2.5 kW and 25 m head.

8- Feed flow meter, with a maximum reading of 21 l/min.

9- Recycle flow meter, with a maximum reading of 130 l/min.

10- EXTCH digital air flow meter, model 407119A, for controlling the quantity of air supply.

11- Steel piping works of 25 mm diameter including influent, effluent and recycle pipes. PVC piping work of 37.5 mm diameter used to transport and inject the air inside the reactor. The piping works include number of gate valves for controlling the flowrate.

12- Generator of 37 kV, It is used for operating the pilot plant during the periods of main electricity shut off.

13- Steel frame structure for supporting the reactor column. It has bottom base of 2×2 $m^2$ and top base of 0.8×0.8 $m^2$. The height of the frame is 5.5m.

### *C. Sizing of Pilot Plant*

The cross sectional area of the reactor was obtained by equalizing the minimum fluidization velocity to total hydraulic loading using the following equation [3];

$$A_c = Q\left(1+r\right)\big/V_m \qquad \ldots(1)$$

where; $A_c$ is the cross section area of the reactor ( $L^2$), Q is the feed flow rate ($L^3/T$), r is the recirculation ratio and $V_m$ is the minimum fluidization velocity (L/T).

$V_m$ is required for fluidization to occur. It is dependent on fluid mechanic principles and can be obtained as [25];

$$V_m = V_t \varepsilon^n \qquad \ldots (2)$$

where; $V_t$ is the terminal settling velocity of a single solid particle (L/T), $\epsilon$ is the bed Porosity and n is a correction factor.

The terminal settling velocity for a solid particle moving in a fluid under steady state conditions can be determined using Newton's low for settling velocity [26];

$$\left[4gd_s^2\left(s-1\right)\big/3C_D\right]^{0.5} \qquad \ldots(3)$$

where; $d_s$ is the particle diameter (L), s is the sp. gr. of particle, g is gravitational acceleration ($L/T^2$) and $C_D$ is drag coefficient which can be well represented by [25];

$$C_D = 24\left(1+3R/16\right)^{0.5}\big/R \qquad \ldots(4)$$

where *R is* Reynolds number defined by;

$$R = V_t \rho d_s / \mu \qquad \ldots(5)$$

and ρ and μ are water density ($M/L^3$) and absolute viscosity ($ML^{-1}T^{-1}$), respectively.

The cross sectional area of the reactor (which is 0.032 $m^2$) was obtained based on adopting the above equations with specifying $\epsilon$=0.6, n=3, $d_s$= 0.6mm, $\rho_s$= 2643kg/$m^3$, ρ= 1000 kg/$m^3$, μ= 1.002×$10^{-3}$ kg/m.sec., Q=0.01 $m^3$/min and r = 3.

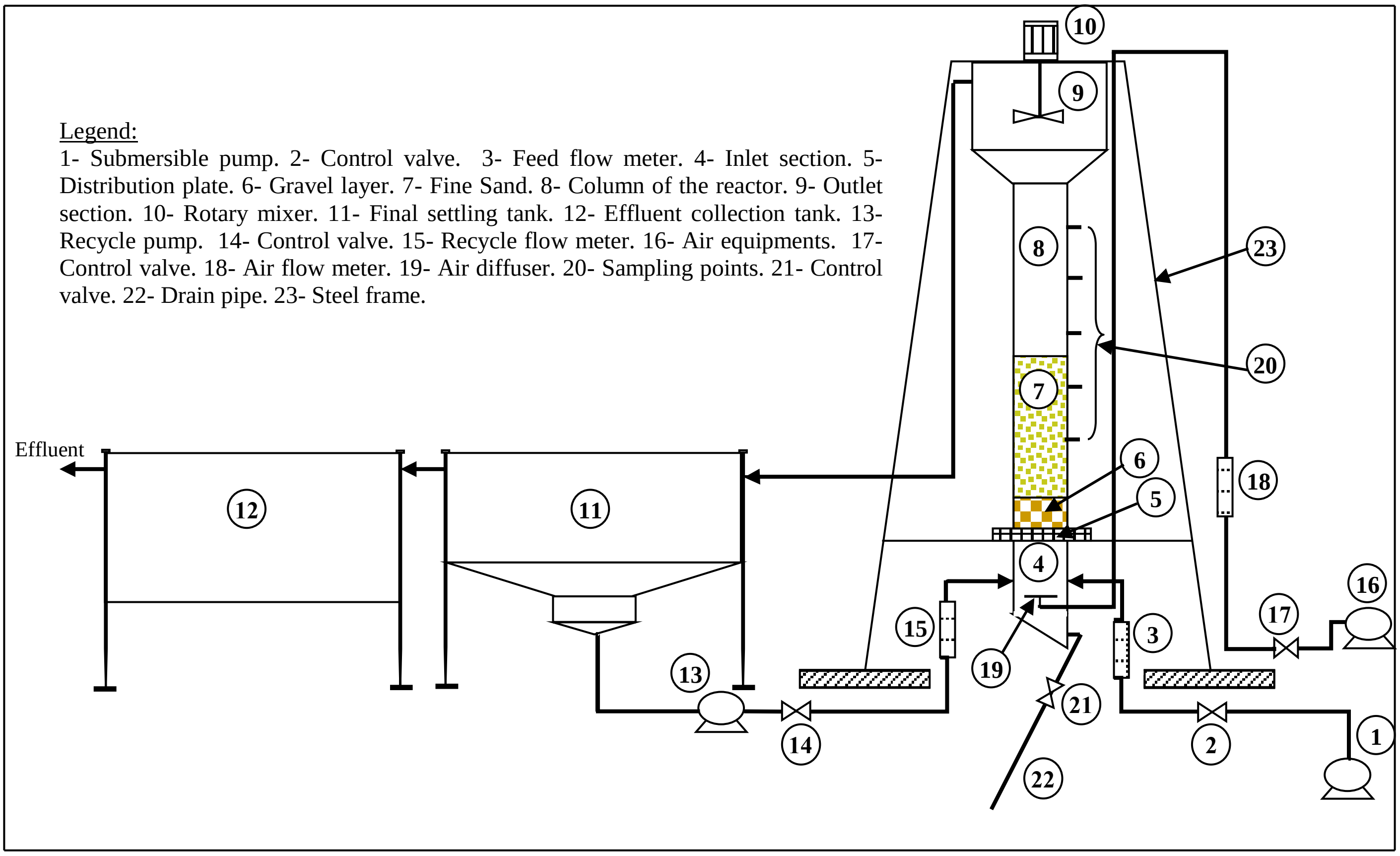



Fig. 1 Schematic flowsheet of FBR pilot plant

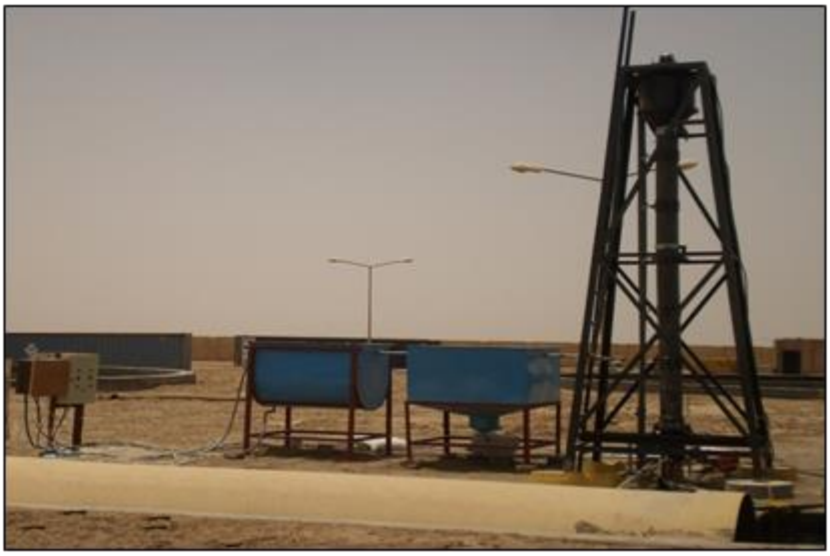


Fig. 2. Photo of Pilot plant of FBR

The height of the reactor was estimated based on the thickness of support media and the expansion ratio of media during fluidization. The thickness of media is usually between 1.5 to 2.0 m [27] and taken to be 1.8 m and the expansion ratio is 100% [28]. Thus, a reactor height of 3.75m is dependent.

*D. Startup and Operation of Pilot Plant*

After the complete installation of the FBR pilot plant, the column of FBR was filled with 1.8 m of sand, which has an effective size of 0.6 mm. Then, the settled sewage obtained from the effluent of the primary settling tanks in HSTP was pumped into the inlet section of the reactor by the submersible pump. The feeding sewage in this section was mixed with a ratio of recycle flow from the final settling tank, and a variable quantity of air was injected in the inlet section and spreads by the air diffuser. The mixed solution passes upward through the distribution plate, via gravel and sand bed, at a minimum fluidization velocity enough to expand the particles of media and make turbulent moving of sand particles.

The operation of FBR pilot plant started on 22/9/2011 and ended on 24/11/2011. Therefore, the pilot plant was operated, continuously, for a period of two months. During this period, the fieldwork was divided into three phases and the conditions of plant operation during each phase are shown in Table I.

TABLE I
OPERATION CONDITIONS OF FBR PILOT PLANT.

| Operation phase | Period | | Feed Flowrate (l/min) | Recycle Flowrate (l/min) | Air Flowrate (l/min) |
|---|---|---|---|---|---|
| | From | To | | | |
| 1 | 22-Sep. | 18-Oct. | 10 | 30 | Variable |
| 2 | 23-Oct. | 2-Nov. | 15 | 45 | Variable |
| 3 | 4-Nov. | 24-Nov. | 5 | 45 | Variable |

During each phase, the plant was operated continuously and during the periods of power shut down, the generator was operated to maintain the plant work. Each phase was ended when the effluent of plant reached a steady state condition at which the BOD of plant effluent does not vary with operation time (the percent of BOD removal remains constant).

After a period of pilot plant operation startup, the sand granules, used as a nucleus for the biological growth, were covered with layers of microorganisms and formed bio-particles. The thickness of the bio-particles decreases with the increase of biofilm thickness. This causes the upward movement of bio-particles with the effluent flow. The increasing of biofilm thickness was controlled by shearing the growth of biological mass from bio-particles using the rotary mixer installed at the outlet section. This allows the return of bio-particles into the reactor and maintains the effective biomass inside the reactor, while, the excess biomass escaping with the effluent flow.

### *E. Laboratory Work*

During the period of pilot plant operation, samples of influent, effluent and from different heights of the reactor were collected and analyzed in laboratory of HSTP. The measured parameters include biochemical oxygen demand (BOD), total suspended solids (TSS), volatile suspended solids (VSS), dissolved oxygen (DO), pH, total dissolved solids (TDS), and temperature. The instruments used to measure these parameters are shown in Table II.

The BOD was measured in samples collected from the influent (sampling point No.1) and effluent (sampling point No.7) of the pilot plant. While, the remaining parameters were measured for samples collected from five locations along the reactor (sampling points No.2, 3, 4, 5, and 6). The locations of these sampling points are at distances of 85, 150, 215, 280, and 345 cm, respectively, from the inlet section of the pilot plant.

TABLE II
LABORATORY WORK INSTRUMENTS

| Parameter | Used Instrument/ Adopted Method |
|---|---|
| BOD | Oxi Top meter |
| TSS | Gravitational method according to standard methods, section 2540D.[29] |
| VSS | Gravitational method according to standard methods, 2540 E. [29] |
| DO and Temp. | DO meter type Oxi 330i |
| Ph | pH meter type HANNA |
| TDS | TDS meter type HANNA |

## III. RESULTS AND DISCUSSION

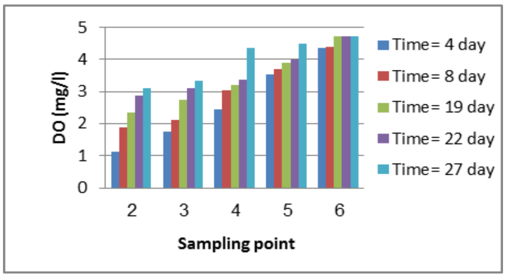
The results of FBR pilot plant operation during the three phases are presented along with evaluating the performance of the plant as a biological treatment system.

### *A. Dissolved Oxygen Distribution*

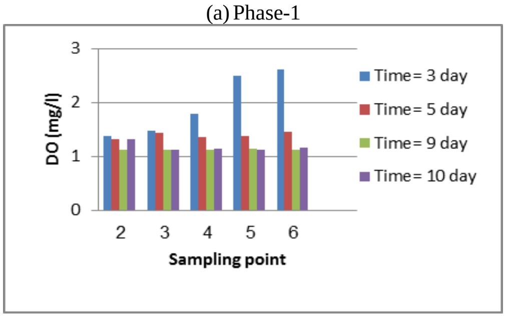
To insure aerobic oxidation of organic matter, it is necessary to measure the concentration of residual dissolved oxygen (DO), where, the residual DO is the concentration difference between oxygen supply and demand. The measured values of DO during the three phases of FBR operation were plotted verses the location along the reactor and at different operation times. The obtained results are shown in Figs. 3- a, b, and c for phases-1, 2, and 3, respectively.

Fig. 3-a shows that for specific operation time, the concentration of DO has a minimum value at sampling point No.2 (head of reactor) and a maximum value at sampling point No.6 (end of reactor). This result is reasonable since the BOD near the influent is of maximum value and as sewage flows along the reactor, it reduces due to biological oxidation process (i.e., the oxygen demand at reactor head is greater than that near the end).

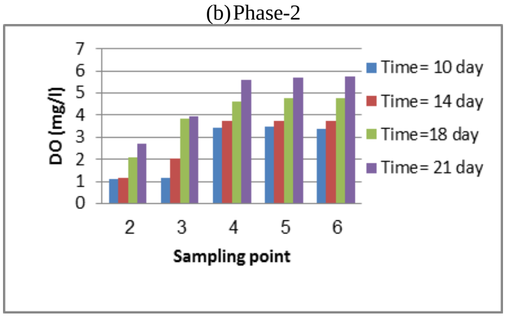
The above result was noticed, also, during operation phases- 2 and 3, Figs. 3-b and c. If the levels of DO measured during the three phases are compared, it can be shown that the minimum DO was during phase-2. Where the values of DO during this phase vary over the range (1.12 – 2.88 mg/l), While, the DO value during phase-1 reached a value of 4.73 mg/l. Also, it can be noticed in Figs. 3- a and b, that the level of DO increases with the increase of operation time. However, in phase-2, as the operation time increases the level of DO decreases.




(a) Phase-1




(b) Phase-2




(c) Phase-3

Fig. 3 Residual DO concentration verses sampling point location along the FBR during the three operation phases

To explain the above result, it is referred to the operation parameters of the pilot plant, Table III. In this table, the up flow velocity was obtained by dividing the total sewage flow, including the recirculation, by the cross sectional area of the reactor. The recirculation ratio is the ratio of recycle flow to the influent flow, while the temperature and BOD ranges are the measured values. Table III shows that the variation of the influent BOD during the three phases is insignificant and the up flow velocity of liquid during phase-2 is greater than that during phase-1. Since, as the up flow velocity increases, the contact time between liquid and air decreases, the oxygen transfer from gas phase to liquid phase decreases [30]. That may cause the lower level of DO during phase-2, although, the oxygen demand is the same as that of phase-1.

TABLE III
OPERATION PARAMETERS OF PILOT PLANT

| Phase | Upflow Velocity (m/min) | Recirculation Ratio | Temp. of Sewage (°C) | Influent BOD (mg/l) |
|---|---|---|---|---|
| 1 | 1.27 | 3 | 28.1-33.5 | 130-160 |
| 2 | 1.91 | 3 | 24.9-27.8 | 135-155 |
| 3 | 1.59 | 9 | 17.5-23.2 | 140-160 |

During phase-3, the maximum value of DO is 5.75 mg/l which is greater than that of phase-1 which is 4.73 mg/l, although, the up flow velocity of liquid in phase-3 is greater than that of phase-1. The reason behind that can be explained as; when the temperature decreases, oxygen saturation concentration increases, as shown in the following equation [31];

$$C_{sat} = 14.61996 - 0.40420T + 0.00842T^2 - 0.00009T^3 \qquad \ldots(6)$$

$C_{sat}$ is oxygen saturation concentration in mg/l at liquid temperature T in ℃. And when the oxygen saturation concentration increases, the rate of oxygen transfer from the gas phase to the liquid phase increases as in the following equation [cited in 32];

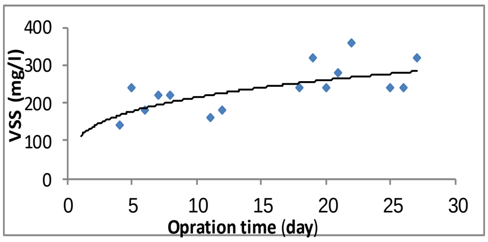
$$dC/dT = k_{la}\left(C_{sat} - C\right) \qquad \ldots(7)$$

where dC/dt is the rate of oxygen transfer from the gas phase to liquid phase in mg/(l.h), C is DO concentration in mg/l, and $k_{la}$ is oxygen transfer coefficient (1/h). Thus, the decrease in temperature increases the rate of oxygen transfer. From Table II, it can be noted that the temperature of sewage during phase-3 is lower than that during phase-1 and subsequently the oxygen transfer rate during this phase has a maximum value. In addition, the recirculation ratio of phase-3 is 9, i.e., the influent is diluted with water of high DO content and this can lead to the increase of DO during this phase.

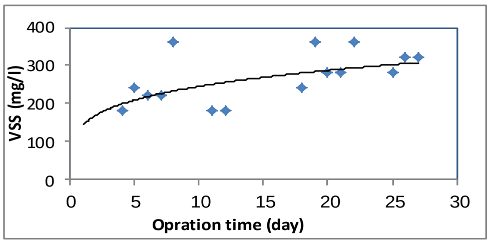
Generally, the measured values of DO during all the operation phases did not decrease below 1 mg/l and subsequently the oxygen supply is adequate to maintain aerobic condition and growth of aerobic bacteria.

### *B. Temporal Variation of Active Mass*

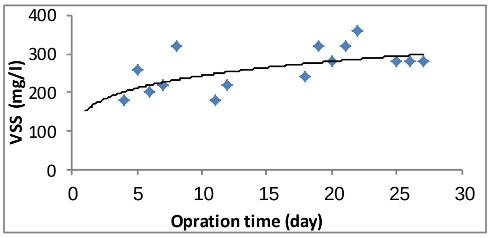
Since in this study FBR is used as a biological treatment system, then, the growth of active mass (bacteria) reflects the success of the system. The active mass is usually measured in terms of volatile suspended solids (VSS) concentration. The measured values of VSS during the three phases of plant operation were plotted verses the operation time as shown in Figs. 4, 5, and 6. In each of these figures, the temporal variations of VSS are plotted for the sampling points.



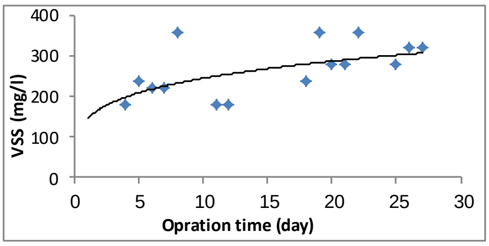

(a) Sampling point No.2



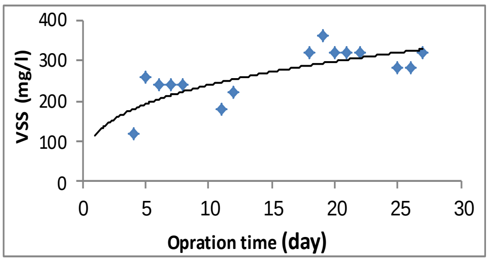

(b) Sampling point No.3




(c) Sampling point No.4




(d) Sampling point No.5

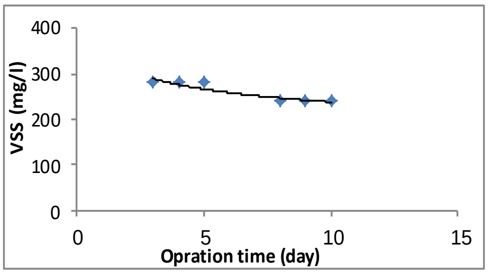



(e) Sampling point No.6

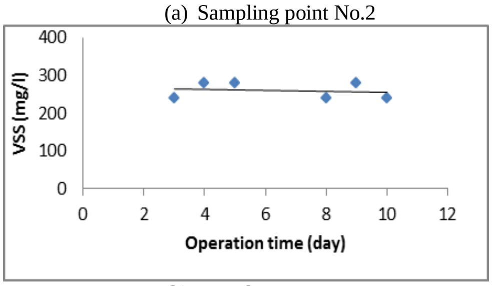
Fig. 4. VSS concentration verses operation time at different locations of FBR during phase-1

In Figs. 4, it can be shown that the VSS increases with the increase of operation time. At the end of operation time, the concentrations of VSS vary over the range (280-360) mg/l. Also, it can be noticed that the growth trend at the locations of the five sampling points is nearly similar.

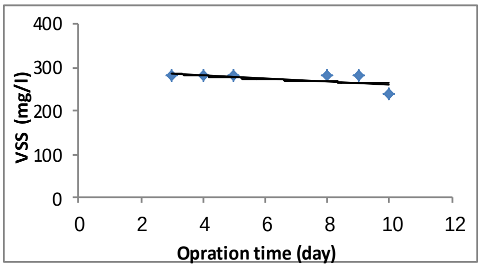
During phase-2, Fig. 5-a shows that VSS concentration decreases with operation time. That may refer to either food or oxygen shortage. In addition, it may refer to the shearing of active mass, accumulated on the sand particles, due to drag forces caused by up flow velocity. Since the BOD and DO are sufficient as shown in the previous section, then, the high up flow velocity during this phase (see Table III) may cause VSS decrease. This result is clear, also, in the other locations of the reactor as shown in Figs. 5-b through e.



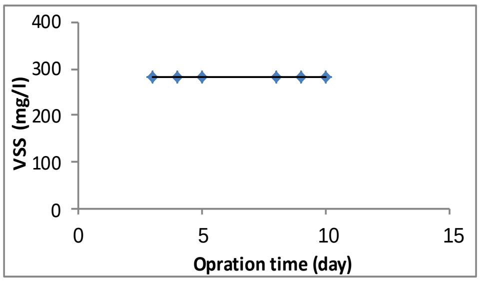

(a) Sampling point No.2



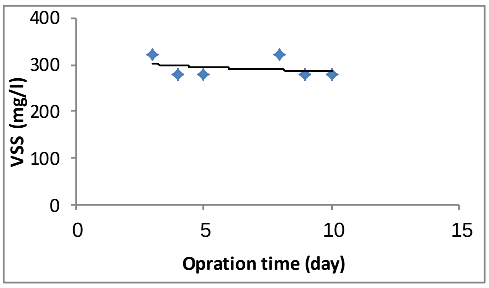

(b) Sampling point No.3




(c) Sampling point No.4

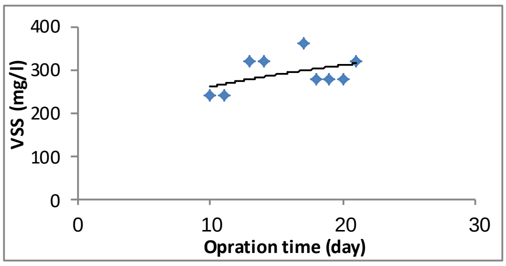



(d) Sampling point No.5

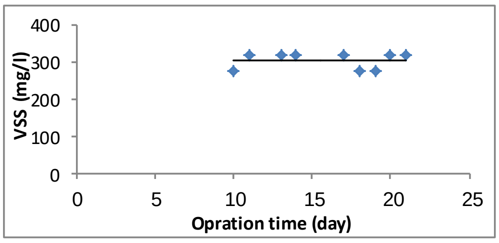



(e) Sampling point No.6

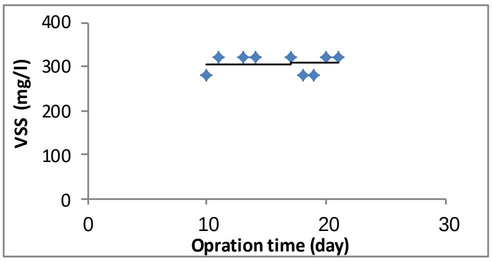
Fig. 5. VSS concentration verses operation time at different locations of FBR during phase-2



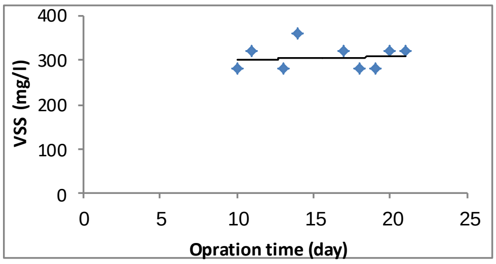

(a) Sampling point No.2



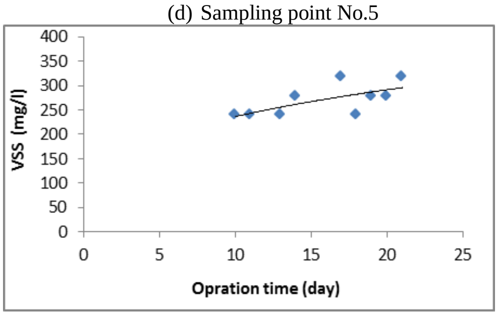

(b) Sampling point No.3




(c) Sampling point No.4




(d) Sampling point No.5




(e) Sampling point No.6

Fig. 6. VSS concentration verses operation time at different locations of FBR during phase-3

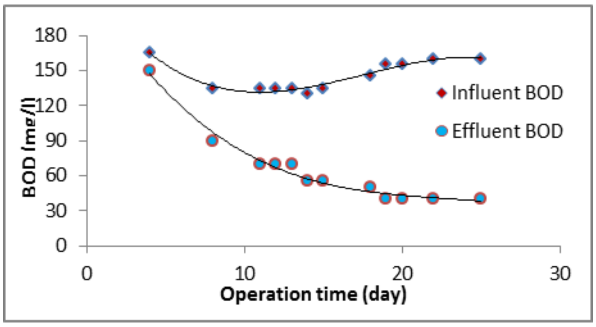
In Figs. 6-a through e, it can be noticed that VSS concentration increases with operation time but at a lower rate as compared with that of phase-1 and this can be referred to low liquid temperature during this phase (Table III) and it is well known that the rate of bacteria growth decreases with temperature decrease [27]. Also, it can be noticed that VSS concentration is nearly constant at locations 5 and 6. The measured values of VSS concentration during phase-3 vary over the range (280-320) mg/l.

## *C. Temporal Variation of Influent and Effluent BOD*

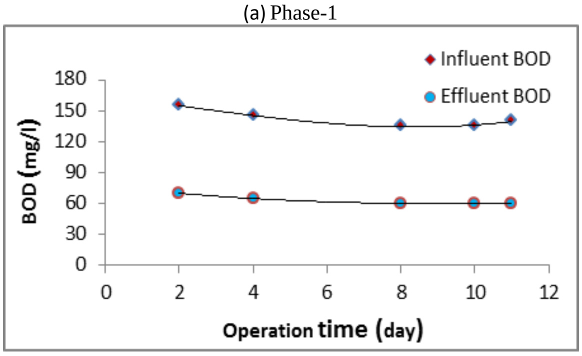
The variation of influent and effluent BOD with operation time during phase-1 of plant operation is shown in Fig. 7-a. This figure shows that, as operation time increases, the effluent BOD decreases until it reaches a steady value of 40 mg/l. Also, it shows that after four days of plant operation, the difference between influent and effluent BOD values (which represents the BOD removed by biological oxidation) is 15 mg/l and this difference increases with the increase in operation time. That is because as operation time increases, the active mass increases (as explained in above section) which enhance the plant performance.

After application of phase-2 operation conditions, the temporal variation of influent and effluent BOD is different from that of phase-1, as shown in Fig. 7-b. In this figure, it can be noted that the minimum value of effluent BOD is 60 mg/l. The temporal variation of influent and effluent BOD during phase-3 is shown in Fig. 7-c, where, it can be shown that at the end of plant operation period, the effluent BOD was 30mg/l.

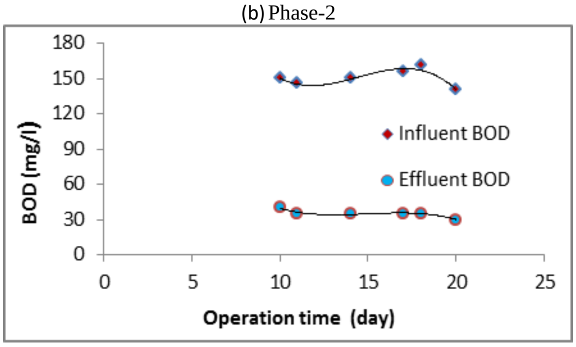
From above it can be noted that the worst operation conditions are those of phase-2, which produced the maximum value of effluent BOD. That can be referred to the high up flow velocity that lowers the concentrations of active mass and dissolved oxygen.




(a) Phase-1




(b) Phase-2




(c) Phase-3

Fig. 7. Variation of influent and effluent BOD with operation time during the three phases of plant operation

### D. *Evaluation of FBR Performance*

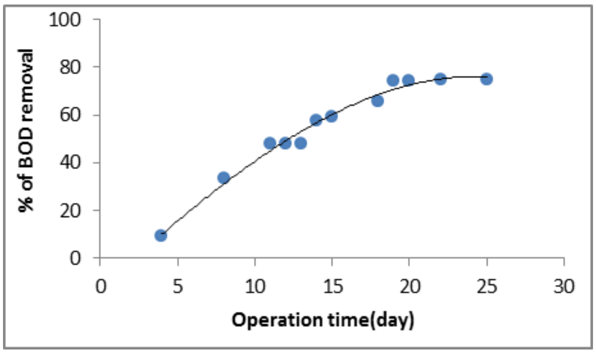
Two bases have been followed to evaluate the performance of FBR as a biological treatment system. The first is based on calculating the percent of BOD removal and the second is based on comparing the quality of effluent with national standards of secondary effluents. Where the percent of BOD removal was calculated as;

$$\%\ BOD_{removal} = \frac{BOD_{influent} - BOD_{effluent}}{BOD_{influent}} \times 100 \qquad \ldots(8)$$

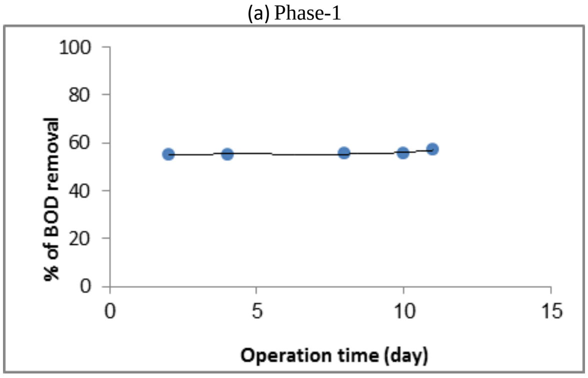
The obtained values of BOD removal percent in their relation with operation time are shown in Figs. 8-a, b, and c for phases-1, 2, and 3, respectively. In Fig. 8-a, it can be shown that the percent of BOD removal during phase-1 increases with operation time with a maximum value of 75%.

During phase-2, the maximum percent of BOD removal is lower than that of phase-1 and equals to 57.1% (Fig. 8-b). While, it reached a value of 78.6% (Fig. 8-c) during phase-3. The national standards of secondary effluents are given in Table IV, while, the obtained quality parameters of FBR pilot plant effluent including the values of BOD and pH are given in Table V. If the corresponding values in Tables IV and V are compared, the followings are noticed;

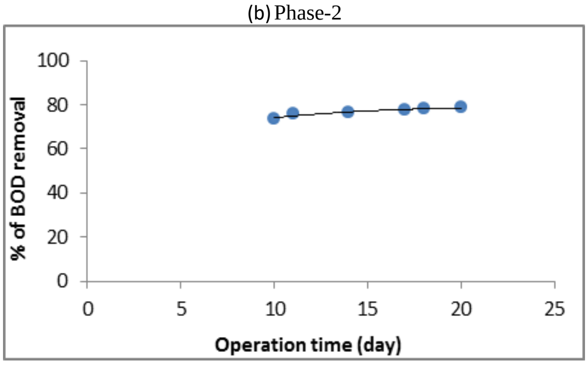
a. During all operation phases, the values of effluent pH are within the limits specified in national standards of secondary effluents.
b. The BOD values of effluent during phases-1 and 2 of plant operation match that of stabilization ponds and trickling filters.
c. The effluent BOD during phase-3 matches that of dispersed growth systems like activated sludge process.




(a) Phase-1




(b) Phase-2



(c) Phase-3

Fig. 8. Percent of BOD removal verses operation time during the three phases of plant operation

TABLE IV

MINIMUM NATIONAL STANDARDS OF SECONDARY EFFLUENTS* [27]

| Effluent characteristics | Average 30-day concentration | Average 7-day concentration |
|---|---|---|
| $BOD_5$ | 30 mg/l | 45 mg/l |
| pH | Within the range at all times | 6.0-9.0 |

* *Present standards allow stabilization ponds and trickling filters to have higher 30-day average concentration (45mg/l) and 7-day average concentration (65 mg/l) of BOD as long as the quality of receiving water is not adversely affected.*

TABLE V

QUALITY PARAMETERS OF FBR PILOT PLANT EFFLUENT

| Phase | BOD (mg/l) | pH Range |
|---|---|---|
| 1 | 40 | 7.1-8.2 |
| 2 | 60 | 7.6-8.1 |

Generally, the performance of the FBR pilot plant is illustrated through the visual comparisons shown in Figs. 9 and 10. The first figure presents a comparison of influent and effluent samples, While, the second figure illustrates the growth of bacteria and formation of bio- particles.

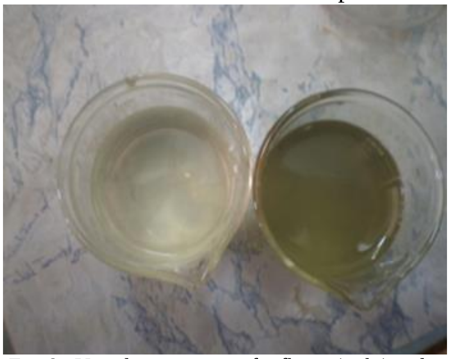


Fig. 9. Visual comparison of influent (right) and effluent (lift) samples

Fig. 10. Visual comparison of clean sand (lift) and bio-particles (right)

## E. *Factors Affecting the Performance of FBR*

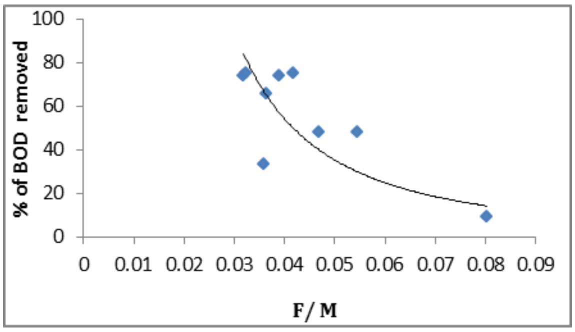
In this study, the effects of both food/microorganisms (F/M) and hydraulic retention time (HRT) on performance of FBR were investigated. F/M is a process parameter commonly used to characterize process designs and operating conditions of biological treatment. This ratio is obtained as [27];

$$\frac{F}{M}=\frac{QS_0}{XV} \qquad \ldots(9)$$

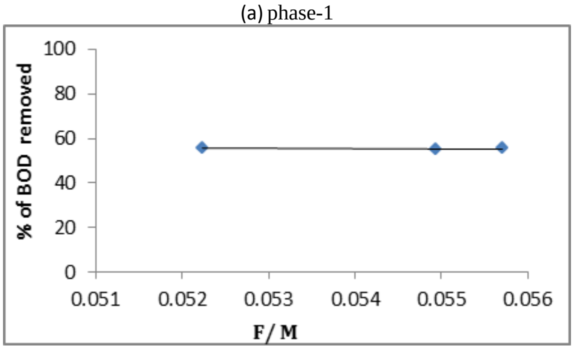
where Q is the influent flowrate ($L^3/T$), $S_O$ is influent BOD ($M/L^3$), X is the mixed liquor volatile suspended solids ($M/L^3$) and V is the reactor volume ($L^3$). In this study, X is obtained as the weighted average of the measured VSS values at different locations along the reactor (locations of sampling points-2 through 6).

The effect of F/M ratio on percent of BOD removal is shown in Figs. 11-a, b, and c for phases-1, 2 and 3. Fig. 11-a shows that the percent of BOD removal decreases with the increase of F/M. In this figure, the F/M values vary over the range (0.032- 0.08 mg BOD/mg VSS.min) and the maximum efficiency of BOD removal (75%) was obtained at F/M equals 0.032 mg BOD/mg VSS.min.

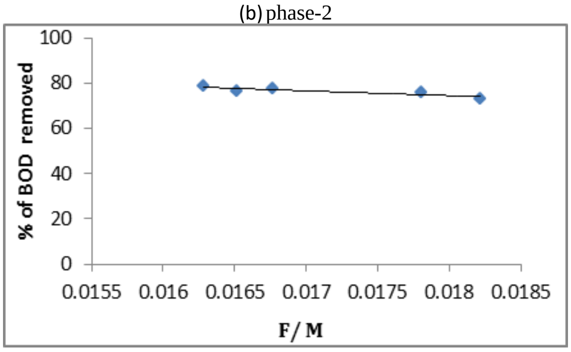
During phase-2, Fig. 11-b shows a narrow range of F/M values (0.052-0.055 mg BOD/mg VSS.min) as compared to phase-1 with nearly constant percent of BOD removal. The same result is obtained during phase-3. Generally, from Figs. 13, it can be noticed that the maximum percent of BOD removal (78.60) was obtained at F/M ratio equals 0.0163 mg BOD/mg VSS.min.




(a) phase-1




(b) phase-2




(c) phase-3

Fig. 11. Effect of F/M on percent of BOD removal

The Hydraulic retention time (HRT) is the period of time that the feed wastewater (Q) remains in the reactor. It is calculated as;

$$HRT = \frac{Q}{V} \qquad \ldots(10)$$

The pilot plant was operated at three values of HRT and the obtained percentages of BOD removal for these times are shown in Fig. 12. This figure shows that the maximum efficiency is obtained at HRT of 24 min during phase-3 of plant operation.

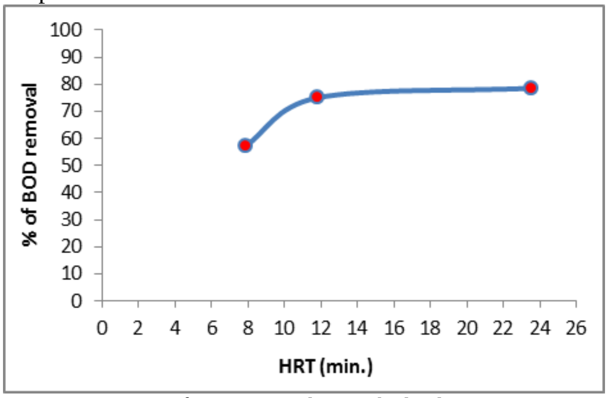



Fig. 12. Percent of BOD removal verses hydraulic retention time

It was shown in Section-D that the effluent quality of FBR during phase-3 matches that of activated sludge system. The values of HRT for activated sludge systems are on order of hours. For example, the HRT of conventional activated sludge systems varies over the range (4-8) hours [27], while, the HRT of FBR was 24 min. Therefore, the required volume of FBR is so small as compared to that of activated sludge system. In addition, it was stated that the performance of FBR during phases-1 and 2 matches that of stabilization ponds. It is important to mention here, that the HRT of stabilization ponds is on the order days [27].

## IV. Conclusions

Based on operation results of fluidized bed reactor treats sanitary sewage, the followings are concluded;

a. The maximum percent of BOD removal is 78.6%. This efficiency was obtained for a hydraulic retention time of 24min and upflow velocity of 1.59 m/min. Although, 75% of BOD removal was obtained at hydraulic retention time of 12 min.
b. The effluent BOD during phases-1 and 2 of plant operation matches that of stabilization ponds and trickling filters and during phase-3 matches that of dispersed growth systems like activated sludge process.
c. During all operation phases, the values of effluent pH are within the limits specified in national standards of secondary effluents.
d. As F/M increases, the efficiency of BOD removal decreases. The maximum efficiency of BOD removal (78.6%) was obtained at F/M ratio of 0.0163 mg BOD/mg VSS.min.

## V. Acknowledgment

The researchers would like to acknowledge the administrator and lab team work of Hamdan Sewage Treatment Plant for allowing the construction and operation of the pilot plant and for facilitating the laboratory work.

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## VII. Biographies

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