Simulation of Dynamic Performance of Biochemical Fermenter

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Abstract

In this work, it was attempted to estimate the mass transfer coefficient (k_La) and study the dynamic behavior of continuous stirred tank fermenter (CSTF) using Matlab program and to solve for mathematical model describing this system. Simultaneous non-linear differential equations were obtained from mass balance and solved for various conditions of agitation speed, gas flow rate, and substrate concentrations. The developed mathematical model describing the production of biomass as a single cell protein (SCP) by treating backer yeast (biomass) with glucose (substrate 1) and oxygen (substrate 2) as an example for CSTF was used in simulation programs. This model based on mass balance equations for each of biomass, glucose, and oxygen which describe the characteristics of the system, and from which the steady state calculation was obtained and then the stability examined by phase plane technique. Dynamic behavior for biomass production was studied under different step changes in initial conditions ± 50% and ± 25% step change in feed oxygen concentration, glucose concentration, and in feed dilution rate (D). It was found that the positive step changes in the feed concentration leads to increase the biomass production while positive step changes in dilution rate leads decrease the biomass production and vice versa. The system was found to be stable from phase plain analysis. Comparison between theoretical results and experimental works of previous authors showed good agreement.

Introduction

Mass transfer in tarried-tank reactors is very important process in chemical and biochemical industry. The oxygen-transfer capacity in specific reactor fundamentally depends on its mechanical design, the geometry of air distributor, and impellers, and the operating conditions, such as the agitation velocity and the agitation rate. These variables, together with the physicochemical and rheologic properties of the absorption medium, are synthetised in a unique parameter, called volumetric gas-liquid mass transfer coefficient (k_La). This parameter is essential in the analysis of the design and scale-up of chemical and biological reactors.

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Mechanism of Mass Transfer

The volumetric oxygen mass transfer coefficient, k_La, (based on specific surface area) plays an important role towards carrying out the design, scaling up, and economics of the process. Several different mechanisms have been proposed to provide a basis for a theory of interphase mass transfer. The mass-transfer coefficient k_L is ratio to the diffusivity D_{AB} to the film thickness z_f as

$$k_L = \frac{D_{AB}}{z_f} \quad (1)$$

Glucose and oxygen are major factors in aerobic fermentation. When oxygen is limited, glucose is metabolized to ethanol, and less cell mass is produced. If the objective is to produce cell mass such as baker's yeast or single cell protein (SCP), oxygen is the most important determinate of cell mass yield. The stoichiometry of glucose oxidation for single-cell organism respiration is

$$C_6H_{12}O_6 + 6O_2 \rightarrow 6H_2O + 6CO_2 \quad (2)$$

Respiration requires 192 grams of oxygen to oxidize 180 grams of glucose. Both glucose and oxygen must be dissolved in the liquid before microorganisms can use them for cell replication. The solubility of oxygen in liquid media is 6000 times lower than that of glucose [1]. Thus, it is not possible to initially provide a microbial culture with all the oxygen it will need for the complete respiration of the glucose. Oxygen must be supplied during

growth at a rate sufficient to satisfy the organism's demand. In glucose fermentation by yeast, it is very important to increase the oxygen transfer rate to the fermentation media because oxygen can be regarded as a nutrient for aerobic microorganisms. Growing microorganisms are capable of removing (metabolizing) oxygen from the liquid faster than it can be supplied as the solubility of oxygen in liquid media is quite low, while its demand for the growth is high. If oxygen supply drops too low, the system may become anaerobic. Oxygen must be supplied to the liquid by aeration and agitation so that cell growth remains aerobic. Aeration supplies the necessary oxygen to the microorganisms, and agitation maintains uniform conditions within the fermenter. Altogether, the aeration and agitation are important in promoting effective mass transfer to liquid medium in the fermenter [2]. The aeration velocity of supplied air and the agitation velocity of the medium will affect greatly the biomass production rate with time. Oxygen is normally introduced into the reactor by supplying compressed air through a circular ring (sparging ring). The resulting air bubbles are distributed by agitating with an impeller powered by an external motor. During fermentation, the transfer of oxygen from an air bubble to the cell can be represented by a number of steps as shown in Fig.1[3].

Ahmad et. al. (1994) [4] found that the volumetric oxygen transfer coefficient for Candida utilize fermentation increased with increasing agitation rate. The agitation rate of the fermenter affects the coalescence and breakup of bubbles, bubble-size distribution, and bubble residence time [3]. Cooper et al. [5] stated that the k_La may be empirically linked to the gassed power consumption per unit volume of broth, P_g/V_L and the superficial air velocity, u_g, as expressed by the following equation:

$$k_L a = b_1 \left(P_g / V\right)^{b_2} u_g^{b_3} \qquad (3)$$

Constant b_2 represents the level of dependence of k_La on the agitation, while, constant b_3 represents the level of dependence of k_La on sparging rate applied to the system. In this equation, the values of the constants b2 and b_3 may vary considerably; depending on the bioreactor geometry and operating conditions. This equation represents the important correlation that correlates k_La and power consumption where P_g is the power absorption in an aerated system (in Watts) , V is the liquid volume in the fermenter (m^3) , and u_g is the superficial air velocity (m/s). P_g is obtained from power number defined as follows [6]:

$$N_P = \frac{P_g}{\rho_c N^3 D_T^5} \qquad (4)$$

Using Fig. 2, N_P is evaluated as a function of Reynolds number according to the type of impeller, N_{Re}, which is defined as a function of rpm, N [6, 48]:

$$N_{\mathrm{Re}} = \frac{\rho_c N D_T^2}{\mu_c} \qquad (5)$$

Cooper et al (1944) [5] investigated the relationship between k_La and power consumption in one impeller stirred tank using the sulfite oxidation technique. They found that the values of b2 and b3 were 0.95 and 0.67, respectively. Riet (1983) summarized the above correlation in two systems: pure water and strong electrolyte solution. For pure water, the values of b_1, b_2, and b_3 were 0.026, 0.4 and 0.5, respectively.

The most widely used continuous culture system is the chemostat, which is characterized by growth control exercised through a growth - limiting substrate. The major feature of a chemostat is a culture vessel containing a fixed volume, (V), of growing culture. A defined growth medium containing one of the constituent substrates S_o at a concentration which is known to be growth-limiting is pumped into the growth vessel at a constant volumetric flow rate (F). By mechanisms such as a weir overflow or another pumping system, culture unused, substrates and products are removed at exactly the same flow rate (F). A certain biomass concentration (C_X) is setup and this also gives the biomass concentration in the effluent culture results through growth, thereby reducing the initial concentration of the growth-limiting substrate, S_o, to a value S_1 [7]. Thus, the growth-limiting substrate concentration depends on a ratio, known as the dilution rate, D:

$$D = F/V \qquad (6)$$

The reciprocal of the dilution rate is known as the mean residence time and denotes the average time that the cell remains within the culture vessel [7].

Hensirisak (1997) [2] observed that the growth rates for MBD at 150 rpm were essentially equivalent with air sparged fermentations at 500 rpm as well as the total power consumption per unit volume is reduced when using MBD. Parakulsuksatid (2000) [8] investigated experimentally a microbubble dispersion (MBD) method to improve oxygen transfer for the production of baker's yeast at low

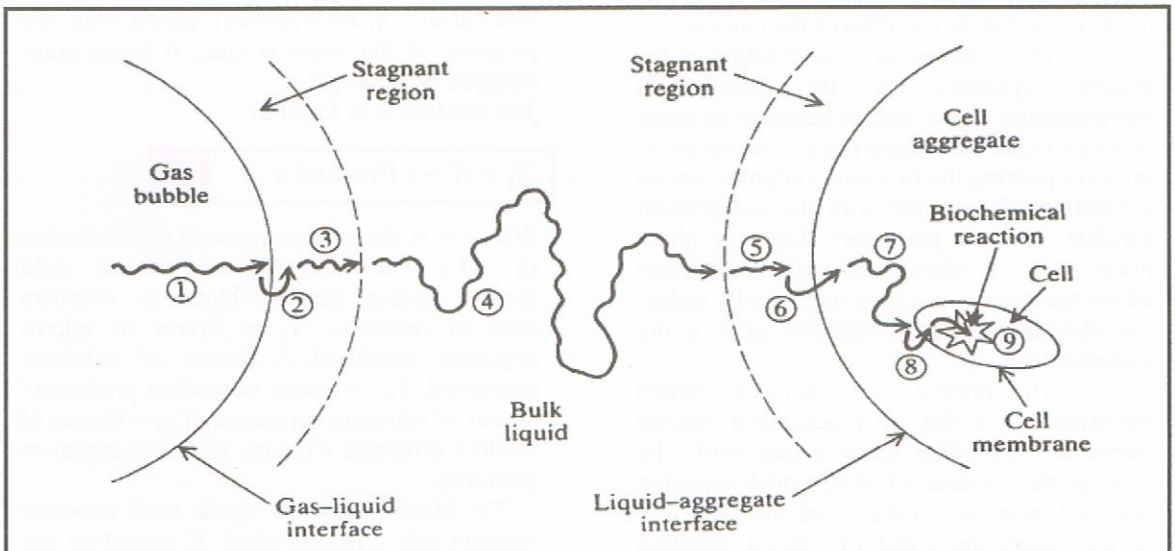

Figure 1: Schematic diagram of oxygen transport from a gas bubble inside a cell [3].

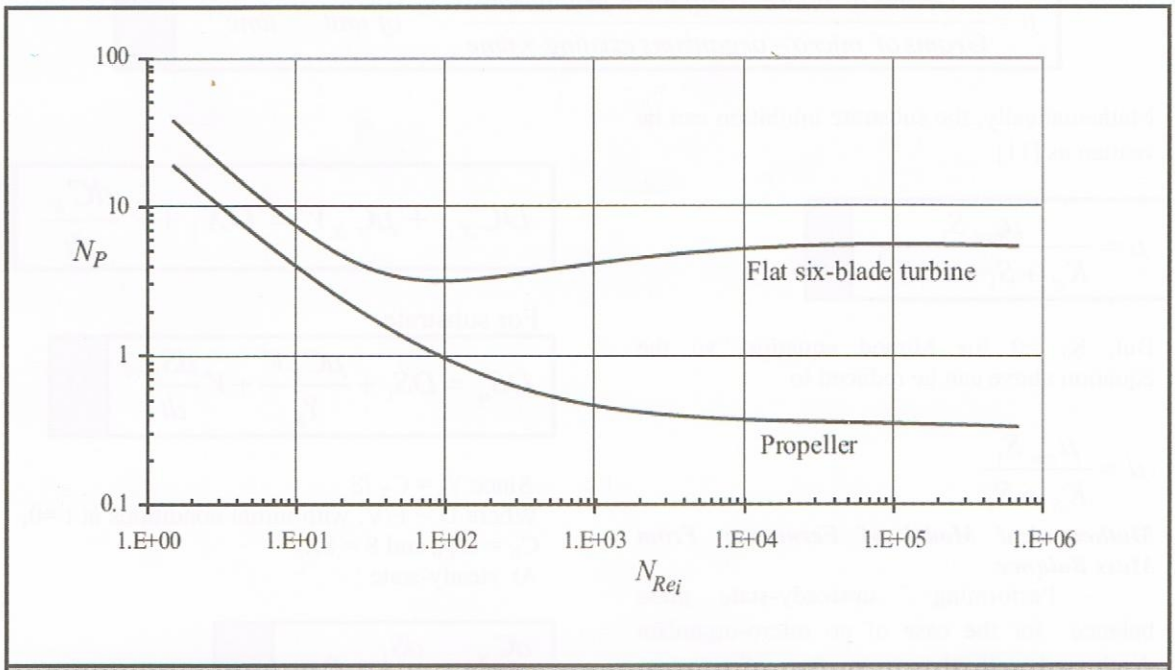

Figure 2 Power number-Reynolds number correlation in an agitator with four baffles [6].

agitation rates (150 rpm) and thus reduce power consumption and shear stress on the microorganisms. Rocha (2003) [9] developed model-based strategies to improve the performance of a high-density recombinant E-coli fed-batch fermentation by contracting a mathematical model framework as well as deriving optimal and adaptive control. The mathematical model of the process is composed of mass balance equations to the most relevant state variables of the process.

The phase-plane representation of the system dynamics is the geometrical representation of the system behavior in terms of trajectories. The phase-plane representation involves plotting the two state variables, one as a function of the other with the independent variable, time, as parameter along the phase plane curve. It shows not only the domain where the steady state is asymptotically stable, but also the transient behavior after a big disturbance.

The present work aims to obtain mathematical model of biochemical reactor (fermenter for baker yeast production) by solving the system of differential equation obtained from mass balance on the fermenter to investigate the effect of liquid agitation velocity, gas flow rate, time, and substrates concentrations (oxygen and glucose) on the mass transfer coefficient and biomass production rate in continous stirred tank fermenter.

Theoretical Analysis and Mathematical Modeling

Modeling of Microbial Systems

Microbial bioreactors differ from enzyme reactors with regard to the fact that the biocatalyst (micro-organism) grows with the progress of the reaction (i.e., it is an auto-catalytic process) [3].

The reaction is as follows:

$$S_1 + X \rightarrow \text{Product} + X \qquad (7)$$

Where x is the micro-organism concentration (g / L). The following biochemical yield factors replace the stoichiometric numbers used in reactions, Y_s = Grams of micro-organism produced / Grams of substrate consumed, Y_{PS} = Grams of product produced / Grams of substrate consumed, Y_{PX} = Grams of product produced / Grams of micro-organism produced.

The Monod equation is the most common reaction rate equation used. It resembles the enzyme rate equation, but it is more empirical [10]:

$$\mu = \frac{Grams\ of\ micro-organism\ produced}{Grams\ of\ micro-organisms\ existing \times time} \quad of\ unit \quad time^{-1} \qquad (8)$$

Mathematically, the substrate inhibition can be written as [11]

$$\mu = \frac{\mu_{max} S_1}{K_S + S_1 + K_I S_1^{\,2}} \qquad (9)$$

But, K_I =0 for Monod equation, so the equation above can be reduced to

$$\mu = \frac{\mu_{max} S_1}{K_S + S_1}$$

Mathematical Model of Fermenter From Mass Balance

Performing unsteady-state mass balance for the case of no micro-organism death and with the assumption of constant volume, yields the following mathematical models: For micro-organisms

$$DC_{X\,o} + \mu C_X V = DX_1 + V\frac{dC_X}{dt} \qquad (10)$$

For substrate

$$DS_o = DS_1 + \frac{\mu C_X V}{Y_S} + V\frac{dS_1}{dt} \qquad (11)$$

Since $Y_s = C_X / S$

Where D = F/V, with initial conditions at t =0, $C_X = C_{X\,o}$ and $S = S_o$.

At steady-state :

$$\frac{dC_X}{dt} = \frac{dS_1}{dt} = 0 \qquad (12)$$

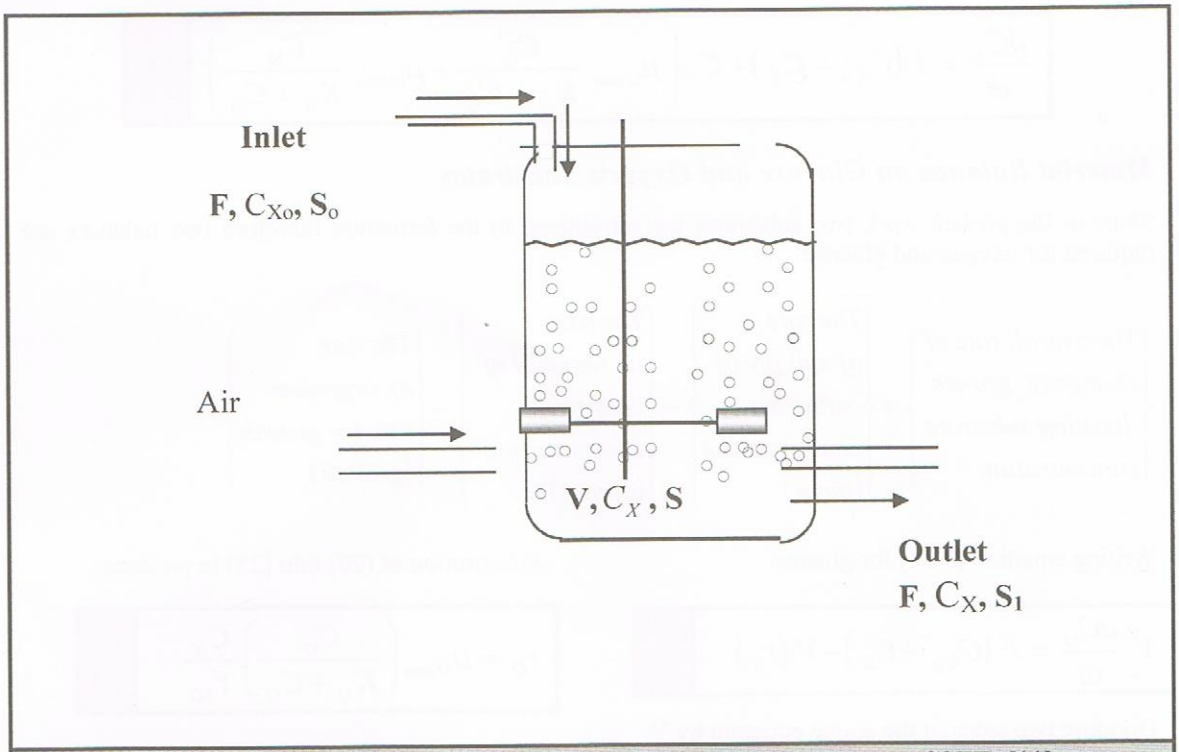

Figure 3: Schematic diagram of continuous stirred tank fermenter (CSTF) [11].

Material Balance on Biomass

For a growing culture, making balance on the biomass according to the following expression:

$$\begin{Bmatrix}\text{The overall rate of}\\ \text{change of biomass}\\ \text{concentration}\end{Bmatrix} = \begin{Bmatrix}\text{The rate of}\\ \text{biomass}\\ \text{input}\\ (in)\end{Bmatrix} - \begin{Bmatrix}\text{The rate of}\\ \text{biomass}\\ \text{removal}\\ (washout)\end{Bmatrix} + \begin{Bmatrix}\text{The rate of}\\ \text{biomass}\\ \text{production}\\ (growth)\end{Bmatrix} \quad (13)$$

$$V\frac{dC_X}{dt} = F(C_{X_o} - C_X) + Vr_x \quad (14)$$

Where $r_x = \mu * C_X$ (15)

There are two substrates

$$r_X = (\mu_G + \mu_o) * C_X \quad (16)$$

Substituting equation (16) into equation (14)

$$V\frac{dC_X}{dt} = F(C_{X_o} - C_X) + V(\mu_G + \mu_o)C_X \quad (17)$$

Dividing equation (17) by V

$$\frac{dC_X}{dt} = \frac{F}{V}(C_{X_o} - C_X) + C_X(\mu_G + \mu_o) \quad (18)$$

Where μ_G, and μ_O (Monod growth kinetics) for glucose and oxygen respectively are given by [12]

$$\mu_G = \mu_{G\max}\left(\frac{C_G}{K_G + C_G}\right) \quad (19)$$

$$\mu_O = \mu_{O\max}\left(\frac{C_O}{K_O + C_O}\right) \quad (20)$$

and $D = \frac{F}{V}$ in h^{-1} (6)

Therefore substituting (19), (20), and (6) in equation (18) to obtain the material differential equation for the biomass:

$$\frac{dC_X}{dt} = D(C_{X_o} - C_X) + C_X\left(\mu_{G\max}\frac{C_G}{K_G + C_G} + \mu_{O\max}\frac{C_O}{K_O + C_O}\right) \quad (21)$$

Material Balance on Glucose and Oxygen Substrates

Since in the present work two substrates are introduced to the fermenter therefore two balances are required for oxygen and glucose

$$\left\{\begin{matrix}\text{The overall rate of}\\ \text{change of growth}\\ \text{limiting substrate}\\ \text{concentration}\end{matrix}\right\} = \left\{\begin{matrix}\text{The rate}\\ \text{of sup ply of}\\ \text{substrate}\\ \text{concentration}\\ (input)\end{matrix}\right\} - \left\{\begin{matrix}\text{The rate}\\ \text{of removal of}\\ \text{substrate}\\ \text{concentration}\\ (output)\end{matrix}\right\} - \left\{\begin{matrix}\text{The rate}\\ \text{of organism}\\ \text{use for growth}\\ (growth)\end{matrix}\right\}$$

Writing equation (3.24) for glucose

$$V.\frac{dC_G}{dt} = F.(C_{G_o} - C_G) - V.(r_G) \quad (22)$$

Dividing two sides of the above equation by V

$$\frac{dC_G}{dt} = \frac{F}{V}(C_{G_o} - C_G) - (r_G) \quad (23)$$

$$r_G = \frac{\mu_G.C_X}{Y_{XG}} \quad (24)$$

Substituting of (19) into (24) leads to

$$r_G = \mu_{G\max}\left(\frac{C_G}{K_G + C_G}\right).\frac{C_X}{Y_{XG}} \quad (25)$$

Substituting equation (25) into (23) to obtain the material balance equation for the glucose:

$$\frac{dC_G}{dt} = D(C_{G_o} - C_G) - \mu_{G\max}\left(\frac{C_G}{K_G + C_G}\right).\frac{C_X}{Y_{XG}} \quad (26)$$

Material Balance on oxygen

For oxygen

$$\frac{dC_O}{dt} = k_L a(C_o^* - C_O) - r_O \quad (27)$$

$$r_O = \frac{\mu_O.C_X}{Y_{XO}} \quad (28)$$

Substitution of (20) into (28) to produce:

$$r_O = \mu_{O\max}\left(\frac{C_O}{K_O + C_O}\right).\frac{C_X}{Y_{XO}} \quad (29)$$

Then equation (27) becomes:

$$\frac{dC_O}{dt} = k_L a(C_o^* - C_O) - \mu_{O\max}\left(\frac{C_O}{K_O + C_O}\right).\frac{C_X}{Y_{XO}} \quad (30)$$

A model of system of three ordinary non linear first order interacting differential equations is produced from the material balance, i.e., Eqs. (21), (26), and (30). Analytical solution of this system of differential equations can be attempted by linearzing the three differential equations but it would be lengthily. Another way for solving this system of differential equations is to use "Mtlab Program". A program is written in Matlab language using commands and functions which are built in Matlab for steady state, dynamic behavior and stability analysis. This programs was written for general user application allowing to represent the dynamics for continuous stirred tank fermenter (CSTF). Eqs. (21), (26), and (30) were solved for various values of agitation velocity, dilution rate, glucose concentration, oxygen concentration, and reactor volume. The volumetric mass transfer coefficient of oxygen was calculated using Wang and Fan K_La Correlation [11]:

$$k_L a = 0.021 * u_{ls}^{0.624}(u_{gs}/1.99 * u_{gs} + 47.1) \quad (31)$$

With

$$u_{ls} = rps * \frac{D}{2} \text{ m/s, and } u_{gs} = F_g/A \text{ m/s}$$

The volume of fermenter is calculated by:

$$V = \frac{\pi}{4} * D_t^{\,2} * L \qquad (32)$$

The power consumption is calculated as follows:

$$Pn = P_g / \left(\rho_c . n^3 . Di^5\right) \qquad (33)$$

From which

$$P_g = Pn . \rho_c . n^3 . Di^5 \qquad (34)$$

Where ρ_c is the density of liquid (kg/m^3), n rps (s^{-1}), D_i Fermentation tank impeller diameter (m), μ_c the liquid viscosity, P_g power, and P_n power No. from Fig 2.

Finding the Steady-State Values

The steady state values can be gotten by solving equations 21, 26, and 30 simultaneously using numerical methods by setting all these equations equal to zero. In other ward there is no accumulation in steady-state, so dC_X/dt, dC_G/dt and dC_O/dt are equal to zero. Matlab program was employed to solve these equations by the functions which are built in it, then check these points (steady-state points) using phase plane technique. Many models of continuous culture systems are formulated in terms of only two dependent variables (e.g. average cell and substrate concentration), and are thus ideally suited for phase-plane analysis [35] 13.

Results and Discussion

Table 1 lists the values of mass transfer coefficient is calculated using Wang and Fan Correlation (Eq. (31)) [11].

Table 1 k_La calculations for 1 liter volume

rpm	u_{ls} (m/s)	u_{gs} ms^{-1}	k_La s^{-1}
144	0.043348	0.000181	0.000629
150	0.045154	0.000181	0.0006453
476	0.14329	0.000181	0.0013264
500	0.150515	0.000181	0.0013677

Table 2 lists the values of steady state concentration of biomass, glucose, and O_2 as calculate by solving Eqs. (21), (26), and (30) numerically using Matlab.

Table 2: Steady State Results at different volumes

Agitation speed (rpm)	Fermentation volume (Liter)	C_{Xs} (g / L)	C_{Gs} (g / L)	C_{Os} (g / L)
144	1	2.4498	1.3628	0.0057
476	1	4.9353	1.6451	0.0058
150	50	2.3734	1.3469	0.0058
500	50	4.4127	1.6078	0.0058

Mass Calculation with Stability Analysis

The second part for studying equilibrium points is mass balance for biomass, substrate 1 to get the best equilibrium point for biomass, substrate 1 (glucose) and substrate 2 (oxygen) concentrations from three equations (21, 26, 30) using "fsolve" Matlab function to get the equilibrium point by using different initial points of C_X, C_G and C_O in a program (RRRR2), by using this program, three steady state (Equilibrium points) were obtained, which were:

X_S =2.4498, G_S = 1.3628, and O_S =0.0057

Before starting the study of the dynamic behavior, the equilibrium points must be checked. The method of stability was used in order to check and analyze the equilibrium points, is the phase plane technique by plotting different initial points of biomass concentration (C_X) with substrate concentration (C_G). Fig. 4 shows the phase plane plot between C_X and C_G. The figure reveals that all initial conditions approach the equilibrium point. Figures 5 and 6 show the interaction of dynamic behavior between the cell mass, dissolved oxygen, and glucose profiles also it illustrates the relations between the biomass with glucose and oxygen respectively. The Figs indicate that the biomass concentration increase with time because it is produced by biological reaction. Also the glucose is decreased with time because it is consumed by the biomass. After about 40 min both biomass and glucose reaches their asymptotic values. After 40 min the biomass increases about 6 times and glucose decreases by 6 times.

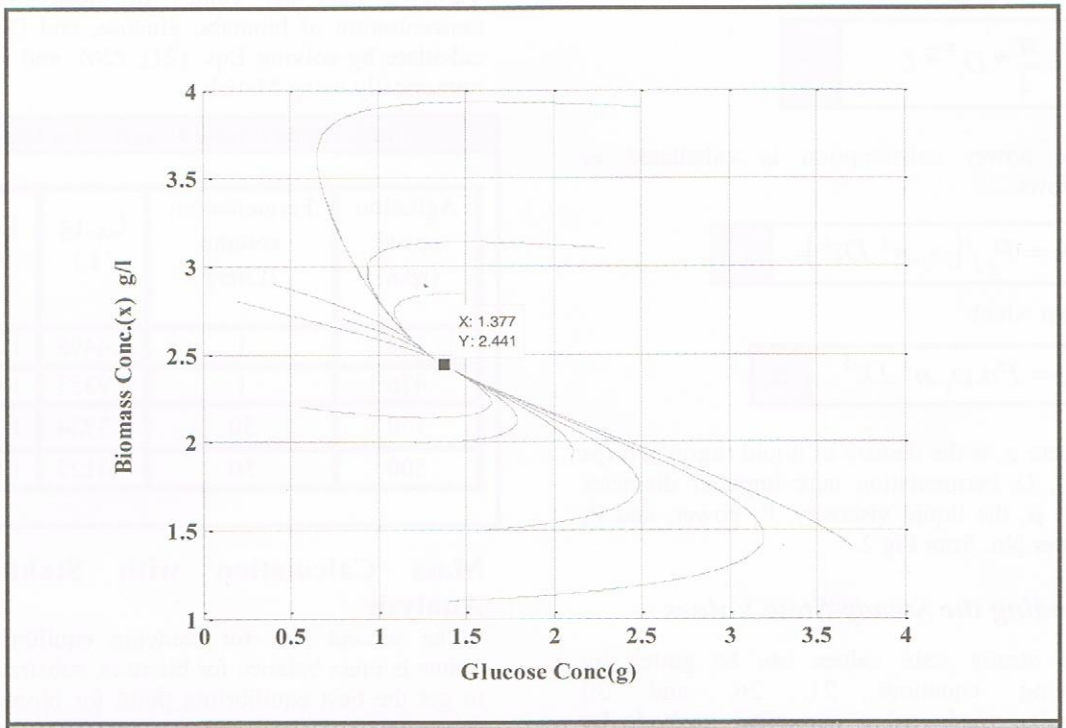

Figure 4: Phase plane plot between Glucose concentration (C_G) and Biomass Concentration (C_X).

Fig. 6 shows the variation of oxygen concentration and biomass concentration with time. The figure reveals that the asymptotic values of oxygen concentration and biomass concentration are after 50 min where the biomass concentration increases by 5% and the oxygen concentration decreases by 5%. It is to be noticed that the effect of oxygen on biomass production is much lower than that of glucose.

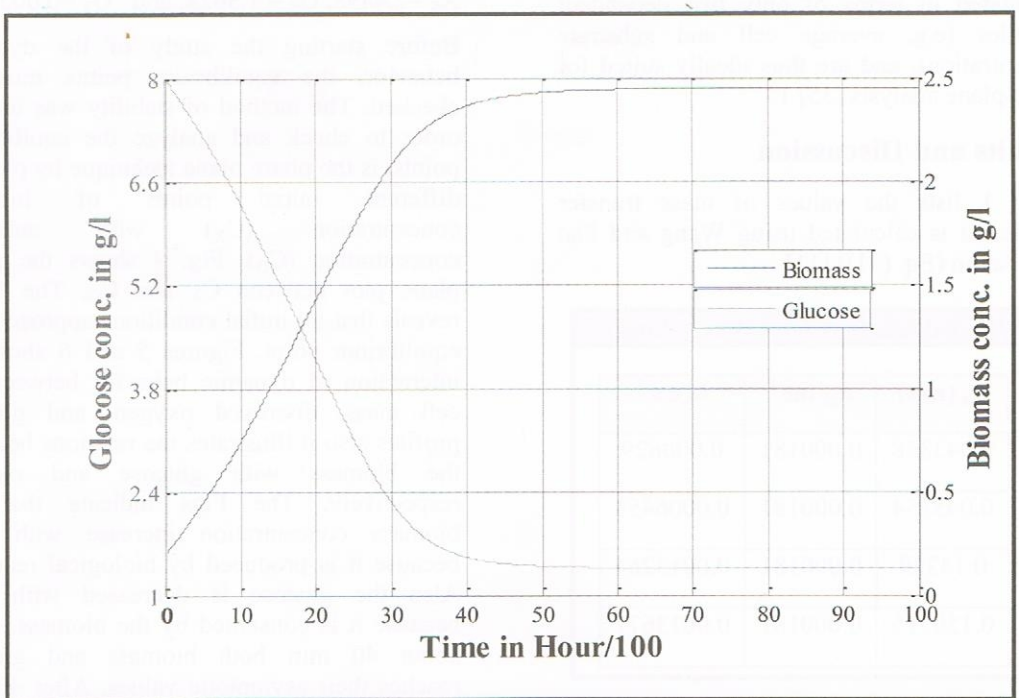

Figure 5 Dynamic Behavior for both of Biomass concentration (C_X) and Glucose concentration (C_G) in normal case

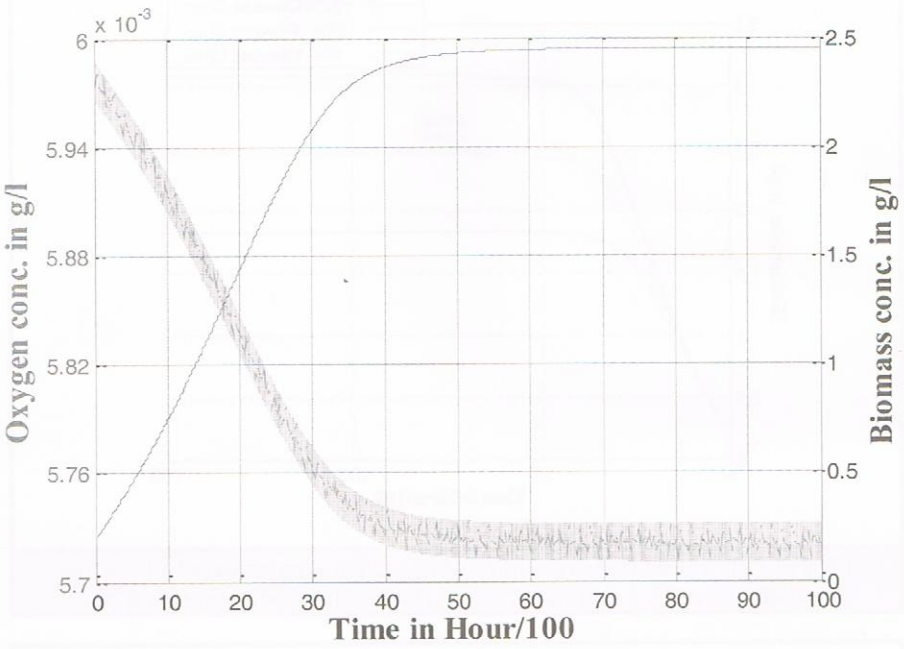

Figure 6 Dynamic Behavior for both of Biomass concentration (C_X) and Oxygen concentration (C_O) in normal case.

For $\mp$ 25% and $\mp$ 50% step changes in feed substrate concentration (C_{Go}). Figures 7 and 8 show the variation of biomass concentration with time for step change in feed glucose. It is evident that the biomass concentration increases with time due to the biological production reaction. Also, these figures indicates that as the glucose concentration in the feed increases the biomass production increases and vice versa. The step change in glucose concentration of 25% cause 30% change in the biomass production rate. The change is due to the change of the nutrient that accelerate the growth of biomass. Figure 7 shows the variation of oxygen concentrations with time for with step change in the glucose concentration. The figure reveals that the glucose concentration decreases with time due to its oxidation by chemical reaction 3.20. The oxygen concentration decreases by respiration from the biomass till reach steady state curve. It is clear that increase in glucose concentration in the feed by 25% leads to decrease the oxygen concentration by 1%. Also the decrease in the glucose concentration by 25% leads to increase O_2 concentration by about 1%. the reason behind this trend is that the increased glucose concentration leads to decrease O_2 by promoting reaction 3.20. Figure 8 indicates that the step change in the of glucose concentration of ±50% cause the oxygen concentration to change by about ±2.1%.

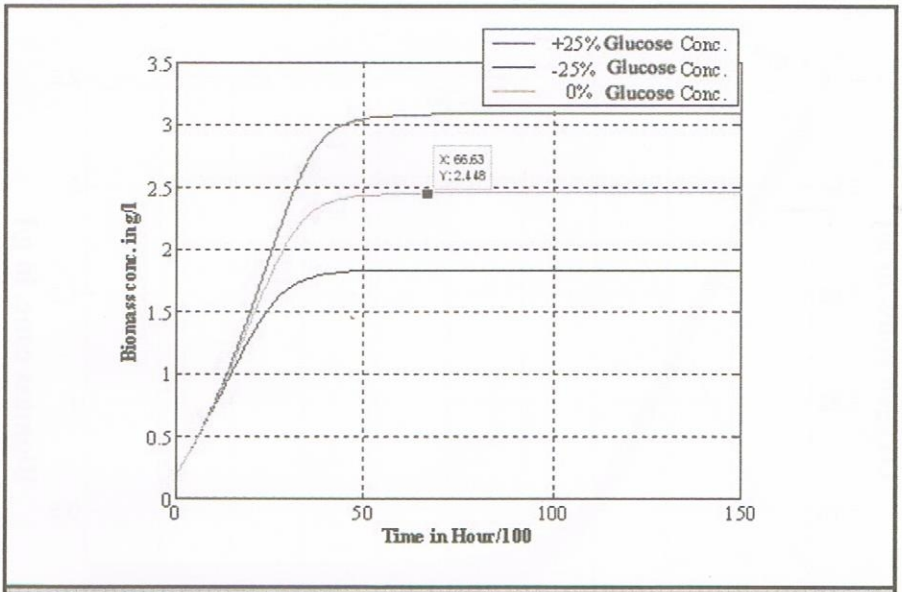

Figure 7 Dynamic Behavior of adding ± 25% step change in Glucose concentration (C_G) and its effect on the Biomass concentration (C_X)

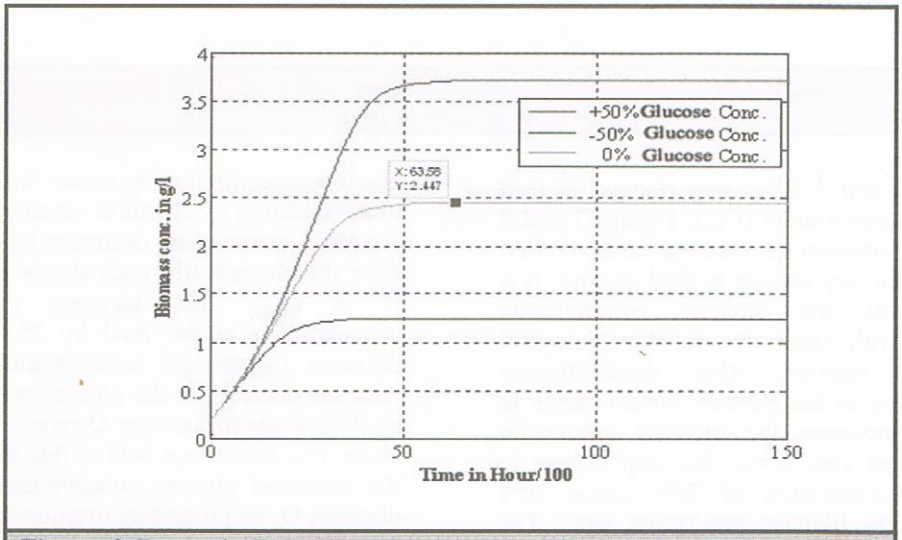

Figure 8 Dynamic Behavior of adding ± 50% step change in Glucose concentration (C_G) and its effect on the Biomass concentration (C_X)

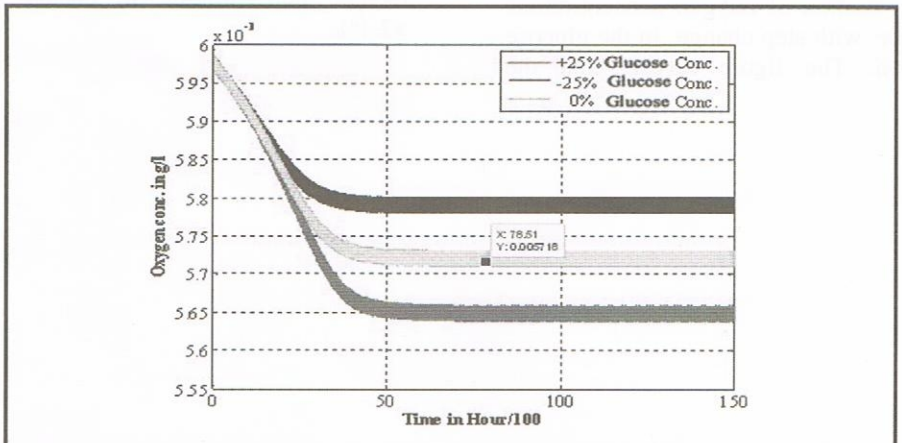

Figure 9 Dynamic Behavior of adding ± 25% step change in Glucose concentration (C_G) and its effect on the Oxygen concentration (C_O)

Step Change in Input Dilution Rate (D)

Figures 11 and 12 show the effect of step changes in dilution rate (D) on biomass concentration. For $\mp$ 25% and $\mp$ 50% step change in input dilution rate (D). From Fig. 11 it can be seen that the step change in dilution rate by +25% leads to change the biomass concentration in the reactor -36%. While the step change by -25% leads to change the biomass concentration by +16%. Step changes in D by +50% and -50% leads to change the biomass concentration by -68% and by 48%. The decrease in biomass concentration with D increase is due to washing out of cells [7] .

Figures 13 and 14 show the effect step change in D on the glucose concentration. It is evident that increasing dilution rate by 25% and 50% leads to increase the glucose concentration by 53% and 300% respectively. Also decreasing dilution rate by 25% and 50% leads to change the glucose concentration by 50% and by 77% respectively.

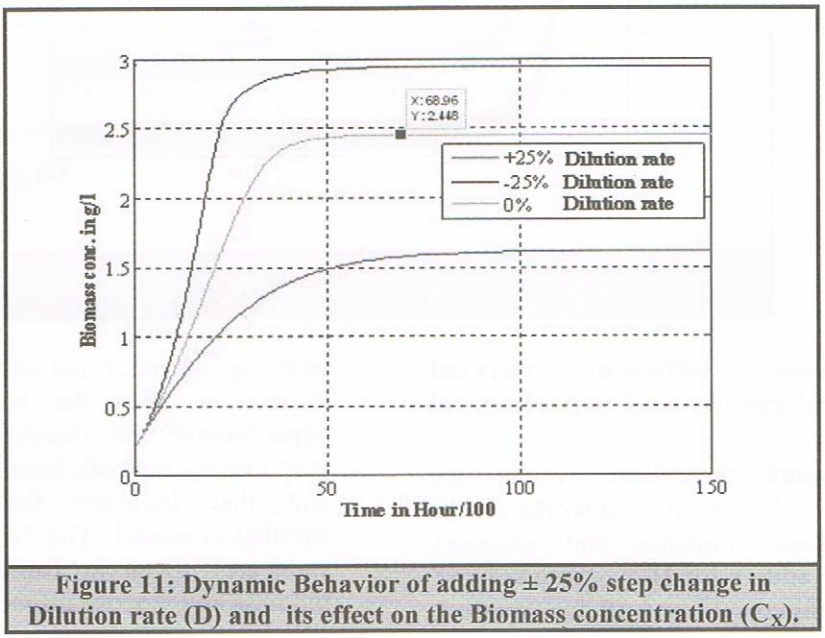

Figure 11: Dynamic Behavior of adding ± 25% step change in Dilution rate (D) and its effect on the Biomass concentration (C_X).

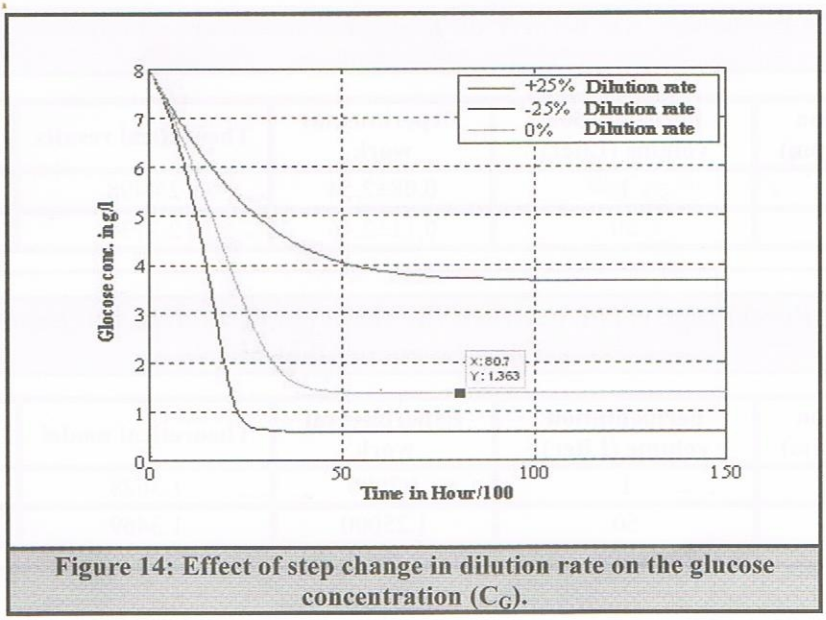

Figure 14: Effect of step change in dilution rate on the glucose concentration (C_G).

Figures 14 and 15 shows the effect of step changes in the dilution rate on the glucose concentration and oxygen concentration. Step change in D by 25% and 50% leads to increase the increase the O_2 concentrations by 1.7% and by 3.5% respectively.

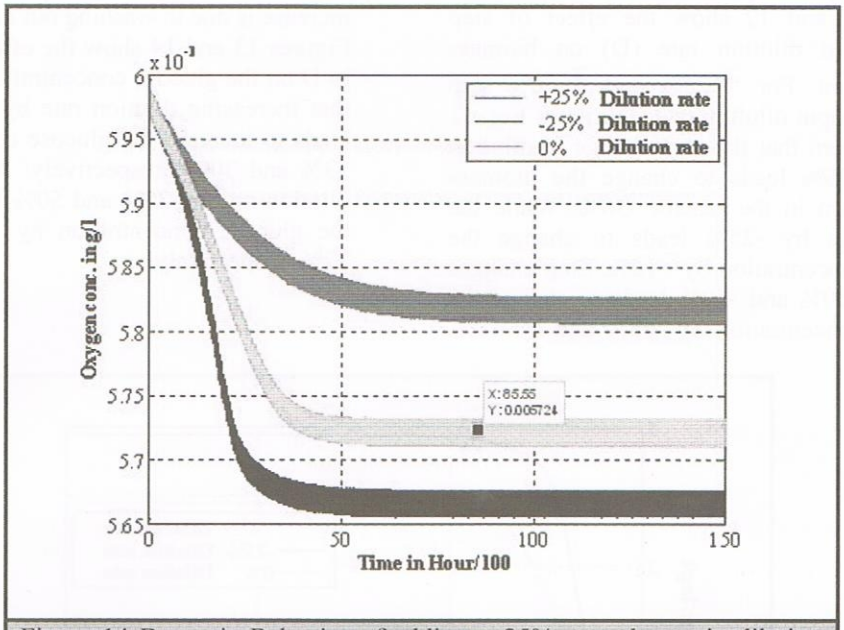

Figure 14 Dynamic Behavior of adding ± 25% step change in dilution rate (D) and its effect on the Oxygen concentration (C_O).

Comparison between current theoretical results and experimental results

The presents theoretical results are compared with experimental works for the same system (biomass and glucose). Figures 15 and 16 for 1L working volume, while the figures 17 and 18 for up 50 L working volume. From all of these figures it can be seen the behavior of the experimental and theoretical works are very closely for both biomass and glucose and that indicates the accuracy of simulation model. The % error at study-state are given in Tables 1 and 2 for biomass and glucose respectively:

Table 1 Percentage error between the theoretical models and experimental for Biomass concentrations

Agitation speed (rpm)	Fermentation volume (Liter)	Experimental work	Theoretical results	% error
144	1	0.08±2.54	2.4498	3.55
150	50	0.11±2.46	2.3734	3.52

Table 2 Percentage error between the theoretical models and experimental for Glucose concentrations

Agitation speed (rpm)	Fermentation volume (Liter)	Experimental work	Theoretical model	% error
144	1	1.2000	1.3628	13.567
150	50	1.25000	1.3469	7.752

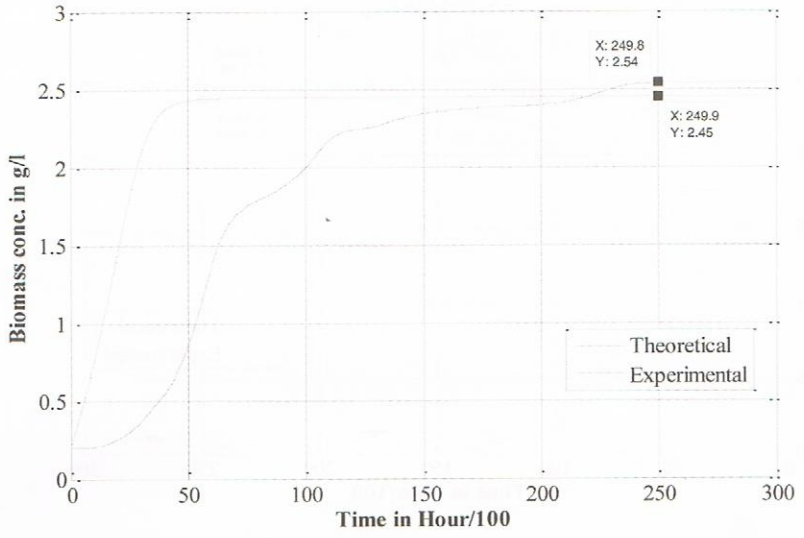

Figure 15 Comparison between an experimental biomass concentration (Blue) and theoretical biomass concentration (Red) for 1-liter fermentation with air sparging at 144 rpm.

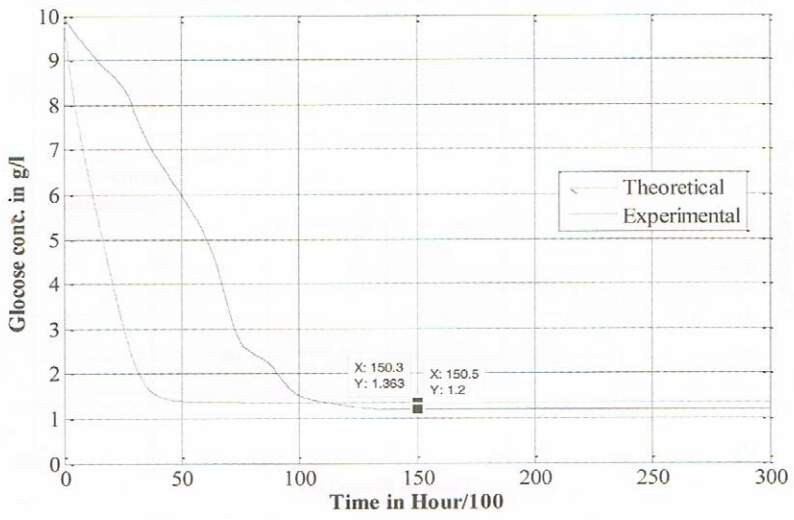

Figure 16: Comparison between an theoretical experimental glucose concentration for 1-liter fermentation with air sparging at 144 rpm.

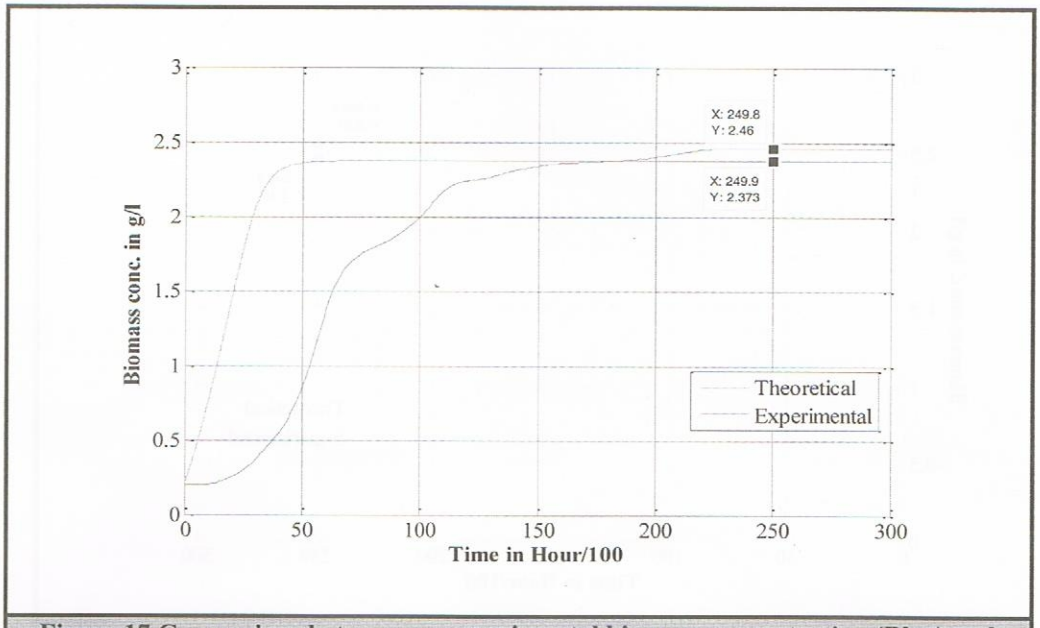

Figure 17 Comparison between an experimental biomass concentration (Blue) and theoretical biomass concentration (Red) for 50-liter fermentation with air sparging at 150 rpm

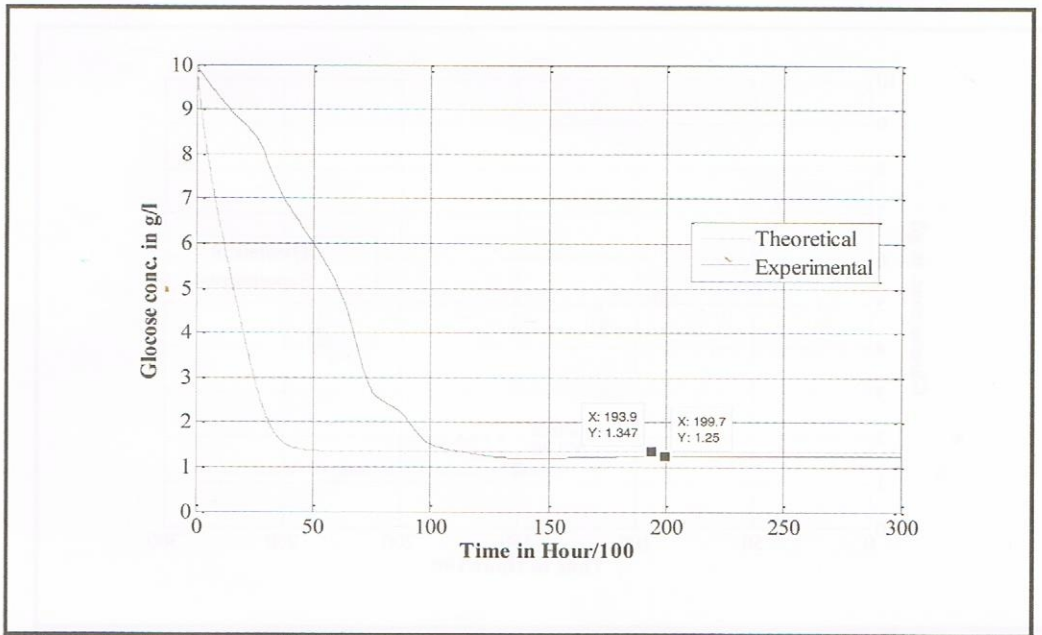

Figure 18 Comparison between an experimental glucose concentration (Blue) and theoretical glucose concentration (Red) for 50-liter fermentation with air sparging at 150 rpm.

Conclusions

The following conclusions are conducted from the present work:

1. The liquid agitation velocity and gas flow rate affect the mass transfer coefficient of oxygen and therefore the biomass production.
2. Step change in glucose concentration by ± 25% causes change in the biomass production by ± 20%. Step change in glucose concentration by ± 50% causes change in the biomass production by ± 45%. The increase in dilution rate by 25% leads to increase the biomass production by 23% while the decrease in dilution rate by 25% leads decrease the biomass production by 36%. Also

increasing dilution rate leads to increase the concentration of glucose and oxygen, while increasing glucose concentration leads to decrease th oxygen concentration.

3. There are single steady state (equilibrium points) exist in each of different cases that varies with volume and agitation velocity when Monod kinetics is applied in a mathematical model.
4. The equilibrium points are tested according to the phase plane technique and show the system is stable.
5. The theoretical results are in a good agreement with experimental results of previous works for the same system, i.e, CSTF, backer, yeast and Glucose).
6- The biomass production rate increases asymptotically with time while the glucose and oxygen concentrations decrease asymptotically with time. The asymptotic values are reached after 4000 h.

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Nomenclature

a_0 Gas-liquid interfacial area

A Cross sectional area of Fermenter

C_* Existing oxygen concentration

C Dissolved oxygen saturation concentration in liquid, gmol/Liter

C concentration, gmol/liter

D Dilution rate of inlet stream

D_{32} Sauter-mean diameter

D_{AB} Diffusivity of component A through B

D_i Impeller diameter

D_t Tank/vessel diameter

F Air Flow rate

G Substrate glucose

$K_L a$ Volumetric mass transfer coefficient , h^{-1}

Greek Letters

Δ Difference between two values

μ Specific growth rate , h^{-1}

μ_{max} Maximum specific growth rate, h^{-1}

μ_c Viscosity of process liquid, kg / m.s

μ_{Gmax} Glucose maximum growth rate, h^{-1}

μ_{Omax} Oxygen maximum growth rate h^{-1}

μ_G Glucose specific growth rate, s^{-1} or $kg/m^3 s$

μ_O Oxygen specific growth rate s^{-1} or $kg/m^3 s$

ρ_c Density of process liquid, kg / m^3

Subscript

c	physical properties of water
g	Gas phase (air)
G	Glucose
S	Steady state
i	inlet conditions
max	Maximum
l	liquid phase
O	Oxygen or for outlet conditions

Abbreviations

CSTF	Continuous Stirred Tank Fermenter
MATLAB	MATrix LABoratory
MBD	Microbubble dispersion

المحاكاة النظرية للأداء الحركي لمفاعل التخمير الحيوي

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الخلاصة:

تم اجراء دراسة نظرية لحساب معامل انتقال الكتلة والسلوك الحركي لخزان تخمير هزاز (CSTR) باستخدام برنامج ال (MATLAB) لحل المعادلات التفاضلية التي تصف النظام. تم الحصول على معادلات تفاضلية انية لاخطية من خلال اجراء موازنة الكتلة وتم حلها لظروف مختلفة من سرع الخلط وسرعة الهواء وتركيز الخلايا الحية والاوكسجين والكلوكوز. تم استخدام المعادلات المستحصل عليها لحساب متغرات النضام. هذه المعادلان اتي تم اتوصل اليها والمبنية على موازنة الكتلةللخلايا الحية والاوكسجين والكلوكوز استخدمت للحصول على قيم الحالة المستقرة. وجد ان زيادة تركيز الاوكسجين والكلوكوز بمقدار 25% و 50% يؤدي الى زيادة انتاج الخلايا الحية بينما زيادة معدل التخفيف (D) بنفس النسب يؤدي الى نقصان معدل انتاج الخلايا الحية والعكس بالعكس. كذلك وجد ان النضام مستقر من خلال مستوي الحالة. من خلال مقارنة النتائج النظرية مع النتائج العملية لباحثين اخرين وجد انه هنالك تطابق جيد.