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Design of PI Controller for Bioreactors for Maximum Production Rate

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Abstract: Bioreactor control has become an active area of research in recent years. In the present work, a conventional PI controller is developed for controlling a bioreactor in which cell growth follows Monod kinetics. The performance of the tuning scheme is studied for maximum biomass production rate by simulating the non-linear model equations of the bioreactor for both servo and regulatory problems.
Keywords: Bioreactor, pi controller.

Introduction

Biochemical reactors are used in a wide variety of processes from waste treatment to fermentation for the production of biochemical. They are inherently nonlinear. In spite of the knowledge that one of the characteristic is inherent nonlinearity of the process, it is traditionally controlled using linear control design techniques The ability of proportional integral (PI) controllers (1-3) to compensate most practical industrial processes has led to their wide acceptance in industrial applications. In particular these controllers perform well for processes with benign dynamics and modest performance requirements. The objective of the present work is to design a conventional PI controller for bioreactor with Monod kinetics.

The dynamic model of a continuous stirred tank reactor

In this section we present the dynamic model of a continuous stirred tank bioreactor where a single population of microorganism is cultivated on a single limiting substrate[4]. A typical control and instrumentation diagram of the bioreactor with biomass concentration as the measured output is shown in figure 1.

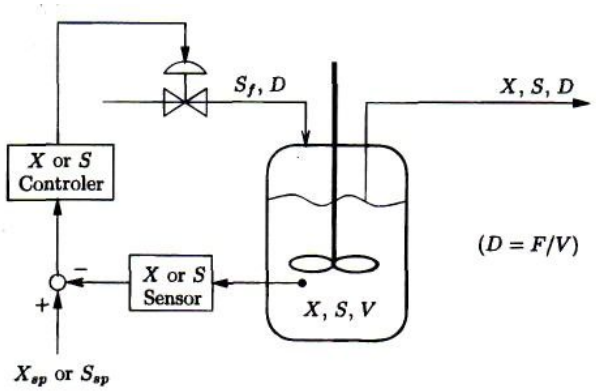


Figure 1 Schematic diagram of a continuous bioreactor

A variety of fermentation processes can be described by the unstructured model[5]

$$\frac{dx}{dt} = (\mu - D).x. \qquad \text{... (1)}$$

$$\frac{dS}{dt} = D(S_f - S) - \mu.x - \frac{1}{Y_{x/s}} \qquad \text{... (2)}$$

$$\frac{dP}{dt} = \mu.x.\ Y_{p/x} - D.P \qquad \text{...(3)}$$

Where x, S, P and μ are the biomass concentration, substrate concentration, product concentration and the specific growth rate respectively.

$$D = \text{dilution rate} = F/v = \frac{Vol.\ flow\ rate}{reactor\ volume}$$

S_f is the substrate feed concentration, $Y_{x/s}$ is the yield coefficient for cell mass and $Y_{p/x}$ is the yield co-efficient for product[6]

Many empirical expressions have been proposed for the function μ(s) and we consider the Monod model (7) which is the most commonly used classical function:

$$\mu(s) = \frac{\mu_{max}.S}{Ks + S} \qquad \text{.... (4)}$$

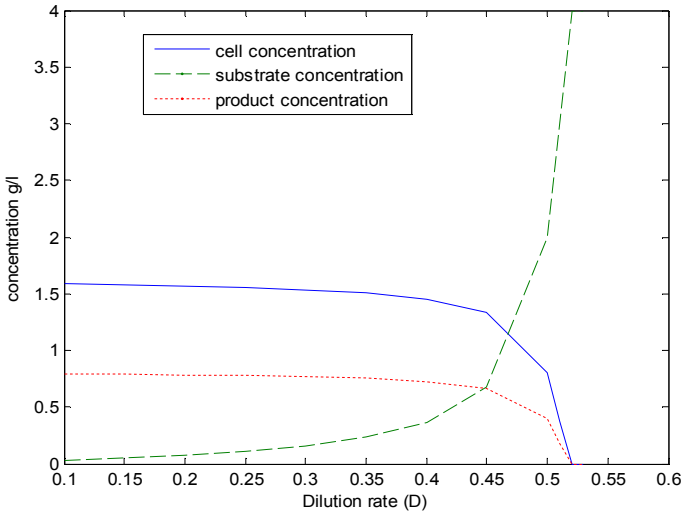
where μ_{max} is the maximum growth rate constant and *Ks* is saturation constant.

Steady state conditions

For systems obeying Monod's model, Equations (1) to (4) have been solved for steady state conditions and the results are shown in figure2.

Figure 2. Dependence of effluent cell concentration x,substrate concentration S,product concentration P on continuous culture dillution rate D as computed from the Monod model

The figure indicates that cell mass concentration is high only at very low dilution rates(D). The cell mass production rate (g/l/h) is obtained by multiplying x and D. Since dilution rate represents the volumetric flow rate(F) (D=F/V where V is the volume of the bioreactor),at very low flow rates even though cell mass concentration is high, the production rate xD will be low. As D increases xD will increase since x is nearly constant. At very high dilution rates, the cells will not reside inside for sufficient time and hence lesser amount of cells will be formed leading to lower cell mass concentration. Even though D is very high xD decreases since x is very low. This trend indicates that the production rate will pass through a maximum when the dilution rate is varied between 0 and $\mu_{max,}$ Since D can be raised only up to $\mu_{max.}$, It is preferable to operate the bioreactor close to this maximum in order to achieve highest production rate. This is shown in figure 3.



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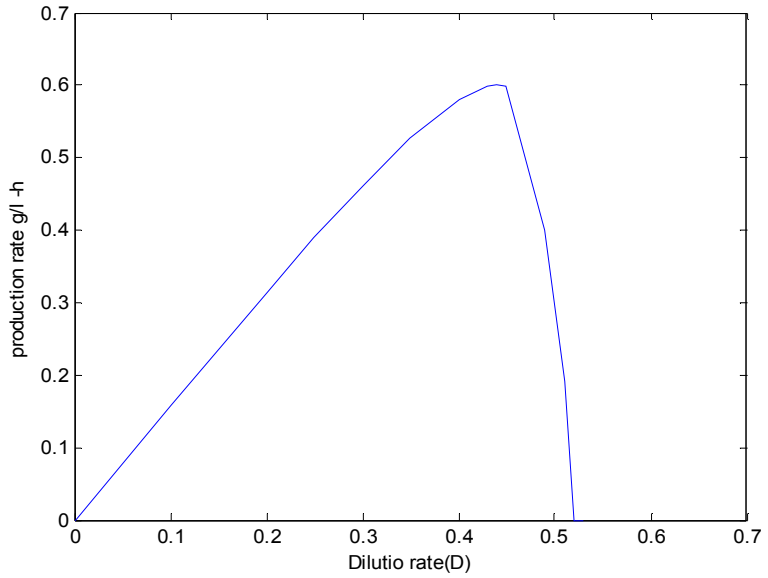

Figure 3. Dependence of production rate xD on continuous culture dilution rate D computed from Monod model

In the present control study the parameters used for the Monod model are

μ_{max} = 0.53 h^{-1}, Ks = 0.12 *g/l*, $Y_{x/s}$= 0.4, $Y_{p/x}$ = 0.5, S_f= 4.0 *g/l*,

The nonlinear process has the following steady state for a dilution rate of 0.43 h^{-1} at which the production rate xD is maximum,

Biomass concentration x = 1.3936 g/l
Substrate concentration s = 0.5160 g/l
Product concentration p = 0.6968 g/l

The maximum production rate is given by :

$$D_{max.output} = \mu_{max}\left[1 - \sqrt{\frac{K_S}{K_S + S_F}}\right]$$

The process is controlled at this operating point.

Controller Design

The state space formulation (8) is used to linearize the nonlinear equations around the steady state operating point. The transfer function relating the dilution rate to the biomass concentration is,

$$Gp(s) = \frac{-1.39s^2 - 1.199s - 0.2577}{s^3 + 1.408s^2 + 0.656s + 0.1013}$$

In process control the step response of highly over-damped system is usually approximated with a first order model. On approximation, the above transfer function is modeled as a first order system as shown in figure 4.

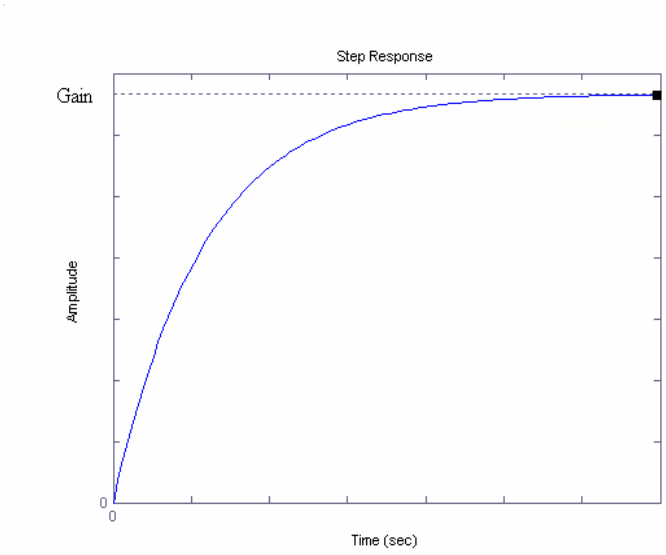

Figure 4. Response of a first order system

From the first order response the process gain (Kp) and the time constant (τ_p) are determined. The process gain is the ratio of the steady state step response to the magnitude of a step response. The time constant (τ_p) of the system is the time at which the response is 63.2% of final value. The values of Kp and τ_p are -2.544 and 1.8538 respectively. Direct Synthesis method for a first-order process is given by (7),

$$K_c \quad \frac{\tau_P}{K_p}$$

$$T_i \quad \tau_p$$

where Kc and Ti are **c**ontroller gain and integral time constant respectively.
From **the** values of Kp and τ_p **the** controller settings are determined as

K_c = – 2.544, T_i = 1.8538 and λ is the only tuning parameter and λ is taken as half of the time constant.

Results and discussions

In this study we present the simulation results of servo and regulatory responses of a PI controller for a dilution rate D = 0.43 h^{-1} (at which the production rate xD is maximum) is presented.
Servo responses of the bioreactor for positive set-point changes in cell concentration for 5% and 10% from stable steady state (x=1.3936 g/l) are shown in figure 5. For the feed concentration 4 g/l the steady state cell mass concentration X cannot be more than 1.6. since, the steady state cell mass concentration is the product of feed concentration and yield co-efficient. Hence in the positive direction the set point change cannot be more than 14.8 %.

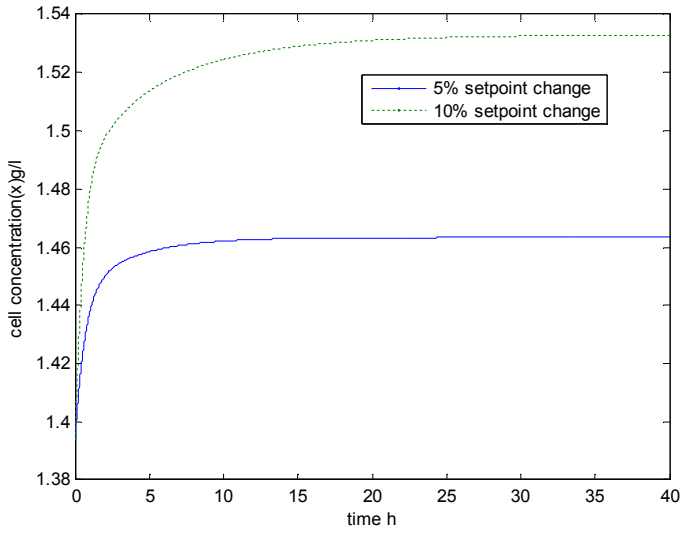

Figure 5 . Servo responses of the bioreactor for positive set-point changes in cell concentration for 5% and 10% from stable steady state (x=1.3936 g/l)

Servo responses of the bioreactor for negative set-point changes in cell concentration for 5%, 7%, 10% and 15% from stable steady state (x=1.3936 g/l) are shown in figure 6.

Regulatory responses of the bioreactor was studied by giving step change in feed concentration of the substrate. S_f was varied from 4.0 to 5.0 for step up and for step down from 4.0 to 3.9 g/l is shown in figure 7.

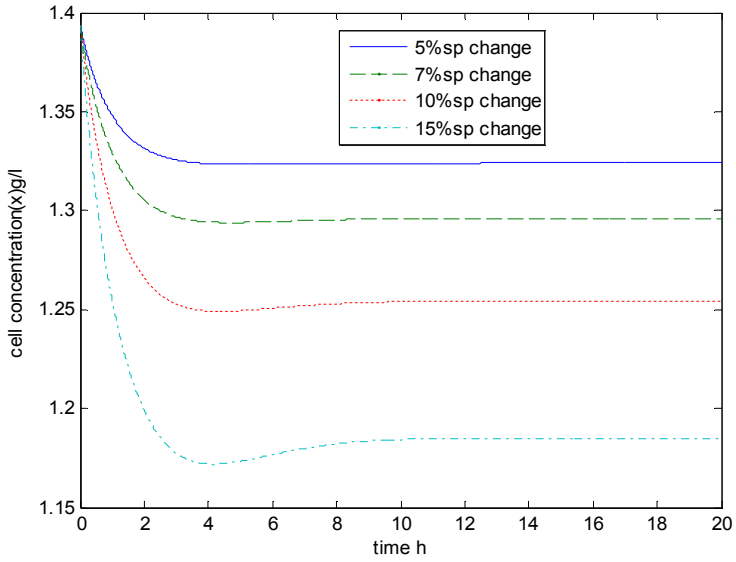

Figure 6. Servo responses of the bioreactor for negative set point change in cell concentration for 5%,7%,10% and 15% from stable steady state (X=1.3936g/l)

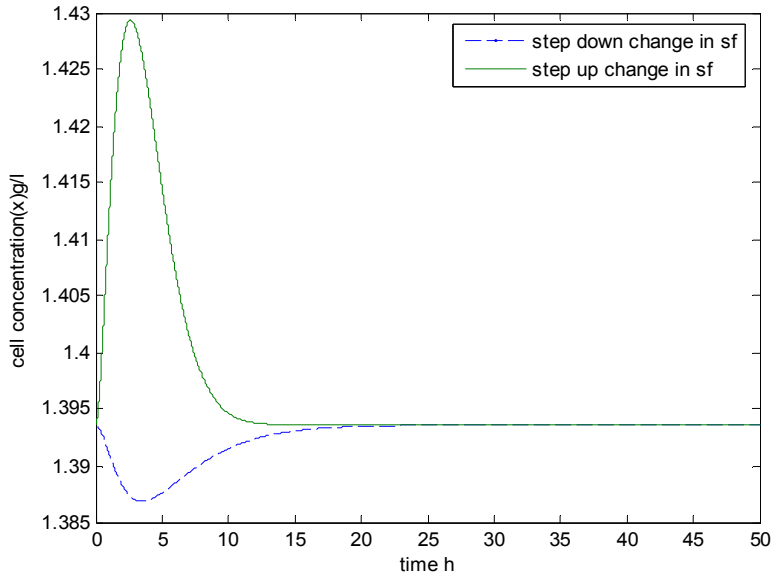

Figure 7. Regulatory responses of the bioreactor for step up in s_f from 4.0 to 5.0 and for step down in s_f from 4.0 to 3.9 g/l

Performance evaluation

For servo responses, for step changes in the positive direction there is no overshoot and the response is good. For step changes in negative direction there is no significant overshoot up to 10%.The magnitude of overshoot increases with increase in size of step change. The regulatory response shows overshoot for both positive and negative step change in substrate feed concentration.

Performance evaluation for both servo and regulatory responses are given in tables 1-3.

Table 1: Servo response-performances of controllers for positive set-point change in cell concentration (X) from 1.3936 g/l.

Change in cell concentration g/l	% change	Settling Time(h)	% overshoot	ISE	IAE
1.4633	5	10	0	0.0125	0.3621
1.5330	10	20	0	0.0436	0.7974

Table 2: Servo response-performances of controllers for negative set point change in cell concentration(X) from 1.3936 g/l.

Change in cell concentration g/l	% change	Settling Time(h)	% overshoot	ISE	IAE
1.3240	5	6	0	0.0022	0.0648
1.2960	7	6	0	0.0043	0.0917
1.2542	10	8	0.34	0.0089	0.1357
1.1846	15	10	16.8	0.0201	0.2171

Table 3: Regulatory response-performances of controllers for both positive and negative change in feed concentration(S_f) from 4.0 g/l

Change in cell concentration g/l	% change	Settling Time(h)	% overshoot	ISE	IAE
5.0	25	15	2.540	0.0042	0.1677
3.9	-2.5	20	0.481	0.00020	0.0512

Conclusion

In the present work the design and implementation of conventional PI controller for biochemical reactors with Monod kinetics have been presented. The performance of the closed loop system with conventional PI controller is evaluated and the performances are tabulated. As a future work the fuzzy logic and neuro controllers can be tested on the bioreactors.

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