



Temperature control in packed-bed solid-state bioreactors

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Abstract

Normal operation of packed-bed bioreactors used for solid-state fermentation involves a static bed aerated from the bottom throughout the fermentation. Due to the rapid heat generating dynamics, temperature control of the bed is very important. In this paper two control algorithm namely conventional proportional- integral (PI) and globally linearizing control (GLC) are used for temperature control of a packed-bed bioreactor, in which *Aspergillus niger* grows on wheat bran. The study was done using the kinetic and distributed dynamic heat transfer models to reproduce the main operating features of this type of bioreactor. In the selected control loop, the average bed temperature was controlled by manipulating the inlet air temperature. The simulation results showed that, although the PI controller has the simpler structure, the GLC controller has the slightly better performance. In addition, this work showed that with controlling the average bed temperature instead of controlling the top-bed temperature, the whole temperature of the bed can be kept in the suitable range of *Aspergillus niger*.

Keywords: Solid-state fermentation; Packed-bed bioreactor; PI control; Nonlinear control

1. Introduction

Solid-state fermentation (SSF) is defined as the growth of microorganisms on moistened solid substrate, in which enough moisture is present to maintain microbial growth and metabolism, but there is no free-moving water and air is the continuous phase [1]. This process is particularly well adapted to the metabolism of fungi [2]. During the latest decade, SSF has experienced renewed interest due to many potential advantages of this bioprocess in comparison with submerged fermentation (SmF). These advantages include smaller bioreactor volumes, low cost media, less effort in downstream processing, superior productivity, reduced energy and cost requirements, simpler techniques and low wastewater output [3,4]. SSF has emerged as a potential technology for the production of microbial products such as feed, fuel, food, industrial chemicals and pharmaceutical products. Current trends on SSF have focused on application of SSF for the production of value-added products such as enzymes, organic acids, biopesticides and hormones [5].

Despite these potentials, very few of the many SSF processes that are studied in the laboratory make the transition to production scale. This is largely due to the difficulty in controlling the key process variables in large-scale bioreactors at the optimal values for growth and product formation. One of the most important variables is the bed temperature. In fact, the temperature of the bed is very critical in SSF. High temperatures affect spore germination, growth, product formation and sporulation, whereas low temperatures are not favorable for growth of the microorganisms and for the other biochemical reactions. The low moisture content and poor conductivity of the substrate make it difficult to achieve good heat transfer in SSF. In fact, limiting the heat dissipation is one of the major drawbacks of SSF in comparison with conventional SmF, where good mixing provides for efficient dispersal of

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sparged oxygen also serves to give better temperature control [6,7]. Note that, very few works have been carried out on the temperature control in packed-bed bioreactors.

Agitation and rotation in SSF were often carried out to improve mass and heat transfers, but the shearing force caused by agitation and rotation has adverse effects on medium porosity and disrupts fungal mycelia [4]. As a result, the substrate bed must be maintained static during the fermentation of such cases. In tray bioreactors reaching to high temperatures and lack of oxygen in the centre of trays is unavoidable. Although tray bioreactors can be used for such processes, packed-beds are more appropriate because the forced aeration allows some control over fermentation parameters through manipulation of the flow rate and the temperature of the air used in the fermentation [8]. In a static packed-bed, it is impractical to add water in a well-distributed manner to the bed during the fermentation. Therefore the packed-bed bioreactor should be operated so as to minimize drying of the bed, because drying can eventually lead to the moisture content in the bed reaching values which restrict the growth of the microorganism. This necessitates the use of saturated air at the air inlet. Note that, even with saturated inlet air, evaporation still occurs because the increase in air temperature between the air inlet and outlet increases the water-holding capacity of the air. These drying considerations remove manipulation of the relative humidity of the inlet air as an operating variable for packed-bed bioreactors, leaving only the inlet air temperature and superficial velocity [9,10].

The degree of control achievable by a simple control schemes is limited due to the complexity of the processes occurring within the SSF bioreactors [7]. Unfortunately, traditional control systems are not robust enough to ensure reproducible operation [11]. In fact, the application of control strategies to SSF bioreactors has proved to be a very difficult task, and is still far from a satisfactory solution [12,13]. Even at laboratory scale, the application of control schemes has been poorly explored, and the work that has been done has focused on continuously mixed bioreactors [14-16]. Due to the difficulties in heat removal and non-linear nature of SSF processes, temperature control is the major challenge in packed-bed bioreactors.

The aim of SSF bioreactor models is to describe how the performance is affected by the various operating variables that can be manipulated in an attempt to control the process, and how these system variables will affect microbial growth and product formation [9]. Hence, monitoring and control of SSF bioreactors require an effective model because an effective model can provide important insights into the control of SSF bioreactors that would be very hard and expensive to obtain experimentally. Although some authors have previously used simulations to show the advantage of using advanced control strategies in SSF reactors, they have used simple lumped models that did not represent the complexity of the real system [12,14].

Present study concentrates on temperature control for the growth of *Aspergillus niger* on wheat bran in a packed-bed bioreactor, and two single-input single-output (SISO) controller were used to control the average bed temperature namely PI and GLC controls. To this aim the kinetic and distributed dynamic heat transfer models were used to assess control performance via simulations that were completely investigated in the previous study since temperature changes and also the influence of temperature on the growth kinetics can be reasonably described by these models. Our modeling work showed that significant temperature gradients are unavoidable in SSF bioreactors of this type, and the temperature at the top of the bed can reach values that are deleterious to the microorganism.

In the following sections, firstly, the mathematical models are described briefly. Secondly, the controller structures are represented and finally performances of the control algorithm are compared through the simulation studies.



2. Mathematical model

2.1. Kinetic model

The growth of biomass occurs according to the logistic equation. This equation can almost give an adequate approximation of the whole growth curve in a single equation, including the lag, exponential growth and stationary phases, but it cannot provide a complete representation that includes the death phase [17,18]. Therefore, we modified the logistic equation as follows:

$$\frac{dX}{dt} = \mu\phi X \left(1 - \frac{X}{X_m}\right) \quad (1)$$

Where X , X_m and ϕ are the biomass concentration, the maximum possible biomass concentration and the level of a physiological factor, respectively. In fact, the physiological factor represents a dimensionless quantity related to the physiological state of the cells that not only plays a key role in the growth process, but also is responsible for its own synthesis. Thermal denaturation of this factor is expressed as a first-order process, and the rate of auto-synthesis of the factor is expressed according to the power-law version of the logistic equation as follows:

$$\frac{d\phi}{dt} = \gamma_s \phi (1 - \phi^\alpha) - \gamma_D \phi \quad (2)$$

Where α is the exponent in the power-law version of the logistic equation, and γ_s and γ_D are the rate coefficients of the synthesis and denaturation reactions, respectively. These rate coefficients are expressed as functions of temperature according to the Arrhenius equation:

$$\gamma_s = \gamma_{s_0} \exp\left(\frac{-E_s}{R(T + 273)}\right) \quad (3)$$

$$\gamma_D = \gamma_{D_0} \exp\left(\frac{-E_D}{R(T + 273)}\right) \quad (4)$$

Where γ_{s_0} and γ_{D_0} are the frequency factors for the synthesis and denaturation reactions, E_s and E_D are the activation energies for the synthesis and denaturation reactions, and T is the bed temperature, respectively. The specific growth rate constant is expressed empirically as a function of temperature [19]:

$$\mu = \left(\frac{s + (T_{\max} - T_{opt})}{(T_{\max} - T_{opt})}\right) \left(\frac{\mu_{opt} (T_{\max} - T)}{s + (T_{\max} - T)}\right) \quad (5)$$

Where μ_{opt} is the specific growth rate constant at the optimal temperature for growth (T_{opt}) and T_{\max} is the maximum temperature at which growth can occur. The parameter s describes the sensitivity of the specific growth rate to increases in temperature. The parameter values used in the kinetic model are given in Table 1 [3,19-22].



Table 1. Parameter values used in the simulations with the kinetic model

Symbol	Value
E_D	294516 J mol ⁻¹
E_S	68138 J mol ⁻¹
R	8.314 J mol ⁻¹ °C
s	6.275 (dimensionless)
t	$t_0 = 0$ hr
T_{opt}	35 °C
T_{max}	52 °C
X	$X_0 = 0.0071$ kg-biomass kg-substrate ⁻¹
X_m	0.22 kg-biomass kg-substrate ⁻¹
α	11 (dimensionless)
ϕ	$\phi_0 = 1$ (dimensionless)
γ_s	hr ⁻¹
γ_{s0}	9.761×10^8 hr ⁻¹
γ_D	hr ⁻¹
γ_{D0}	8.74×10^4 hr ⁻¹
μ	hr ⁻¹
μ_{opt}	0.173 hr ⁻¹

2.2. Dynamic heat transfer model

The packed-bed bioreactor modeled is cylindrical, and it consists of a bed of substrate supported on a perforated base plate through which saturated air with water vapor is blown. The moist wheat bran is inoculated with *Aspergillus niger* and placed in the bioreactor at time zero. Models concentrate on the heat transfer phenomena. Due to the increase in temperature as the air flows through the bed, heat removal phenomena are considered in the axial direction. Also the air and the moist solid at any particular location within the bed are assumed to be in thermal equilibrium.

2.2.1. Distributed dynamic heat transfer model

Sangsurasak and Mitchell [19] described the development of a model which was identical with our distributed dynamic heat transfer model except that it described heat transfer in both the vertical and horizontal directions. Distributed dynamic heat transfer model includes terms for the generation of heat from microbial growth, convective and evaporative heat removal, and axial conduction:

$$\rho_b C_{pb} \left(\frac{\partial T}{\partial t} \right) = \rho_s (1 - \varepsilon) Y_Q \frac{dX}{dt} - \rho_a C_{pa} V_z \left(\frac{\partial T}{\partial z} \right) - \rho_a f \lambda V_z \left(\frac{\partial T}{\partial z} \right) + k_b \left(\frac{\partial^2 T}{\partial z^2} \right) \quad (6)$$



Where C_{pb} , C_{pa} , ρ_b , ρ_a , ρ_s , Y_Q , V_z , ε , T_a , f , λ and k_b are heat capacity of the bed, heat capacity of the moist air, bed density, moist air density, substrate density, metabolic heat yield coefficient, superficial velocity of the moist air, void fraction, temperature of inlet air, water carrying capacity of air, latent heat of evaporation of water and thermal conductivity of the bed, respectively. In above equation each term has the units of $\text{J hr}^{-1} \text{m}^{-3}$. Air moves only in the axial direction, with a constant velocity profile across the bed. In Eq. (6), values for density, thermal conductivity and heat capacity of the bed were calculated as weighted averages of the properties of the air and substrate within the bed.

The parameter values for the distributed dynamic heat transfer model are given in Table 2 [20,22-26].

Table 2. Parameter values used in the simulations with lumped and distributed heat transfer models

Symbol	Value
C_{pa}	$1180 \text{ J kg}^{-1} \text{ }^\circ\text{C}^{-1}$
C_{pb}	$\text{J kg}^{-1} \text{ }^\circ\text{C}^{-1}$
C_{ps}	$2500 \text{ J kg}^{-1} \text{ }^\circ\text{C}^{-1}$
f	$0.00246 \text{ kg-water kg-air}^{-1} \text{ }^\circ\text{C}^{-1}$
k_a	$74.16 \text{ J hr}^{-1} \text{ m}^{-1} \text{ }^\circ\text{C}^{-1}$
k_b	$\text{J hr}^{-1} \text{ m}^{-1} \text{ }^\circ\text{C}^{-1}$
k_s	$1080 \text{ J hr}^{-1} \text{ m}^{-1} \text{ }^\circ\text{C}^{-1}$
T	$T_0=30 \text{ }^\circ\text{C}$
V_z	51 m hr^{-1}
Y_Q	$8.366 \times 10^6 \text{ J kg-biomass}^{-1}$
ε	0.35 (dimensionless)
λ	$2414300 \text{ J kg-water}^{-1}$
ρ_a	1.14 kg m^{-3}
ρ_b	kg m^{-3}
ρ_s	700 kgm^{-3}

C_{ps} , k_s are heat capacity and thermal conductivity of substrate. k_a is thermal conductivity of air.

3. Control algorithms

High performance of SSF processes can only be attained by the application of effective control strategies over the environmental conditions within the bioreactor [11]. The conventional techniques and concepts used for temperature control in SmF are not easily adaptable to SSF [7]. The aim of the current work is to test different control types and understand the influence of the different variables on the bioreactor performance. In this paper, two control algorithms are used: the conventional PI as a linear control and a globally linearizing control algorithm as a non-linear control. The average bed temperature was controlled by manipulating the inlet air temperature, based on temperature measurements made at various different bed heights. Temperature measurements can be made with sensors that can be inserted at different bed heights. Temperature measurements invariably contain noise, which may come from either variations in the process itself or from the sensors. As a result, Random noise was added to these cases to simulate the measurement noises that occur in the real process.



3.1. PI control

A velocity form of the PI algorithm [27] was used:

$$T_a(t) = T_a(t-1) + K_c \left(1 + \frac{t_s}{\tau_I} \right) e(t) - K_c e(t-1) \quad (7)$$

Where e is the controlled variable error, K_c is the proportional gain, τ_I is the integral time constant (hr) and t_s is the sample time (hr).

3.2. GLC control

The GLC method is a nonlinear control algorithm based on differential geometric approach. The first step in the GLC synthesis is the calculation of a state feedback, under which the closed loop input/output system is exactly linear. Then for linearized system, a controller with integral action such as PI can be designed.

To implement the state feedback of the GLC, all the process state variables should be measured or estimated. Open loop or closed loop observers such as extended Kalman filter can be used for estimation of unmeasured state variables. Consider SISO processes with the following model:

$$\begin{cases} \frac{dx}{dt} = f(x) + g(x)u \\ y = m(x) \end{cases} \quad (8)$$

With a finite relative order r (the relative order is the smallest integer for which, $L_g L_f^{r-1} m(x) \neq 0$). Here x is the vector of state variables, u and y are the manipulated input and the controlled output, respectively. Under the state feedback:

$$u = \frac{v - m(x) - \sum_{i=1}^r \beta_i L_f^i m(x)}{\beta_r L_g L_f^{r-1} m(x)} \quad (9)$$

Where, β_i 's are tunable parameters, the closed loop $v - y$ behavior is linear and described by the following equation:

$$v = y + \sum_{i=1}^r \beta_i \frac{d^i y}{dt^i} \quad (10)$$

Some guidelines for tuning of β_i 's parameters and other remarks for using GLC method are described by Soroush and Kravaris [28].

In the temperature control loop, the manipulated variable is the inlet air temperature (T_a) and the controlled output is the average bed temperature (T_{av}). The control law has the following form:



$$T_a = \frac{v - T_{av} - \frac{\beta_1}{n} \sum_{i=1}^n F_i(x)}{\beta_1 / n\tau} \quad (11)$$

In the above equation, n is the number of grid points and τ is the time constant assumed for inlet air temperature. Also the functions $F_i(x)$ are defined as below:

$$F_1(x) = -\frac{T_1}{\tau} \quad (12)$$

$$F_i(x) = \frac{dT_i}{dt} = \frac{\rho_s(1-\varepsilon)Y_Q}{\rho_b C_{pb}} \frac{dX}{dt} - \frac{1}{2h} \frac{\rho_a C_{pa} V_z}{\rho_b C_{pb}} (T_{i+1} - T_{i-1}) - \frac{1}{2h} \frac{\rho_a f \lambda V_z}{\rho_b C_{pb}} (T_{i+1} - T_{i-1}) + \frac{1}{h^2} \frac{k_b}{\rho_b C_{pb}} (T_{i+1} - 2T_i + T_{i-1}) \quad \text{for } i = 2, 3, \dots, n \quad (13)$$

A PI controller can be used to generate the input of the linearized system as below:

$$v(t) = v(t-1) + K_c \left(1 + \frac{t_s}{\tau_I} \right) e(t) - K_c e(t-1) \quad (14)$$

4. Results and discussion

In all simulations a discrete form of controllers with 0.025 hr sampling time was used. The value of τ and the set point for the average bed temperature were considered 0.035 hr and 33°C, respectively. The parameters of PI and GLC controls were tuned according to the trial and error method until satisfactory results are obtained. The resulted values of these parameters are given in [Table 3](#).

Table 3. Tunable parameters of PI and GLC

Controller	K_c	τ_I (hr)	β_1 (hr)
PI	0.4	0.1	---
GLC	0.4	0.1	0.1

In implementation of the above control algorithms assumed that the temperature at different bed heights can be measured online, but the values of X and ϕ can not be measured on-line. Therefore, in all simulations, an open loop observer is used for estimation of these values. In fact, controllers would use the model to compute the values of X and ϕ . It is necessary to note that the biomass concentration is unobservable from substrate concentration measurements and therefore a closed loop observer can not be used.

In GLC algorithm two cases are considered and then compared in order to compute the values of X and ϕ . In the first case, differential equations for temperature are inserted to the process model (open loop observer case). In the next case, the aforementioned differential equations are not inserted to the process model that this case is called "open loop reduced order observer". The results obtained from simulations of these cases were almost equal. As a result, the open loop reduced order observer is the better case than the other one, because in spite of simplicity, it does not depend on temperature differential equations.

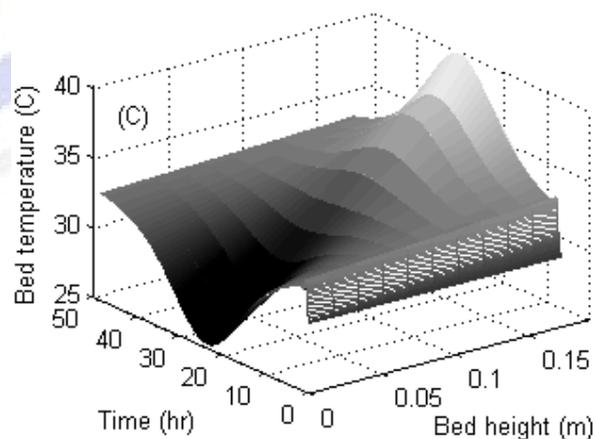
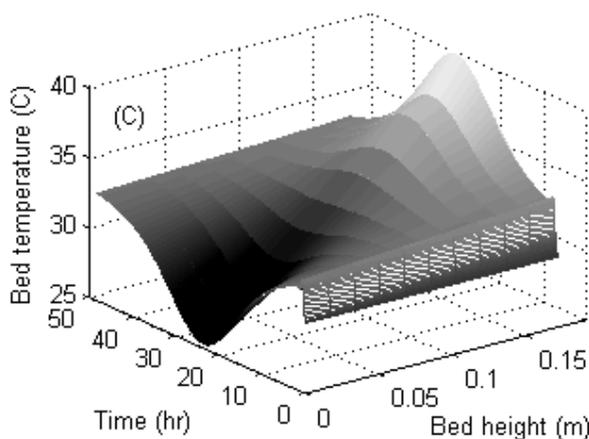
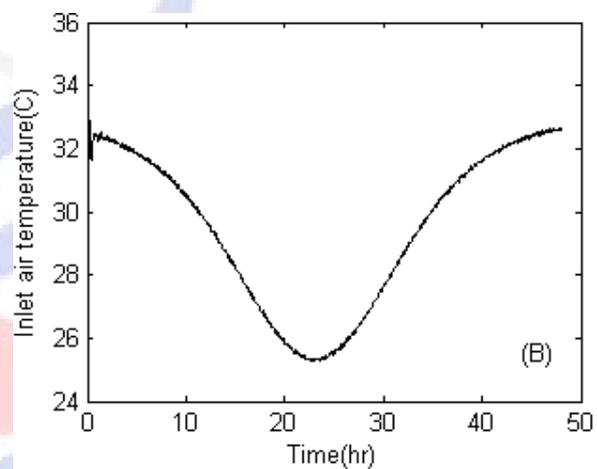
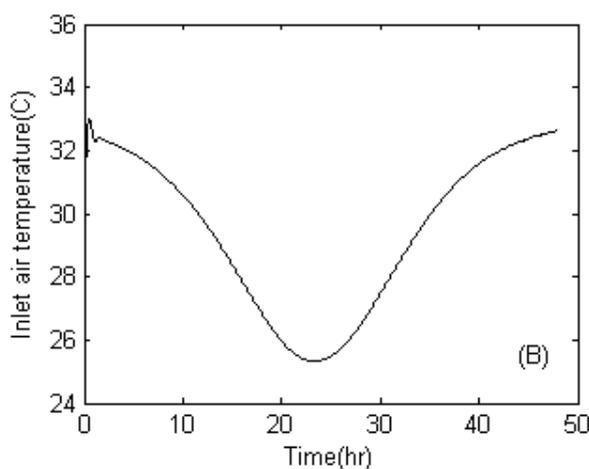
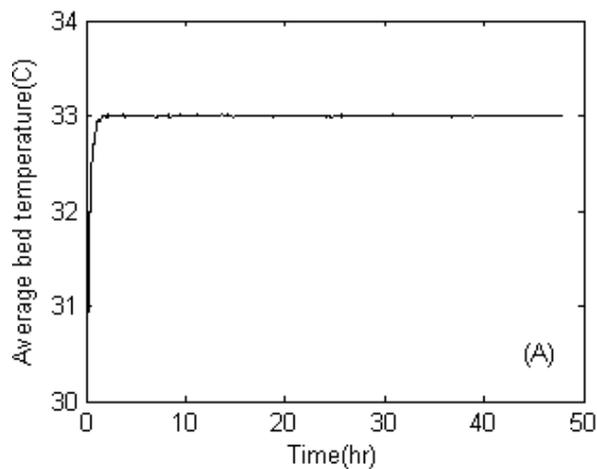
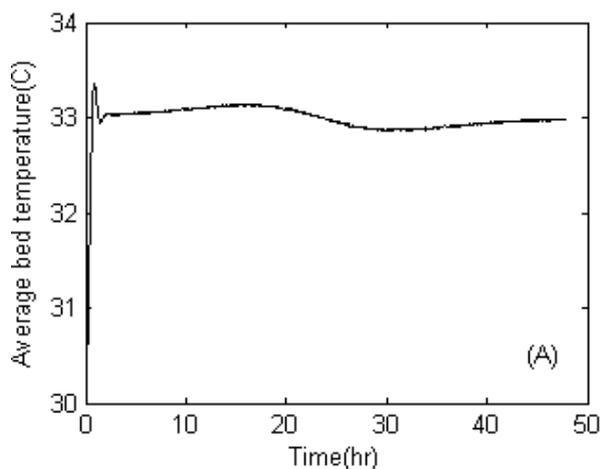


Fig. 1. (A) Resulting profile of the average bed temperature, (B) Inlet air temperature profile and (C) Bed temperature at different heights, for PI controller.

Fig. 2. (A) Resulting profile of the average bed temperature, (B) Inlet air temperature profile and (C) Bed temperature at different heights, for GLC controller.

A comparison of the PI and GLC controllers is shown in Figs. 1 and 2. It is clear that the GLC leads to better results than the PI control. Since the average bed temperature is accurately controlled at the set point by GLC controller. In the case of the PI control (Fig. 1A), there is



an oscillatory behavior in average bed temperature that is not observed in the GLC control (Fig. 2A).

Inlet air temperature can be set several degrees below the optimum temperature for growth, but there are practical limits on how low T_a can be. It must be sufficiently high to support reasonable growth rates, since the region near the base of the column will be maintained near T_a by the incoming air. To this point, the inlet air temperature was not allowed to be manipulated to values below 25°C in above control schemes as shown by Figs. 1B and 2B.

Reaching to a temperature 5°C above T_{opt} is undesirable for any part of the bioreactor. This temperature will be referred to as the critical temperature. The critical temperature for *Aspergillus niger* is 40°C. The temperature higher than critical temperature could trigger sporulation, or could have adverse effects on growth and product formation. In fact, these temperatures impose a limit on growth. The objective of the above control strategies was to keep the bed temperature below the critical temperature for the growth of *Aspergillus niger* throughout the entire fermentation. As Figs. 1C and 2C show, the bed temperature was controlled between 25 and 40°C that *Aspergillus niger* grows well over this temperature range.

To investigate the performance of controllers, a reduction at set point was done during the period of high heat generation, and the results are shown in Figs. 3 and 4. Again GLC control gave superior results than PI control.

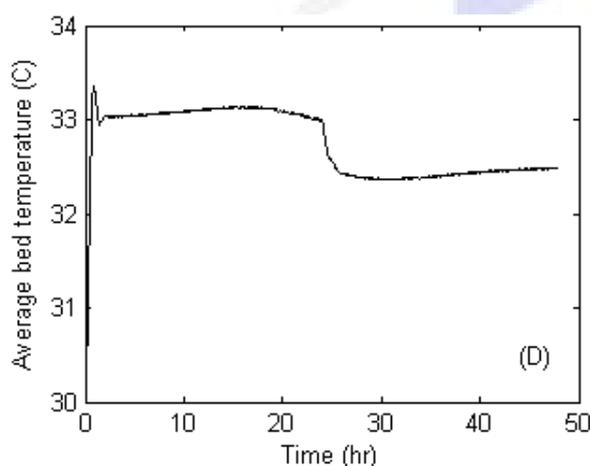


Fig. 3. Response of PI controller with 0.5°C setpoint change.

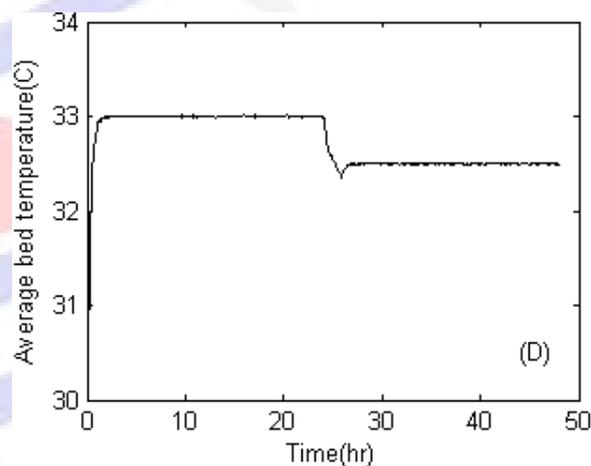


Fig. 4. Response of GLC controller with 0.5°C setpoint change.

Existence of error in the values of model parameters is unavoidable. To this point $\pm 20\%$ error was inserted to the model parameters, but the simulation results were not change significantly. Thus the GLC and PI controller is robust enough for temperature control in packed-bed bioreactors.

In operating a bioreactor, it is not possible simply to fix the conditions within the bioreactor at desired values; the only means available to influence these conditions is to manipulate the input operating variables. The operating variables, which can be manipulated to achieve this aim, depend on the design of the bioreactor. As described above in packed-bed bioreactors the only operating variables which can be manipulated are the temperature and superficial velocity of the inlet air [7,29]. In this study, we did not use the superficial air velocity as an operating variable because if the bed temperature is too low, then decreasing the superficial air velocity might cause to fall the oxygen concentration below the critical level that would adversely affect metabolic activity of the cells, and if the temperature of the substrate is high, increasing the superficial air velocity promotes cooling of the substrate, which is not



favorable for the growth of the organism [7]. On the other hand, greater superficial velocities in order to maintain bed temperatures within a given range lead to higher pressure drops, which are undesirable from both technical and economical points of view [30]. Therefore, practical limits on superficial velocity might be imposed by the maximum pressure drop through the bed with which the aeration system can cope. Furthermore, high pressure drops can reduce air flows, and favor the phenomenon of 'air-channelling', where vertical cracks appear within the bed, or between the bed and the bioreactor wall, with air flowing preferentially through these cracks. Pressure drop considerations might impose limits on the heights that can be used for packed-bed bioreactors [10].

5. Conclusion

Many control strategies can be assessed and good estimates of their parameters can be obtained by computer simulation. The current study has given insights into the control of packed-bed bioreactors for SSF that would have been very expensive and time consuming to undertake experimentally. In our work, two control types were tested namely PI and GLC controls. The results showed that the performance of GLC control is slightly better than PI control. In addition, this work showed that with controlling the average bed temperature instead of controlling the top-bed temperature, the whole temperature of the bed can be kept in the suitable range of *Aspergillus niger*. These investigations suggest that reasonable control of the bed temperature can be achieved by a GLC or PI algorithm and manipulating the inlet air temperature.

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