24 Models of Packed-Bed Bioreactors

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24.1 Introduction

This chapter provides two case studies to show how modeling work can provide insights into how to design and operate traditional packed-beds and Zymotis-type packed-beds. Various other mathematical modeling case studies have been undertaken with packed-bed bioreactors:

- Saucedo-Castaneda et al. (1990). A model of a thin packed-bed in which axial convection is not taken into account, rather only horizontal conduction through the walls.
- Gutierrez-Rojas et al. (1995). A mathematical model for solid-state fermentation of mycelial fungi on inert support.
- Hasan et al. (1998). Heat transfer simulation of solid-state fermentation in a packed-bed bioreactor.
- Oostra et al. (2000). A model of a packed-bed bioreactor was used as part of the process of bioreactor selection.
- Weber et al. (1999). A simplified material and energy balance approach for process development and scale-up of *Coniothyrium minitans* conidia production by solid-state cultivation in a packed-bed reactor.
- Weber et al. (2002). Validation of a model for process development and scale-up of packed-bed solid-state bioreactor.

The aim of this chapter is not to review these models. Readers with a deeper interest in modeling of packed-beds are recommended to find the original papers.

24.2 A Model of a Traditional Packed-Bed Bioreactor

The system modeled is a traditional packed-bed bioreactor, as shown in Fig. 24.1. It is assumed that the bioreactor is sufficiently wide such that heat transfer to the side walls is negligible, and therefore only heat transfer in the axial direction is included in the equations.

Note that this model is relatively limited. It is aimed only at predicting temperatures within the bed. It does not aim to describe what happens with water in the bed. As such, it is only useful for processes in which the substrate can undergo large decreases in water content with only minor changes in water activity, such that growth is never water-limited. In any case, such substrates must be used if strict packed-bed operation is to be used, that is, with absolutely no mixing events. Such a substrate is nutrient impregnated hemp, used by Weber et al. (1999).

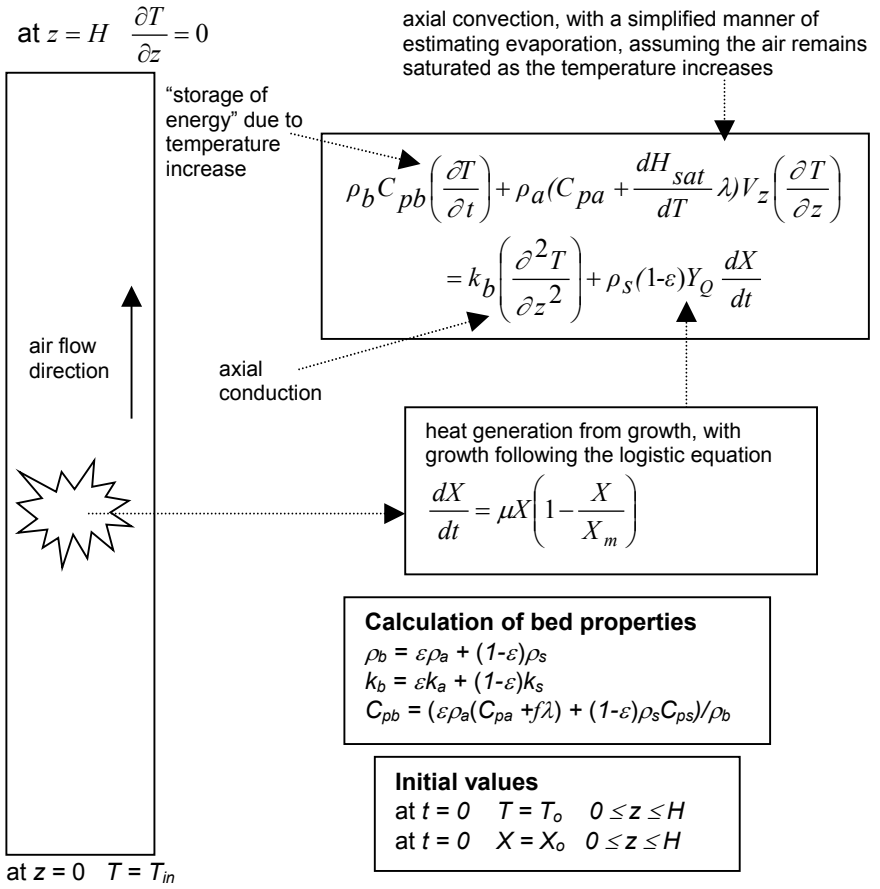

Fig. 24.1. Summary of the model used in the first packed-bed case study. The subscripts for the parameters ρ, k, and C_p are "*s*" for substrate particle, "*a*" for air, and "*b*" for the weighted average calculated for the bed

24.2.1 Synopsis of the Mathematical Model and its Solution

The model is based on the work of Sangsurasak (Sangsurasak and Mitchell 1995, 1998). It is almost identical to the version used by Mitchell et al. (1999), although the equation describing the effect of temperature on growth has been changed. The full model is not reproduced here; Fig. 24.1 summarizes its main features. Growth occurs according to the logistic equation, where the specific growth rate constant (μ) is affected by the temperature of the solid in a manner identical to that described by Eq. (22.1) (see Sect. 22.2). The energy balance takes into account axial convection, conduction, and evaporation and the production of metabolic heat. These two differential equations allow the biomass and temperature to be predicted as functions of time and space. The fact that the model is so simple means that it has many implicit assumptions and simplifications. Amongst these, some of the most important are:

- Growth depends only on biomass density and temperature. The bed does not dry out sufficiently during the fermentation to limit the growth;
- The bed is treated as a single pseudo-homogeneous phase that has the average properties of the gas and solid phases;
- Biomass does not move in space;
- Growth does not affect the void fraction;
- The substrate bed properties do not change with temperature or during consumption of substrate and production of biomass;
- Flow phenomena arising from increased pressure drop are not important;
- The air is always in thermal and moisture equilibrium with the solid (i.e., as the air heats up as it passes through the bed, water evaporates from the solid to maintain saturation of the air).

Table 24.1. gives the values of parameters, initial values of state variables and values of operating variables used in the base-case simulation. Note that some parameters are considered as constants even though they are not truly constant. For example the heat capacity of the air will change as the air passes through the bed, due to the increase in water content.

The mathematical model summarized in Fig. 24.1 contains partial differential equations. This model is solved by application of orthogonal collocation to convert each partial differential equation into a set of ordinary differential equations, which can then be integrated numerically. The principles of orthogonal collocation are beyond the scope of this book.

Note that the differential term dH_{sat}/dT in the energy balance is given by Eq. (19.20) (see Sect. 19.4.1). It has been used to replace the constant value of 0.00246 kg-H_2O kg-dry-air $°C^{-1}$ used by Mitchell et al. (1999).

Table 24.1. Values of the parameters and variables used for the base case simulation

Symbol	Significance	Base case value and units[a,b]
Design and operating variables (can be varied in the input file for the model)		
T_o	Initial bed temperature*	38°C
T_{in}	Inlet air temperature*	38°C
K	Control factor for the inlet air temperature	1 (see Eq. (24.1))
V_Z	Superficial velocity of the air*	5 cm s^{-1}
H	Height of the bioreactor*	1.0 m
Microbial parameters that you can vary		
X_o	Initial biomass concentration	0.001 kg-biomass kg-substrate^{-1}
X_m	Maximum possible biomass concentration	0.125 kg-biomass kg-substrate^{-1}
μ_{opt}	Maximum possible value of μ (at $T=T_{opt}$)	0.236 h^{-1}
T_{opt}	Optimum temperature for growth	38°C
Parameters related to the effect of temperature on growth (cannot be varied)		
A_1	Constant in the equation describing $\mu=f(T)$	8.31×10^{11}
A_2	Constant in the equation describing $\mu=f(T)$	70225 J mol^{-1}
A_3	Constant in the equation describing $\mu=f(T)$	1.3×10^{47}
A_4	Constant in the equation describing $\mu=f(T)$	283356 J mol^{-1}
Other parameters and constants (cannot be varied)		
C_{pa}	Heat capacity of the air	1180 J kg^{-1} °C $^{-1}$
C_{ps}	Heat capacity of the substrate particles	2500 J kg^{-1} °C $^{-1}$
k_a	Thermal conductivity of the air phase	0.0206 W m^{-1} °C^{-1}
k_s	Thermal conductivity of the substrate	0.3 W m^{-1} °C^{-1}
R	Universal gas constant	8.314 J mol^{-1} °C^{-1}
Y_Q	Yield of metabolic heat from growth	8.366×10^6 J kg-dry-biomass^{-1}
ε	Void fraction in the bed	0.35
λ	Enthalpy of vaporization of water	2.414×10^6 J kg-water^{-1}
ρ_a	Density of the air phase	1.14 kg-dry-air m^{-3}
ρ_s	Density of the solid particles*	700 kg-substrate m^{-3}

[a] Note that those values highlighted with an asterisk must be supplied in the input file. The remaining values are already in the program and cannot be changed.

[b] The program converts all variables and parameters to a consistent set of units.

24.2.2 Base-Case Predictions

Figures 24.2 and 24.3 represent the type of information that such a model provides. Both axial and temporal temperature variations occur (Fig. 24.2). The axial temperature profile is a result of convective cooling (see Fig. 4.3), with the steepness of this profile depending on the rate of growth and the superficial air velocity.

Note that the temperature variations in time are relatively slow compared to the temperature variations along the length of the bed. For example, in Fig. 24.2(b), the difference in temperature between 10 and 11 h is much smaller than the temperature difference between the inlet and outlet of the bed. This has implications for understanding the behavior of intermittently stirred beds (see Chap. 25).

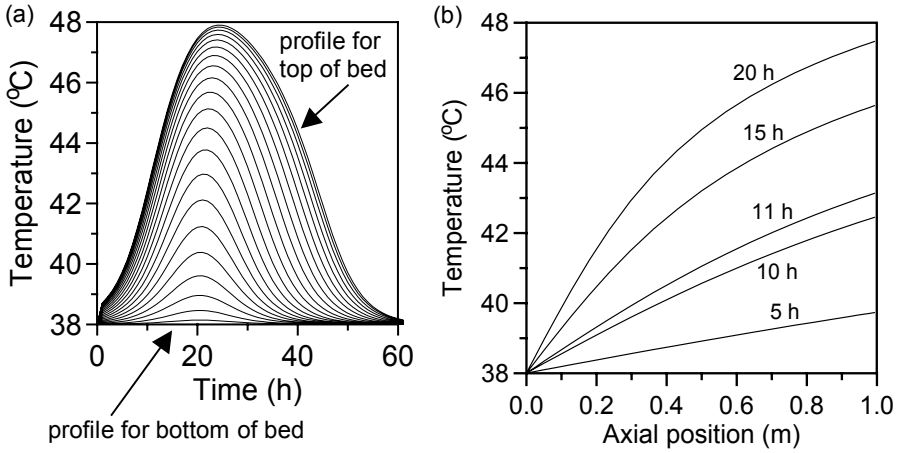

Fig. 24.2 (a) Typical output of the model, showing axial and temporal temperature profiles. Each curve represents the temperature profile against time at a particular location in the bioreactor. Moving upwards, the curves are for greater and greater heights in the bed. **(b)** The same data as shown in Fig. 24.2(a), but plotted against axial distance at several different times, in order to highlight the spatial temperature gradients. After 21 h the axial temperature profile begins to decrease again

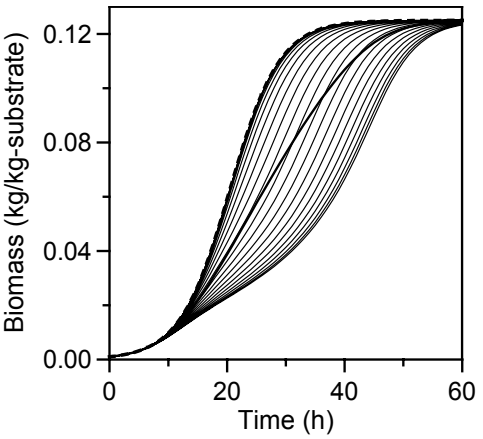

Fig. 24.3. Predicted growth profiles at different heights within the bed. The lines at the top of the graph represent the bottom of the bed. The lines at the bottom represent the top of the bed, which is much hotter and therefore causes growth to be slower. The dashed line at the left represents logistic growth with $\mu=\mu_{opt}$ throughout the growth phase. The darker solid line in the middle of the lighter lines represents the average biomass concentration in the bed

Note the similarity of the general appearance of Fig. 24.2(a) to the experimental results obtained by Weber et al. (2002) in Fig. 7.6(a). Of course, the temperatures and times are different because the organism simulated by the model is quite different from that which they used. One feature that is common to the two graphs is the lack of symmetry around the peak, that is, the decrease in temperature takes slightly longer than the initial rise in temperature.

As a result of the spatial temperature profiles, growth will occur at different rates in the different regions of the bed (Fig. 24.3).

24.2.3 Insights that Modeling Has Given into Optimal Design and Operation of Traditional Packed-Beds

Section 7.2 showed that the design and operational variables that can be manipulated for traditional packed-bed bioreactors are the inlet air temperature and flow rate, the presence of a water jacket and the temperature of the water in this jacket, and the height and width of the bioreactor. The model can be used to investigate the effect of some of these design and operating variables on bioreactor performance. It cannot describe the effect of the bioreactor width or of a water jacket since it does not describe heat removal by conduction normal to the direction of air flow.

In the simulations presented in this section, in which only the effect of temperature on growth is considered, the aim is to minimize temperature gradients in order to maintain the average temperature as close as possible to the optimum temperature for growth and product formation. The effects of bed height, aeration rate, and air temperature are interrelated, but, in the subsections that follow, one-at-a-time changes will be made in order to make the contribution of each individual variable clear.

Effect of inlet air temperature. The inlet air temperature can be reduced below the optimum for growth in order to combat the temperature rise that occurs within the bed. However, it is important not to maintain the air temperature at a constant low value during the fermentation. During the initial stages of the fermentation the air temperature must remain near the optimum in order not to retard the initial growth. Therefore, in the simulations shown in Fig. 24.4, a simple temperature control scheme was included in the model:

$$T_{in} = T_{opt} - K(T_{ou} - T_{opt}), \tag{24.1}$$

where K is a factor that determines by how much the temperature of the inlet air (T_{in}) is decreased for a given rise in the outlet air temperature (T_{out}) above the optimum temperature for growth (T_{opt}).

This strategy causes the temperature in the bed to vary around the optimum for growth (Fig. 24.4(a)). At the time of maximum heat production, the axial temperature profile is actually steeper than for aeration with the inlet air at T_{opt}: Fig. 24.4(a) shows a difference of almost 15°C between the air inlet and outlet, com-

pared to 10°C in Fig. 24.2(a). However, the maximum deviation from the optimum temperature for growth (38°C) is only 7.5°C, and the average deviation from T_{opt} is also smaller, because the axial gradient straddles the optimum temperature. This leads to corresponding predictions of better growth (Fig. 24.4(b)).

Effect of inlet airflow rate. Doubling the airflow rate (i.e., increasing the superficial velocity of the air from 0.05 to 0.1 m s^{-1}) in the absence of any control of the air temperature, the performance of the bioreactor is predicted to improve significantly. Increasing the air flow rate decreases the gradient of the axial temperature profile: the highest temperature reached decreases from 48°C (Fig. 24.2(a)) to 45°C (Fig. 24.5(a)) and, as a result, the growth profiles in the different regions of the bed are closer to the optimum profile (Fig. 24.5(b)).

No work has been done to investigate the upper limits on the superficial velocities that can be used in packed-bed bioreactors, and this model does not take pressure drops into account. Obviously, the higher the airflow rate, the greater the operating cost, not only because more air must be supplied, but also because the pressure drop is greater. Therefore an economic optimum will need to be found between improved packed-bed performance and increasing operating costs. The best strategy might be to increase the air flow rate only during the period of peak heat production. In this case fluidization of the particles in the bed will not be a problem, because the microorganism will bind the particles together before high air flow rates are used. However, it is possible for the pressure drop to be sufficiently high that the whole knitted bed is ejected from the bioreactor!

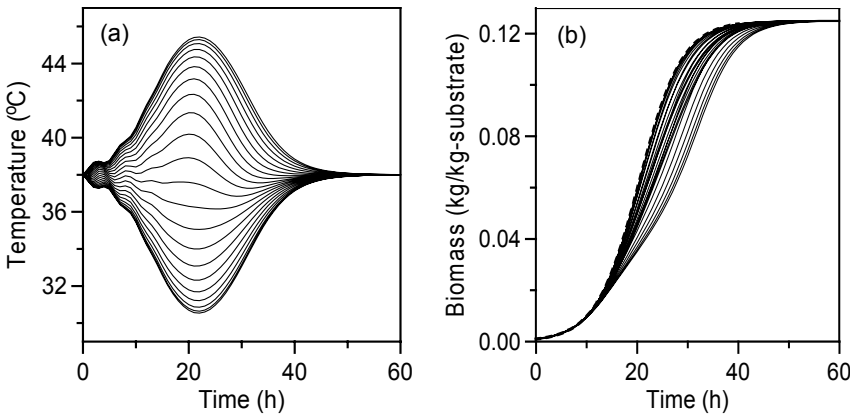

Fig. 24.4. (a) Predicted temperatures within the bed for a bioreactor in which the inlet air temperature (bottom-most line) is reduced in response to the temperature at the bed outlet (topmost line) according to Eq. (24.1). The other operating conditions are as in Table 24.1. Each curve represents the temperature profile against time at a particular location in the bioreactor. Moving upwards, the curves are for greater and greater heights in the bed. **(b)** Predicted growth profiles at different regions in the bed. Compared with Fig. 24.3, the average growth rate in the bed is much closer to the maximum possible rate

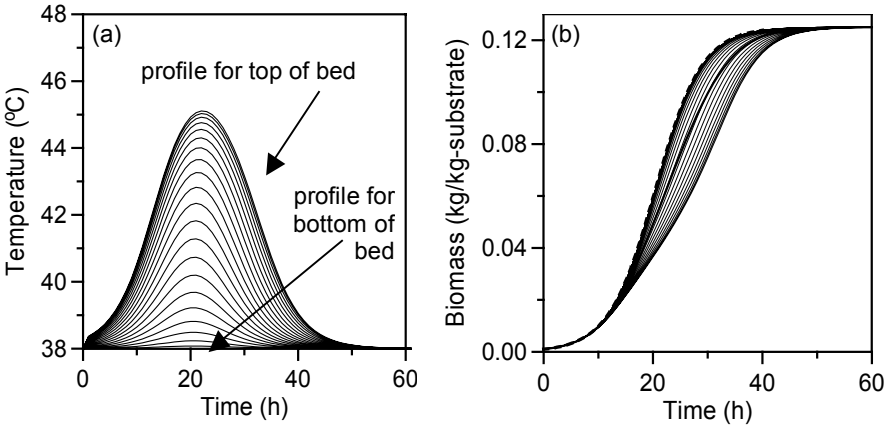

Fig. 24.5. (a) Temperature as a function of time and position for a superficial velocity (V_Z) of 0.1 m s^{-1}, twice as high as the value used to produce Fig. 24.2(a). All other parameters are the same as in Table 24.1. For ease of comparison, the same Y-axis range has been used as was used in Fig. 24.2(a). **(b)** Growth as a function of position for a superficial velocity (V_Z) of 0.1 m s^{-1}

Effect of bioreactor height and fungal specific growth rate. Obviously, with all operating conditions held constant, the height of the bioreactor will affect the maximum temperature reached, due to the unavoidable presence of an axial temperature gradient. In turn, this will affect the performance of the bioreactor. Figure 24.6 shows how the bioreactor height affects the time for the average biomass concentration (i.e., averaged over the whole bed) to reach 90% of the maximum biomass concentration. This time is denoted as t_{90}: the larger the value of t_{90}, the poorer the performance of the fermentation. This criterion is used to compare bioreactors since, for logistic growth kinetics, over a wide range of microbial and system parameters, the productivity of the fermentation, in terms of g-biomass m^{-3}-bioreactor h^{-1}, reaches a maximum when the biomass reaches around 90% of its final value (Mitchell et al. 2002b). In fact, t_{90} is inversely proportional to the productivity. The simulations were done for different specific growth rates.

The value of t_{90} increases approximately linearly with bioreactor height. This occurs because the greater the height, the greater the average deviation of the temperature from the optimum for growth. The value for zero height is the time that it would take for the biomass to reach $0.9X_m$ if the whole of the bioreactor remained at the optimum temperature for growth throughout the entire period.

Summary of strategies for optimizing the operation of traditional packed beds. The previous sections involved only one-by-one changes of variables. Obviously it is possible for more than one variable to be changed simultaneously. Simulations will not be shown for simultaneous changes (readers can use the program supplied with this book to undertake their own explorations), but, in general terms, to improve the performance of a traditional packed-bed, it is necessary to

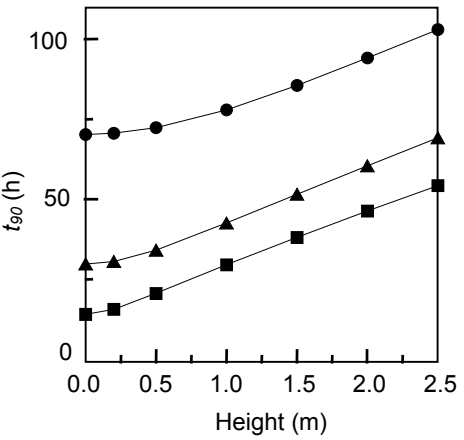

Fig. 24.6. The time taken for the average biomass concentration to reach 90% of X_m (t_{90}), as a function of the bed height within the packed-bed bioreactor, shown for various values for the parameter μ_{opt}. Key: (●) μ_{opt} = 0.1 h^{-1} (▲) μ_{opt} = 0.236 h^{-1} (■) μ_{opt} = 0.5 h^{-1}

decrease the height of the bed within the bioreactor, increase the superficial velocity and use a control system to reduce the temperature of the inlet air in response to temperature increases in the outlet air. Note that, in order to minimize operating costs, it would be preferable not to have to refrigerate the inlet air.

The implications of changes in bed height might need to be considered either at the bioreactor design stage or in an attempt to optimize the performance of a bioreactor that has already been built. At the design stage, decreasing the height while maintaining the superficial velocity constant means that the bioreactor will need to be wider to hold the same amount of substrate, occupying more floor space. Once a bioreactor is built, decreasing the height will mean that the unutilized volume within the bioreactor will increase and therefore the volumetric productivity of the bioreactor will fall, if the calculation is based on the total bioreactor volume and not the bed volume. Therefore, provided the aeration system has the capacity, it would be preferable to increase the superficial velocity than to decrease the bed height, although problems may occur with high pressure drops.

A model similar to the one used in the simulations above was used by Ashley et al. (1999) to investigate whether reversing the airflow direction would help to overcome the problem of overheating at the top of the bed. Figure 24.7 shows that the model predicts that indeed such a strategy will prevent the temperature at the ends of the beds from reaching deleteriously high values. Unfortunately, it is not a useful strategy since the cooling of the middle sections of the bed is very inefficient, allowing them to reach very high temperatures.

Insights into scale-up of traditional packed-beds gained from modeling work. If you are considering using a traditional packed-bed bioreactor due to the inability of your microorganism to tolerate agitation, then, on the basis of the results in

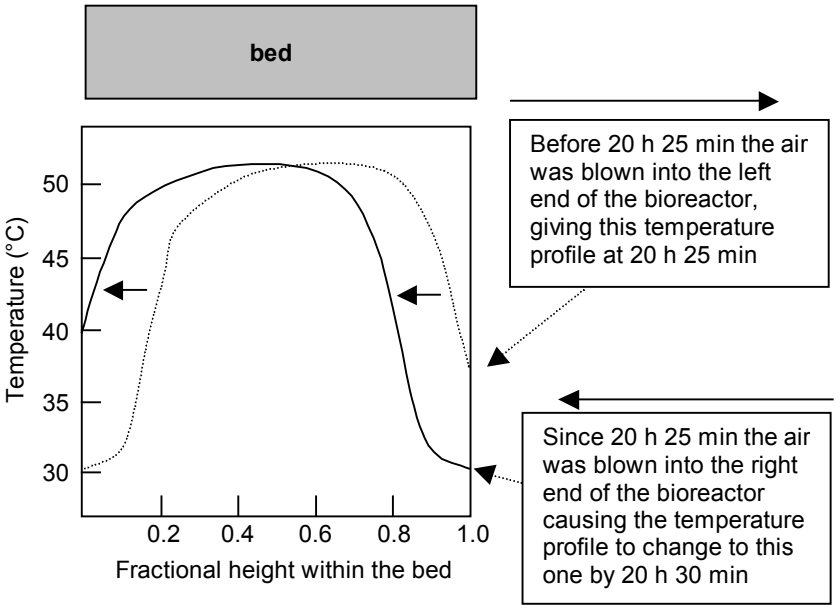

Fig. 24.7. Predicted axial temperature profiles at the time of peak heat production, for a fermentation in a packed-bed, in which there is no heat removal through the side walls and for which the direction of the air flow is reversed every five minutes. The inlet air temperature is 30°C. The bed height is 0.345 m and the superficial air velocity is 2.36 cm s^{-1}. This figure predicts that the temperature at the ends of the beds never exceeds 40°C, however, there is essentially no cooling effect in the central regions of the bed, which reach temperatures of more than 50°C. Adapted from Ashley et al. (1999) with kind permission of Elsevier

the previous section, it is possible to give advice about the research and development program for scale-up.

After the preliminary kinetic investigation in Raimbault columns (Sect. 15.1), experiments should be done in a pilot-scale packed-bed. A reasonable scale would be of the order of 15 cm diameter and as much as 1 m height. The walls of the column should be insulated well, in order to mimic the situation in the large-scale bioreactor, in which radial heat removal will be relatively minor. The 1 m height will allow studies to be done at bed heights that might actually be used in large-scale bioreactors. As such, this pilot bioreactor will represent a vertical section of the full-scale bioreactor (Fig. 24.8). This enables a study of those phenomena that depend on bed height, such as axial temperature profiles and pressure drops, and biomass and product formation as functions of height, and how these are affected by the temperature and velocity of the inlet air.

The advantage of this approach is that you might identify limitations on performance that are not predicted by the mathematical model. For example, the pressure drop may be excessive in your particular system. It is better to identify such problems, and to modify the bioreactor to overcome them, in a pilot-scale bioreac-

tor than it is to build a full-scale bioreactor only to find that it does not work properly.

In the scale-up process, there will be a limit on bed height, in the sense that very tall beds will lead to unacceptably poor performance, due to axial temperature gradients or other considerations. Once this limit is reached, the capacity of the bioreactor can only be increased by making the bed wider. This "critical height" is not a constant, since it depends on the growth rate of the organism and the operating conditions, especially the superficial velocity of the air. An estimate of the "critical height" of a traditional packed-bed can be calculated for logistic growth kinetics as (Mitchell et al. 1999):

$$H = \frac{\rho_a(C_{pa} + f)V_z(T_{out} - T_{in})}{0.25\rho_s(1-\varepsilon)Y\mu_{opt}X_m}, \tag{24.2}$$

where the symbols have the meanings given in Table 24.1. T_{out} is the maximum temperature allowable in the bed while f is an estimate of dH_{sat}/dT.

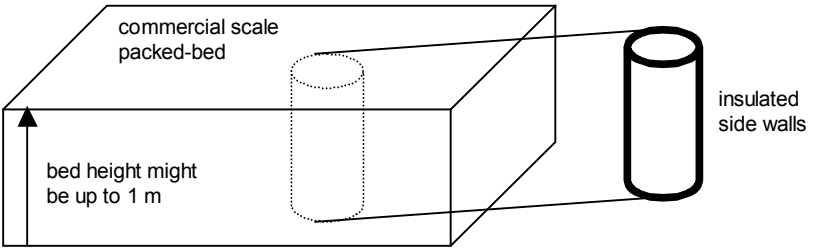

Fig. 24.8. A relatively thin packed-bed can be used for pilot-scale investigations into packed bed design. If its sides are insulated, this will mimic the presence of "hot" substrate around an identical section within the bed of the commercial-scale bioreactor

24.3 A Model of the Zymotis Packed-Bed Bioreactor

The model described here is based on that developed for the Zymotis bioreactor of Roussos et al. (1993) by Mitchell and von Meien (2000). The version used here has been modified by the inclusion of a water balance.

24.3.1 The Model

The model of the Zymotis packed-bed must account for heat transfer in two directions in the substrate bed: (1) the direction that is co-linear with the air flow, which causes convective and evaporative heat removal; and (2) the horizontal conduction to the cooling plates, which is normal to the direction of the air flow. Typically front-to-back gradients will be negligible (Fig. 24.9).

In this model the same growth kinetic equations are used as described by Eqs. (22.1), (22.2), and (22.3) (see Sect. 22.2). The solids and gas phases are assumed to remain in thermal and moisture equilibrium with one another. In other words, the gas phase remains saturated at the temperature of the bed. The differential term dH_{sat}/dT in the energy balance is given by Eq. (19.20) (see Sect. 19.4.1).

The parameter values used in the base-case simulation are the same as those given in Table 24.1 for the traditional packed-bed. There are several extra parameters that do not appear in that model, all of which are associated with the heat transfer plates. The spacing between plates (L) was varied, the overall heat transfer coefficient for heat transfer from the edge of the substrate bed across the plate wall to the cooling water (h) was taken as 95 W m^{-2} $^oC^{-1}$ a value that is typical of heat exchangers and, finally, the cooling water temperature (T_w) was set at 38°C in various simulations and varied according to a control scheme in others.

24.3.2 Insights into Optimal Design and Operation of Zymotis Packed-Beds

The model can be used to explore the effect of operating conditions on bioreactor performance. Simulations will not be shown for the effects of superficial air velocity, inlet air temperature, or bioreactor height. The general principles are the same as for the traditional packed-bed discussed in Sect. 24.2, although the effects are not exactly the same, because of the extra heat removal by the heat transfer plates. These parameters are therefore discussed generally, without new simulations being done. Readers with greater interest are encouraged to consult Mitchell and von Meien (2000) and also to use the simulation program provided to explore the performance of Zymotis packed-bed bioreactors in more detail.

After this, the effects of the new design and operating variables introduced by the internal heat transfer plates, namely the spacing between the plates and the temperature of the cooling water, are explored.

General principles (overall trends). The model predictions in Fig. 24.10(a)). show clearly the temperature gradients vertically through the bed (i.e., parallel to the direction of air flow) and horizontally (i.e., normal to the direction of air flow). Figure 24.10(b) compares the central axis temperatures for the Zymotis bioreactor and a traditional packed-bed that is wide enough for heat removal through the side walls to be negligible. The comparison is done for the same microorganism, that is, in both cases μ_{opt} is set at 0.236 h^{-1}, and for the same operating conditions. Along the central axis of the bioreactor, due to the greater heat removal rate in the Zymotis bioreactor, the temperature does not reach such high values in the top half of the bioreactor, although the performance is reasonably similar in the bottom half. Note, however, that for the Zymotis bioreactor the curve represents only the central axis temperature, while for the traditional packed-bed it represents the temperature at all radial positions for that height. In the Zymotis packed-bed the remainder of the bed is cooler than the central axis at the corresponding height, and therefore growth is correspondingly better.

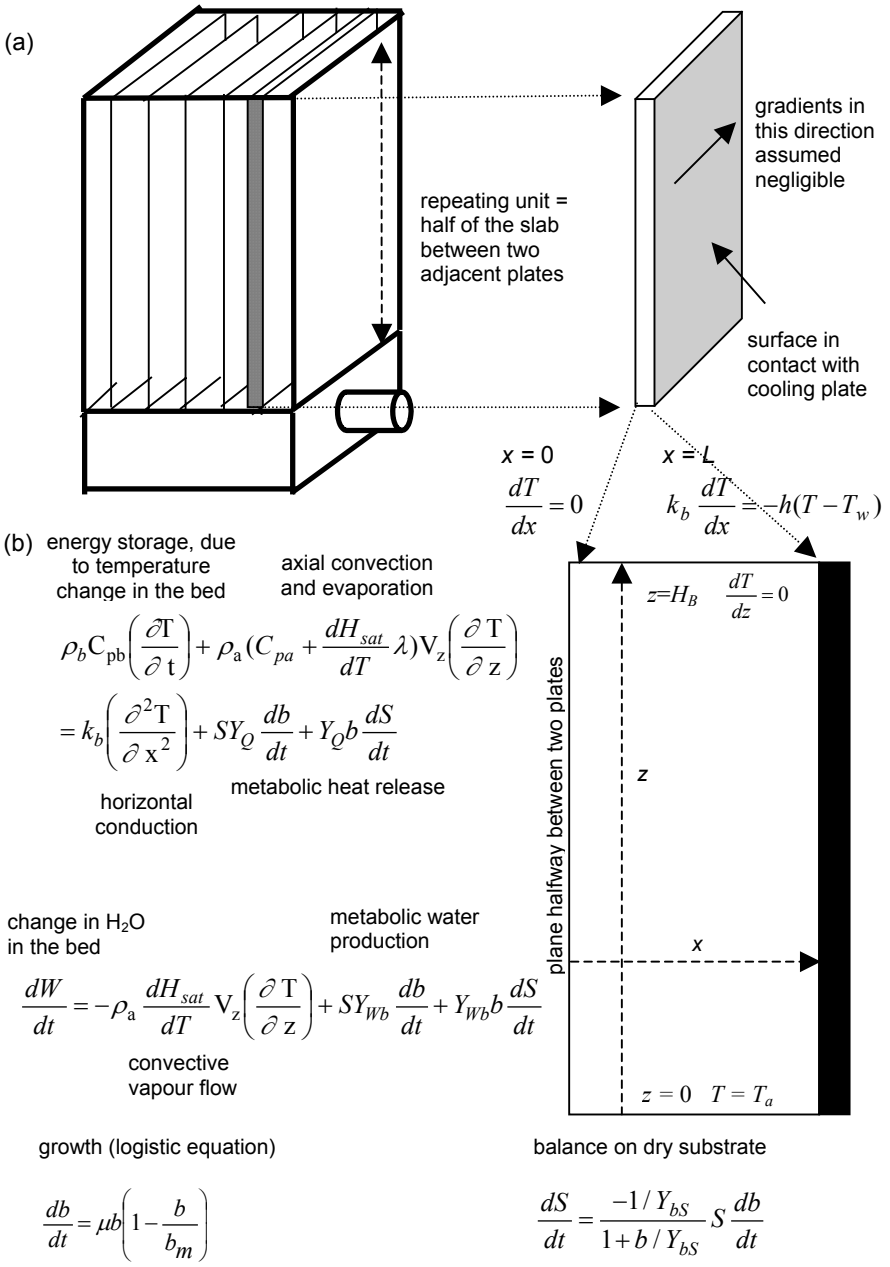

Fig. 24.9. Summary of the model of the Zymotis packed-bed bioreactor used in the second case study. **(a)** The Zymotis bioreactor can be treated as consisting of repeating units. **(b)** Summary of the mathematical model used to model one of the repeating units

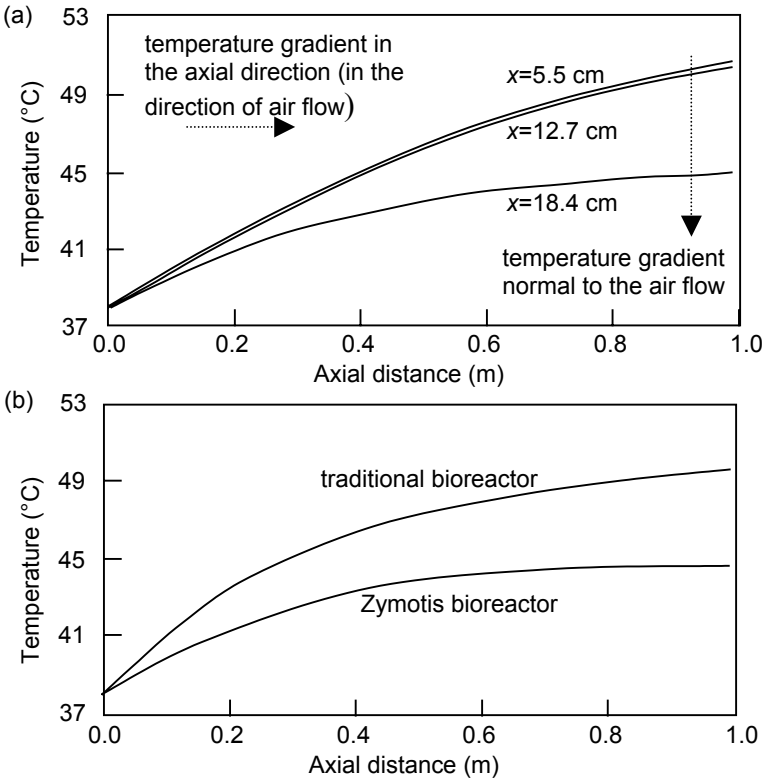

Fig. 24.10. (a) Predictions of the mathematical model about the axial temperature profile at several horizontal positions within the Zymotis packed-bed at the time of peak heat production at 23 h. Note that in the upper regions of the bed the temperature is high from the central plane to 12.7 cm from the central plane (7.3 cm from the plate), only reducing significantly at positions close to the plate (which is at x=20 cm). **(b)** Comparison of the central axial temperature profile in a wide traditional packed-bed and the central plane temperature profile of the Zymotis bioreactor, at 20 h, in the case in which L = 0.03 cm. In both cases the bioreactor is 1.0 m high, the superficial air velocity is 5 cm s^{-1}, the inlet air temperature is 38°C and the specific growth rate constant (μ_{opt}) is 0.236 h^{-1}

Regarding bioreactor dimensions, the maximum practical value for the front-to-back depth will depend on the size of heat transfer plates that can be constructed. Given that more plates can be added to extend the width of the bioreactor, theoretically there is no limitation on the width. One advantage of the Zymotis bioreactor is that, with the flattening out of the temperature profile, larger heights are theoretically possible than for traditional bioreactors, at least based on temperature considerations, although pressure drop may become problematic at large heights.

In the upper regions of the bed much of the waste metabolic heat is removed by conduction to the heat transfer plates, as evidenced by the flattening out of the temperature profile (Fig. 24.10(b)); in these regions only a relatively minor proportion is removed by axial convection and evaporation. Note that the predicted profiles are similar to the experimental temperature profiles that Saucedo-Castaneda et al. (1990) measured in a water-jacketed packed-bed bioreactor of 6 cm diameter (Fig. 7.5). The position at which the axial temperature profile flattens out in the Zymotis bioreactor depends on the heat production rate and on the various operating parameters, including the superficial air velocity. In fact, due to the acceleration and deceleration in the growth rate during the various phases of a fermentation, and the corresponding changes in the rate of production of waste metabolic heat, the extent of the "flat zone" will fluctuate during the fermentation.

Effect of a cooling water control scheme and of the gap between the plates. In these simulations only data about overall predicted performance is presented; no information is given about the gradients within the bioreactor. However, clearly the best performance must correspond to those operating conditions that minimize the temperature deviations from the optimum temperature, in both space and time.

The spacing between the cooling plates makes a large difference to the predicted performance, with all other parameters and operating variables being held constant (Fig. 24.11). Performance worsens rapidly as the size of the gap increases from 1 to 10 cm. That is, t_{90} increases rapidly over this range. Further increases in gap size worsen performance even further.

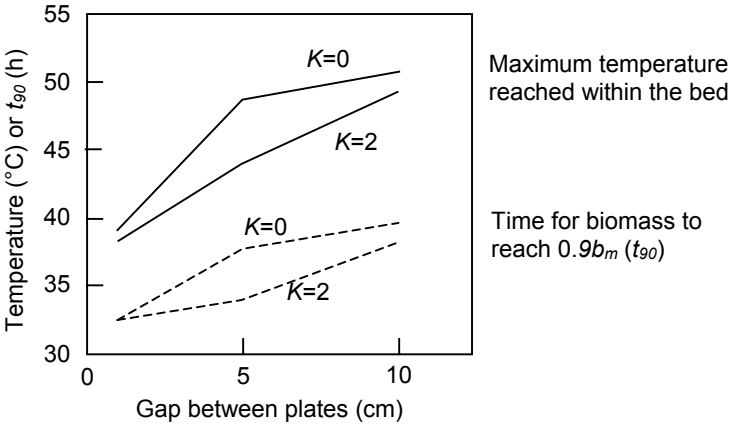

Fig. 24.11. Predictions of the mathematical model about the combined effect on the performance of the Zymotis bioreactor of the gap between the heat transfer plates (the gap is equal to $2L$) and the use of a scheme for the control of the cooling-water temperature according to Eq. (24.3). In this graph the performance of the bioreactor is evaluated on the basis of two criteria: (——) the maximum temperature reached within the bed during the fermentation and (- - - -) the time for the biomass to reach 90% of its maximum value (t_{90})

Two lines are shown on the graph. One corresponds to the case in which the cooling water temperature is maintained at 38°C (K=0). In the other case (K=2) the temperature of the cooling water (T_w) is manipulated in response to the temperature measured at the top of the bed, halfway between the heat transfer plates (this temperature being denoted T^*), according to the following equation:

$$T_w = T_{opt} - K\,(T^* - T_{opt}), \tag{24.3}$$

where T_{opt} is the optimum temperature for growth. This equation calculates the number of degrees by which the measured temperature exceeds the optimum for growth and then decreases the temperature of the cooling water by this temperature difference multiplied by a factor K. If possible, the value of K should be chosen so as not to require refrigeration of the cooling water to values below the temperature at which it is normally available. This will avoid the costs of building and operating a water refrigeration system. However, the ability to do this will depend on the optimum growth temperature of the organism in relation to the temperature of the available cooling water.

Whether or not it is advantageous to use this strategy to control the cooling water temperature depends on the spacing between the plates. If the spacing between the plates is large, of the order of 20 cm (L=10 cm), the cooling water has relatively little effect on much of the bed, and therefore the temperature control scheme brings little advantage (Fig. 24.11). If the spacing between the plates is small, of the order of 2 cm (L=1 cm), temperature control is reasonably efficient even without the temperature control scheme, so there is little advantage in having it. The temperature control scheme is most advantageous at intermediate plate spacings.

In fact, intermediate plate spacings are probably preferable. Although a 2 cm gap between plates gives near optimum performance (the minimum possible value of t_{90} for X_o=0.001 kg kg^{-1}, X_m = 0.125 kg kg^{-1}, and μ_{opt} = 0.236 h^{-1} is 29.7 h), it is not a reasonable value, because a significant volume of the bioreactor will be occupied by the plates, leading to a low overall productivity and, additionally, the capital costs of the bioreactor will be much higher. On the other hand, the wider the space between the plates, the less effective they are in cooling the bed, and therefore the higher is the superficial air velocity that is needed to achieve the same cooling effect and, consequently, the higher are the operating costs. Essentially these will need to be balanced against each other. Mitchell et al. (2002b) use a model similar to the one presented here to explore these issues in more depth, identifying a gap of 6 cm (L=3 cm) as optimal in terms of productivity of the bioreactor, calculated per m^3 of overall bioreactor volume, for a microorganism with a specific growth rate of 0.324 h^{-1} and a superficial air velocity of 1 cm s^{-1}. Obviously the optimal plate spacing will differ for different organisms and under different operating conditions. The mathematical model provides a tool that allows the optimum to be determined for any particular combination of growth kinetics, bioreactor design and operating conditions.

24.4 Conclusions on Packed-Bed Bioreactors

Packed-bed bioreactors are necessary in those cases in which the bed must not be mixed during the fermentation. Such bioreactors will always suffer from axial temperature gradients. These can be minimized, although not eliminated, by selection of appropriate design and operating variables. The question as to what combination of design and operating conditions will lead to best performance is not simple, and is best answered with the use of a mathematical model.

The models that were presented in this chapter for the traditional packed-bed bioreactor and Zymotis packed-bed bioreactors could be much more powerful tools for use in the design process if they were improved. Some possible improvements include:

- introduction of a water balance. This would require a description of the effect of the temperature and the water content of the substrate on its water activity and a description of the effect of water activity on the growth kinetics of the microorganism. Note that the model of the intermittently-stirred bioreactor presented in Chap. 25 can be used to explore the water balance during packed-bed operation, simply by suppressing the mixing events;
- incorporation of the effect of growth of the microorganism into the interparticle spaces on the pressure drop through the bed and the resulting effects on air flow and heat transfer phenomena;
- description of changes in the bed due to the shrinkage of substrate particles as dry matter is converted into CO_2 by the microorganism.

Further Reading

Mathematical models of packed-bed bioreactors

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