Heat Transfer Simulation in Solid Substrate Fermentation

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A mathematical model was developed and tested to simulate the generation and transfer of heat in solid substrate fermentation (SSF). The experimental studies were realized in a 1-L static bioreactor packed with cassava wet meal and inoculated with *Aspergillus niger.* A simplified pseudohomogeneous monodimensional dynamic model was used for the energy balance. Kinetic equations taking into account biomass formation (logistic), sugar consumption (with maintenance), and carbon dioxide formation were used. Model verification was achieved by comparison of calculated and experimental temperatures. Heat transfer was evaluated by the estimation of Biot and Peclet heat dimensionless numbers 5-10 and 2550-2750, respectively. It was shown that conduction through the fermentation fixed bed was the main heat transfer resistance. This model intends to reach a better understanding of transport phenomena in SSF, a fact which could be used to evaluate various alternatives for temperature control of SSF, i.e., changing air flow rates and increasing water content. Dimensionless numbers could be used as scale-up criteria of large fermentors, since in those ratios are described the operating conditions, geometry, and size of the bioreactor. It could lead to improved solid reactor systems. The model can be used as a basis for automatic control of SSF for the production of valuable metabolites in static fermentors.

A large quantity of metabolic heat[3] is produced during SSF, up to 3200 kcal/kg dry matter (DM) in composting systems[4] and a temperature gradient of 3°C/cm in tempeh fermentation.[5] Heat generation is directly related to the metabolic activities of the microorganisms, particularly respiration during growth, which is related to oxygen consumption and CO_2 formation.

In SSF of cassava by *Aspergillus niger*,[6] nearly 35% of consumed sugars are oxidized by respiratory metabolism, producing approximately 80 kcal/h kg DM. The heat generated must be dissipated since microbial growth is very sensitive to a rise in temperature, affecting spore germination, growth, product formation, and sporulation. Different methods have been used to remove heat from the fermenting media.[3]

In fact, the use of stirred solid substrate fermentors can reduce system heterogeneity in comparison to static bed fermentation, and the macrogradients of a static bed could be decreased. However, the abrasion encountered in rotating systems can disrupt mold development That is, in certain cases the use of static SSF reactors could be required by using forced-air currents as a cooling mean.[7] In such

simultaneous ordinary differential equations that describes the physiological processes, i.e., biomass growth, substrate consumption, and carbon dioxide production, as function of temperature, and (b) an energy balance which takes into account the accumulation, conduction, convection, and generation phenomena through dimensionless numbers. Further, one of the features of this model is to use the temperature records as the experimental basis to follow up the spatial and temporal distribution of microbial activity. It can predict carbon dioxide evolution which can be used as a continuous signal of microbial metabolism. Such a model should contribute to our understanding of heat transfer and generation mechanisms in SSF and could lead to improved design and control techniques in reactor systems.

MATERIALS AND METHODS

Microorganism

A strain of *Aspergillus niger,* var. henebergii (No. 10), described by Raimbault and Alazard[2] was used.

Inoculum

Spore suspension was prepared as described elsewhere.[9] The inoculum size was 2×10^7 spores/g DM.

Pretreatment of Raw Material

Whole cassava (*Manihot esculenta,* var. cubana) meal was prepared using dry root chips obtained from Tabasco State, Mexico. Industrial grade mineral salts were added as follows (g/kg DM): ammonium phosphate 25 and ammonium sulfate 26. Salts were dissolved in tap water to a final moisture content of 38% adjusted to pH 3; the wet chips were gelatinized in a wet steam autoclave at 1 kg/cm^2 for 30 min. The mixture was then dried and milled at a mean particle size of 0.4–0.5 cm for its utilization.

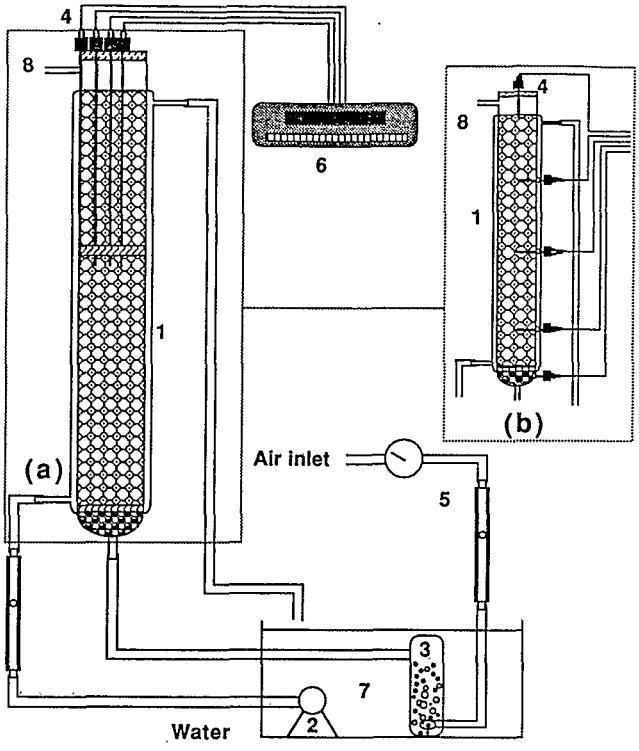

Figure 1. Experimental device for the monitoring of temperature gradient in solid fermentation of cassava by *A. niger.* (a) Radial gradient. (b) Axial gradient. Column fermentor; volume 1 L, 6 cm radius, 35 cm length: (1) jacket fermentor, (2) water pump, (3) humidifier, (4) thermocouples, (5) pressure and flow air controls, (6) temperature display device, (7) water bath.

mocouple, Eirelec, with a precision of 0.1°C. Temperature was measured at four positions along the radius at 0, 1.2, 2.4, and 3 cm. A fine double spiral acrylic grid plate was used to avoid interferences among thermocouples. Due to experimental limitations, either the radial or axial gradient was measured. Initial and final total true protein contents were measured by the Lowry method,[12] as applied earlier.[10]

concentration because the residual amount of sugar (around [redacted] specific growth constant (μ_{max}) and the maximum biomass [redacted]

Since the experimental data were not complete, Equations (4) and (5) were used only as assumptions required for model development. The validity of this would be tested by looking at the agreement between observed and calculated data by the model.

Transport Model

$$t = \frac{t'}{\Phi} \tag{10}$$

$$\Phi = \frac{L}{u} \tag{11}$$

The geometric reactor ratios L/R_a and R_a/d_p, the thermal diffusivity (α), and the dimensionless numbers Peclet (Pe),

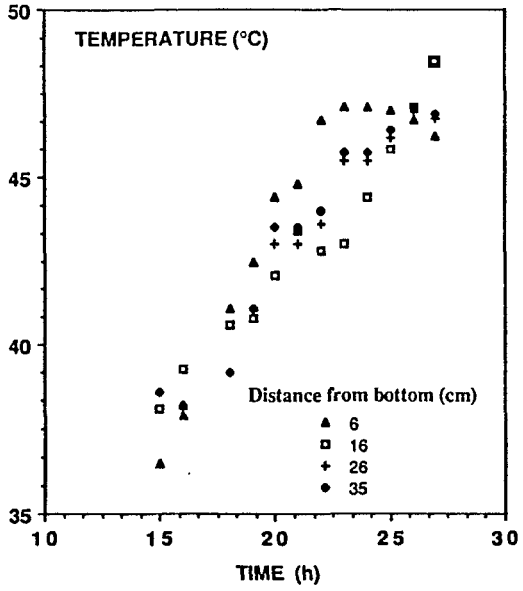

Figure 3. Evolution of the experimental temperature at different axis positions and at the center of the bioreactor.

tor. Thus for simulation studies the reactor length was considered to be 30 cm.

In similar experiments but monitoring the radial temperature gradient at 15 cm of the reactor top, a maximum difference of nearly 5°C/cm was found (Fig. 6) between the center and the wall of the reactor, that is, nearly 30 times greater than those compared with the axial gradient.

Reactor design and operating conditions can enhance the temperature gradient in one direction rather than the other. Rathbun and Shuler[5] encouraged the gradients of temperature in the axial direction by using a reactor of cartesian geometry placed in a dry fan-type incubator and using a high air flow rate. On the other hand, our reactor design and operating conditions enhanced the radial gradient of temperature rather than the axial one. These results are explained by the low air flow used, which also reduces the problems of air humidification and drying of fermenting mass. In other words, we have set conditions in such a way that the main heat transport resistance was conduction rather than convection.

Even when this simplification appears contradictory, the Peclet number remains in the model considering the ratio of convection and conduction heat transfer mechanisms in the packed bed bioreactor. The equation is

$$\frac{\partial T}{\partial t} = \frac{Ld_p}{R_a^2 \mathrm{Pe}}\left(\frac{\partial^2 T}{\partial r^2} + \frac{1}{r}\frac{\mathrm{d}T}{\partial r}\right) + F_{\mathrm{dm}}\mathrm{Da}_{\mathrm{III}}\frac{1}{R_{gi}}R_g \qquad (16)$$

using the boundary and initial conditions defined by Equations (12a–c) and the reaction rate defined by Equations (13)–(15). It is called the simplified model.

Model Verification

Temperature is a function of radius and time simultaneously [Eq. (16)]. Model verification was realized by comparison of real and calculated temperature. Testing points were at 4 positions in the reactor radius during the fermentation time. A total of 64 experimental data points were used. Two independent fermentation data sets were used in parameter estimation. Temperatures were measured from 15 to 30 h of fermentation time, which is the period of most active mold growth. The spatial term on the right side of Equation (16) was discretized using the collocation orthogonal method.[22] The resulting set of ordinary differential equations, from Equation (16), were integrated by a Runge–Kutta method.[23] Parameters were estimated through the Marquardt algorithm,[18] which minimizes the sum of squares. In Figure 5 is

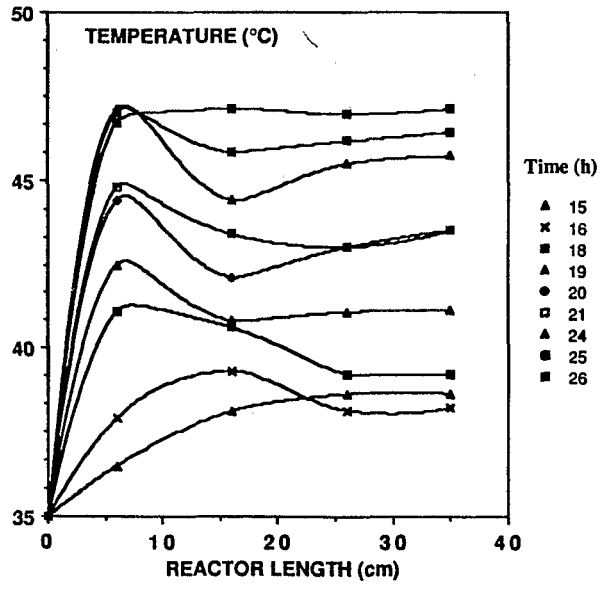

Figure 4. Variation of the experimental temperature as a function of the reactor length at different fermentation times.

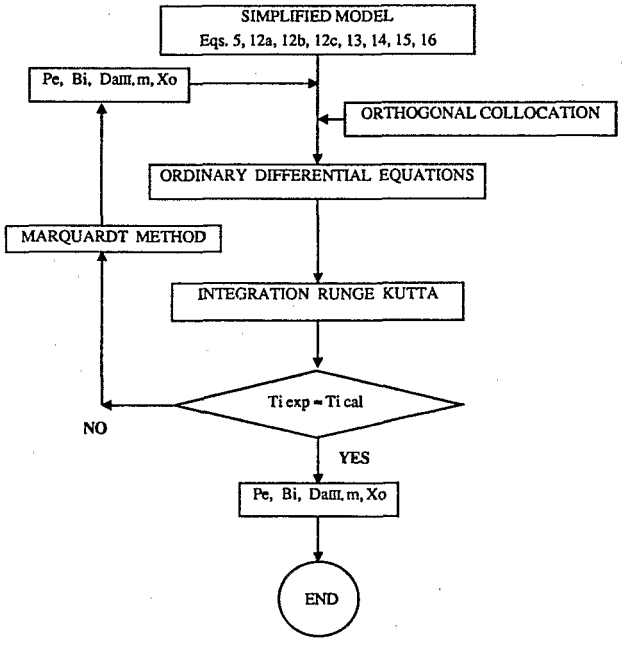

Figure 5. Program flow chart used for parameter estimation.

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In order to verify this model, we carried out a number of regressions using a factorial-like design, considering μ_{max}, X_{max}, and the maintenance coefficient [Eq. (14)] in the estimation of Pe, Bi, and Da_{III} (Fig. 5). For example, μ_{max} and X_{max} could have assigned mean values of 0.3 h^{-1} and 30% DM, respectively or changes with temperature according to Equations (4) and (5), respectively. The maintenance coefficient could have values 0 and 0.07 or an estimated value provided by the algorithm in Figure 5. Constant μ_{max} and X_{max} in the same regression were not accepted since it does not simulate different growth profiles in the fermenting media.

Unsatisfactory results in regression analysis were obtained by using μ_{max} defined by Equation (4), probably due to the sharp μ_{max} variation beyond 37°C (Fig. 2). Best results as judged by the minimum sum of squares were obtained by holding μ_{max} equal to 0.3, X_{max} as function of temperature [Eq. (5)] and allowing an estimation of the coefficient of maintenance by the algorithm (Fig. 5). Estimates of initial biomass (X_0) were also found to be critical and were of the order of less than 1 mg/100 g DM.

Finally, the regressions were realized by changing inoculum size, maintenance energy, and Peclet, Biot, and Damköhler numbers in the simplified model [Eqs. (5), (12a)–(12c), and (13)–(16)]. Regression results are shown in Table I for two independent fermentation data sets. For both cases parameters values were in the same order of magnitude. Figures 6–8 correspond to run 1; the results for run 2 were very similar and are not shown. In Table II are shown the constant values used in the regressions.

As was expected, the main effect on calculated temperature profiles was in Pe, Bi, and Da_{III} estimations. Maintenance energy and inoculum size presented a lower effect in minimizing the sum of squares. Estimation of inoculum size, X_0 (Table I), were in good accordance with those values previously calculated by Raimbault[6] (1.2×10^{-3}).

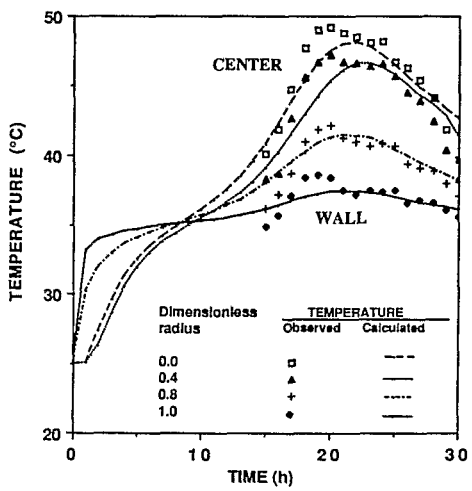

Figure 6. Comparison between calculated (——) and observed (symbols) temperature at different reactor radius during fermentation time.

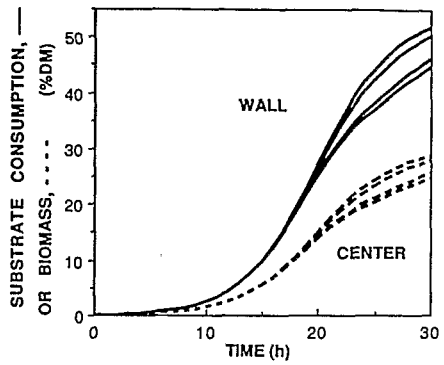

Figure 7. Evolution of the simulated biomass and sugar consumption at different reactor radius positions.

Nevertheless the calculated maintenance values were lower than those calculated by the same author (0.07).

Model verification was based on experimental and calculated temperature comparison. How well this worked is demonstrated in Figure 6, where the goodness of fit can be judged. Real data and predicted curves were in a good accordance since differences between calculated and observed values were very small (0.14°C). It suggests that the simplifications and principles used in model development and evaluation are correct.

The model simulates the biomass growth and sugar consumption (Fig. 7) inside the bioreactor. As expected, the biomass concentration and sugar consumption were more important near the wall than close to the reactor center. On the other hand, simulated final mean protein was 13.65% DM, considering 47% biomass content, which was in agreement with the experimental determination, 11.47% DM.

Further, carbon dioxide formation was simulated as a function of radius position, and a radial mean as a function

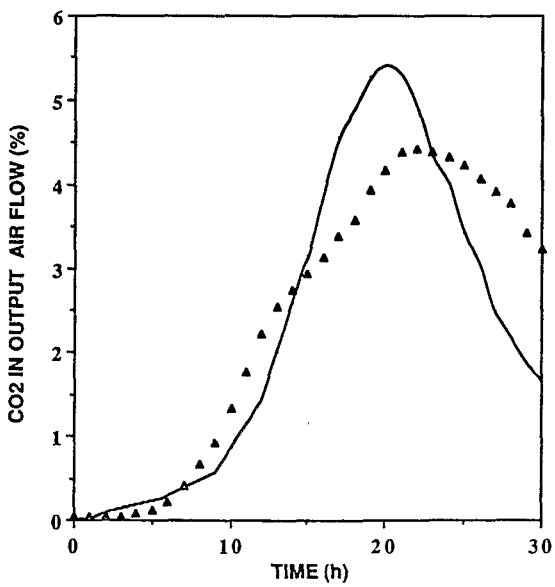

Figure 8. Comparison between calculated (——) and experimental (△) CO_2.

of time was calculated. Even when CO_2 production was not a criterion for model testing, comparison of simulated and real gas production gives satisfactory results (Fig. 8).

R_x	biomass reaction rate (g % DM/h)
r'	dimensional radius (cm)
S	consumated glucose (g/100 g DM, g % DM %)