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Model-based control of enzyme yield in solid-state fermentation

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Abstract

Solid-state fermentation is gaining increasing importance for the production of high-added value products, for instance enzymes, from agro-industrial by-products. The main difficulty in operating this process in packed-bed bioreactors is the development of temperature and moisture gradients along the reactor length which decrease the overall productivity. In this work a practically relevant feedback control scheme is developed that allows influencing and equalizing the moisture and temperature distributions along the packed-bed bioreactor during solid-state fermentation. This prevents locally critical process conditions which may necessitate the shutdown of the operation of the bioreactor.

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Keywords: solid-state fermentation; feedback control; distributed parameter system; heat and mass transfer; drying

1. Introduction

Solid-state fermentation (SSF) is a general term to designate a biotechnological process for cultivation of microorganisms on moist solid substrates with a continuous gas-phase in inter-particle space [1]. One of the most positive points of SSF is the possibility of using low-cost agro-industrial by-products as substrates in order to produce high-added value products, such as polymers, chemical compounds and mainly enzymes, which can be applied in many industries, for instance in foods or textile or even wastewater treatment [2].

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Cellulolytic enzymes are a special group of enzymes, which can be produced by SSF and has received attention from scientific and technological communities in recent years, since these enzymes are able to depolymerize biomass ecologically friendly into fermentable sugars with the final target of producing biofuels. Solid-state fermentation has been used to produce cellulases, hemicellulases and ligninases from a variety of biomasses, such as green or dried grasses, sugar cane bagasse, wheat bran, rice straw, soybean hulls, sawdust, orange pulp and peel, corncob and corn straw [3, 4, 5]. Hence, in regions with a high output of organic material of low value, such as agricultural economic-based countries, SSF provides means of adding value to the by-products and minimizes the problem of solid waste disposal.

For enzyme production, SSF is often implemented as a cylindrical packed bed bioreactor (PBB) with the (wetted) substrate forming the solid matrix and an initial inoculation of the solid material with the fermentative agent, usually shear sensitive filamentous fungi [2]. Due to the consumption of the substrate, an increase in the concentration of the fungal biomass and of the valuable product, for example an enzyme, is achieved. However, the fungal growth is accompanied by heat release due to respiratory activities. One of the main drawbacks while operating a PBB for SSF is the deficient removal of the metabolically generated heat, due to the low effective thermal conductivity of the substrate and to the low air flow rates employed, leading to hot spots above the ideal fermentation temperature, inhibiting the microorganism growth and affecting the production of metabolites [6, 7].

Moreover, water is removed from the solid matrix by the gas flow which has to percolate the bioreactor in order to supply oxygen to the cells and remove carbonic gas, heat and water vapor from the surroundings, in order to maintain the organism-specific environmental conditions, since the microorganisms are quite sensitive to changes in temperature and water activity of the solid matrix. However, temperature and consequently moisture content profiles are difficult to be avoided in PBB due to end-to-end aeration and the use of convective cooling with unidirectional flow of air, yielding sections that are too dry for fermentation and sections that are too moist, thereby decreasing the yield of the process. Even if the inlet air is saturated, the axial temperature gradients will change its saturation moisture content, giving to the air driving force to remove water from the solid-phase [8].

Solid-state fermentation is therefore a complex system, involving multiple phases and multiple components, with heat and mass transfer taking place between the phases and the components as well as reaction processes. Various attempts to control the thermal conditions in SSF to enhance the processe performance, either in open-loop or closed-loop operation, can be found in the literature with varying success and a strong dependency on the process equipment. Von Meien et al. [9], for example, presented a MISO feedback control approach using PID and DMC for the mean bed temperature in a mixed SSF bioreactor, where the mixing achieves implicitly an equalization of the temperature and moisture gradients. Chen et al. [10] considered the control of the temperature distribution by a pulsating operating pressure and an internal air circulation. They were able to obtain good results on a lab-scale plant with a height of a few centimeters, however, the choice of manipulated variable (the operating air pressure in the bioreactor) makes it difficult to implement this scheme on pilot or plant scale. In other works, e.g. Abbas Shojaosadati et al. [11], open-loop control is considered to achieve, amongst other aspects, an optimal temperature distribution in the bioreactor.

In this contribution a model-based control scheme for a packed bed SSF process is developed with the aim of manipulating the thermal conditions in such a way to avoid operating conditions throughout the bed that may inhibit microorganism growth and negatively affect the production of enzymes. After a summary of the process model, the feedback control system is designed, including an analysis of the different manipulated variables. A main aspect of the design is the use of manipulated variables that can be implemented easily on all plant scales. This contribution closes with simulation results of the closed-loop operation and a discussion of the achieved performance.

Nomenclature

b	fungal biomass concentration
d_p	particle diameter
ha	heat transfer coefficient
l_p	particle length
t	time
v_0	superficial air flow velocity

z	spatial coordinate (length)
B	wet solids concentration
Cp_a, Cp_s	specific heat capacity air (a) and solid (s)
Cp_v, Cp_w	specific heat capacity water vapour (v) and liquid water (w)
D	diameter of packed bed
L	length of packed bed
R_s, R_w, R_Q	reaction/conversion yield coefficients
S	dry solids concentration
T_g, T_s	temperature of gas (g) and solid (s)
X	solid moisture content
Y, Y*	gas moisture content, saturation moisture content
βa	mass transfer coefficient
ε_0	packed bed porosity
μ	biomass growth rate
ν’	normalized drying rate
ρ_a,ρ_s	mass density air (a) and solid (s)
ΔH_{vap}	specific heat of evaporation

2. Process modeling

2.1. Balance equations

Several mathematical models have been proposed in SSF literature to provide insights into the competing phenomena and the resulting process performance. Mitchell et al. [8] and Sangsurasak and Mitchell [12], for instance, presented a model taking into account the two-dimensional heat transfer and evaporation in the SSF, enabling to predict drying properly during fermentation. Von Meien and Mitchell [13] and Schutyser et al. [14] proposed two-phases models to predict temperature and moisture gradients in packed bed bioreactors for SSF, which in addition allows to represent water transfer from the solid phase to the gas phase.

The model used in this contribution is motivated by the approaches presented in [13, 14] and considers a fermentative system composed of the thermophilic fungus *Myceliophthora thermophila* I-1D3b cultivated in a mixture of sugar cane bagasse and wheat bran (7:3 dry weights) [5]. The bioreactor was supposed to operate as packed bed bioreactor without mixing during the whole fermentation. Heat and mass transfer as well as the production of biomass from the substrate are considered in axial direction of the cylindrical packed bed.

The model therefore consists of mass and energy balances for: (a) mass of water vapor in the gas, (b) mass of liquid water on the substrate, (c) biomass, (d) gas temperature and (e) substrate temperature. Due to the spatial dependencies of the balanced quantities, the resulting equations are partial differential equations whose solution gives the temporal evolution of the masses and temperatures along the packed bed.

The mass balance of water in the gas phase is expressed by equation (1):

$$\rho_a \varepsilon_0 \frac{\partial Y}{\partial t} + \upsilon_0 \ \rho_a \frac{\partial Y}{\partial z} = \nu'(X)\, \beta a\, \rho_a \varepsilon_0 (Y^* - Y) \tag{1}$$

First term on left-hand side represents the change in water content of the gas, while the second term represents flow of water vapor by advection. The term on the right-hand side represents solid-to-gas interface water transfer.

Similarly, the mass balance of water in the solid phase is given by equation (2):

$$S\frac{\partial X}{\partial t} + X\frac{\partial S}{\partial t} = -\nu'(X)\, \beta a\, \rho_a \varepsilon_0 (Y^* - Y) + R_w \left[S\frac{\partial b}{\partial t} + b\frac{\partial S}{\partial t} \right] \tag{2}$$

The first term on the right-hand side describes the water removal by evaporation, the second term represents water production due to fungal growth.

The energy balance for the gas phase is given by equation (3):

$$\rho_a \varepsilon_0 (Cp_a + Y\, Cp_v)\frac{\partial T_g}{\partial t} + \upsilon_0\, \rho_a (Cp_a + Y\, Cp_v)\frac{\partial T_g}{\partial z} = \Delta H_{vap}(T_s)\, \nu'(X)\, \beta a\, \rho_a \varepsilon_0 (Y^* - Y) - ha\,(T_g - T_s) \quad (3)$$

First term on left-hand side represents accumulation of energy in the gas phase and the second term represents the convective heat transfer due to advection. The first term on right-hand side represents the amount of energy associated to water evaporation, whereas the second describes the heat transfer from the gas to the solid phase. Heat transfer by conduction is neglected, as the thermal conductivity of air is low.

The energy balance in the solid-phase was expressed by equation (4):

$$S(Cp_s + X\, Cp_w)\frac{\partial T_s}{\partial t} + T_s(Cp_s + X\, Cp_w)\frac{\partial S}{\partial t} = ha\,(T_g - T_s) - \Delta H_{vap}(T_s)\, \nu'(X)\, \beta a\, \rho_a \varepsilon_0 (Y^* - Y) + R_Q \left[S\frac{\partial b}{\partial t} + b\frac{\partial S}{\partial t} \right] \quad (4)$$

The third term on right-hand addresses the production of metabolic heat during fungal growth. Similar to the energy balance of the gas phase, heat transfer by conduction in the substrate is neglected as the thermal conductivity is also low.

The increase of fungal biomass (b) was assumed to follow the logistic model [15]:

$$\frac{db}{dt} = \mu b \left(1 - \frac{b}{b_{max}} \right) \quad (5)$$

Fractional specific growth rates, μ_T (depending on T_s) and μ_W (depending on a_{ws}), were defined as fractions of μ_{opt}, as already assumed by Von Meien and Mitchell [13]. The growth rate μ_W was obtained by empirical fit to experimental data of Glenn and Rogers [18] for *Rhizopus oligosporus*, while the growth rate μ_T was given by empirical fit of an Arrhenius-type equation proposed by Saucedo-Castañeda et al. [16]. The growth rate μ was then calculated by geometrically averaging μ_W and μ_T [17].

The dry solids concentration (S) changes due to substrate consumption and biomass production:

$$\frac{\partial S}{\partial t} = R_S \frac{\partial B}{\partial t} = R_S \frac{\partial (Sb)}{\partial t} \quad (6)$$

The saturation moisture content of the gas phase (Y^*) required to calculate the evaporation mass flow was calculated based on saturation pressure of water vapor in air, given by an Antoine equation and depending on T_g [19]. Because of larger proportion of sugar cane bagasse in substrate composition, it was assumed that the substrate sorption isotherm would be represented well by the sorption isotherm of this component [20].

The interface gas-solid heat and mass transfer coefficients were calculated from classical correlations for Nusselt (Nu) and Sherwood (Sh) correlations. Substrate fibers were assumed to be of cylindrical shape, 70% of them with air in cross-flow and 30% in parallel flow. Hence, Nusselt numbers were calculated for cylinders in cross-flow and in parallel flow and then a weighted Nu_w number was obtained. For the calculation of the corresponding Sherwood number Sh_w the analogy between heat and mass transfer was used. The interface coefficient for heat transfer (h) was obtained from Nu_w and for water transfer (β) from Sh_w. For cross-flow, a correlation given by Churchill and Bernstein [21] was employed; for parallel-flow, cylinders were considered as flat plates [22]. The obtained values were then scaled by the ratio of total transfer area in the bioreactor to total bioreactor volume (a).

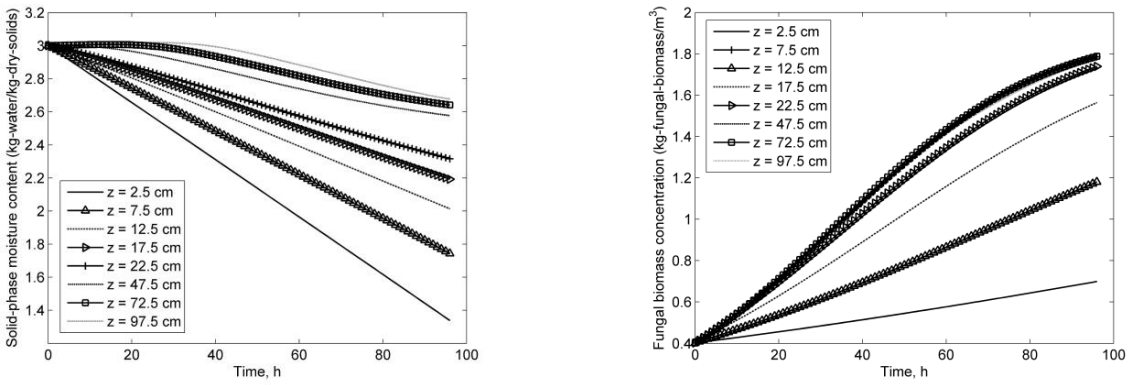

Fig. 1. (left) Solid-phase moisture content at different positions in the packed-bed bioreactor over time. (right) Fungal biomass concentration at the same positions over time. It can be seen that low solid-phase moisture content corresponds to a low fungal biomass production.

2.2. Open-loop process dynamics

In Figs 1 and 2 process results for the solid moisture content, fungal biomass production, as well as the solid temperature along the packed-bed bioreactor obtained in typical SSF operation (Tab. 1) are shown. It can be seen that over time a temperature gradient develops in the bed with the lowest temperature at the inlet of the PBB (equal to the gas inlet temperature) and the highest temperature at the outlet. The temperature maximum depends on the process conditions, the available amount of substrate and the reaction rate. In the present case a maximum increase of 4.7K is obtained. The temperature gradient influences the drying significantly as can be seen in the plot of the solid-phase moisture content. The moisture content up to approximately one quarter of the total length of the PBB is significantly lower than in the later sections of the reactor. This is due to the drying potential of the unsaturated inlet air which is further increased by the exothermic reaction. In the later sections, the air is saturated and the moisture content is reaction-controlled. Same trend can be seen in the plot of the fungal biomass concentration along the PBB. The concentration is quite non-uniform, with significantly lower concentrations in the inlet section of the reactor.

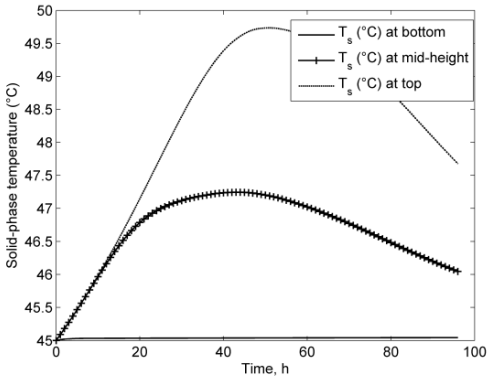

Fig. 2. Temperature distribution at the inlet, in the middle and at the outlet of the packed-bed bioreactor over time. It can be seen that the temperature increases with increasing distance from the inlet. The temperature at the inlet of the bioreactor is constant as it is a fixed process parameter in open-loop operation.

In order to increase the efficiency of SSF, in terms of the biomass concentration, the thermal management influencing the solid-phase moisture content has to be improved. One approach to achieve this is presented in the following section.

Table 1. Process parameters.

Symbol	Value
b_0 [kg biomass/kg dry solid]	0.0041
b_{max} [kg biomass/kg dry solid]	0.0204
d_p [m]	0.0046
l_p [m]	0.015
t_{end} [h]	96
v_0 [m/s]	0.007309
ε_0 [m^3 void/m3 total]	0.75
ρ_a [kg/m^3]	1.11
B_0 [kg biomass/m^3]	0.4
Cp_a [J/kg/°C]	1006
Cp_s [J/kg/°C]	1760
Cp_v [J/kg/°C]	1880
Cp_w [J/kg/°C]	4184
D [m]	0.0762
L [m]	1
R_s [kg dry solids/kg biomass]	-2
R_w [kg water/kg biomass]	0.3
R_Q [J/kg biomass]	8366000
S_0 [kg dry solids/m^3]	98.4
T_{g0}, T_{gin},T_{s0} [°C]	45
X_0 [kg water/kg dry substrate]	3

3. Feedback control design

3.1. Control configuration and choice of manipulated variables

Although there is a huge number of process parameters influencing the process performance, the set of possible manipulated variables is limited and covers mostly gas-side parameters, i.e. gas inlet temperature, gas inlet moisture content and gas mass flow. The solid properties can in general not be changed during the process. An exception is the possibility to install a jacket, by which the wall temperature of the PBB during SSF can be influenced. An effective influence on the process is usually reported for “radially thin” reactors, i.e. reactors with a ratio of diameter-to-length considerably smaller than unity, where the heat and mass transfer along the radial coordinate can be neglected.

Additionally, due to the low conductivity of the materials, the heat and mass transfer in the PBB is convection-dominated, that is changes in the quantities of interest are transported with finite speed along the reactor. This leads to significant delays (dead times) in the response of measured quantities, e.g. gas moisture content at the outlet, with respect to changes in a manipulated variable, for instance at the inlet of the PBB.

Standard measurements for the control of SSF in packed-bed bioreactors are thermocouples placed along the length of the reactor and positioned in the center, measuring a mixing temperature of gas and solid phase. In some cases, mostly at lab-scale plants, the outlet gas moisture content is also measured. Apart from suffering from the

delay times in the response of this measurement to changes in the input or along the length of the PBB due to the convective transport, it is currently not clear whether on industrial scale this measurement will be generally available.

The gas-side manipulated variables also suffer from the time delays and due to the complex interaction with the solid phase, changes at the inlet of the packed-bed in reaction to changes near the outlet may worsen the situation in the solid phase even more. But due to the lack of alternatives, the gas-side inputs as well as the jacket temperature are chosen as manipulated variables.

3.2. Controller design

In the following a simple and easily implemented approach is proposed which allows influencing the temperature and moisture profiles along the length of the packed-bed reactor. The main idea is to cut down delay times inside the reactor by extending the plant to allow for switching the flow direction of the gas. This can be realized controlled hatch in the air distributor. The temperatures along the bed can be obtained from the spatially distributed measurements and thus the position of the current hot-spot can be detected. Assuming an initial flow direction from-left-to-right, the direction is switched to from-right-to-left if the hotspot is detected in the second half of the reactor and vice versa. The switching is controlled by a bang-bang feedback controller with a certain threshold in the temperature maximum to react on. Measurements of the maximum temperature in the reactor are fed to the bang-bang controller with a user-defined sampling time based on the time scale of the heat production, to avoid excessive switching of the flow direction. This strategy will yield a forced movement of the hot-spot from either end to the middle of the bioreactor. During this movement, by an underlying temperature feedback control scheme including the jacket temperature, the maximum peak is decreased. This can be realized by standardized PI-controllers based on the measurement of the maximum temperature in the reactor and the jacket temperature. The reference value for the reactor temperature in the controller corresponds to the optimum value obtained from the sorption isotherms and water activity.

4. Results and discussion

In Figs 3 and 4, results are shown for the process considered in Sec. 2.2 with the design changes made as explained in Section 3.2. In a first step, switching is performed every four hours, if necessary. The gas-side manipulated variables are kept constant, i.e. a pure change of flow direction takes place. The temperature of the jacket is adjusted by a simple proportional controller. By direct comparison, it can be seen that the solid moisture and fungal biomass concentration profiles are much more uniform, additionally the temperature peak is reduced. Drying of the substrate in the former inlet section is considerably smaller and an increase in fungal biomass production in this section can be observed. As a side effect, due to the switching of the flow direction, now a small drying effect and a decrease in biomass production can also be observed at the either end of the packed-bed bioreactor. Although this signifies a decrease in productivity of the PBB, the more uniform distributions along the reactor allow for longer process times as locally critical process states which may require the shutdown of operation are prevented. Also shown in Fig. 4 is the flow direction of the gas with positive values signalizing a flow from left-to-right and negative values from right-to-left.

In Fig. 5 the influence of the switching frequency on the process results is shown exemplary for the fungal biomass concentration. It can be seen that an increase in switching frequency, i.e. a decrease of the time interval between switching decisions, yields mainly in a smoother change of the concentration but has only small influence on the obtained value.

It has to be noted at this point that the proposed scheme is indirect in terms of the solid moisture content, meaning that the need for preliminary testing of the control scheme is not rendered completely obsolete and that re-tuning of the controller parameters may be required at the plant.

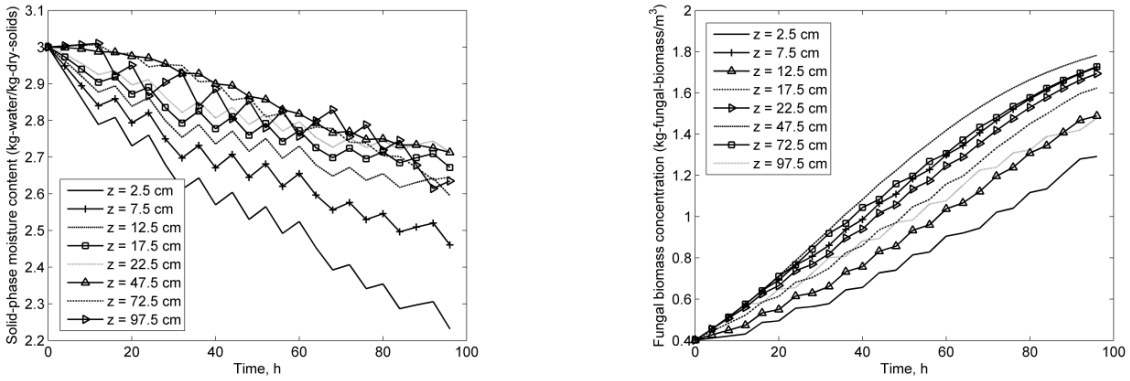

Fig. 3. (left) Solid-phase moisture content over time under feedback control. Compared to Fig. 1 (left), drying is less extreme at the inlet of the reactor and more uniform over the reactor length. (right) Fungal biomass concentration at various positions along the reactor. Compared to Fig. 1 (right) an increase in fungal biomass production in the inlet section is achieved, but a certain decrease at the former outlet section is observed. However, the biomass distribution along the reactor is more uniform.

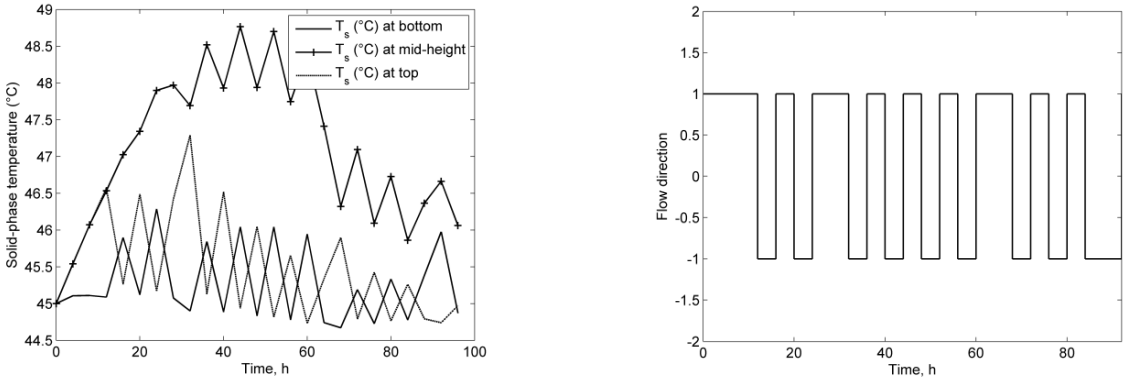

Fig. 4. (left) Temperatures at either end and in the middle of the reactor. As predicted, the maximum temperature is moved towards the middle of the reactor and is decreased by heat transfer to the jacket. The changes in the temperatures correspond to the change in flow direction, as either end is used as inlet (low temperature) and outlet (high temperature). (right) Profile of the switching of the flow direction during SSF operation under feedback control. A change in sign denotes a change in flow direction, "+1" corresponds to a flow from left-to-right (as in open-loop operation), "-1" denotes a flow direction from right-to-left.

4. Summary and outlook

In this work, based on a rigorous one-dimensional spatially distributed model of heat and mass transfer, a feedback control scheme for the solid-phase moisture content and the fungal biomass concentration in a packed-bed bioreactor in solid-state fermentation operation was devised. It was shown that by a change in apparatus design and the use of industrially available standard feedback controllers a significant improvement in the spatial distribution of the solid-phase moisture content can be achieved. This is a key element for a spatially homogeneous production of fungal biomass which is preferred in many applications.

Currently, only the reactor temperature is directly influenced by a jacket, in future works also the gas mass flow and its temperature will be controlled to increase the operation range and to increase the speed of the fermentation process. As these process parameters do not have an independent influence on the solid-phase moisture content and the fungal biomass concentration, advanced multivariable feedback control schemes will be designed.

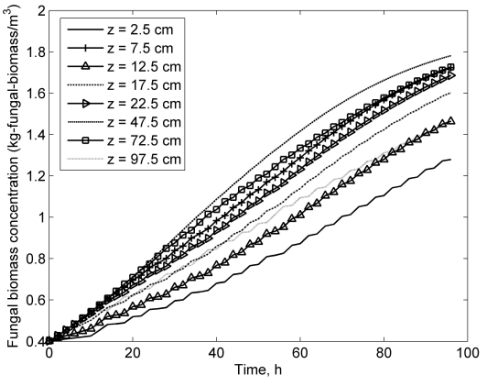

Fig. 5. Influence of switching frequency on the process result. Compared to Fig. 3 (left), the time interval between switching decisions was decreased from 4 hrs to 2 hrs. A smoother change in concentration can be observed, the final values are not significantly changed.

Acknowledgements

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