PRODUCTION OF BIOMASS FROM SULPHITE LIQUORS IN TOWER FERMENTOR WITH FORCED CIRCULATION

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Cultivation of the yeast *Candida utilis* on sulphite liquors was studied in a tower fermentor with ejector-type gas distributor and additional bubble bed showering. The aim of the work was to test the fermentor applicability for the cultivation process studied and to obtain data characterizing the fermentation in the tested unit. The effect of medium composition and dilution rate on selected cultivation parameters was determined in the fermentor with active volume $0{\cdot}1\ m^3$ and the hydrodynamic and energy characteristics of the apparatus were evaluated. The experimental results proved the fermentor to be suitable for the cultivation process studied. Favourable effect of the optimized fermentation medium was observed both on the values of characteristic cultivation parameters and on the efficiency of the utilization of energy supplied to the system.

Due to the increase of crude oil prices on the world market, production of fodder proteins from crude-oil derivatives (n-alcanes, alcohols) has been becoming uneconomical. As a result, trends towards utilization of waste materials as carbon sources for various microbial synthesis can be observed worldwide. In our conditions application of sulphite liquors (namely those of magnesium-bisulphite type) seems to be one of the most perspective possibilities. Apparently the utilization of sulphite liquors is important not only from the viewpoint of the economics of fodder proteins (SCP) production but also due to ecological contribution of this technology of sulphite liquors processing. In comparison with crude-oil derivatives used as carbon sources represent however the sulphite liquors considerably more complex medium containing wide spectrum of species. From the chemical engineering point of view it means that the relations obtained on the basis of experiments with model media or in other fermentation systems cannot be used for description of hydrodynamics and mass transfer in sulphite-liquor based fermentation broths. Experimental data characterizing hydrodynamic behaviour of the complex gas–liquid system as well as data basis for the estimation of the effect of fermentor construction parameters on the hydrodynamics and mass transfer in the system have to be therefore obtained (if possible) directly under conditions of the cultivation process.

Regarding the existence of mutual relations between hydrodynamic characteristics of fermentors and corresponding cultivation process parameters it is obvious that the proper choice of the fermentor type and its construction parameters values can considerably influence the productivity and economy of specific cultivation processes. Beside the search for new raw-material sources for biomass production considerable attention has been therefore paid lately to the development of new types of fermentors and to optimization of their construction parameters according to specific demands of various fermentation processes.

In last few years several modifications of fermentors with forced medium circulation were tested in our research group and their application for the continuous cultivation of the yeast *Candida utilis* on synthetic ethanol was studied[1,2]. Experimental results proved that the intensity of interfacial mass transport achieved in the fermentor with Venturi tube – type ejector gas distributor was sufficient even for the process with relatively high demand on oxygen supply. Special modification[3] of such fermentor was also designed and tested. Part of the liquid was taken off the circulation circuit to the top of fermentor and sprayed above the bed surface. It was proved that such arrangement reduced substantially formation of the stable foam in the system[1,2]. It was the aim of our present work to test the applicability of this modified ejector-distributor fermentor for cultivation of *Candida utilis* yeast on sulphite liquors and to obtain data characterizing the cultivation process in the tested unit *i.e.* data which could serve as a base for process- regime optimization. To achieve these goals, the effect of cultivation medium composition and of the dilution rate on the decisive cultivation parameters was determined and values of parameters characterizing hydrodynamics and energy effectiveness of the fermentor were evaluated.

EXPERIMENTAL

Equipment

Tower fermentor with forced medium circulation[3] was used for experiments. Its construction principles are apparent from the schematic chart of experimental set-up (Fig. 1). The fermentor body consisted of glass cylinders mounted between metal flanges. Apparatus diameter was 0·29 m, overall height was 3·5 m. The ejector (Venturi-tube type) was used for gas–liquid dispersion formation (Fig. 2), diffuser length was 0·43 m, diffuser cross-section at the outlet was 0·02 m^2, ratio of outlet and inlet diffuser cross-section was 2·5. Nozzle of diameter $d_s = 0{\cdot}011$ m was used in all cultivation experiments, several preliminary measurements of hydrodynamic parameters were performed with nozzle diameter 0·009 m. Volumetric flow rate of medium in the circulation circuit was constant in all experiments, $Q_L = 7\ m^3/h$, corresponding pressure drop over the ejector was 0·12 MPa. Volumetric flow rate of gas sucked corresponding to the experimental value $Q_L\,\Delta P_e$ was 3·7 m^3/h, gas sucking was supressed to 2·85 m^3/h during some experimental runs. Corresponding values of superficial gas flow rate were 0·015 and 0·012 m/s respectively. Part of the liquid (about 10%) was taken off the circulation circuit and sprayed at the top of fermentor above the gas–liquid bed.

Cultivations were performed with the yeast culture *Candida utilis* from the collection of the Institute of Microbiology, Czechoslovak Academy of Sciences, the culture was adapted to high concentration of free SO_2. Calcium-bisulphite liquors from South-Bohemian paper mills Větřní were used as a base of the fermentation medium. Liquors were obtained in a thickened form and contained also a certain unknown amount of sulphite stillage. Two sets of experiments were performed for two different compositions of the fermentation broth (Table I). Composition of medium A corresponded to the medium used by Vogelbusch company[4], medium B was prepared according to results of optimization cultivation experiments performed by Křen[5] at the Prague Institute of Chemical Technology. Cultivation media were prepared in a storage tank and pumped *via* the overflow vessel into the column, the volumetric flow rate was measured by a rotameter. Medium was not sterilized, storage tanks and fermentor were sterilized before

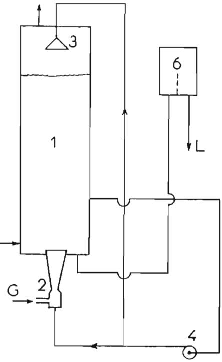

FIG. 1

Experimental set-up. 1 column, 2 ejector, 3 spraying nozzle, 4 pump, 5 storage tank, 6 overflow vessel

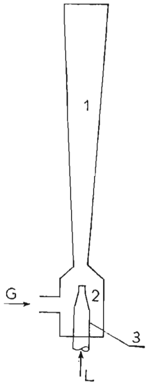

FIG. 2

Schema of Venturi-tube type ejector. 1 Diffuser, 2 suction chamber, 3 nozzle experiments by 3% formaldehyde solution. Constant pH = 4·3 and temperature 30°C were kept in fermentor during cultivation experiments. Defoaming agent (Kontramin) was dosed continuously to the system during all cultivation runs, dosing rate was 16 cm^3/h. Reactor was filled by the medium at the beginning of each run, initial dry biomass concentration in the bed was approximately 1 kg/m^3. After that the aeration was started, the fermentor content was tempered and pH value was fixed. When the dry biomass concentration reached about 10 kg/m^3, the growth period was considered to be finished and continuous operation was started. After the steady state was reached values of selected process parameters were measured. The length of measuring period equaled to triple value of mean residence time of medium *i.e.* 5—10 h in various runs.

Experimental Program

Two sets of experiments were performed using two different fermentation media (1. set — medium A, 2. set — medium B, see Table I). Dilution rate, D, defined as a ratio of medium flow rate to its volume in fermentor, $D = \dot{V}_L / V_L$, was chosen as the second independent variable. Dilution rate values were varied in both sets of experiments between 0·10 and 0·23 h^{-1} by the change of medium flow rate while constant liquid holdup 0·089 or 0·101 m^3 was kept during all experimental runs of the first and second set respectively. Medium flow rates ranged between 0·009 and 0·021 or 0·01 and 0·0235 m^3/h for the first and second set of experiments respectively.

Concentration of oxygen dissolved in medium, concentrations of oxygen and CO_2 in the gas phase at the fermentor outlet, dry biomass content and content of reducing substances in medium and gas holdup (bubble bed porosity) were determined experimentally during cultivation runs, while values of other selected dependent variables (cultivation parameters) were calculated from the measured experimental data. The calculated characteristics included respiration coefficient, RQ, rates of oxygen consumption and CO_2 production, $q(O_2)$, $q(CO_2)$, relative utilization of oxygen, R_v, yield of biomass related to the consumed oxygen, $Y_{X/O}$, overall productivity, p, and volumetric liquid-side mass transfer coefficient, $k_L a$, characterizing the rate of interfacial oxygen transfer in the system. All dependent variables evaluated during cultivation experiments are summarized in Table II. Due to the high content of non-reducing assimilable substances in cultivation medium, caused by the presence of sulphite stillage in liquors used, the yield of biomass related to the consumption of reducing substances, commonly used as a cultivation process characteristics, was not evaluated in our work.

TABLE I
Composition of cultivation media (related to 0·1 m^3 of a medium)

Composition	Medium A	Medium B
H_3PO_4 (solution 55% vol.)	0·072 l	0·097 l
NH_4Cl	0·29 kg	0·19 kg
KCl	0·078 kg	0·078 kg
$MgCl_2.6\,H_2O$	0·038 kg	0·038 kg
NH_4OH	0·415 l	0·591 l
Reducing substances	2·3 kg	3·1 kg
Dry solids	11% (mass)	16% (mass)

Measuring Methods

The outlet oxygen concentration was measured by the oxygen analyzer Permolyt 2 (Junkalor Dessau), concentration of CO_2 by the infrared analyzer Infralyt 5 (Junkalor Dessau). Concentration of oxygen dissolved in the fermentation broth was monitored continuously by the polarographic electrode of the Clark type and registered by the recorder TZ 21S. Dry biomass content in the medium was determined gravimetrically, concentration of reducing substances was determined by a titration method, detailed description of both methods can be found in the thesis of Kostka[6]. Other dependent variables listed in Table II were then calculated from values of these measured parameters. Values of the respiration coefficient, RQ, were read off the nomogram published by Fiechter and Meyenburg[7] for corresponding experimental values of inlet and outlet concentrations of O_2 and CO_2 in the gas phase. Total rates of oxygen consumption and CO_2 production related to the volume of liquid medium, $q(O_2)$, $q(CO_2)$, and the relative oxygen utilization, R_v, were also calculated from concentration differences $(y_1 - y_2)$ and $(z_1 - z_2)$ determined experimentally for O_2 and CO_2 using the relations

$$q(O_2) = (y_1 - (1 - A)\,y_2)\,M(O_2)\,P\dot{V}_G/V_L RT \tag{1}$$

$$q(CO_2) = ((1 - A)\,z_2 - z_1)\,M(CO_2)\,P\dot{V}_G/V_L RT \tag{2}$$

$$R_v = 100(y_1 - y_2)/y_1\,, \tag{3}$$

where A in Eqs (*1*) and (*2*) is a correction factor defined as

$$A = (y_1 - y_2) + (z_1 - z_2)\,. \tag{4}$$

The yield of biomass related to the oxygen consumption, $Y_{X/O}$, and total productivity, p, were for individual steady states, characterized by medium composition and by specific values of the dilution rate, calculated from relations

$$Y_{X/O} = D\,.\,X/q(O_2)\,, \tag{5}$$

$$p = D\,.\,X\,, \tag{6}$$

where X denotes steady-state dry biomass concentration in the fermentation broth.

TABLE II

Dependent variable parameters determined in cultivation experiments

Measured quantities	X kg/m^3	c_L kg/m^3	y % vol.	z % vol.
Calculated quantities	$Y_{X/O}$ kg/kg	RQ —	$q(O_2)$ kg/m^3 h	$q(CO_2)$ kg/m^3 h
Calculated quantities	R_v %	p kg/m^3 h	k_La s^{-1}	ε_G —

Steady-state values of k_La were determined from the oxygen balance in the system according to the equation[1]

$$k_La = q(O_2)/(c_L^+ - c_L), \quad (7)$$

where the value of equilibrium oxygen concentration at the interphase $c_L^+ = 7{\cdot}365 \cdot 10^{-3}$ kg/m^3 was determined experimentally for the fermentation medium used. The form of Eq. (7) corresponds to the assumption of complete mixing of the liquid phase and to the case when dissolved oxygen is consumed only by cells. Values of gas holdup (gas–liquid bed porosity), ε_G, were determined from the difference of clear liquid height and the height of aerated bed, $\varepsilon_G = (H - H_0)/H$. The height of stable foam was not included into the aerated bed height, calculated ε_G values thus represented the dynamic gas holdup.

RESULTS AND DISCUSSION

Cultivation Parameters

The effect of medium composition and of the dilution rate on the values of selected cultivation characteristics was studied during cultivation experiments. Dry biomass concentration in the fermentor, X, overall process productivity, p, relative oxygen utilization, R_v, and yield of biomass related to the oxygen consumption, $Y_{X/O}$, were selected as decisive parameters for the evaluation of efficiency and economics of the fermentation process. In Figs 3 and 4 experimental data X *vs* D and p *vs* D are compared for both sets of experiments *i.e.* for the two cultivation media used. It is clearly apparent from Fig. 3 that with the optimized medium B, having both higher

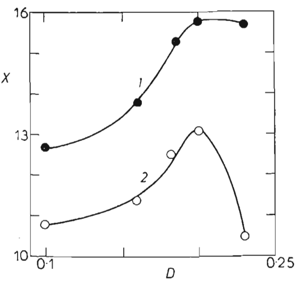

FIG. 3

Dependence of steady-state dry-biomass concentration on dilution rate. 1 Medium A, 2 medium B

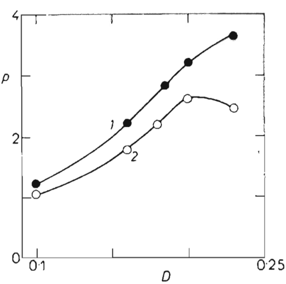

FIG. 4

Dependence of productivity on dilution rate. 1 Medium A, 2 medium B initial content of reducing substances and higher overall content of assimilable species, higher values of dry biomass concentration were achieved in the whole experimental region of dilution rates. For medium A, the dependence X *vs* D exhibited a distinctive maximum at $D \approx 0{\cdot}20\ h^{-1}$ while only inexpressive extreme can be observed at $D \approx 0{\cdot}22\ h^{-1}$ on the curve corresponding to medium B. Similar course as graphs X *vs* D exhibited also dependences of overall productivity on the dilution rate plotted for both media studied in Fig. 4. Values of productivity for the optimized medium B were higher than those for medium A in the whole experimental region of dilution rates. The shape of the graph for medium B suggests further that even higher values of productivity could be achieved with this medium at higher dilution rates ($D > 0{\cdot}23\ h^{-1}$). For medium A, the dependence p *vs* D had a maximum at $D \cong$ $\cong 0{\cdot}20\ h^{-1}$. Corresponding maximum value of productivity for this medium was $2{\cdot}62\ kg/m^3\ h$, the highest productivity achieved for medium B was $3{\cdot}65\ kg/m^3\ h$ at $D = 0{\cdot}23\ h^{-1}$. Significantly higher values of oxygen utilization factor R_v were also observed in experimental runs with the optimized medium B, values R_v for this medium ranged between 22·6 and 32·8% in the experimental region of dilution rates, in comparison with the values 9·3–21·9% for medium A. Respiration coefficient values were almost identical for both media studied, ranging between 0·83 and 1·07 for medium A and 0·84 and 1·03 for medium B. In the whole region of experimental conditions the system never approached the oxygen limitation state, concentrations of oxygen dissolved in fermentation medium, c_L, ranged between 43 and 99% or 53 and 70% of the saturation value for medium A and B respectively.

For comparison, some results can be presented obtained previously[4] with unthickened Ca-bisulphite liquors in the fermentor Vogelbusch (fermentor with immersed jet), $V_L = 0{\cdot}2\ m^3$. Maximum productivity value achieved in that study was $5{\cdot}65\ kg/m^3\ h$ at $D = 0{\cdot}41\ h^{-1}$, oxygen utilization factor equaled to 26·1% at $D =$ $= 0{\cdot}36\ h^{-1}$. Due to different character of cultivation media used, comparison of these data with our experimental results can however serve only to a qualitative estimation of efficiency of the fermentor studied in our work.

Hydrodynamic Parameters

Preliminary experiments proved that in sulphite liquors based media, formation of the stable foam, inefficient for the mass transfer in the bed, was substantially reduced compared to cultivations on ethanol[1]. Consequently, formation of such a foam could be almost completely hindered by the apparatus arrangement used *i.e.* under conditions of combined bubble-spray regime. Significantly higher dynamic gas holdup was however observed in the system studied than in ethanol-based media causing thus low fermentor volume utilization. Understandably this dynamic holdup could not be reduced by bed-surface showering and anti-foaming agent was therefore dosed continuously during all cultication runs. Constant dosing rate ($16\ cm^3/h$)

was kept in all cultivation experiments. It was adjusted on the basis of preliminary test experiments and was not minimized in following cultivation runs. Related to the produced biomass, the rate of anti-foaming agent supply corresponded approximately to 45 cm^3/kg at maximum productivity. This dosing rate was considerably higher than in the case of fermentation on ethanol in the same unit and corresponded rather to the column fermentor without spraying device[1]. It can be explained by the character of gas–liquid bed described above *i.e.* by the ratio of dynamic and static gas holdup. In the presence of anti-foaming agent the gas-hold up varied according to experimental conditions between 0·22 and 0·37 or 0·35 and 0·44 for the first and second set of experiments respectively.

Table III summarizes steady-state $k_L a$ data for individual cultivation experiments. Each value of $k_L a$ in the table represents an average value of five measurements made during steady-state periods, mean variance of $k_L a$ data was 6·1%. It is obvious from Table III that, for both media used, $k_L a$ values were independent on the dilution rate and thus apparently also on values X and S which varied in individual experimental runs with D. It is also apparent that values of $k_L a$ were significantly higher for the optimized medium B than for medium A. The average value of $k_L a$ for medium A was 0·143 s^{-1} while the corresponding average value for the medium B was 0·198 s^{-1}, mean deviations from these average values were 12 and 4% for the first and second set of experiments respectively. Comparison with the value $k_L a =$ $= 0{\cdot}033\ s^{-1}$ obtained in the same unit and at the same superficial gas-flow rate ($w_G = 0{\cdot}015$ m/s) in the air-water system demonstrates significantly higher intensity of interfacial oxygen transfer in the fermentation broth. Due to the system complexity reliable identification of factors responsible for the $k_L a$ increase and for the difference of $k_L a$ values obtained in the two experimental sets performed with different media is not feasible on the basis of experimental evidence obtained in this work. An experimental study is therefore to be recommended aimed primarily at the determination of the effect of cultivation medium composition and of the concentration of biomas in the system on $k_L a$.

TABLE III

Steady-state $k_L a$ values for individual cultivation runs

D h^{-1}	$k_L a$, s^{-1} medium A	 medium B
0·10	0·129	0·212
0·16	0·171	0·197
0·18	0·127	0·184
0·20	0·127	0·205
0·23	0·159	0·195

Energy Effectiveness

Our former studies[8,9] proved that if a Venturi-tube type ejector was used as a gas distributor the quality of gas–liquid dispersion created in the unit and consequently the rate of interfacial mass transport in the bed depended on the overall rate of energy dissipation in the place of dispersion formation (*i.e.* in this case in the ejector) defined as

$$e_d = \Delta P_e Q_L / \dot{V}_L \varrho_L \,. \qquad (8)$$

According to definition e_d is related to a mass unit of the bed, ΔP_e denotes the ejector pressure drop. After substitution of appropriate experimental values $\Delta P_e = 0.12$ MPa, $Q_L = 0.194 \,.\, 10^{-2}\ m^3/s$, $\varrho_L = 1\,060\ kg/m^3$, $V_L = 0.101\ m^3$, value $e_d = 2.18$ W/kg was obtained from Eq (*8*) for the second set of experiments (for the first set $V_L = 0.089\ m^3$, $e_d = 2.47$ W/kg.) To compare the efficiency of different ways of gas–liquid bed formation, a theoretical ("internal") energy effectiveness was defined[9], $\Phi = k_L a / e_d$. According to its definition parameter Φ characterizes the efficiency of utilization of the dissipated energy from the viewpoint of demands on maximum intensity of interfacial contact in the bed. After multiplication by the mass transfer driving force (concentration gradient) parameter Φ represents the amount of transported mass on the unit of dissipated energy. Values $\Phi = 0.058$ kg/J and 0.091 kg/J were obtained for the first and second set of experiments respectively from corresponding average values of $k_L a$ ($k_L a = 0.143$ or $0.198\ s^{-1}$). Comparison of these values with values of energy effectiveness factor corresponding in the same unit to identical $k_L a$ values for the air–water system ($\Phi = 0.013$ or 0.0075) shows that under fermentation process conditions the dissipated energy is utilized very efficiently for the dispersion formation *i.e.* for creation of intensive interfacial contact. As can be seen comparison of energy effectiveness factors indicates even more pronounced difference of the fermentation system from the standard air–water system than just comparison of $k_L a$ values corresponding to equal superficial gas-flow rates.

One of the decisive criteria for the choice of proper type of fermentor for a specific process is the real energy supply (power input to the equipment) related usually to the amount of oxygen transferred or to the biomass production. In some cases also the specific energy consumption related to the liquid phase volume in fermentor, P_m, is reported as a fermentor characteristic, corresponding to appropriate intensity of interfacial mass transport (oxygen transfer), characterized by $k_L a$ value. In our previous work[1] criterion $k_L a / P_m$ was introduced for comparison of fermentors from the viewpoint of demands on maximum intensity of interfacial contact achieved at minimum energy consumption. Experimental values of specific energy input, P_m, for the first and second set of our present experiments were 6.7 a 5.9 kW/m^3 resp., corresponding values of $k_L a / P_m$ were thus 0.021 and 0.034 m^3/kWs for average

k_La values from both experimental sets. These reported values k_La/P_m are comparable with data reported by Sittig and Heine[10] for units of similar type — $k_La/P_m = 0{\cdot}028$ m^3/kWs for fermentor with a dispersing nozzle and internal central tube; $k_La/P_m =$ $= 0{\cdot}039$ m^3/kWs for the unit with an ejector gas distributor and internal central tube. Values of energy consumption related to a mass unit of oxygen transferred, $P_m/q(O_2)$, corresponding to maximum rates of oxygen transfer in both sets of cultivation experiments were 3·85 and 2·33 kWh/kg for the first and second set respectively. For comparison, energy consumption 1·75–2·0 kWh per kg of oxygen transferred has been reported[4] for a laboratory-scale Vogelbusch immersed-jet fermentor. In our case however the rate of cultivation was not limited by the oxygen transfer. Apparently thus the energy consumption related to biomass production was a more adequate criterion for energy effectiveness evaluation. Value of this criterion, P_m/p, was in our case 1·64 kWh/kg for maximum productivity achieved ($p = 3{\cdot}65$ kg/m^3 h). In the survey of Vogelbusch fermentors used industrially for various cultivation processes values of this criterion are reported[4] to be in the range 0·55–2·2 kWh/kg for fermentor volumes 0·3–400 m^3. The authors[4] also claim that the energy consumption related to biomass production decreases with increasing size of the unit and that it is consequently always higher in laboratory-scale and pilot-plant fermentors. Regarding this, values of energy characteristics achieved in our fermentor model can be considered as promising. It is also obvious from the graph in Fig. 4 that further increase of productivity and thus more favourable values of criterion P_m/p can be achieved with medium B at higher dilutation rates. The fact that the apparatus studied has not been optimized from the viewpoint of energy consumption and utilization by this state of research should be also borne in mind in evaluating its energy effectiveness. Reported values of energy characteristics has to be therefore considered only as preliminary estimates. Further increase of fermentor energy effectiveness can be apparently achieved by minimizing the power input of the circulation pump engine (using the properly sized engine) and by optimizing the Venturi-tube parameters due to fermentor size (to its efficient volume).

CONCLUSIONS

Extremely high content of gaseous phase in the bed exhibited by the fermentation system studied was formed predominantly by the dynamic gas holdup whereas the contribution of static-foam region located in the upper part of the bed was of minor importance only. The fermentation broth composition influenced favourably the k_La values which were under comparable conditions significantly higher than in the standard air–water system. No oxygen transfer limitation of the yeast growth was observed in the whole region of experimental conditions; this apparently points out to existing reserves for the fermentation process in the apparatus studied. Results of cultivation experiments proved positive effect of the optimized medium studied (medium B) on the parameters selected for cultivation process assessment *i.e.* on $k_L a$, biomas concentration, oxygen utilization, and overall cultivation productivity. Similarly, more favourable values of energy characteristics were achieved in experiments with the optimized medium, considering both theoretical effectiveness of bed formation and real efficiency of energy supply utilization.

Results of our experimental study has proved that the fermentors of the tested type can be suitably used for fermentation on sulphite liquors considering hydrodynamic characteristics of the unit and values of cultivation parameters achieved as well as the energy effectiveness of the fermentor. The analysis of experimental data further suggests that even more favourable values of individual energy characteristics can be expected after optimization of construction parameters of the apparatus due to energy consumption and due to utilization of dissipated energy for interfacial mass transport enhancement. Due to the specific properties of gas–liquid bed (high values of dynamic gas holdup ratio and only limited static foam formation) the additional bed showering did not influenced its character so significantly as *e.g.* in the case of fermentation on ethanol based media. Dosing of an anti-foaming agent was therefore inevitable in the whole range of experimental conditions to achieve proper fermentor volume utilization. In addition to the "energetic" optimization of fermentor construction parameters, as discussed above, further research steps should be thus aimed at the experimental determination of the effect of fermentation broth composition and of the presence of anti-foaming agent in the bed on the rate of interfacial oxygen transfer and at the determination of optimum process conditions for the optimized medium B.

LIST OF SYMBOLS

A	correction factor defined by Eq. (*4*)
a	specific interfacial area
c_L	concentration of oxygen dissolved in fermentation broth
c_L^+	equilibrium concentration of dissolved oxygen
D	dilution rate
D_K	fermentor diameter
d_s	ejector nozzle diameter
e_d	rate of energy dissipation related to a unit of liquid mass in bed
H	height of aerated liquid
H_0	height of clear liquid in bubble bed
k_L	liquid-side mass transfer coefficient
P	pressure
P_c	power input of the circulation pump motor
P_m	specific power input related to a unit of liquid phase volume
p	overall productivity of the fermentation process
ΔP_e	ejector pressure drop
Q_L	volumetric flow rate of medium through ejector

$q(O_2)$, $q(CO_2)$	rate of oxygen consumption (or CO_2 production) related to the volume of liquid phase in fermentor
R	gas constant
RQ	respiration coefficient
R_v	relative utilization of air oxygen
T	temperature
$\dot{V}_G$	volumetric flow rate of gas in fermentor
$\dot{V}_L$	volumetric flow rate of cultivation medium
V_L	volume of medium in the fermentor
w_G	superficial gas velocity
X	steady-state dry biomass concentration
$Y_{X/O}$	yield of biomass related to oxygen consumption
y_1, y_2	oxygen mole fractions in gaseous phase at the fermentor inlet and outlot
z_1, z_2	mole fractions of CO_2 in gaseous phase at the fermentor inlet and outlet
ε_G	relative gas holdup
Φ	energy effectiveness
ϱ_L	liquid phase density

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