APPLICATION OF TOWER BIOREACTORS WITH FORCED CIRCULATION IN BIOMASS PRODUCTION

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Characteristic hydrodynamic parameters of three selected types of bubble-type column reactors with forced liquid circulation were studied experimentally. Application of these units for the fermentation process — production of fodder protein by continuous cultivation of *Candida utilis* yeast on the synthetic ethanol was tested. Two modifications of the column reactor with the ejector gas distributor and the bubble type tower reactor with the central tube and forced internal circulation of the gas–liquid dispersion were selected. Diameter of all three reactors used was 0·3 m. The volumetric mass transfer coefficient ($k_L a$) and gas holdup (ε_G) were determined both for the standard water–air system and under actual fermentation process conditions. The effect of the inlet ethanol concentration and of the dilution rate on the cultivation characteristics was determined during the cultivation experiments. The obtained results have proved that the studied reactor types are suitable for fermentation. Equipment with the ejector gas distributor can be recommended even for processes with high requirements on the oxygen consumption, while by the proposed construction arrangement the undesired foam formation in the system was succesfully eliminated.

Application of bubble-type column reactors for aerobic biotechnological processes-first of all for fermentations, represents one of largest potential regions of their use. A number of experimental studies and survey papers have been published recently[1–4] where possibilities of intensification and improvement of economics of industrially performed fermentations were considered by use of bubble-type column reactors. In these studies prospects of these reactors for fermentation processes are discussed in detail, together with their advantages and disadvantages, regarding the assumed application. From the nature of the considered biologic processes (aerobic microbial cultivation) and from the demand of maximum economy of production of biomass, two basic requirements arise on fermentation reactors: *a*) maximum supply of oxygen into the system (reaching the maximum transport rate of oxygen into the liquid phase, *b*) the lowest possible energetic consumption.

From the engineering point of view and thus from the point of view of hydrodynamics of the considered system this means to endeavour after the highest possible liquid phase volumetric mass transfer coefficients $k_L a$ (characterising the rate of interphase mass transfer) at the minimum requirements on the energy supplied into the system. Since it is known from theory that the magnitude of interfacial area and mass transfer coefficients in the bubbled gas–liquid bed are directly related to the energy dissipated in the bed or in the point of formation of the gas-liquid dispersion[5-9] it is necessary in design of these units to try to reach an optimum solution in the frame of both above stated opposite requirements. In certain specific cases one of the above mentioned requirements can prevail; *e.g.* if the process considered requires the maximum supply of oxygen or, on the contrary, if the requirement of maximum savings in energy will force construction of a less efficient device even if at the costs of larger dimensions. It is also necessary to take into consideration that requirement of the largest possible supply of oxygen into the system need not in all cases appear as imperative, considering different fermentations (according to the demands of used substrate). With regard to all these facts and to relatively wide spectrum of other requirements put on the unit for individual processes it is natural to consider various modifications of bubble type reactors so that the final design of the unit satisfy as well as possible the requirements of the process. Beside the classical bubble type column reactor (single and multi-stage) with gas distribution by perforated or porous plates, reactors with forced liquid circulation[10-12] are frequently considered. In these reactors sufficient supply of oxygen as well as other positive properties are assumed as regards the undergoing process *i.e.* the relatively well defined streaming character of both phases in the system, suppresion of formation of the undesired foam *etc.* These units are more complex as concerns the design than classical bubble type reactors while it is obvious that their efficiency and energy consumption considerably depend on construction parameters of the unit.

A sufficient set of design information for an optimal design of these units is thus the first assumption for their eventual application. In this study three types of bubble type column reactors with forced circulation were tested. The aim was to obtain information on characteristic hydrodynamic parameters of these units and to verify the possibility of their application for a specific fermentation process-production of the fodder protein by continuous cultivation of the yeast *Candida utilis* on synthetic ethanol.

EXPERIMENTAL

Equipment and Procedure

All three types of used reactors are schematically illustrated in Fig. 1 (*A*, *B*, *C*). The equipments *A* and *B* represent the different construction arrangements of the reactor in which an ejector of the Venturi-tube type is used for gas distribution (for formation of dispersion). In the reactor *B* has been separated, in the circulation loop of the ejector at the pressurised outlet of the pump, a side-stream of liquid medium by which was sprayed from above the surface of the gas-liquid bed[13]. The third type of equipment is the modification of the reactor with oriented internal circulation (air-lift). In the proposed arrangement the gas has been fed into the annular space between the wall of the reactor and the inbuilt central tube in which was situated an axial screw mixer. Its task was to induce the circulation of the gas-liquid dispersion in the system.

The experiments with the first two types of equipments (reactors *A*, *B*) have been linked to our earlier studies devoted to application of ejector distributors in bubble-type reactors[14-15]. The measurements in the reactor *A* have been used for verification of suitable application of this method of gas-liquid bed formation in fermentations, *i.e.* in systems with complex liquid medium containing further the dispersed solid phase. The arrangement of the type B has been proposed on basis of our previous experience to eliminate formation of the undesired static foam whose existence negatively affects efficiency of the unit. Similarly, also the proposed modification of the air-lift reactor (equipment *C*) has been aimed first of all to suppression of the formed foam which is in this arrangement sucked into the central cylinder. Prolongation of the residence time of gas in the unit due to forced circulation of the gas phase contributes also to improvement of utilization of oxygen in this type of reactor.

The diameter of all three column reactors was 0·29 m, height of the bed in reactors has been fixed by use of a downcomer. The effective volume of reactors was 0·070 (reactor *A*), and 0·131 m^3 (reactors *B* and *C*). The inbuilt central tube in the unit *C* had the diameter 0·16 m. It was situated in the height 0·06 m above the bottom of the reactor and its upper end reached 0·05 m below the overflow pipe of the reactor. The screw mixer had two threads with lead 0·08 m, outside diameter of the mixer was 0·12 m, diameter of the axis 0·035 m. The mixer was situated in the bottom part of the central tube. For formation of dispersion in reactors *A* and *B* the Venturi tube with the nozzle diameter 0·006 m was used. For optimal function of the spraying device in the reactor *C* about 10% of the total flow rate of the circulating medium was separated as the spraying jet. Other details on the used units and their detailed drawings can be found in the thesis of Klekner[16], Bydžovský[17] and Mikyška[18].

The values of characteristic hydrodynamic parameters of the bubbled bed, volumetric mass transfer coefficient ($k_L a$) and gas holdup (ε_G) for the standard air–water system were determined in all three units. Similar characteristics were obtained during fermentations for complex fermentation medium containing live yeast. In all three studied units, within the cultivation experiments were observed the effect of inlet concentration of ethanol and dilution rate (defined as the ratio of cultivation medium flow rate to the volume of liquid phase in the bed), on parameters of the cultivation process. In these measurements the concentration of the solids in the medium, concentrations of oxygen in the liquid phase and concentration of oxygen and carbon dioxide in the gas phase at the outlet from the reactor were recorded. After reaching the steady state the ethanol and acetic acid concentrations were also determined in the medium leaving the reactor. On basis of these data the values of other characteristic cultivation parameters *i.e.* of the respiration coefficient, rate of oxygen consumption and formation of CO_2, utilization of air oxygen, yield of biomass related to the substrate and total productivity were calculated. All cultivations were performed with the yeast *Candida utilis*. Table I gives the composition of the cultivation

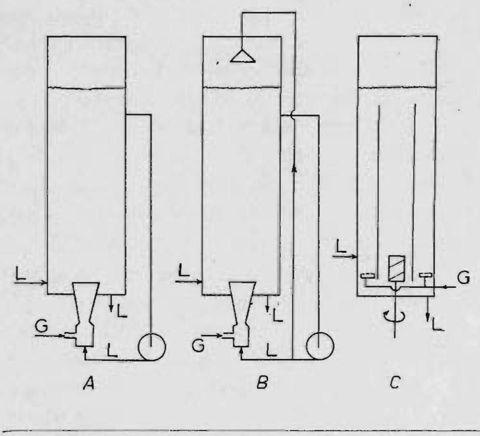

FIG. 1
Used reactor types

medium for one selected ethanol concentration (10 kg/m^3). The medium was not sterilised, storage tanks for preparation of the cultivation medium and the fermentor were before the experiments sterilized by 50% solution of ethanol or 3% solution of formaldehyde. These units were temperred for cultivation experiments and equipped with automatic control of pH and in all cultivations steady conditions ($t = 30°C$, pH = 4·0) were kept in the reactors. Experimental conditions under which the cultivation experiments have been performed in individual units are summarised in Table II, other details concerning the methods and arrangement of the cultivation experiments can be again found in the above mentioned thesis[16–18].

Measuring Methods

Hydrodynamic parameters. In the water–air system the k_La values were determined by the dynamic method based upon evaluation of the response of the oxygen probe to saturation of the reactor volume by oxygen from air. Liquid in the reactor was deprived of dissolved oxygen by desorption by nitrogen in advance. For measurement of the oxygen concentration was used the polarographic probe of the Clark type produced in the Research workshops of Czechoslovak Academy of Sciences with the teflone membrane of thickness 0·015 mm. At evaluation of the probe response was assumed ideal mixing of the liquid phase in the bed, for description of behav-

TABLE I

Composition of fermentation broth for ethanol concentration 10 kg/m^{-3}

Compounds	c, kg m^{-3}
$(NH_4)_2SO_4$	3·45
$NaHPO_4.12\ H_2O$	1·62
K_2SO_4	0·39
$MgSO_4$	0·11
$CaCl_2.2\ H_2O$	0·10
fermentation autolysate	0·10

TABLE II

Range of experimental conditions at cultivation experiments

Unit type	$F \cdot 10^3$ $m^3\ h^{-1}$	S_0 kg m^{-3}	D h^{-1}	$\dot{V}_G \cdot 10^3$ $m^3\ s^{-1}$
A	8·8—28·7	8·0—30·6	0·131—0·429	2·22—2·78
B	19·9—39·8	7·0—17·4	0·182—0·365	2·08—2·14
C	15·9—46·3	7·6—25·2	0·121—0·353	2·22

iour of the probe was used the two-regional model according to Linek[19]. Details concerning the methodics of measurement and evaluation of data can be found in the report of Kratochvíl[20]. For comparison the values $k_L a$ were in some experiments determined also by the classical Winkler titration method. Under the fermentation conditions the values of $k_L a$ were calculated on basis of the oxygen balance in the system from relation

$$k_L a = q(O_2)/(c_L^+ - c_L) \tag{1}$$

into which was substituted for equilibrium oxygen concentration the value $c_L^+ = 7{\cdot}615 \cdot 10^{-3}$ kg . . m^{-3}, taken from the paper of Páca and Grégr[21]. The form of Eq. (*1*) corresponds to ideal mixing of the liquid phase and to the case when the dissolved oxygen is consumed only by cells. The rate of oxygen consumption in the bed, $q(O_2)$ was determined from the difference of the inlet and outlet oxygen concentration on air; these values were measured by the oxygen analyser Permolyt 2 (Junkalor Dessau). Oxygen concentration dissolved in the fermentation medium, c_L, was continuously measured by the polarographic probe of the Clark type and was registered by the recorder TZ 21S.

Gas holdup in the bed, ε_G, was both in the water–air system and under the fermentation conditions determined from the difference of heights of clear liquid and of the bubbled bed.

Parameters of cultivation. Dry biomass content was determined turbidimetricaly on the spectra colorimeter Spekol (Carl Zeiss, Jena), after reaching the steady state was performed the controlling gravimetric measurement. The outlet oxygen concentration was measured by the oxygen analyser Permolyt 2, concentration of CO_2 was determined by the infrared analyser Infralyt-5 (Junkalor, Dessau). The outlet ethanol and acetic acid concentration in the fermentation medium were determined chromatographically on the gas chromatograph CHROM 3 with the flame-ionisation detector. The value of the respiration quotient RQ, for the measured concentrations of O_2 and CO_2 in the outlet and inlet gas was read off the nomogram published by Fiechter and Meyenburg[22]. On basis of the found concentration differences $(y_1 - y_2)$ or $(z_2 - z_1)$ for oxygen or CO_2 were then calculated the total rates of oxygen consumption or of CO_2 formation related to the volume of the liquid medium $q(O_2)$ or $q(CO_2)$ and the relative utilization of oxygen from air R_v from relations

$$q(O_2) = [y_1 - (1 - A)\, y_2]\, M(O_2) P \dot{V}_G / V_L \boldsymbol{R} T \tag{2}$$

$$q(CO_2) = [(1 - A)\, z_2 - z_1]\, M(CO_2) P \dot{V}_G / V_L \boldsymbol{R} T \tag{3}$$

$$R_v = 100(y_1 - y_2)/y_1\,, \tag{4}$$

where A in Eqs (*2*) and (*3*) is the correction factor defined as

$$A = (y_1 - y_2) + (z_1 - z_2)\,. \tag{5}$$

Yield of biomass related to the substrate $Y_{X/S}$ and total productivity p were for individual fermentation regimes, characterised by dilution rate $D(D = F/V_L)$ and the inlet ethanol concentration S_0, calculated from experimental data of dry substance content in steady state X and the outlet ethanol concentration S from relations:

$$Y_{X/S} = X/(S_0 - S) \tag{6}$$

$$p = D\,.\,X\,. \tag{7}$$

RESULTS AND DISCUSSION

Hydrodynamic Parameters

In Table III are given values of ε_G and k_La obtained for the system water–air in the equipment with the ejector distributor. As in these experiments the spraying of the bed has not been used, these data hold for both reactors A and B. Analogical data obtained in the unit C for the speed of the stirrer 1 700 min^{-1} are presented in Table IV. Comparison of relation between the values of gas holdup and k_La in reactors with the ejector distributor, or with the forced internal circulation (reactor C) indicates that a large gas holdup in the unit C is partly formed by bubbles with reduced oxygen concentration. From Table IV is also obvious a comparison of data k_La obtained by the dynamic method by use of the oxygen probe and values determined by the titration method according to Winkler, when for deoxidation of water Na_2SO_3

TABLE III

Experimental data ε_G and k_La for water–air system in units A and B

$Q_L . 10^3$ $m^3 s^{-1}$	$\dot{V}_G . 10^3$ $m^3 s^{-1}$	w_G $m s^{-1}$	ε_G %	k_La s^{-1}
0·945	1·667	0·025	9·6	0·055
1·055	2·278	0·034	13·5	0·078
1·167	2·555	0·038	16·1	0·093
1·278	2·833	0·042	17·5	0·120
1·388	3·000	0·045	18·3	0·142

TABLE IV

Experimental data ε_G and k_La for water–air system in unit C (1 700 min^{-1})

$\dot{V}_G . 10^3$ $m^3 s^{-1}$	w_G ms^{-1}	ε_G %	$(k_La)_{dyn}$ s^{-1}	$(k_La)_{titr}$ s^{-1}
1·111	0·017	16·1	0·042	0·050
1·389	0·021	18·3	0·046	0·053
1·667	0·035	19·1	0·049	0·059
1·944	0·029	20·0	0·055	0·076
2·222	0·033	20·9	0·058	0·085

has been used. Higher values k_La obtained by this latter method correspond to the expected effect of presence of sulphite ions and character of the bed. According to our experience[20] it is not possible to use the chemical method in systems with higher values of k_La and thus with fast rate of oxygen saturation.

Under the fermentation conditions the values of studied hydrodynamic parameters of the bed (k_La, ε_G) are affected by composition of the medium in the reactor namely by concentration of added inorganic salts and by the value of the residual ethanol concentration, and by the biomass content in the reactor *i.e.* by factors changing with the conditions of the fermentation process. The values k_La and ε_G obtained in individual units during the cultivation experiments were thus dependent on values of the inlet ethanol concentration and on dilution rate *i.e.* on independent variables determining the cultivation conditions, and on the growth ability of used cultures. For comparable gas velocities ($\dot{V}_G = 2{\cdot}22 \,.\, 10^{-3}\ m^3\ s^{-1}$ for reactors A and C, and $\dot{V}_G = 2{\cdot}08 - 2{\cdot}14 \,.\, 10^{-3}\ m^3\ s^{-1}$ for the reactor B) the values k_La measured in individual units within the range of experimental conditions of performed cultivations (Table II) were within the limits $0{\cdot}105 - 0{\cdot}174\ s^{-1}$ (reactor A) $0{\cdot}056 - 0{\cdot}140\ s^{-1}$ (reactor B) and $0{\cdot}044 - 0{\cdot}097\ s^{-1}$ (reactor C). The magnitude of these ranges and comparison of given values of k_La with the data obtained at similar hydrodynamic conditions ($\dot{V}_G$, n) in the water–air system (Tables II and IV) confirm a considerable effect of ethanol and biomass concentrations in the medium on values of k_La. In agreement with other authors[23] it is thus necessary to warn against mechanical transfer of the data obtained experimentally in model systems onto biologic media.

The effect of cultivation conditions on values of gas holdup is illustrated in Table V on an example of data from the unit B. The values ε_G measured in all three units at the fermentation conditions were, in agreement with the literature data[1–2], significantly higher than the corresponding data for the system water–air. Comparison of experimental data of ε_G and k_La obtained in the water–air system and under conditions of cultivation experiments indicate that the value ε_G and thus obviously of the interfacial area are growing due to the presence of the surface active compounds and by action of other factors determining the behaviour of bubbled beds at fermentation conditions faster than the values of k_La. It is thus possible to conclude that the growth of ε_G or a takes place partially at the account of the decrease of k_L, caused by the decrease of microturbulence in the bed. The given results correspond to our data obtained for various bubbling regimes in the water–air[24] system where a considerable increase in the gas holdup at the regime of homogeneous bubbling ("foam" regime) was not accompanied by the corresponding rise in k_La.

Cultivation Parameters

Within the cultivation experiments the effect of the inlet ethanol concentration (S_0) and the dilution rate (D) on the fermentation process *i.e.* on values of individual parameters of the process has been studied. All dependent variables whose values were obtained at cultivation experiments for different pairs of values S_0 and D were summarised in Table VI. In Figs 2 and 3 is then given the typical example of the primary graphical treatment of results demonstrating the effect of S_0 and D on individual studied parameters. As the decisive factors for considerations on the efficiency and economics of the fermentation process in individual units can be considered the yield of biomass related to the consumption of substrate $(Y_{X/S})$, concentration of biomass in the reactor (X), total productivity of the process (p), relative utilisation of oxygen (R_v) and the residual ethanol concentration in the medium at the outlet from the reactor (S).

The average yield of biomass related to the consumed substrate was in all three used units for the whole studied range of value S_0 and D approximately constant, while there were not found significant differences between the average values $Y_{X/S}$ in individual reactors ($Y_{X/S} = 0{\cdot}690$; $0{\cdot}711$; $0{\cdot}698$ for reactors A, B or C).

The relative utilisation of the oxygen was in all three cases relatively low, the average values were 10·4% for the unit A, 9·4% for the unit B and 9·5% for the unit C which point to considerable reserves of individual reactors in this respect. Beside the possible construction arrangements such as are increase of the bed height or use of the multistage arrangement of reactors A or B there is a possible way toward greater utilization of oxygen through the increase of values S_0 and D. However the problem of selection of these independent variables has to be considered in connection with all factors characterising the efficiency and economics of fermentation. As limiting factors

TABLE V

Gas holdup in dependence on fermentation conditions in unit B $w_G = 0{\cdot}031$—$0{\cdot}032\ \mathrm{m\,s^{-1}}$

S_0 kg m^{-3}	D h^{-1}	S kg m^{-3}	X kg m^{-3}	ε_G —
7·0	0·185	0	4·57	0·145
7·05	0·246	0	4·57	0·145
7·8	0·182	0	5·83	0·145
9·2	0·365	0·02	6·42	0·168
13·7	2·212	0·04	10·2	0·282
13·6	0·301	0·07	9·95	0·282
15·0	0·221	0·11	10·6	0·305
14·9	0·313	0·23	10·5	0·313
17·4	0·239	0·74	11·8	0·359
17·1	0·345	2·12	11·3	0·374

TABLE VI

Dependent variable parameters recorded in cultivation experiments

Measured quantities	S kg m^{-3}	X kg m^{-3}	c_L kg m^{-3}	y % vol.	z % vol.	
Calculated quantities	D h^{-1}	$Y_{X/S}$ kg_B/kg_{et}	$Y_{X/O}$ kg_B/kg_{O_2}	RQ —	$q(O_2)$ kg_{O_2} m^{-3} h^{-1}	$q(CO_2)$ kg_{CO_2} m^{-3} h^{-1}
Calculated quantities	q_{et} kg_{et} m^{-3} h^{-1}		R_v %	p kg_B m^{-3} h^{-1}		

can be active in this respect *e.g.* the total productivity or the residual ethanol concentration. The maximum value R_v reached in the unit B at conditions complying with appropriate limitations was 15% (for $S_0 = 14{\cdot}9\ \text{kg/m}^3$, $D = 0{\cdot}31\ \text{h}^{-1}$; the corresponding values of productivity and residual ethanol concentration were $p = 3{\cdot}3\ \text{kg/m}^3\ \text{h}$, $S = 0{\cdot}23\ \text{kg/m}^3$).

In Figs 4 and 5 effects of S_0 and D on concentration of biomass, total productivity and residual ethanol concentration are demonstrated. From these graphs is obvious an agreement of data obtained in reactors A and B which correspond to the identical way of formation of the bed in both these units. Lower values X and p and considerably larger values S measured in the same range of independent variables in the reactor C can be also partially affected by lower activity of the yeast culture used in experiments in this reactor[17]. From these graphs is obvious a relatively fast increase in the residual ethanol concentration observed in reactor A and B after exceeding some limiting values S_0 and D. With regard to the fact that the expenses on substrate form a significant part of total production costs, it is necessary to respect the requirement of low outlet ethanol concentration in selection of operating conditions in the unit

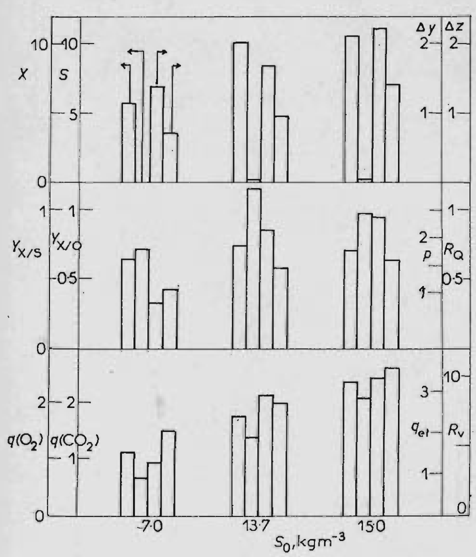

FIG. 2

Effect of initial ethanol concentration on selected cultivation parameters. Unit B, $D = 0{\cdot}21\ \text{h}^{-1}$

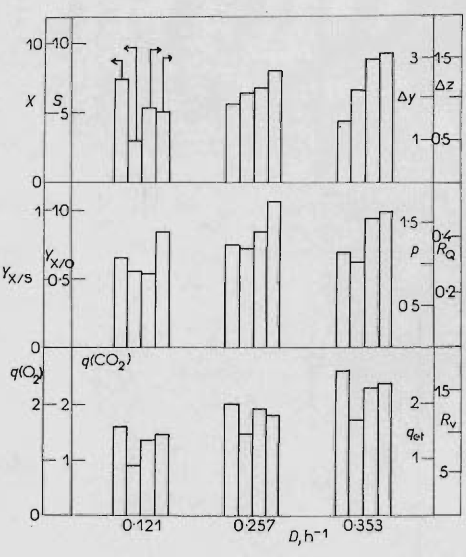

FIG. 3

Effect of dilution rate on selected cultivation parameters. Unit C, $S_0 \approx 14\ \text{kg m}^{-3}$

operating without recirculation of media. The maximum productivity reached in the reactor B at the usually specified condition of economically feasible residual ethanol concentration $S_{max} \leqq 0.5\ kg/m^3$ was $p = 3.3\ kg/m^3\ h$ (for $S_0 = 14.9\ kg/m^3$, $D = 0.31\ h^{-1}$). For conditions at which a higher productivity was reached $p = 3.9\ kg/m^3\ h$ ($S_0 = 17.1\ kg/m^3$, $D = 0.34\ h^{-1}$) was the residual ethanol concentration $2.12\ kg/m^3$ and at the conditions corresponding to the maximum reached productivity $p = 5.5\ kg/m^3\ h$ ($S_0 = 26.6\ kgm^3$, $D = 0.35\ h^{-1}$) the residual ethanol concentration rose to $7.3\ kg/m^3$.

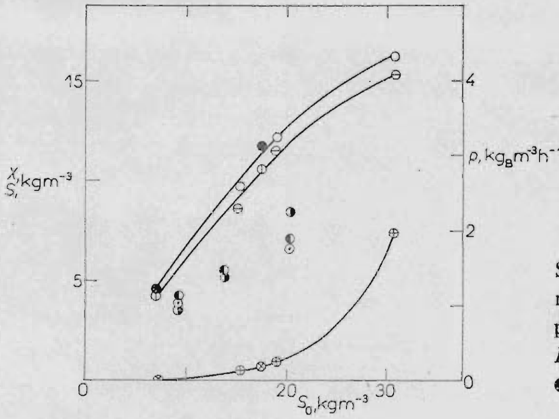

FIG. 4

Steady-state dry biomass concentration, ethanol concentration and productivity in dependence on inlet ethanol concentration $D = 0.25\ h^{-1}$. Parameter X ○ unit A, ● B, ◐ C; S ⊕ A, ⊗ B, ◑ C; p ⊖ A, ⦶ B, ⊙ C

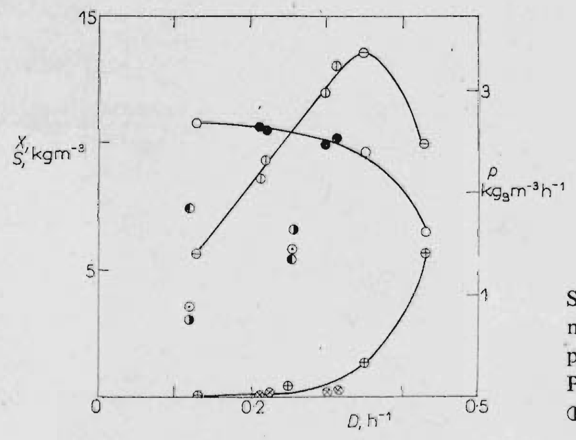

FIG. 5

Steady-state dry biomass concentration, ethanol concentration and productivity in dependence on dilution rate. $S_0 \approx 15\ kg\ m^{-3}$. Parameter X ○ A, ● B, ◐ C; S ⊕ A, ⊗ B, ◑ C, p ⊖ A, ⦶ B, ⊙ C

In the reactor C the maximum productivity was reached $p = 1{\cdot}83\ \text{kg/m}^3\ \text{h}$ for $S_0 = 20{\cdot}4\ \text{kg/m}^3$, $D = 0{\cdot}26\ \text{h}^{-1}$. The residual ethanol concentration was under these conditions $9{\cdot}6\ \text{kg/m}^3$, while in the whole measured range of values S_0 and D the values S did not decrease below $1{\cdot}04\ \text{kg/m}^3$.

Comparison of Used Units

Experiments in the unit A have confirmed, that the use of the ejector distributor enables to reach relatively homogeneous fine gas dispersion in the bed which ensures a relatively big supply of oxygen into the liquid phase. The k_La values determined in the cultivation experiments in this unit have been within the limits 0·105 to $0{\cdot}174\ \text{s}^{-1}$. According to expectations the negative feature of this type of distributor in fermentation media was a considerable formation of a stable foam layer on the surface of the bubbled bed. Existence of this foam decreases the effective volume of the reactor and deteriorates the parameters of cultivation. To prevent the foam formation it was necessary to charge into the system considerable amount of defoaming agent (according to the type of used defoaming agent this was 30 to $180\ \text{cm}^3/\text{kg}$ of produced cell solids). It is known that the defoaming agents decrease the rate of oxygen transfer into the liquid and also negatively affect the physiological state of yeast cells thus negatively affecting the cultivation process.

Both units designed with respect to suppression of undesirable formation of the foam have proved that the construction arrangement is suitable. In the unit C with the central pipe and forced internal circulation the consumption of defoaming agent according to the cultivation conditions was $10-25\ \text{cm}^3/\text{kg}$, in the combined spray reactor B it was necessary to add the defoaming agent only at the ethanol concentration in the feed above $13{\cdot}7\ \text{kg/m}^3$ at the rate $3-5\ \text{cm}^3/\text{kg}$. In both cases the defoaming fat R 05 from SETUZA, Ústí nad Labem was used.

The comparison of energetic requirements of individual studied reactors and k_La values which were there reached are given in Table VII. The corresponding values of the specific input (P_m) and k_La given in Table VII were obtained at cultivation experiments under comparable conditions in individual units. From comparison of similar data published for different industrially tested types of fermentors *e.g.* by Sittig and Heine[10] results that the unit C was as concerns the k_La values and corresponding specific power input comparable with units of similar type, while in the units A and B the needed specific power input was slightly higher then is given in literature for reactors in which comparable k_La values have been obtained. It is however possible to assume that the Venturi tube used as the distributor in the reactors A and B would enable with its parameters further increase of the effective reactor volume which would obviously lead to the decrease of the specific power input without significant deterioration of hydrodynamic parameters of the bed (see comparison of reactors A and B with effective volumes 0·070 or $0{\cdot}131\ \text{m}^3$).

Recently performed testing experiments point to a similar possibility of decrease of the specific power input even for the reactor C. Values of specific inputs given in Table VII for reactors A, – and C should be therefore treated only as preliminary estimates.

For an objective evaluation of suitability of the studied reactors from the view of postulated requirements of maximum oxygen transfer at minimum energy consumption the parameter k_La/P_m has been defined, characterising the efficiency of utilisation of the power input supplied into the unit. From values of these ratios given in Table VII is obvious that the unit C is less suitable than the units A and B regarding the optimal utilisation of energy.

CONCLUSIONS

The performed experimental study has proved the possibility of application of the tested types of reactors for microbial cultivations on non-classical sources of carbon. The proposed unit C with the central tube and forced internal circulation of the gas–liquid dispersion can be recommended on basis of the obtained results for application in systems with lower requirements on oxygen delivery. In comparison with units of similar type and similar hydrodynamic parameters (k_La, P_m) the advantage of the used arrangement can be seen in the considerable limitation of foam formation enabling to reduce the undesirable charging of defoaming agents into the system.

The experiments in units A and B have proved that the ejector used as a gas dispersing device is capable to supply sufficient amount of oxygen into the system even for processes with relatively high requirements on oxygen consumption. The construction arrangement used in the reactor B has practically eliminated the undesirable formation of foam. The way toward the decrease of the energetic consumption of units of this type can be seen both in determination of the optimum size of the used ejector distributor for the given fermentor and in detailed measurement of the effect

TABLE VII

Comparison of studied fermentor types as concerns the energy requirements

Type of unit	P_m kW m^{-3}	k_La s^{-1}	k_La/P_m kW m^{-3} s^{-1}
A	12	0·158	0·0132
B	10	0·142	0·0142
C	6	0·058	0·0097

of the type of used ejector and its geometrical construction parameters on efficiency of the ejector and on the character of the dispersion it is producing. From our testing experiments and from data of other authors it can be concluded that utilisation of reserves which exist in this respect could make this type of unit even more competitive. As concerns the considerations on wide application of these reactors, performance of further experimental studies has to be recommended, aimed at obtaining bases for the estimate of limits of application of ejector distributors (*e.g.* of the Venturi tube type) in large-scale units. Beside studies on scaling-up of ejector distributors it would be suitable to consider in this respect even the alternative simultaneous application of several ejectors in large diameter units.

LIST OF SYMBOLS

A	correlation term defined by Eq. (*5*)
a	specific interfacial area
c_L	concentration of oxygen dissolved in the liquid phase
c_L^+	equilibrium concentration of dissolved oxygen
D	dilution rate
F	volumetric flow rate of medium
k_L	liquid side mass transfer coefficient
k_La	liquid side volumetric mass transfer coefficient
M	molecular mass
n	speed of stirrer
P	pressure
P_m	specific power input supplied into the reactor
p	total productivity of the fermentation process
Q_L	volumetric liquid flow rate in the circulation loop of ejector
$q(O_2)$, $q(CO_2)$	rate of oxygen consumption (or CO_2 production) related to the volume of liquid phase in the reactor
q_{et}	rate of ethanol consumption related to the volume of liquid phase in reactor
$\mathbf{R}$	gas constant
R_Q	respiration coefficient
R_v	relative utilization of air oxygen
S	residual ethanol concentration
S_0	inlet ethanol concentration
T, t	temperature
V_L	volume of liquid phase in reactor
$\dot{V}_G$	volumetric gas flow rate in reactor
w_G	superficial gas velocity in reactor
X	steady state dry biomass concentration
$Y_{X/S}$	yield of biomass related to consumed substrate
$Y_{X/O}$	yield of biomass related to consumed oxygen
y_1, y_2	mole fractions of oxygen in gaseous phase at the inlet or outlet of reactor
z_1, z_2	mole fractions of CO_2 in the gaseous phase at the inlet or outlet of reactor
ε_G	relative gas holdup in the system

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