DESIGN AND SCALE-UP OF VENTURI-TUBE GAS DISTRIBUTORS FOR BUBBLE COLUMN REACTORS

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General principles of ejector distributors performance are surveyed and demonstrated for two particular cases of Venturi tubes commonly employed for gas dispersion in tower reactors with forced liquid circulation. Design recommendations for the two types of Venturi-tube gas distributors are presented and a general method is outlined for ejector distributors scale-up, based on the decisive effect of energy dissipation rate on the distributors performance. As an illustration, the specific case of Venturi-tube gas distributor design for an industrial reactor for catalytic hydrogenation of rape-seed oil is treated in detail. The procedure included design of a small-scale laboratory reactor for kinetic experiments at real process conditions (scale-down step) and subsequent ejector distributor scale-up to dimensions corresponding to the industrial reactor (vessel diameter 1·6 m, effective reactor volume ~ 5 m^3). Comparison with other modes of gas dispersion proved superiority of Venturi-tube distributors both on the laboratory- and industrial--scale level, regarding the overall rate of reaction process achieved and/or catalyst load requirements.

Various types of ejectors (two-phase nozzles) have been increasingly used as gas distributing devices in bubble-bed tower reactors for chemical and biochemical processes, namely in cases when the interfacial mass transfer is the rate controlling step of the reaction process. It has been established that the ejector distributors utilizing the kinetic energy of the liquid jet for dispersion of gaseous feed can ensure high intensity of interfacial mass transfer in two- and three-phase aerated beds and thus bring a sufficiently large amount of gas reactant into the contact with the liquid phase. Unlike in the case of other gas distributing devices commonly used in gas-liquid reactors, the decisive part of energy is in this case supplied for the liquid phase circulation through the ejector while gas can be either fed to the ejector under pressure (forced-supply operation mode) or can be sucked into the ejector due to the pressure decrease in its expansion (suction) chamber (free-suction operation mode).

It has been the aim of the present paper to review general principles of ejector distributors performance and to formulate basic rules for ejector distributors design and scale-up. In this respect, special attention has been paid to the Venturi-tube type ejectors. Due to their compact design and superior distributing performance, gas distributors of this type have been widely used in industrial units and accordingly they have been receiving extensive coverage in literature[1-8]. In this paper, two basic modification of Venturi-tube gas distributors are compared (Fig. 1) and recommendations for their optimum design are presented based upon data from literature as well as on the results of our own experiments. The procedure of ejector distributors design from laboratory data is demonstrated for the practical case of a Venturi-tube gas distributor for an industrial reactor for catalytic rape-seed oil hydrogenation.

Gas Dispersion with Ejector Distributors

Mechanism of gas–liquid dispersion formation in units with Venturi-tube gas distributors can be in principle deduced from the analysis of gas dispersion with two-phase nozzles presented by Kürten and Maurer[9] and by Schügerl and co-workers[10,11]. Liquid phase is pumped to the ejector nozzle and fast liquid jet formed by the nozzle enters the suction chamber to which gas phase is simultaneously fed or sucked due to the pressure decrease at the nozzle outlet. From there, gas flows to the momentum exchange tube or directly to the diffuser (see Figs 1a or 1b, respectively) where large bubbles are formed at first. The fast liquid jet having pulsating character decomposes in the momentum exchange tube (or diffusor) into fine droplets which hit the gas–

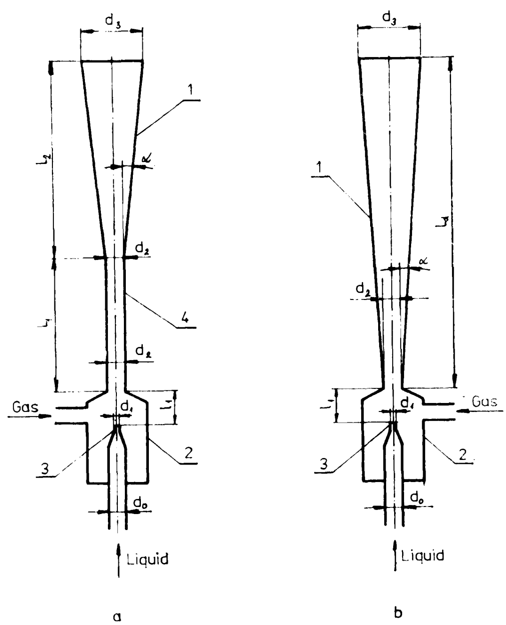


FIG. 1

Two types of Venturi-tube ejector distributors, a VT-1: modification with the mixing tube, b VT-2: modification without the mixing tube; 1 diffuser, 2 suction chamber, 3 nozzle, 4 mixing tube

–liquid interface (bubbles surface) causing subsequently its break-up. Such formed small bubbles are then further disintegrated into even finer bubbles due to highly turbulent dynamic pressure in the momentum exchange tube and/or in the diffuser. Resulting two-phase jet having high kinetic energy then enters the liquid layer in the reactor and having penetrated a certain distance into the surrounding liquid loses its energy and breaks up into cloud of bubbles. Apparently, this two-step mechanism of bubble dispersion formation resulting in the extremely small size of primary bubles and at the same time in the large extent of turbulence induced in the gas liquid layer explains large efficiency of gas phase utilization observed in our former study[12]. Comparison with sieve-tray bubble columns proved significantly higher values of gas holdup and k_La achieved in the ejector distributor reactor at constant values of superficial gas velocity. It is apparent that the positive effect of ejector distributors on the quality of gas dispersion is of essential importance namely in non-coalescing systems in which the size of primary bubbles (and thus the large interfacial area) created by these distributors remains constant during their rise through the bed. On the other hand, the improvement of gas dispersion due to the ejector performance is of lesser importance in coalescence promoting systems as the equilibrium bubble size establishing in such systems as a result of bubble coalescence and break-up processes is always significantly larger than the size of primary bubbles.

It has been proved that the enhancement of the mass transfer and high efficiency of gas phase utilization in ejector–distributor reactors are generally achieved at the expense of a considerable increase of energy supply (both absolute and relative) and consequently the total energy effectiveness of phases contacting in such units is appreciably lower than e.g. in sieve-tray bubble column reactors[12]. The final decision on the application of ejector-type gas distributors should be therefore always based on the consideration of all related process aspects as well as of the investment and operating costs of various reactor (gas distributor) types[13]. As to the operating costs, sieve-tray bubble columns are apparently superior to ejector–distributor reactors in cases when pressurized gas phase is available from a previous operation step (i.e. when gas supply does not require any additional power input). In other cases it can be however often more suitable to supply the energy to the reactor by liquid pumping instead of gas pumping or compressing[13]. Understandably, all these considerations have to be preceded by the initial estimation of the effect of mass transfer on the overall (effective) rate of the particular reaction process (supposing indeed that information on the intrinsic reaction kinetics is available) i.e. by the estimation of possible reaction rate enhancement due to mass transfer intensification. Apparently, the largest profit stemming from the application of Venturi-tube gas distributors can be expected in cases of very fast and instantaneous reactions. Also, these gas distributors have a promising application area in bioreactors for aerobic cultivation processes with large demands on the rate of oxygen supply.

Design and Performance of Venturi-tube Gas Distributors

Various variants of Venturi-tube installation and internal reactor arrangement are shown in Fig. 2. The choice of an appropriate reactor modification has to be based on considerations of specific features of each particular process (required rate of gas reactant transport to the liquid phase, foaming or non-foaming system nature, demands on the gas utilization degree etc.) and/or of operation limitations (e.g. due to safety precautions, type and capacity of the pump available etc.). Flow diagram of the basic and most commonly used reactor arrangement is given in Fig. 3 from which three alternative operation modes of ejector–distributor reactors are clearly apparent. As can be seen from the scheme, liquid phase can be either continuously pumped to the ejector nozzle from an external storage tank (direct liquid feed) or, most commonly, withdrawn from the reactor (usually at the bottom) and circulated through the ejector in a closed loop. Apparently the advantage of the latter operation mode consists in the fact that the liquid flow rate through the ejector and thus consequently the ejector distribution performance and efficiency are independent of the liquid feed rate determined by the reaction kinetics. Obviously, the liquid circulation mode is the only possible one in semi-batch reactors (i.e. at zero net liquid throughput), it can be however employed even in continuous-flow reactors. In this latter case, the liquid phase has to be fed to and withdrawn from the reactor at separate feed and exit points, respectively. The circulation rate is usually much higher than the net liquid throughput[14]. Similarly, gas phase can be either supplied to the ejector mixing chamber from an external source (fed under pressure or sucked from the atmosphere) or alternatively, it can be sucked from the space above the gas–liquid bed (gas circulation mode). Apparently, such an arrangement further increases the efficiency of gas phase utilization without any additional demands on energy supply and also it enables higher-than-stoichiometric amount of gas to be used for the interfacial area formation without losses of the active gas component. Indeed, an appropriate amount of the fresh gas has to be continuously supplied in this latter operation mode to compensate for the gas reactant exhaustion due to reaction.

In principle, two different types of Venturi-tube gas distributors can be distinguished, shown schematically in Fig. 1. Apparently, these two Venturi tube types differ by the presence or absence of the cylindrical mixing (momentum exchange) tube located between the suction chamber and diffuser, the diameter of this tube being equal to the diffuser inlet diameter. Type VT-1 with the mixing tube (Fig. 1a) has been commonly described in literature[1–3,7] while the performance of the simplified version VT-2 shown in Fig. 1b (diffuser directly attached to the suction chamber) was thoroughly investigated in our previous studies[6,8]. In both cases, diffuser can be alternatively conical or slot-shaped[4]. Considering the decisive effect of energy dissipation rate on the intensity of interfacial contact in bubble beds

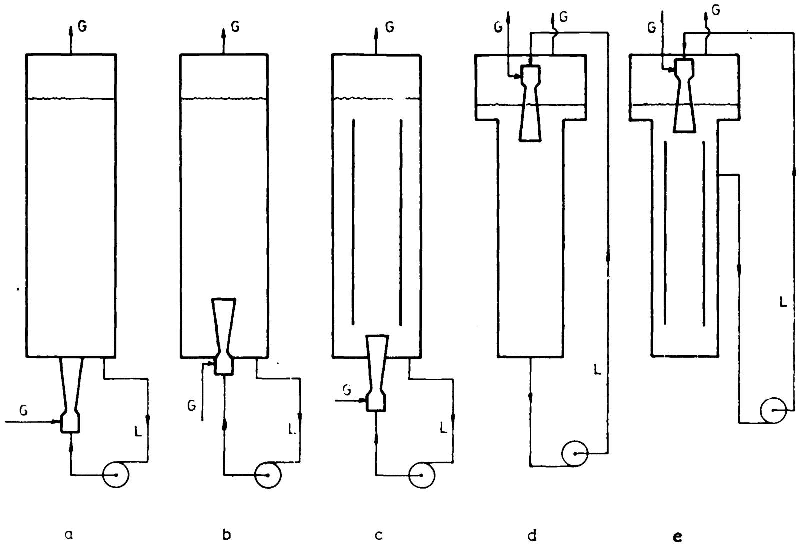


FIG. 2

Construction modifications of tower reactors with Venturi-tube gas distributors. a Upflow arrangement — Venturi tube located at the reactor bottom, b upflow arrangement — Venturi tube inserted into the reactor vessel, c upflow arrangement — reactor with a central draught tube, d downflow arrangement, e downflow arrangement — reactor with a central draught tube

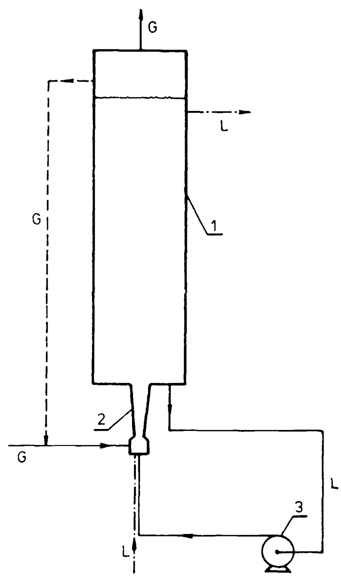
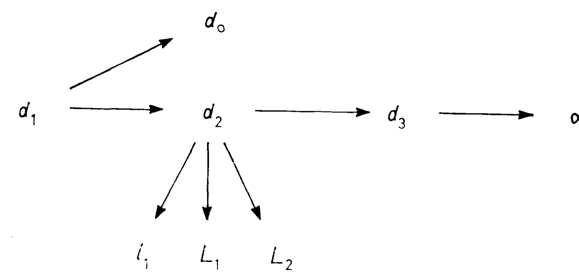


FIG. 3

Alternative flow schemes of ejector-distributor reactors. 1 Reactor vessel, 2 ejector distributor, 3 liquid circulation pump; flow streams: full lines — liquid circulation loop, fresh gas supply; dashed line — gas circulation loop; dot and dash line — external liquid feed generated by ejector distributors[12], nozzle diameter can be regarded as the basic ejector parameter determining for given liquid flow rate value of the ejector pressure drop and thus the rate of energy dissipation in the ejector, $E_d = \Delta P_e Q_L$. For given demands on the ejector performance (required mass transfer rate and corresponding value of superficial gas velocity in the reactor), value of nozzle diameter d_1, has to be determined with respect to the capacity of liquid pump available and to its pressure drop limitations. Selection of further ejector parameters then evolves from the d_1 value according to the Scheme 1, where meaning of individual symbols is apparent



SCHEME 1

from Fig. 1. For practical design purposes, geometrical parameters of Venturi-tube distributors with the momentum exchange tube (VT-1) should be kept within limits given by hereinafter recommended values of their respective ratios, proved[3,7,8] to ensure efficient performance of such gas distributing devices:

$$d_o/d_1 = 1{\cdot}2 - 3{\cdot}5 \qquad l_1/d_2 = 0{\cdot}9 - 2{\cdot}2$$
$$d_2/d_1 = 1{\cdot}5 - 4{\cdot}5 \qquad L_1/d_2 = 5 - 8$$
$$L_2/d_2 = 8 - 12\,.$$

Liquid velocity related to the nozzle-throat cross-section should range between 20 and 50 m/s, the angle of diffuser walls inclination, α, between 2 and 5°. In common applications, the ejector pressure drop should not exceed 1 MPa and as a rule, it should vary between 0·5 and 0·8 MPa. Specific features of Venturi-tube distributors without the momentum exchange tube (VT-2) were investigated in our previous studies[6,8,12] devoted namely to the effect of nozzle and diffuser geometry on the quality and efficiency of gas dispersion. It has been proved that the absence of the mixing tube has no negative influence on the distribution performance of Venturi tubes and that the energetic efficiency of gas–liquid dispersion formation varies for both distributor types (VT-1 and VT-2) with the diffuser length only. Also, no effect of the diffuser shape on the energy effectiveness of distributor performance was observed and almost identical data were obtained for both conical and slot-shaped diffusers at constant mixing tube and diffuser lengths[7,8]. These results are comprehensively illustrated by Fig. 4, in which gas holdup data are plotted against the specific rate of energy dissipation in the Venturi tube related to a unit of liquid mass in the bed, $e_d = \Delta P_e Q_L / V_L \varrho_L$. As can be seen from the graph, no significant difference was observed between ε_G vs e_d data obtained with the Venturi tube VT-1 for conical and slot-shaped diffuser at constant diffuser length ($L_2 = 0{\cdot}18$ m) and inlet-to outlet cross-section ratio. Also, it is apparent from the figure that the data for the Venturi-tube distributor with the mixing tube (VT-1) agreed well with those obtained for the VT-2 type at corresponding diffuser length ($L_2 = 0{\cdot}12$ m, $L_d = 0{\cdot}10$ m or $L_2 = 0{\cdot}18$ m, $L_d = 0{\cdot}20$ m) despite the fact that the total length $L_1 + L_2$ was in the former case considerably larger (for $L_1 = 0{\cdot}11$ m) than the diffuser length only in the case of the distributor modification VT-2. In other words, the equivalence of ε_G vs e_d data was observed at constant diffuser length and not as could be expected at $L_d = L_1 + L_2$. In experiments with Venturi-tube distributors with the mixing tube no further increase of gas holdup was observed with increasing L_2 values from 0·20 to 0·50 m (this apparently being in agreement with Henzler's conclusion concerning

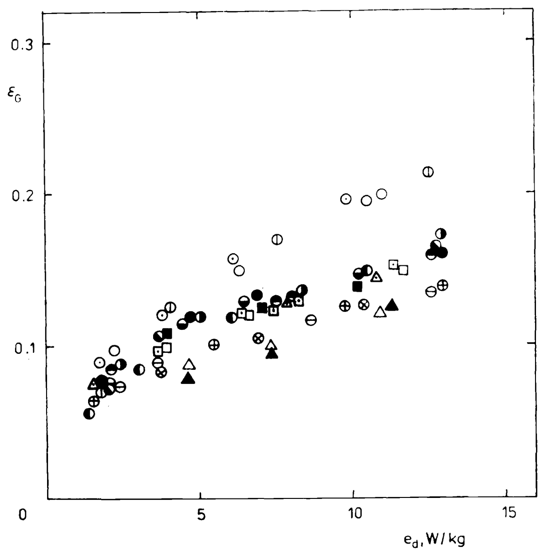


FIG. 4

Gas holdup as a function of the energy dissipation rate. Distributor VT-1: $L_1 = 0{\cdot}11$ m, $d_0 = 0{\cdot}015$ m, $d_1 = 0{\cdot}005$ m, nozzle *a*: △ $L_2 = 0{\cdot}12$ m, $\alpha = 5°$; ▲ $L_2 = 0{\cdot}12$ m, $\alpha = 2{\cdot}5°$; ◭ $L_2 = 0{\cdot}18$ m, $\alpha = 5°$; ◮ $L_2 = 0{\cdot}18$ m, $\alpha = 2{\cdot}5°$; ◬ $L_2 = 0{\cdot}50$ m, $\alpha = 2{\cdot}5°$; ◬ $L_2 = 0{\cdot}18$ m, slot-shaped diffuser; distributor VT-2: $d_0 = 0{\cdot}016$ m, $d_1 = 0{\cdot}008$ m; $L_d = 0{\cdot}4$ m, $\alpha = 1{\cdot}7°$: ○ *a*, ⊙ *b*, ⦶ *c*; $L_d = 0{\cdot}2$ m, $\alpha = 3{\cdot}4°$: ● *a*, ◐ *b*, ◑ *c*; $L_d = 0{\cdot}2$ m, $\alpha = 1{\cdot}7°$: ◒ *a*, ◓ *b*, ◔ *c*; $L_d = 0{\cdot}1$ m, $\alpha = 6{\cdot}8°$: ⊕ *a* ⊗ *b* ⊖ *c*; *a*, *b*, *c* nozzle types[6,8] the optimum length of this type of Venturi tubes[3]) and consequently, E_G data for the latter diffuser length were at comparable e_d values significantly lower than those obtained with the distributor type VT-2 for $L_d = 0{\cdot}4$ m. Apparently this result suggests that the distributing performance of this latter type of Venturi tubes (and consequently the quality of gas dispersion) can be influenced by the diffuser geometry in wider range than in the case of Venturi tubes VT-1 by variations of both mixing tube and diffuser geometry. Correspondingly, higher efficiency of dissipated energy utilization and thus even higher energy effectiveness of gas distribution can be ensured at comparable flow conditions by the application of Venturi-tube distributors without the mixing tube which seem to be advantageous even due to their simpler construction. As to appropriate design recommendations, optimum values of ratios d_2/d_1, d_3/d_2, l_1/d_1, and d_o/d_1 given above for Venturi tubes VT-1 can be employed even for design of Venturi tubes VT-2, as for the diffuser length, our experimental results proved that L_d/d_2 ratio could be raised up to 25–30 with a positive effect on the quality of gas dispersion. Correspondingly, values of the angle of diffuser walls inclination can be even lower than 2° (values 1·7° were used in both laboratory and large-scale distributing devices reported further in this paper). Regarding the direct proportionality between gas holdup (and k_La) and gas flow rate in buble column reactors with ejector gas distributors (e.g. for air–water system respective dependences were well described[12] by linear relations $\varepsilon_G = 3{\cdot}5u_{oG}$ and $k_La = 2{\cdot}0u_{oG}$), the gas suction rate Q_G has to be viewed as the decisive operating parameter of selfsucking ejectors. It has been proved in our former study[8] that for given rate of energy dissipation in the ejector gas suction rate increases with the diffuser length (see Fig. 5). Our recent experimental study[15] confirmed that it was really the diffuser length rather than the L_d/d_2 ratio which decisively influenced gas suction efficiency. As can be seen from Fig. 5, almost identical dependence Q_G vs E_d was obtained with Venturi tubes A1 and **B** (see Table I) differing in all characteristic dimensions except diffuser length. While the complete presentation of this study results lies beyond the scope of this paper, it can be concluded here that the ejector distributor scale-up can be safely based upon the identity of the specific energy dissipation rate, e_d, while the efficiency of dissipated energy utilization depends primarily on the diffuser length. Indeed, the ratios of characteristic ejector dimensions and flow rates in decisive cross-sections (nozzle outlet, diffuser inlet) should be always kept within the limits recommended above.

Industrial Ejector Design

Design of the ejector (Venturi-tube type) gas distributor for a specific duty has to be evolved from the estimate of gas suction rate corresponding to the demands on the rate of mass transfer (size of the interfacial area) and to the reactor dimensions. Following design steps then include determination of the energy dissipation rate ensuring required rate of gas suction and thus appropriate value of superficial gas velocity and specification of limits imposed on the ejector pressure drop (ΔP_e) and/or liquid circulation rate (Q_L) by process conditions or by functional parameters of accessible pumps. The characteristic ejector dimensions can be then calculated for selected ΔP_e and Q_L values with due regard for recommended optimal ranges of geometrical parameters ratios and flow rates in characteristic ejector cross-sec-

TABLE I

Characteristic dimensions of Venturi tubes compared in Fig. 5

Parameter	A1	A2	A3	B
L_d, m	0·4	0·2	0·1	0·4
d_1, m	0·008	0·008	0·008	0·020
d_2, m	0·016	0·016	0·016	0·032
d_3, m	0·040	0·040	0·040	0·056

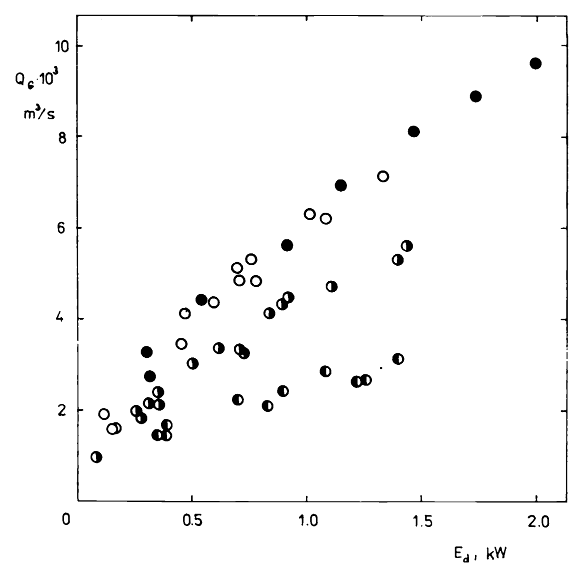
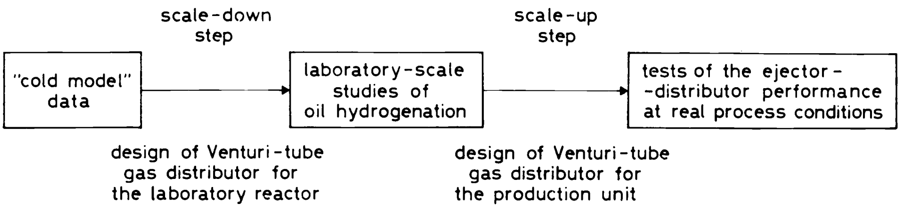


FIG. 5

The effect of diffuser length on gas suction rate. Dimensions of ejector distributors A1, A2, A3, and B are given in Table I ○ A1, ◑ A2, ◐ A3, ● B tions. The „case history" presented further as an illustrative example gives the account of a Venturi-tube gas distributor design for the industrial reactor ($V_r = 8\ m^3$, $D_r = 1{\cdot}6$ m) for rape-seed oil hydrogenation catalysed by Ni on kieselguhr. The design procedure including scale-down and scale-up steps is schematically represented by Scheme 2. The "cold model" data were obtained[6,8] in a tower reactor ($D_r =$



SCHEME 2

$= 0{\cdot}3$ m) with the VT-2 type gas distributor. The experiments were carried out in the semi-batch flow arrangement with the standard air–water system at atmospheric pressure and room temperature. The two studies[6,8] yielded quantitative information about the effect of individual Venturi-tube design parameters on the ejector pressure drop and gas suction rate as well as on the decisive hydrodynamic and mass transfer characteristics of gas–liquid beds generated in the reactor (see e.g. Figs 4 and 5) The „cold model" experiments thus yielded a suitable basis for the optimization of Venturi tubes geometry with respect to the quality of gas dispersion and to the energy effectiveness of phases contacting in units with Venturi-tube gas distributors. In the following *scale-down step* the appropriate optimization principles were employed in the design of the Venturi-tube gas distributor for a bench-scale reactor built for laboratory studies of the rape-seed oil hydrogenation. Experimental data from this reactor proved[16,17] the positive effect of mass transfer intensification on the effective reaction rate (see below) and yielded the optimum value of superficial gas velocity in the reactor ($u_{oG} = 0{\cdot}01$ m/s) ensuring sufficient rate of the process and suitable product composition (i.e. hydrogenation selectivity) at reasonably low rate of energy dissipation in the ejector (i.e. at correspondingly low demands on the energy supply). Gas suction rate for the large-scale unit ($D_r = 1{\cdot}6$ m, $A_r = 2{\cdot}0\ m^2$) was therefore estimated to ensure such superficial gas velocity and the respective value $Q_G = 0{\cdot}02\ m^3/s$ (i.e. $72\ m^3/h$) was then taken as the starting point of the *scale-up step* aimed at determination of decisive design parameters of the Venturi-tube gas distributor for the industrial hydrogenation reactor. The value of total energy dissipation rate pertaining to the required rate of gas suction was subsequently determined from our Q_G vs E_d data[8,15] (shown partially in Fig. 5). Extrapolation of respective graphical dependences yielded value $E_d = 9{\cdot}6$ kW (wherein correction coefficient

1·1 was employed allowing for possible extrapolation error) and for the postulated maximum ejector pressure drop, $\Delta P_e = 0{\cdot}8$ MPa, this E_d value then according to definition yielded the appropriate value of the volumetric flow rate of liquid through the ejector (liquid circulation rate), $Q_L = 12{\cdot}0 \,.\, 10^{-3}$ m^3/s. Determination of Venturi-tube design parameters for such defined working conditions $((\Delta P_e)_{max}, Q_L)$ was then based on the "cold model" data. Comparison of pressure drop data (plotted in Fig. 6 against the correlation term $(u_{L1}^2 \varrho_L/2)$) for different nozzle types proved superior operation flexibility of conical nozzles shown schematically in Fig. 7

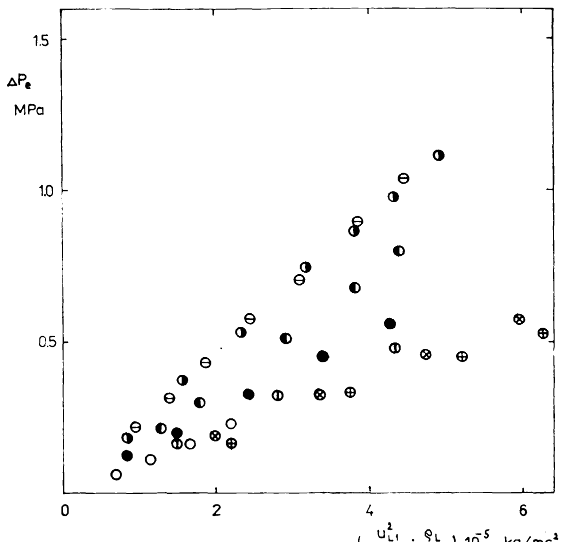


FIG. 6

Ejector pressure drop as a function of the correlation term $(u_{L1}^2 \varrho_L/2)$. Nozzle type *a*: ○ $d_1 = 0{\cdot}011$ m, ● $d_1 = 0{\cdot}009$ m, ◐ $d_1 = 0{\cdot}008$ m, ◑ $d_1 = 0{\cdot}006$ m; nozzle *b*: ⊕ $d_1 = 0{\cdot}010$ m, ⊗ $d_1 = 0{\cdot}008$ m, ⦶ $d_1 = 0{\cdot}006$ m; nozzle *c*: ⊖ $d_1 = 0{\cdot}008$ m

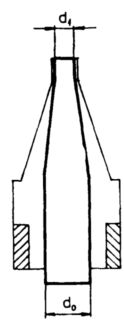


FIG. 7

Schematic chart of the conical nozzle (*b*-type[6,8]) (and denoted as *b*-type nozzles in our previous studies[6,8]). Nozzle of this type was therefore selected for the industrial-scale ejector distributor and the appropriate value of nozzle diameter, $d_1 = 0{\cdot}02$ m, was then determined from the pressure drop correlation,

$$\Delta P_e = \xi u_{L1}^2 \varrho_L / 2 \tag{1}$$

for the average value of pressure drop coefficient, $\xi_{av} = 1$, representing for nozzles of this type the whole experimental range of nozzle outlet-to-inlet cross-sectional areas ratio, $A_1/A_0 = 0{\cdot}18 - 0{\cdot}51$ (as can be seen in Fig. 6, functional dependence $\xi = \xi(A_1/A_0)$ is relatively week for this type of nozzles). For this d_1 value, liquid velocity in the nozzle throat equaled 38·2 m/s and lied thus within the above recommended region, $u_{L1} = 20 - 50$ m/s. Characteristic diffuser dimensions were then derived from those of the "cold model" distributore mploying the geometrical similarity concept i.e. keeping constant values of respective dimensionless ratios, $d_2/d_1 = 1{\cdot}6$; $d_3/d_2 = 2{\cdot}5$; $L_d/d_2 = 25$.

Values of decisive geometrical parameters of Venturi-tube distributors designed for the laboratory- and large-scale hydrogenation reactors are summarized in Table II together with respective dimensions of the referential "cold model" distributor. Apparently, values of all respective ejector parameter ratios fall within the ranges recommended above for the efficient ejector–distributors performance. Certain concessions had to be made in the scale-down step to demands on large variability of operation conditions in the laboratory hydrogenation unit and partially even to the technical problems of such small ejector manufacturing. As a result, rules of ejector scaling were not strictly respected in the laboratory ejector design and certain deviations of its dimensions from the geometrical similarity with the other two Venturi-tube distributors are therefore apparent in Table II.

The configuration of all three reactors considered herein corresponded to the basic scheme a shown in Fig. 2 (reactor without central draught tube with the external ejector–distributor located in the upward position at the reactor bottom). The hydrogenation reactors were operated semi-batchwise in the gas circulation mode where the rate of fresh hydrogen supply was at all working conditions negligible against the rate of hydrogen circulation induced by the ejector performance. The superficial velocity of gas flow in these reactors could be thus always directly related to the gas suction rate of respective Venturi-tube gas distributors. Preliminary tests proved negligible settling rate of catalyst particles at common operation conditions. No provisions were therefore made regarding the presence of the solid phase and the suspension was treated as a pseudo-continuous phase. Kinetic data from the laboratory hydrogenation reactor proved[16,17] that in comparison with other modes of gas distribution tested in this unit (aeration from an open pipe or from a perforated spider-type distributor), application of the Venturi-tube distributor reduced the hydrogenation time per batch approximately to a half within the whole experimental range of reaction temperatures and pressures ($T = 150-190°C$, $P = 0{\cdot}16-1{\cdot}0$ MPa). Comparison of results from individual industrial-unit runs (and thus even the exact evaluation of the positive contribution of the Venturi-tube distributor) was hampered by the fact that constant catalyst concentration and activity could not be guaranteed in these runs due to the technological process limitations[18]. While the profound discussion of related problems and a complete account of experimental results from the full-size production reactor can be found elsewhere[18], it can be concluded here that application of the Venturi-tube distributor increased substantially the overall rate of the process (i.e. reduced reaction time corresponding to a required hydrogenation degree) or alternatively, the catalyst load corresponding to particular process requirements (reaction time and hydrogenation degree) could be appreciably reduced in comparison with the original reactor configuration – mechanically stirred vessel with propeller agitator ($D_m = 0{\cdot}55$ m, $n = 3\,s^{-1}$), stoichiometric amount of hydrogen supplied below the agitator by an open-end tube. Typical hydrogenation data from both laboratory- and industrial-scale ejector–distributor reactors are presented in Figs 8 and 9 in the form of time dependences of the iodine value I.V. commonly employed as the characteristics of oils and fatty acids hydrogenation degree*. Comparison with respective data obtained in the same reactor vessels for

TABLE II

Dimensions of Venturi-tube distributors designed for the laboratory- and industrial-scale hydrogenation reactors and of the referential "cold model" distributor. Reactor types: A glass-wall reactor for hydrodynamic measurements, B laboratory-scale hydrogenation reactor, C industrial reactor for rape-seed oil hydrogenation

Parameter	Reactor type		
	A	B	C
D_r, m	0·3	0·094	1·6
V_L, m^3	0·120	0·003	5
d_0, m	0·014	0·005	0·035
d_1, m	0·010	0·002	0·020
d_2, m	0·016	0·008	0·032
d_3, m	0·040	0·020	0·080
l_1, m	0·032	0·016	0·067
L_d, m	0·4	0·1	0·8
α	1°38′	3°26′	1°38′

* According to its definition[19], the iodine value of an oil or fatty acid can be regarded as a measure of the double bonds content in the respective compound and its time dependence thus appropriately represents the course of hydrogenation.

alternative modes of gas distribution proved that the positive effect of interfacial contact intensification due to the Venturi tube performance was even more pronounced in the large-scale reactor. Apparently, the extremely low hydrogenation rate observed in the industrial stirred-tank reactor at common working conditions can be ascribed to the poor interfacial contact generated in this reactor at conditions of only "stoichiometric" hydrogen feed rate.

Concluding Remarks

While the effect of bubble coalescence degree and liquid viscosity on the ejector distributors performance was briefly treated by Zlokarnik[4] and Henzler[3] respectively, further attention should be paid to determination of the effect of gas–liquid system properties on the quality of gas dispersion in ejector–distributor reactors and on the energy effectiveness of ejector distributors performance. Considering the extremely fine primary bubble dispersion generated by ejectors as compared with perforated or even sintered plates, it can be assumed that the favourable effect of ejector distributors application on the intensity of mass transfer in bubble beds and on the efficiency of gas phase utilization is more significant in coalescence sup-

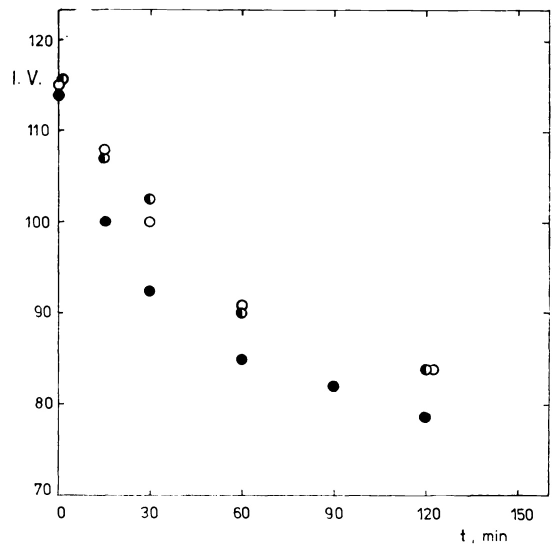


FIG. 8

Iodine value as a function of time — laboratory-scale experiments; $V_L = 0{\cdot}003\ m^3$, $T = 175\ °C$, $P = 0{\cdot}16$ MPa, $c_{cat} = 0{\cdot}07$ wt. %. ○ Gas supply by an open pipe; ◐ perforated-pipe spider-type gas distributor; ● Venturi-tube gas distributor pressing systems. In such cases, data determined with the standard air–water system can be thus apparently used only as the first approximation for the ejector design purposes and laboratory experiments with real systems should be preferred whenever possible.

Considering typical ratios of diffuser outlet to reactor cross-section area, ejectors (Venturi tubes) can be regarded as typical "point" (space concentrated) distributors and their distributing efficiency is thus strongly dependent on the reactor height to diameter ratio, H/D_r (at constant reactor volume). Obviously, higher H/D_r values should be always preferred at given reactor volume to ensure more uniform gas distribution over the whole reactor cross-section. For $H/D_r < 2$, multi-jet arrangement[20,21] can be generally recommended for full-size reactors and similarly such a reactor configuration should be always preferred in large diameter units ($D_r > 1$ or 3 m for upward or downward oriented ejectors respectively). Alternatively, the configurations with a central draught tube (Figs 2c, e) can be also recommended for large-diameter units. From this point of view, the multi-jet or draught-tube configurations would have been apparently appropriate even in the case of the industrial hydrogenation reactor discussed above. Application of the simple single-ejector variant was however in this case necessitated by the strict requirement of the smallest pos-

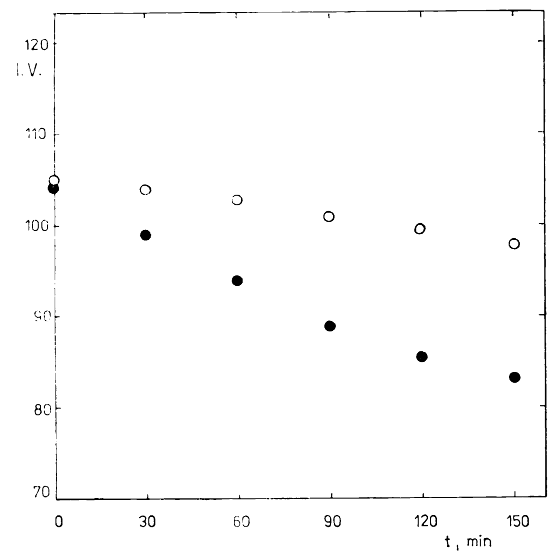


FIG. 9

Iodine value as a function of time for different operation modes of the industrial hydrogenation reactor; $V_L = 5\ m^3$, $P = 0{\cdot}1$ MPa, $c_{cat} = 0{\cdot}1$ wt. %. ○ Stirred vessel ($D_m = 0{\cdot}55$ m, $n = 2\ s^{-1}$), gas supplied by an open pipe below the agitator; ● reactor with Venturi-tube gas distributor sible modifications of the existing production unit with only one inlet throat available at the vessel bottom.

SYMBOLS

A_o	cross-sectional area of nozzle inlet
A_1	nozzle throat area
A_r	cross-sectional area of a reactor
c_{cat}	catalyst concentration
D_m	impeller diameter
D_r	reactor diameter
d_o	diameter of nozzle inlet
d_1	nozzle throat diameter
d_2	diffuser inlet diameter
d_3	diffuser outlet diameter
E_d	rate of energy dissipation in the ejector
e_d	energy dissipation rate related to a unit of bubble bed mass
H	reactor height
k_La	volumetric liquid-side mass transfer coefficient
L_1	mixing tube length
L_2, L_d	diffuser length (VT-1 or VT-2, respectively)
l_1	distance between the nozzle tip and mixing tube inlet
n	impeller rotation speed
P	pressure
ΔP_e	ejector pressure drop
Q_G	gas suction rate
Q_L	liquid throughput (rate of liquid circulation through an ejector)
T	temperature
t	time
u_{oG}	superficial gas velocity
u_{L1}	liquid velocity in a nozzle throat
V_L	volume of liquid in bubble bed
V_r	reactor volume
α	angle of diffuser walls inclination
ε_G	gas holdup
ξ	pressure drop coefficient
ϱ_L	liquid density

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