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A STUDY OF THE PERFORMANCE OF A DOWNCOMER

M. Anwar Ulhaq

Fertilizer Research and Development Institute, Faisalabad

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Performance of a downcomer, using O₂-air-aqueous glycerine system 50% by wt. has been investigated. Mass transfer efficiencies and mean residence time, have been measured as a function of liquid and air flow rates. It has been demonstrated that downcomer does play a significant role in the mass transfer, in a unit, therefore a serious thought should be given while designing a downcomer.

Key words: Downcomers, Sieve trays, Mass transfer plates.

INTRODUCTION

Although, the importance of the downcomer, as a contributor to the mass transfer in a distillation unit was pointed out as early as [1], but very limited data is reported in the literature and that too is very much contradictory. It has been stated by Thomas and Campbell [2] that the behaviour of the tray plus downcomer, as a unit may be very important in certain circumstances. Due to paucity of data the design of downcomer is still mainly based on empirical equations.

The present study was carried out to establish the overall performance of a downcomer.

Downcomer theory. The mechanism of bubble and froth formation in a downcomer is unexplored. Small bubbles of almost uniform sizes are collected at the base of the downcomer. As we pass upwards these bubbles grow and a gradual transition to froth occurs. A large deal depend upon the mode of entry of the liquid from the tray above.

The earlier workers [1, 3, 4] were of the opinion that column would flood, when the aerated liquid reaches the top of the exit weir. Thomas and Shah [5] has shown that the most important design factor is clear liquid height, the froth height contributes little to the tendency of flooding. The clear liquid height in the downcomer can be calculated from the following equation.

$$Z_d = P + L_b + (h_f/\rho_L) - h \dots \dots \dots (1)$$

The liquid throw over the weir is no longer considered to be a limiting factor in tray designing [2].

Relative foam density values in the downcomer are not available in the literature. Since specific foam density varies approximately from ρ_L at the bottom of the downcomer to ρ_f at the foam vapor interface. Therefore, a conservative

average value of the relative of the density $\phi = 0.5$ is widely used in the design of the downcomers [3].

A certain residence time of liquid in the downcomer is necessary in order to allow collapse of foam. It is common practice to base this residence time on the total downcomer volume. The minimum allowable residence time should be based on the foamability of the system.

A true residence time of the aerated mass in the downcomer is given as [3].

$$t = 0.083 \cdot \frac{(A \cdot Z_t)}{q/\phi} = 0.083 \frac{A \cdot Z_d}{q} \dots \dots \dots (2)$$

Where $Z_t = \phi = Z_d$ and $A \cdot Z_d$ is equivalent to the clear liquid volume. q is the liquid flow rate. Then t as given in the above equation 2 represent to "Plug flow".

A value of 5 seconds for plug flow in the downcomer is suggested by Davis [1]. Observations made by Thomas *et al* [2, 5] certainly leads one to expect anything but a plug flow in the downcomer.

EXPERIMENTAL

Since the circular downcomer provides very low downflow area and poor vapor disengaging space and usually constitute the first bottle neck to column capacity, therefore a segmental downcomer was selected for the present studies. The downcomer was 12.5 cm deep, 30.5 cm wide and 60 cm long with the provision of measuring points.

The pilot plant [2] was operated, as under the normal conditions, using O₂ desorption from aqueous glycerine (50% by wt.) solution. Froth height and clear liquid heights were measured visually, as the downcomer was constructed from transparent material. For mass transfer efficiency studies, liquid samples were withdrawn at inlet, and outlet

and were analysed for dissolved O_2 concentration in the liquid phase continuously with the help of O_2 detection cell, (manufactured by Cambridge Instrument Ltd. of England).

For mean residence time studies, a dye injection (Nigrocine) technique was selected. The dye was injected in inlet-weir and its concentration in solution against time was measured at downcomer outlet, with the help of a photoelectric cell and was recorded by an ultra violet recorder continuously [6].

RESULTS AND DISCUSSION

In Fig. 1 is shown the effect of liquid flow rates on the froth height and static liquid head in the downcomer at two different air flow rates. The froth height, on top of clear liquid increases with liquid and air flow rates which gradually tails off at higher liquid rates. This is expected, since with increasing liquid and air flow rates more froth will be generated on the tray above, which will pass over the exit weir into the downcomer. At higher liquid flow the cascading liquid from the exit weir disrupt and mechanically breaks the froth, which results in lowering its height in the downcomer.

The clear liquid height in downcomer increases with increasing liquid and air flow rates Fig. 1(b). This build up of liquid in the downcomer is mainly due to the increase in liquid flow over the exit weir into the downcomer. Secondly, the increasing flow rates, increases the pressure drop across the trays, which results in additional hold up of the liquid in the downcomer. The liquid hold up in the downcomer can be calculated very accurately by equation 1. It was observed that there is no possibility of closing of the mouth of downcomer by cascading liquid flowing over the exit weir into the downcomer. The clear liquid height should be considered as the main design parameter, as the light froth contributes very little to the total head, hence to the flooding of the downcomer [5, 7].

The pressure drop in the downcomer is very small and is almost independent of the air flow rates. This is expected since very narrow flow rates were investigated. The pressure drop is slightly dependent on the liquid flow rates i.e. increases with increasing flow rates, but is not significant as shown in Fig. 2.

Mean residence time (MRT) is highly dependent on the liquid flow rates. It decreases with increasing liquid flow rates Fig. 3. For high flow rates (in case of industrial operation) MRT in downcomer of 2s seems to be more realistic figure rather than 5s as suggested by Davis [1].

The mass transfer efficiency of the downcomer increases with liquid flow rates Fig. 4. At 511 LPM/m weir liquid flow rates, mass transfer efficiency as high as 12% is

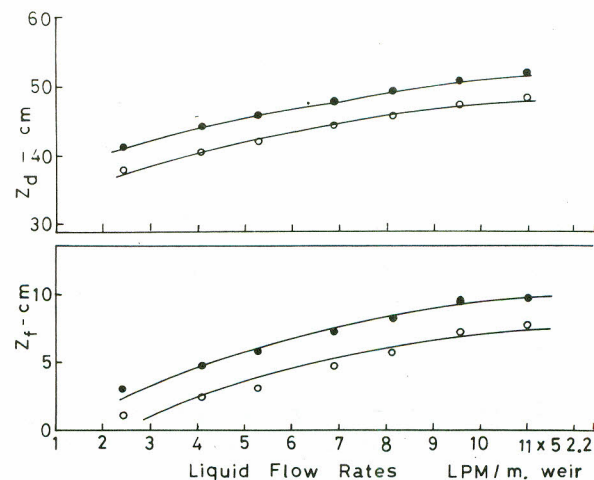


Fig. 1. A plot of (a) froth height and (b) clear liquid height verses liquid flow rates at different air flow rates.

(a) Froth height (Z_f) (b) Static liquid height (Z_s)
 O- $F_A = 2.1$ O- $F_A = 2.1$
 ●- $F_A = 2.43$ ●- $F_A = 2.43$

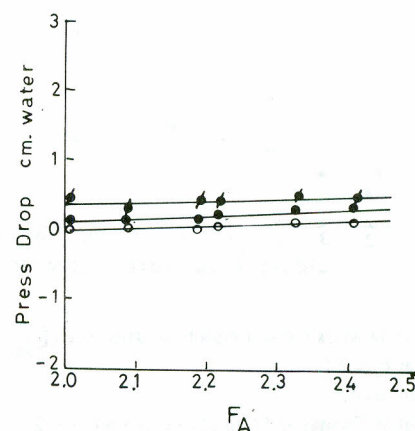


Fig. 2. Plot of pressure drop in the downcomer verses air flow rates at different liquid flow rates.

O - 269.1 LPM/m of weir; ● - 401.2 LPM/m of weir; ◆ - 536.7 LPM/m of weir.

obtained, which is a high contribution to mass transfer, when the combined unit of tray plus downcomer is considered. This is supported by the data of earlier workers [2] which is reported alongwith the present data in Fig. 4. This suggests that role of the downcomer should be considered carefully while designing a distillation column.

Number of transfer units, N_L , are shown as a function of air flow rate and MRT in Fig. 5 and 6 respectively. The positive role of the downcomer towards mass transfer is demonstrated as shown in Fig. 5, N_L is independent of air flow rates over the narrow range investigated. Number of transfer units, increase with increase in MRT of the liquid Fig. 6. The scatter in data at the lower values of MRT (or

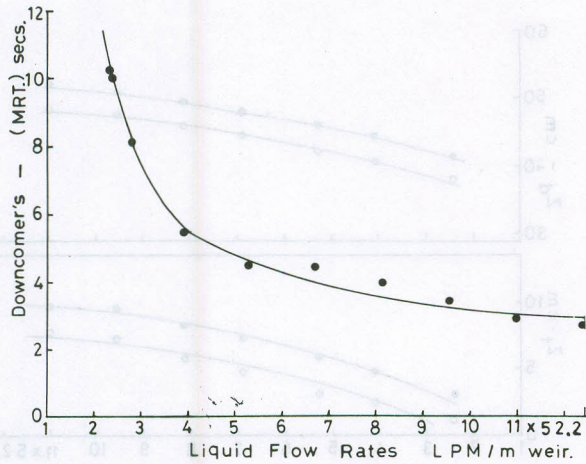


Fig. 3. Plot of MRT against liquid flow rate at $F_A = 2.1$ air flow rate.

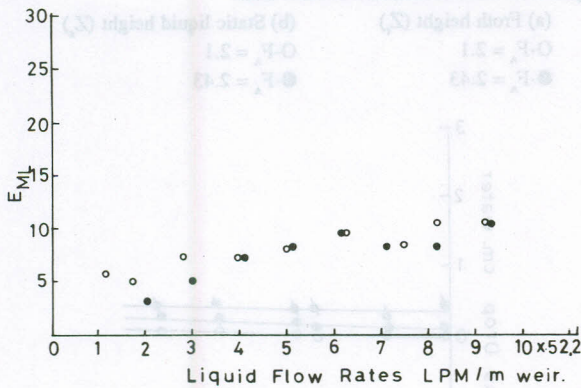


Fig. 4. Plot of Mass transfer murphree efficiency (E_{ML}) against liquid flow rates at $F_A = 2.1$
 ● - Present study;
 O - Report at Thomas and Campbell's data ref. No. 2.

high liquid flow rates) is less when compared with the higher values of MRT (or low liquid flow rates). A comparison of number of transfer units, N_L for tray alone and tray plus downcomer shows that they are of the same order. This phenomena can not be explained in term of N_L alone since dimensionless and MRT have also to be taken into consideration as they effect the liquid phase number of transfer units, N_L as well. It is also likely, that this is predominantly due to a fall in interfacial area in the downcomer itself [8].

CONCLUSIONS

It is concluded from the present study that the downcomer plays a significant role in the mass transfer process, the relative importance of which depends upon the flow rates and the nature of the system. Secondly, the design of a downcomer should be based on the clear liquid

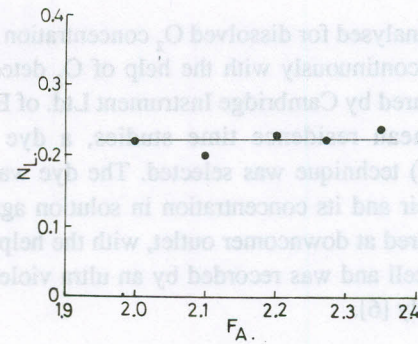


Fig. 5. Plot of number of transfer unit (N_L) versus air flow rate at ● - 269.1 LPM/m weir liquid flow rates.

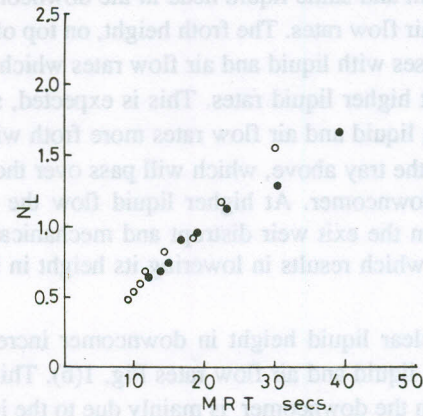


Fig. 6. Plot of number of transfer unit (N_L) against liquid mean residence time (MRT) at air flow rate $F_A = 2.1$; O - Tray alone; ● - Tray plus downcomer.

head rather than on the froth height as previously reported in the literature.

NOMENCLATURE

- A = Downcomer area cm^2
- E_{ML} = Mass transfer murphree efficiency
- F_A = F-factor based on the perforated area of the tray air flow rate.
- h = Pressure build up in downcomer cm of water.
- h_t = Total pressure drop across wet try in cm of water.
- L_b = Static liquid seal on the lower tray in cm.
- MRT = Mean residence time of liquid. s.
- N_L = Number of mass transfer units, liquid based.
- P = Pressure drop through the clearance between downcomer and lower tray in cm.
- q = Liquid flow rates in cm^3/sec .
- t = Time. s.
- Z_d = Height of clear liquid in downcomer - cm of liquid.
- Z_f = Froth height in downcomer - cm.
- ρ_f = Average foam density gm/cm^3 .

- ρ_L = Density of clear liquid gm/cm³
 ϕ = Froth density factor on the tray.

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Results are reported on recovery of MgSO₄ by chilling of sea water bittern using the design of the chiller as well as cooling temperatures as the two important variables. Data are presented for chillers of four different capacities from which it appears that MgSO₄ can be effectively recovered by cooling bittern to -2°, the temperature at which the concentration of this salt reduces to zero. These studies also indicate that recoveries can be enhanced if [1] arrangements are made for speedy removal of the crystals from the bulk phase of the liquid and [2] jackets are provided instead of coils in the chilling units.

Statistical analysis of the data reveals that during cooling of MgSO₄, there exist linear relationships between [1] concentration and temperature and [2] recovery and temperature which holds good to 95% confidence level. In the former case, the relationship is independent of initial feed concentration and the equipment design whereas in the latter the concentration range is a parameter. Nevertheless, heating of

Key words: Chilling, Sea bittern, Reaction design.

INTRODUCTION

Sea water bittern, the residual concentrated liquor after the recovery of solar salt is allowed to be discharged back into the sea in Pakistan, although it is known to contain quite valuable chemicals. Earlier publications [1, 2] dealt with the status of sea salt production and reasons for deterioration of its quality. It was pointed out that admixtures due to formation of metastable phases are invariably noted if solar evaporation is continued beyond 29-29.5°C. Be. They consequently impart impurities to the sea salt. Since the chemicals which form the impurities are recoverable from bittern, an integrated process was reported [2] for obtaining them on pilot plant scale.

Sea-water bittern is a very sensitive and complex ionic mixture of typical inorganic compounds. The crystallization behaviour and subsequent composition of various salts and residues in the mother liquor at different stages of processing are controlled by the previous history, particularly the weather conditions. Changes in the composition raise a number of problems relating to equilibrium data on phase separation. This is perhaps the reason for adopting conventional methods such as the following for obtaining the various chemicals from this important but neglected resource: chilling [3-10], solvent extraction [11-13], passage through ion-exchange resins [14, 15], treatment with lime/caustic soda [16, 17], or an odd combination of these methods [18-

Chilling and solvent extraction from the basis of some of the major studies in this area. Experiments on low temperature separation of MgSO₄ have been reported for 12°, 10°, 8°, 10°, 0°, or -2°, [3, 5, 6]. These investigations are rather silent on optimizing the temperature and composition conditions which are pertinent to recovery of magnesium sulphate and of the chemicals extracted subsequently. The only study [7] specific for the production of MgSO₄ by chilling, though elaborate, lacks appropriate correlation with composition and is restricted to bench scale experiments. Additionally, studies by various authors [2, 21, 22, 23] on the recovery of MgSO₄ by chilling to different low temperatures were aimed at evolving an integrated process for the recovery of different chemicals from sea bittern. The present studies have been undertaken to establish low temperature as a driving force for the separation of MgSO₄ since this step is critical for the success of the subsequent operations. Accordingly experiments have been designed to establish recovery parameters from laboratory scale (1-6 litre batches) to different pilot plant scale operations viz. 25 to 400 litres, to strike correlation of temperature with the other parameters such as composition and recovery and also to note the effect, if any, of the design of the equipment thereon.

EXPERIMENTAL

Laboratory and pilot plant studies were carried out in