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MEASUREMENT OF SIEVE TRAY EFFICIENCY

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Murphree tray efficiency (E_{ML}) and mass transfer coefficient ($k_L a$) have been measured, using a 61 x 30.5 cm sieve tray with large perforation diameter i.e. 9.5 mm for an absorption system using [illegible]$_2$-air and water system. Dependence of E_{ML} and $k_L a$ on the flow rates have been demonstrated. The relationship between E_{ML}, N_{OL}, $k_L a$, $\bar{t}$ and δ^2 has also been examined.

Key words: Sieve tray efficiency.

INTRODUCTION

Extensive work has been done on the measurement of sieve tray efficiency in the past and is summarised by A.I.Ch.E. in the form of *Bubble Tray Design Manual* [1]. The basic work of Gerster *et al.,* [2, 3] considered inter-phase mass transfer and its relationship to efficiency.

The effect of liquid mixing on tray efficiency has been examined on the basis of "pool theory". This was conceived by Kirschbaum [4] and further developed by Gautreux and O'Connel [5]. An eddy diffusion mechanism has been discussed by other earlier workers [6, 7]. The experimental techniques used are of dye or tracer injection with calculation of values of eddy diffusion coefficient. The measurement of residence time distribution functions as described by Foss *et al.,* [8] used by Thomas *et al* [9, 10] has much to commend it. The continuous function appears to be more representative of the real situation than a stage-wise consideration such as exists in the pool theory.

It has been found that maloperation of small hole perforated trays could occur because of rust and sediment deposits. Where heavy tray fouling could occur or where sedimentation normally occurred in the process, it has been the practice to use buuble-cap trays. To avoid these problems larger hole perforated trays are recommended. Owing to the limited information available on the performance of these trays, the present study with larger perforation diameter i.e. 9.5 mm, was initiated in order to collect more mass transfer data so that design methods could be improved.

Theory

The Murphree overall efficiencies (E_{MV}, E_{ML}) and the corresponding point efficiencies (E_{OG}, E_{OL}) are well described in the literature [1, 2]. The transfer unit concept applied to a tray gives N_G, N_L, N_{OG} and N_{OL} for point and overall conditions. Using a mass transfer film model $K_L a$, $k_L a$, $K_G a$ and $k_G a$ are related by the additivity rule for resistance.

It can be shown that for a perfectly mixed liquid on a tray that

$$E_{ML} = N_{OL} / 1 + N_{OL} \tag{1}$$

When there is plug flow of liquid, then

$$E_{ML} = e^{N_{OL}} - 1/e^{N_{OL}} \tag{2}$$

When there is partial mixing of the liquid on a tray the situation is much more complex, the distribution of residence time having to be taken into account in the form of a distribution function $f(\theta)$.

Assuming that gas passes uniformly up through the liquid and that the operating and equilibrium conditions are linear, a relationship can be derived between Murphree overall and point efficiencies, λ and the distribution function expressed either in units of time (t) or dimensionless time (θ). This relationship was developed by Foss *et al.,* [8]. The equation is as follows;

$$E_{ML} = \left[\frac{1 - \int_0^\infty \exp(-\lambda E_{OG}\theta) \cdot f(\theta) d\theta}{1 - \lambda^{-1} \cdot (1 - \int_0^\infty \exp(-\lambda \cdot E_{OG}\theta) \cdot f(\theta) \cdot d\theta} \right] \tag{3}$$

for a liquid film controlled system where $\lambda \gg 1$, a simplified equation is therefore,

$$E_{ML} = 1-\int_0^{\infty} \exp\,(-\lambda E_{OG}\theta).\ f(\theta)\ d\theta \qquad (4)$$

It can be shown that,

$$N_L = \lambda E_{OG} \qquad (5)$$

It follows that,

$$E_{ML} = 1-\int_0^{\infty} \exp(-N_L\theta) \cdot f(\theta)\ d\theta \qquad (6)$$

Foss *et al.* [8] proposed a function of the form,

$$f(\theta) = \alpha.\theta.\beta.e^{\gamma\theta} \qquad (7)$$

and to yield the correct mean time and normalisation, the constant are such that

$$f(\theta)=[\theta^b \cdot e^{-\theta/\delta^2}] \,/\, [b\cdot\delta^{2(b+1)}] \qquad (8)$$

where $b = \delta^{-2} -1$.

Thomas and Campbell [9] has derived an equation for $f(\theta)$ using equation–8 as a suitable function for correlation. The derived equation is,

$$f(\theta) = \left[\frac{\theta^{(1-\delta^2)/\delta^2} \cdot e^{-\theta/\delta^2}}{\delta^{2/\delta^2} \cdot \Gamma(1+\delta^2)}\right] \qquad (9)$$

A continuous function is examined in both cases, but equation 9 is preferable, because in practice it is unlikely that (b) will be an integral and tables of gamma function are readily available. Substitution for the values of $f(\theta)$ from equation 9 into equation 3 yields the relationship between E_{ML} and N_L and when λ is much greater than one. This can be shown to be,

$$E_{ML} = 1-(1+N_L \cdot \delta^2)^{-1/\delta^2} \qquad (10)$$

By rearrangement of equation 10 we have,

$$N_L = \left[\frac{(1-E_{ML})^{-\delta^2}}{\delta^2} - 1\right] \qquad (11)$$

From this equation, N_L can be obtained from the experimentally measured values of the variance and efficiency under identical operating conditions in the same apparatus. The number of transfer units is also given by Foss *et al.*, [8].

$$N_L = k_L\, a\, Z_f/L = k_L\, a\, \bar{t} \qquad (12)$$

From the knowledge of the N_L and mean residence time $\bar{t}$ values $K_L a$, can be calculated for given operating conditions.

EXPERIMENTAL

The pilot plant consisted of three large stainless steel columns, each fitted with a large perspex window and an external downcomer. Two trays were used, the lower tray for the even distribution of gas phase, to the upper tray, which was used for experimental tests, once through liquid flow and recirculated gas flow with control of temperature. The important details of the tray and downcomer are given in (Table 1).

Table 1. Physical data for tray and downcomer.

Tray	
Length	61 cm
Width	30.5 "
Hole dia.	9.5 m.m
% Free area	10.8%
Spacing between trays	60 cm
Weir height	7.6 "
Weir length	30.5 "
Downcomer	
Height	60 cm
Width	28.57 "
Inside length	12.7 "

Great care was taken to ensure a constant flow of CO_2 gas into the air stream, so as to maintain a constant gas inlet concentration to the test tray. The gas was passed through a heater, stablizer tank, rotameter and a needle valve before injecting into the suction side of the blower. Similarly, sampling of gas for analysis was done with extreme care. The sampled gas was sucked in by a vacuum pump through a drop separator fitted with demister into another drop separator. Then the stream was cooled and the condensate collected. After this, part of the dry gas was passed through a U-tube containing the self indicating silica gel at controlled rate, to ensure that the moisture has been eliminated completely. The dried samples were fed into the infrared analyser. The gas samples were analysed for CO_2 concentration just before and after the test tray [10, 12]. Liquid samples were collected at tray inlet, tray outlet and at downcomer outlet and were analysed by automatic titrator. The sampling and analysis were continuous as was the recording [10]. The circulating gas and liquid was maintained at 25 ± 0.5°.

RESULT AND DISCUSSIONS

The mean residence times and variances were obtained at the same time as absorption rates and are reported elsewhere [10]. From the equations described, N_L was calculated from E_{ML} and $\delta^2 \cdot k_L a$ can be calculated from $\bar{t}$ and N_L obtained at different operating conditions.

[illegible]rphree plate efficiency (E_{ML}) is greatly affected [illegible] the liquid flow rates. It increases with decreasing liquid flow rates (Fig. 1). At low liquid flow rates, the residence time of the liquid on the tray increases, and so is the mass transfer between the phases [2, 3]. E_{ML} increases with increase of gas flow rates over the narrow range investigated [8, 9], i.e. F_A = 0.42 + 0.496. CO_2 concentration (3.6 to 12.6% V) in the incoming gas phase has no effect on the plate efficiency. This might be expected, since a small change in gas concentration may not change the physical properties of the system considerably.

Effect of liquid flow rates on $k_L a$ is given in Fig. 2. It is clear from these results that $k_L a$ is nearly independent of liquid flow rates. A drop in $k_L a$ value at low liquid flow rates could be artibuted to gas channeling, i.e. poor contact between the two phases, which could result in lower mass transfer rates and $k_L a$ values. A slight increase in mass transfer coefficient $k_L a$ was observed with increasing gas flow rates [1, 9], but the range examined was too small to warrant any generalized statement.

It has been stated earlier that the downcomer under certain circumstances may have a significant effect on the mass transfer taking place. In Figs. 1, 2 are shown the typical results for CO_2-water system for combined tray and downcomer and the tray alone. It is clear from the results that efficiencies are higher for the combined tray and downcomer compared with the tray alone. Thus it can be said that the downcomer does enhance the process of mass transfer in the unit. A comparison of the mass transfer coefficient, $k_L a$, for the sieve tray alone and the combined tray plus downcomer shows that the liquid phase $k_L a$ decreases for the combined tray plus downcomer when compared with the tray alone [9]. It is likely that this is predominantly due to a fall in interfacial area 'a' in the downcomer, but this could be stated purely as $k_L a$ itself may be less. It is not immediately possible to numerically compare the change in $k_L a$ with the change in Murphree efficiency, because the dimensionless variance δ^2 and mean

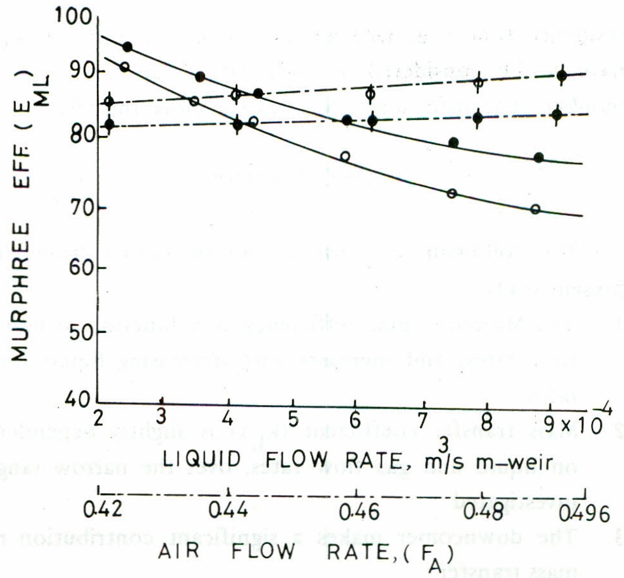

Fig. 1. Murphree tray efficiency (E_{ML}) as a function of liquid and gas flow rates.

(a) Liquid flow rates
F_A = 0.44
○ = Tray alone
● = Tray plus downcomer

(b) Gas flow rates
L = 4.56 m^3/s.m of weir
ϕ = Tray alone
ϕ = Tray plus downcomer

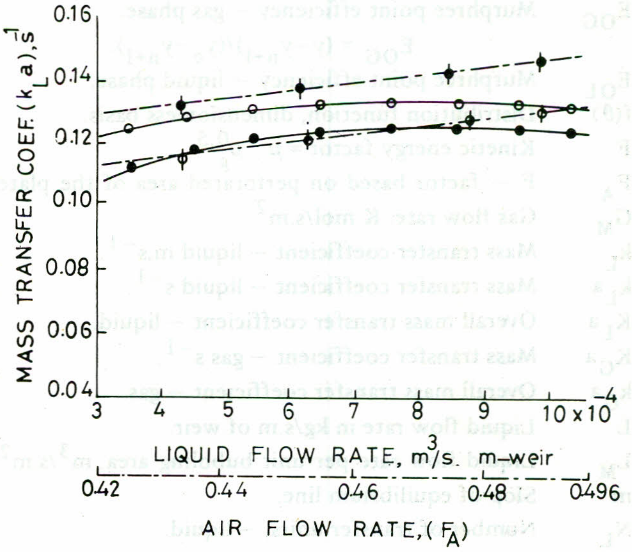

Fig. 2. Mass transfer coefficient ($k_L a$) as a function of liquid and gas flow rates. CO_2 composition in gas phase = 5.5% by volume.

(a) Liquid flow rates
F_A = 0.44
○ = Tray alone
● = Tray plus downcomer

(b) Gas flow rates
L = 4.5 m^3/s.m of weir
ϕ = Tray alone
ϕ = Tray plus downcomer

residence time t as well as mass transfer coefficient $k_L a$ have to be considered as they affect the liquid phase number of transfer units, N_L and therefore the efficiency.

CONCLUSIONS

The following conclusions can be drawn from the present study:

1. The Murphree plate efficiency is a function of liquid flow rates, and increases with decreasing liquid flow rates.
2. Mass transfer coefficient ($k_L a$) is slightly dependent on liquid and gas flow rates, over the narrow range investigated.
3. The downcomer makes a significant contribution to mass transfer.

Symbols

a.	Interfacial area per unit volume m^2/m^3.
E_{ML}	Murphree plate efficiency – liquid phase $E_{ML} = (X_{n-1} - X_n)/(X_{n-1} - X_{en})$
E_{MV}	Murphree plate efficiency – gas phase.
E_{OG}	Murphree point efficiency – gas phase. $E_{OG} = (y - y_{n+1})/(y_e - y_{n+1})$.
E_{OL}	Murphree point efficiency – liquid phase.
$f(\theta)$	Distribution function, dimensionless basis.
F	Kinetic energy factor = $\mu \cdot \rho_g^{0.5}$
F_A	F – factor based on perforated area of the plate.
G_M	Gas flow rate, K mol/s.m^2.
k_L	Mass transfer coefficient – liquid m.s^{-1}.
$k_L a$	Mass transfer coefficient – liquid s^{-1}.
$K_L a$	Overall mass transfer coefficient – liquid.
$K_G a$	Mass transfer coefficient – gas s^{-1}.
$k_G a$	Overall mass transfer coefficient – gas.
L	Liquid flow rate in kg/s.m of weir.
L_M	Liquid flow rate per unit bubbling area, $m^3/s.m^2$.
m	Slop of equilibrium line.
N_L	Number of transfer units, – liquid.
N_G	Number of transfer units, – gas.
N_{OL}	Overall number of transfer units – liquid.
N_{OG}	Overall number of transfer units – gas.
t	Residence time s.
$\bar{t}$	Mean residence time s.
X	Mole fraction of solute in liquid.
X_n	Value of x leaving the nth tray.
X_{n-1}	Value of x leaving the n–1th tray.
X_{en}	Equilibrium liquid composition corresponding to y_n
y	Mole fraction of solute in gas phase.
y_e	Equilibrium vapour composition corresponding to X.
y_{n+1}	Value of y leaving n+1th tray.
Z_f	Height of froth above the tray floor, m.
θ	Dimensionless time.
$\delta\theta^2$	Variance of residence time distribution, s^2.
δ^2	Dimensionless variance.
λ	$m.G_M/L_M$.
$\alpha \cdot \beta \cdot \gamma$	Constants in Eq. 7.
ρ_g	Gas density in kg/m^3.
Γ	Γ – Function.

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