Stripping of Trihalomethanes from Drinking Water in a Bubble-Column Aerator

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The desorption of trihalomethanes (THMs) in a cocurrent bubble column has been studied experimentally to validate a previously developed model. The experiments were carried out in a 0.2-m-diameter, 2.4-m-high, plexiglas bubble column, equipped with seven samplers alongside. The phases in contact were tap water and air. Three out of the four THMs normally found in drinking water were spiked in the feed tank to assure their detection in the analyses. After preparing the feed, each run was begun by setting the flow rates of water and air, and waiting until a steady state was attained. Then, samples from each sampler were taken and sent for chemical analysis. A total of 12 runs were performed with varying superficial gas and liquid velocities, and up to four concentration profiles were obtained for each run. The experimental results were then compared to the predictions of the model proposed by Estévez (1991). The agreement between calculated and experimental concentration profiles was quite good, indicating that the model can be used reliably either in the simulation or design of industrial scale equipment.

Introduction

In 1974, some trihalomethanes (THMs) were discovered to be formed during drinking water chlorination due to the presence of organic precursors. These THMs (chloroform, bromoform, bromodichloromethane, and dibromochloromethane) must be controlled because of their carcinogenic effects in human health (Tuthill and Moore, 1980). This and other reasons put the USEPA under obligation to amend the National Interim Primary Drinking Water Regulations, specifying a maximum contaminant level of 100 μg/L for total trihalomethanes (TTHMs) contents.

The problem of reducing the THM level in the drinking water delivered to the public can be approached in different ways: changing the disinfection method, eliminating the precursors for the THM formation, or eliminating the THMs once they are formed. Although some studies considering adsorption to remove THMs have been carried out, aeration or stripping is currently the preferred removal method. Culp (1984) and AWWA (1982) presented excellent reviews describing the various alternatives to approach this problem. Aeration techniques may also be used to remove other volatile organic contaminants (VOCs) present in water, as indicated in another review by Fleming (1989), or in the series of articles by Wilson and coworkers, the last one being Harkins et al. (1988). Most stripping studies (of THMs or other VOCs) have been carried out either in packed tower or in batch sparged aerators. Nirmalakhandan et al. (1987) presented a cost study of stripping five VOCs from water in a packed tower and found optimal operating conditions in each case. Kavanaugh and Trussell (1981) compared various alternatives for air-stripping treatment processes, including spray towers, diffused aeration, stirred-tank aeration, and packed towers, and found that the selection of the equipment depends on the desired removal level and the value of the Henry's constant. Roberts and Levy (1985) examined and compared the energy requirements for stripping THMs in stirred-tank aerators and packed columns. They recommended the former in general unless there are space limitations or air-quality problems. Umphres et al. (1983) conducted two pilot studies to remove THMs by packed-tower aeration. One of them was particularly important due to the unusually high amount of bromoform, which accounted for 60% to 80% of the TTHM contents. They found that packed-tower aeration appeared to be the most cost-effective alternative to meet USEPA regulation in this case. Models for

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stirred- and sparged-tank aerators have also been presented, for example, by Roberts et al. (1984), Harkins et al. (1988), and Cho and Wakao (1988).

Since packed towers or columns are operated in a countercurrent mode, their main limitation is the air-to-water ratio resulting in flooding. This is particularly important for those VOCs (including THMs), which are less volatile such as bromoform and dibromochloromethane. Bubble-column contactors are devices that have been increasingly used as multiphase reactors and mass-transfer equipment for a number of applications (Shah et al., 1982; Fleming, 1989; Pino et al., 1990; Mixon et al., 1990). They are simple and offer a higher interphase mass-transfer surface, and their operation is quite versatile. While in reaction engineering applications modeling bubble columns is quite case-specific (Pinares and Estévez, 1991), for mass-transfer applications a general model can be developed. Estévez (1990) presents a model for air-stripping volatile components from water in bubble columns which can be readily modified for other mass-transfer applications.

This work has been undertaken to study the feasibility of using bubble columns to reduce the THM contents of drinking water to meet USEPA regulations. This article's goal is to verify experimentally the model proposed by Estévez (1990, 1991) to use it in the simulation and design of bubble columns for this specific application.

Mathematical model for bubble column aerator

The mathematical model presented here follows that of Estévez (1990, 1991). This model predicts the concentration of the trihalomethane as a function of the height of a column operated in the cocurrent mode. The model has been developed from a differential balance in the liquid phase coupled with an overall mass balance in both phases. This approach is similar to standard procedures followed, for example, by Treybal (1980). The basic difference between the two approaches (this and Treybal's) is the axial dispersion phenomenon. In conventional mass transfer equipment, such as packed-tower absorbers and strippers, axial dispersion is not important and thus neglected. Backmixing of the liquid phase, however, is rather high in bubble columns, leading in some extreme cases to nearly perfect mixing. This term causes the differential equation to be second order, and therefore, the solution and design method are quite different from those used for the above-mentioned conventional units. Mass transfer between phases is taken into account by overall mass-transfer coefficients so that the resulting model applies to both gas- and liquid-phase controlled mechanisms.

The details of the development can be found elsewhere (Estévez, 1990, 1991), and the differential equation obtained is:

$$D_L \frac{d^2C}{dZ^2} + \left(\frac{K_L a S D_L - V_L}{V_L}\right)\frac{dC}{dZ} - K_L a(S+1)C = -K_L a S C_1 \qquad (1)$$

which is a second order, linear, ordinary, nonhomogeneous differential equation, requiring two boundary conditions. The boundary conditions follow from Danckwerts and have been extensively cited in the literature (for example, Froment and Bischoff, 1990):

$$C_1 = C(0^+) - \frac{D_L}{V_L}\frac{dC(0^+)}{dZ} \qquad (2)$$

and

$$\frac{dC(L^-)}{dZ} = 0 \qquad (3)$$

The solution of the differential equation subjected to the boundary conditions is:

$$C = \frac{C_1}{S+1}\left[S - \frac{R_2 e^{R_2 L} e^{R_1 Z} - R_1 e^{R_1 L} e^{R_2 Z}}{R_1 e^{R_1 L}\left(1 - \frac{D_L R_2}{V_L}\right) - R_2 e^{R_2 L}\left(1 - \frac{D_L R_1}{V_L}\right)}\right] \qquad (4)$$

where R_1 and R_2 are the roots of the characteristic equation of Eq. 1 given by:

$$R_{1,2} = \frac{V_L - \frac{K_L a S D_L}{V_L} \pm \sqrt{\left(V_L - \frac{K_L a S D_L}{V_L}\right)^2 + 4 D_L K_L a (S+1)}}{2 D_L} \qquad (5)$$

where S is the conventional stripping factor, which in this case is given by:

$$S = \frac{\rho_L R T}{H M_w}\frac{V_L}{V_G} \qquad (6)$$

As seen in Eqs. 5 and 6, the concentration profile in the liquid phase is affected by operating variables (V_L, V_G), geometric parameters (L, D_c), and mass-transfer and axial dispersion coefficients $(K_L a, D_L)$.

Experimental Studies

Apparatus

Figure 1 shows the experimental apparatus, which consists mainly of a plexiglas column, a self-metering pump, two feed tanks, filter devices (Parker Pneumatic), and a rotameter. The column is 20 cm in diameter and about 2.4 m in height, and has seven sampling ports alongside, which can be used either to take samples or to measure local pressure. The column was designed to be operated in the cocurrent mode by mixing the two phases at a tee located below the bottom cone; no gas distributor was used in this work. The pump is Milton-Roy, model HMRB1-0711-140SM. The two feed tanks are Nalgene, 400 L each with 5/16 in. (8 mm) wall thickness, one of which is equipped with a stirrer. The rotameter is Brooks Instrument and has a flow rate range of 1.34 to 13.4 std. ft^3/min (0.6 to 6.3 L/s). The gas line is 3/8 in. (9.5 mm) stainless steel tubing, and most fittings are Swagelok brand. The liquid line is PVC pipe, schedule 80, with nominal diameter of 1/2 in. (13.9 mm ID) before the column, and 1 in. (24.3 mm ID) after the column. All fittings are PVC schedule 80.

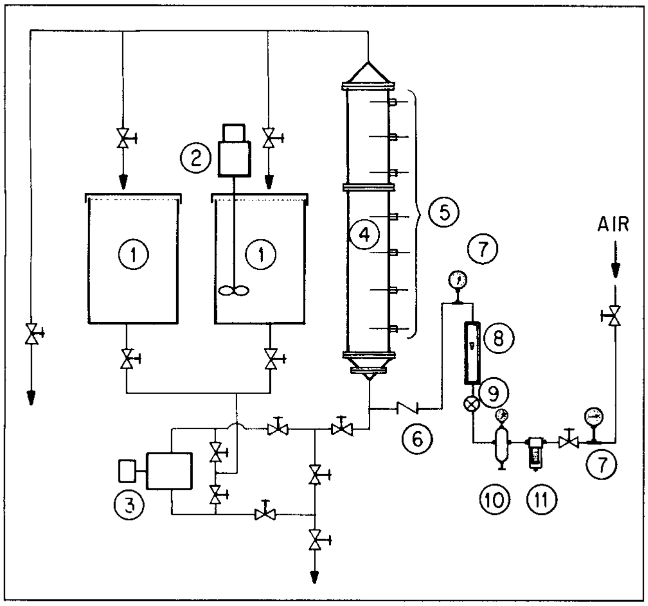

Figure 1. Experimental Apparatus.

(1) Feed tanks
(2) Stirrer
(3) Pump
(4) Column
(5) Sample ports
(6) Check valve
(7) Bourdon gauge
(8) Rotameter
(9) Metering valve
(10) Pressure regulator
(11) Air filter

Materials

The experiments were carried out using water, chloroform, dibromochloromethane, bromoform, air, and sodium thiosulfate to stabilize the samples. Tap water was used with no further treatment. THMs and sodium thiosulfate were analytical-grade reagent (chloroform is from Mallinckrodt Chemicals, while bromoform and dibromochloromethane are from Fluka Chemicals). Utility compressed air was treated in-line by passing it through a water separator and an air filter, as indicated in Figure 1.

Experimental procedure

Vials to take the samples were prepared beforehand by adding about 10 mg of sodium thiosulfate to halt any further reaction to form THMs. Once they were prepared, the experiment began by spiking chloroform, bromoform, and dibromochloromethane into the feed tank. Only the tank equipped with a stirrer was used in this work. Various samples of the tap water were taken as blank or control samples. Then, the stirrer was turned on to assure a uniform concentration in the tank. This was done a few minutes before pumping the water through the column, and the stirrer was kept running throughout the experiment. Various samples were taken from the feed tank to determine the feed concentration. Water was then pumped through the column until it was filled. At this point, the air was allowed to flow cocurrently with the water to the column until a steady state was reached, which took about 15 minutes. After about 15 to 30 min, samples were taken at each sampling port. Variability and reproducibility were randomly tested. For variability, two samples were taken one right after the other from a chosen port. For reproducibility, a larger sample was taken, well mixed, and divided into two vials. Then, the temperature of the water, pressure at the rotameter and column inlet, water and air flow rates, and absolute atmospheric pressure were read and recorded. The gas holdup was measured at the end of each run by closing the air and water valves at the same time, and by measuring the position of the interphase after air and water had been completely disengaged. Finally, samples were analyzed by gas chromatography. Due to the low concentrations of THMs in the samples, a purge and trap technique had to be used to concentrate each sample before injecting it into the gas chromatograph. The analyses were carried out at an EPA-certified external laboratory, in the Drinking Water Laboratory of the Puerto Rico Department of Health. The detectability limit of the technique is 0.5 ppb.

Results

Operating conditions

Table 1 summarizes the operating conditions used in this work: the liquid velocity, the gas velocity, approximate volume of THMs added (spiked-in) to the feed tank, and resulting concentrations in the feed tank. The experimental values of gas holdup are also shown. To characterize the THM concen-

Table 1. Experimental Conditions

						Feed Concentrations				
Run	V_G cm/s	V_L cm/s	v_c μL	v_{db} μL	v_b μL	$CHCl_3$ ppb	BDCM* ppb	DBCM** ppb	$CHBr_3$ ppb	ϵ_G
11	2.89	0.31	40	0	0	56.7	4.5	-	-	0.118
12	2.89	0.31	40	0	0	56.7	4.5	-	-	0.118
13	2.89	0.31	40	0	0	56.7	4.5	-	-	0.118
14	2.89	0.31	40	0	0	41.2	5.9	-	-	0.114
20	3.61	0.31	40	0	0	37.8	6.1	-	-	0.131
30	4.39	0.31	40	0	0	30.5	4.8	-	-	0.141
40	2.89	0.62	40	40	40	20.4	5.1	43.5	106.2	-
50	3.61	0.62	40	40	40	20.4	5.1	43.5	106.2	0.137
60	4.39	0.62	40	40	40	25.4	5.2	50.4	170.3	0.141
70	2.89	0.70	40	15	15	38.4	6.5	12.1	13.9	0.137
80	3.61	0.70	40	15	15	36.5	5.1	4.7	15.5	0.125
90	4.39	0.70	40	20	20	25.8	5.1	37.8	44.0	-

*bromodichloromethane
**dibromochloromethane

Table 2. Results of Tap Water Analysis

Run	Sample No.	$CHCl_3$ ppb	$CHCl_2Br$ ppb	$CHClBr_2$ ppb	$CHBr_3$ ppb
11, 12, 13	1	44.6	5.7	<0.5	<0.5
	2	41.5	3.1	<0.5	<0.5
	3	56.3	2.7	<0.5	<0.5
40, 50, 60	1	50.2	12.0	1.8	<0.5
	2	17.9	5.7	2.4	<0.5
	3	18.5	6.0	1.4	<0.5
	4	17.8	5.8	1.3	<0.5
70, 80	1	38.9	7.4	<0.5	<0.5
	2	62.7	7.4	15.8	89.6
	3	42.4	7.7	2.4	18.6
	4	42.4	7.9	<0.5	<0.5

trations of the tap water, control samples were taken and analyzed for every run. The corresponding results are shown in Table 2. None of the individual values therein exceeds the EPA limit of 100 μg/L. Water temperature and atmospheric pressure did not vary much from run to run, so average values of 25°C and 94 kPa were used in the computations.

Superficial gas velocity and gas holdup

In bubble-column literature, V_G is usually reported with no indication of the conditions at which it has been calculated. This may lead to inconsistencies among research results, as indicated, for example, by Reilly et al. (1986). In this work, the gas velocity has been computed as:

$$V_G = \frac{Q_G}{A_c} \tag{7}$$

where A_c is the column cross-sectional area, and Q_G is the volumetric gas flow rate. The rotameter reading is given in ft³/min measured at the standard conditions of 14.7 lb_f/in^2. (0.01 kPa) and 70°F (21°C). Thus, Q_G is to be computed as:

$$Q_G = Q_R \frac{P_o P_R}{P_c^2} \tag{8}$$

where Q_R is the rotameter reading, and P_o, P_R, and P_c are the standard, rotameter, and column absolute pressures, respectively. The pressure of the column is measured at the first pressure tap; therefore, the superficial gas velocity computed by Eq. 7 represents V_G at inlet conditions.

Gas holdups were measured as indicated above and were correlated using the Zuber-Findlay theory (Zuber and Findlay, 1965), giving the following equation for ϵ_G:

$$\epsilon_G = \frac{V_G}{3.365(V_G + V_L) + 0.142} \tag{9}$$

The correlation coefficient, R^2, for this correlation was 0.9992.

Predicted concentration profiles

The predicted concentrations as a function of position along the column were computed by using Eq. 4. The properties of the gas and liquid phases were taken as those of air and water, respectively. A number of correlations are available for bubble column design (Shah et al., 1982). The most recommended ones have been chosen for illustrative purposes in this work. The mass-transfer coefficient was estimated from the correlation by Akita and Yoshida (1974), which requires, among other variables, the gas-phase holdup and the molecular diffusivity of the solute in the liquid phase. The latter was computed by the Wilke/Chang equation (as cited, for example, by Treybal, 1980). Equation 9 was used to compute the gas holdup. The infinite-dilution activity coefficient for chloroform was taken from Barr and Newsham (1987) who measured it experimentally. The one for bromodichloromethane was computed using the UNIFAC method. For bromoform and dibromochloromethane, Henry's constants were obtained from Fleming (1989). The axial-dispersion coefficient was calculated by the correlation of Joshi (1980) which agrees with preliminary measurements by Cayuela and Estévez (1990) using tracer experiments and the residence-time-distribution analysis. The various properties used in the calculations are presented in Table 3.

Table 3. Parameters for the Analytical Solution at 298.15 K

Contaminant	γ^∞	p^{sat} kPa	H MPa	$V_b \times 10^6$ m³/mol	$D_i \times 10^{10}$ m²/s
Chloroform	821	20.8	17.1	80.5	9.86
Bromodichloromethane	204	101.3	20.7	82.7	10.00
Dibromochloromethane	38	117.4	4.45	85.2	9.97
Bromoform	9.8	33.7	0.33	89.6	9.80
References	(a)	(b)	(a)	(c)	(d)

(a) H or γ^∞ from literature or estimated. The other computed by $H = \gamma^\infty p^{sat}$.
(b) Fleming (1989).
(c) Perry's *Chemical Engineers' Handbook*, 6th edition.
(d) From Wilke/Chang correlation.

Figures 2 to 7 show the experimentally-measured, and computed concentration profiles (normalized with respect to the feed concentration) for the various cases studied.

Discussion

The reproducibility tests showed that the analyses were indeed reproducible. For example, for chloroform the dupli-

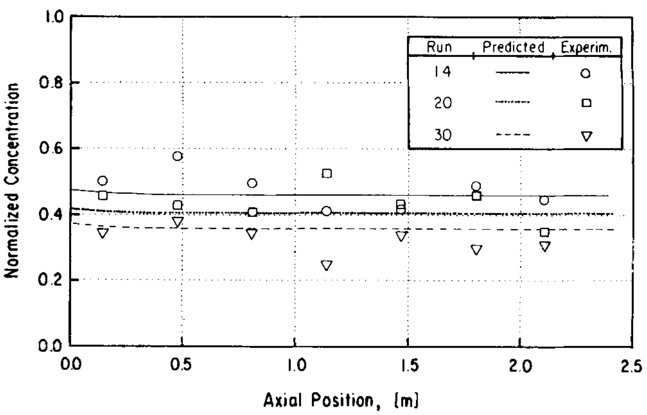

Figure 2. Experimental and predicted concentrations for chloroform in runs 14, 20 and 30.

Effect of gas superficial velocity for $V_L = 0.31$ cm/s.

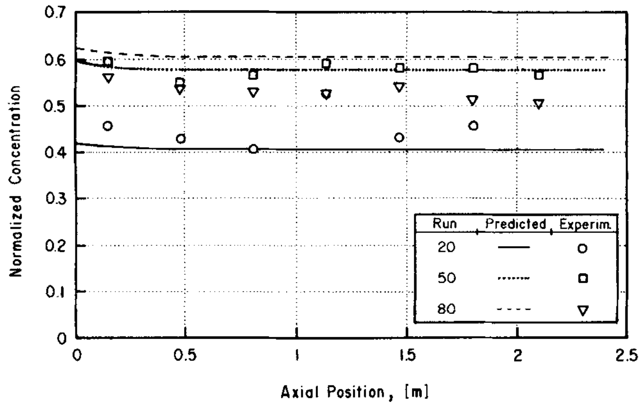

Figure 3. Experimental and predicted concentrations for chloroform in runs 20, 50 and 80.

Effect of liquid superficial velocity for $V_G = 2.89$ cm/s.

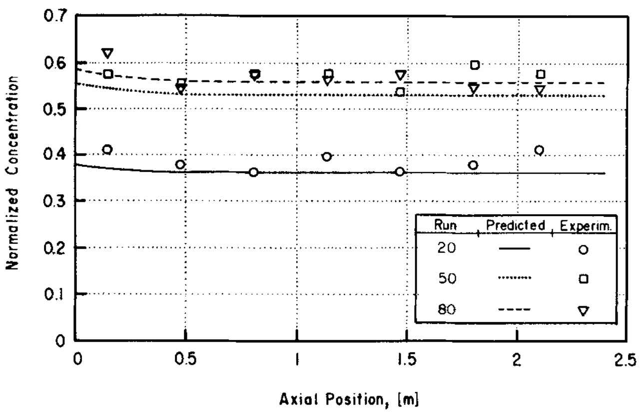

Figure 5. Experimental and predicted concentrations for bromodichloromethane in runs 20, 50 and 80.

Effect of liquid superficial velocity for $V_G = 3.61$ cm/s.

cations were within 3.5%, whereas the variability was about 9%. This is an average of eight variability and 12 reproducibility tests. These figures are the main cause of the scatter shown in the experimental concentration profiles. The scatter tends to be higher for those runs with concentrations closer to the detectability limit.

An inspection of the concentration profiles presented reveals that there is good agreement between experimental and predicted values. They show the following general characteristics:

- There is a jump in concentration at the entrance of the column.
- There is a region in which concentration of THMs in water decreases.
- There is a plateau region in which no further mass transfer occurs.

This general pattern is followed by the predicted curves and to a less clear extent by the experimentally-measured profiles, the latter being somewhat hidden by the aforementioned scatter.

The agreement between experimental and predicted results demonstrates that the main physical aspects of the process have been taken into account by the model. At the same time, it is possible to interpret the general characteristics of the profiles on physical grounds. First, the jump in concentration is a consequence of the difference in mixing level in the incoming tube and in the column. This is mathematically taken into account by the Danckwerts boundary condition at $Z=0$, Eq. 2. There are obviously two extreme cases: $D_L=0$ which

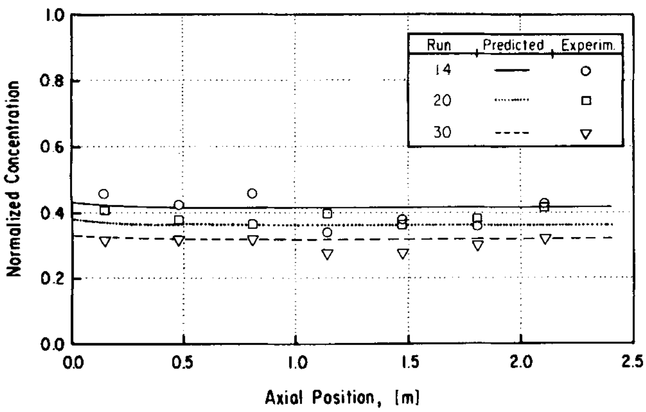

Figure 4. Experimental and predicted concentrations for bromodichloromethane in runs 14, 20 and 30.

Effect of gas superficial velocity for $V_L = 0.31$ cm/s.

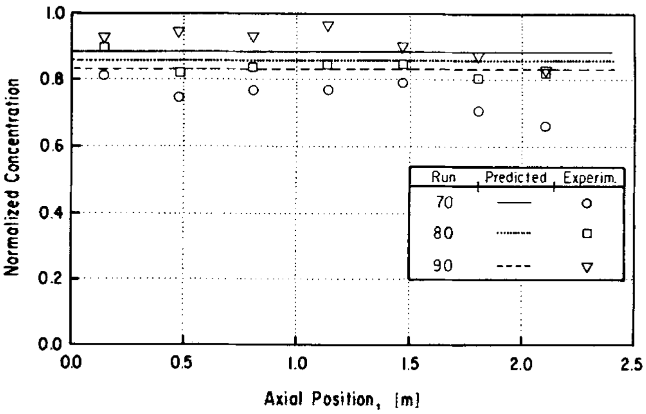

Figure 6. Experimental and predicted concentrations for dibromochloromethane in runs 70, 80 and 90.

Effect of gas superficial velocity for $V_L = 0.70$ cm/s.

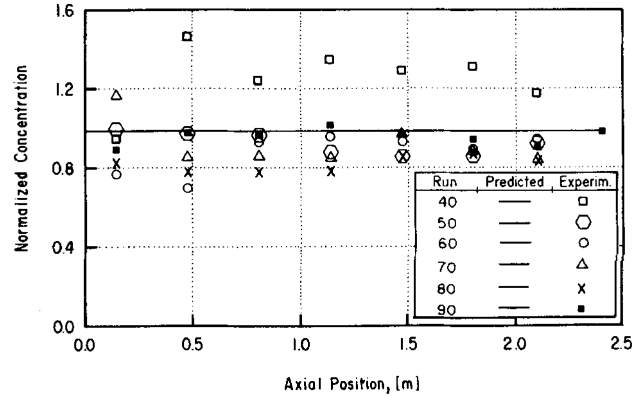

Figure 7. Experimental and predicted concentrations for bromoform in runs 40 to 90.

would cause no jump; and $D_L \rightarrow \infty$ which would result in a flat concentration profile or a jump in concentration equal to $C_1 - C(L^+)$.

The shape of the concentration profile is typical of cocurrent contact. Near the entrance there is a marked concentration change. As driving force decreases, mass-transfer rate decreases until equilibrium is reached, which is reflected in the curves by a plateau region in which no mass transfer takes place. The predicted concentration in this region corresponds to a negligible second term in the bracket of Eq. 4. Under these conditions, the exit concentration is:

$$C_{eq} = C_1 \frac{S}{S+1} \tag{10}$$

Incidentally, Eq. 10 can also be obtained directly from Eq. 1 by crossing out the derivative terms that are zero in the plateau region.

For the conditions tested in this work, all concentration profiles reached the plateau region. In practical applications, this means that equilibrium between phases is reached rather rapidly, thus two or more in-series columns could be used to achieve any removal level. This is particularly important for those less volatile THMs such as dibromochloromethane and bromoform, which might not be removed in one theoretical stage. Figures 2, 4 and 6 show the effect of the gas superficial velocity: V_L is the same for all data in each plot for chloroform, bromodichloromethane, and dibromochloromethane, respectively. The trend is as expected: lower concentration (higher removal) at higher V_G. Similarly, Figures 3 and 5 show the effect of V_L on the concentrations profiles for chloroform and bromodichloromethane, respectively, and again data follow the trend as expected. Finally, for bromoform, all data were plotted in Figure 7 since computed concentrations were about 1, and in effect of either phase supercritical velocity would be observed.

Figures 8 and 9 show the removal percent, that is, the fraction of the THMs entering the column that is removed by the air as a function of the gas and liquid velocities. It is computed as:

$$R_P = 100 \frac{C_1 - C(L^+)}{C_1} \tag{11}$$

where the exit concentration $C(L^+)$ is obtained from the experimental data. There are two factors influencing the exit concentration: the mass-transfer rate and the overall mass balance, which determines the equilibrium concentration. The mass-transfer rate determines how fast the plateau region is reached, but does not affect the final concentration in the plateau region. Therefore, this effect is important only in the case where the contact time is not long enough for the exit concentration to reach the plateau value. As mentioned above, the plateau concentration was reached for all cases in this work; therefore, the mass-transfer rate did not affect the removal percent. Under these conditions, the removal percent is affected only by the mass-transfer balance equations coupled with the equilibrium relationships. The exit concentration in this case is given by Eq. 10, which upon substitution in Eq. 11 gives:

$$R_P = \frac{100}{S+1} \tag{12}$$

From Eq. 6 it is observed that S is directly proportional to V_L and inversely proportional to V_G. Hence, the trends observed in Figures 8 and 9 are in agreement with what Eq. 12 predicts: higher removal percents at higher gas velocities and lower removal percents at higher liquid velocities. It is interesting to note that the mass-transfer rates always increase as either phase velocity increases. This means that exit concentration will always decrease when gas velocity increases, since it enhances mass transfer *and* lowers the equilibrium concentration. The effect of liquid velocity, however, cannot be predicted when the mass-transfer effect and equilibrium effects compete with each other, since the effect of liquid velocity on these factors is in the opposite direction.

Finally, Figure 10 is intended to verify whether the experimental removal percent data follow Eq. 12. To this end, the corresponding data for chloroform and bromodichloromethane were chosen. The data points fall on a straight line that in turns falls quite close to the diagonal line representing Eq. 12. This plot is useful particularly in design problems, where the removal percent is usually specified and the required value of S can be read from Figure 10 or calculated from Eq. 12. To illustrate this, consider the following case. Umphres et al. (1983) found that bromoform could be removed to the extent

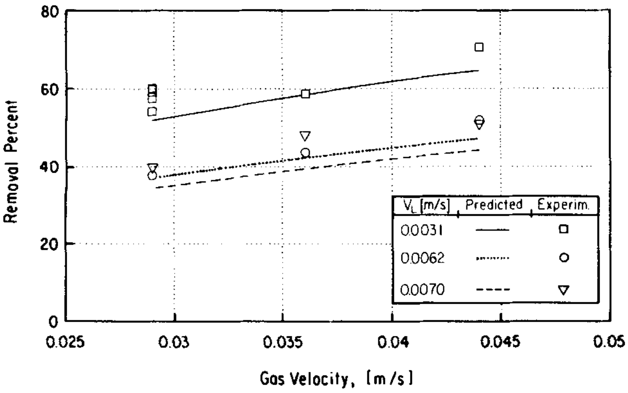

Figure 8. Effect of superficial velocities on removal percent for chloroform.

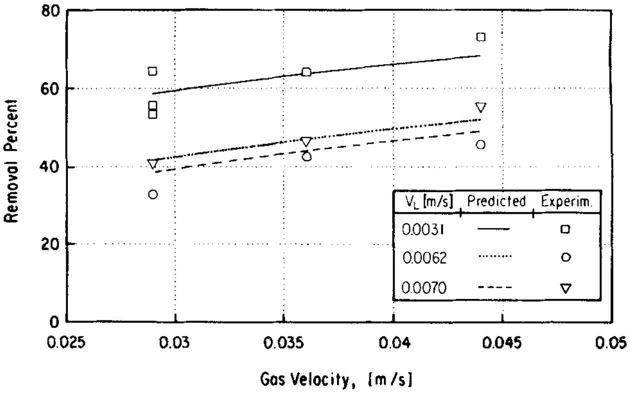

Figure 9. Effect of superficial velocities on removal percent for bromodichloromethane.

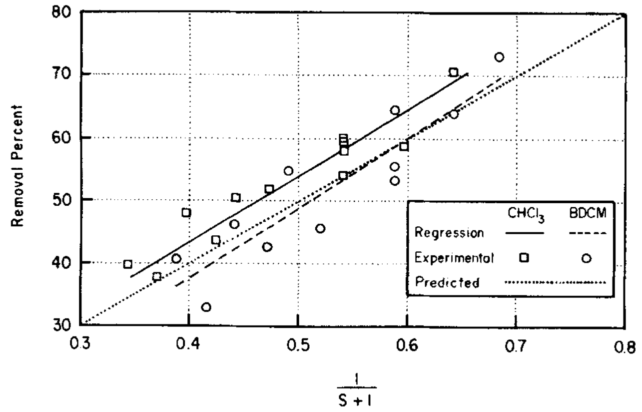

Figure 10. Removal percent as function of the stripping factor for chloroform.

of 42% in a packed-bed aerator. Although no detectable removal was observed in this work, the model predicts that this can be achieved with a stripping factor of $S=1.38$. For the lowest liquid velocity used, a gas velocity of 0.92 m/s would be required for a removal percent similar to that reported by Umphres et al. (1983).

Conclusions

The main goal of this work was to test the mathematical model developed by Estévez (1990, 1991) by comparing it to experimental results. From the discussion in the previous section, it can be concluded that the model indeed represents the experimental data. It can be further concluded that bubble columns can be used as equilibrium stages which can be set up in series for higher removal levels. The model can predict the trends of the percent removal with operating variables that will facilitate any further studies of the feasibility of using bubble-column aerators to remove THMs from drinking waters.

Acknowledgment

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Notation

a = gas-liquid interfacial area per unit total volume, m^2/m^3
A_c = column cross-sectional area, m^2
C = liquid-phase concentration of the THM, kg/m^3
$C(0^+)$ = value of C at $Z=0$, kg/m^3
C_{eq} = equilibrium concentration, kg/m^3
C_1 = feed concentration (tank), kg/m^3
$C(L^+)$ = outlet concentration, kg/m^3
$C(L^-)$ = concentration in the column at $Z=L$, kg/m^3
D_c = inside diameter of the column, m
D_i = diffusion coefficient of the THM in the liquid phase, m^2/s
D_L = axial dispersion coefficient in the liquid phase, m^2/s
e = 2.71828...
H = Henry's constant, Pa
K_L = overall mass-transfer coefficient based on $(C\text{-}C^*)$, m/s
L = length of the bubble column, m
M_w = molecular weight of water, kg/kmol
p^{sat} = vapor pressure of the THM, Pa
P_c = inlet column pressure, Pa
P_o = standard pressure, Pa
P_R = rotameter pressure, Pa
Q_G = gas volumetric flow, m^3/s
Q_R = rotameter reading of gas volumetric flow, m^3/s
R = universal gas constant, J/mol·K
R_P = removal percent
R_1, R_2 = roots of characteristic equation of Eq. 5, 1/m
S = stripping (desorption) factor
T = absolute temperature, K
v_c = chloroform volume added, μL
v_{db} = dibromochloromethane volume added, μL
v_b = bromoform volume added, μL
V = specific molar volume of liquid THM at its boiling point
V_G = superficial gas velocity, m/s
V_L = superficial liquid velocity, m/s
V_b = specific molar volume, m^3/mol
Z = axial position on the column, m

Greek letters

γ^∞ = infinite-dilution activity coefficient
ϵ_G = gas holdup
ρ_L = density of the liquid phase, kg/m^3

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