

GROWTH OF MIXED CULTURES AND OXYGEN TRANSFER, IN TOWER SYSTEMS
WITH MOTIONLESS MIXERS

2148-
56088

by

KENNETH H. HSU

B. S., Kansas State University, 1972

A MASTER'S THESIS

submitted in partial fulfillment of the
requirements for the degree


MASTER OF SCIENCE

Department of Chemical Engineering

KANSAS STATE UNIVERSITY
Manhattan, Kansas

1974

Approved by:


Major Professor

**THIS BOOK
CONTAINS
NUMEROUS PAGES
WITH THE ORIGINAL
PRINTING BEING
SKEWED
DIFFERENTLY FROM
THE TOP OF THE
PAGE TO THE
BOTTOM.**

**THIS IS AS RECEIVED
FROM THE
CUSTOMER.**

LD
2668
T4
1974
H855
C. 2

TABLE OF CONTENTS

Document

CHAPTER I	INTRODUCTION	1-1
CHAPTER II	PRESSURE DROP, GAS HOLD UP, AND OXYGEN TRANSFER IN TOWER SYSTEMS	2-1
	Introduction	2-1
	Materials and Methods	2-3
	Results and Discussion	2-4
	Conclusions	2-9
	References	2-10
CHAPTER III	BIOLOGICAL WASTE TREATMENT IN TOWER SYSTEMS	3-1
	Introduction	3-1
	Materials and Methods	3-4
	Results and Discussion	3-6
	Concluding Remarks	3-21
	Nomenclatures	3-22
	References	3-24
CHAPTER IV	OXYGEN TRANSFER TO MIXED CULTURES IN TOWER SYSTEMS	4-1
	Introduction	4-1
	Theoretical	4-2
	Experimental	4-5
	Results and Discussion	4-8
	Conclusions	4-12
	References	4-14
CHAPTER V	CONCLUDING REMARKS	5-1
	Conclusions	5-1
	Recommendations	5-2

ACKNOWLEDGMENT

CHAPTER I

INTRODUCTION

Continuous biological reactor systems are becoming more popular as new processes are developed for single cell protein production, fermentation, and waste treatment. Many of these biological processes are aerobic and depend upon the rate at which oxygen is being supplied to the liquid phase.

The use of mixing to enhance oxygen transfer rates is well known. Mechanical agitation using surface aeration units and turbine aeration systems to create good mixing are common in biological waste treatment. Extensive use is also made of diffused aeration devices. For a bubble column without mechanical agitation good mixing is achieved only at high aeration rates. To improve the mixing and oxygen transfer at low aeration rates motionless mixers may be used. The functions of these motionless mixers are to create better liquid mixing and to redistribute the gas phase.

The purpose of this study is to investigate the application of a sieve tray column and a Koch mixer column for activated sludge treatment. The thesis is divided into three parts: (1) physical characteristics of the systems, (2) biological waste treatment characteristics of the systems, and (3) oxygen transfer characteristics of the systems during activated sludge treatment.

In the physical characteristics section, the results of gas hold up and pressure drop measurements at various gas and liquid flow rates are presented. Steady state dissolved oxygen concentration measurements at the inlet and outlet of the column are used to determine mass transfer coefficients, $K_L a$. The results are compared with those obtained from a bubble column.

In the waste treatment characteristics section, the effects of dilution rate on the mixed liquor suspended solids (MLSS) and chemical oxygen demand (COD) concentrations are presented. The effects of wall growth, cell sedimentation, and plugging on the column operation and washout are also presented. Mathematical models are employed to interpret and simulate some of the experimental results.

In the oxygen transfer section, a modified dynamic method which was utilized to obtain values of $K_L a$ from the oxygen concentration-time curve is introduced. Values of $K_L a$ at various air flow rates are presented for the two columns and compared with the available literature. Values for the oxygen utilization term, r_X , from direct sampling and by calculation using the determined values of $K_L a$ are presented and compared.

CHAPTER II

PRESSURE DROP, GAS HOLD UP, AND OXYGEN TRANSFER IN TOWER SYSTEMS

Introduction

Oxygen transfer is an important consideration in aerobic fermentations. The large scale production of single cell protein and aerobic biological waste treatment processes use substantial amounts of energy for oxygen transfer. As energy costs increase, oxygen transfer efficiency becomes more important. Oxygen transfer rates vary significantly; in single cell protein production, British Petroleum [1] has reported values of 15 kg/(cu m hr) for hydrocarbon fermentations. In biological waste treatment oxygen transfer rates are frequently much lower [2] with values often being only 1 to 2% of those discussed by British Petroleum. The present study is directed toward oxygen transfer in biological waste treatment and other aerobic processes where oxygen transfer rates are less than 1.5 kg/(cu m hr). The effect of motionless mixers on pressure drop, hold up, oxygen transfer rate, and oxygen transfer efficiency is investigated using a 3 inch column.

Oxygen transfer in bubble columns and columns with motionless mixers to promote oxygen transfer depends on a number of factors. Mass transfer rates are very high during bubble formation. The values of the mass transfer coefficient varies significantly in a bubble column with large values being associated with high surface renewal rates. Coalescence and break up, bubble formation, bubble shear and a turbulent liquid surface are examples of conditions where the mass transfer coefficient is large. Small rigid air bubbles less than about 2 mm in diameter frequently have been found to have

much smaller mass transfer coefficients [3]; however, the interfacial area is usually large when bubble size is small. Important factors in oxygen transfer are usually bubble formation, bubble size, gas hold up, bubble shear, multiphase flow behavior, and rate of coalescence and break up. As energy costs increase more capital can be allocated to motionless mixers which increase oxygen transfer efficiency.

Mass transfer in bubble columns has been investigated extensively [4-11]. Akita and Yoshida [8] have investigated the effects of liquid properties on mass transfer coefficients in bubble columns. Dimensionless correlations for bubble size and liquid phase mass transfer coefficient have been developed [9] by experimentally measuring bubble size distribution, gas hold up, and oxygen absorption. The effects of gas flow rate, nozzle diameter, column diameter, temperature, and liquid flow rate on gas hold up and oxygen transfer were investigated.

Gas hold up in a perforated plate column has been investigated by Fair et al. [12] and Kitai et al. [13,14]. Fair et al. found that the hold up increased when 20 sieve trays were inserted into the bubble column and that movement (reciprocating motion at 1050 cycles/min) further increased hold up.

Gas hold up and oxygen transfer data were also reported by Kitai et al. [13,14] for perforated plate multi-stage tower systems. The effects of the column diameter on the gas hold up and oxygen transfer coefficient were found to be very small. Values of the oxygen transfer coefficient were found to be higher than those obtained by Yoshida and Akita [7], at the same superficial gas velocity.

In this study, a bubble column, a sieve tray column, and a Koch mixer column are investigated for their gas hold up, pressure drop and mass

transfer characteristics. The data obtained from the sieve tray column and the Koch mixer column are to be compared with those of the bubble column.

Materials and Methods

The sieve tray column system consisted of a 3.664 liter bubble column divided into 3 compartments by horizontally positioned sieve trays. A schematic diagram of the system set up is shown in Figure 1. Each sieve tray was made of a 1/16 inch thick brass plate with hole diameter of 3/64 inch and free open area to column cross sectional area ratio of 3%.

The Koch mixer column set up was similar to that of the sieve tray column as shown in Figure 2a. Instead of using sieve trays, Koch mixers were used to distribute the gas. The Koch mixers were made of stainless steel sheets bent into wave shape and put together with the layers parallel to each other and the corrugation angle of the adjacent layers reversed with respect to the mixing axis. The mixers used in this study were of the BY type [15]; they are 3 inches tall and 3 inches in diameter with a layer height of 1/4 inch (see Figure 2b).

The gas flow rate to the column was measured by a gas rotameter and was controlled by a valve attached to the rotameter.

The gas was supplied by a compressed oxygen bottle. The gas entered the column through a 1/4 inch diameter nozzle. A pressure gauge was installed between the gas rotameter and the gas inlet to the column to determine the oxygen pressure. Liquid flow rate was controlled by checking the volumetric flow rate at each gas flow rate because of the change in pressure drop at various gas flow rates. A manometer was connected near the inlet and outlet of the column to measure the pressure drop over a 25 inch column section.

Values of gas hold up were obtained by simultaneously shutting off the gas and liquid flows and measuring the space occupied by the gas phase.

Dissolved oxygen was measured with a lead-silver galvanic cell oxygen probe (Precision Scientific). Liquid temperatures were measured with a thermister for the purpose of saturation calculations. All experimental data were taken at an average temperature of 19°C and the deviations were in the range of $\pm 1.0^\circ\text{C}$.

Dissolved oxygen concentration was measured in the inlet storage reservoir and at the top of the column close to the outlet under steady state conditions. No oxygen concentration differences could be detected between the column outlet and a distance of about 5 inches from the outlet. Also, the dissolved oxygen concentration was measured across the column cross section, and no radial concentration gradient was observed.

Gas flow rates were in the range of $2 \text{ ft}^3/\text{hr}$ (56.64 l/hr) to $20 \text{ ft}^3/\text{hr}$ (566.4 l/hr). The corresponding gas superficial velocities ranged from 0.346 cm/sec to 3.46 cm/sec. Liquid flow rates and superficial velocities ranged from values of 10.8 l/hr and 0.0658 cm/sec to 78 l/hr and 0.476 cm/sec, respectively. The liquid flow ranges were chosen so that oxygen transfer could be accurately measured. All experiments were carried out under co-current flow, and an oxygen-water system was used.

Results and Discussion

Gas hold up

Gas hold up data are shown in Figures 3, 4 and 5 for the bubble column, the sieve tray column, and the Koch mixer column, respectively. The results show that the gas hold up in the column increases as the gas flow rate

increases, and decreases as the liquid flow rate increases. Gas hold up is much greater in the sieve tray column and the Koch mixer column. These higher values suggest that the motionless mixers help to retain more gas in the column and, thus, increase the gas phase residence time. Similar results have been reported by Fair et al. [12] and Kitai et al. [13].

The gas hold up data at 10.8 l/hr liquid flow rate for the bubble column are very close to those reported by Yoshida and Akita [7]; however, Yoshida and Akita [8] found that hold up was not significantly affected by changes in liquid flow rate. The present results appear to differ significantly from those of Yoshida and Akita [8] with respect to the effect of liquid flow rate on gas hold up.

Pressure drop

Figures 6, 7, and 8 show the pressure drop data for the bubble column, the sieve tray column, and the Koch mixer column, respectively. The data are those measured over a 25 inch column section. Pressure drop over the whole column length can be estimated by assuming constant pressure drop per unit column length.

The pressure drop due to liquid and gas flow through the systems is very small relative to the pressure drop due to the liquid head. At the liquid flow rate of 78 l/hr and the gas flow rate of 20 ft³/hr the pressure drop due to factors other than the liquid head amounts to 0.4%, 4.3%, and 1.1% of the total pressure drop for the bubble column, the sieve tray column, and the Koch mixer column, respectively.

In general the pressure drop of the systems decreases as the gas flow rate increases and it increases as the liquid flow rate increases. At higher gas flow rates, a large volume of the column is occupied by the gas so that

the liquid head decreases. This decrease of the liquid head results in a decrease in the total pressure drop. At higher liquid flow rates, the pressure drop increases because the friction loss is larger and the liquid head is greater. For the range of liquid flow rate investigated, the pressure drop in the bubble column and the Koch mixer column is not as sensitive to the liquid flow rate as that in the sieve tray column.

Oxygen transfer

Figures 9, 10, and 11 show the oxygen transfer coefficient as a function of the gas flow rate with liquid flow rate as the parameter. The oxygen transfer coefficients were calculated based on the liquid phase following the plug flow model. Figures 12, 13, and 14 show the values of the oxygen transfer coefficient as calculated assuming the liquid phase follows the completely mixed model. The actual transfer coefficients should fall within the range of these two extreme cases. In this work, the axial mixing is expected to be greater in the Koch mixer column than in the sieve tray column. The division of the sieve tray column by the horizontally positioned trays provides an additional physical barrier, which reduces the axial mixing.

The oxygen transfer coefficient was found to be a function of both the gas and the liquid flow rates. The values of the transfer coefficient generally increases as the gas and the liquid flow rates increased. However, at higher values of the gas flow rate the slopes of the log-log plot of $K_L a$ vs. gas flow rate decreased. A similar decrease in slope has been observed by Kitai et al. [13].

Two observations should be made with respect to the values of $K_L a$ reported here. First, the accuracy of the estimates for $K_L a$ should improve

as liquid flow rate increases and as gas flow rate decreases because the concentration driving force is larger under these conditions. Second, axial mixing of the liquid phase is probably a function of gas and liquid flow rate as well as motionless mixer design. Axial mixing of the liquid phase is probably greatest in the bubble column and least in the sieve tray column.

The values of $K_L a$ obtained at a liquid flow rate of 78 l/hr in the bubble column were very close to those reported by Yoshida and Akita [7] except that their data were obtained at the condition of no liquid flow. Figure 15 shows such a comparison. Akita and Yoshida [8] reported that the $K_L a$ values were independent of the liquid flow rate up to 160 m/hr; however, our results indicate that $K_L a$ varies with liquid flow rate. Since the estimated values of the oxygen transfer coefficient depend on axial mixing, part of the variation may be the result of more liquid phase axial mixing as liquid flow rate decreases and liquid residence time increases.

The estimated values of the oxygen transfer coefficient in the sieve tray column and the Koch mixer column are compared with those of the bubble column in Figure 16. At low gas flow rates, the values of $K_L a$ for both the sieve tray column and the Koch mixer column are substantially higher than those for the bubble column. A direct comparison of actual values of $K_L a$ is not possible because knowledge of the axial mixing for each system is not available.

Oxygen transfer efficiency in terms of pounds of O_2 transferred/hp hr is plotted against the gas flow rate in Figures 17, 18, and 19 for the bubble column, the sieve tray column and the Koch mixer column, respectively. The results show that there are no significant differences between the sieve tray column and the Koch mixer column with respect to the oxygen transfer

characteristics. Although the $K_L a$ values for the sieve tray column are slightly higher than those for the Koch mixer column, the pressure drop in the sieve tray column is higher, also.

Figure 20 shows a comparison of the oxygen transfer efficiencies for the three column systems at a liquid flow rate of 78 l/hr. At low gas flow rates, the oxygen transfer efficiencies for the sieve tray column and the Koch mixer column are almost twice as large as those of the bubble column. However, at high gas flow rates, the efficiencies are about the same for the three column systems; this is because saturation is approached. Oxygen transfer efficiency depends on the concentration driving force as well as power input, surface renewal, and interfacial area. Since pure oxygen was employed in this study, the values of oxygen transfer efficiency are considerably larger than those summarized elsewhere [2].

Since the oxygen transfer efficiency depends on the concentration driving force, this should be considered in projecting the efficient use of these systems in specific applications. Axial mixing of the liquid phase significantly reduces oxygen transfer efficiency in the experimental cocurrent system investigated in this work; however, in growth processes and the work of Kitai et al. [13] where the dissolved oxygen concentration profile is constant, axial mixing of the liquid phase does not influence oxygen transfer efficiency. If the column with Koch motionless mixers has more axial mixing than the sieve tray column, its oxygen transfer efficiency may be greater than the sieve tray system in other applications where axial mixing does not influence oxygen transfer efficiency.

Conclusions

The results indicate that both the sieve tray column and the Koch mixer column are superior to the bubble column with respect to oxygen transfer, especially at low gas flow rates. The presence of the trays and mixers increases the gas hold up and the oxygen transfer rate.

For all three systems studied, the pressure drop is mainly due to the liquid head. Friction losses due to the gas and liquid flow are very small; they amount to less than 5% of the total pressure loss. The results indicate that friction losses are highest in the sieve tray column and lowest in the bubble column.

This work provides a qualitative comparison of some of the characteristics of the three systems. Other characteristics, such as residence time distributions, need to be investigated further before the systems can be fully understood.

The motionless mixers which were investigated significantly improved oxygen transfer efficiency. Additional research on the use of motionless mixers to achieve efficient oxygen transfer should be carried out in order to fully understand their advantages.

References

1. Chemical Engineering, Nov. 26, 62 (1973)
2. C.L. Cooney and D.I.C. Wang, Biological Waste Treatment, John Wiley & Sons, Inc., 63 (1971)
3. P.H. Calderbank, Biochemical and Biological Engineering Science, Vol. 1, Academic Press, 101 (1967)
4. F.H.H. Valentin, Absorption in Gas-Liquid Dispersions, E. & F. N. Spon Ltd (1967)
5. Chemical Engineer's Handbook, 5th edition, McGraw-Hill, 18-67 (1973)
6. S. Hashimoto, M. Fujita, and M. Fukuchi, J. Ferment. Technol., 51, 58 (1973)
7. F. Yoshida and K. Akita, J. A.I.Ch.E., 11, 9 (1965)
8. K. Akita and F. Yoshida, Ind. Eng. Chem., Process Des. Develop., 12, 76 (1973)
9. K. Akita and F. Yoshida, Ind. Eng. Chem., Process Des. Develop., 13, 84 (1974)
10. W.W. Eckenfelder, J. Sanitary Eng. Div., Proceedings ASCE, 2090, July 1959
11. S. Aiba, A.E. Humphrey, and N.F. Millis, Biochemical Engineering, Academic Press (1965)
12. J.R. Fair, A.J. Lambright, and J.W. Andersen, Ind. Eng. Chem., Process Des. Develop., 1, 33 (1962)
13. A. Kitai, S. Goto, and A. Ozaki, J. Ferment. Technol., 47, 356 (1969)
14. A. Kitai, R. Okamoto, and A. Ozaki, Ferment. Technol. Today, Proc. IVth Intern. Ferment. Symposium, Kyoto, 147 (1972)
15. Koch Static Mixer, Bulletin KSM-2, Koch Engineering Co., Wichita, Kansas

**THIS BOOK
CONTAINS
NUMEROUS PAGES
WITH DIAGRAMS
THAT ARE CROOKED
COMPARED TO THE
REST OF THE
INFORMATION ON
THE PAGE.**

**THIS IS AS
RECEIVED FROM
CUSTOMER.**

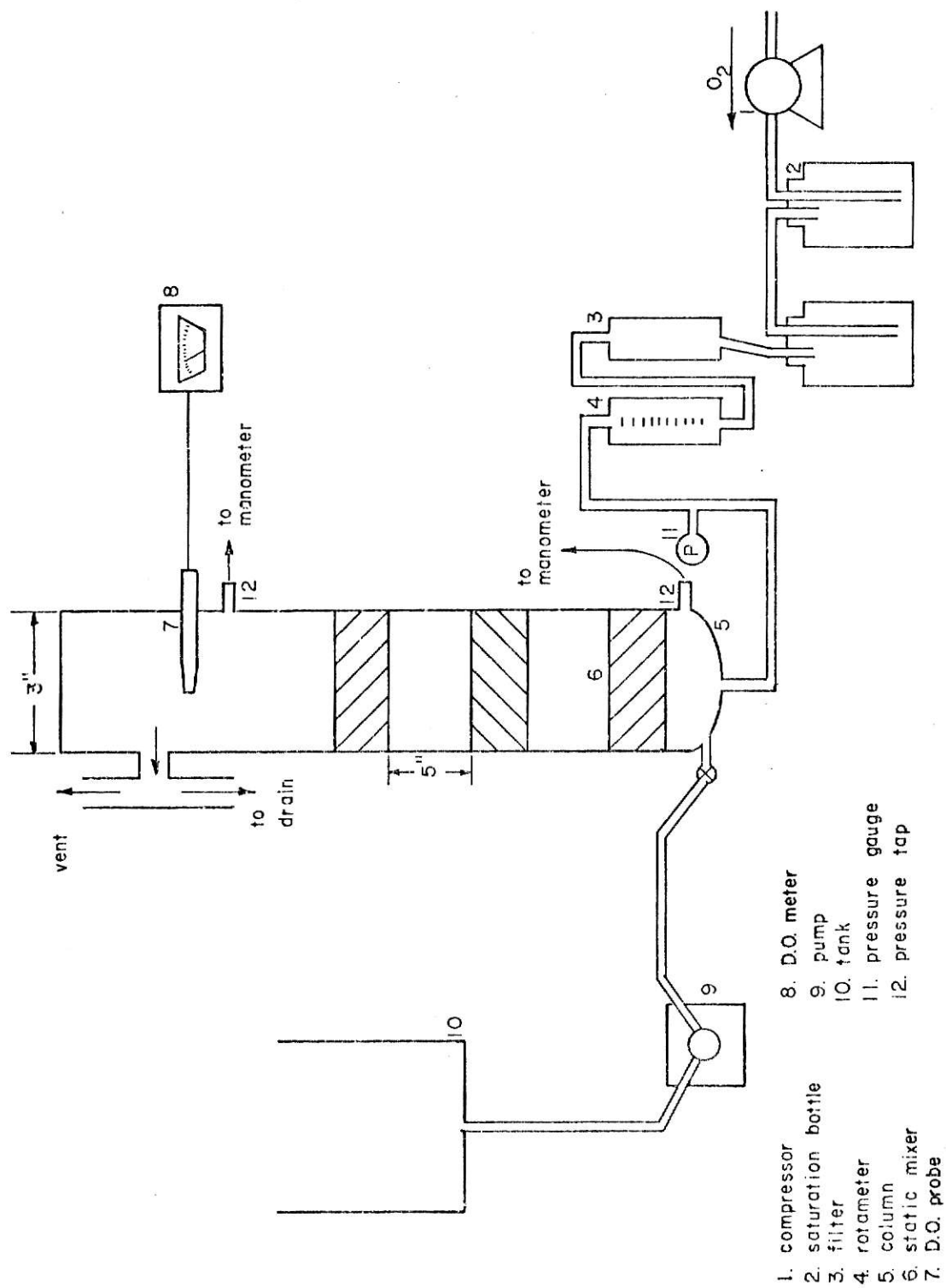


Fig.2a. Schematic Diagram of Koch Mixer Column System.

**THIS BOOK
CONTAINS
NUMEROUS PAGES
WITH ILLEGIBLE
PAGE NUMBERS
THAT ARE CUT OFF,
MISSING OR OF POOR
QUALITY TEXT.**

**THIS IS AS RECEIVED
FROM THE
CUSTOMER.**

**THIS BOOK
CONTAINS SEVERAL
DOCUMENTS THAT
ARE OF POOR
QUALITY DUE TO
BEING A
PHOTOCOPY OF A
PHOTO.**

**THIS IS AS RECEIVED
FROM CUSTOMER.**



Fig. 2b. Koch mixer

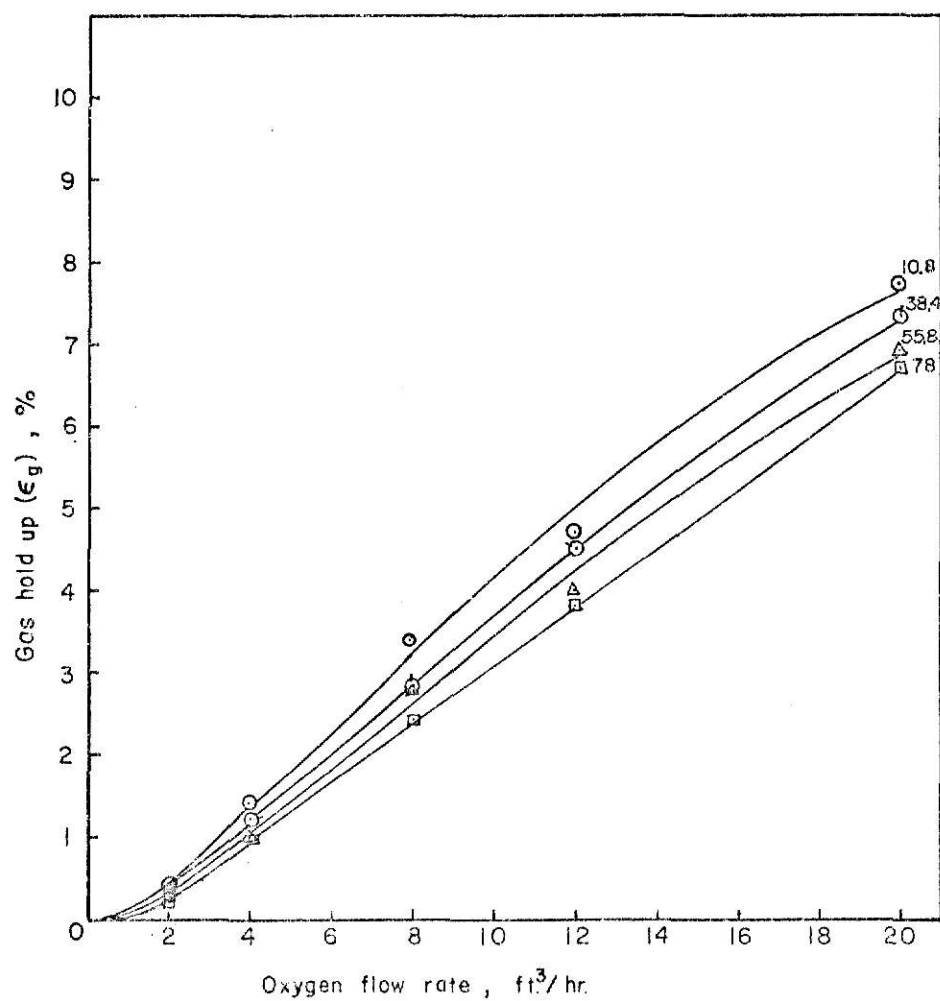


Fig. 3. Gas hold-up in bubble column with liquid flow rate (l./hr.) as parameter.

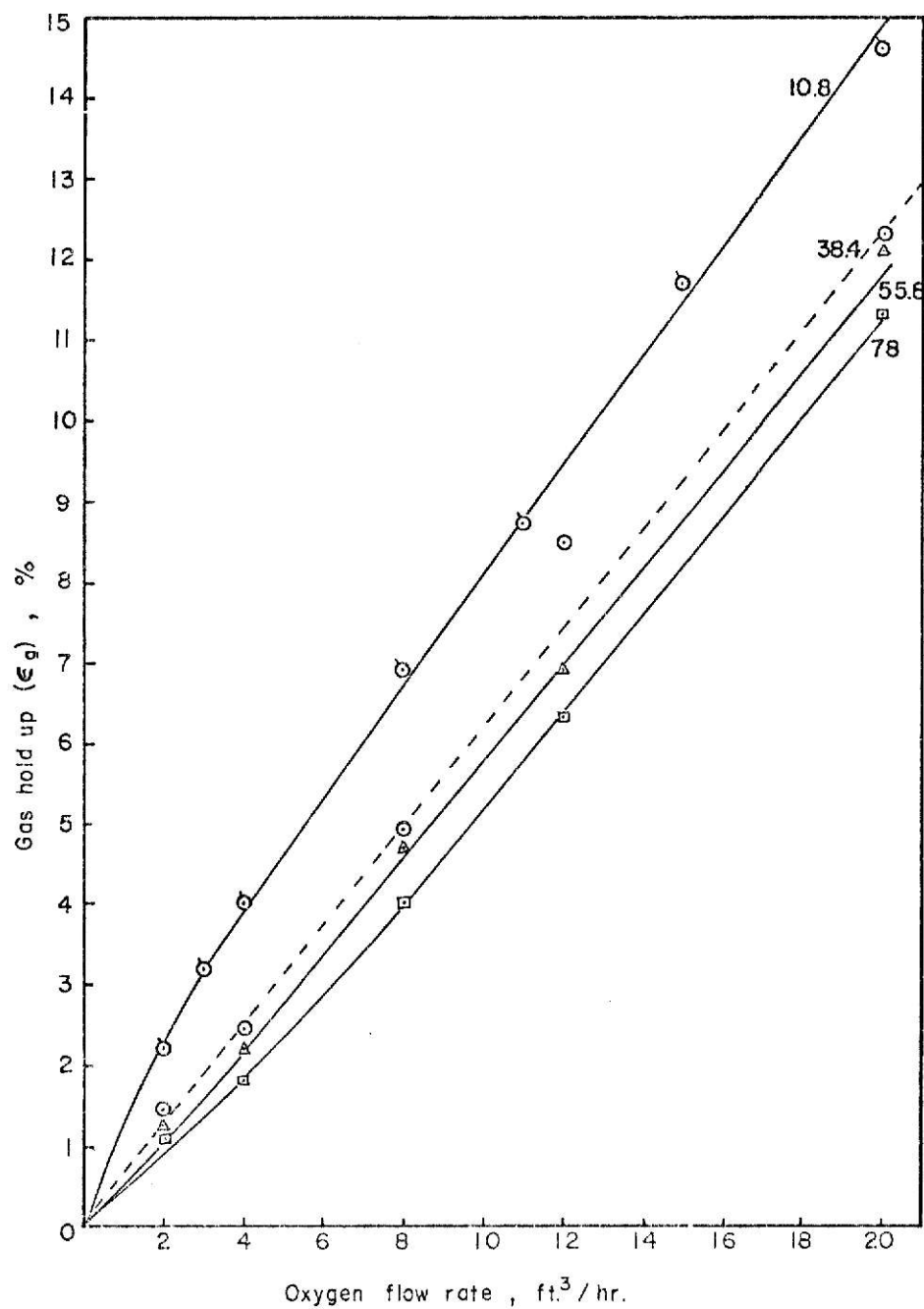


Fig.4. Gas hold up in sieve tray column with liquid flow rate (l./hr.) as parameter.

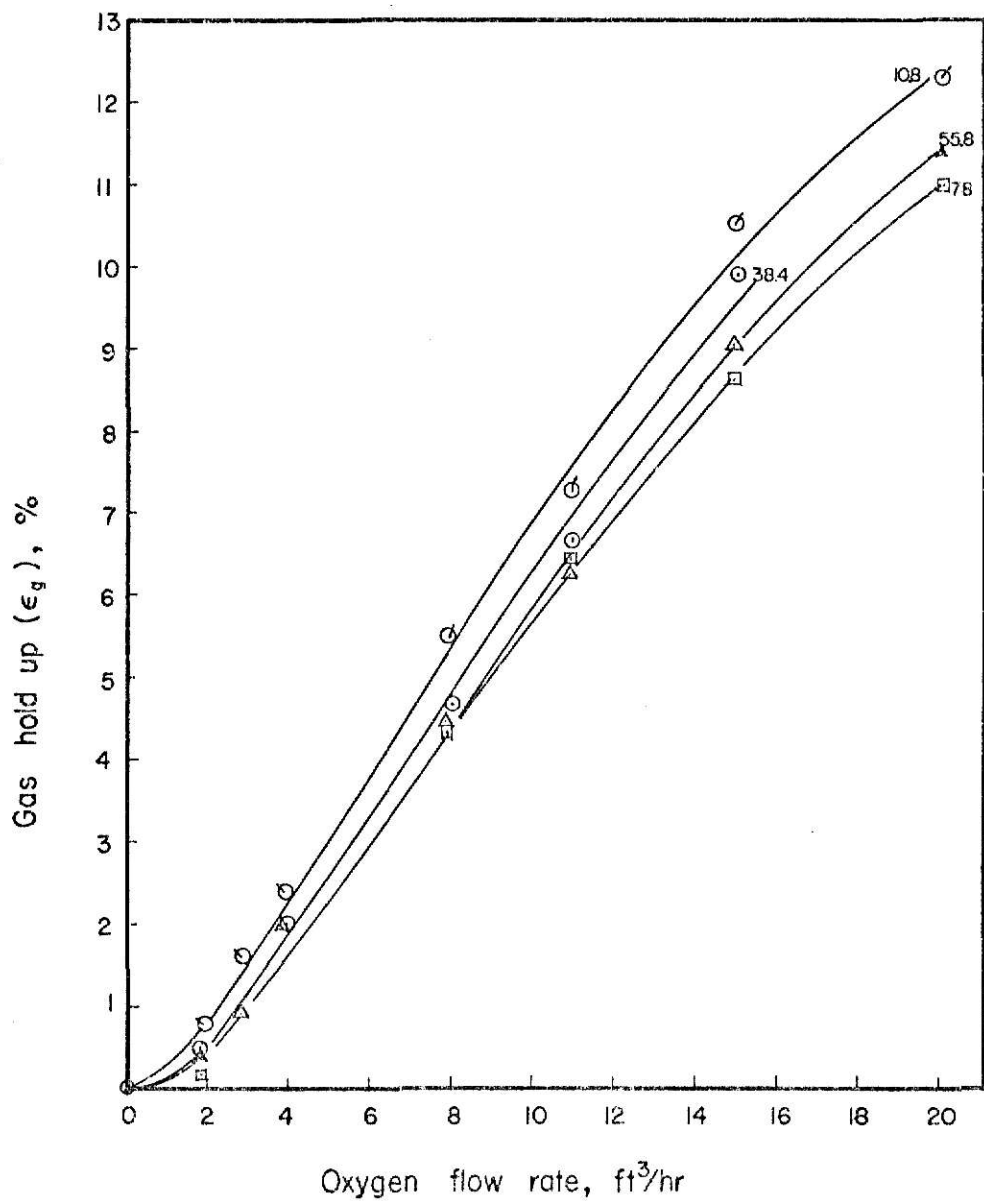


Fig. 5. Gas hold up in Koch mixer column with liquid flow rate (l/hr) as parameter.

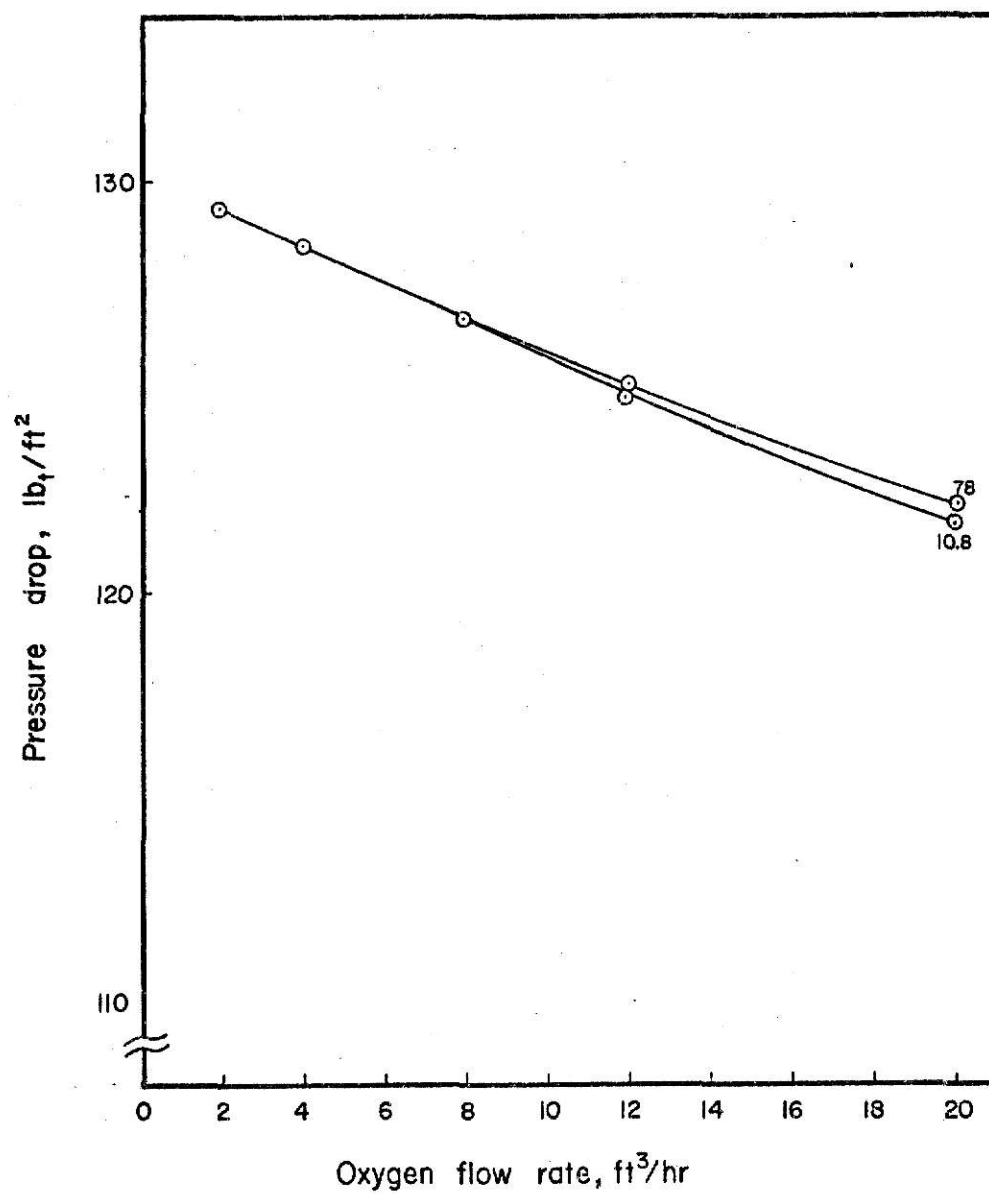


Fig. 6. Pressure drop in bubble column with liquid flow rate (l/hr) as parameter.

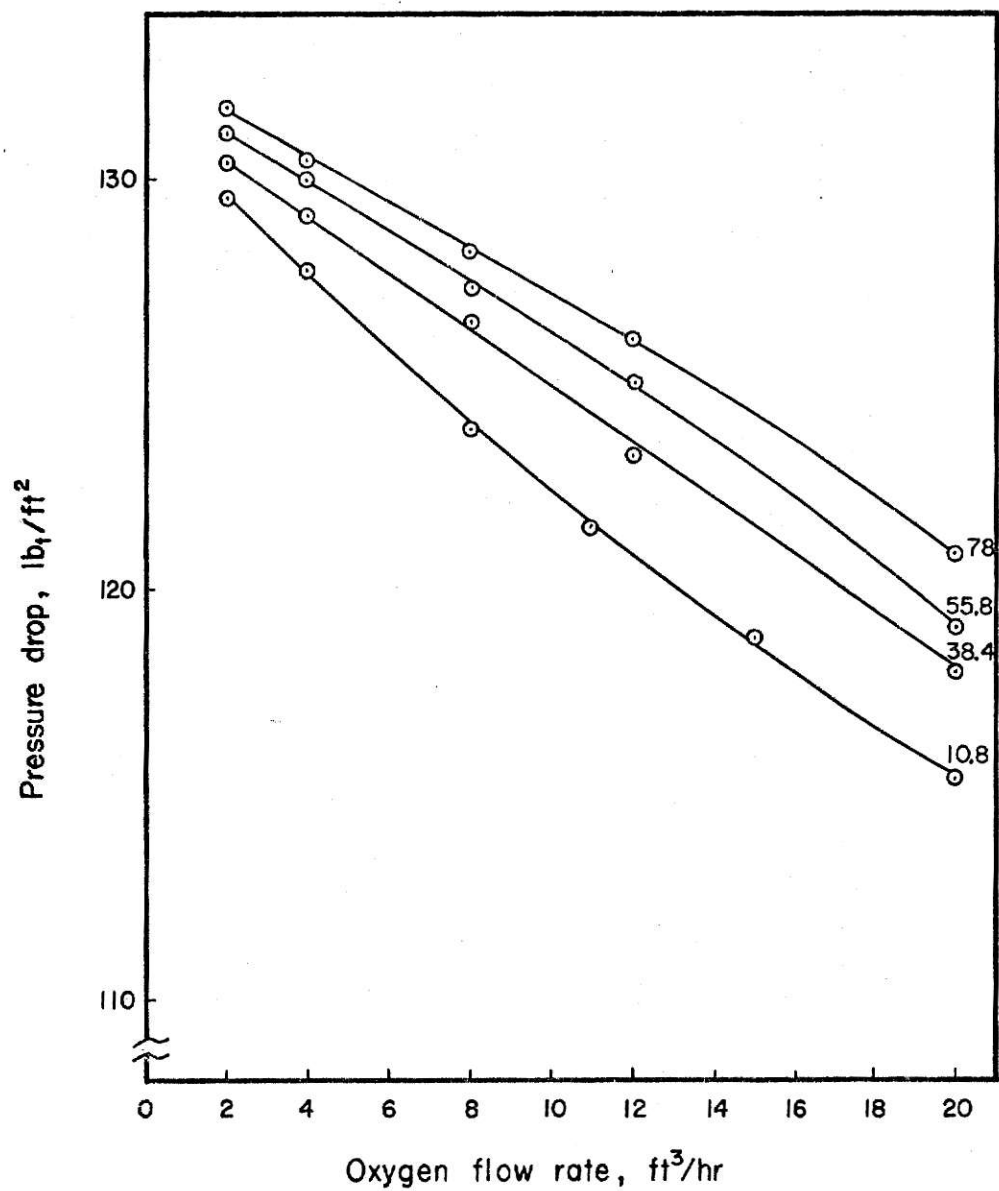


Fig. 7. Pressure drop in sieve tray column with liquid flow rate (l/hr) as parameter.

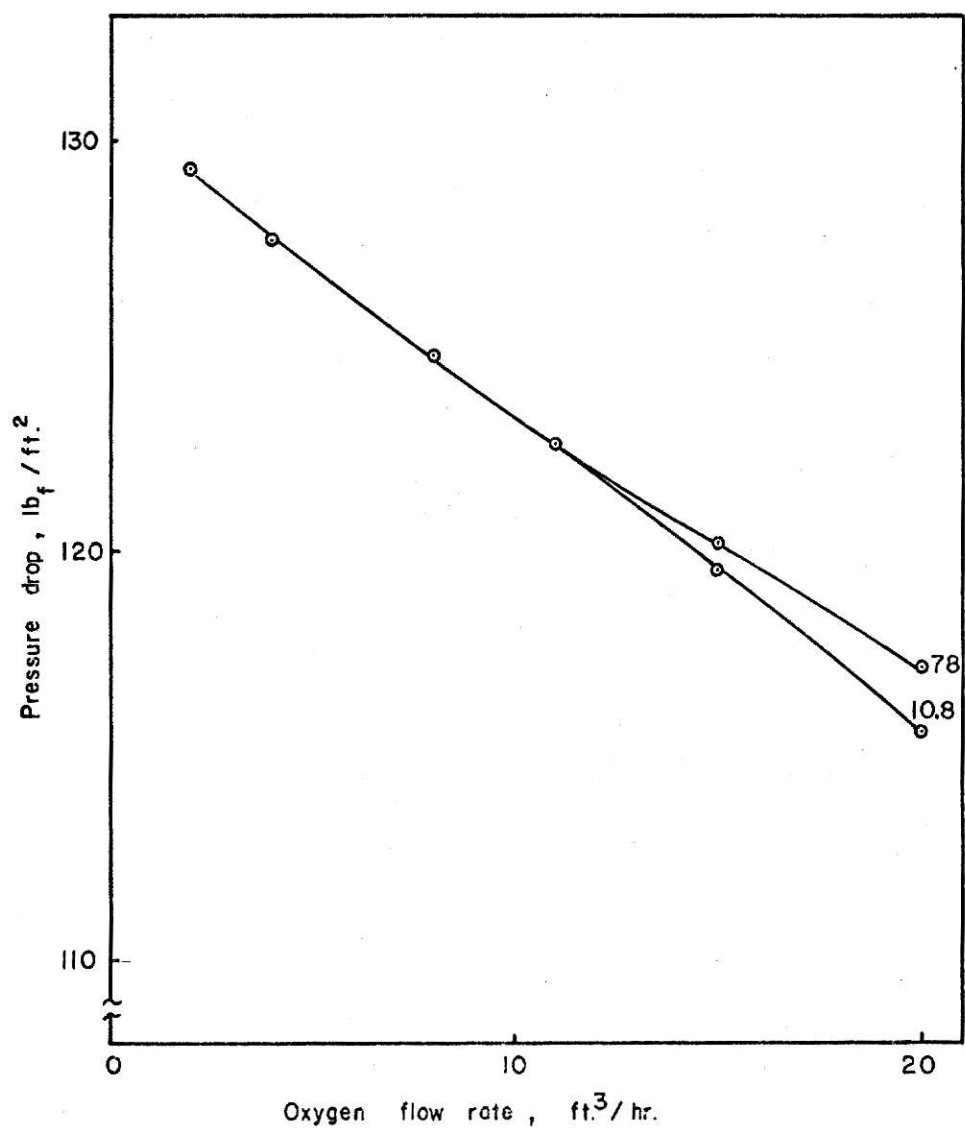


Fig. 8. Pressure drop in Koch mixer column with liquid flow rate (l./hr.) as parameter.

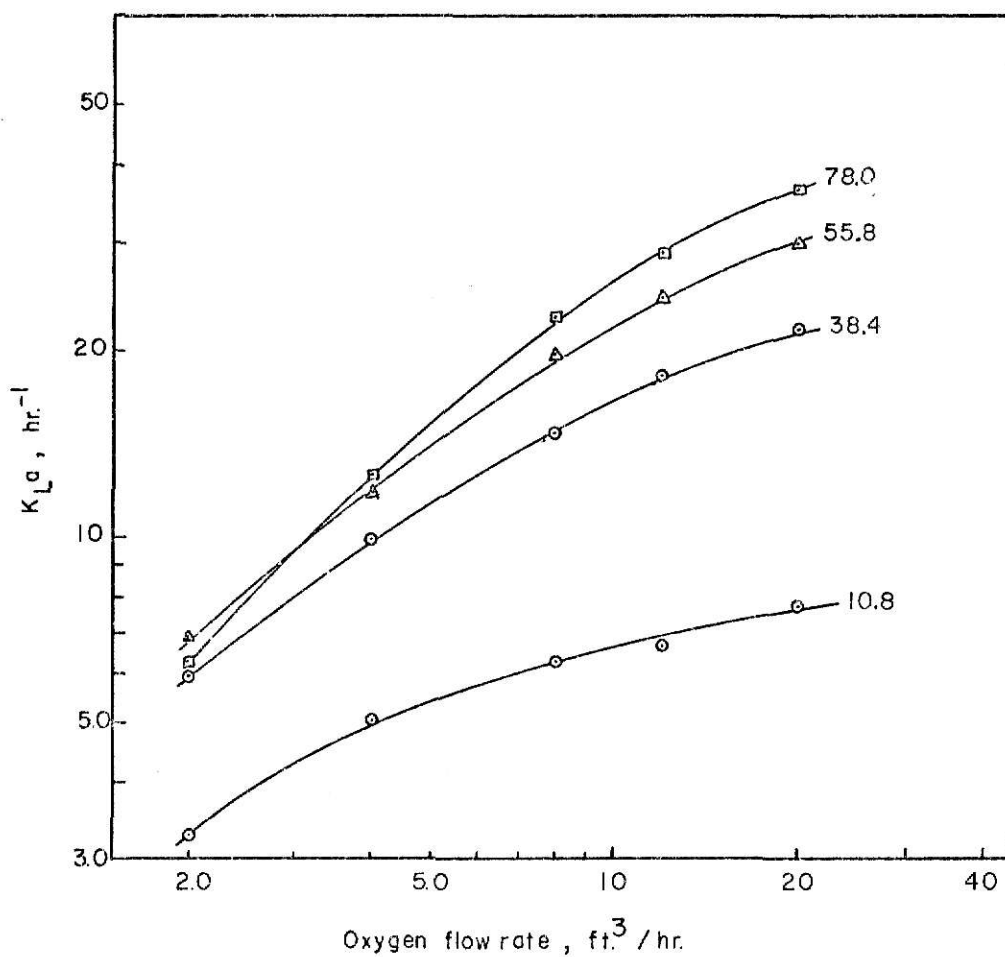


Fig. 9. Bubble column oxygen transfer coefficient based on the plug flow model with liquid flow rate (l./hr.) as parameter.

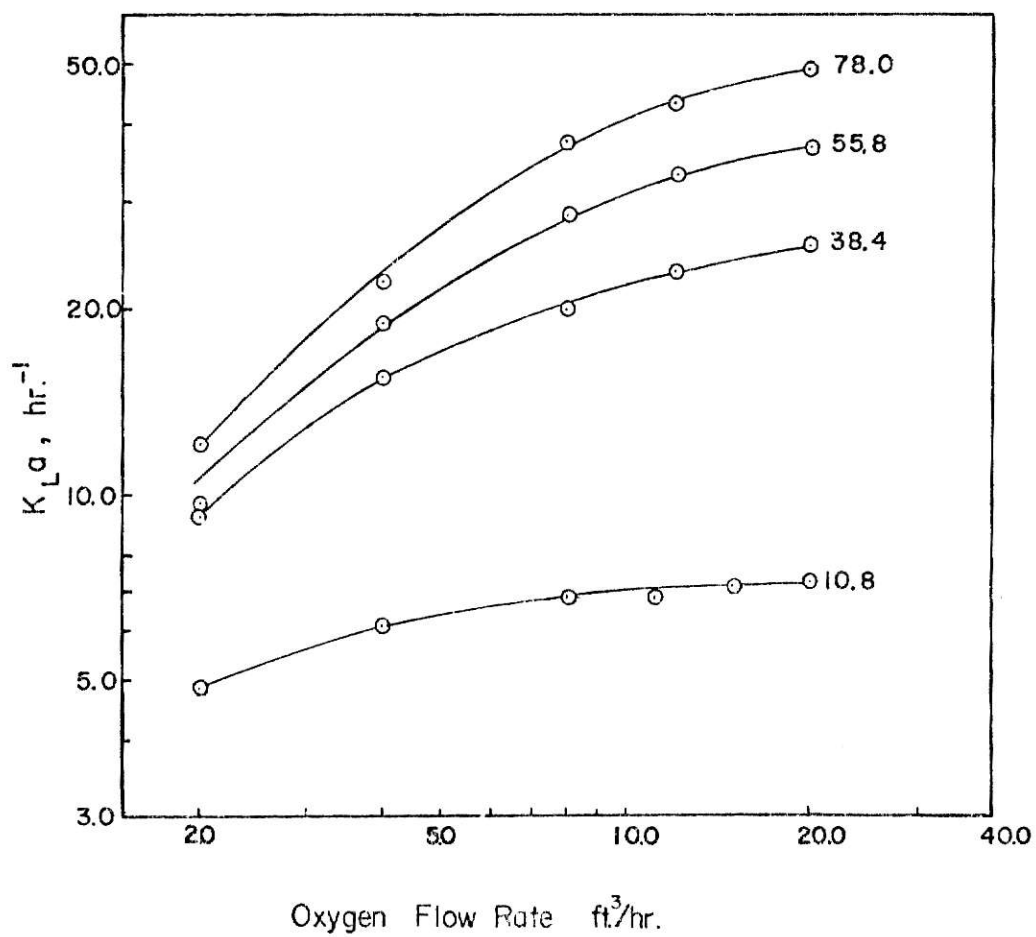


Fig.10. Sieve tray column oxygen transfer coefficient, based on the plug flow model with liquid flow rate (l./hr.) as parameter.

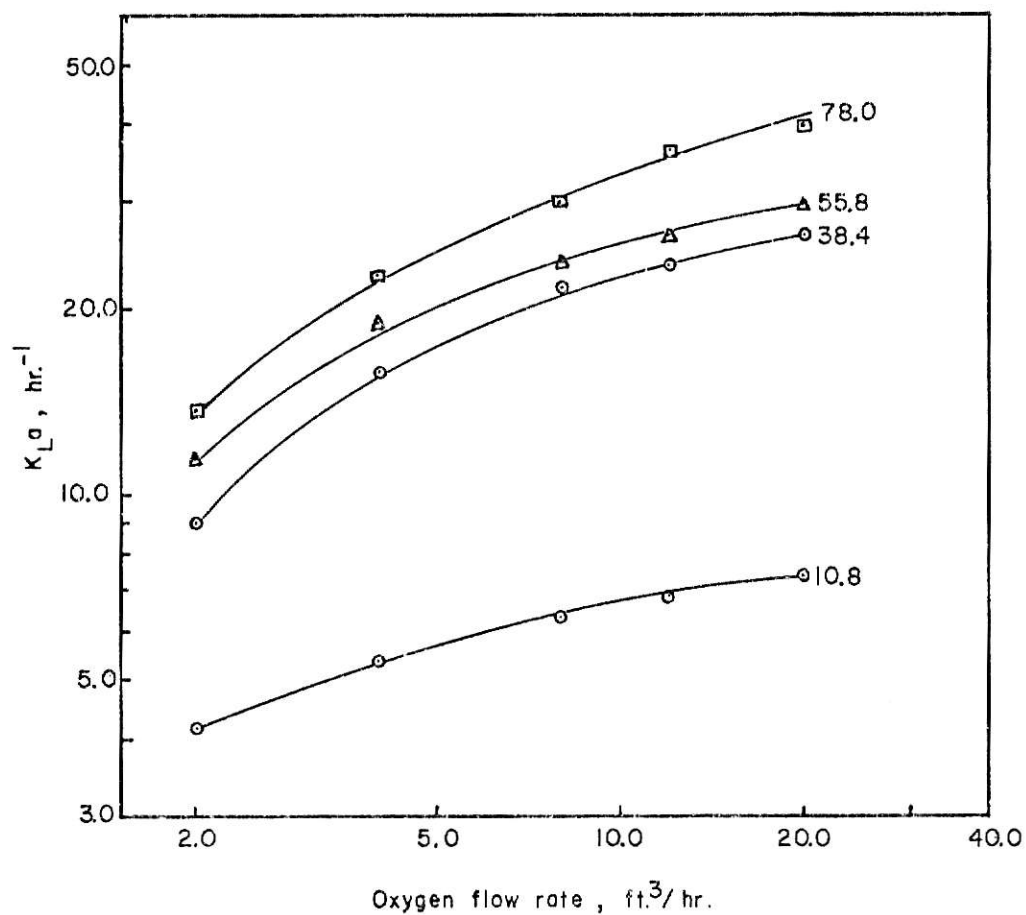


Fig.11. Koch mixer column oxygen transfer coefficient, based on the plug flow model with liquid flow rate (l./hr.) as parameter.

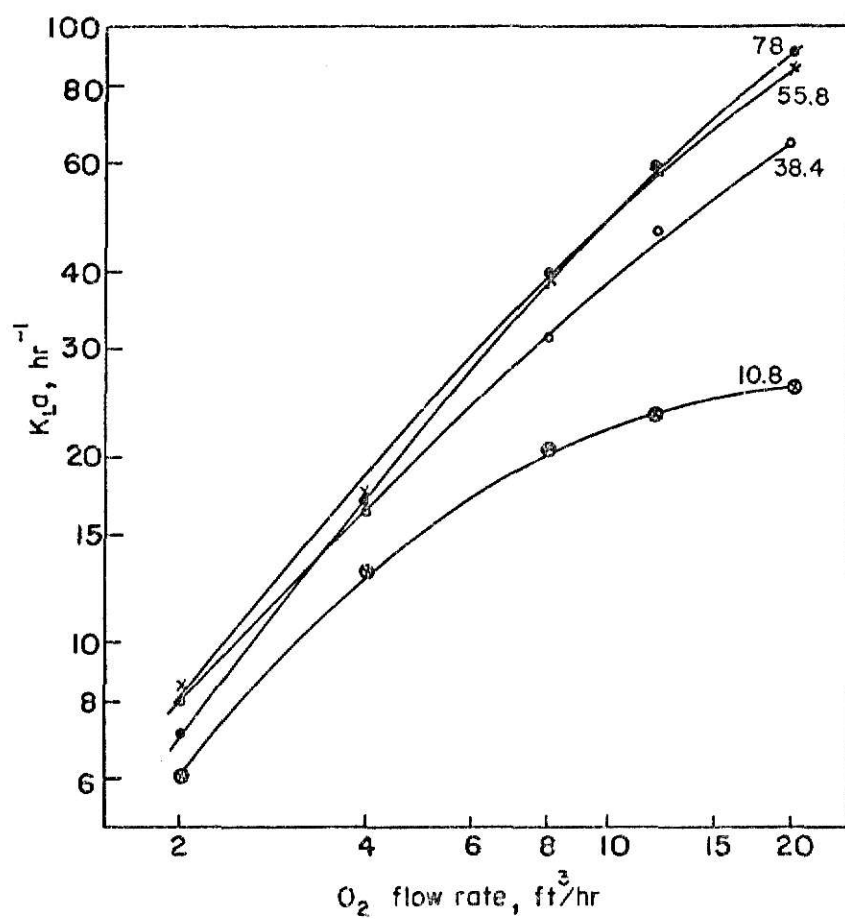


Fig. 12. Bubble column oxygen transfer coefficient based on complete mixed model with liquid flow rate (l./hr.) as parameter.

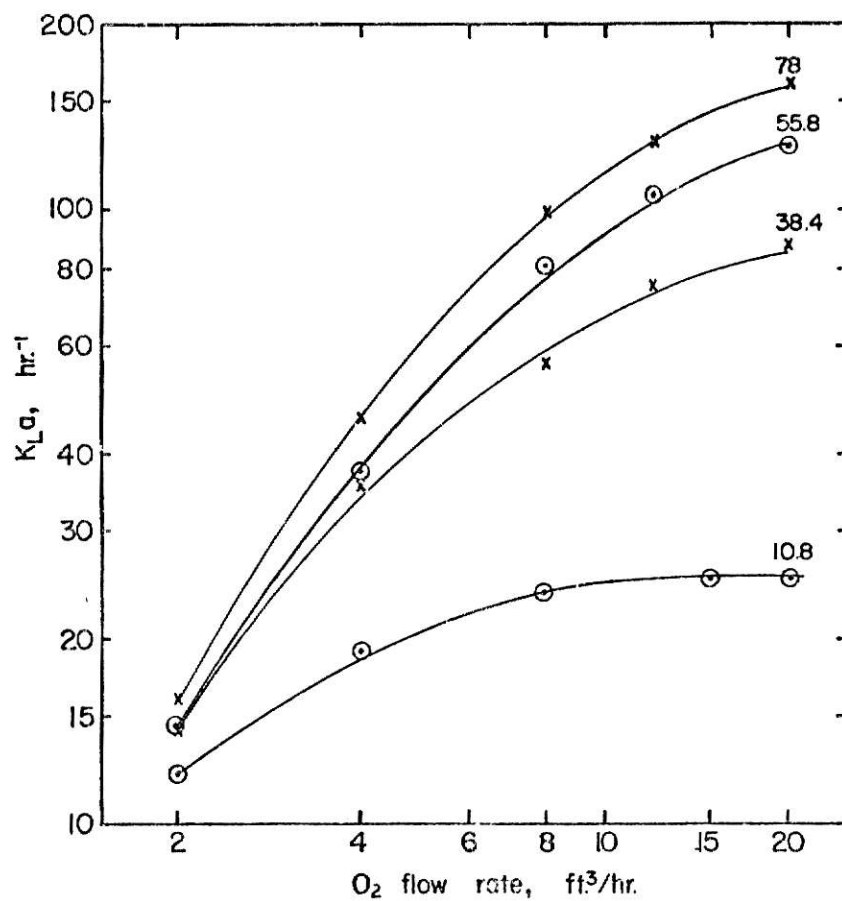


Figure 13. Sieve tray column oxygen transfer coefficients, based on complete mixed model with liquid rate (l./hr.) as parameter.

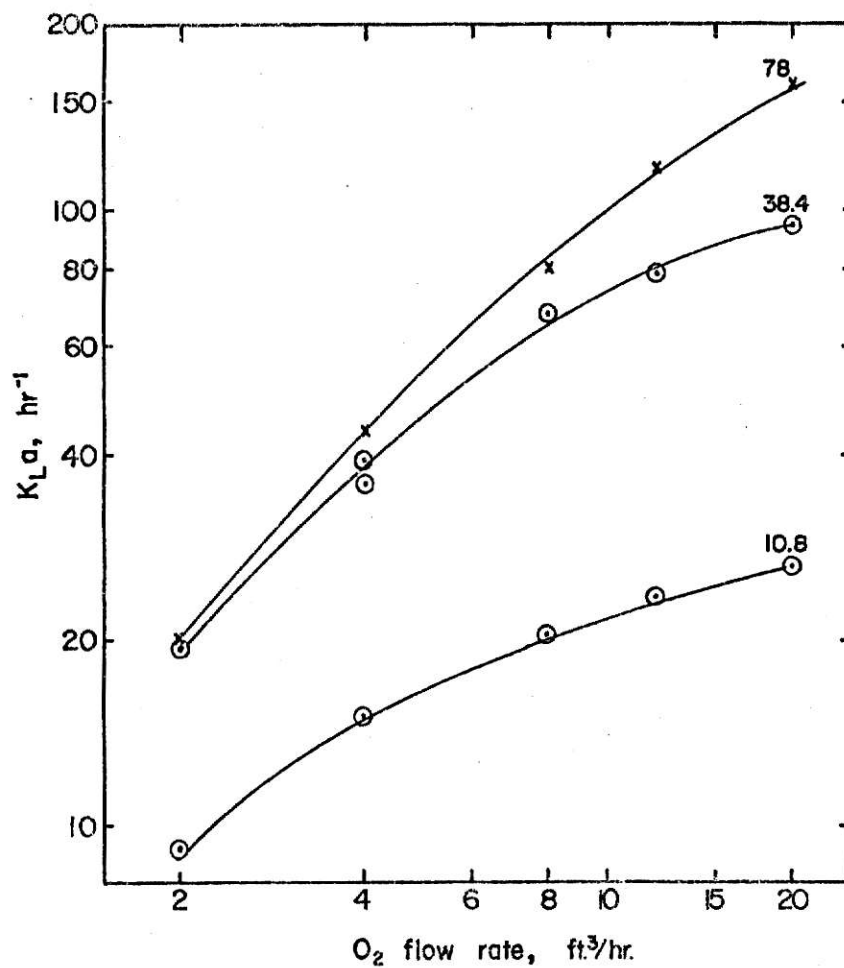


Figure 14. Koch mixer column oxygen transfer coefficients, based on complete mixed model with liquid flow rate (l/hr) as parameter.

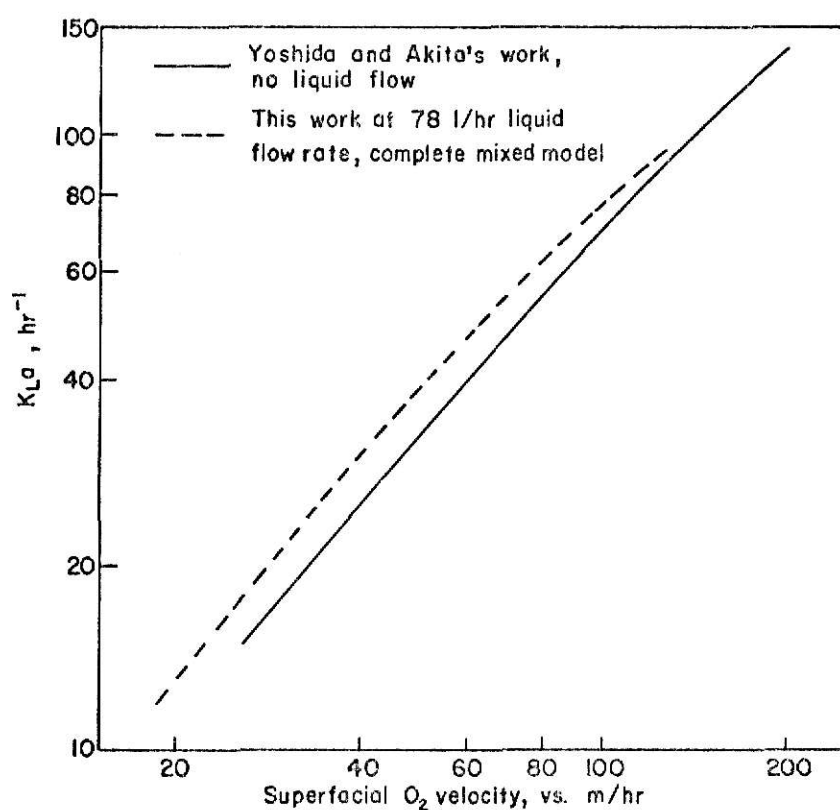


Fig. 15. Comparison of the $K_L a$ values for the bubble column obtained in this work with those reported by Yoshida and Akita [7].

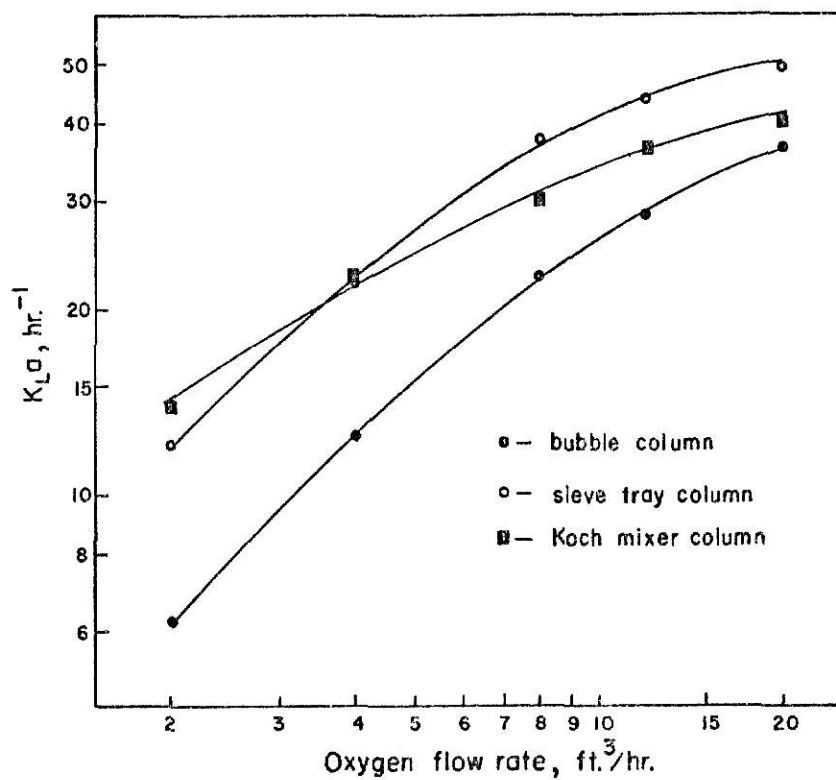


Fig. 16. Oxygen transfer coefficient based on the plug flow model at a liquid flow rate of 78 l/hr.

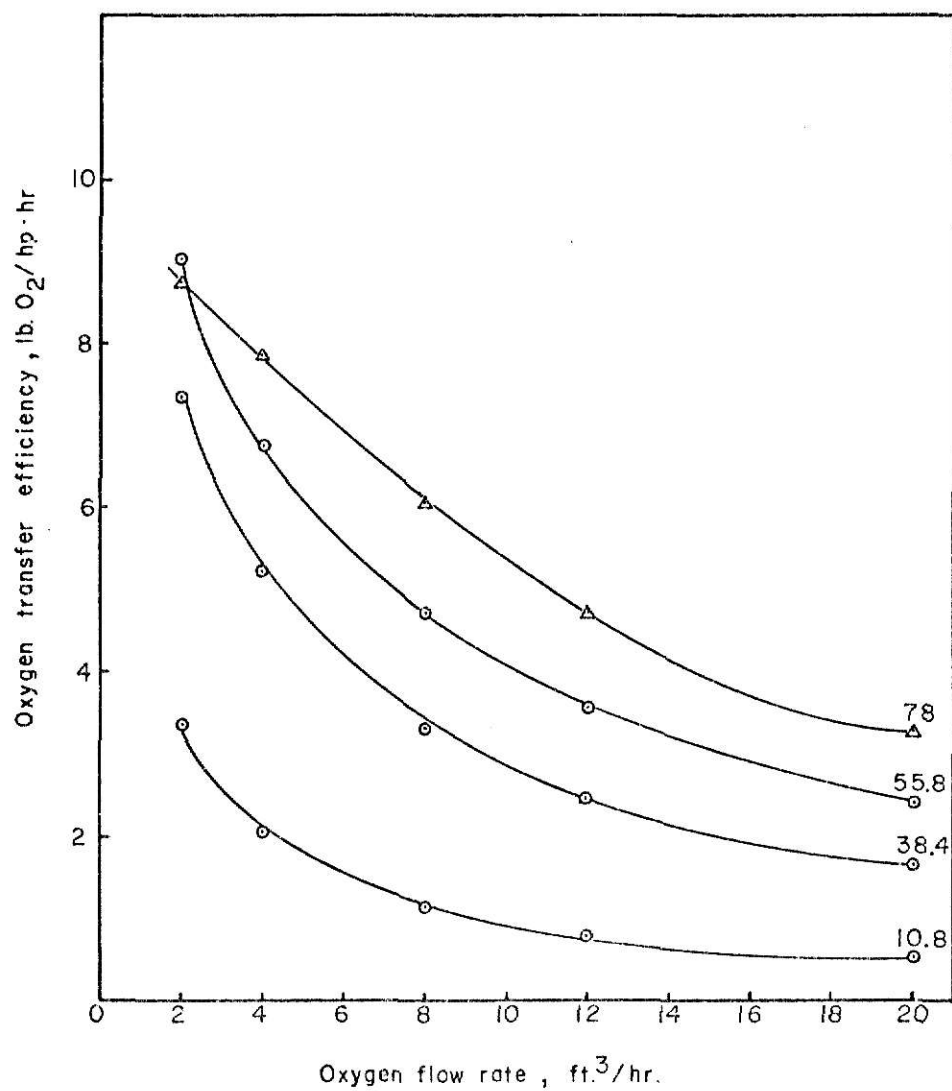


Fig.17. Oxygen transfer efficiency in the bubble column with liquid flow rate (l./hr.) as parameter.

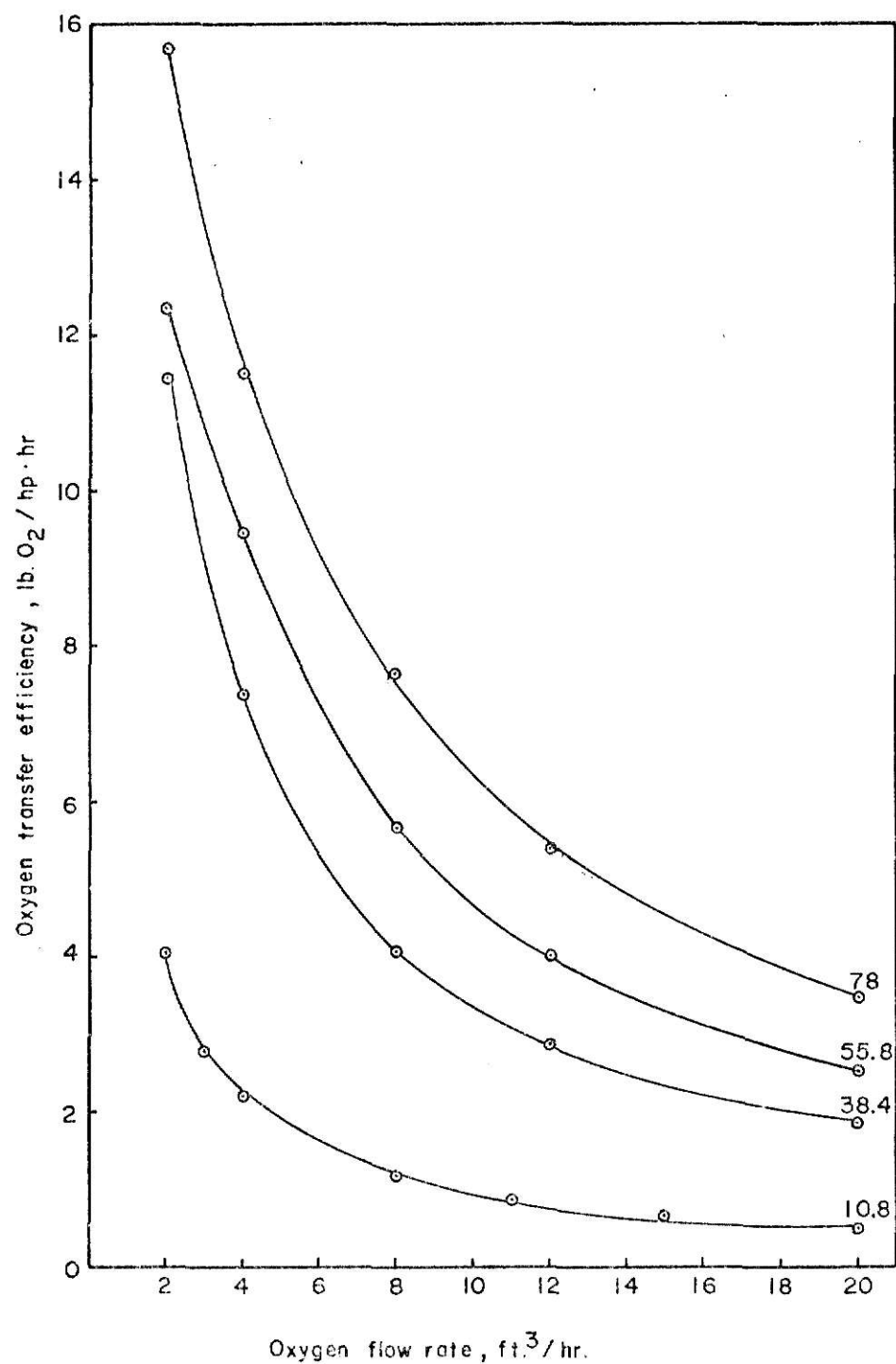


Fig. 18. Oxygen transfer efficiency in the sieve tray column with liquid flow rate (l./hr.) as parameter.

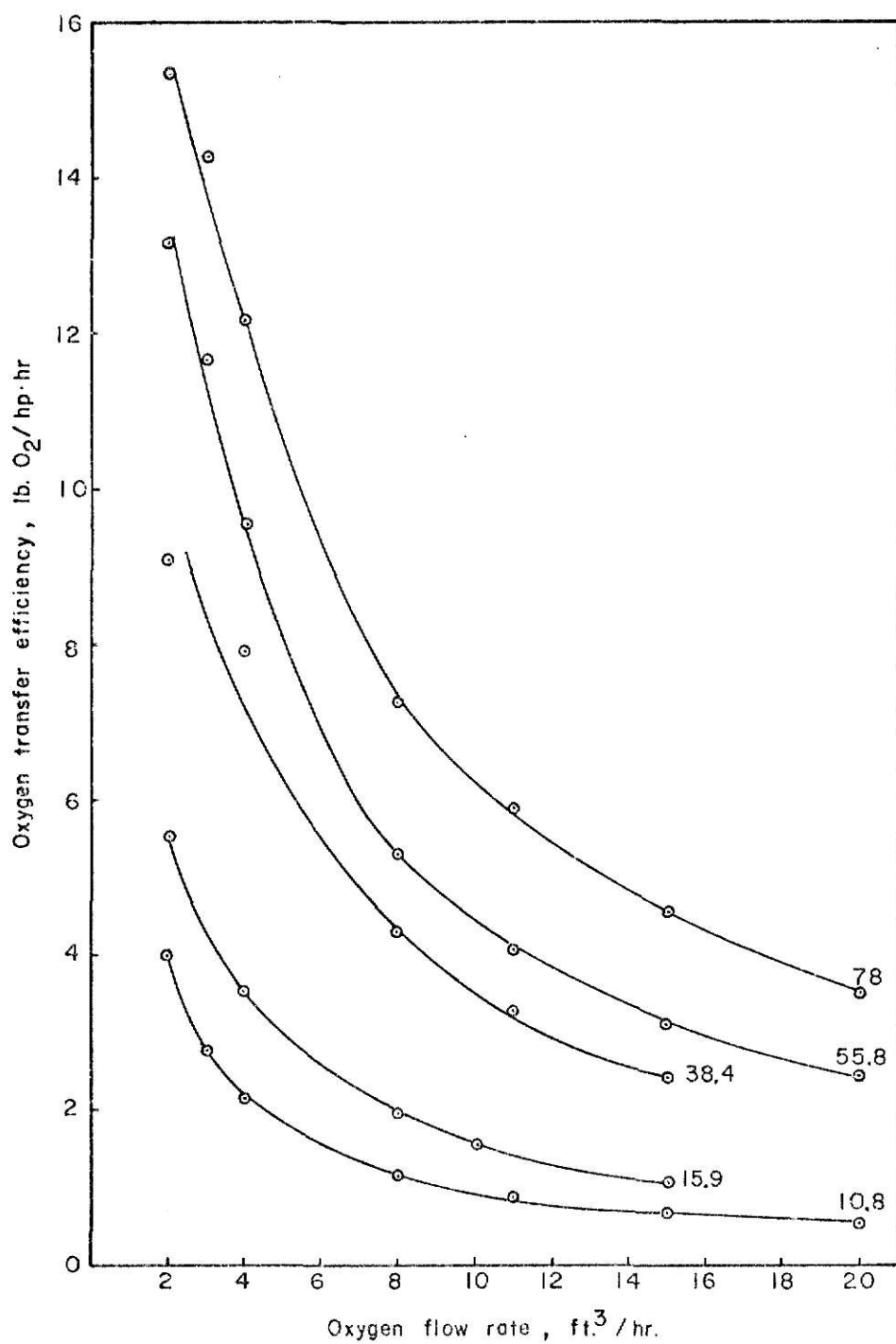


Fig.19. Oxygen transfer efficiency in the Koch mixer column with liquid flow rate (l./hr.) as parameter.

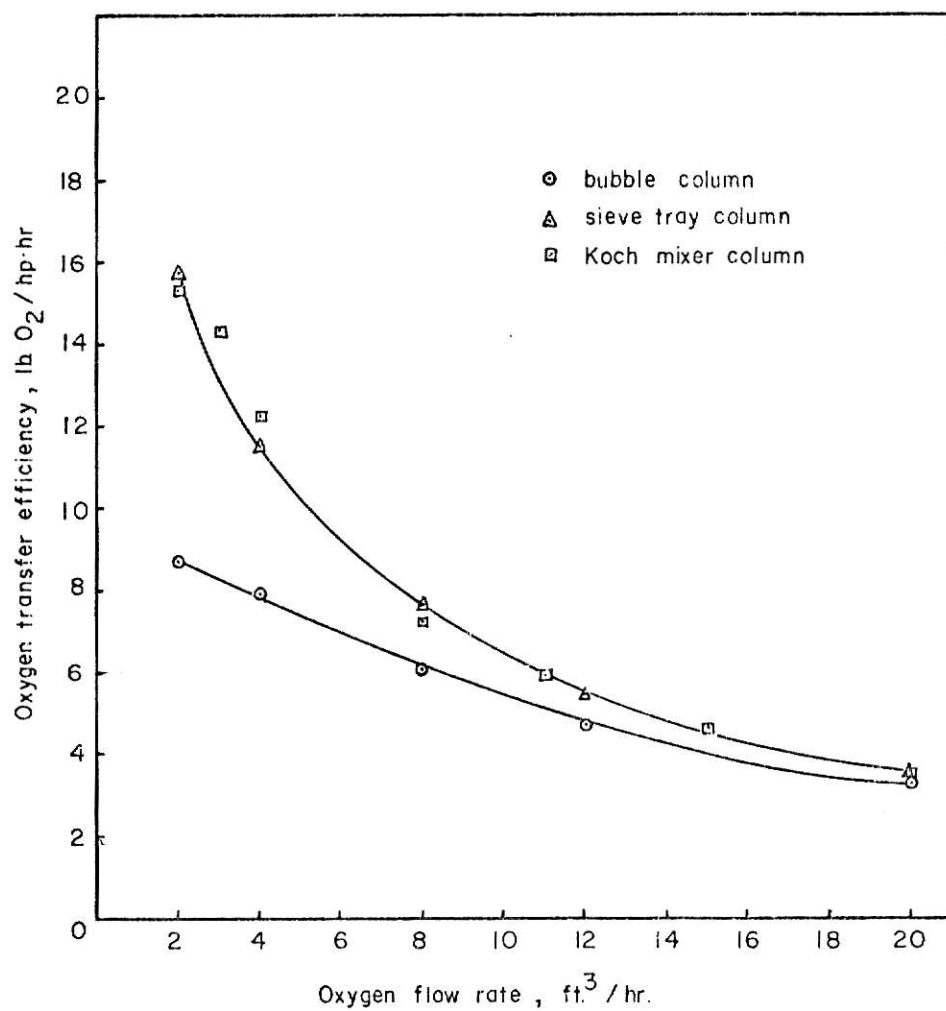


Fig.20. Comparison of oxygen transfer efficiencies of the three systems at a liquid flow rate of 78 l./hr.

CHAPTER III

BIOLOGICAL WASTE TREATMENT IN TOWER SYSTEMS

Introduction

Continuous biochemical reactor systems are becoming more popular as new processes are developed for single cell protein production, fermentation, and waste treatment. The continuous stirred tank reactor (CSTR) is the most common reactor for these applications. Many of the important biological growth processes are aerobic and large volumes of air or oxygen are required. The difficulties in maintaining adequate aeration and the long residence times make the design of plug flow systems difficult.

Tower type reactors with mixing devices designed to promote mass transfer are able to provide the required aeration for biological processes. The utilization of these tower type reactors for biological processes offers an alternative reactor design with liquid flow characteristics intermediate between the ideal backmix and plug flow models.

Several workers [1-9] have investigated the performance of multi-stage tower fermentors. In 1969, Kitai et al. [1-4] reported on the continuous cultivation of *E. Coli* using a multi-stage vertical column with horizontally positioned sieve trays. Growth differentiation, washout, flow characteristics, and oxygen transfer were studied under the condition of cocurrent flow of air and liquid phases. They found that the liquid flow characteristics were approximately those of the completely mixed tanks in series model, if trays with hole diameters of 2 mm were used. Cell and substrate concentration differences were observed to occur from the bottom to the top in the flow direction. They also found that with the help of the sieve trays,

oxygen transfer coefficients comparable to those of agitated vessels could be obtained.

Prokop et al. [5] studied the continuous fermentation of baker's yeast in an eight stage tower fermentor similar in design to the one used by Kitai et al. Under cocurrent flow conditions, they found that the sedimentation of yeast cells played an important role in the performance of the tower fermentor. Unlike the results of Kitai et al., they found that the yeast cell concentration decreased in the direction of flow. By conducting a residence time distribution (RTD) study under actual fermentation conditions, they found that substantially amounts of mixing among the stages existed, and that the liquid flow characteristics approached those of the CSTR model if sieve trays of 15% open area were used. To reduce the back mixing, the open area on the sieve trays was reduced to about 10% of that for the original design. Separate RTD results for the cell suspension and the liquid media were presented.

Falch and Gaden [6,7] developed a multi-stage tower fermentor equipped with downcomers and mechanical agitators. Fermentation of *E. Coli* was carried out counter currently with air being introduced from the bottom and substrate from the top of the column. Back flowrates up to 120% of the net flow rate to the column were measured. The back flow was found to be strongly dependent upon the agitator speed, and the size and position of the impellers. The cell concentration was observed to increase in the liquid phase flow direction. The effects of dilution rate on the concentration levels of cells and substrate were also studied.

The early applications of the tower system were limited to fermentation of pure cultures. It was not until later that the application to waste

treatment was investigated. In 1971, Lee et al. [8] reported on the modeling and optimization of a tower type activated sludge system. A CSTR in series model was employed along with cell sedimentation and back flow. The study showed that the performance of the tower system was affected by back flow and cell sedimentation.

Besik [9] in 1973, reported on a pilot plant study with a multi-stage tower activated sludge process. The pilot system consisted of an activated carbon bed, a counter current sieve tray tower with downcomers for aeration, a clarifier with contained flocculation chamber, and two media filters. Some of the aerated culture leaving the tower was recirculated through the bed of activated carbon while the remainder flowed to the clarifier. The sludge from the clarifier was recycled to the activated carbon bed. The capacity was 15,000 GPD. The system was found to have a BOD removal efficiency of 98%. Other operation data were compared with various activated sludge processes. The results indicated that the multi-stage tower type activated sludge process was more efficient than any other known activated sludge process.

Sonoda et al. [10] reported on the application of a bubble column and a 6-stage perforated plate column for activated sludge treatment. The experiments were carried out at low dilution rates ($0.05-0.2 \text{ hr}^{-1}$). They found that at $D = 0.15 \text{ hr}^{-1}$ and $D = 0.2 \text{ hr}^{-1}$, the sludge in the bubble column caused bulking, while sludge from the 6-stage perforated plate column had good settling properties.

In this study, a multi-stage tower fermentor with sieve trays and a Koch mixer column are investigated for their applications in waste water treatment. Results from steady state and transient experiments are presented

and discussed. Also, mathematical models are used to simulate some of the results which were observed experimentally.

Materials and Methods

Equipment

The sieve tray column system consisted of a 3.664 liter bubble column divided into 4 compartments by horizontally positioned sieve trays. A schematic diagram of the system is shown in Figure 1. Each sieve tray was made of a 1/16 inch thick brass plate with hole diameter of 3/64 inch and free open area to column cross sectional area ratio of 3%.

The system was operated cocurrently with liquid being pumped into the column by means of a sigma motor pump (type AL-4-E-40, by Sigma Motor, Inc.). Air was pumped by an air compressor through the humidifiers, an air filter, and a flow meter to the nozzle at the bottom of the column. A pressure gauge was installed between the flow meter and the inlet nozzle to the column to detect any pressure build up.

The equipment arrangement for the Koch mixer column system was similar to that of the sieve tray column system. Instead of using sieve trays, the Koch mixers were used to distribute the air. A schematic diagram of the system is shown in Figure 2a. The mixers were made of stainless steel sheets bent into wave shape and put together with the layers parallel to each other and the corrugation angle of the adjacent layers reversed with respect to the mixing axis. The mixers used in this study were of the BY type [11]; they were 3 inches tall and 3 inches in diameter with a layer height of 1/4 inch (see Figure 2b). This system was also operated cocurrently.

Synthetic waste

The composition of the glucose-limiting media contained, per liter:

Glucose	1000 mg
$(\text{NH}_4)_2\text{SO}_4$	500 mg
$\text{MgSO}_4 \cdot 7\text{H}_2\text{O}$	100 mg
$\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$	0.5 mg
CaCl_2	7.5 mg
$\text{MnSO}_4 \cdot \text{H}_2\text{O}$	10 mg
KH_2PO_4	527 mg
K_2HPO_4	1070 mg

Organisms

Seed organisms were obtained from the municipal sewage treatment plant at Manhattan, Kansas. Two hundred milliliters of primary clarifier effluent were added to 800 ml of a 3 fold concentration of medium and placed into a glass jar and aerated. Each day, 80% of the culture solution was discarded and replaced by an equal amount of fresh medium. After 3 days, the stock was used as inoculum to the continuous flow system.

Analytical procedures

Ten-milliliter samples were collected at each of the sampling ports and at the exit of the column. In order to obtain a more representative sample, the first 10 ml of the solution withdrawn from each sampling port was discarded and the next 10 ml was used to analyze for cell and substrate concentrations.

Cell concentration in terms of mixed liquor suspended solids (MLSS) was determined by filtering a 10 ml sample through 0.45 μ pore size Millipore

membrane; the nonvolatile solids retained were washed and dried to a constant weight at 105°C. Substrate concentration in terms of dissolved chemical oxygen demand (COD) was determined by employing the dichromate method [12] to analyze the filtrate collected from the filtration of the sample.

Dissolved oxygen was frequently measured at the top of the column with a Galvanic Cell Oxygen Analyzer (Precision Scientific Co.). The air flow rate for all the experimental runs was set at 14 ft³/hr, which was approximately 2 vvm. At this air flow rate the dissolved oxygen concentration of the media was maintained at a range from 7 mg/l to 7.5 mg/l; therefore, sufficient oxygen was present during the treatment.

The pH was measured by using an Accumet pH meter (Fisher Scientific Co.). Throughout the experiment pH was quite constant at a range of 6.8 ± 0.2 , and no control was needed.

Results and Discussion

Visual observations during operation

In the sieve tray column, under the condition of cocurrent flow of liquid and gas, both phases existed as continuous and dispersed phases in each of the compartments divided by the sieve trays. Underneath each sieve tray, the gas phase was continuous, and froth and foam were present at the surface of the liquid phase. The gas phase hold up beneath the sieve trays was strongly dependent upon the gas flow rates. In general, the higher the gas flow rate the larger the space occupied by the froth. At a constant gas flow rate the hold up does not necessarily stay constant, because physical properties of the medium, such as surface tension, also affect cocurrent two

phase flow behavior. Kitai et al. [1,2] has reported similar observations for the fermentation of E. Coli in the same type of column.

The air bubbles inside the compartments varied in size. By visual observation most bubbles ranged from about 1/16" to 3/8" in diameter. The medium in the compartments was vigorously mixed by the bubbling action when the gas flow rates were 14 ft³/hr (approximately 2 vvm) or higher. Complete mixing of the liquid phase within each of the compartments may be safely assumed in this case.

Usually after about 15 hours of continuous operation, a steady state was reached. This period varied depending upon the system dilution rate at which the system was operated; a lower dilution rate usually took longer.

Wall growth inside the compartments became significant usually after 15 hours to 1 week of continuous operation, depending upon the dilution rate. For instance, the wall growth was clearly observed after 15 hours of continuous operation at a system dilution rate close to the washout dilution rate (0.303 hr⁻¹), while it took at least 5 days to observe the same amount of wall growth at the dilution rate of 0.1 hr⁻¹. Wall growth was always observed first in the lowest compartment of the sieve tray column.

For the Koch mixer column, under the same flow conditions as that of the sieve tray column, continuity in the liquid phase was observed. There was no froth section underneath any of the Koch mixers. The only place that foaming occurred was at the region close to the exit of the column. Good mixing was observed everywhere above the lowest positioned mixer. Bubble sizes were very uniform in the Koch mixer column; they were approximately 1/4" in diameter. By visual observation, the space occupied by the gas bubbles appeared to be larger at the top section of the column than at the bottom.

Wall growth inside the Koch mixer column was similar to that of the sieve tray column, except that the organisms attached not only to the column wall but also to the surfaces of the mixers.

COD and MLSS concentrations

At the lower dilution rates, steady state measurements could be made for several days under conditions in which wall growth and cell sedimentation were not visually detected. The results presented in this section are the averages of four consecutive daily measurements. At higher dilution rates, it was not possible to obtain a steady state for four days in which wall growth was not visually detected. Figure 3 shows the steady state MLSS concentrations at each stage in the sieve tray column with dilution rate as parameter. The fluctuations of the MLSS concentration between stages may be partly due to the fact that a mixed culture of organisms was present. Another possibility may be the change in flow pattern among the stages. Except for the lowest system dilution rate ($D = 0.046 \text{ hr}^{-1}$), the MLSS concentrations were in the range of 500-600 mg/l for all stages. The low MLSS concentration level at the lowest dilution rate was probably due to endogenous metabolism. Chiu et al. [13] observed the same phenomena for a continuous activated sludge process in a completely mixed tank system.

For the sieve tray column system at dilution rates lower than the washout dilution rate, the dissolved COD concentrations for all stages were between 0 and 35 mg/l.

The steady state MLSS profiles for the Koch mixer column system are shown in Figure 4 with dilution rate as parameter. The MLSS concentrations for these dilution rates were quite evenly distributed along the column length. The MLSS concentrations inside the Koch mixer column were somewhat

higher than those of the sieve tray column; they ranged from 600 to 700 mg/l.

COD removal for the Koch mixer column was as efficient as the sieve tray column. For the experimental runs with dilution rates as indicated in Figure 4, the dissolved COD concentrations were between 0 to 40 mg/l. No COD concentration differentiation could be measured throughout the column at these dilution rates.

The results in Figures 3 and 4 were collected when wall growth and cell sedimentation were not visually apparent. Continued operation for extended periods, however, frequently led to visually observable wall growth.

Wall growth, sedimentation, and plugging

Some of the organisms had the tendency to adsorb to surfaces and to grow on surfaces. As mentioned before, wall growth frequently became significant during continuous operation. Wall growth usually became visible sooner, at higher dilution rates, especially at dilution rates close to the washout dilution rate. This suggested that the adsorbed organisms can compete for substrate better at higher dilution rates where the substrate concentration is higher. At low dilution rates the substrate concentration may be so low that the adsorbed organisms are washed from the wall as fast as they are produced. If the adsorbed species of organisms have a significantly higher saturation constant, K_s , than the freely suspended organisms, which dominate at low dilution rates [13], very little wall growth would be expected at low dilution rates.

Extensive wall growth usually also resulted in plugging (blockage of some of the air and medium passages). In case of plugging, organisms could not pass through the column freely, and accumulation of cells became evident.

For the sieve tray column system, the most severe wall growth occurred right beneath the tray between the compartments. This was due not only to the attaching property of the organisms on a surface but also to the transport of the organism cells by the froth. Deposits beneath the sieve trays resulted in plugging of some of the holes in the sieve tray column. Once the wall growth underneath a sieve tray accumulated to a certain thickness, flocs of cells would desorb from the tray into the medium of the lower stage.

Figure 5 shows the MLSS concentrations inside the sieve tray column with plugging in the lowest compartment. The concentrations indicated were the averages of three consecutive daily measurements. For $D = 0.115/\text{hr}$ a very concentrated MLSS concentration in the first (lowest) compartment was observed, followed by a sharp decrease in the next stage.

With Koch mixers in the column, wall growth appeared not only on the column wall but also on the mixer surfaces. The Koch mixer system did not have a plugging problem; however, sedimentation and extensive wall growth did occur. Figure 6 shows the MLSS concentration profile, at $D = 0.303 \text{ hr}^{-1}$, for the Koch mixer column under the conditions of extensive wall growth. The MLSS concentration was very high (about 1500 mg/l) at the lower part of the Koch mixer column; however, it did not have a sharp drop like that observed in the sieve tray column. Instead, the MLSS concentration decreased gradually in the direction of flow. The extent of wall growth on the mixer surfaces was the highest for the mixer positioned at the bottom of the column; the mixers positioned at the upper portion of the column had less wall growth. The growth on the mixer surfaces gives the system some of the advantages of a trickling filter while retaining the advantages of the activated sludge system. Throughout the operation of the Koch mixer column, no large flocs in

the medium were observed; however, small flocs of about 1/16" in diameter were present.

Figure 6 also indicates that with wall growth the MLSS concentrations in the lower section of the column are much larger than those measured when no wall growth is visible. Improved sedimentation may be due to a change in a dominant species as well as the increased cell mean residence time which results when wall growth is present. It is known that long cell residence time and endogenous metabolism usually improve flocculation and settling.

For both the sieve tray column and the Koch mixer column systems, more than one steady state for the MLSS were observed over a two week period. At low dilution rates, a steady state usually was reached before any significant amount of wall growth was visible. This steady state without wall growth could be maintained for about five to six days for both systems. For the sieve tray column system, the occurrence of wall growth would frequently lead to plugging and another steady state would eventually be reached. For the Koch mixer column system, a new steady state was reached after the wall growth on the mixer surfaces came to its steady thickness, and no further change seemed to take place after that.

The operation of the Koch mixer column system under the conditions of extensive wall growth was more stable than that of the sieve tray column. This is understandable, since the cell flocs in the Koch mixer column system were in a small finely dispersed form and the problem caused by plugging and the loss of large flocs from the sieve trays was non-existent.

Modeling of plugging

In order to validate the assumption that plugging inside of the sieve tray column prevented the cells from flowing to the next stage with the

medium, completely mixed tanks in series model, similar to that suggested by Lee et al. [8], was used to estimate the steady state MLSS and COD concentrations in the column, under the plugging condition.

The steady state material balances based on the completely mixed tanks in series model are:*

1st stage:

$$Q_0 S_0 + f_2 S_2 - Q_1 S_1 - V_1 \left[\frac{\mu_{\max} S_1 X_1}{Y(K_S + S_1)} \right] = 0 \quad (1)$$

$$f_2 \bar{X}_2 - Q_1 \bar{X}_1 + V_1 \left[\frac{\mu_{\max} S_1 X_1}{(K_S + S_1)} - k_d X_1 \right] = 0 \quad (2)$$

2nd stage:

$$Q_1 S_1 + f_3 S_3 - Q_2 S_2 - f_2 S_2 - V_2 \left[\frac{\mu_{\max} S_2 X_2}{Y(K_S + S_2)} \right] = 0 \quad (3)$$

$$Q_1 \bar{X}_1 + f_3 \bar{X}_3 - Q_2 \bar{X}_2 - f_2 \bar{X}_2 + V_2 \left[\frac{\mu_{\max} S_2 X_2}{(K_S + S_2)} - k_d X_2 \right] = 0 \quad (4)$$

3rd stage:

$$Q_2 S_2 + f_4 S_4 - Q_3 S_3 - f_3 S_3 - V_3 \left[\frac{\mu_{\max} S_3 X_3}{Y(K_S + S_3)} \right] = 0 \quad (5)$$

$$Q_2 \bar{X}_2 + f_4 \bar{X}_4 - Q_3 \bar{X}_3 - f_3 \bar{X}_3 + V_3 \left[\frac{\mu_{\max} S_3 X_3}{(K_S + S_3)} - k_d X_3 \right] = 0 \quad (6)$$

* See nomenclature section for definition of symbols.

4th stage:

$$Q_3 S_3 - Q_4 S_4 - f_4 S_4 - V_4 \left[\frac{\mu_{\max} S_4 X_4}{Y(K_S + S_4)} \right] = 0 \quad (7)$$

$$Q_3 \bar{X}_3 - Q_4 \bar{X}_4 - f_4 X_4 + V_4 \left[\frac{\mu_{\max} S_4 X_4}{(K_S + S_4)} - k_d X_4 \right] = 0 \quad (8)$$

The flow balances are:

$$Q_1 = Q_0 + f_2 \quad (9)$$

$$Q_4 = Q_3 - f_4 \quad (10)$$

$$Q_1 + f_3 = Q_2 + f_2 \quad (11)$$

$$\text{and} \quad Q_2 + f_4 = Q_3 + f_3 \quad (12)$$

The cell concentrations leaving stage i in the upward flow \bar{X}_i and the downward flow (backflow) X_i are related to the average concentration X_i using two sedimentation parameters as follows:

$$\bar{X}_i = \delta_i X_i$$

$$X_i = \frac{X_i}{\epsilon_i}$$

The relationship between δ_i and ϵ_i is

$$1/2(\delta_i + \frac{1}{\epsilon_i}) = 1$$

The kinetic constants (μ_{\max} , K_S , k_d , and Y) in these equations were obtained from the kinetic study of activated sludge, by Chiu et al. [13].

Table I lists the values of these kinetic constants. For simplicity back flow rates were assumed constant throughout the operation. In this case, back flow rates were fixed at 20% of the inlet flow rate.

In obtaining the solutions to these equations, the non-linear growth terms in Equations (1) through (8) were linearized using a Taylor series expansion around a set of initial concentrations. The second and higher order terms in the expansions were ignored. The scientific subroutine SIMQ was employed to solve these simultaneous linear equations. The resulting solutions from the calculations were compared with the previous set of concentrations. If the first stage COD and MLSS concentrations of these new solutions differed more than 1% from those of the previous iteration, then, the whole process of computation was repeated using the latest concentrations in the series expansion. The computation was stopped only when the termination criteria was satisfied.

Since the sedimentation factor, δ , was the parameter that determined the amount of cell mass being transported to the next compartment with the upward flow, this factor was adjusted to describe the plugging phenomena. By fixing the sedimentation factor in the first stage, δ_1 , at 0.35 and the values for the other stages at 1.0, the solutions of the simultaneous equations predicted concentrations similar to those of the experimental measurements. The results are shown in Figure 7. Even though the calculated MLSS concentrations in the second and higher stages did not match the experimental data as closely as one would like, the model did validate the assumption that cells were being screened off in the presence of plugging.

Washout

An analysis of washout in tower systems has been reported by Erickson et al. [14]. Without back flow between stages, for the sieve tray column, washout should occur at a system dilution rate of about $D = 0.175 \text{ hr}^{-1}$, if Chiu's [13] kinetic constants were assumed. Sedimentation, wall growth, and back flow would each act to increase the washout dilution rate. Experimentally, washout conditions were approached at a system dilution rate of 0.303 hr^{-1} . Good indications of washout conditions are that the exit substrate concentration is at or close to the feed substrate concentration, and the column is free of cell mass.

Figure 8 shows the COD concentrations and Figure 9 shows the MLSS concentrations at different instants of the transient period for the sieve tray column at $D = 0.303 \text{ hr}^{-1}$. At about 20 hours after starting the continuous run, high COD concentrations (between 800 and 900 mg/l) and low MLSS concentrations (less than 100 mg/l) were measured. These results indicated that the organisms were being washed out. However, wall growth throughout the entire column began to pick up at this point, and the system went through a period (about 55 hours) where the MLSS concentration slowly increased until a new steady state was attained.

This dynamic behavior was probably the result of a change in dominant species of the mixed culture organisms which were utilized in this study. Fast growing organisms were present in the seed solution, initially, in a very small amount; however, after most of the slow growing organisms were washed out, these fast growing organisms utilized the available substrate for growth. Wall growth would be expected to help keep these organisms from washing out completely; however, the hypothesis of a change in dominant

species provides a rational explanation of the return to a new steady state with a low substrate concentration. This change in dominant population was also observed by Chiu et al. [15] at $D = 0.57 \text{ hr}^{-1}$ in a chemostat.

For the Koch mixer column system, an experimental run at a system dilution rate of $D = 0.69 \text{ hr}^{-1}$ also initially showed rapid washout was taking place; however, recovery before complete washout was again observed. The phenomena of washout being avoided through wall growth and changes in dominant species was similar to that of the sieve tray column system; the only difference was that the washout dilution rate was higher for the Koch mixer column system. Figures 10 and 11 show the COD and the MLSS concentrations, respectively, for the Koch mixer column at $D = 0.69 \text{ hr}^{-1}$. It took only about 12 hours for the slower growing organisms to be washed out from the column. The recovery period was about the same as that of the sieve tray column system. Again, sedimentation was observed. Thus far, the sedimentation effects have been observed at extensive wall growth conditions; it is very likely that the factors which cause cells to adhere to the walls also promote the flocculation and settling of cells.

Simulation of washout and recovery

A simple two species model was used to simulate the experimental observations of washout and recovery in the sieve tray column operated at the system dilution rate, $D = 0.303 \text{ hr}^{-1}$. The model assumed the presence of two organism species Z and W, where Z represented the slower growing organisms which were washed out, and W represented the fast growing organisms. The two species were assumed to compete for the limited amount of substrate, S. At the beginning, the stock was assumed to be dominated by the slower growing

organisms; only a few of the fast growing organisms were assumed to be present. Again, the model was based on the completely mixed tanks in series assumption with constant back flow rates. The unsteady state material balances based on this model are as follows:

1st stage:

$$V_1 \frac{dZ_1}{dt} = -Q_1 \bar{Z}_1 + V_1 \left[\frac{\mu_Z S_1 Z_1}{K_Z + S_1} - k_d Z_1 \right] + f_{2Z_2} \quad (13)$$

$$V_1 \frac{dW_1}{dt} = -Q_1 \bar{W}_1 + V_1 \left[\frac{\mu_W S_1 W_1}{K_W + S_1} - k_d W_1 \right] + f_{2W_2} \quad (14)$$

$$V_1 \frac{dS_1}{dt} = Q_0 S_0 - Q_1 S_1 + f_{2S_2} - V_1 \left[\frac{\mu_Z S_1 Z_1}{Y_Z (K_Z + S_1)} + \frac{\mu_W S_1 W_1}{Y_W (K_W + S_1)} \right] \quad (15)$$

2nd stage:

$$V_2 \frac{dZ_2}{dt} = Q_1 \bar{Z}_1 + f_{3Z_3} - Q_2 \bar{Z}_2 - f_{2Z_2} + V_2 \left[\frac{\mu_Z S_2 Z_2}{K_Z + S_2} - k_d Z_2 \right] \quad (16)$$

$$V_2 \frac{dW_2}{dt} = Q_1 \bar{W}_1 + f_{3W_3} - Q_2 \bar{W}_2 - f_{2W_2} + V_2 \left[\frac{\mu_W S_2 W_2}{K_W + S_2} - k_d W_2 \right] \quad (17)$$

$$V_2 \frac{dS_2}{dt} = Q_1 S_1 + f_{3S_3} - Q_2 S_2 - f_{2S_2} - V_2 \left[\frac{\mu_Z S_2 Z_2}{Y_Z (K_Z + S_2)} + \frac{\mu_W S_2 W_2}{Y_W (K_W + S_2)} \right] \quad (18)$$

3rd stage:

$$V_3 \frac{dZ_3}{dt} = Q_2 \bar{Z}_2 + f_{4Z_4} - Q_3 \bar{Z}_3 - f_{3Z_3} + V_3 \left[\frac{\mu_Z S_3 Z_3}{K_Z + S_3} - k_d Z_3 \right] \quad (19)$$

$$V_3 \frac{dW_3}{dt} = Q_2 \bar{W}_2 + f_4 \bar{W}_4 - Q_3 \bar{W}_3 - f_3 \bar{W}_3 + V_3 \left[\frac{\mu_W S_3 W_3}{K_W + S_3} - k_d W_3 \right] \quad (20)$$

$$V_3 \frac{dS_3}{dt} = Q_2 S_2 + f_4 S_4 - Q_3 S_3 - f_3 S_3 - V_3 \left[\frac{\mu_Z S_3 Z_3}{Y_Z (K_Z + S_3)} + \frac{\mu_W S_3 W_3}{Y_W (K_W + S_3)} \right] \quad (21)$$

4th stage:

$$V_4 \frac{dZ_4}{dt} = Q_3 \bar{Z}_3 - Q_4 \bar{Z}_4 - f_4 \bar{Z}_4 + V_4 \left[\frac{\mu_Z S_4 Z_4}{K_Z + S_4} - k_d Z_4 \right] \quad (22)$$

$$V_4 \frac{dW_4}{dt} = Q_3 \bar{W}_3 - Q_4 \bar{W}_4 - f_4 \bar{W}_4 + V_4 \left[\frac{\mu_W S_4 W_4}{K_W + S_4} - k_d W_4 \right] \quad (23)$$

$$V_4 \frac{dS_4}{dt} = Q_3 S_3 - Q_4 S_4 - f_4 S_4 - V_4 \left[\frac{\mu_Z S_4 Z_4}{Y_Z (K_Z + S_4)} + \frac{\mu_W S_4 W_4}{Y_W (K_W + S_4)} \right] \quad (24)$$

The flow balances between the stages are given by Equations (9) through (12).

The kinetic constants were obtained from the work by Chiu et al. [13]; the fastest and the slowest growth kinetics were selected. The kinetic constants for the slower growing and faster growing organisms are presented in Tables II and III, respectively. These unsteady state ordinary differential equations were solved by employing the IBM Continuous System Modeling Programming (CSMP). By utilizing the two species model with the presence of back flow, the general trend of washout and recovery of the organisms was simulated to some success. The difficulty was to get the MLSS and COD concentrations to match those of the experimental data at the right time. After a few trials with $\delta = 0.9$, $c = 0.9$ and several different values of back flow rates, a

rather close fit of the experimental data was found at a back flow rate equal to 150% of the inlet flow rate, Q_0 . The simulated results are compared with the experimental results in Figures 12 and 13.

In Figure 12, the first stage simulated MLSS concentrations were higher than those of the experimental measurements. At this dilution rate ($D = 0.303 \text{ hr}^{-1}$), the organism cells tended to diffuse to the column wall and attach there. Hence, the measured cell concentration in the solution was probably smaller than the average concentration actually existing in the stage. In Figure 13 the simulated results for the COD concentrations at 39 hours and at 75 hours were higher than the experimental values. The model assumed the presence of only one fast growing organism species; in the experimental study there may be other organisms with higher growth rates than the species assumed to be present. These organisms with higher growth rates would consume the substrate at a faster rate, and bring the substrate concentration down to the lower level indicated in the experimental data.

The effects of back flow rate on washout have been discussed. Simulations using the two species model were carried out at several back flow rates. Figure 14 shows the simulated first stage total MLSS concentration as a function of time with back flow rate as the parameter. With no back flow, the organisms in the first stage were washed out completely in less than 10 hours. If a significant amount of back flow was present, then, instead of complete washout, the fast growing organisms would achieve a steady state concentration. A larger back flow rate resulted in a higher organism concentration inside the compartment throughout the transient period as indicated in Figure 14.

The effect of sedimentation on the washout behavior of the sieve tray column is shown in Figure 15 where simulated values of the first stage total MLSS concentration are plotted as a function of time with the sedimentation factor as the parameter. A constant back flow rate equal to 150% of the inlet flow rate is used. The results show that the cell sedimentation affects the washout of the column significantly. Without cell sedimentation, the first stage organisms were washed out in a short period of time (about 20 hours), and the recovery did not pick up until after 20 more hours of continuous operation. With significant sedimentation ($\delta = 0.8$), the simulated results showed a steady increase in the first stage cell concentration during the first 30 hours; the increase slowed down for a short period and then picked up again at a slower rate than that of the original increase. The concentration increase at the beginning of the operation was due to organism sedimentation from the upper stages plus those generated by growth. As the concentration of the slower growing organisms in the upper stages was reduced, the amount of the settling organisms decreased, so the rate of increase slowed down. After a time lag, the faster growing organisms picked up the growth and the cell concentration started to increase again.

Comparison of Figures 14 and 15 shows that the effect of cell sedimentation on the organism concentration level in the first stage is greater than that of back flow.

No model simulation was carried out for the Koch mixer column system; however, the effects of back flow and sedimentation on its performance near the washout dilution rate should be similar to those of the sieve tray column system.

Concluding Remarks

The results showed that both the sieve tray column and the Koch mixer column could be used for the growth of mixed cultures of sewage origin. Although the MLSS concentration was observed to be affected by wall growth and sedimentation, the removal of COD was stable for a period of two weeks of continuous operation. The Koch mixer column system was operated without any significant problems; however, plugging of the holes in the sieve trays was experienced using the sieve tray column.

Plugging in the sieve tray column was observed to be caused by the transport of cells by froth. To reduce the plugging, the hole size of the sieve trays may have to be enlarged.

Wall growth and sedimentation were observed in some of the experiments, especially at higher dilution rates. The sedimentation of cells increased as the wall growth increased. Under conditions of extensive wall growth, the MLSS concentration in the lower portion of the Koch mixer column more than doubled because of sedimentation. Sedimentation within the column is advantageous in activated sludge processes. Wall growth and sedimentation also improve the stability of the system with respect to washout.

The competition between different species of organisms was considered in computer simulation studies which were used to help interpret some of the experimental results. The experimental and computer results suggest that slow growing organisms are initially present in large quantities in the inoculum, and that these organisms wash out while faster growing organisms, which are present in the inoculum in much smaller numbers, grow and multiply in the tower system. Computer simulation was found to be a helpful tool in interpreting experimental results.

Nomenclatures

- $D = Q_0/V$ = system dilution rate based on the total tower volume, hr^{-1} .
 f_i = backflow rate from stage i , hr^{-1} .
 K_s = the saturation constant of the mixed culture organisms, mg/l .
 K_Z = the saturation constant of the slow growing organisms, mg/l .
 K_W = the saturation constant of the fast growing organisms, mg/l .
 k_d = organism endogenous decay constant, hr^{-1} .
 Q_0 = inlet substrate flow rate, l/hr .
 Q_i = upward flow rate from stage i , mg/l .
 S_0 = inlet substrate concentration, mg/l .
 S_i = substrate concentration at stage i , mg/l .
 V = total tower volume, l .
 v_i = liquid volume at stage i , l .
 W_i = cell concentration of fast growing organisms at stage i , mg/l .
 \bar{W}_i = fast growing organism concentration of the upward flow stream from stage i , $\text{mg/l} = \delta_i W_i$.
 \underline{W}_i = fast growing organism concentration of the backflow stream from stage i , $\text{mg/l} = W_i/\epsilon_i$.
 X_i = cell concentration of the mixed culture organisms at stage i , mg/l .
 \bar{X}_i = $\delta_i X_i$ = mixed culture organism concentration of the upward flow stream from stage i , mg/l .
 \underline{X}_i = X_i/ϵ_i = mixed culture organism concentration of the backflow stream from stage i , mg/l .
 Y = yield coefficient of the mixed culture organisms, dimensionless.
 Y_Z = yield coefficient of the slow growing organisms, dimensionless.
 Y_W = yield coefficient of the fast growing organisms, dimensionless.
 Z_i = cell concentration of the slow growing organisms at stage i , mg/l .

\bar{Z}_1 = slow growing organism concentration of the upward flow stream from stage 1, mg/l = $\delta_1 Z_1$.

\underline{Z}_1 = slow growing organism concentration of the backflow stream from stage 1, mg/l = Z_1/δ_1 .

δ_1 = sedimentation factor at the top of stage 1, dimensionless, $0 \leq \delta_1 \leq 1$.

ϵ_1 = $1/(2 - \delta_1)$ = sedimentation factor at the bottom of stage 1, dimensionless, $0 \leq \epsilon_1 \leq 1$.

μ_{\max} = the maximum specific growth rate for the mixed culture organisms, hr⁻¹.

μ_Z = maximum specific growth rate of the slow growing organisms, hr⁻¹.

μ_W = maximum specific growth rate of the fast growing organisms, hr⁻¹.

References

1. A. Kitai, H. Tone, and A. Ozaki, J. Ferment. Technol., 47, 333 (1969)
2. A. Kitai, S. Goto, and A. Ozaki, J. Ferment. Technol., 47, 340 (1969)
3. A. Kitai, S. Goto, and A. Ozaki, J. Ferment. Technol., 47, 348 (1969)
4. A. Kitai, S. Goto, and A. Ozaki, J. Ferment. Technol., 47, 356 (1969)
5. A. Prokop, L.E. Erickson, J. Fernandez, and A.E. Humphrey, Biotechnol. Bioeng., 11, 945 (1969)
6. E.A. Falch and E.L. Gaden, Biotechnol. Bioeng., 11, 927 (1969)
7. E.A. Falch and E.L. Gaden, Biotechnol. Bioeng., 12, 465 (1970)
8. S.S. Lee, L.E. Erickson, and L.T. Fan, Biotechnol. Bioeng. Symp. No. 2, 141 (1971)
9. F. Besik, Water and Sewage Works, Sept., 120, 122 (1973)
10. Y. Sonoda, S. Tanaka, and Y. Ishida, J. Ferment. Technol., 51, 813 (1973)
11. Koch Static Mixing Systems and Reactor Internals, Bulletin KSM-2, Koch Engineering Co., Wichita, Kansas
12. Standard Methods for the Examination of Water and Waste Water, 12th ed., American Public Health Assoc., Inc. (1965)
13. S.Y. Chiu, L.T. Fan, I.C. Kao, and L.E. Erickson, Biotechnol. Bioeng., 14, 179 (1972)
14. L.E. Erickson, S.S. Lee, and L.T. Fan, Biotechnol. Bioeng. Symp. No. 4, 301 (1973)
15. S.Y. Chiu, L.E. Erickson, L.T. Fan, and I.C. Kao, Biotechnol. Bioeng., 14, 207 (1972)

Table I. The kinetic constants used in solving the simultaneous equations.

μ_{\max}	0.69 hr^{-1}
K_S	26.5 mg/l
k_d	0.0189 hr^{-1}
Y	0.577

Table II. The kinetic constants for the slower growing organism, Z.

μ_Z	0.237 hr^{-1}
K_Z	3.2 mg/l
k_d	0.02 hr^{-1}
y_Z	0.481

Table III. The kinetic constants for the faster growing organism, W.

μ_W	0.951 hr^{-1}
K_W	246.2 mg/l
k_d	0.02 hr^{-1}
Y_W	0.551

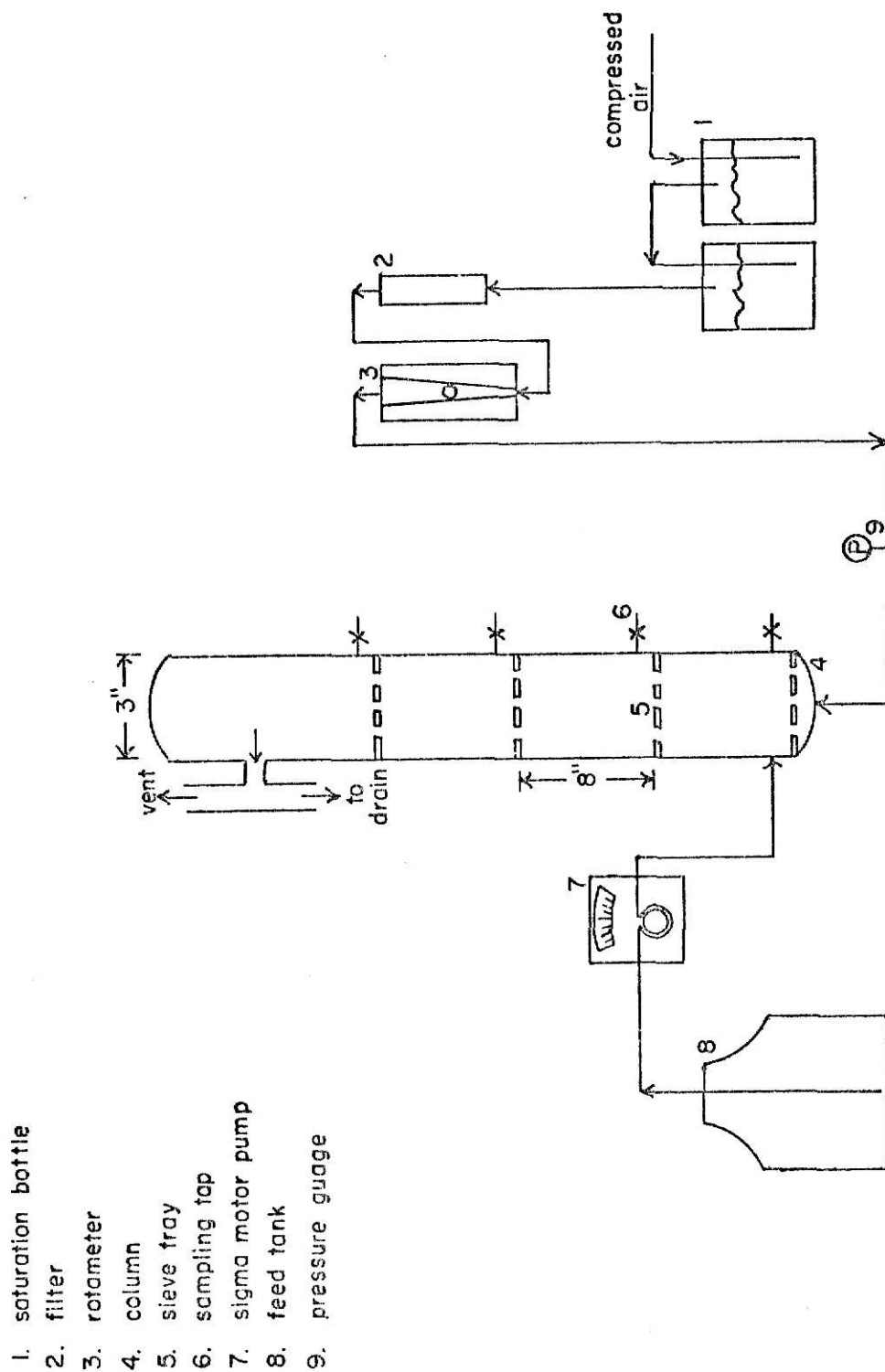


Fig. 1. Schematic diagram of the sieve tray column system.

1. saturation bottle
2. filter
3. rotameter
4. column
5. Koch mixer
6. sampling tap
7. sigma motor pump
8. feed tank
9. pressure guage

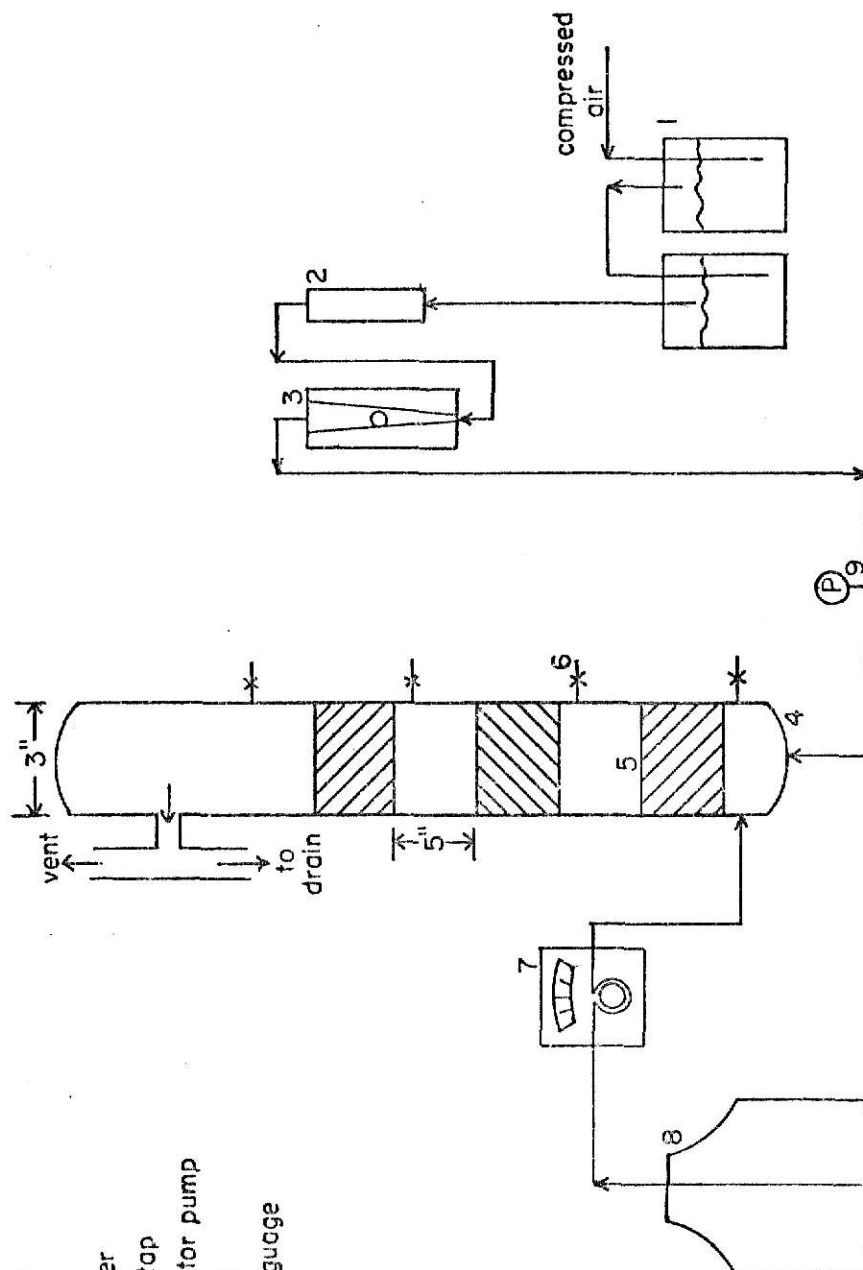


Fig. 2a. Schematic diagram of the Koch mixer column system.



Fig. 2b. Koch mixer

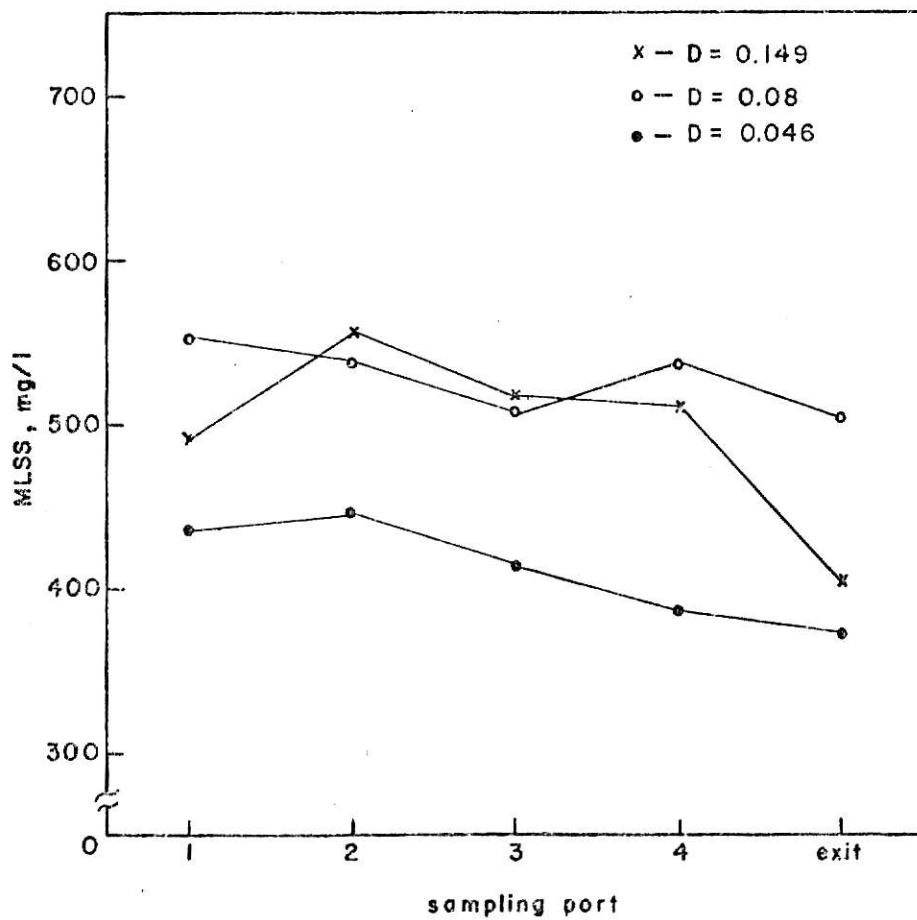


Figure 3. MLSS profile in the 4 stage sieve tray tower, with dilution rate (hr^{-1}) as parameter, Air flow rate = $14 \text{ ft}^3/\text{hr}$.

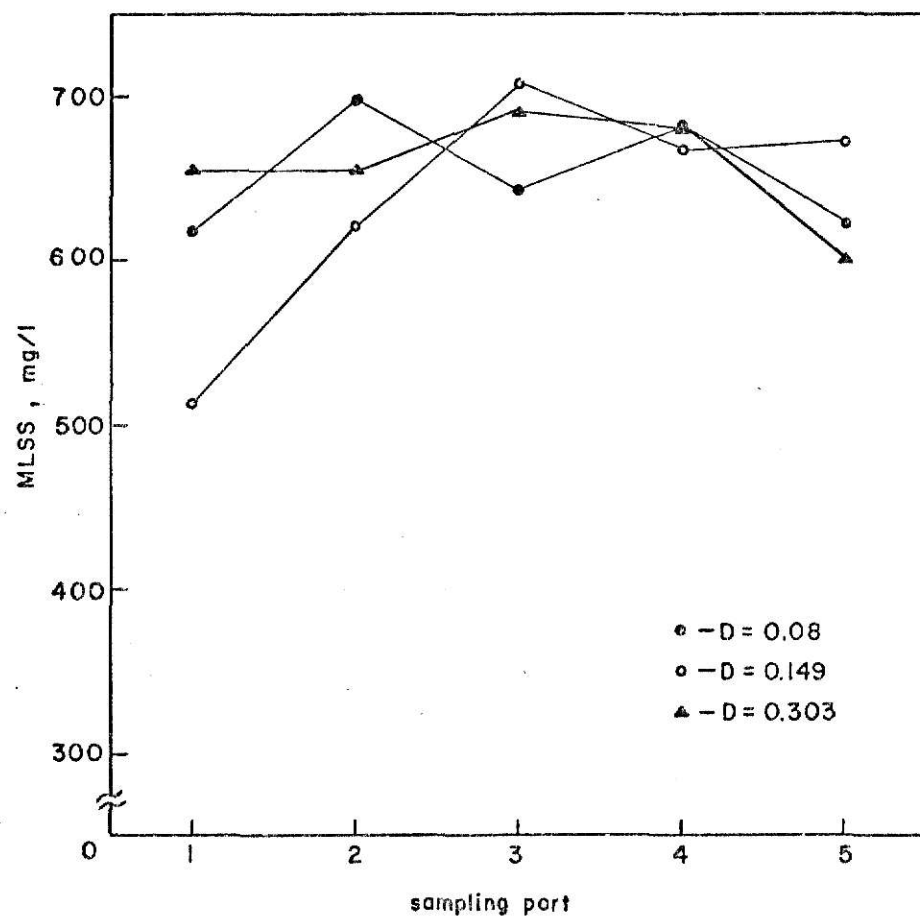


Figure 4. MLSS profile in the Koch mixer tower, with dilution rate (hr^{-1}) as parameter. Air flow rate = $14 \text{ ft}^3/\text{hr}$

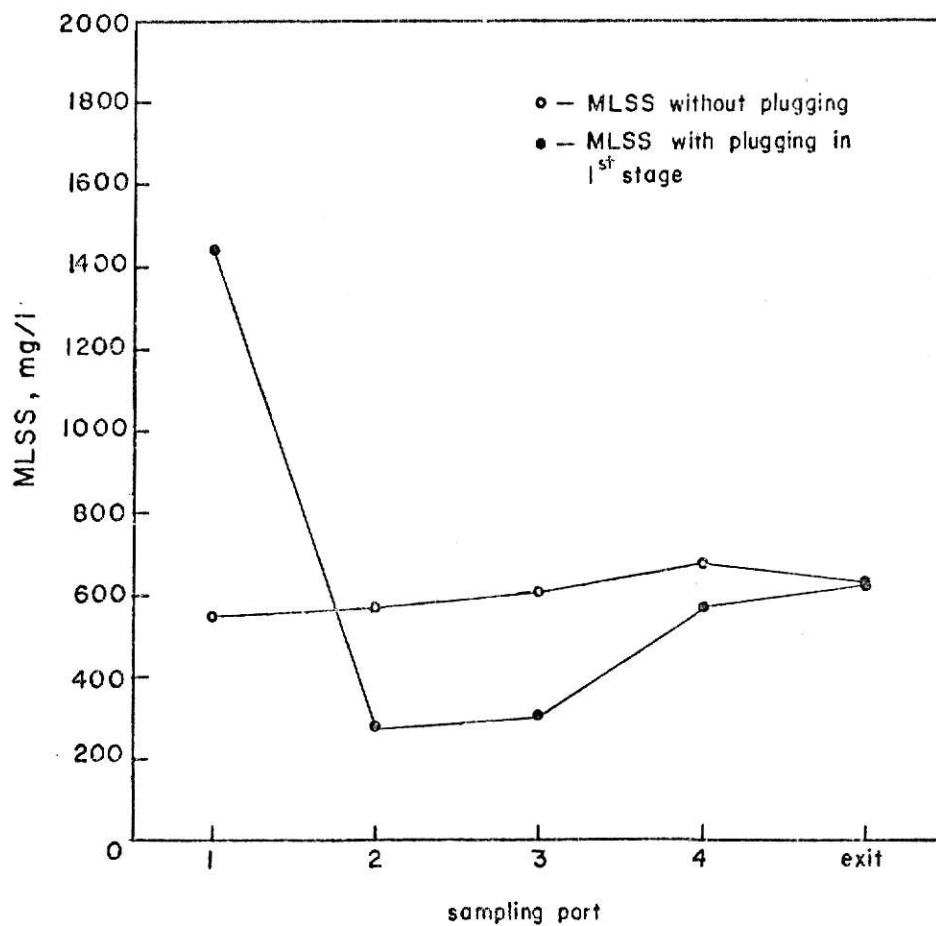


Figure 5. Effect of plugging on the performance of 4 stage sieve tray tower. $D = 0.115 \text{ hr}^{-1}$, air flow rate = $14 \text{ ft}^3/\text{hr}$.

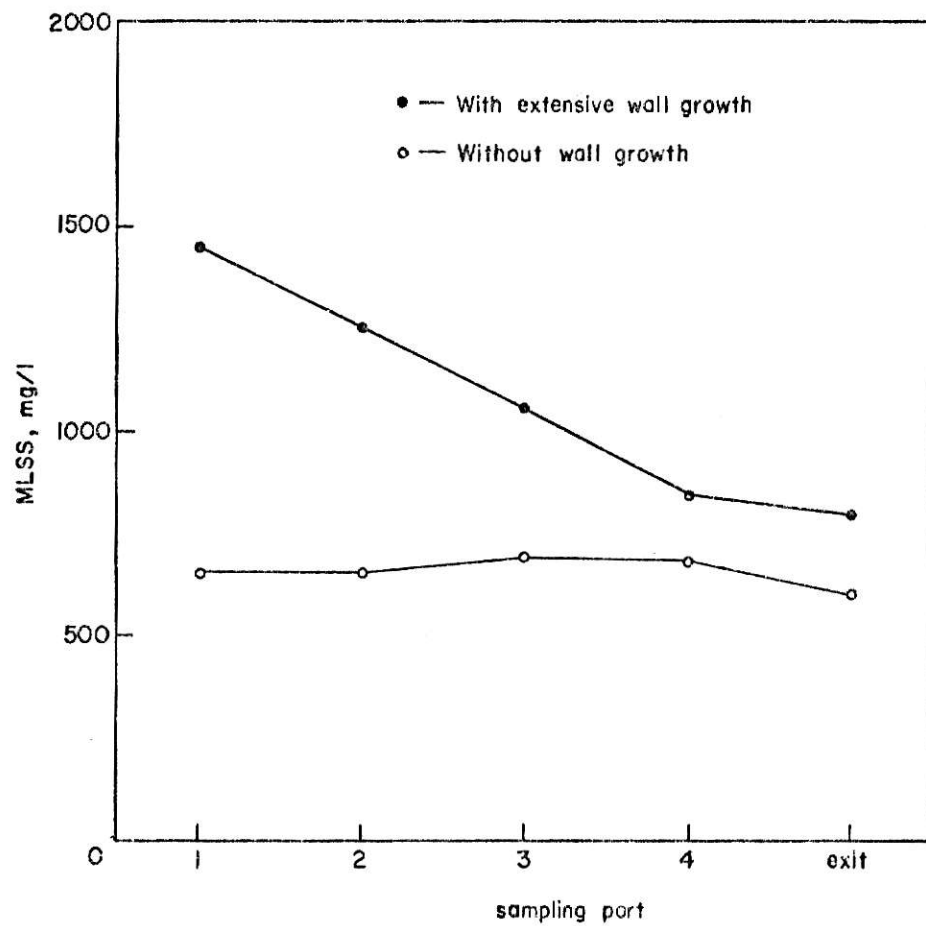


Figure 6. Effect of extensive wall growth on the MLSS profile in the Koch mixer tower. $D = 0.303 \text{ hr}^{-1}$, air flow rate = $14 \text{ ft}^3/\text{hr}$.

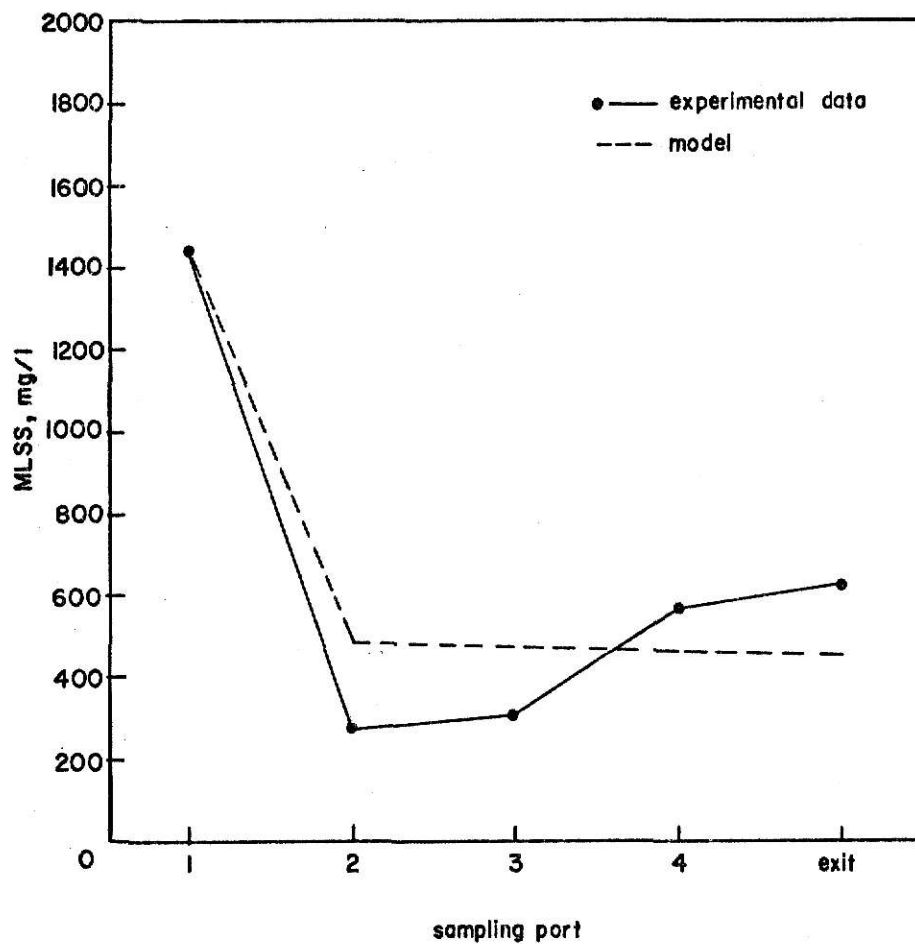


Figure 7. Comparison of the experimental data and the proposed model under the plugging condition for the sieve tray tower system. $D = 0.115 \text{ hr}^{-1}$, air flow rate = $14 \text{ ft}^3/\text{hr}$.

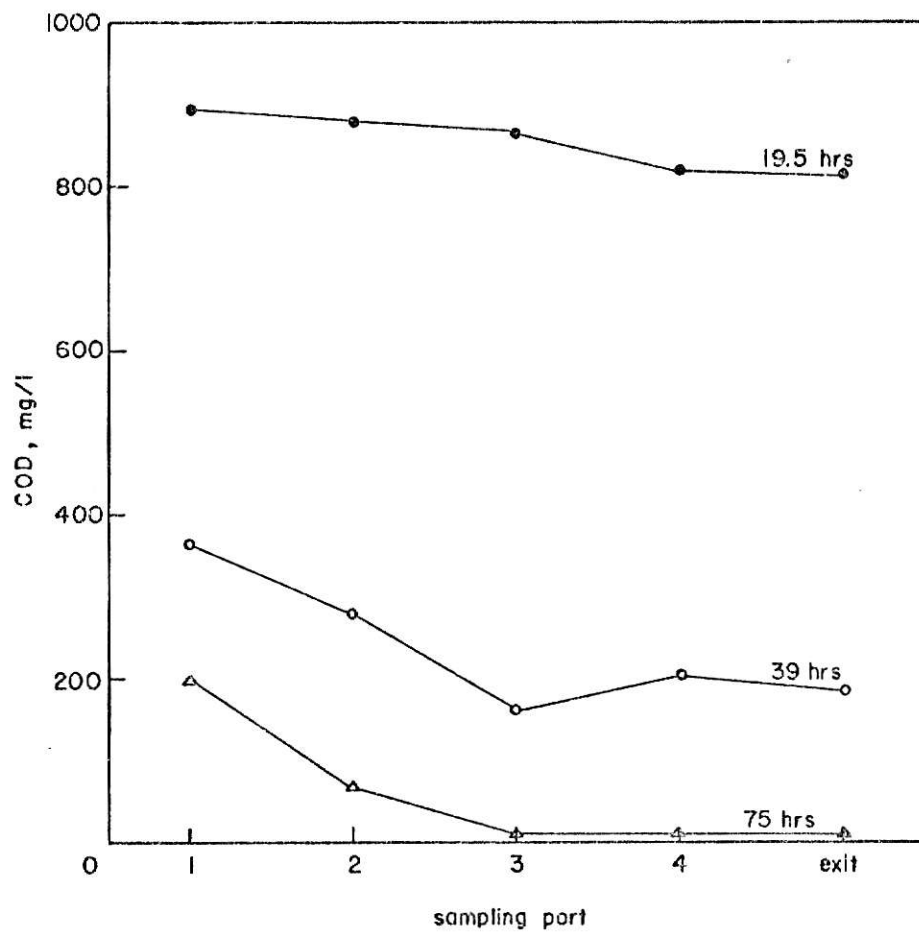


Figure 8. Dissolved COD profiles for the sieve tray column for $D=0.303 \text{ hr}^{-1}$ with time after inoculation as a parameter.

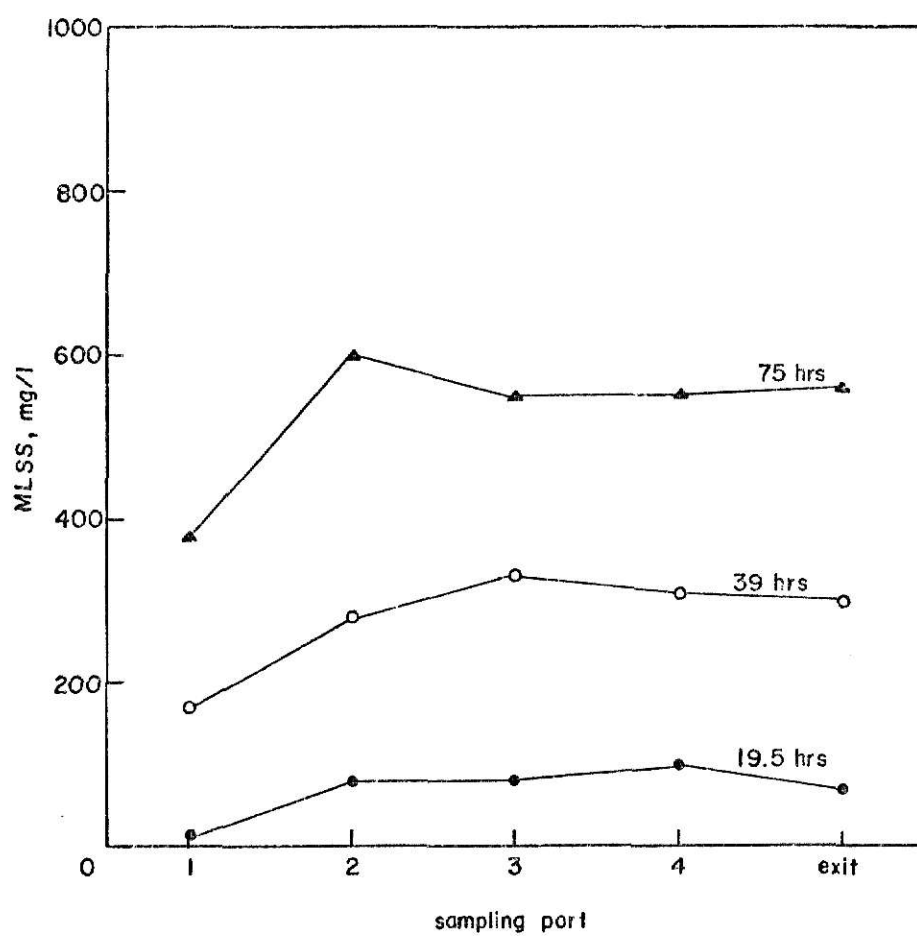


Figure 9. MLSS profiles for the sieve tray column for $D=0.303 \text{ hr}^{-1}$ with time after inoculation as a parameter.

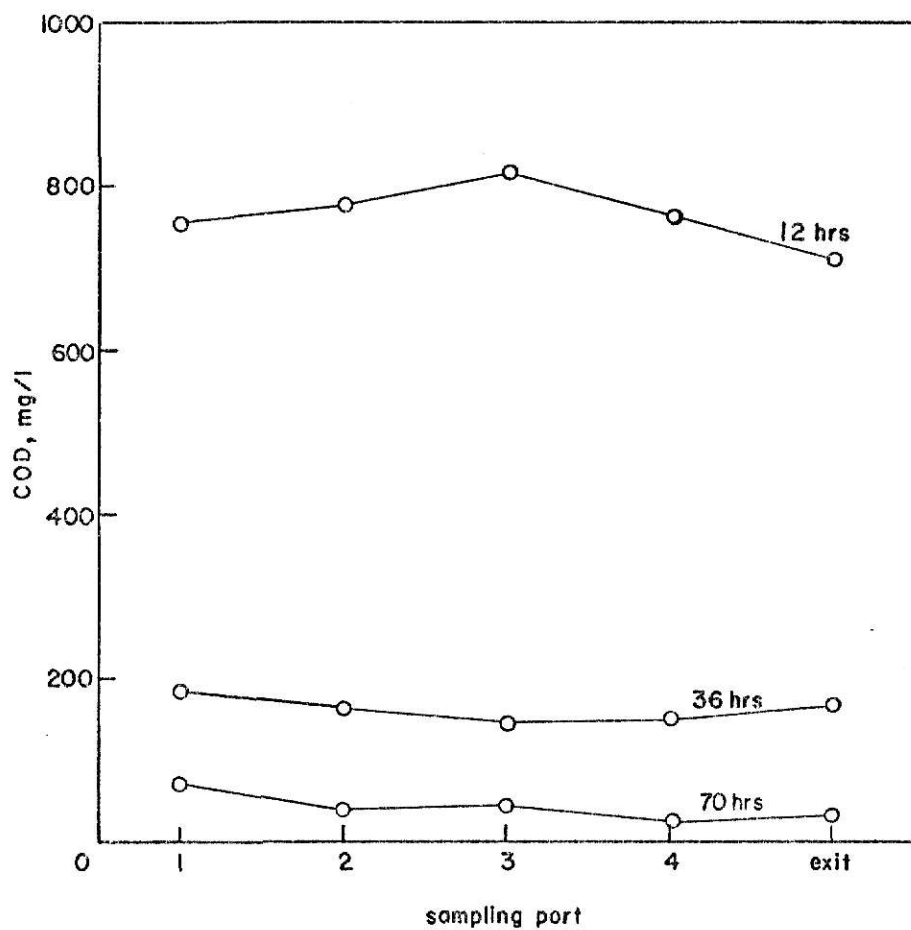


Figure 10. Dissolved COD profiles for the Koch mixer column for $D = 0.69 \text{ hr}^{-1}$ with time after inoculation as a parameter.

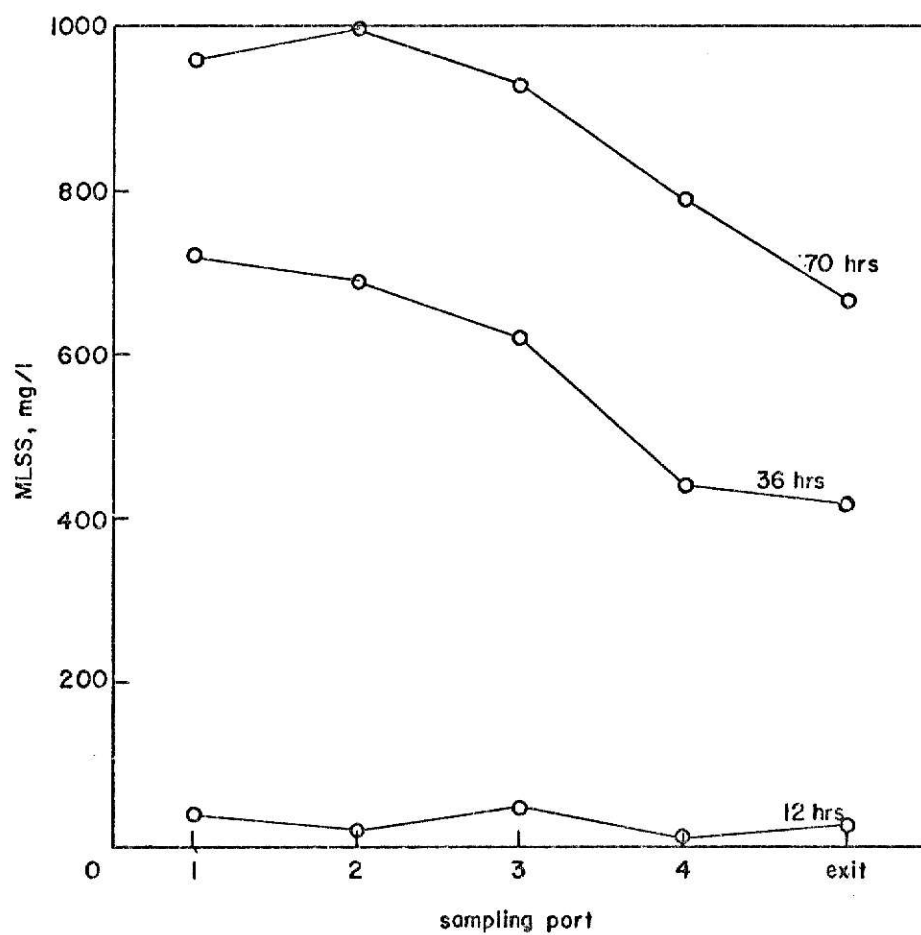


Figure 11. MLSS profiles for the Koch mixer column for $D=0.69 \text{ hr}^{-1}$ with time after inoculation as a parameter.

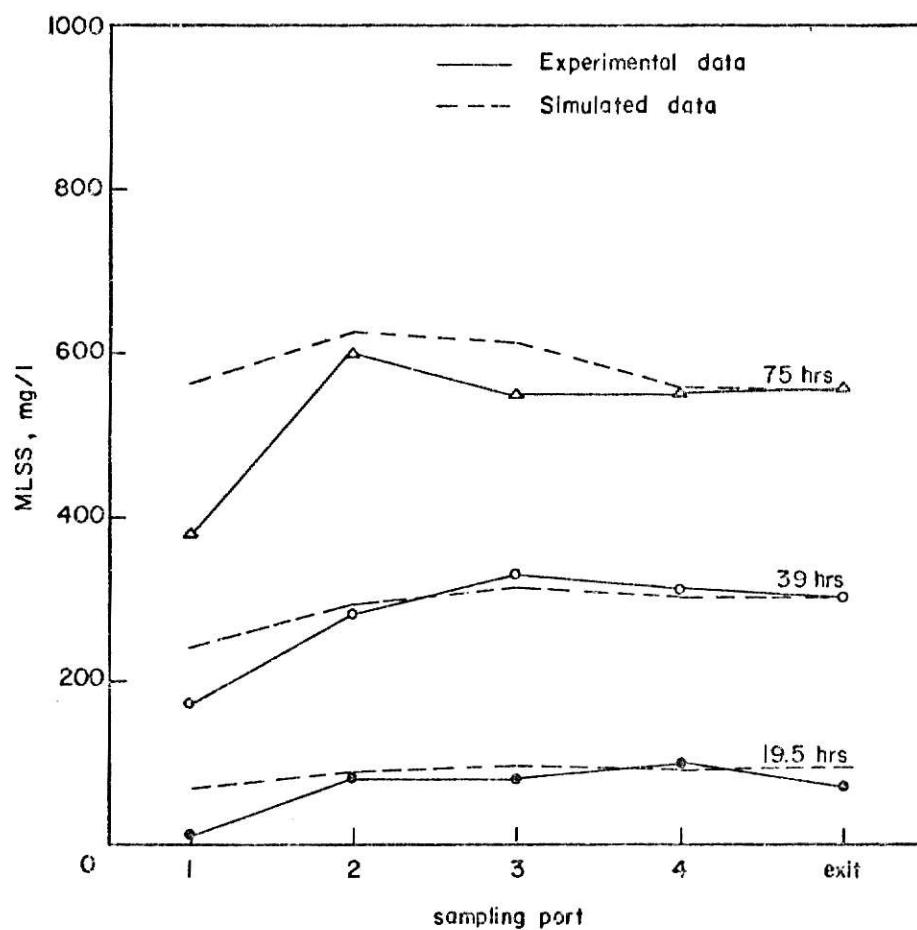


Figure 12. Comparison of the experimentally measured MLSS concentration profiles and the simulated results for the sieve tray column for $D = 0.303 \text{ hr}^{-1}$.

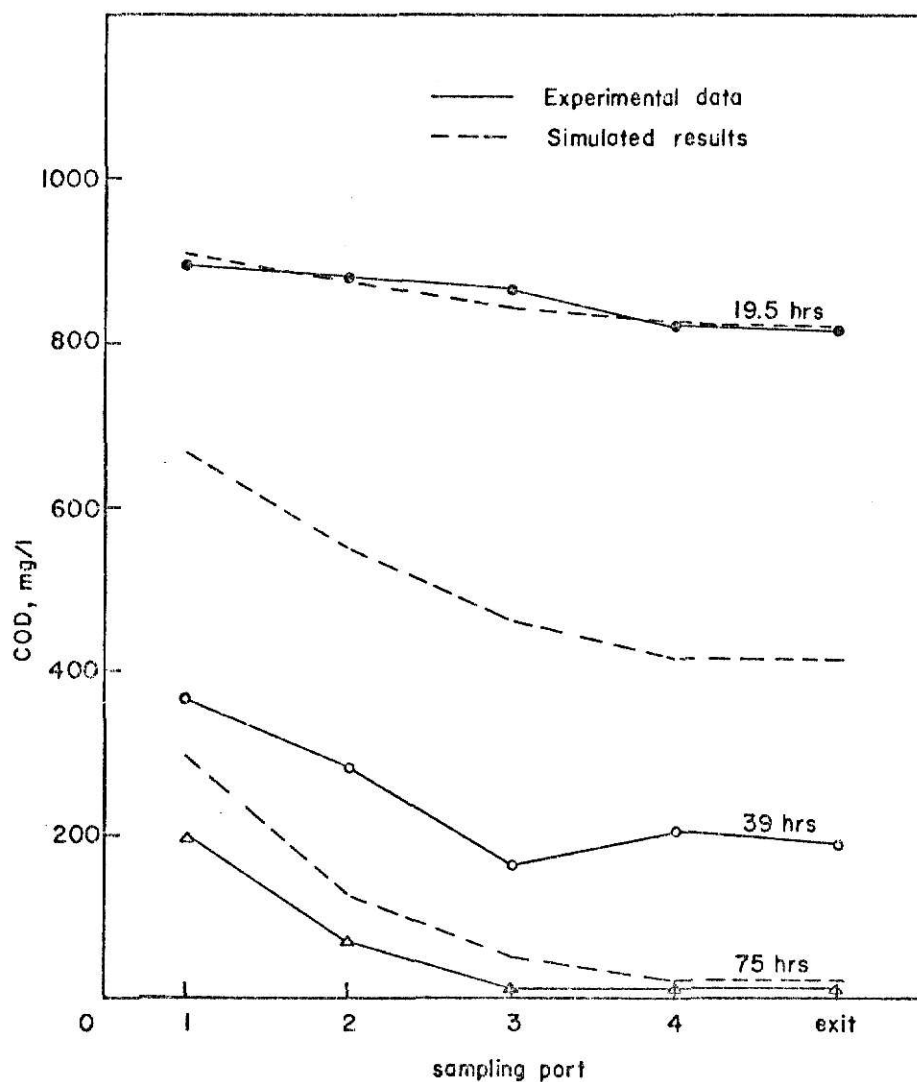


Figure 13. Comparison of the experimentally measured COD profile with the simulated results for the sieve tray column for $D = 0.303 \text{ hr}^{-1}$.

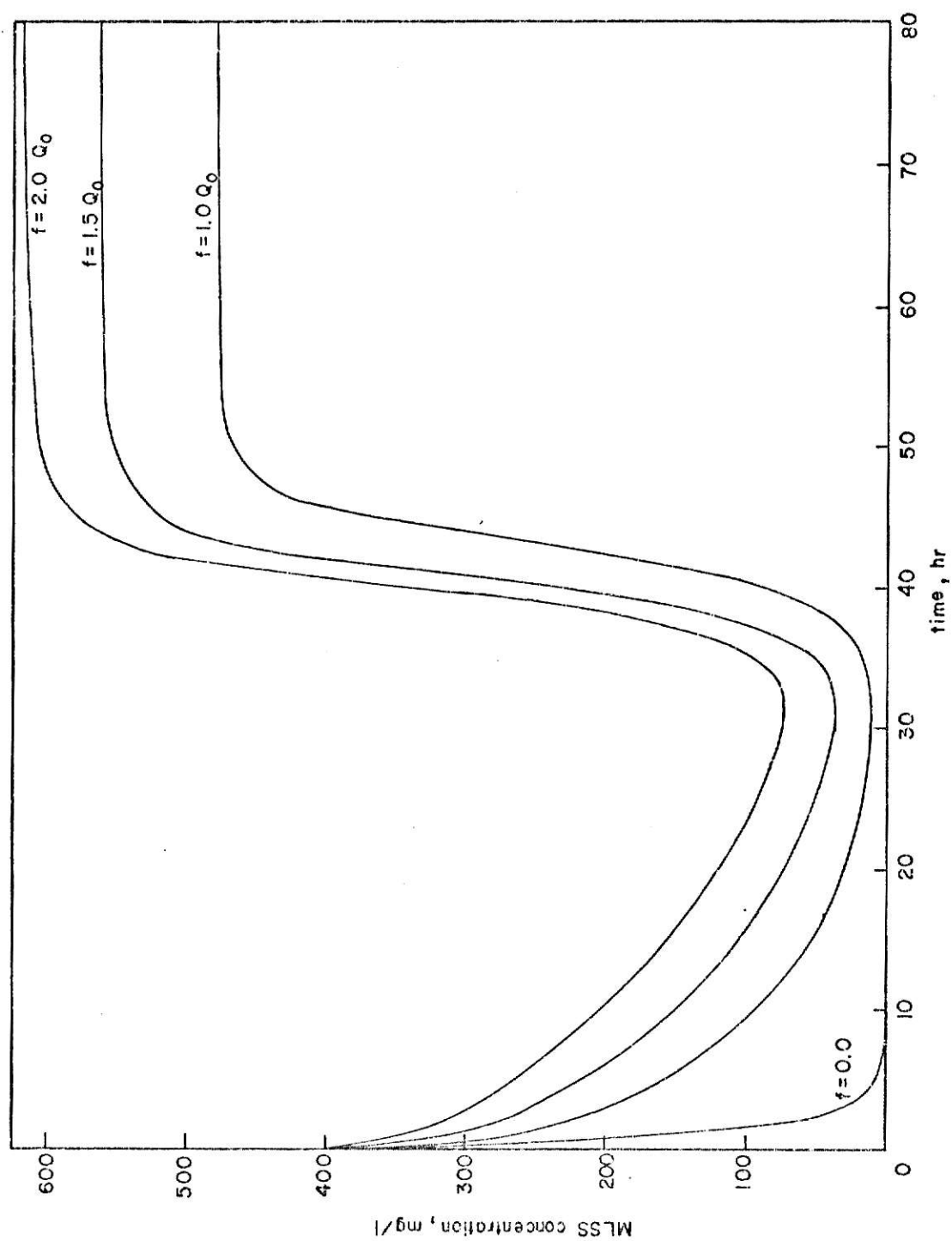


Figure 14. Simulated 1st stage MLSS concentration as a function of time with back flow rate as parameter, $\beta = 0.9$.

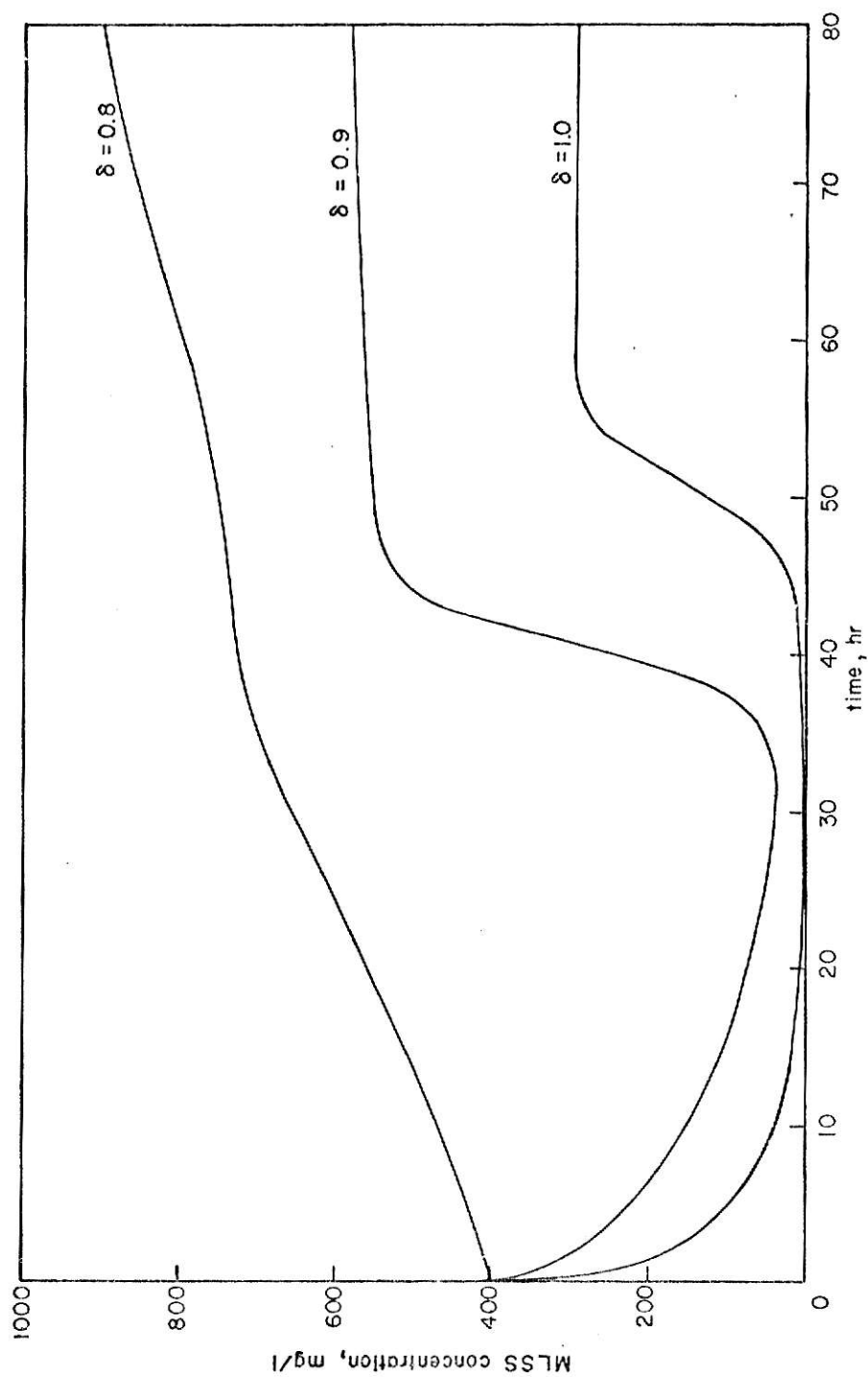


Figure 15. Simulated 1st stage MLSS concentration as a function of time with sedimentation factor as the parameter. Sieve tray column system, $f = 1.5 Q_0$, $D = 0.303 \text{ hr}^{-1}$.

CHAPTER IV

OXYGEN TRANSFER TO MIXED CULTURES IN TOWER SYSTEMS

Introduction

Oxygen is required for aerobic microorganisms to grow and function properly. Unlike most of the other microbial nutrients which can be added in large quantity initially, oxygen has to be supplied continuously because of its low solubility in aqueous solutions. Therefore, the success of an aerobic biological process depends strongly upon the rate at which dissolved oxygen is being supplied.

Many methods for providing sufficient oxygen have been proposed. The most successful one consists of growing organisms throughout a mechanically agitated tank adequately sparged with air. This provides large gas-liquid interfacial area which facilitates the transfer. Most of the oxygen transfer studies in biological processes have utilized this type of system.

Multi-stage tower systems, which have good mass transfer characteristics, can be used to grow aerobic cultures. Previous studies of oxygen transfer in tower systems have been carried out by Kitai et al. [1,2] and Falch and Gaden [3].

Kitai et al. [1,2] investigated oxygen transfer in a multi-stage tower fermentor. In their study, batch experiments were run with respect to the liquid phase. The sulfite oxidation method as suggested by Cooper et al. [4] was used. The effects of gas flow rate, plate design, plate spacing and column size were investigated. They found that the relationship between the oxygen transfer coefficient, $K_L a$, and superficial air velocity was linear on

a log-log scale, and that the $K_L a$ values obtained were competitive to those obtained in an agitated vessel.

Falch and Gaden [3] investigated oxygen transfer in a multi-stage tower fermentor equipped with mechanical agitators. Again, the sulfite oxidation method was employed. They found that the oxygen transfer rate was strongly dependent upon the extent of mechanical mixing while changes in the air flow had a smaller effect.

In this work, oxygen transfer in a sieve tray tower system and a Koch mixer tower system was investigated. A growing mixed culture of sewage origin was present during the oxygen transfer study. The dynamic method [5-7] was used to directly record the dissolved oxygen concentration during the transient period following a step change in gas flow rate. The data obtained were used to determine the oxygen transfer coefficient, $K_L a$, using a modified technique. The results are compared with those obtained from other literature.

Theoretical

In the mixed culture system where oxygen is supplied by bubbling air into the culture medium and simultaneously consumed by the organisms, the overall oxygen balance in the aqueous solution can be simply expressed as:

$$\begin{aligned} (\text{oxygen accumulated}) = (\text{oxygen transferred in}) - (\text{oxygen consumed} \\ \text{by the organisms}). \end{aligned}$$

For a tower system with air bubbling to the liquid, one can assume that at a particular point in the tower the liquid phase is well mixed for the area closely surrounding this point. If local complete mixing is assumed, the above balance can be mathematically expressed as

$$V \frac{dC}{dt} = N_A V - rXV \quad (1)$$

where

V = liquid volume, l.

C = dissolved oxygen concentration, mg/l.

t = time, hr.

N_A = oxygen transfer rate, mg/hr·l.

r = oxygen consumption rate per unit weight of organisms,
mg O_2 /mg cell·hr.

X = cell concentration, mg/l.

The oxygen transfer rate, N_A , can be expressed as:

$$N_A = K_L a (C^* - C) \quad (2)$$

where

$K_L a$ = local overall transfer coefficient based on the liquid film resistance assumption, hr^{-1} .

C^* = oxygen concentration in the medium which is in equilibrium with the oxygen partial pressure in the gas phase at the system temperature, mg/l.

$(C^* - C)$ = concentration driving force which promotes the transfer, mg/l.

Equation (1) now becomes:

$$\frac{dC}{dt} = K_L a (C^* - C) - rX \quad (3)$$

At steady state, the derivative with respect to time is equal to zero; this gives

$$\frac{dC}{dt} = 0 = K_L a (C^* - C_s) - rX \quad (4)$$

or

$$rX = K_L a (C^* - C_s) \quad (5)$$

where

C_s = steady state dissolved oxygen concentration, mg/l.

The term rX can be assumed constant as long as the dissolved oxygen concentration is above the critical oxygen concentration for the culture and the organisms concentration, X , remains constant. Equation (5) can be used to estimate the oxygen consumption rate by the organisms if $K_L a$, C^* , and C_s are known.

Substituting Equation (5) into Equation (3) gives,

$$\frac{dC}{dt} = K_L a (C^* - C) - K_L a (C^* - C_s) \quad (6)$$

Equation (6) can be further reduced to

$$\frac{dC}{dt} = K_L a (C_s - C) \quad (7)$$

According to Equation (7), a graphical estimation of the slope, dC/dt , at various values of C along the dynamic oxygen trace, followed by a plot of dC/dt vs. $(C_s - C)$ should give a straight line which passes through the origin. The slope of the resulting line should be equal to $K_L a$. Equation (7) allows the calculation of $K_L a$ from a recording of the transient dissolved oxygen concentration without knowledge of either C^* or rX . Both C^* and rX are assumed to be constant during the dynamic experiment.

Experimental

Design

Recently, with the development of fast responding oxygen probes, instantaneous measurement of the dissolved oxygen in fermenting medium is possible. With this improvement in the measuring device, measurement of oxygen transfer by recording the transient dissolved oxygen concentration during dynamic operation can be easily achieved. Several methods involving the measurement and analysis of the dynamic trace have been discussed by Bandyopadhyay and Humphrey [5], Wang and Humphrey [6], and Fujita and Hashimoto [7]. The system used in their studies has been limited to agitated vessels where the mixing of the fermenting media was mainly provided by mechanical agitation. Even under non-gassing conditions good mixing of the liquid phase can still be achieved with mechanical agitation.

For tower systems where no mechanical agitation is used, mixing of the media is provided by aeration and the flow of air through the static mixers. In this case, termination of gas flow to the system greatly reduced the mixing and enhanced the settling of cells; this created regions with different cell concentrations, oxygen uptake rates, and dissolved oxygen concentrations. Because of this, a technique involving non-gassing of the system was not applied here.

Instead of using a non-gassing method to reduce the dissolved oxygen concentration in the medium, dynamic experiments were carried out by increasing (step up) or decreasing (step down) the air flow rate. By using this continuous gassing method, some advantages of the non-gassing method, such as direct measurement of the oxygen uptake rate by the organisms, were lost. However, mixing of the medium and homogeneity of the suspension could be

ensured. Fortunately, the calculation of the oxygen transfer coefficients does not require any knowledge of the oxygen consumption rate by the microorganisms. This continuous gassing measurement provided a direct and reliable means of obtaining data for calculating the oxygen transfer coefficient.

Equipment

The sieve tray column system consisted of a 3.664 liter bubble column divided into 4 compartments by horizontally positioned sieve trays. A schematic diagram of the system is shown in Figure 1. Each sieve tray was made of a 1/16 inch thick brass plate with hole diameter of 3/64 inch and free open area to column cross sectional area ratio of 3%.

The system was operated cocurrently with liquid being pumped into the column by means of a sigma motor pump (type AL-4-E-40, by Sigma Motor, Inc.). Air was pumped by an air compressor through the humidifiers, an air filter, and a flow meter to the nozzle at the bottom of the column. A pressure gauge was installed between the flow meter and the inlet nozzle to the column to detect any pressure build up.

The equipment arrangement for the Koch mixer column system was similar to that of the sieve tray column system. Instead of using sieve trays, the Koch mixers were used to distribute the air. A schematic diagram of the system is shown in Figure 2a. The mixers were made of stainless steel sheets bent into wave shape and put together with the layers parallel to each other and the corrugation angle of the adjacent layers reversed with respect to the mixing axis. The mixers used in this study were of the BY type [8]; they were 3 inches tall and 3 inches in diameter with a layer height of 1/4 inch (see Figure 2b). This system was also operated cocurrently.

Synthetic waste

The composition of the glucose-limiting media contained, per liter:

Glucose	1000 mg
$(\text{NH}_4)_2\text{SO}_4$	500 mg
$\text{MgSO}_4 \cdot 7\text{H}_2\text{O}$	100 mg
$\text{FeCl}_3 \cdot 6\text{H}_2\text{O}$	0.5 mg
CaCl_2	7.5 mg
$\text{MnSO}_4 \cdot \text{H}_2\text{O}$	10 mg
KH_2PO_4	527 mg
K_2HPO_4	1070 mg

Organisms

Seed organisms were obtained from the municipal sewage treatment plant at Manhattan, Kansas. Two hundred milliliters of primary clarifier effluent were added to 800 ml of a 3 fold concentration of medium and placed into a glass jar and aerated. Each day, 80% of the culture solution was discarded and replaced by an equal amount of fresh medium. After 3 days, the stock was used as inoculum to the continuous flow system.

Analytical procedures

Cell concentration in terms of mixed liquor suspended solids (MLSS) was determined by filtering a 10 ml sample through 0.45 μ pore size Millipore membrane; the non-volatile solids retained were washed and dried to a constant weight at 105°C.

Dissolved oxygen concentration was measured by a Beckman Field Lab oxygen analyzer unit. A Bausch and Lomb 10 mv strip chart recorder hooked up to the oxygen analyzer was used to continuously record the dissolved oxygen

concentration during the transient period following a step change in aeration rate. The experiment was continued until a new steady state dissolved oxygen concentration was attained.

The concentration vs. time curve obtained was analyzed according to Equation (7). Slopes of the curve at various concentrations, C , were estimated and plotted against the quantity $(C_s - C)$. A straight line passing through the origin resulted; the slope of this line provided the estimation of the oxygen transfer coefficient, $K_L a$.

Results and Discussion

Figure 3 is a plot of dC/dt vs. $(C_s - C)$ at an air flow rate of $16 \text{ ft}^3/\text{hr}$ and system dilution rate $D = 0.303 \text{ hr}^{-1}$ for the Koch mixer column system. Results are shown for measurements made both at the bottom and at the top of the column. The slopes of the resulting lines shown in Figure 3 are those obtained using a least square fit of the data. The local cell concentration at the time of the dissolved oxygen measurement was 650 mg/l at the bottom and 600 mg/l at the top of the column. Similar plots were also constructed for data collected at other flow rates. The results are summarized in Figure 4 where $K_L a$ values are plotted as a function of the air flow rate on a log-log scale. The results show that the $K_L a$ values measured at the bottom of the Koch mixer column are higher than those measured at the top by approximately 15%.

Figure 5 presents plots of dC/dt vs. $(C_s - C)$ at an air flow rate of $16 \text{ ft}^3/\text{hr}$ and $D = 0.149/\text{hr}$ for the sieve tray column system. The local cell concentration at the time of the dissolved oxygen measurements was 470 mg/l at the bottom and 440 mg/l at the top of the sieve tray column. Similar

plots were constructed for data obtained at other air flow rates. The results are summarized in Figure 6. The $K_L a$ values determined at the bottom of the sieve tray column were also found to be approximately 25% higher than those determined at the top.

Higher values of $K_L a$ were observed at the bottom of the column for both the Koch mixer and the sieve tray columns. These results may be due to several factors. The interfacial area, a , may be greater at the bottom of the column due to the presence of surfactants produced by the growing cells. Since these surfactants may later be consumed by the cells, the interfacial area may be influenced more by these substances at the bottom of the column where considerable growth is taking place. In future experiments, measurements of the surface tension and interfacial area at the top and bottom of the column would be desirable.

The effects of chemical reaction on mass transfer [9] and biochemical oxidation on oxygen transfer [10] have been investigated previously. In general, utilization of the substance being transported increases the mass transfer coefficient. Thus, a larger value of $K_L a$ would be expected at the bottom of the column because the rate of oxygen uptake by the cells is larger there. Tsao [11] has discussed the reasons for this enhancement in oxygen transfer rate in considerable detail.

For both the Koch mixer column system and the sieve tray column system, the relationship between the $K_L a$ values and the air flow rates were linear on a log-log scale. For the Koch mixer column system, the $K_L a$ values were found to be proportional to the 0.95 power of the air flow rate, while for the sieve tray column system the $K_L a$ values were proportional to the 0.89 power of the air flow rate.

In Figure 7 the $K_L a$ values obtained at the bottom of each column are compared with those reported by Kitai et al. [1]. Here, the $K_L a$ values are plotted against the air superficial linear velocity based on the empty column. The results show that the $K_L a$ values reported by Kitai et al. depend more strongly upon the air linear velocity than those obtained in this work. However, it should be noted that in Kitai's work oxygen transfer data were obtained by using the sulfite oxidation method. The results also show that at low air linear velocity the $K_L a$ values obtained in this work were much higher than those reported by Kitai et al. For example, at $v_g = 0.35$ cm/sec the $K_L a$ value obtained in this work was higher than 20/hr for either system while that reported by Kitai was less than 10/hr.

The tower systems investigated in this work are capable of providing as good oxygen transfer as the agitated vessel system. Fujita and Hashimoto [12] reported oxygen transfer coefficients for a stirred tank activated sludge system. The experiment was carried out batchwise with respect to the liquid phase. At a MLSS concentration of 848 mg/l and air flow rate of 1.67 vvm the $K_L a$ value was determined to be 155/hr (2.58/min). At this air flow the $K_L a$ value for the Koch mixer column system was found to be 155/hr and for the sieve tray column system, it is 114/hr. So, even without any mechanical agitation, high oxygen transfer rates could be obtained in these tower systems.

The oxygen consumption rate by the organisms, r_X , was calculated by utilizing Equation (5). Since the actual saturation concentration of oxygen in the medium, C^* , was not measured, the saturation concentration of oxygen in tap water at the same temperature was used. If the assumption of constant oxygen consumption by the organisms is correct, the calculated r_X values at

various air flow rates should be similar, because the dissolved oxygen concentration of the medium was always above 2 mg/l. Tables I and II show the rX values, as calculated by using Equation (5), at the bottom and the top of the Koch mixer column system, respectively. The calculated rX values at various air flow rates do show fairly good agreement with each other; variations are within 10% of the mean values.

Similarly, Tables III and IV show the calculated rX values at the bottom and the top of the sieve tray column system, respectively. Again, the calculated rX values are in fairly good agreement with each other.

Values of rX were also measured experimentally by taking samples from the columns and stirring the samples slightly with the oxygen probe. The oxygen concentration change during the first minute was recorded. The slope of this curve gives rX since during the non-gassing period Equation (3) becomes

$$\frac{dC}{dt} = - rX$$

These experimentally determined rX values are compared with the calculated values in Table V.

For both systems, the measured rX values and the calculated rX values, at the bottom of the column, were rather closely matched. However, at the top of the column, they were quite different. The calculated rX values were more than twice as big as the measured values. It should be noted that the measurement of rX by sampling might not describe the true oxygen consumption rate in the column. As the samples were stirred with the oxygen probe, some oxygen was continuously transferred into the media at the surface of the sample.

For samples taken at the top of the column, the rate of change in the oxygen concentration began to decrease within 1 minute. Surface aeration is more significant in samples collected from the top of the column, because the respiration rate (r_X) is smaller. Moreover, sedimentation of the cells is usually more pronounced at the top of the column because of the physiological state of the culture. The differences between the calculated and measured values could also result from a change in respiration rate which might result from carbon substrate transport limitation in the sample. If the culture at the top of the column were growth limited by substrate transport to the cells the change in agitation intensity with sampling could lead to the reduced values of r_X .

Conclusions

The dynamic method proposed in this work provides an easy means of obtaining $K_L a$ for systems where agitation and mixing depend solely on aeration. Direct measurement of the oxygen consumption rate by the organisms, r_X , is difficult using this method because, without aeration, the homogeneity of the phases is not maintained. If the oxygen consumption rate is constant, it does not come into the calculation of $K_L a$, so the inability of measuring r_X does not limit the application of this dynamic technique.

The results of this study show that high oxygen transfer rates can be obtained in tower systems without any mechanical agitation. Since the tower systems depend only upon the aeration to create the necessary mixing, the energy required for driving the agitator is saved. The economy of the tower system was also verified by Kitai et al. [1,2].

The results also indicated that the Koch mixer column is superior to the sieve tray column with respect to oxygen transfer during activated sludge treatment.

For both the systems studied the $K_L a$ values were found to be higher at the bottom of the column than at the top. Some explanations are provided; however, further experimental verification would be desirable.

References

1. A. Kitai, S. Goto, and A. Ozaki, *J. Ferment. Technol.*, 47, 356 (1969)
2. A. Kitai, R. Okamoto, and A. Ozaki, *Ferment. Technol. Today*, Proc. IVth Int. Ferment. Symp., 147 (1972)
3. E.A. Falch and E.L. Gaden, *Biotechnol. Bioeng.*, 11, 927 (1969)
4. C.M. Cooper, G.A. Fermstrom, and S.A. Miller, *Ind. Eng. Chem.*, 36, 504 (1944)
5. B. Bandyopadhyay and A.E. Humphrey, *Biotechnol. Bioeng.*, 9, 533 (1967)
6. D.I.C. Wang and A.E. Humphrey, *Prog. in Ind. Microbiol.*, 8, 1 (1968)
7. M. Fujita and S. Hashimoto, *J. Ferment. Technol.*, 48, 461 (1970)
8. Koch Static Mixer, Bulletin KSM-2, Koch Engineering Co., Wichita, Kansas
9. P.V. Danckwerts, *Gas-Liquid Reactions*, McGraw-Hill, New York (1970)
10. G.T. Tsao, A. Mukerjee, and Y.Y. Lee, *Ferment. Technol. Today*, Proc. IVth Int. Ferment. Symp., 65 (1972)
11. G.T. Tsao, *Biotechnol. Bioeng.*, 10, 765 (1968)
12. M. Fujita and S. Hashimoto, *J. Ferment. Technol.*, 51, 47 (1973)

Table I. Values of rX , as calculated by equation (5),
at the bottom of the Koch Mixer Column.

air flow rate (ft ³ /hr)	rX (mg/l · min)
2	2.49
5	2.30
10	2.14
16	2.16

Table II. Values of rX , as calculated by equation (5),
at the top of the Koch Mixer Column.

air flow rate (ft ³ /hr)	rX (mg/l · min)
2	1.64
5	1.41
16	1.44

Table III. Values of rX , as calculated by equation (5),
at the bottom of the sieve tray column.

air flow rate (ft ³ /hr)	rX (mg/l · min)
2	1.80
5	2.00
10	2.23
16	2.20

Table IV. Values of r_X , as calculated by equation (5),
at the top of the sieve tray column.

air flow rate (ft ³ /hr)	r_X (mg/l · min)
2	0.98
5	0.84
16	0.79

Table V. Comparison of the experimentally determined rX with the calculated values.

		rX (mg/l · min) experimentally measured	rX (mg/l · min) calculated average
Koch mixer column	bottom	2.18	2.27
	top	0.62	1.50
Sieve tray column	bottom	2.0	2.06
	top	0.4	0.87

1. saturation bottle
2. filter
3. rotameter
4. column
5. sieve tray
6. sampling tap
7. sigma motor pump
8. feed tank
9. pressure gauge
10. D.O. probe
11. D.O. meter

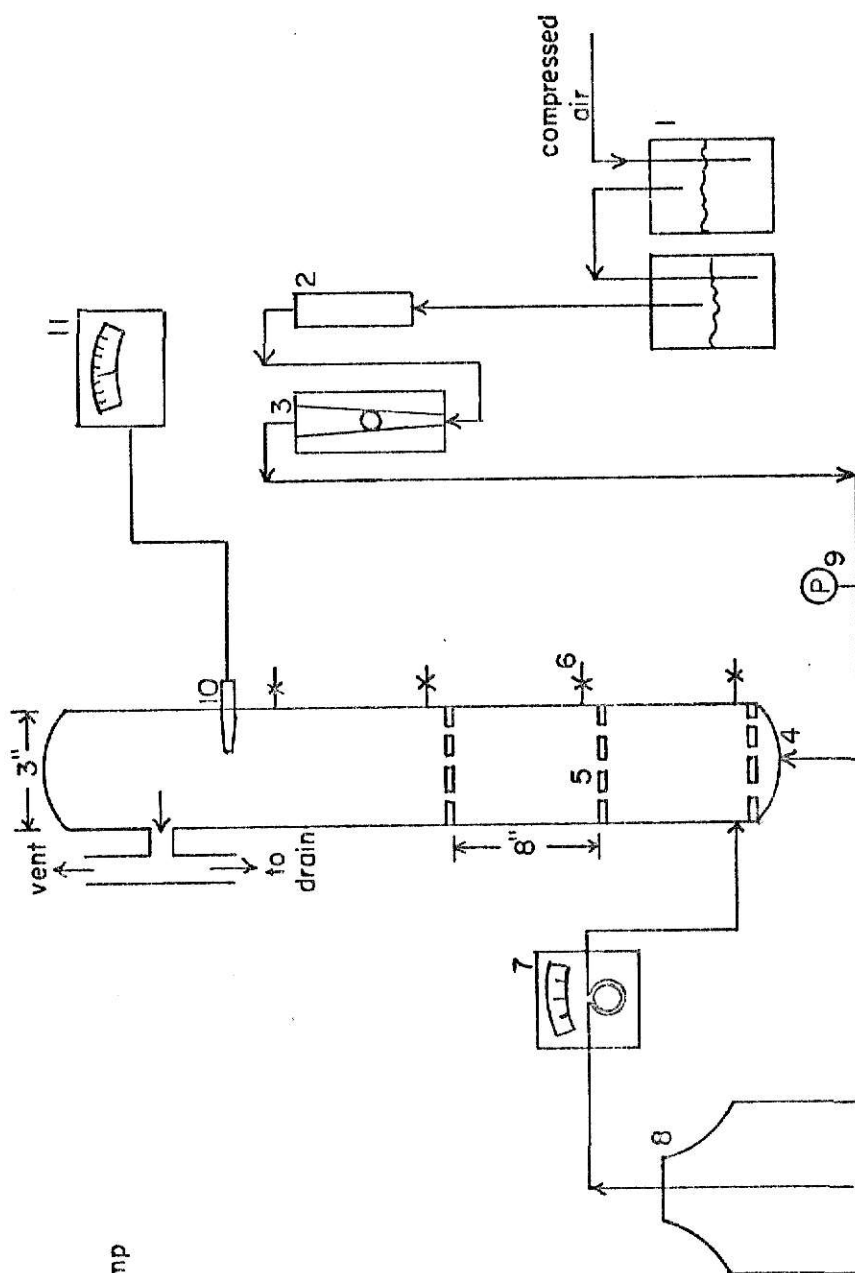


Fig. 1. Schematic diagram of the sieve tray column system.

1. saturation bottle

2. filter

3. rotameter

4. column

5. Koch mixer

6. sampling tap

7. sigma motor pump

8. feed tank

9. pressure guage

10. D. O. probe

11. D. O. meter

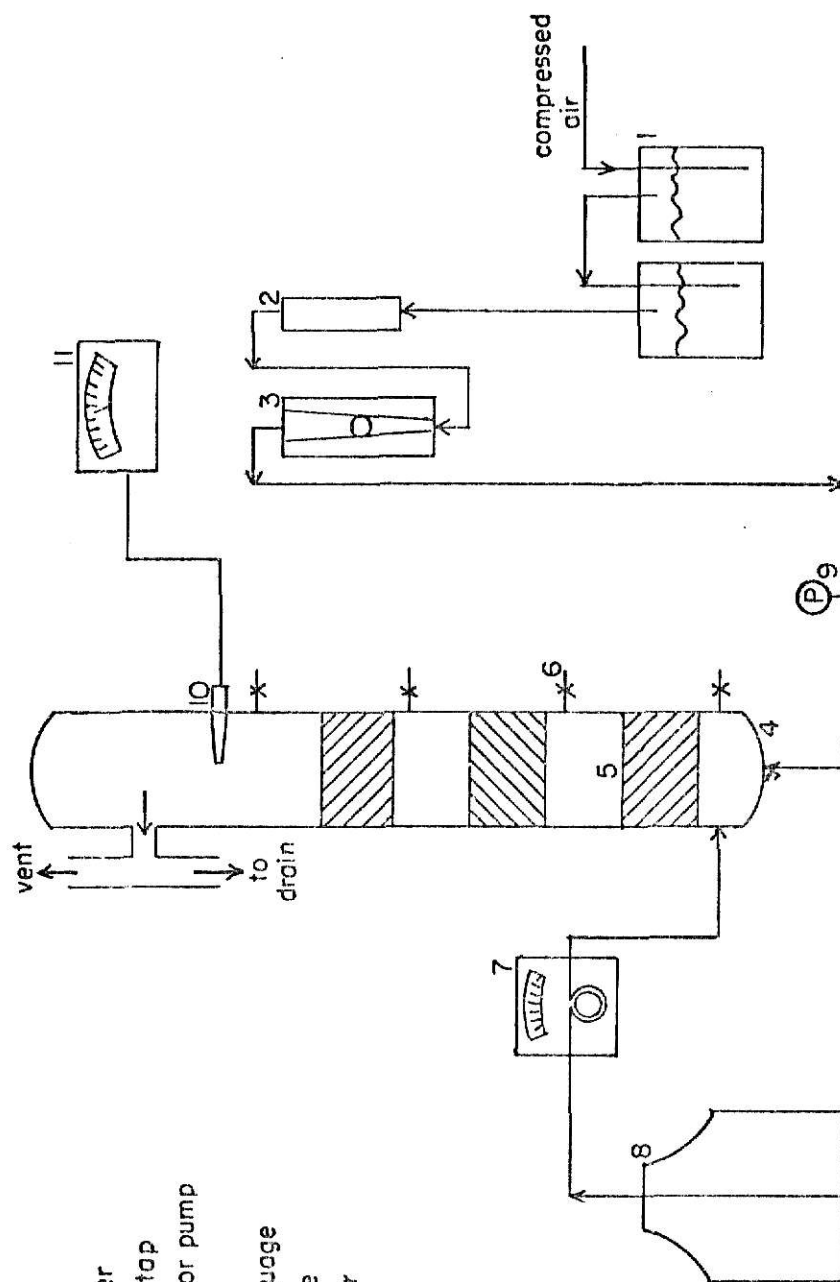


Fig. 2a. Schematic diagram of the Koch mixer column system.

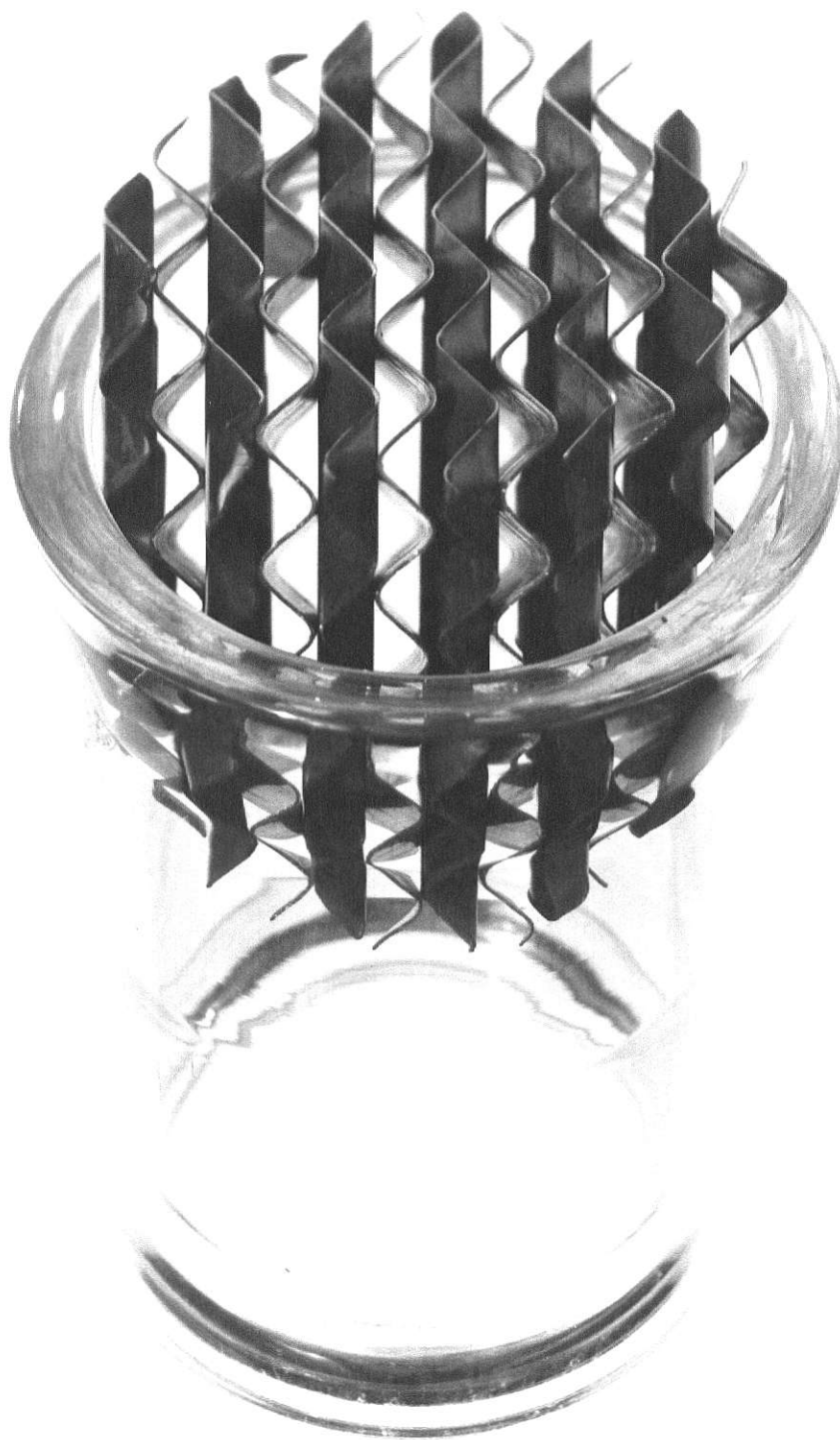


Fig. 2b. Koch mixer

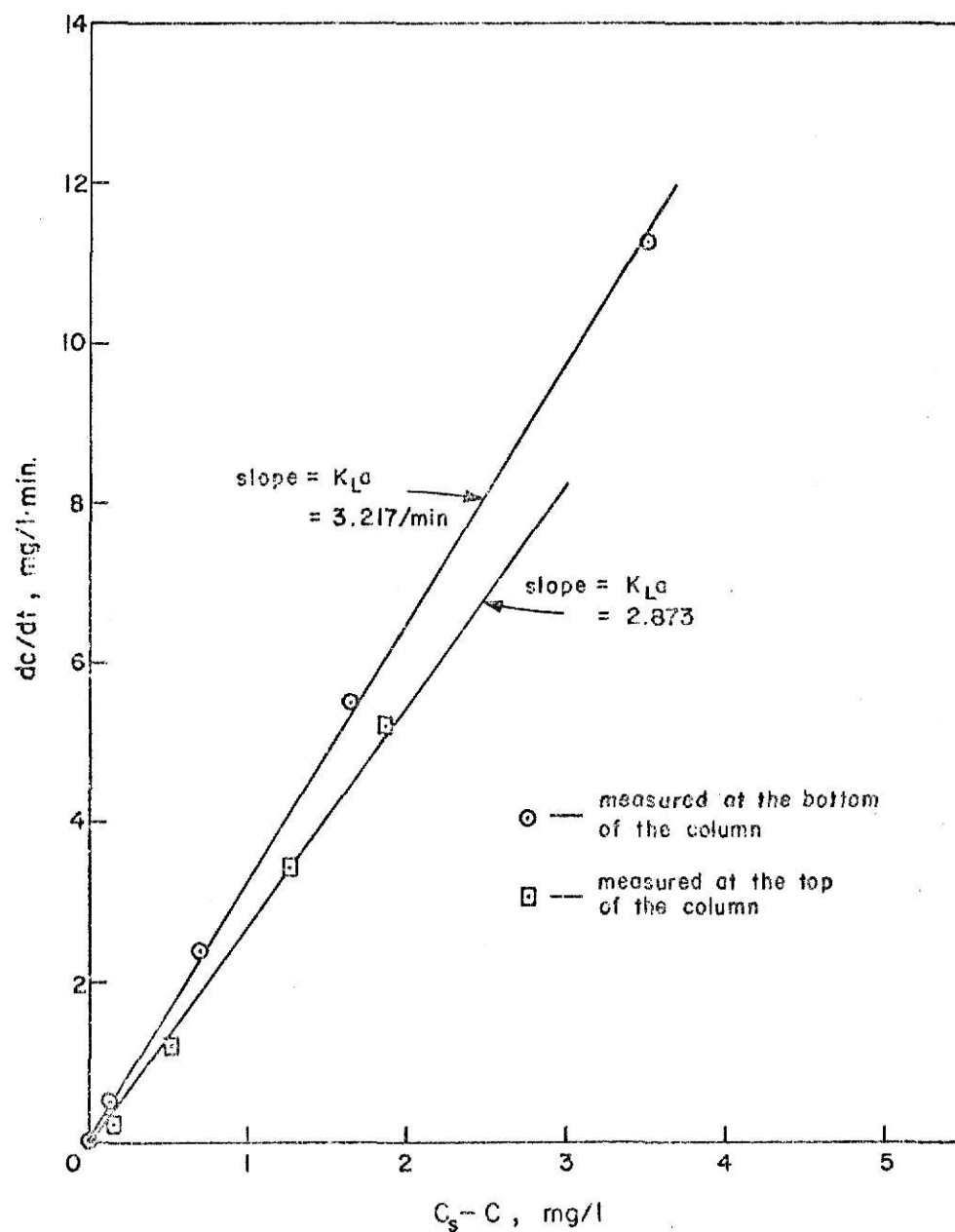


Figure 3. dc/dt vs. $(C_s - C)$ for the Koch mixer column system.
 $D = 0.303 \text{ hr.}^{-1}$, air flow rate = $16 \text{ ft}^3/\text{hr.}$

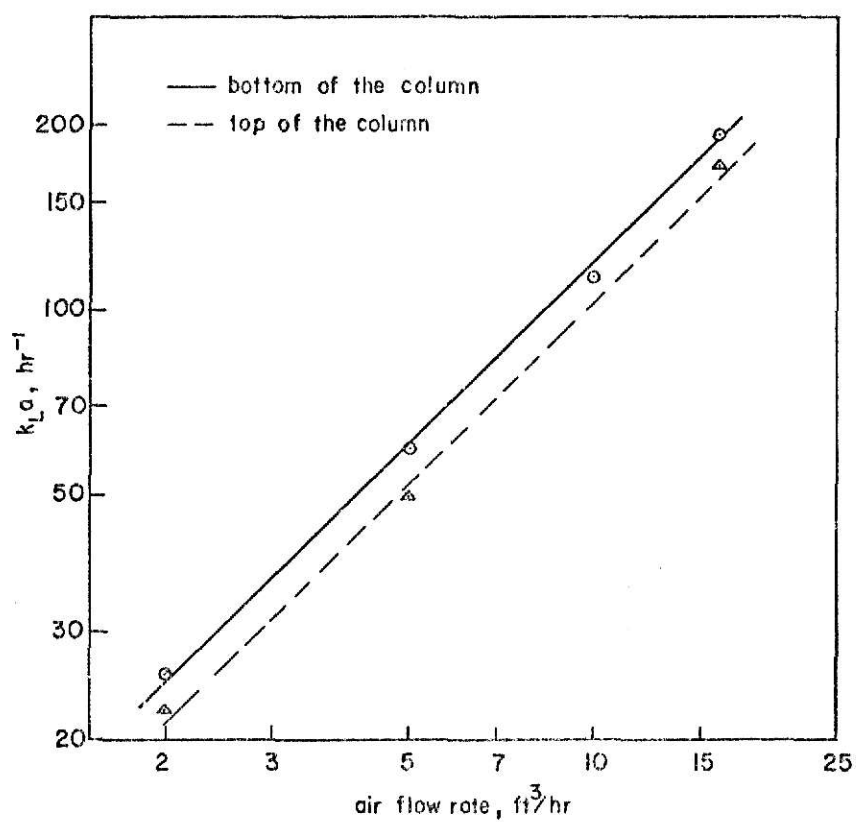


Figure 4 $k_L a$ vs air flow rate for the Koch mixer column system, $D = 0.303 \text{ hr}^{-1}$.

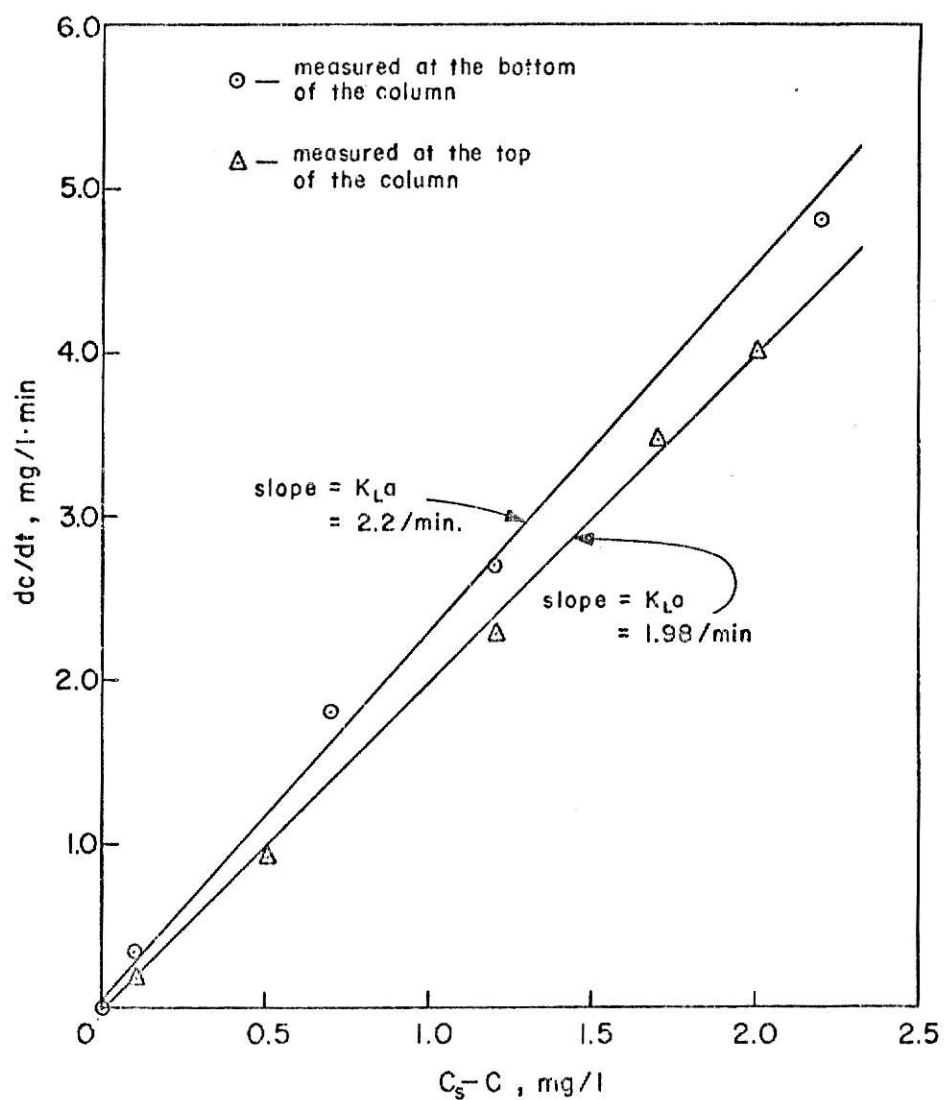


Figure 5. dc/dt vs $(C_s - C)$ for the sieve tray column system.
 $D = 0.149 \text{ hr}^{-1}$, air flow rate = $16 \text{ ft}^3/\text{hr}$.

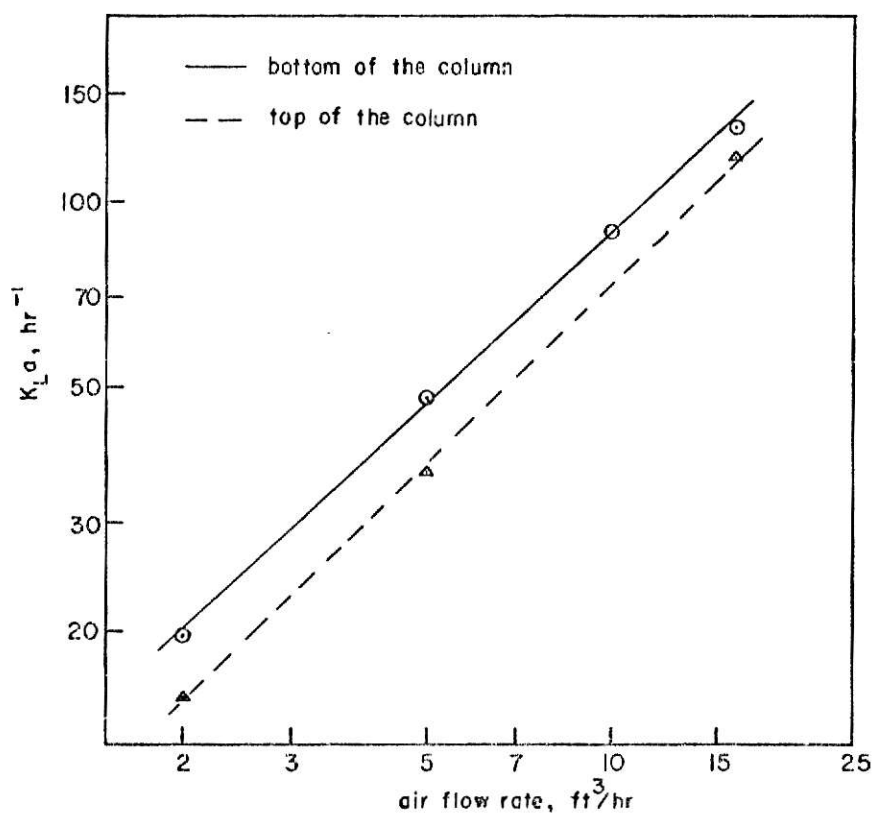


Figure 6 $k_L a$ vs. air flow rate for the sieve tray system, $D = 0.149 \text{ hr}^{-1}$.

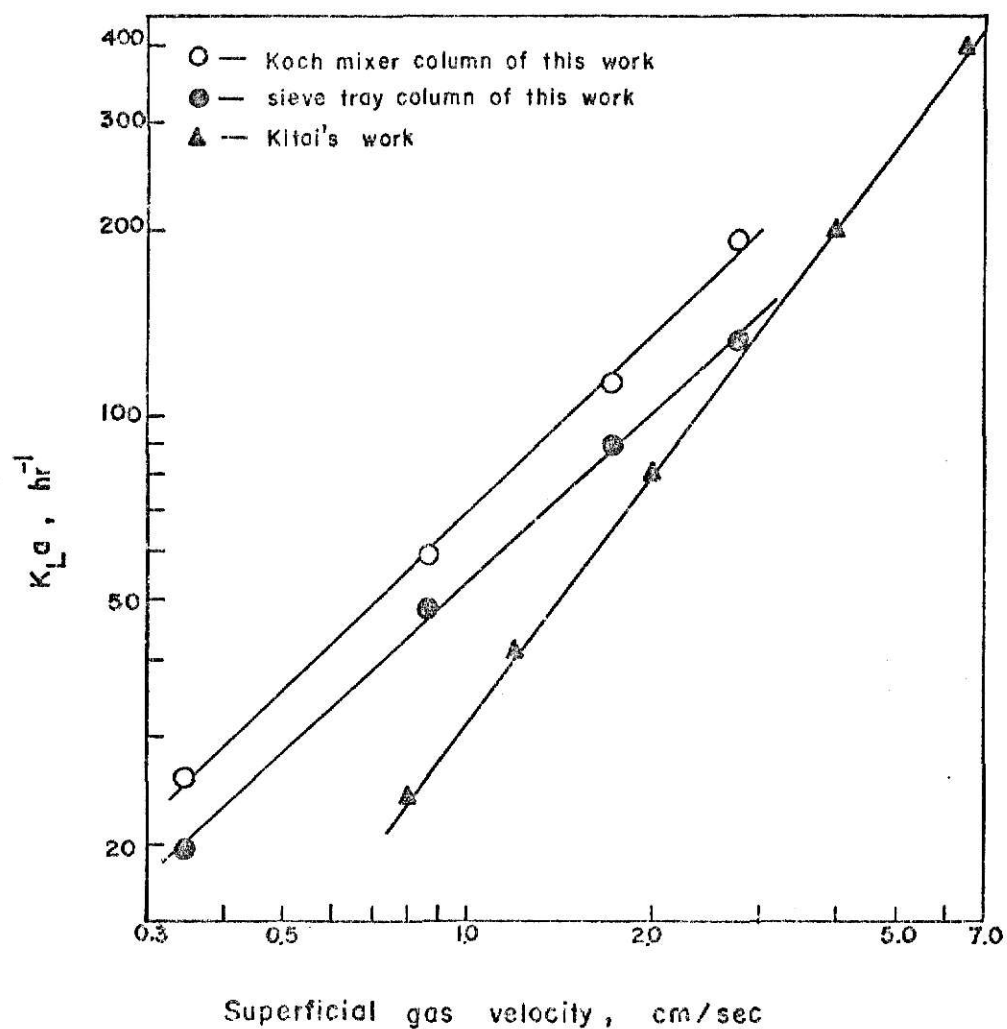


Fig. 7 Comparison of the $K_L a$ values obtained in this work with those reported by Kitai et. al.

CHAPTER V

CONCLUDING REMARKS

Conclusions

For the sieve tray column and the Koch mixer column investigated in this work, the mass transfer performance was found to be superior to that of the conventional bubble column. The total pressure drop in the sieve tray column was greater than that in the conventional bubble column. For the Koch mixer column, the pressure drop due to friction loss was greater than that in the bubble column; however, due to a larger gas hold up in the Koch mixer column, the total pressure drop in the Koch mixer column was smaller.

Although, the sieve tray column and the Koch mixer column were found to be competitive with each other with respect to oxygen transfer, their performance in activated sludge treatment was quite different. The sieve tray column removed the COD of the waste effectively; however, the trays tend to become plugged up. Thus, the operation was less stable. Washout of the sieve tray column occurred at a lower system dilution rate ($D = 0.303 \text{ hr}^{-1}$). The Koch mixer column, on the other hand, removed the COD of the waste effectively with enhanced cell sedimentation and wall growth. Washout of the Koch mixer column occurred at a higher system dilution rate ($D = 0.69 \text{ hr}^{-1}$); also plugging of the mixers was not observed.

The high oxygen transfer rate, low pressure drop, high COD removal efficiency, enhanced cell sedimentation and wall growth, and high washout dilution rate all point toward the Koch mixer column as being superior to the sieve tray column for activated sludge treatment.

A change in dominant species was observed for operation near the washout dilution rate. This phenomena could also be simulated by a mathematical model. The effects of back flow and cell sedimentation on washout were also studied through model simulation. It was found that cell sedimentation has a more significant effect on washout than back flow.

The modified dynamic method proposed in this work for measuring and analyzing oxygen transfer data has proved to be simple and effective. The technique is especially useful for those systems which are not equipped with mechanical agitation devices.

Recommendations

The results of Chapter 2 showed that the sieve tray column and the Koch mixer column each has its own unique gas hold up, pressure drop, and oxygen transfer characteristic. Future works should be directed towards the design of efficient systems. Repeated experiments covering a wider range of gas and liquid flow rates and correlation of the data would be desirable.

In this work, the operation of the column systems was limited to gas-liquid cocurrent flow. Other operations of the systems, such as counter current flow and sludge recycling, are worthy of further study. The results of Chapter 3 suggested that the plugging in the sieve tray column was caused mainly by the transport of cells with the froth. If the sieve tray column were operated counter-currently with downcomers between the stages, then, the liquid level in each stage may be controlled so that the froth does not reach the sieve tray above. Adequate control of the liquid level in each stage may reduce the plugging and pressure drop in the column substantially.

The concept of recycling for a continuous flow system is an important one; this is especially true in biological waste treatment systems. Sludge recycle may improve the waste treatment system in two ways: first, it can raise the cell concentration in the system to a higher level so that the system is capable of handling larger COD or BOD loads; secondly, it can prevent the system from washing out easily, thus, it increases the operation range.

An analysis of flow characteristics of the cells and medium in the column, and an analysis of process economics would also be desirable.

In Chapter 4, values of K_L determined at the bottom of the column were found to be greater than those obtained at the top. Further studies should be directed toward understanding this result. Repeated oxygen transfer studies accompanied by measurements of surface tension, viscosity, and gas-liquid interfacial area at the top and bottom of the column would be most helpful.

ACKNOWLEDGMENT

The author wishes to express his sincere appreciation to his committee members for their support, to Dr. Larry E. Erickson for his constant guidance, and to Dr. Liang Tseng Fan for his encouragement.

This work was supported in part by the National Science Foundation (Grant GK-34693) and the Koch Engineering Company.

GROWTH OF MIXED CULTURES AND OXYGEN TRANSFER IN TOWER SYSTEMS
WITH MOTIONLESS MIXERS

by

KENNETH H. HSU

B. S., Kansas State University, 1972

AN ABSTRACT OF A MASTER'S THESIS

submitted in partial fulfillment of the

requirements for the degree

MASTER OF SCIENCE

Department of Chemical Engineering

KANSAS STATE UNIVERSITY
Manhattan, Kansas

1974

A sieve tray column and a Koch mixer column were investigated for their application in activated sludge treatment. Gas hold up, pressure drop, and oxygen transfer were experimentally measured for cocurrent flow of water and oxygen in these two columns and an equivalent conventional bubble column. The mass transfer performance of the two columns studied was found to be superior to that of the conventional bubble column.

The biological waste treatment characteristics of the columns were investigated using a synthetic waste. The mixed liquor suspended solids (MLSS) and dissolved chemical oxygen demand (COD) concentrations in the column at various system dilution rates were experimentally measured. A study was also made of the effects of wall growth, cell sedimentation, and plugging on treatment performance of the two columns. Mathematical models were employed to simulate some of the experimental conditions. The results of the simulation were found to compare favorably with the experimental results.

Local overall mass transfer coefficients during biological treatment were experimentally measured using a modified dynamic method. The values of the local overall transfer coefficient were found to be higher at the bottom of the column than at the top. The values of the local overall transfer coefficient obtained in the Koch mixer column were found to be higher than those obtained in the sieve tray column.